INNOVATIONS IN GOLD EXTRACTION

THESIS

Submitted in Fulfilment of the
Requirements for
the degree of
DOCTOR OF SCIENCE
of Rhodes University

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December 1992

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FOREWORD

This thesis is submitted in accordance with the regulations for the Degree of Doctor of Science of Rhodes University. It has not been submitted in whole, or in part, for any degree at any other university. The thesis takes the form of a collection of publications and patents concerning gold extraction which were presented during the past 18 years while employed at the Anglo American Research Laboratories.

Acknowledgements are gratefully made to the management and senior metallurgical staff of the Anglo American Research Laboratories, the Gold and Uranium Division of Anglo American Corporation of South Africa Limited, the many operating gold mines where pilot and full scale testwork was carried out. Mention must also be made of contributions made by The Council for Mineral Technology, General Mining Metals and Minerals Limited and Delkor Technik (Pty) Limited without whose support some of the significant improvements to CIP plant operations would not have been possible.

CURRICULUM VITAE

1. Following schooling at Kingswood College, Grahamstown, entered Rhodes university in 1951 majoring in Chemistry and Geology. Graduated after 3 years with B Sc.

2. Student field geologist in Zambia for 3 months in 1953 employed by Rhodesian Selection Trust.

3. Employed by Goldfields of SA as pupil metallurgist at the Vogelstruisbult uranium plant for 6 months in 1954.

4. Graduated from Rhodes University with BSc Hons in chemistry in 1955. Represented the university at both rugby and squash.

5. Employed by Rhokana Corp. as an assay chemist in Kitwe (Zambia) for 6 months in 1956.

6. Returned to Rhodes University in 1957 to undertake an MSc degree in chemistry. In 1958 obtained MSc on thesis on the determination of boron in plant material.

7. 1959-1968. Employed at Leather Research Institute (Rhodes University) as a research officer. Initially worked on aspects related to the chrome tanning and impregnation of freeze dried raw hide (1959-1962). 1963-1968 undertook investigation into the irradiation of soluble collagen proteins with X-rays and γ rays, for which a PhD was awarded. A total of 11 papers were published in overseas journals.

8. 1968 - present time. Employed at the Anglo American Research Laboratories as a research metallurgist in an R and D section. The initial 4 years were involved in investigations into the segregation of low grade copper and nickel oxide ores. The segregation process involved the reaction of either copper or nickel oxide ores at temperatures in the range 800-1100°C in the presence of halide salts under reducing conditions. During this period, 7 patents on various related aspects of the process were taken out, five of which are in force at the present time. One paper on the effect of iron oxides on nickel segregation was published.

   During the remaining period, investigations related to the extraction and recovery of gold were carried out. Promotion to head of the Extraction Metallurgy Section (Principal Metallurgist) at the laboratories (1974) involved directing the programmes of some ten graduate research workers. Both the intensive cyanidation of gold plant concentrates as well as the adsorption of gold onto activated carbon were studied in considerable depth.

   1978 - Promoted to Research Fellow (Divisional Metallurgist).

   1985 - Promoted to Senior Research Fellow (Senior Divisional Metallurgist).

   In all some 29 papers and 25 patents, many related to gold recovery, were published. The more notable patents, and innovations, several of which are in use in the gold industry, include the AARL elution process, the intensive cyanidation of gravity concentrates, the continuous elution of loaded carbon, the pinned bed process, the counter current moving bed and the AAC-Delkor linear screen.

R J Davidson

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   R.J. Davidson
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R.J. Davidson
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R.J. Davidson, W.D. Douglas & J.A. Tumilty
CIM XIV Int. Min. Proc. Congr. (1982) II-6.1 - II-6.19

20. The Recovery of Gold from Plant Effluent by the use of Activated Carbon

R.J. Davidson & B. Strong
J.S. Afr. Inst. Min. Metall (1983) 181-188.

21. Properties of Carbon for Gold Adsorption and their Measurement

R.J. Davidson
S. Afr. Inst. Min. Metall. Carbon School (1985) 10.1-10.21.

22. The Recovery of Gold from Plant Solutions by use of a Countercurrent Moving Carbon Bed

R.J. Davidson, W.D. Douglas & J.A. Tumilty
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13th CMMI Congress - Metallurgy (1986), Singapore, 215-225.

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R.J. Davidson
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J. Moore, P. Smith & R.J. Davidson
S.A. Mech Engineer, Vol. 37, (1987), 150-157.

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R.M. Whyte, P. Smith, D.J. Morris & R.J. Davidson
Soc. of Chem. Ind., Hydrometallurgy 87, The 3rd Int. Symp., (1987) Ch. 30, 298-318.

27. A pre-burial Adsorption Model for the Genesis of Gold in the Witwatersrand.

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28. Carbon Management.

R.J. Davidson & N. Schoeman.
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29. Continuous Elution of Gold from Activated Carbon.

R.J. Davidson, F.J.E. Scholtz, L.P.H. Greenhalgh & R.P. Riley.
Randol Gold Forum 90, Randol International, Sept. 1990, Colorado, U.S.

Note: The candidate was the major author in all of the above publications but for publication Nos. 4, 5, 7, 8, 12, 25 & 26.

ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

MEMORANDUM FOR DR R.J. DAVIDSON

INVENTIONS IN THE NAME OF DR R.J. DAVIDSON

... As requested, I have now updated the list of patents/patent applications for which you are mentioned as an inventor or co-inventor.

I confirm that this schedule is up to date as of today's date.

If you require any further assistance, please do not hesitate to contact me.

CRAIG R. OPPERMAN
DIVISIONAL PATENTS ADVISER

JOHANNESBURG
Our Ref: Mr Opperman/cz
PAT 8006

13 March 1991

0035B/21

SCHEDULE OF PATENTS/PATENT APPLICATIONS
WITH R.J. DAVIDSON CITED AS INVENTOR/CO-INVENTOR

S.A. PAT. NUMBER	TITLE	FOREIGN EQUIVALENT	NUMBER
68/6879	Chlorodization reagent	-	-
69/3158	Improvement in the segregation	Boph.	69/3158
69/3159	Improvements in the segregation process	-	-
70/2937 (cognating 70/6384)	Improvements in the segregation process	Boph.	70/2937
70/5890	Segregation process	Boph.	70/5890
		Canada	954 307
		Chile	29 733
		France	71/39906
		OAMPI	3846
		Phillip.	8812
		USA	195 008
70/8391	Beneficiation of ores	Australia	36679/71
		Boph.	70/8391
		Brazil	PI710822
		Canada	956119
		Chile	28042
		Colombo	133 800
		France	71/44481
		Greece	44509
		Guatemala	-
		India	133 912
		Indonesia	3672
		Japan	100588/71
		OAMPI	3934
		Phillip.	10615
		Transkei	70/8391
		UK	1375006
		USA	3914123
		Yugoslavia	P3097/71
		Zaire	1333A
		Zambia	183/71/156
		Zimbabwe	489/71

2./...

S.A. PAT. NUMBER	TITLE	FOREIGN EQUIVALENT	NUMBER
73/8939 (cognating 74/1182)	Gold recovery	Australia	482773
		Boph.	73/8939
		Brazil	7409844
		Canada	1021573
		France	74/38545
		Germany	2455301
		India	140947
		Israel	45955
		Italy	1025713
		Japan	976738
		Switzerl.	611171
		Transkei	73/8939
		UK	1450486
		USA	3970737
		USSR	2079146/22
74/4204	Recovery of metal values	Australia	508816
		Boph.	76/4204
		Brazil	PI7704649
		Canada	1097508
		France	7721323
		Germany	P27310298
		India	147 865
		Israel	52438
		Italy	25766A/77
		Japan	11102/78
		Switzerl.	631488
		Transkei	76/4204
		UK	1573685
		USA	811117
		USSR	2505302/02
77/0995	Uranium extraction	Argentina	271 363
		Australia	513 412
		Boph.	77/0995
		Brazil	PI7801884
		Canada	1093831
		Iran	19411
		Spain	467780
		SWA	78/0015
		Sweden	78018546
		USA	878809
77/1777	Apparatus for carrying out flotation and/or leaching processes	Boph.	77/1777

3./...

S.A. PAT. NUMBER	TITLE	FOREIGN EQUIVALENT	NUMBER
77/1778	Gold and uranium extraction	Boph. Brazil	77/1778 PI7802430
77/4285	Activation of carbon	Boph.	77/4285
77/4287	Regeneration of used carbon	Boph.	77/4287
78/0688	Gold extraction	Australia Boph. Canada SWA Spain USA USSR Zimbabwe	518768 79/0018 1107076 79/0016 477505 008965 2730253/02 31/79
78/5147	The recovery of gold and silver values	Australia Boph. Brazil Canada India Swaziland USA USSR Zimbabwe	521281 79/0139 PI7905698 1125033 151 975 RP/10/1980 157 345 2814349/02 174/79
78/5463	Metal recovery	Australia Brazil Canada Europe (France, Germany, Italy, Switz, UK) India Israel Japan USA USSR	527 451 PI7906146 1125519 0010367 153180 58246 44600/80 076024 2818699/02
78/5464	Noble metal recovery	-	-
78/5864	Regeneration of activated carbon	Australia Boph. Botswana Brazil Canada Europe (France, Germany, Switz, UK)	524296 79/0150 P340 PI7906677 1125729 0010381

4./...

S.A. PAT. NUMBER	TITLE	FOREIGN EQUIVALENT	NUMBER
	Regeneration of activated carbon (continued)		
		Israel	58329
		Japan	100213/80
		Lesotho	270
		SWA	79/0120
		Swaziland	RP5/1980
		Transkei	79/0124
		USA	4267069
		USSR	904506
		Zimbabwe	210/79
83/3952	Gold adsorption	Australia	15350/83
		Brazil	PI8303528
		Canada	429 573
		USA	500 094
		Zimbabwe	124/83
83/8659	Recovery of metal values from a pulp containing such values in dissolved form	Australia	21645/83
		Boph.	83/0185
		Botswana	P466
		Brazil	PI8306618
		Canada	442 399
		Chile	830/83
		Finland	834 399
		N. Zealand	206 461
		Phillip.	29 913
		Spain	527 716
		SWA	83/0110
		Sweden	8306607-7
		USA	556132
		USSR	3691036.03
		Zambia	84/83
85/2404	Metal recovery	Australia	40415/85
		Brazil	PI8301425
		Canada	477793
		USA	715 504
		Zimbabwe	59/85

5./..

S.A. PAT. NUMBER	TITLE	FOREIGN EQUIVALENT	NUMBER
85/4746	Wood Filter	Australia	44206/85
		Boph.	85/0088
		Brazil	PI8503104
		Canada	485636
		Chile	449/85
		Europe	85304549
		Finland	852530
		Portugal	80730
		Spain	544590
		SWA	85/048
		USA	4692240
		Zambia	47/85
		Zimbabwe	104/85
90/8081	Method for the recovery of metal values		

Note: The candidate was the major inventor in all of the above patents.

CHAPTER I

GOLD AND SILVER RECOVERY USING ACTIVATED CARBON

1.0 INTRODUCTION

With the advent of a higher gold price in the 1970's, coupled with a need for containing escalating capital and operating costs in the South African gold mining industry, considerable attention has been focused on the carbon in pulp (CIP) process as a means of recovering gold. Exciting developments in this regard are currently taking place, with the process gaining wide acceptance and becoming firmly established as an alternative to the filtration-Merrill-Crowe process. While the programme of testwork that preceded this acceptance of the CIP process in South Africa was conducted over some eight years, the process itself has undergone a considerable metamorphosis in a remarkably short period of time and is likely to change even more so in the foreseeable future. During this phase of development a better understanding of both the metallurgical and the engineering aspects of the process have become apparent although such an understanding is far from complete due primarily to the complex chemistry of the system.

Metallurgical interest in activated carbon as a means of recovering gold from cyanide leach liquors dates back to the Yuanmi Mine in Western Austrtalia in 1916, where wood charcoal displaced zinc as a precipitant for gold. Since the powdered charcoal was prepared from wood in rather a crude fashion, very low gold loadings were achieved as this product could not have developed the large surface area typical of modern activated carbons. Another serious disadvantage of this earlier process was the lack of a suitable desorption procedure which necessitated the ashing of the loaded charcoal. Furthermore, as the technology of zinc cementation improved, so the interest in carbon for gold recovery declined. During this early period a detailed investigation into the fundamental aspects of gold adsorption on activated carbon was undertaken at the U S Bureau of Mines. This investigation also included a comprehensive bibliography of early publications and patents.

It was not until the 1950's that further interest in activated carbon as an adsorbent for gold was aroused. Successful investigations into gold elution and electrowinning at the US Bureau of Mines by Zadra et al provided much of the technology on which the CIP process is based, and it was this technology which culminated in the installation in 1973 of the first large-scale CIP plant by the Homestake Mining Company in the USA. The process involves the direct addition of granular carbon to a cyanided pulp in a four stage contacting circuit in which carbon is moved counter current to the flow of pulp. Gold loaded carbon from this circuit is stripped using a simultaneous elution and electrowinning procedure, and the carbon returned to the adsorption circuit following thermal regeneration. This plant has operated successfully from its inception and has effectively demonstrated the viability of the CIP process at a commercial level.

In South Africa, exciting developments on the use of activated carbon for gold extraction have been, and are currently taking place. Interest was initiated in 1972 at the Anglo American Research Laboratories (AARL) and later the National Institute for Metallurgy in 1974. During this early phase of development, a better understanding of both the metallurgical and engineering aspects of the process became apparent.

2.

Following several successful piloting operations notably at Durban Roodepoort Deep, Fairview and Grootvlei a 250 t/d plant was erected at Golden Dumps by a private operator to treat reclaimed rock and sand dump material.

NIM and AARL assisted in the design and commissioning of this plant which paved the way for the 30 large and 40 small CIP operations currently operating in South Africa. Much credit for the rapid development of this technology must be given to the NIM (now Mintek) for the sustained effort which this organisation maintained during this early development.

As may be expected, due to such rapid development, CIP technology is very much still in a state of flux. Many improvements in interstage carbon screens have taken place. Following the successful introduction of airswept and equalised pressure screens developed by Mintek, recent innovations of note relate to the use of improved electrowinning cells, and here the Mintek type cell has gained wide acceptance. With regard to feed preparation, the presence of wood chips and wood fibre originating from mining operations presented unforeseen problems in CIP operations. While the use of vibrating screens and the recirculating Mintek tank may have alleviated such problems considerably, the recent development of the AAC-Delkor linear screen has gone a long way towards solving these problems.

No review on the application of activated carbon in the gold mining industry would be complete without commenting on the treatment of gold bearing effluents and solution in so-called CIS circuits. During the last decade, many such CIS circuits have successfully treated dump leach liquors, U-plant primary filtrates, gold plant filtrates and CCD solutions. Even the economical treatment of dam return waters containing as little as 0,02 g/t gold has been demonstrated with carbon loadings approaching 1 kg/t which may be equated with a concentration factor of some 50 000:1 ! Once again the CIS circuits currently in use manifest a wide range of contacting columns. The use of upflow expanded bed carousel circuits, downflow packed columns and even moving bed columns have all demonstrated a high degree of efficiency in recovering gold from such liquors.

What of the future with regard to the use of activated carbon in the gold mining industry ? Seen in the light of the rapid development over the last decade, many more exciting developments may be expected in the near future. Very much smaller, high carbon concentration carousel plants may well develop now that more efficient interstage screens are available. The use of coarser carbons and even extruded carbons would improve this possibility considerably. Even the pinned bed concept with the entire carbon inventory packed in a single carbon column may yet produce the next revolutionary development. Similar thoughts may be applied to both carbon elution and to gold electrowinning. While both elution and electrowinning operations are currently batch type systems, both circuits are amenable to more efficient continuous counter current modifications.

The high degree of metallurgical efficiency and lower gold lock-up demonstrated by many CIS circuits would also indicate more of such circuits being used on existing gold plants. Together with improved thickening technology, many effluent and even pregnant streams may be amenable to carbon treatment. A judicious balance between CIS and CIP circuits would appear to be necessary in optimising overall plant performance.

3.

Investigations related to the use of activated carbon were initiated at the AARL by the author in 1972. Numerous aspects related to this subject have since been studied which will be broadly outlined below. The co-operation and contribution made by co-workers and colleagues in these studies is gratefully acknowledged.

1.1 REVIEW ON GOLD ADSORPTION

A review on the mechanism of gold adsorption and the physical and chemical properties affecting adsorption is presented. This review is the author's contribution to Chapter 9 - The Extractive Metallurgy of Gold in South Africa, edited by G G Stanley and published by The South African Institute of Mining and Metallurgy (SAIMM), Johannesburg 1987.

ACTIVATED CARBON

"Charcoal is a strange and interesting substance. The fact that it has the power to abstract gold and silver from cyanide solutions to the extent of 7 per cent of its weight in gold and 3 per cent of its weight in silver without showing the slightest change in appearance, even under the microscope, clothes it with a mystery that has long interested metallurgists. Even now, with the mass of data made available by various investigators, much remains to be learned". This statement[1] was made in 1927 but may well apply to the present day due to the complex physical and chemical character of activated carbon. In spite of the importance of the carbon-in-pulp process to the gold mining industry, no consensus has been reached regarding the mechanism by which gold (silver) is adsorbed onto activated carbon. Nevertheless, in order to optimise the metallurgical performances of a process using activated carbon for gold or silver recovery, some understanding of the involved and sometimes rather complex chemistry is most necessary. This philosophy will apply not only to the adsorption and elution processes, but also to the effective regeneration of the carbon for repeated use.

In many ways activated carbon is a unique material with properties not possessed by any other class of organic substance. Since activated carbon usually contains considerable quantities of adsorbed oxygen and hydrogen, the term carbon is perhaps a misnomer. It is essential to regard activated carbon as a complex organic polymer and not merely as an amorphous form of the element carbon. Activated carbon however differs from other organic polymers in having an exceptionally large surface area and in being a good conductor of electricity. Many of its intriguing properties stem from the latter fact, as well as being related to the various functional groups present on the carbon surface. A better understanding of both the physical and the chemical characteristics of activated carbon is thus a prerequisite to understanding the factors influencing gold adsorption as well as in formulating any adsorption mechanism.

Physical and chemical properties of activated carbon

Activated carbon may be made by heating virtually any carbonaceous material in a mildly oxidizing atmosphere. The physical and chemical properties of the activated carbon are influenced to a large degree by both the nature of the source material and the conditions of activation. During the course of activation the carbon develops a porous graphitic structure of molecular dimensions with an extraordinarily large internal surface area on which adsorption may take place. The quantity, size and shape of these pores distinguishes the various types of activated carbons together with their subsequent application. In general, raw materials with a dense cellular structure produce a hard brittle product and it is for this reason that carbons made from coconut shell are used almost exclusively in the gold mining industry.

The pore structure and surface area of carbon are developed during the charring and activation process. The sizes of the pores are important, because they may exert a screening influence, which prevents molecules from being absorbed, or else promote adsorption when the pore diameters are of optimal size. For example, the small pores of coconut-shell carbons are ideal for adsorbing small gaseous molecules, which fit more snugly than into the larger pores of other types of carbon. This has led to the use of adsorption of molecules of different sizes by activated carbons as an indication of the distribution of their internal free volume among pores of different sizes. Thus, the small iodine molecule is adsorbed in pores down to 10 A diameter, so that the 'iodine number' is an indication of the number of pores above this size, while, at the other end of the scale, molasses is adsorbed only in pores above about 30A, and the 'molasses number' gives an indication of the number of pores larger than this. In cross-section the pores in activated carbons may be cylindrical or rectangular, or a variety of irregular shapes, and constrictions may occur. These pores are generally classified in terms of their equivalent diameters, viz,

macropores	500 - 20 000 A
transitional pores or mesapores	100 - 500 A
micropores	8 - 100 A

The most important division of carbons is into what are called H- and L-carbons[2]. H-carbons are formed at temperatures in excess of 700°C (typically around 1 000°C) and are characterised by taking up acid on immersion in water. L-carbons are activated well below 700°C, typically around 300 to 400°C, and are characterised by taking up base on immersion in water. An important aspect of the production of both types is the exposure to oxygen, which is irreversibly adsorbed by the carbon, and only comes off again at high temperatures as CO and CO_2 showing that it is present on the carbon surface as a 'surface oxide' of the carbon.

Although these surface oxides play an important role in the chemical nature of the carbon, their identity can by no means be regarded as well established. Because the respective temperatures, at which the gases are evolved upon outgassing samples of carbon under vacuum, correspond with the temperature ranges involved in the production of H- and L-carbons, it has been suggested[3,4] that the surface oxides evolved as CO_2 at lower temperature are responsible for the physico-chemical behaviour of L-carbons, and those evolved as CO at higher temperatures are responsible for H-carbon behaviour. Attempts to identify these surface oxides[5-7] are rather complex and specialised, suffice to say that the low-temperature oxides appear to be carboxylate in nature, whereas those formed at higher temperatures appear to be phenolic, or some derivative of this group, such as a lactone. The importance of these groups probably lies in their ability to alter the hydrophobic versus hydrophilic properties of the carbon surface.

The activation of carbon usually involves two distinct stages, viz, a preliminary charring step at lower temperatures (300 to 600°C), which drives off the volatile matter of the raw material to leave a hydrophobic carbon skeleton, followed by heating of the char at higher temperatures (700 to 1 000°C) in the presence of air, carbon dioxide, or water vapour (or mixtures of these). This is the activation stage, and the reactive oxygen burns away part of the carbon skeleton as carbon monoxide and carbon dioxide, thereby increasing the internal surface area of the carbon and developing the pore structure. If oxygen is excluded, the resulting carbons have no ability to adsorb acids.

Apart from the strong interactions involved in forming chemical bonds between atoms, molecules exert weak attractive and repulsive forces on each other, which differ from those in chemical bonding only in that they are much weaker. These weak van der Waals forces are responsible for the adsorption of non-polar molecules onto surfaces, and are of the utmost importance in the adsorption of organics by activated carbon.

In a like fashion, more polar molecules like water are attracted to the carbon surface via hydrogen bonding thus rendering the surface hydrophilic. Both of these forces are involved in the adsorption of gases onto carbons, while adsorption from aqueous solution may be simply understood as a balance between the much stronger hydrogen-bonding forces and the weaker van der Waals forces. The lattice of liquid water is held together by hydrogen bonds, and non-polar organic molecules generally have only a low solubility in water because the water molecules form strong hydrogen bonds with each other rather than weak van der Waals interactions with the non-polar organic molecule, or a hydrophobic surface incapable of forming hydrogen bonds. It is thus energetically favourable for the organic molecule to escape from the water structure and adsorb onto a hydrophobic surface. However, if the surface is made hydrophilic, the water will adsorb preferentially and the organic molecule will remain in the bulk liquid. Should the organic molecule have a hydrophilic group, e.g. a carboxylate as well as a hydrophobic chain, the best arrangement then will be with the hydrophobic portion attached to the hydrophobic surface, with the hydrophilic end oriented into the bulk aqueous solution.

The above factors lead to a fairly simple behaviour of organic molecules adsorbing onto activated carbons. In an homologous series of organic molecules, e.g. methyl, ethyl, propyl alcohols, the adsorption becomes stronger as more methylene groups are added because the hydrophobic part of the molecule becomes larger, lowering its affinity for the solvent water, and increasing the extent of van der Waals interactions with the hydrophobic surface of the carbon. This gives rise to Traube's rule, which states that the larger homologues in a series will be the most strongly adsorbed. At the same time, because ions are very strongly hydrated (i.e., hydrophilic), activated carbon has, in general, only a small affinity for ions. Because of this, it is probable that theories of adsorption of ions such as $Au(CN)_2^-$, as ions per se, can be largely discounted. The key to understanding the properties of activated carbon lies in the solubility of the adsorbed species.

The weak adsorption of acids that does take place is postulated to take place via specific adsorption of anions into the electrical double layer, with non-specific adsorption of cations such as the proton or Na^+ so as to preserve electroneutrality. It is postulated[7] that the anions are in the innermost, specifically adsorbed layer, because the adsorption of ionic substances is more sensitive to the nature of the anion than of the cation in these substances.

On the freshly activated carbon, stronger take-up of acid occurs, which is not anion-dependent. It has been proposed[8] that oxygen reacts with the freshly activated surface to form surface complexes of the type C_xO, which then react with water according to equation (1):

$$C_xO + H_2O \rightleftharpoons C_x^{2+} + 2OH^- . \qquad (1)$$

The net effect of this reaction is the removal of protons from the bulk aqueous solution. The formulation C_xO indicates that the exact nature of these groups remains unknown. Garten et al[9] proposed that these groups were chromene groups. This would account for the appearance of peroxide ions after reaction with oxygen, and also a very important property of the carbon, namely its reducing ability. It is also possible that phenolate groups are involved in a quinone-hydroquinone type of redox couple, or that the graphitic structure of the carbon may act as an 'electron sink' and that electrons are delocalized over this aromatic structure and are available for reducing purposes.

Another important property of carbon is its catalytic ability. This derives from adsorption of the reactants onto the surface, where they react more readily than when in the bulk aqueous solution. This catalytic activity of carbon is caused by a variety of factors. The process of adsorption may activate one of the reactants owing to electronic interaction with the surface, which alters the electron distribution within the molecule, or orients it correctly for reaction. Most important, in practice, is that activated carbon can act even as an oxidation catalyst and, for example, can catalyze the oxidation of ferrous ion to ferric by atmospheric oxygen.

An excellent literature survey on activated carbon and gold has recently been published by McDougall and Hancock[10].

Uses of activated carbon

Carbon has been known throughout history as an adsorbent, its usage dating back centuries before Christ. The ancient Hindus filtered their water with charcoal. In the thirteenth century, carbon materials were used in a process to purify sugar solutions. Then, late in the eighteenth century, Scheel discovered the gas adsorptive capabilities of carbon and Lowitz noted its ability to remove colours from liquids. Carbon adsorbents have since been subjected to much research resulting in numerous development techniques and applications.

One of these applications was begun in England in the mid-nineteenth century with the treatment of drinking waters for the removal of odors and tastes. From these beginnings, water and wastewater treatment with carbon has become widespread in municipal and industrial processes, including wineries and breweries, paper and pulp, pharmaceutical, food, petroleum and petrochemical, and other establishments of water usage. Likewise, activated carbons adsorb gases and vapours and as such are extensively used for air pollution control and vapour recovery operations in such applications as gas masks, cigarette filters, solvent recovery, cooker hoods and catalytic supports. The use of activated carbons for precious metal recovery constitutes a very small part of a diverse and extensive field of application.

Basically there are two forms of activated carbon: powdered and granular. Powdered carbons are usually used once on a throw-away basis and are kinetically very active because of their fine particle size. On the other hand, granular carbons, and this includes moulded and pelletised products, are generally re-used following regeneration (see below). Because of this, the use of granular carbons is often more cost effective even though the powdered carbons are invariably lower priced.

Carbon regeneration

The economics of granular carbon are such that costs would be prohibitive if it could not be reused. Thus, spent carbon passes through a regeneration process in which contaminants (usually organic) are desorbed from the carbon. This process is usually a thermal one though chemical procedures have been used.

Thermal regeneration is carried out in a rotary kiln or multiple-hearth furnace at temperatures 850-950^{o}C during which time the carbon is dried of residual water and the organic contaminants subsequently volatilized and oxidised. Combustion conditions within the furnace are controlled to limit the oxygen content to effect oxidation of the adsorbed material rather than the carbon.

After thermal regeneration, the carbon is quenched in a water bath, washed of carbon fines and recycled back to the adsorbtion circuit, or sent to storage. Air pollution devices, such as scrubbers and afterburners, are installed on the furnace to control off-gas pollutants.

During each treatment cycle carbon losses can vary between 2 and 10%. A carbon make-up hopper and bin are included in-line to provide the additional carbon necessary for the purpose. The entire thermal regeneration process usually lasts about 30 min.

Regeneration of carbon without heat has been effected, but usually with a recovered adsorptive capacity of low levels. However, the cost disadvantages of a low adsorptive capacity recovery level may be offset via by-product recovery.

In the case of carbon regeneration following gold adsorption as currently practised in South Africa, the technical necessity for carbon regeneration is complicated by the value of the gold recovered. For this reason, used carbon is invariably thermally regenerated following each adsorption cycle albeit at lower temperatures than are used in most wastewater applications (see carbon regeneration). While in many instances thermal regeneration as currently practised may not be necessary as the carbon is chemically regenerated in the elution procedure, this practice does afford some guarantee of maintaining a high level of carbon activity in such circuits. However, because of this practice, both carbon evaluation and quality control have to date not received the attention they perhaps deserve.

Carbon Evaluation

To obtain the optimum advantage from activated carbon use, experimental analysis utilizing actual operating conditions is often necessary. The operating parameters of wastewater characteristics, treatment facility configuration and effluent requirements are controlling factors in the selection of carbon type and mode of application.

The following list of characteristics can be evaluated for each activated carbon to determine its suitability for a particular application:

1. Surface area: generally, the larger the surface area, the more adsorption can take place.

2. Apparent density: a measure of the regenerability of a carbon.

3. Bulk density: used to determine carbon quantities necessary to accomplish certain jobs.

4. Effective size, mean particle diameter, uniformity coefficient: used to determine hydraulic conditions of an adsorber column.

5. Pore volume: can be used to determine the adsorbability of a particular waste entity.

6. Sieve analysis: used to check carbon breakdown from plant-handling.

7. Ash per cent: shows the activated carbon's residue.

8. Iodine number: an important parameter to be determined because it can indicate a carbon's ability to adsorb low molecular weights.

9. Molasses number, value and decolorizing index: for indication of ability of a carbon to adsorb high molecular weights.

10. Pore size: used to obtain a carbon which can adsorb specific molecules.

The relationship between a carbon's adsorbability of a substance and that substance's concentration in a wastestream (or other liquid) is the adsorption isotherm. Adsorption isotherms are determined at constant temperatures and controlled operating conditions. From the adsorption isotherm, the amount of carbon required to effect the desired effluent characteristics can be estimated.

Before discussing the possible adsorption mechanisms, it is necessary to first outline the factors affecting gold adsorption onto activated carbon, as any acceptable adsorption mechanism has to account for these effects. Due to the sometimes complex confounding effects of such variables it is not always possible to quantify each parameter in simple terms. At best, only general trends may be highlighted in a review of this nature. Furthermore, as many effects may be best described in terms of equilibrium data, such effects do not necessarily relate to the kinetics of gold adsorption in the same way.

Since various workers have used either the Freundlich or the Langmuir models in order to describe equilibrium adsorption data, it is most necessary to establish the reversibility of the adsorption reaction. In exhaustive tests lasting 3 months it was shown that carbon continued to adsorb gold from a solution containing no free cyanide and an irreversible mechanism was concluded[11]. Likewise, equilibrium experiments using different initial gold solution concentrations containing no free cyanide showed a considerable difference in adsorption capacities[12]. Similar tests however, when carried out in the presence of free cyanide yielded almost identical adsorption isotherms indicative of a reversible adsorption mechanism. The reversibility of the adsorption reaction would thus appear to be dependent on the presence of free cyanide[12].

Ionic strength of the adsorption medium. The enhanced adsorption of gold and silver in the presence of added salts, and particularly Ca^{2+}, was first demonstrated by Davidson[13] and subsequently confirmed by other workers[11,14,15]. Figure 1 shows the effect of various ionic media on the adsorption of gold where it may be considered that the very significant effects demonstrated are due to an ionic strength effect. However, concentrations of KCL three times as high as $CaCl_2$ do not produce a loading enhancement of anything like as much. Furthermore, the enhancement of loading has a dramatic effect with only a very small quantity of added electrolyte which is not in keeping with a simple ionic-strength effect.

Further investigations[15] related to the adsorption of gold in the presence of various mono and divalent cations indicated the adsorption of these counter cations which appeared to be very pH dependent. At low gold loadings (< 2 kg/t) there was a negligible adsorption of counter cations between equilibrium pH 2-10,5 indicating that the gold on the carbon was present as a neutral species or ion-paired with H^+. However, at higher gold loadings (ca 10 kg/t) there was a significant adsorption of cations above pH 6 and a corresponding increase in gold loading when compared with pure solutions (Figure 2). Once again, there was very little gold loading enhancement in the presence of Li^+ and K^+ when compared with Ca^{2+} and Mg^{2+}. The enhancement in gold loading appeared to correlate with the charge of these cations and not on their size. Although any estimation regarding the molar adsorption ratio of gold and Ca^{2+} or Mg^{2+} was complicated by the possible precipitation of $CaCO_3$ or $MgCO_3$, one is forced to conclude that a divalent cation such as calcium is specifically involved in the adsorption of gold. Also, this sensitivity of gold adsorption to low concentrations of added electrolyte may well explain in part at least the many confusing observations and conclusions reported by earlier investigators.

Calcium carbonate deposition. While the presence of a divalent cation like calcium would appear to play a significant role in enhancing gold adsorption, the simultaneous deposition of calcium carbonate which occurs on the carbon in a typical plant situation has a passivating effect on the rate of gold adsorption. Although this detrimental effect has not been satisfactorily documented and quantified, it would appear that calcium loadings of 10-20 kg/t have only a marginal effect on gold adsorption rates while loadings in excess of 50 kg/t result in a serious decline in adsorption rates. The presence of calcium carbonate does not affect the equilibrium capacity of the carbon to adsorb gold which suggests that the passivating effect is purely physical.

It is well known that activated carbon readily adsorbs oxygen and acts as an oxidation catalyst for the oxidation of free cyanide[16].

$$CN^- + \tfrac{1}{2}O_2 \rightleftharpoons CNO^-$$
$$CNO^- + 2H_2O \rightleftharpoons HCO_3^- + NH_3$$

It may thus be presumed that the free cyanide and oxygen present in typical plant circuits are the main source of such $CaCO_3$ fouling. Pilot CIP testwork[12] in which initial cyanide concentrations were controlled and cyanide degradation monitored certainly indicated poor metallurgical performance at high cyanide levels (ca 300 g/t NaCN) together with high loadings of calcium when compared with the exceptionally good metallurgical performance and low calcium loadings observed at lower cyanide levels (ca 100 g/t NaCN). This testwork further indicated that the marginal lowering of pH of the feed pulp had a most beneficial effect on both cyanide destruction as well as on gold adsorption while dramatically reducing calcium carbonate fouling (see below).

Based on solubility considerations the deposition of calcium carbonate is most sensitive to pH. As $[CO_3^=] \approx [HCO_3^-]$ at pH 8,3 it is evident that the loadings of calcium onto activated carbon becomes very pH dependent in the region pH 9-11 which is commonly found in gold plant dissolution circuits.

The acid washing of loaded carbon with dilute hydrochloric acid (3-5 vol.%) has been shown to effectively remove calcium carbonate from the carbon without eluting any gold or silver values. Acid washing of carbon prior to elution and thermal regeneration is also advisable as calcium is a particularly effective catalytic oxidant for the water gas reaction (see carbon regeneration).

Cyanide concentration and temperature. It has been amply demonstrated from a thermodynamic point of view that gold adsorption onto carbon is adversely affected by the presence of free cyanide in the adsorption medium and this effect becomes more pronounced with any increase in temperature. This effect is graphically illustrated in Figure 3 where the almost linear effect of temperature may be seen to increase from 0,50 kg Au/tC/^oC in the absence of free cyanide to 0,78 kg/tC/^oC in the presence of 200 g/t NaCN. Further equilibrium data showing the effect of free cyanide on gold adsorption at 25^oC are presented in Figure 4 where it is evident that small additions of cyanide have a pronounced effect on gold loadings but that further increases in cyanide concentration above ca 100 g/t NaCN have a much smaller effect.

From kinetic considerations it may be expected that an increase in temperature would increase the initial rate of gold adsorption but that at gold loadings approaching equilibrium a serious decrease in adsorption rates would result. These effects are well illustrated in Figure 5 where the use of the Langmuir equation would appear to depict the rate of gold adsorption onto carbon fairly well.

The free cyanide concentration also has a significant effect on the selectivity of carbon for gold when dealing with typical plant solutions or pulps but this aspect will be discussed later (see Other plant solution constituents).

Acidity of the adsorption medium. It is well established that gold adsorption onto carbon is strongly enhanced by an increase in the acidity of the adsorption medium. Equilibrium data obtained in the absence of free cyanide (Figure 6) indicate a marginal decrease in gold capacity in the region pH 6-12, whereas the gold capacity increases strongly at lower pH levels. An equilibrium gold loading of 60 kg/t of carbon at pH 12 was reported which may be compared with a gold loading of 200 kg/t obtained at pH 1.

As many of the investigations related to pH have been carried out in synthetic solutions it is perhaps not wise to assume a similar response when dealing with complex solutions or pulps (see Other plant constituents). Both the presence of free cyanide and lime together with a heterogenous mixture of metal cyanide complexes have a profound effect on gold adsorption. In the case of free cyanide, CN^- is protonated to the neutral HCN species at relatively high pH levels (pKa HCN = 9,5) thus lowering cyanide concentrations significantly when marginally lowering pH below pH 10,3 while at the same time increasing the catalytic oxidation of free cyanide to $CO_3^=$ and NH_3. The marginal lowering of pH below pH 10 will also have a significant effect on calcium carbonate fouling of the carbon (see Calcium carbonate deposition).

Other plant constituents. A number of base metals are usually leached with gold in cyanide solution, notably copper, nickel, cobalt, iron and zinc. Under typical plant conditions activated carbon is most selective in adsorbing gold from such complex mixtures. Selective metal adsorption

appears to take the form of a complex displacement sequence with gold followed by silver being the most strongly adsorbed metal species. No well defined sequence of metal displacement may be formulated as the adsorption of such metal values appears to be most sensitive to such processing parameters as pH, free cyanide concentration, temperature and of course the relative concentration of these metal values in a particular plant solution or pulp. In general, the selectivity of carbon for gold adsorption will increase with any increase in pH or free cyanide concentration.

In view of carbon's limited capacity to adsorb metal cyanide complexes, the adsorption of such values has a detrimental affect on gold adsorption by virtue of the selective replacement observed. While this effect is difficult to express in simple terms, any loss in selectivity for gold adsorption which may result from the lowering of pH or free cyanide concentration must be offset by the very much higher capacity of carbon effected by such changes.

The selective adsorption of metal cyanide complexes onto activated carbon would appear to relate to the solubility and charge of the adsorbed complex concerned (see Mechanism of Gold Adsorption). While the less soluble monovalent dicyano complexes of gold, silver and copper have all been shown to be strongly adsorbed by carbon, the more soluble divalent and trivalent tri and tetracyano complexes (e.g. Cu, Ni and Zn) are only weakly adsorbed, while the very soluble hexa and octacyano complexes (e.g. Co and Fe) are not adsorbed to any marked degree[17,18,19].

The adsorption of copper on carbon would appear to play a rather unique role in possibly aiding gold adsorption in some instances via its ability to catalyse the oxidation of free cyanide[16]. Under typical plant conditions copper is present as the $Cu(CN)_3^{2-}$ and $Cu(CN)_4^{3-}$ complexes and as such is only weakly adsorbed on carbon. On acidification or lowering of cyanide levels, however, $Cu(CN)_2^-$ is formed which is strongly adsorbed by carbon resulting in an increase in the rate of free cyanide destruction and subsequent improvement in gold adsorption efficiency. Obviously, a compromise between the beneficial effect of lower free cyanide concentrations and/or lower pH levels and the competative adsorption of copper rather than gold will pertain[19]. The subsequent cold acid washing of such gold/copper loaded carbon in effluent treatment circuits low in free cyanide (Figure 7) is also of considerable interest when related to the chemistry of gold adsorption[18]. Under such conditions, the irreversible

formation of an insoluble AuCN CuCN precipitate on the carbon followed by further gold adsorption in excess of any equilibrium loading capacity and akin to the formation of insoluble AuCN using hot acid as demonstrated by McDougall[20] is strongly suggested.

Effect of oxygen. The role played by oxygen in gold adsorption is rather complex but nevertheless most significant. By virtue of carbon's ability to adsorb oxygen from the adsorption medium, oxygen may be expected to play a significant role in cyanide oxidation and calcium carbonate deposition. Adsorbed oxygen may also influence the character of the surface active groups involved in gold adsorption if some sort of oxygen regeneration mechanism is operative. As may be expected, the roles played by oxygen dissolved in the adsorption medium as well as the oxygen adsorbed on the carbon surface are particularly difficult to define and quantify.

Because of the complexities involved in an investigation of this nature, only limited information concerning the role played by oxygen is currently available. It has been demonstrated that gold adsorption is adversely affected on treating solutions devoid of oxygen[12,20] e.g. Merrill barren solutions[20], while Tsuchida et al[15] showed that "deoxygenated" carbon displayed significantly lower gold and silver adsorption capacities when compared with normal carbon (Figure 8). Furthermore, when the "deoxygenated" carbon was exposed to oxygen bubbling through the solution, the loading of gold increased to that found on normal carbon.

Organic fouling. Activated carbon's ability to scavenge organic constituents from industrial effluents is well known and as such, the presence of such contaminants in a gold adsorption circuit will have some detrimental effect on gold adsorption. Recent studies[19,22,23] have identified a number of such poisons such as flotation reagents, lubricating oils and kerosene which are commonly found in gold plant circuits. Invariably, the rate of gold adsorption rather than the loading capacity of the carbon is seriously impairied by such poisons[19]. One possible mechanism of carbon "poisoning" that affects the rate of loading but not the loading capacity could be the adsorption of a film of the solvent onto the superficial surface of the carbon granule (which thereby creates a barrier to mass transfer) without significant penetration of the solvent into the pores of the granule. It is also significant that the influence

of the organic solvents on the rate of loading is substantially reduced in typical gold plant pulps[19], indicating that a portion of the organic solvents is adsorbed by the inert solids, thereby reducing the effective concentration in solution or on the carbon. The presence of the poisons in a CIP plant can be identified by a low extraction efficiency in the first few stages and, as the poison becomes adsorbed on the carbon, a gradually increasing extraction efficiency towards the latter stages of the plant. Since these poisons affect the rate of loading far more than the loading capacity, their influence can be counteracted to a certain extent simply by an increase in the concentration of carbon in those stages where the extraction efficiency is poor.

Physical effects. In most CIP operations where the loading of gold is not allowed to approach equilibrium, film diffusion predominates in limiting the rate of gold extraction. Under such conditions, such physical aspects as mixing, pulp density, carbon particle size and carbon concentration play a most significant role and are likely to overshadow any chemical considerations. Features such as contactor design, choice of agitation and pulp density are particularly significant[19] while adsorption rates increase linearly with the reciprocal of the carbon particle size.

Conversely, in an adsorption circuit where the loading of gold does approach equilibrium, those factors influencing loading capacity become operative. As the loading in any stage approaches equilibrium, the rate of extraction decreases and therefore any factor that influences the loading capacity has a bearing on plant performance under these conditions. Such variables as ionic strength, pH value and temperature[19] are likely to play an important role.

GOLD ELUTION

Zadra Process

Gross and Scott[1] investigated the eluant power of a wide range of inorganic compounds for gold cyanide adsorbed onto carbon. Potassium cyanide and sodium sulphide showed the highest solvent power for the adsorbed gold. This was pursued more recently by Zadra[24] who found that a gold-loaded carbon containing a negligible silver loading could be eluted efficiently with a hot (93°C) alkaline solution of sodium sulphide. This elution technique is not applicable to carbons carrying silver, since insoluble silver sulphide (Ag_2S) is precipitated in the pores of the carbon and so is retained. However, in the modified Zadra procedure[25] (the first practical elution procedure to be developed at Homestake), which utilizes a hot (90 to 93°C) alkaline cyanide solution (1,0 per cent NaOH + 0,1 per cent NaCN) at atmospheric pressure as the eluant, both gold and silver can be eluted efficiently. In the circuit that was originally proposed, the eluant is pumped upflow through the bed of loaded carbon, and the eluant containing the desorbed gold and silver is then fed to a circular electrowinning cell fitted with a stainless-steel anode and a pervious cathode containing steel wool. As the electrowinning cell and the elution unit form a closed circuit, the eluted gold and silver values are removed from the eluant by electrodeposition as rapidly as they are desorbed from the carbon, and the barren eluant is recycled to the elution circuit. Thus, continuous elution is achieved with the minimum volume of eluant. The gold and silver values are finally recovered by smelting the wool to bullion.

On implementing the modified Zadra elution procedure at the Carlton Mill, Cripple Creek, Colorado, Ross and coworkers[26] found that from 24 to 48 hours was required to strip carbon. This prompted them to investigate elution techniques that would strip the gold more rapidly, and led to the development of a pressure method for stripping the gold from activated carbon by simultaneous elution and electrolysis in a cyclic process.

It was found that the elution of gold in a pressurized column at high temperatures resulted in a marked improvement in the rate of gold elution compared with that observed at atmospheric pressure. In a comparative study the results indicated that, whereas only 8 per cent of the adsorbed gold would be eluted from the carbon at 90°C, more than 90 per cent would be eluted at 180°C, i.e., the rate of elution is about 12 times greater at 180°C. The effect of the composition of the eluant on elution was also investigated, and it was shown that gold could be eluted efficiently with tap water at 150°C, after preconditioning of the loaded carbon with one column volume of 1 per cent NaOH + 0,1 per cent NaCN for 30 minutes.

AARL Process

Using softened or deionised water as the eluant for gold, Davidson and coworkers [27,28] at the Anglo American Research Laboratories (AARL) investigated elution following the pretreatment of loaded carbon with various relatively strong caustic/cyanide reagents. Although efficient gold elution using only NaOH as a pretreatment reagent was demonstrated (little or no elution of Ag, Cu, Ni and Zn was observed) significant improvements were later obtained using a NaCN/NaOH pretreatment reagent. As found previously gold elution was particularly sensitive to temperature (Figure 9), as well as the hardness of the water used as eluant. Acid washing of the loaded carbon which removed calcium carbonate, nickel and zinc values but not gold, silver or copper from the carbon was also shown to improve gold elution. In the absence of acid washing, elution of metal values followed the sequence copper, zinc, nickel, silver and then gold suggesting a reversal of the selective adsorption mechanism previously discussed (see also Mechanism of Gold Adsorption). This elution procedure using an elution column having a relatively high length/diameter ratio when compared with the Zadra elution columns was first piloted at the Fairview Gold Mine and then at Golden Dumps (Pty) Ltd., and is now in general use in South Africa. In many of these operations 2-3 elution cycles including acid washing are carried out daily in each of one or more columns operating at ca 110°C and ca 300 kPa (gauge). Eluate from the elution columns is generally processed in an electrowinning circuit, although zinc cementation is currently receiving considerable attention. Significant savings in energy and reagents when compared with the Zadra process are claimed.

Organic Solvents

Work by Heinen et. al.,[29] and Martin et. al.,[30] around 1976, established that when ethanol and certain other water miscible solvents were added to aqueous cyanide solutions, they had a remarkable effect on the desorption of gold from carbon. For example, it was found that gold was stripped rapidly and efficiently at 60^{o} using 20 per cent ethanol solutions in contrast to the slow high temperature aqueous cyanide strip conditions originally developed by Zadra. However, the reason for this solvent effect was not established.

In a further study by Tsuchida and coworkers[15] using a variety of organic solvents containing different functional groups, it was found that aqueous acetonitrile (AN) was the best solvent for desorbing gold. When a solution containing 40 per cent v/v AN and 0,25 M NaCN, was equilibrated at room temperature with carbon loaded with gold, over 80 per cent of the gold was desorbed from carbon in less than 10 hours. Acetonitrile was slightly better than acetone, and far better than dimethylformamide and ethanol as solvents at the same concentration. Further studies[31] established that aqueous acetonitrile could be effectively applied to elute gold in 1-2 bed volumes from carbon columns, which compared favourably with the Zadra and Anglo American Procedures. The main reason why organic solvents, and acetonitrile in particular, promote the desorption of gold would seem to relate to the solvent effect on the activity of CN^{-} and $Au(CN)_2^{-}$ due to the preferential adsorption of solvent on the carbon surface.

While the elution of gold using organic solvents is undoubtedly most efficient at lower operating temperatures when compared with other elution procedures, both the recovery of solvent and the attendant fire hazards have perhaps mitigated against their wider use particularly on large scale operations. Nonetheless, considerable interest in organic elution processes are receiving attention particularly for small operations. It is also claimed that organic solvent elution effectively regenerates the carbon thus removing the need to thermally regenerate carbon, but this claim is perhaps rather speculative at the present time.

THE MECHANISM OF GOLD ADSORPTION ON ACTIVATED CARBON

Early workers[32-34] postulated a reduction mechanism for the adsorption of aurocyanide on carbon. The reducing agent was usually assumed to be carbon monoxide occluded by the carbon. Feldtmann[33] proposed the adsorption of an addition compound of the type AuCN, $CO,(CN)_2$ on the carbon. However, Allen[35-36] subsequently advanced the hypothesis that gold cyanide was adsorbed on carbon without chemical change as $Na\ Au(CN)_2$. This mechanism was generally accepted until Williams[37] found that the sodium content of a carbon loaded with $Na\ Au(CN)_2$ and then burnt was not sufficient to account for the amount of gold present.

In the 1920s, Gross and Scott[1] undertook the first detailed investigation of the fundamentals of the adsorption and elution of gold and silver cyanide on carbon. They suggested the following mechanism for the adsorption of gold on activated pinewood carbon, which contains $Ca(OH_2)$ (50 per cent of the ash of this carbon was found to be CaO):

$$2K\left[Au(CN)_2\right] + Ca(OH)_2 + CO_2 \rightleftharpoons Ca\left[Au(CN)_2\right]_2 + 2KHCO_3 \quad \ldots\ldots(1)$$

For a sugar carbon that contains no metal ions, they proposed the mechanism given in equation (2). They attributed the improved loading at low pH values to the decomposition of the $Au(CN)_2^-$ and $Ag(CN)_2^-$ complex anions in the presence of acid, according to equations (3) and (4):

$$K\left[Au(CN)_2\right] + H_2O + CO_2 \rightleftharpoons H\left[Au(CN)_2\right] + KHCO_3 \qquad (2)$$

i.e.

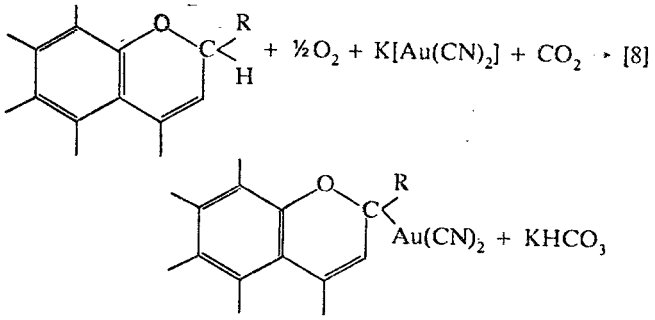
$$Au(CN)_2^- + H^+ \rightleftharpoons AuCN(s) + HCN \qquad (3)$$

$$Ag(CN)_2^- + H^+ \rightleftharpoons AgCN(s) + HCN \qquad (4)$$

They therefore thought that highly insoluble AuCN and AgCN were precipitated on the carbon. They attributed the enhanced adsorption in the presence of acid to precipitation, and not to an adsorption mechanism such as that proposed for loading under alkaline conditions.

Numerous publications have appeared subsequently in which the mechanisms advanced involve adsorption of the $Au(CN)_2^-$ anions on the carbon. Carten

and Weiss[38] proposed an alternative mechanism to explain the results of Gross and Scott[1]. They proposed electrostatic attraction of $Au(CN)_2^-$ anions to positively charged carbonium sites, as in equation (5); as written, this amounts to the rather unlikely oxidation of Au(I) to Au(III):



.....(5)

Kuzminykh and Tjurin[39] argued that since the presence of simple anions in the cyanide solution, even at high concentrations such as 1,5 M, did not affect the capacity of birch and aspen carbons for gold cyanide, the interactions between the $Au(CN)_2^-$ anion and the carbon were not electrostatic in nature. They proposed that, under acidic conditions, the neutral molecule $HAu(CN)_2$ is concentrated on the surface of the carbon by a capillary condensation mechanism, whereas, under neutral or alkaline conditions, the $Au(CN)_2^-$ anion is retained on the carbon together with an equivalent amount of an alkali cation such as K+ or Na+, which is held on the carbon electrostatically. They proposed 'dispersion forces' as being responsible for the retention of $H\,Au(CN)_2$ and $Au(CN_2^- - Na^+$ species on the carbon. Therefore, according to this theory the nature of the adsorbing form of gold cyanide is different in alkaline and acidic solutions.

Davidson[13] investigated some of the factors that influence the adsorption of gold on activated coconut-shell carbon under controlled and reproducible experimental conditions. Attempts to reproduce adsorption isotherms under conditions of constant pH value with deionized water failed, and reproducibility could be achieved only in a 0,20 M buffer medium of boric acid and 0,20 M KCI+NaOH, at a pH value of 10. He established that the degree of gold adsorption is strongly dependent on the nature and concentration of the 'spectator' cations present in the adsorption medium, and that gold cyanide cannot be adsorbed on carbon in the absence

of stabilizing cations such as Na^+ and Ca^{2+}. He advanced a mechanism that involves the adsorption of gold as a metal dicyanoaurate complex: $M^{n+}\left[Au(CN)_2\right]_n$. In support of this hypothesis, he claimed that, when the cation of the complex M^{n+} is a divalent alkaline earth metal, the complex is bound more firmly to the carbon than when M^{n+} is a monovalent alkali metal. He found the strength of adsorption of the $M^{n+}\left[Au(CN)_2\right]_n$ adsorbate to follow the sequence.

$$Ca^{2+} > Mg^{2+} > H^+ > Li^+ > Na^+ > K^+.$$

Cho, Dixon, and Pitt[40-42] investigated the kinetics and thermodynamics of the adsorption of gold and silver cyanide on activated coconut-shell carbon. They found the adsorption rates for both gold and silver cyanide to be controlled by the rate of pore diffusion in the temperature range 25 to 55°C, and the activation energies to be small - 8 to 13 kJ mol^{-1}. They found that, although the size of the carbon granules affects the initial rate of adsorption, observing faster rates for the smaller size fractions, it has no effect on the equilibrium amount of gold adsorbed. They observed that the equilibrium gold loading, in agreement with the results of other investigators[1,13,38,39], is strongly dependent on the pH value of the adsorption medium, the amount of gold adsorbed in the pH range 4 to 7 being about twice that adsorbed in the pH range 8 to 11.

The mechanism suggested by Cho et al, for the adsorption of gold cyanide on carbon involves an electrostatic interaction between $Au(CN)_2^-$ anions and positively charged sites on the surface of the carbon, as previously suggested by Plaksin[43]. They proposed that the positive sites on the carbon are formed in accordance with the theories postulated by Frumkin et al.[44-46] to account for the adsorption of acid on carbon. According to these theories, active carbon behaves like an oxygen electrode, and molecular oxygen reacts with the surface of the carbon to form a surface functional group, C_xO or C_xO_2, which reacts with water according to equation (6):

$$C_xO + H_2O \rightleftharpoons C_x^{2} + 2OH^- \quad \ldots\ldots\ldots\

Cho et al. suggested the reaction represented in equation (6) as an explanation for the higher gold capacity observed in acidic solutions, since it would be expected that, at low pH values, the equilibrium would be shifted towards the right, and more positively charged adsorption sites would be produced that would attract an equivalent number of $Au(CN)_2^-$ anions.

With regard to the adsorption of silver cyanide, they found that, as the loading of silver on the carbon was increased, the zeta potential of the carbon surface became more negative. This led to the postulate that the adsorbate is an anion, $Ag(CN)_2^-$.

Claus and Weiss[47], in studying the adsorption of gold cyanide on various activated carbons, concluded that gold adsorption is not a property of the carbon surface, since certain chemical treatments of the carbon (for example, with a mixture of nitric and hydrochloric acids at 80°C) prevent gold adsorption. Although it is believed that gold adsorption is related to specific functional groups or 'points of attachment' on the carbon, they excluded the involvement of basal planes, carboxylic acid groups and basic oxides. However, they pointed out that a special type of micropore might be involved.

Recently, Cho and Pitt[48,49] made a detailed investigation of the factors that include the adsorption of silver cyanide on activated coconut-shell carbon. Although they concluded that the $Ag(CN)_2^-$ anion is adsorbed, they excluded an electrostatic interaction between the $Ag(CN)_2^-$ anions and positive sites on the carbon surface, at least for adsorption from alkaline solutions. Instead, they advanced an adsorption mechanism that is in accordance with the ionic solvation-energy theory developed by Andersen and Bockris[50] to account for the specific adsorption of anions on metal electrodes. According to this model, the degree and type of ionic hydration are the principal factors that determine specific adsorption. Applied to carbon, the theory predicts that a large, weakly hydrated anion such as $Ag(CN)_2^-$ will be specifically adsorbed on the carbon surface after losing some of its primary-hydration water molecules, whereas some anions with a large number of strongly bound water molecules in their primary hydration shells, for example CN^-, will not be specifically adsorbed and will therefore remain in the outer part of the electrical double layer.

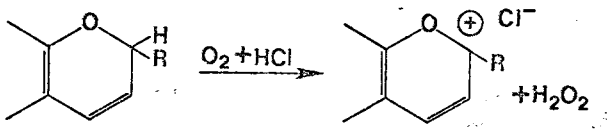
The capacity of carbon for the $Ag(CN)_2^-$ anion is considerably higher than for the CN^- anion, and Cho and Pitt regarded this as proof of the applicability of the Andersen and Bockris model to the adsorption of anions on carbon. They also found that the small, highly hydrated Na^+ cation is adsorbed on carbon only if it is present in solution with $Ag(CN)_2^-$ anions, and not if it is present alone with small, highly hydrated anions such as CN^-. They obtained similar results for the Ca^{2+} cations.

Confirming the results of Davidson[13] on the adsorption of gold cyanide, Cho and Pitt[48,49] found that an increase in the concentration of K^+, Na^+, and Ca^{2+} ions in solution enhances the adsorption of silver cyanide on carbon. To account for these 'ionic strength' effects, they proposed that the Na^+, K^+, and Ca^{2+} ions are non-specifically adsorbed on the silver cyanide complex, which in turn is specifically adsorbed on the carbon surface, and that these ions provide additional sites for the adsorption of anions such as $Ag(CN)_2^-$ and CN^-. Therefore, the picture that emerges is of a multilayer adsorption (supposedly consistent with the modern view of the electrical double layer) as depicted in Figure 10.

The deleterious effect of the free CN^- anion on the adsorption of the $Ag(CN)_2^-$ anion (although CN^- is not specifically adsorbed on the carbon) is believed to be related to competitive non-specific adsorption on the cation sites between the $Ag(CN)_2^-$ and CN^- anions.

The capacity of carbon for gold cyanide is about three times higher than for silver cyanide. Cho and Pitt attributed this to the larger ionic radius of gold (13,7 nm) compared with that of silver (12,6 nm), consistent with the solvation-energy theory.

The enhanced adsorption of silver cyanide in the presence of an acid was attributed to a different mechanism, consistent with the model postulated by Garten and Weiss[38] to account for the adsorption of hydrochloric acid on carbon in the presence of molecular oxygen, as represented in equation (7):



..... (7)

This mechanism assumes the presence on the carbon of chromene functional groups that are readily oxidized in the presence of acid to carbonium ions, and Cho[48] postulated that the $Ag(CN)_2^-$ anion is adsorbed on these positive sites by an electrostatic interaction. However, he pointed out that such positive sites can also be formed in accordance with the theories of Frumkin et al.[44-46]. Therefore, he assumed the adsorption of $Ag(CN)_2^-$ on carbon from acidic medium to be dependent on the surface charge of the carbon.

On the other hand, Grobovskii and his co-workers[51], after investigating the adsorption of gold on carbon manufactured from phenol-formaldehyde resin, concluded that the adsorption of gold is accompanied by reduction of the $Au(CN)_2^-$ complex to metallic gold. Their investigation[52] into the elution of gold adsorbed on coal carbon from cyanide solution showed that complete elution of the gold can be achieved only if a gold-complexing agent such as CN^- or CNS^- is present in the eluant. This gives credence to the hypothesis of a reduction mechanism as the responsible reactions. They consider that the role of carbon as the reducing agent is in accordance with the reaction shown in equation (8), where C_xO---Na represents the oxidized surface of the carbon with an adsorbed cation:

$$NaAu(CN)_2 + C_x + 2NaOH \rightleftharpoons C_xO\text{---}Na + Au(o) + CN2Na + H_2O \quad \ldots\ldots\ldots\ldots\ldots\ldots (8)$$

McDougall et al.[11] using X-ray photoelectron spectroscopy showed that the adsorption of gold cyanide on carbon in the presence or absence of electrolytes and acids proceeds by the same mechanism. Although the identity of the gold cyanide adsorbate was not known, the mechanism appeared to have an initial adsorption stage that involved the adsorption of an ion pair: $M^{n+}\left[Au(CN)_2^-\right]_n$, where M^{n+} is a metal ion having a lower solubility than the $Au(CN)^-$ anion in the adsorption medium. This initial stage is then followed by a reduction step in which either a sub-stoichiometric $Au(CN)_x$ surface species or a cluster-type compound of gold is formed.

In a later study Tsuchida and co-workers[15] also suggest the initial adsorption of an ion paired aurocyanide species involving charge separated carboxylate and chromene type active groups on the carbon surface. If Ca^{2+} or Mg^{2+} is present, then the ion-pair formed with the carboxylate group would have a nett positive charge and ion-pair with further $Au(CN)_2^-$. This initial stage however, is then followed by an oxidation step involving a quinone group in which the $Au(CN)_2^-$ adsorbed species is decomposed to the neutral AuCN species thus allowing for the further adsorption of $Au(CN)_2^-$.

The mechanismsof gold adsorption from cyanide solution which have been proposed may thus be summarised as follows:

(a) reduction of $Au(CN)_2^-$ to gold metal

(b) adsorption of $Ca\left[Au(CN)_2\right]_2$ ion pairs

(c) the initial adsorption of ion pairs followed by a partial reduction step to $Au(CN)_x$ or cluster type species

(d) the initial adsorption of ion pairs followed by an oxidation step in which $Au(CN)_2^-$ is degraded to AuCN.

Support for mechanism (a) stems from the obvious deposition of gold metal which occurs when carbon is added to $AuCl_4^-$ solutions. It is easy to rationalise this mechanism in acidic chloride media if we assume quinone/hydroquinone groups are associated with the active sites since the reduction potential of $AuCl_4^-$ is 0,995v (nHe), compared with quinhydrone

0,699 v(nHe). The same rationale however, cannot be applied to cyanide solutions where the reduction potential for $Au(CN)_2^-$ is much lower (E^o= - 0,6v(nHe)). Mechanism (b) gains support from the enhanced loading of gold onto carbon from solutions containing Ca^{2+} and the absence of simple ion-exchange properties. Whilst mechanism (c) is a compromise, likening the adsorption behaviour of carbon to charged electrode surfaces.

In spite of all these studies there is little agreement over whether carbon acts as a reductant or an oxidant and what role air or oxygen plays in the process. McDougall et. al., [11] support the view that Au(I) is partially reduced to a mixture of Au(I) and Au(0) in cluster-type compounds, yet the quinone group, if present, would have a higher reduction potential than $Au(CN)_2^-$ and CN^-. Furthermore, it is well known that activated carbon readily adsorbs oxygen and acts as an oxidation catalyst both for oxidation of Fe(II) to Fe(III) and for oxidation of CN^- to CNO^- and CO_3^{2-}.[16] Tsuchida et al.[15] believe that oxygen and/or quinone groups oxidise $Au(CN)_2^-$ to insoluble AuCN as suggested by Feldtmann[53] and oxidise cyanide via CNO^- to CO_3^{2-} and NH_3.

While considerable support for the adsorption of gold as the ion pair $Ca\left[Au(CN)_2^-\right]_2$ as proposed by Davidson would appear to explain many experimental observations and to have gained considerable acceptance, the subsequent oxidation of $Au(CN)_2^-$ to AuCN as proposed by Tsuchida et al.[15] remains rather speculative when related to typical plant solutions containing excess free cyanide. The reversibility of the gold adsorption reaction [12,19] under such conditions together with the irreversible nature of gold adsorption carried out in the absence of free cyanide[11] would suggest that the mechanism (d) proposed by Tsuchida et al will only take place in a system devoid of free cyanide.

From the evidence available at the present time, the adsorption of metal values onto activated carbon in a typical cyanide circuit takes place via a very reversible ion pair mechanism involving a divalent cation likely to be calcium and the metal cyanide complex. The adsorption of the ion paired metal complex is likely to follow an homologous series depending on the solubility of the adsorbed species; in particular the monovalent dicyano complexes of gold, silver and copper would appear to be strongly adsorbed onto activated carbon. The more specific nature of the functional group(s) on the carbon surface which are involved in the adsorption process are perhaps rather speculative.

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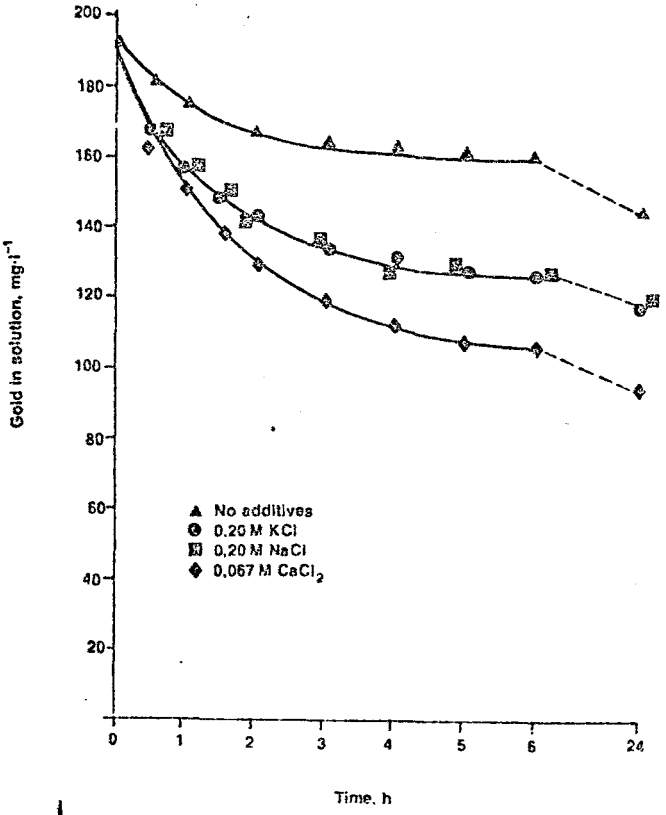


Fig. 8—Rate curves for the adsorption of aurocyanide on carbon in the absence or presence of electrolytes at the same ionic strength (experimental conditions: volume of solution 300 ml, mass of carbon 0,25 g, pH 10,0, nitrogen atmosphere, ionic strength 0,20 M, initial concentration of gold 192 mg.l $^{-1}$)

(from McDougall et al [11])

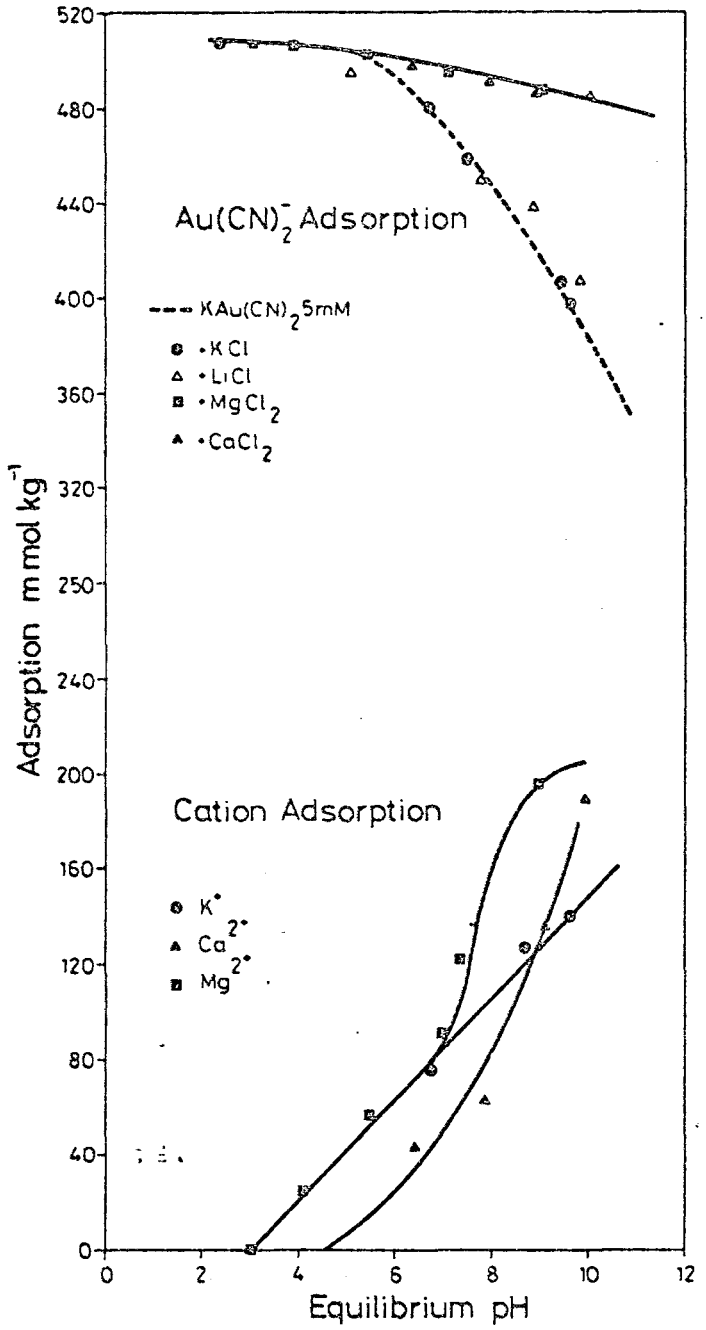


FIGURE 9.

Adsorption of gold and cations on carbon at 25°.

(pH 10; 5 mM $KAu(CN)_2$; 10 mM added salts).

(from Tsuchida et al[15])

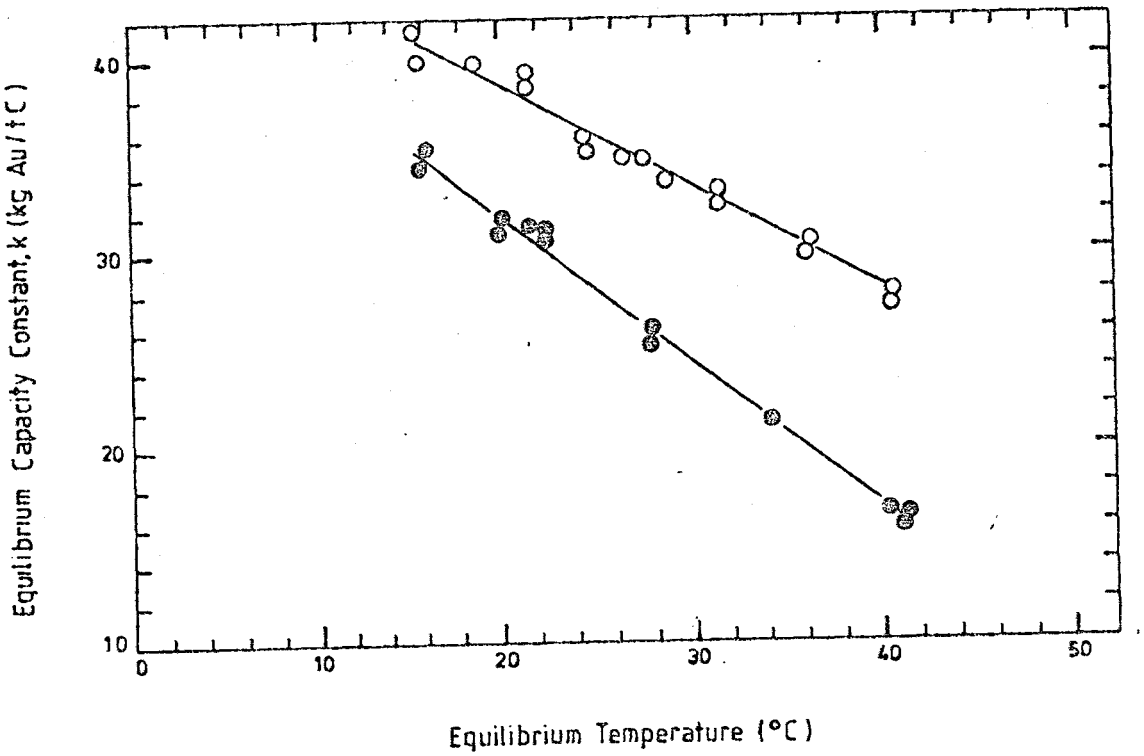


Figure 3 The Effect of Temperature on Equilibrium Gold Loadings in Synthetic Plant Solution at pH 10.

(-o- Series A - no cyanide present; -o- Series B - 200 g/t NaCN added)

(from Davidson et al[12])

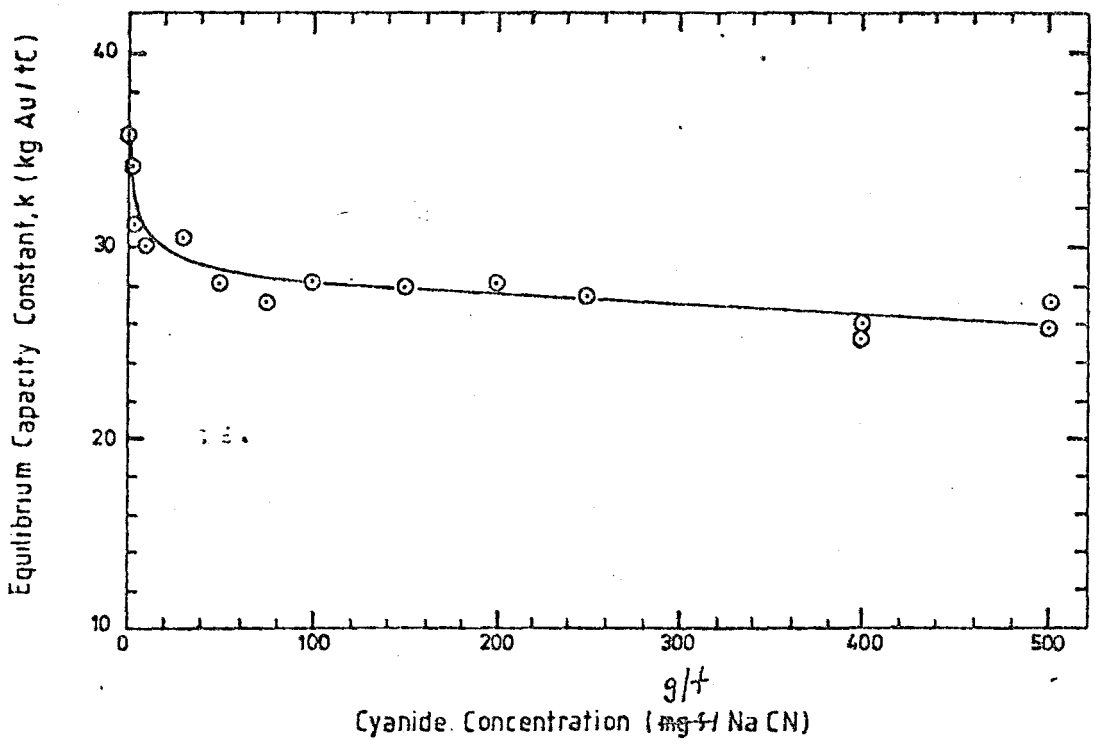


Figure 4 The Effect of Cyanide Concentration on Equilibrium Gold Loadings at 25°C in Synthetic Plant Solution at pH 10.

(from Davidson et al[12])

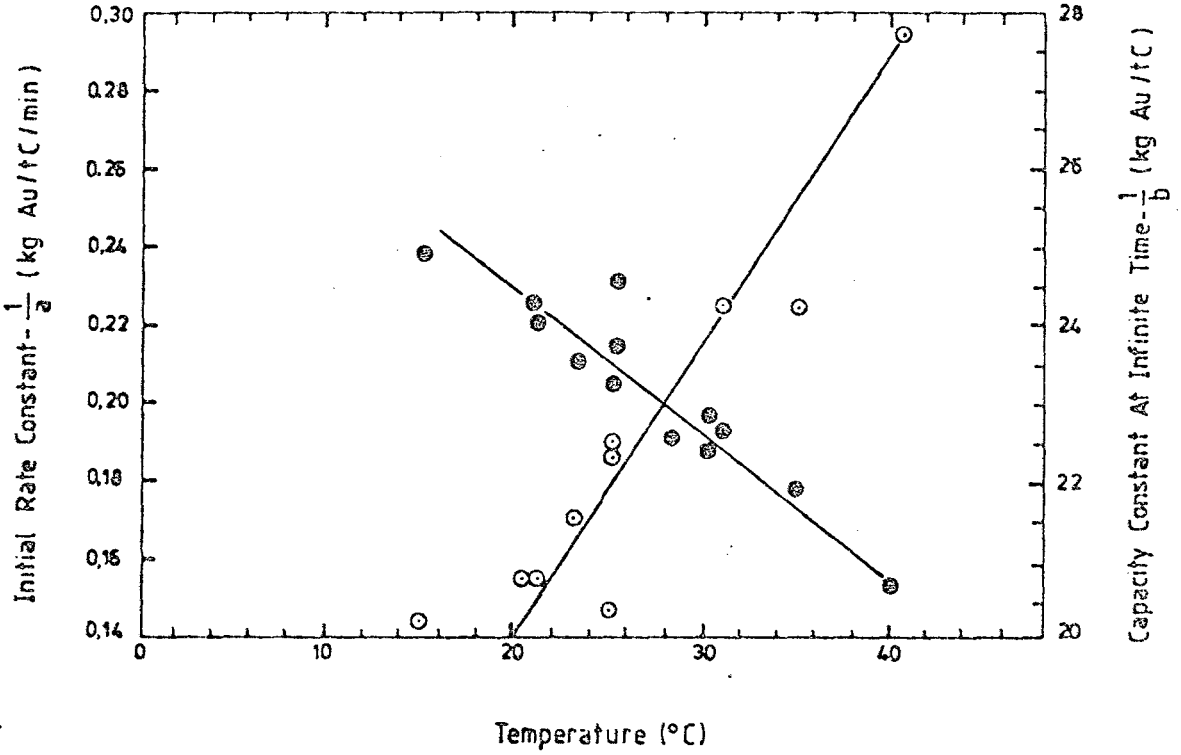


Figure 5 The Effect of Temperature on Gold Adsorption Rates in Synthetic Plant Solution at pH 10 (O initial rate constant; ● capacity constant at infinite time) (from Davidson et al [12])

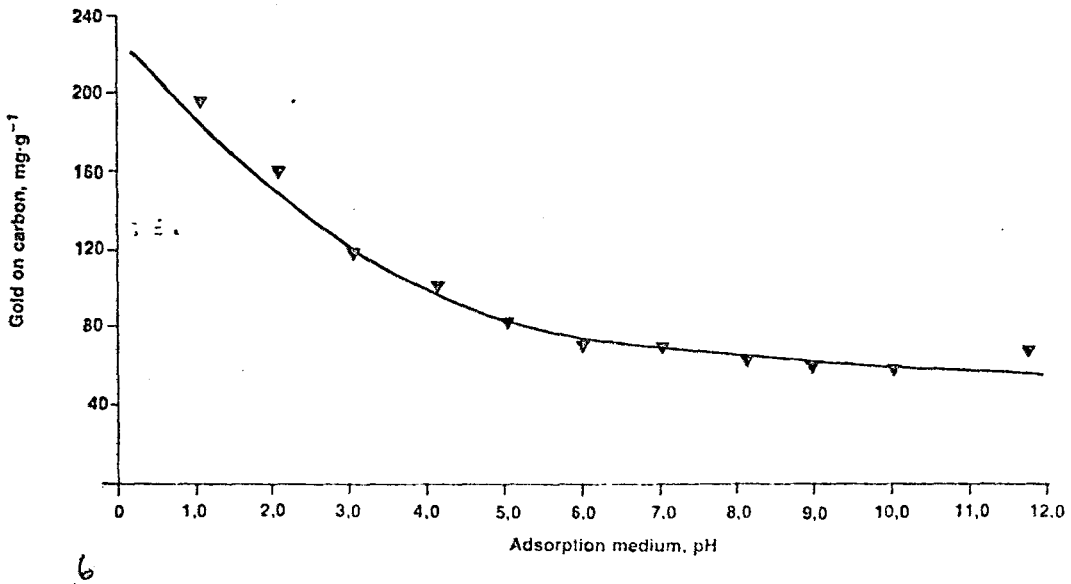


Fig. 6—Effect of the pH value of the adsorption medium on the gold capacity of the carbon (experimental conditions: volume of solution 300 ml, mass of carbon 0,25 g, nitrogen atmosphere, initial concentration of gold 190 mg.l $^{-1}$)

(from Mc Dougall et al [11])

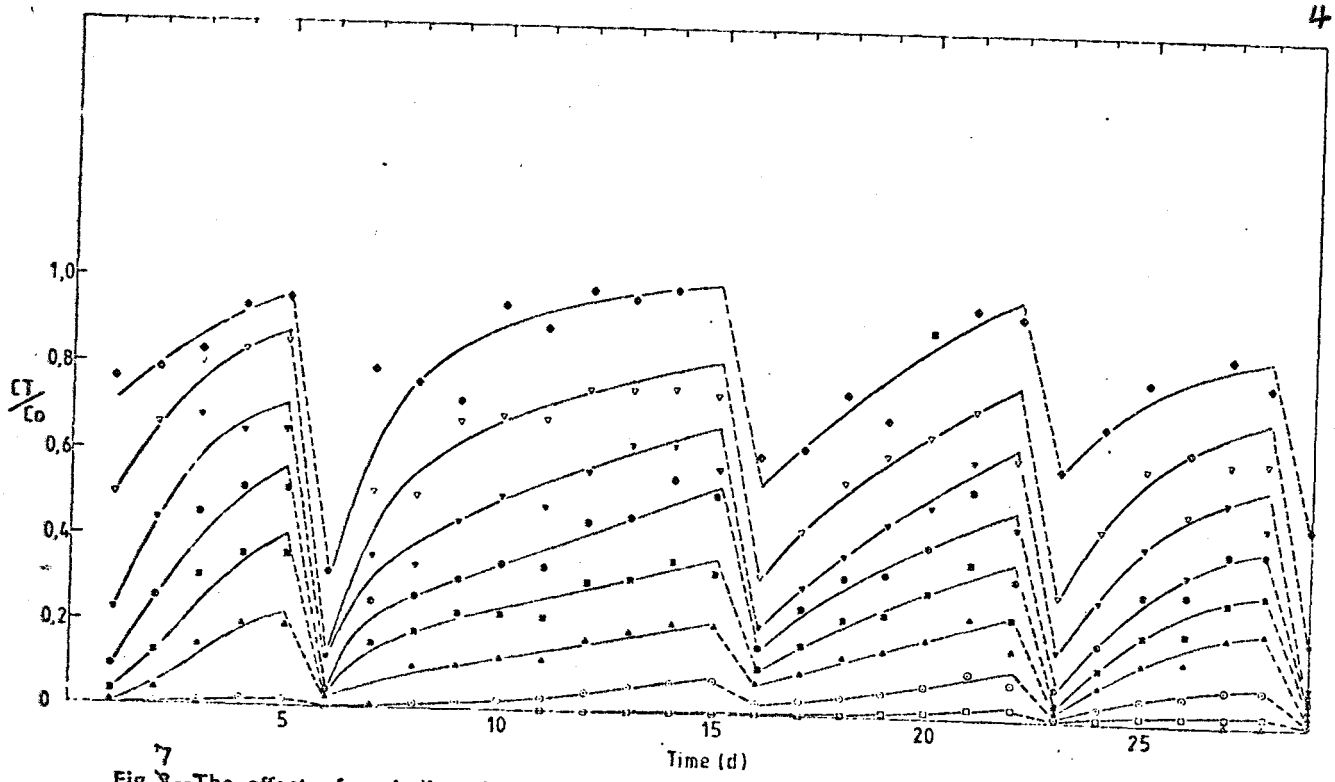
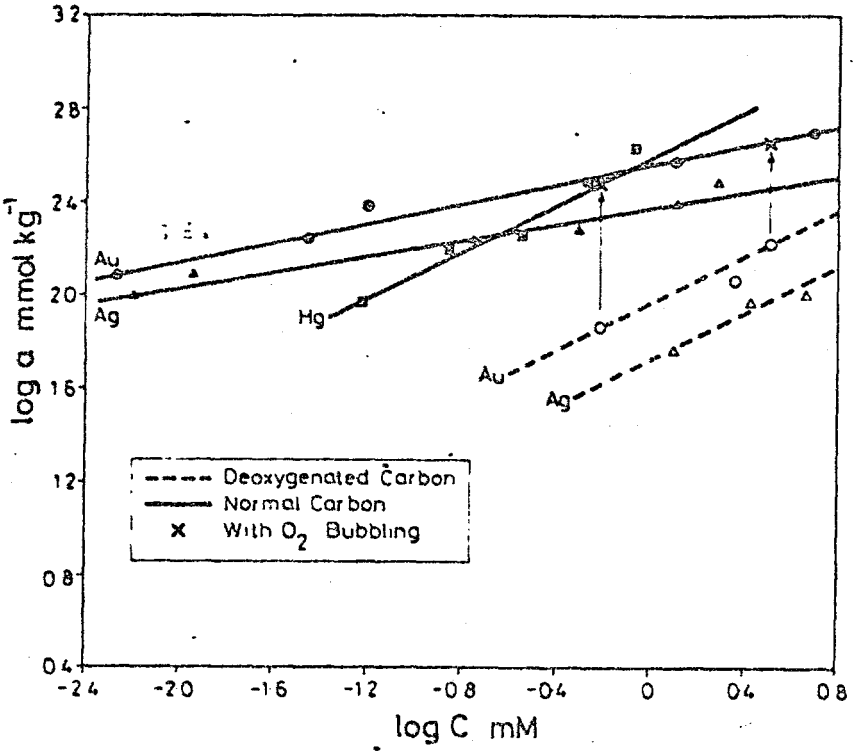


Fig. 7—The effect of periodic acid washing on the gold adsorption in a fixed-bed pilot plant
CT = gold concentration in effluent at specified depth of carbon bed; Co = gold concentration in column influent; breakthrough curves correspond to each metre of bed depth of carbon in the circuit as follows:
◆ = 1 m ▽ = 2 m ▼ = 3 m ◉ = 4 m
▣ = 5 m ▲ = 6 m ○ = 7 m □ = 8 m △ = 9 m

(from Davidson & Strong[18])

8

X. Adsorption isotherm for loading of gold, silver and mercury cyanide complexes onto normal and deoxygenated carbon at 25°.



(from Tsuchida et al[15])

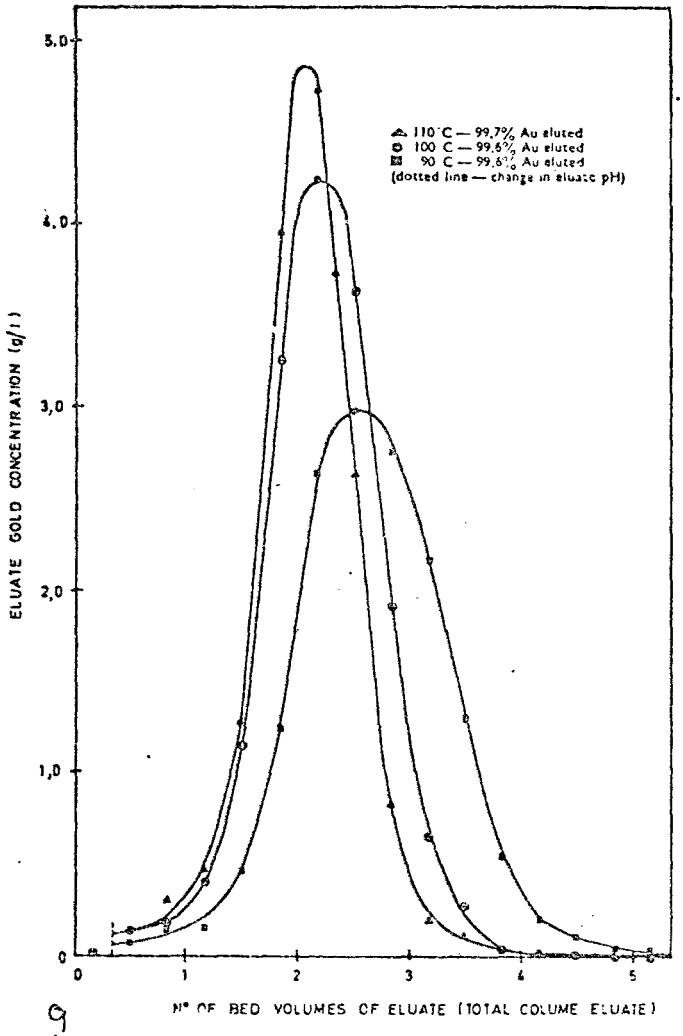


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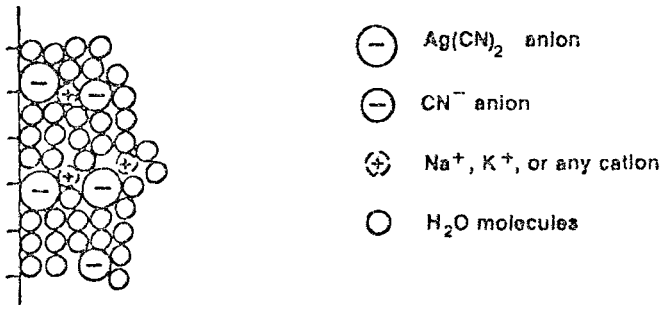
Fig. 1—The effect of temperature on the elution of carbon sample C

(from Davidson + Veranese[28])



10

Fig. 2—Schematic representation of the Cho model for the mechanism of adsorption of $Ag(CN)_2^-$ on carbon

1.2 MECHANISM OF GOLD ADSORPTION

From the evidence available at the present time, the adsorption of metal values onto activated carbon in a typical cyanide circuit takes place via an easily reversible ion-pair mechanism involving a divalent cation likely to be calcium and the metal cyanide complex. The adsorption of the ion-paired metal complex is likely to follow a homologous series, depending on the solubility of the adsorbed species; in particular the monovalent dicyano complexes of gold, silver and copper appear to be strongly adsorbed onto activated carbon. The more specific nature of the functional group(s) on the carbon surface that are involved in the adsorption process is perhaps rather speculative.

Evidence supporting this ion-paired mechanism was first published in 1974. Some of the factors that influence the adsorption of gold onto activated coconut-shell carbon were investigated (Davidson, 1974). Attempts to reproduce adsorption isotherms under controlled and reproducible experimental conditions of constant pH value with deionised water failed, and reproducibility could be achieved only in a O,20 M buffer medium of boric acid and 0,20 M KCl + NaOH, at a pH of 10. It was established that the degree of gold adsorption is strongly dependent on the nature and concentration of the 'spectator' cations present in the adsorption medium, and that gold cyanide cannot be adsorbed on carbon in the absence of stabilising cations such as Na^+ and Ca^{2+}. A mechanism that involves the adsorption of gold as a dicyanoaurate complex, $M^{n+}(Au(CN_2)_n$, was advanced. In support of this hypothesis, it was claimed that, when the cation of the complex M^{n+} is a divalent alkaline earth metal, the complex is bound more firmly to the carbon than when M^{n+} is a monovalent alkali metal. It was also found that the strength of adsorption of the $M^{n+}(Au(CN)_2)_n$ adsorbate followed the sequence:

$$Ca^{2+} > Mg^{2+} > H^+ > Li^+ > Na^+ > K^+$$

Considerable support for the adsorption of gold as the ion pair $Ca(Au(CN)_2)_2$ would appear to explain many experimental observations and has gained considerable acceptance, when related to typical plant solutions containing excess free cyanide. The reversibility of the gold adsorption reaction has also been amply demonstrated (Davidson, Douglas and Tumilty, 1982).

The mechanism of gold adsorption on activated charcoal

by R. J. DAVIDSON,* M.Sc., Ph.D. (Rhodes) (Visitor)

SYNOPSIS

The adsorption of gold onto activated charcoal is most dependent on the concentration and character of the 'spectator' cations present in the adsorption medium. In the presence of calcium ions, gold is adsorbed as the calcium aurocyanide complex. The degree to which the aurocyanide complexes are adsorbed follows the series Ca> Mg> H> Li> Na> K, calcium aurocyanide being the most stable complex and potassium aurocyanide the least firmly bound. If a loaded charcoal, where the gold is present as the calcium complex, is treated initially with either sodium carbonate or potassium carbonate solutions, the gold can subsequently be eluted from the charcoal with hot deionized water as either the sodium or the potassium complex respectively. The elution of gold with deionized water appears to be most attractive practically and economically, and has several potential applications in the recovery of gold from dilute solutions.

SAMEVATTING

Die adsorpsie van goud op geaktiveerde houtskool hang in 'n groot mate af van die konsentrasie en aard van die „toeskouerkatione" wat in die adsorpsiemiddel aanwesig is. In die aanwesigheid van kalsium-ione word goud as die kalsiumgoudsianiedkomplekse geadsorbeer, volg die serie Ca> Mg> H> Li> Na> K. Die kalsiumgoudsianied is die mees stabiele kompleks terwyl kalsiumgoudsianied die minste stabiel is. As 'n belaste houtskool, waar die goud as die kalsiumkompleks teenwoordig is, aanvanklik met natriumkarbonaat- of kaliumkarbonaatoplossings behandel word, kan die goud later met warm gedeioniseerde water as onderskeidelik die natrium- of die kaliumkompleks uit die houtskool gewas word. Die uitwassing van goud met gedeioniseerde water lyk prakties en ekonomies baie belowend en het verskeie moontlike gebruike by die herwinning van goud uit verdunde oplossings.

INTRODUCTION

Activated carbon has been used in purification schemes for centuries because of its remarkable adsorption properties. While these properties are certainly related to the porous nature of the carbon, the detailed mechanism of adsorption remains rather speculative at the present time. In the case of non-electrolytes, physical or 'specific' adsorption[1] takes place at the solid-liquid interface, whereas with electrolytes a chemical or 'non-specific' adsorption takes place in the electrical double-layer surrounding each carbon surface.

It has been proposed[1] that active carbon behaves like a gas (oxygen) electrode in which irreversibly fixed molecular oxygen reacts with the surface of the carbon to form a surface complex or functional group that is sufficiently active to cause the oxidation of water as follows:

$$C_xO + H_2O \longrightarrow C_x^{2+} + 2OH^-.$$

The presence of such charged sites on the carbon surface, together with the nature of the surrounding electrical double-layer, does explain, at least qualitatively, many of the observed influences affecting the adsorption of certain charged ions. The present study is confined to some of the factors influencing the adsorption of gold onto activated charcoal as the aurocyanide anion. The motivation behind such a fundamental study is threefold: (a) as a means of optimizing the conditions of gold recovery in actual practice, (b) to assess quantitatively the selection of charcoals suitable for gold recovery, and (c) to provide a means of assessing the activity of regenerated charcoals. A literature survey of the above requirements revealed very little relevant information.

It has been amply demonstrated that, even on a commercial scale, activated charcoal is a most efficient and selective scavenger for gold from gold-plant effluents, the major problem to date being the recovery of the gold from the loaded charcoal to allow re-use of the latter. Methods involving expensive and hazardous reagents such as sodium sulphide/sodium sulphite/sodium hydroxide[2], boiling sodium cyanide/sodium hydroxide[3], pressure leaching at 150°C with sodium cyanide/sodium hydroxide[4], and anhydrous ammonia[5], have all been used but have not yet proved to be attractive. This has perhaps inhibited the wider use of activated charcoal as a means for the economic concentration of gold from dilute cyanide solutions.

The present investigation has demonstrated the critical role played by both the character and concentration of the 'spectator' cations in effecting gold adsorption and, consequently, the feasibility of using hot deionized water as an efficient eluant in the recovery of gold from activated charcoal.

EXPERIMENTAL METHOD

Type G210 granulated coconut-shell charcoal (0,5 to 2 mm) supplied by Le Carbone SA (Pty) Ltd was used throughout the present investigation. All other chemicals used were of C.P. grade.

Gold Adsorption

The gold capacity of activated charcoal was evaluated as follows. A given weight of the air-dry charcoal was shaken with 100 ml of an approximately 100 p.p.m. solution of potassium aurocyanide for 17 hours under ambient conditions, the slurry was filtered through Whatman No. 540 filter paper, and the filtrate was analysed for gold.

The degree of gold adsorption occurring at equilibrium was measured using the empirical Freundlich Isotherm. The curve obtained by plotting the amount of solute adsorbed per unit mass of adsorbent against residual solute

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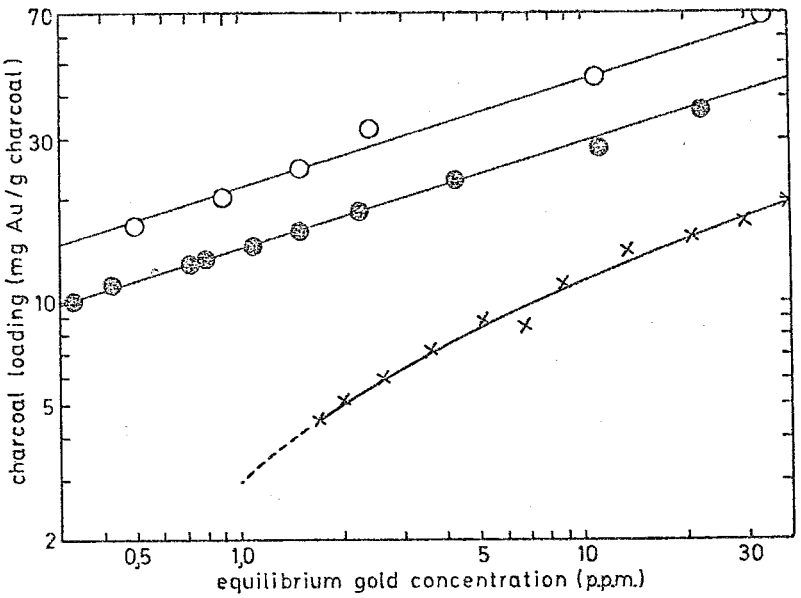

Fig. 1—The effect of ionic strength on the adsorption of gold at pH 10 (O 0,2M boric acid/0,2M potassium chloride/sodium hydroxide buffer solution, ⊛ municipal tap water, × deionized water)

concentration at a given temperature results in an isotherm equilibrium curve (the mass of the adsorbent is assumed to be proportional to its surface area). By the Freundlich isotherm relationship,

$$\frac{X}{M} = k\underline{c}^{1/n},$$

where X=mass of solute adsorbed or removed from solution,
M=mass of adsorbent,
$\underline{c}$=concentration of solute remaining in solution, and
k,n are constants, k being a measure of the fundamental effectiveness of the adsorbent, and n a measure of the change in rate of effectiveness with relative dosage.

The Freundlich equation was found to describe the mechanics of gold adsorption on activated charcoal more accurately than the more theoretical Langmuir equation.

Reaction Rates

Data on adsorption reaction rates were obtained from a given weight of air-dried charcoal that had been shaken with 100 ml of an approximately 100 p.p.m. solution of potassium aurocyanide for a given time interval before filtration and analysis. The pH of the adsorption medium was adjusted by the addition of 0,1N sodium hydroxide or 0,1N hydrochloric acid as required.

Gold Elution

15 g of air-dried loaded charcoal (ca 0,5% Au) was placed in an oil-jacketed chromatography column (0,9 cm internal diameter by 40 cm) in series with an LKB 17000 Minirac fraction collector. After wetting back the charcoal, deionized water was pumped through the column, and the subsequent eluate fractions were analysed. A high-quality deionized water was used in the present study (conductivity ca 1500 μS/m, Na 0,5 p.p.m., K 0,2 p.p.m., Ca 0,15 p.p.m., Mg 0,01 p.p.m.).

FACTORS AFFECTING GOLD ADSORPTION

Effect of Ionic Strength on Gold Adsorption

Early attempts using tap water, and finally deionized water, to reproduce adsorption isotherms under conditions of constant pH and temperature proved futile. Good reproducibility was, however, achieved in a strong 0,2 M boric acid/0,2 M potassium chloride/sodium hydroxide buffer solution (pH 10,0). Fig. 1 illustrates a wide divergence of capacity constants, $\underline{k}$, although the $\underline{n}$ values remain virtually constant at 3,2. A progressive increase in $\underline{k}$ values from 5 mg/g with deionized water to 22 mg/g in the standard boric buffer solution suggested that the ionic strength of the adsorption medium played some important role in the adsorption

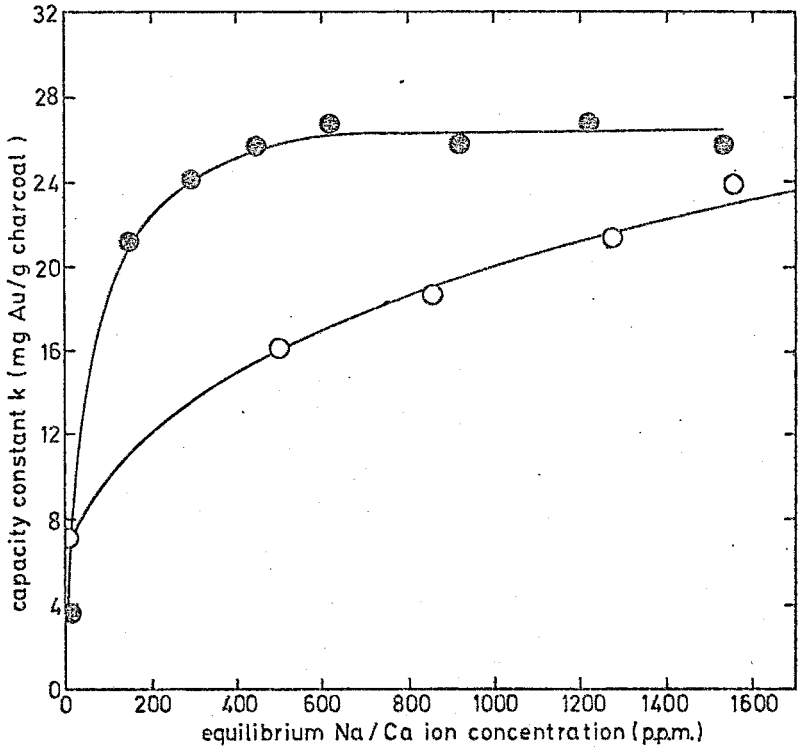

Fig. 2—The effect of sodium and calcium ions on the adsorption of gold at pH 7,2 (⊛ Ca^{2+} added as calcium chloride O Na^{+} added as sodium chloride)

mechanism. The mechanism was thought to involve stabilizing the charge on the relatively large aurocyanide anion by means of a complementary small cation, indicating the important role played by 'spectator' cations comprising the charged electrical double-layer surrounding charcoal surfaces.

Fig. 2 shows the effect of an increase in the ionic strength of the adsorption medium with sodium and calcium ions, respectively. A very marked increase in the capacity constant, k, of the charcoal can be observed in each instance, although the addition of calcium ions has a far greater effect than the addition of sodium ions. This very large difference in k produced by the presence of sodium and calcium ions cannot be explained in terms of ionic strength alone, and it is considered that the adsorbed calcium aurocyanide complex is, in addition, more stable than the sodium aurocyanide complex.

The influence of excessive additions of calcium ions to the adsorption medium likewise had a noticeable effect on both reaction rates and on adsorption isotherms when compared with the effects produced by the addition of sodium ions. Figs. 3 and 4 illustrate the reaction rates for gold adsorption at various pH levels (3 to 11) obtained in the presence of excess sodium and calcium ions respectively. It can be seen that the addition of 1000 p.p.m. of calcium ions has a greater influence on the kinetics of gold adsorption than does the presence of 2000 p.p.m. of sodium ions.

Fig. 5 illustrates the equilibrium adsorption isotherms obtained at pH 10 for gold solutions in various ionic media in contact with activated charcoal. Both the capacity constant, k, and the slope (n=3,2) for the isotherm carried out in the presence of excess sodium ions (graph O in Fig. 5) were typical of solutions containing only sodium and potassium cations. Results were easily reproducible, and the adsorption of gold was found to obey the Freundlich equation between equilibrium concentrations 0,01 to 100 p.p.m. of gold. Conversely, gold

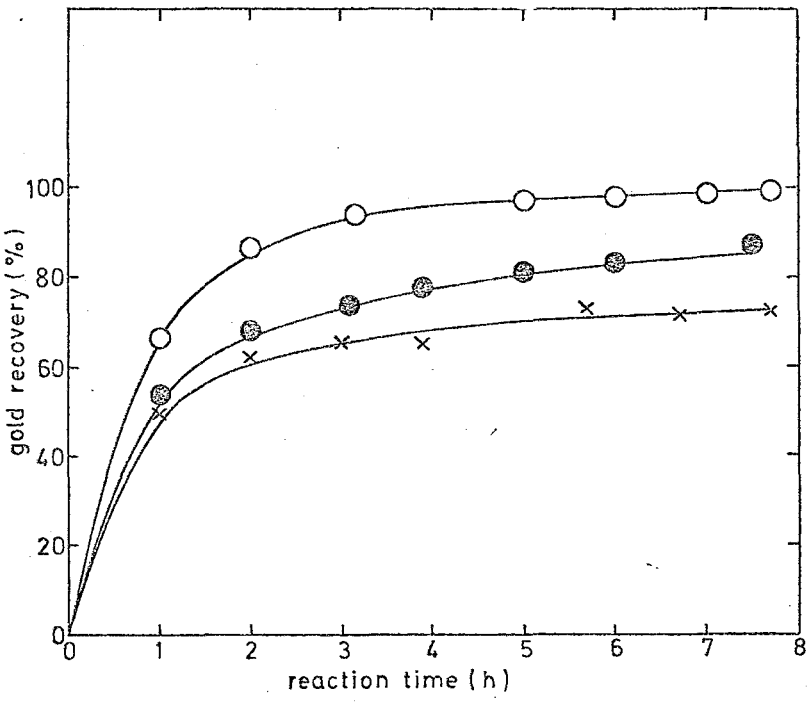

Fig. 3—The effect of pH on reaction rates in the presence of 2000 p.p.m. of sodium added as sodium chloride
(O pH 3,3, ⊛ pH 7,4,
× pH 11,3)

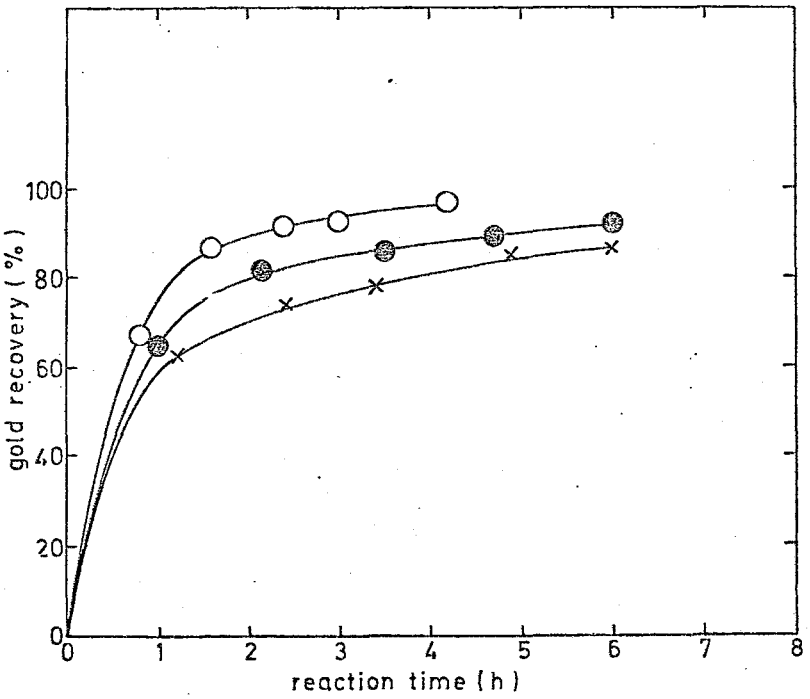

Fig. 4—The effect of pH on reaction rates in the presence of 1000 p.p.m. of calcium added as calcium chloride
(O pH 3,1, ⊛ pH 7,2,
× pH 10,6)

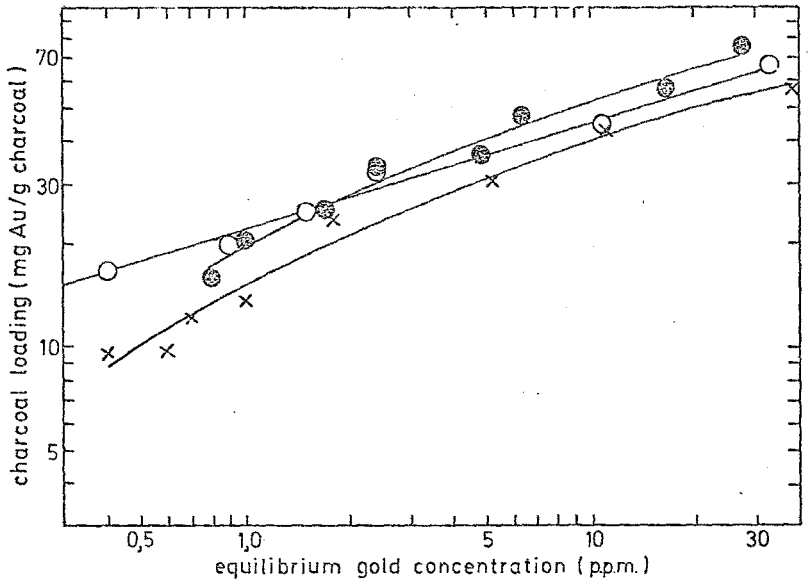

Fig. 5—The effect of the ionic media on the adsorption of gold at pH 10 (○ 0,2M boric acid, 0,2M potassium chloride/sodium hydroxide buffer solution, ◍ 1000 p.p.m. Ca^{2+} added as calcium chloride, × gold-plant effluent)

adsorption carried out in the presence of excess calcium ions (graphs ◍ and × in Fig. 5) yielded poor reproducibility, the isotherm resulting in a different value for n ($n \approx 2$) and not displaying a linear log-log relationship. The similarity between the isotherm obtained in the presence of 1000 p.p.m. of calcium ions and that obtained with gold-plant effluent to which potassium aurocyanide had been added should be noted.

Effect of pH on Gold Adsorption

Because both hydrogen and hydroxide ions tend to be adsorbed fairly strongly by activated carbons, the adsorption of other ions is influenced to a considerable extent by the pH of the solution medium. Fig. 6 shows the effect of pH on the adsorption capacity of activated charcoal for gold in the presence of 2000 p.p.m. of sodium ions added as sodium chloride. The pH was adjusted by the addition of either sodium hydroxide or hydrochloric acid to the system. An increase in the capacity constant, k, from 18 mg/g at pH 10 to 44 mg/g at pH 4 should be noted. This observed decrease in k at higher pH values can be expected if it is considered that both the hydroxide and aurocyanide anions are competing for active adsorption sites on the charcoal surface. Likewise, the reaction rates of gold adsorption were found to increase considerably when the pH of the adsorption medium was lowered (Figs. 3 and 4).

Effect of Temperature on Gold Adsorption

Being an exothermic reaction, adsorption tends to increase with decreasing temperature. The energy changes accompanying adsorption are usually very small, and thus small ambient variations in temperature tend not to alter the adsorption process to any significant extent. Studies of equilibrium gold adsorption carried out under isothermal conditions in a strong boric acid/potassium chloride/sodium hydroxide buffer solution at pH 10 as a function of temperature (20 to 30°C) indicated a linear variation in the capacity constant, k, of only 0,3 mg/g/°C.

Because earlier findings had demonstrated that gold adsorption was most dependent on both the concentration and character of the 'spectator' cations present in the adsorption medium, the use of deionized water was immediately suggested as an eluant for the subsequent recovery of gold from activated charcoal. Studies of equilibrium adsorption carried out under ambient conditions (22°C) with deionized water yielded a capacity constant, k, of 5 mg/g, which was significantly less than the values observed at higher ionic strengths (Fig. 1), where k was 22 mg/g. This effect when deionized water is used can be expected to be greatly magnified at elevated temperatures. Fig. 7 illustrates the equilibrium isotherm for gold adsorption carried out at 94°C (refluxed at the boil for 7 hours) with deionized water. A

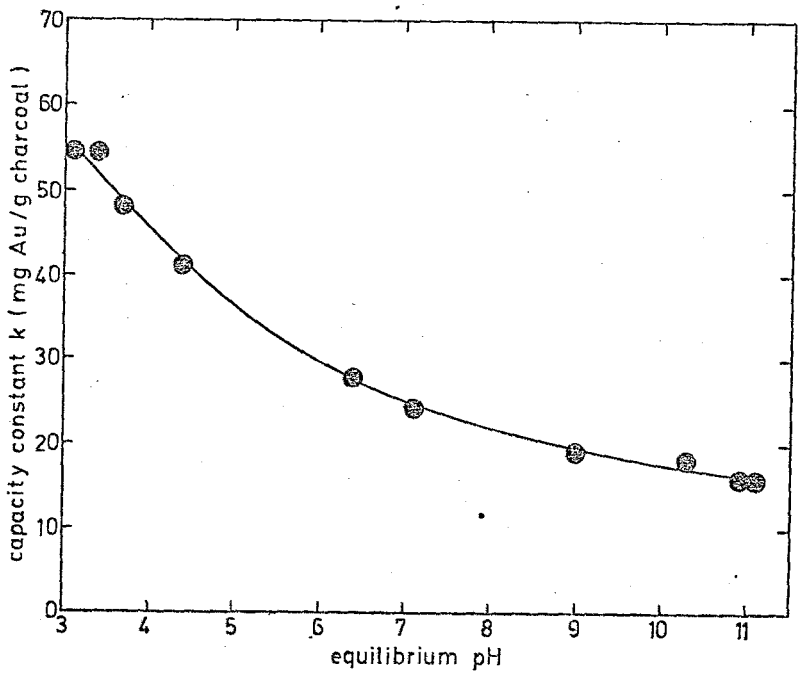

Fig. 6—The effect of pH on the adsorption of gold in the presence of 2000 p.p.m. of sodium

capacity constant, k, of only 0,13 mg/g was obtained. It can also be noted that concentrations of sodium and potassium ions in solution increased progressively to 5 and 40 p.p.m. respectively as larger amounts of charcoal were used. The increase was attributed to the leaching of impurities from the original charcoal (see Table I).

The above findings suggested the use of hot deionized water as an eluant in the recovery of gold from activated charcoal. Further investigation was confined to a demonstration that such a possibility did in fact exist.

ELUTION OF GOLD FROM ACTIVATED CHARCOAL WITH DEIONIZED WATER

Elution from Charcoal Loaded with Potassium Aurocyanide

Activated charcoal was loaded to 0,63 per cent gold by being shaken in a solution of potassium aurocyanide and then being air-dried prior to use. The subsequent elution of the loaded charcoal at a flow velocity of 2,4 cm/min and 90°C with deionized water as eluant yielded a gold recovery of 43 per cent in 21 bed volumes (Fig. 8).

In a similar exercise, the loaded charcoal was soaked in a 5 per cent calcium chloride solution overnight before being washed and eluted at 90°C as above. After elution with 22 bed volumes, a recovery of only 1 per cent gold was recorded (Fig. 8).

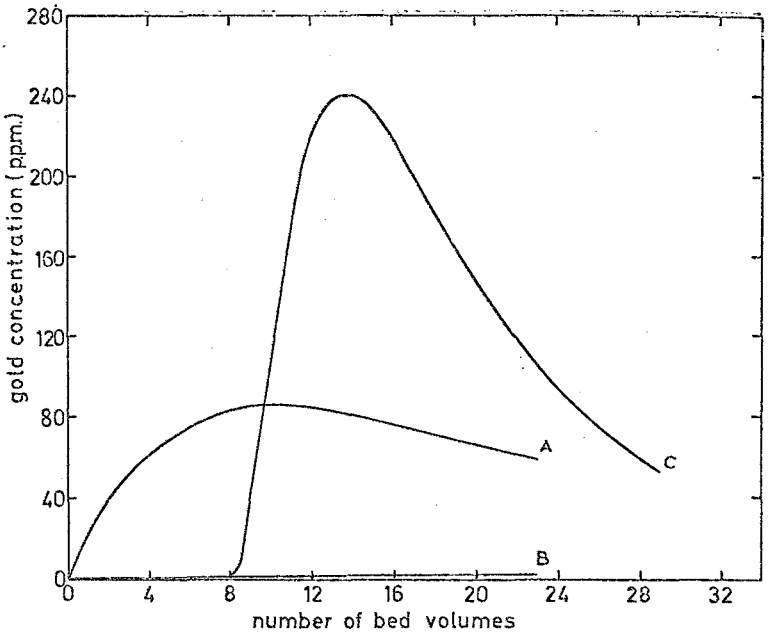

Fig. 8—The elution of charcoal loaded with potassium aurocyanide (0,63 per cent Au) at 90°C using deionized water
A no carbonate pretreatment
B pretreated with 5 per cent calcium chloride
C pretreated with 10 per cent sodium carbonate)

The same charcoal sample was then treated with 8 bed volumes of a 10 per cent sodium carbonate solution at 50°C before being eluted with deionized water at 90°C. A gold recovery of 80 per cent in 21 bed volumes of deionized water was recorded (Fig. 8).

Elution from Charcoal Loaded from Gold-plant Effluent

Activated charcoal was loaded to 0,33 per cent gold from a gold-plant effluent containing ca 0,1 p.p.m. of gold. Table I, which compares the composition of the original charcoal with that of the loaded charcoal, shows that there was a large increase in the calcium content of the loaded charcoal.

The elution of the above loaded charcoal, when only deionized water at 90°C was used, yielded a gold recovery of only 2 per cent in 8 bed volumes.

Samples of the loaded charcoal were then each pretreated, before elution at 90°C, with 8 bed volumes of the following reagents at 50°C with deionized water as before: 10 per cent potassium carbonate, 10 per cent sodium carbonate, 1,5 per cent lithium carbonate (saturated solution), 10 per cent ammonium carbonate, and 10 per cent sodium chloride.

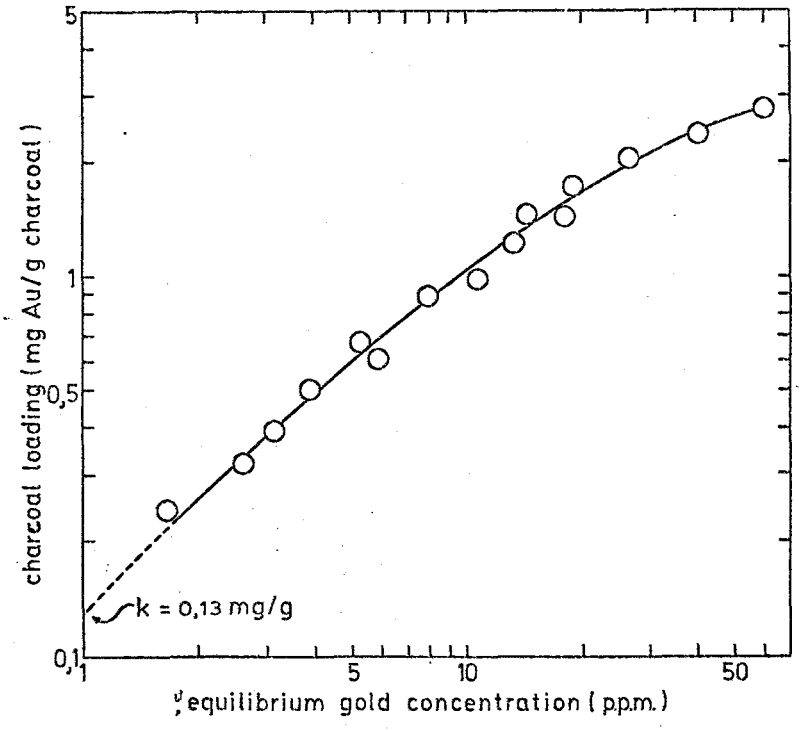

Fig. 7—The equilibrium adsorption isotherm for gold adsorption at 94°C using deionized water

TABLE I

THE COMPOSITION OF ACTIVATED CHARCOAL LOADED TO 0,3 PER CENT GOLD FROM GOLD-PLANT EFFLUENT

Element	Original charcoal p.p.m.	Loaded charcoal p.p.m.	Acid-washed loaded charcoal p.p.m.
Na	18 500	4 500	—
K	10 400	1 900	—
Ca	640	4 800	410
Mg	300	100	—
Au	0	3 300	3 500
Ag	< 10	45	25
Fe	1 500	1 500	1 600
Ni	10	2 500	290
Cu	20	240	170
Co	4	2	—
Zn	120	380	80

The semi-quantitative spectrographic analysis, reported in p.p.m., for the loaded charcoal is as follows:
Sr 30, Bi < 5, Pb 10, Sb < 10, Sn 5, Y < 10, Be < 0,1, Mo < 10, Nb < 1000, La < 100, V 10, Zr 10, Cd < 5, Ti 50, Al 100, Ba < 100, Ge < 5, Si 300, Cr < 50, Mn 10, B 10, As 100

Fig. 9 illustrates the resulting elution patterns, which indicate that only potassium, lithium, and sodium carbonate have any significant effect on the subsequent elution of gold. The gold recovery was 84 per cent in 20 bed volumes of deionized water when 10 per cent potassium carbonate was used.

In a similar exercise, a sample of loaded charcoal was eluted with 4 bed volumes of 5 per cent potassium carbonate at 50°C, followed by elution with deionized water at 90°C, the flow velocity being maintained at 0,74 cm/min. The subsequent analysis of the eluate fractions (Table II) indicates the simultaneous elution of gold and silver in the form of their potassium cyanide complexes, together with the elution of nickel and copper. The antipathetic relationship between gold and silver with respect to calcium and magnesium is striking, indicating that gold does not elute as a calcium or magnesium aurocyanide under such conditions.

Effect of Acid Pretreatment on Charcoal Loaded from Gold-plant Effluent

A sample of charcoal loaded from a gold-plant effluent was acid-washed by means of continuous elution overnight with 2,5 per cent (v/v) nitric acid/2,5 per cent (v/v) hydrochloric acid in order to remove the majority of the adsorbed elements (see Table I). No gold was eluted during this process.

The charcoal was then eluted with 4 bed volumes of 5 per cent potassium carbonate at 50°C, followed by 23 bed volumes of deionized water at 90°C, a flow velocity of 0,74 cm/min being maintained throughout the elution. A gold recovery of only 47 per cent was recorded, which can be compared with a recovery of 91 per cent achieved in the elution, under similar conditions, of charcoal that had not been washed in acid. A considerably reduced gold recovery can be noted.

A sample of the acid-washed

TABLE II

THE CONTINUOUS ELUTION OF VARIOUS ADSORBED ELEMENTS FROM ACTIVATED CHARCOAL LOADED FROM GOLD-PLANT EFFLUENT

Elution volume ml	K p.p.m.	Na p.p.m.	Ca p.p.m.	Mg p.p.m.	Au p.p.m.	Ag p.p.m.	Fe p.p.m.	Ni p.p.m.	Cu p.p.m.	Co p.p.m.	Zn p.p.m.
19	16 000	200	4,6	0,8	1,3	< 1	5	9	2	< 1	3
39	22 000	41	2,8	0,6	1,5		3	11	1		5
60	24 000	25	2,0	0,6	1,6		2	14	2		5
80	24 000	23	1,8	< 0,5	1,8		2	17	3		5
100	25 000	22	1,8		2,0		2	18	4		5
121	20 000	22	< 0,5		28	—	2	—	—	—	—
141	2 700	5			96	12	< 1	860	35	< 1	5
160	660	5			144	16		200	12		2
179	660	5			185	13		60	6		2
198	510	6			213	9		20	5		1
217	400	6			235	6		11	3		< 1
236	340	6			228	3		7	2		
255	290	6			209	2		4	2		
275	260	7	1,0		192	1		3	2		
293	240	7	1,0		170	1		2	1		
312	220	7	0,8		145	1		2	1		
332	180	7	0,8		121	1		1	1		
349	160	7	0,8		92	< 1		1	1		
370	130	8	1,0		82			1	1		
389	100	8	1,2		65			1	1		
440	108	8	1,2		52			1	1		
423	83	8	1,2		39			1	< 1		
442	73	9	1,6		29			1			
461	57	8	2,4		20			1			
480	41	8	3,6		13,3			1			
498	30	6	6,6		9,1			1			
517	17,6	6	10,0	0,6	7,0			< 1			
536	10,4	5	12,0	0,6	5,8						
555	6,9	5	13,0	0,6	5,1						
575	4,7	5	13,8	0,8	4,8						

NOTE: 1. Fractions of 0 to 100 ml were eluted with 5 per cent K_2CO_3 at 50°C.
2. Fractions of 100 to 575 ml were eluted with deionized water at 90°C.
3. 25 ml = 1 bed volume.

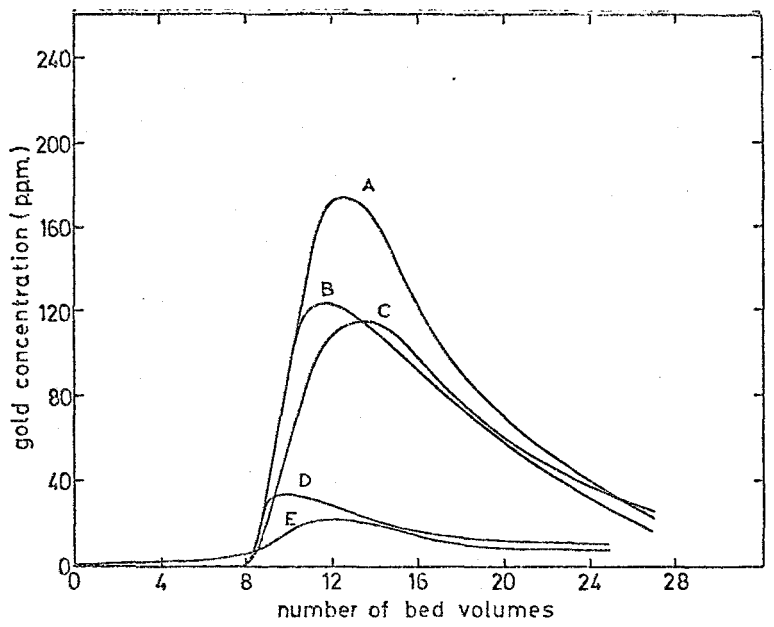

Fig. 9—The elution of loaded charcoal (0,33 per cent Au) at 90°C using a carbonate pretreatment at 50°C
(A pretreatment 10 per cent potassium carbonate, gold recovery 84 per cent
B pretreatment 1,5 per cent lithium carbonate, gold recovery 68 per cent
C pretreatment 10 per cent sodium carbonate, gold recovery 60 per cent
D pretreatment 10 per cent ammonium carbonate, gold recovery 18 per cent
E pretreatment 10 per cent sodium chloride, gold recovery 12 per cent)

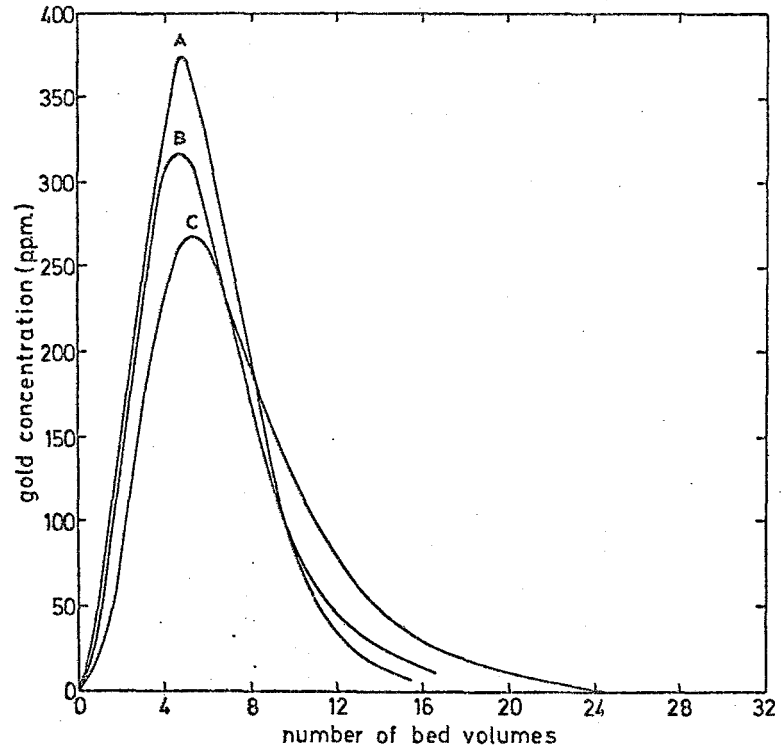

Fig. 10—The effect of flow velocities on the elution of gold at 90°C using deionizd water
(A eluted at 0,18 cm/min, B eluted at 0,45 cm/min, C eluted at 0,72 cm/min)

charcoal was then eluted as above, save for the pretreatment stage, where a 5 per cent potassium carbonate/3 per cent potassium hydroxide solution was used, instead of the 5 per cent potassium carbonate solution. A subsequent gold recovery of 87 per cent in 16 bed volumes of deionized water was obtained, indicating the important role played by the potassium hydroxide in effecting gold elution under the prescribed conditions.

Optimization of the Carbonate Pretreatment

Initially, the application of 8 bed volumes of 10 per cent potassium carbonate eluted at 50°C resulted in good gold recoveries from charcoal loaded from gold-plant effluent. During the course of the investigation, the temperatures of pretreatment were raised to 90°C, and the carbonate treatment was limited to ca ½ bed volume (the dry loaded charcoal was placed in the column before 12 ml of the carbonate solution was pumped in, and then the saturated charcoal was allowed to stand for 2 hours before being eluted with water). Potassium hydroxide was also added to the pretreatment solution with beneficial results.

Table III summarizes the gold recoveries obtained as a function of the potassium carbonate treatment used. Gold recoveries were highest when a 5 per cent potassium carbonate/10 per cent potassium hydroxide solution was used in a continuous elution procedure.

Effect of Flowrates

Fig. 10 illustrates the effect of decreasing the flowrate of the eluant using a 3 per cent potassium carbonate/1 per cent potassium hydroxide pretreatment as indicated in Table III. The sharper elution patterns obtained as flow velocities decreased from 0,74 to 0,20 cm/min suggest that the kinetics of gold desorption are diffusion controlled under the prescribed conditions.

Gold Elution at Higher Temperatures

As is to be expected, the elution of gold from activated charcoal is greatly enhanced at higher temperatures, affording the use of faster flowrates and thus shorter elution times, as well as yielding more-concentrated gold eluates.

Fig. 11 shows the elution of gold

TABLE III

THE OPTIMIZATION OF THE CARBONATE PRETREATMENT OF ACTIVATED CHARCOAL CONTAINING 0,3 PER CENT GOLD LOADED FROM GOLD-PLANT EFFLUENT

Pretreatment	Flow velocity cm/min	No. of bed volumes (carbonate pretreatment)	Temp. of pretreatment °C	Temp. of elution °C	Gold recovery
1% K_2CO_3	2,37	4	50	90	70% in 24 bed vol.
5% K_2CO_3	0,74	8	50	90	91% in 19 bed vol.
10% K_2CO_3	2,37	8	50	90	48% in 20 bed vol.
0,5% K_2CO_3+ 0,5% KOH	0,74	ca ½	90	90	35% in 24 bed vol.
1% K_2CO_3+ 1% KOH	0,74	ca ½	90	90	73% in 16 bed vol. 82% in 24 bed vol.
3% K_2CO_3+ 1% KOH	0,74	ca ½	90	90	88% in 16 bed vol. 93% in 24 bed vol.
5% K_2CO_3+ 3% KOH	0,74	ca ½	90	90	96% in 16 bed vol. 98% in 25 bed vol.
5% K_2CO_3+ 3% KOH	0,74	1	90	90	98% in 16 bed vol. 99% in 25 bed vol.
10% K_2CO_3+ 5% KOH	0,74	1	90	90	94% in 7 bed vol. 98% in 12 bed vol. 99% in 24 bed vol.
5% K_2CO_3+ 10% KOH	0,74	1	90	90	94% in 7 bed vol. 99% in 12 bed vol. 100% in 22 bed vol.

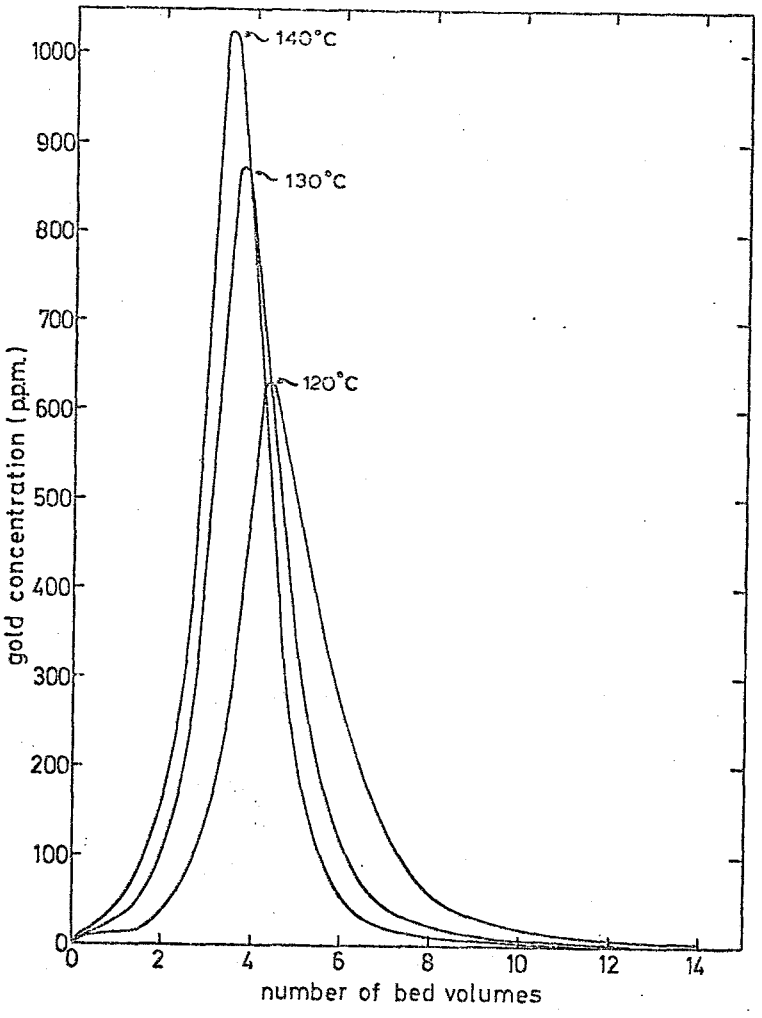

Fig. 11—The effect of temperature on the elution of gold at a flow velocity of 3,6 cm/min

at 120 to 140°C carried out at 3 to 4 bar using a 1 bed volume pretreatment of a 10 per cent potassium carbonate/5 per cent potassium hydroxide solution and eluting at a flow velocity of 3,5 cm/min. Elution at 140°C can be seen to yield a gold recovery of 95 per cent in 10 bed volumes, the complete elution taking 1,9 hours and yielding an eluate averaging 200 p.p.m. of gold. Alternatively, by use of selective elution, a 91 per cent recovery of gold can be obtained in an eluate averaging 500 p.p.m. of gold.

Use of Municipal Water as Eluant

While the present study on elution has been confined to the use of a high-quality deionized water, the presence of relatively high concentrations of potassium in the eluate (Table III) suggest that suitably softened municipal water can be used as eluant. In comparative elution studies carried out at 90°C with 5 per cent potassium carbonate/3 per cent potassium hydroxide as the pretreatment reagent, the following gold recoveries were recorded after elution with 20 bed volumes of different types of water:

97,3 per cent with deionized water,
97,6 per cent with a 'soft' municipal water (25 p.p.m. of calcium),
96,8 per cent with a 'medium' municipal water (103 p.p.m. of calcium),
81,7 per cent with a 'hard' municipal water (234 p.p.m. of calcium), and
93,1 per cent with a 'hard' municipal water suitably softened by the addition of potassium carbonate.

The above study suggests that medium-soft municipal waters can be satisfactorily used as eluants for the elution of gold from charcoal under the prescribed experimental conditions.

Capacity of Regenerated Charcoals

Studies of equilibrium isothermal gold adsorption on charcoal loaded with potassium aurocyanide and subsequently eluted at 90°C by the prescribed process indicated no decrease in the capacity constant, k, when tested in a boric acid buffer solution containing 100 p.p.m. of gold as potassium aurocyanide. The regeneration of charcoal loaded from

gold-plant effluent does, however, result in a decrease in the activity of the charcoal due to the presence of organic poisons. A soluble yellow organic fraction was, in fact, observed to elute with the otherwise colourless gold eluate. Infrared analysis of this organic fraction strongly suggested the presence of a saponified lubricating oil or grease. As the activity of such regenerated charcoal will almost surely be a function of the time period of initial loading, together with the elution characteristics of such organic poisons, further investigation is required in this direction.

MECHANISM OF GOLD ADSORPTION

The investigation into the effects of sodium chloride and calcium chloride additions on the subsequent adsorption of gold onto activated charcoal suggested significant differences in the behaviour of the sodium and calcium ions, which were not completely explained in terms of ionic strength alone. While both species considerably aided gold adsorption, calcium additions produced the greatest effects on the charcoal capacity at concentrations well below that of sodium. Likewise, in the presence of an excess of the halide salts, calcium additions resulted in faster reaction rates within the pH range 3 to 11 when compared with sodium. Furthermore, the slopes and shapes of the equilibrium isotherms indicated some difference in the adsorption mechanism when calcium was present as a 'spectator' ion. The formation of a more strongly adsorbed calcium aurocyanide complex would be consistent with these findings.

What was most evident, however, was the inability of gold to adsorb in the complete absence of stabilizing cations such as sodium and calcium ions, suggesting the use of deionized water as an eluant for the recovery of gold from loaded charcoal. The adsorption isotherm carried out under reflux had shown this to be a distinct possibility.

Initial attempts to elute gold from charcoal that had been exposed to calcium ions met with no success, reinforcing the view that the gold was firmly adsorbed as either calcium aurocyanide or hydrogen aurocyanide, depending on the pH of the adsorption medium. Charcoal to which the gold had been added as potassium aurocyanide under neutral pH conditions responded poorly to water elution at 90°C. A pretreatment with sodium carbonate, however, enhanced the elution significantly. The original adsorption of gold as hydrogen aurocyanide rather than as potassium aurocyanide is suggested, the subsequent addition of sodium carbonate then producing the less firmly adsorbed sodium aurocyanide species:

$$H^+ Au(CN)_2^- + \tfrac{1}{2} Na_2CO_3 \rightleftarrows$$

$$Na^+ (Au(CN)_2^- + \tfrac{1}{2} H_2CO_3.$$

The use of a carbonate pretreatment was then tried on the charcoal loaded from gold-plant effluent that had previously failed to respond to elution with hot deionized water. The good recoveries of gold obtained after such a pretreatment strongly indicated the following reaction mechanism:

$$Ca\ Au_2(CN)_4 + Na_2CO_3 \rightleftarrows$$

$$2Na\ Au(CN)_2 + CaCO_3$$

$$Ca\ Au_2(CN)_4 + K_2CO_3 \rightleftarrows$$

$$2K\ Au(CN)_2 + CaCO_3.$$

The better recoveries obtained with potassium carbonate imply that potassium aurocyanide is less firmly adsorbed than is sodium aurocyanide.

Further evidence supporting, if not proving, the above mechanism was obtained from the analysis of the eluates for other adsorbed species (Table III). Gold was observed to elute as potassium aurocyanide, being antipathetic to the elution of both calcium and magnesium. Moreover, the analysis of the original loaded charcoal (Table I) strongly suggested the adsorption of gold as the calcium aurocyanide complex.

The acid leaching of loaded charcoal also provided some valuable information. A potassium carbonate pretreatment followed by water elution yielded a gold recovery of only 47 per cent, indicating either reduction of the loaded gold on the charcoal under such acid conditions (pH approaching 0), or the adsorption of gold as a more firmly held hydrogen aurocyanide complex. That subsequent pretreatment with a potassium carbonate/potassium hydroxide reagent, followed by a water elution, yielded an 89 per cent gold recovery suggests the latter mechanism. While some decomposition of gold to the metallic state (less than 11 per cent) may have taken place, it is nevertheless surprising that total reduction was not effected under such acid conditions.

It thus appears that the degree to which the various cyanide complexes of gold are adsorbed onto surface-active sites (activated charcoal) depends to a very large extent on the character of the complementary cation. Calcium aurocyanide appears to be the most strongly adsorbed complex in the following series:

$$Ca^{2+} > Mg^{2+} > H^+ > Li^+ > Na^+ > K^+.$$

DISCUSSION

The present findings regarding the mechanism of gold adsorption onto surface-active sites immediately emphasizes the important role played by both the calcium cation and the hydroxyl anion. The two species are antipathetic in that high calcium concentrations will favour gold adsorption, while high pH levels will decrease gold adsorption. As such, the findings may have wider ramifications when related to current metallurgical practice where lime is used to maintain pH levels in the cyanidation circuit. The detrimental effect caused by excess calcium[6] may be explained, in part at least, by the adsorption of calcium aurocyanide by gangue materials[7]. As the ionic strength of the adsorption medium has also been shown to play an important role in gold adsorption, the use of foul cyanidation liquors may also be expected to yield poor gold recoveries. The use of sodium hydroxide as a means of maintaining pH levels in the cyanidation circuit is suggested if increased reagent

costs are off-set by higher gold recoveries.

It has been adequately demonstrated that gold can be efficiently eluted from activated charcoal by use of a carbonate pretreatment followed by elution with hot water low in calcium ions. While the efficiency of potassium or sodium carbonate as pretreatment reagents has been shown to be due to the formation of insoluble calcium carbonate, the use of a variety of alkali metal salts producing insoluble calcium compounds may be possible. The use of fluorides, oxalates, and sulphites can be mentioned as examples. On the basis of cost alone, few of these salts can be expected to compete with the use of carbonates. Furthermore, should an excessive build-up occur after repeated use, calcium carbonate can easily be removed from the charcoal by use of dilute acid.

If the proposed technique is compared with other known methods for the elution of gold from activated charcoal[1-4], it appears that, on the basis of ease of operation, economics, handling of reagents, recovery of gold from pregnant eluate, and overall simplicity, the hot-water process has significant advantages. As the major problem in the elution of gold from activated charcoal has to date been the difficulty involved in the re-use of the charcoal, the development of a simple process using hot water as an eluant immediately enhances the use of activated charcoal as a highly selective ion-exchange medium in the field of gold metallurgy.

ACKNOWLEDGEMENT

The author is indebted to the analytical section of Anglo American Research Laboratories for the chemical analyses.

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XIV INTERNATIONAL MINERAL PROCESSING CONGRESS
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ASPECTS OF LABORATORY AND PILOT PLANT EVALUATION OF CIP WITH RELATION TO GOLD RECOVERY

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INTRODUCTION

With the recent advent of several large carbon-in-pulp plants in the Republic of South Africa, considerable emphasis has been placed on a better understanding of the more pertinent parameters affecting gold adsorption onto activated carbon. Such an understanding is imperative if the modelling of carbon-in-pulp circuits is ever to become a meaningful exercise. The quantification of such operational parameters as cyanide concentration, temperature, pH, and pulp density is initially required prior to relating their confounding effects on the process. Furthermore the effect of both the organic as well as the inorganic constituents of the pulp may also be expected to exert a considerable influence on the process.

The complex and as yet ill-defined reaction mechanism pertaining to the adsorption of the metal cyanide complexes of gold, silver, copper, nickel and zinc which are commonly found in such plant solutions, immediately imposes a tremendous constraint on an investigation of this nature. Not only is the adsorption route together with the nature of the adsorbed species not clearly understood at the present time, but the functional groups partaking in the reaction on the carbon surface remain a subject of considerable conjecture. Obviously, a better understanding of the overall reaction mechanism whereby gold is adsorbed onto activated carbon would greatly assist any proposed modelling exercise.

The present investigation covers a wide spectrum of aspects related to the carbon-in-pulp (CIP) process in an effort to relate certain observations to the adsorption mechanism as well as to attempt to quantify such effects.

EXPERIMENTAL METHOD

FACTORS AFFECTING GOLD ADSORPTION

Preparation of carbon for equilibrium testwork. Type G210-AS carbon (1,18 - 2,36 mm) supplied by Le Carbone S.A. (Pty.) Ltd was milled to 75 - 150 µm and the fines discarded. The milled carbon was then boiled for 10 minutes in an equal volume of 3 per cent (vol/vol) hydrochloric acid in order to simulate the acid washing stage in a CIP operation. This was followed by washing the carbon five times with equal volumes of distilled water. The carbon was subsequently titrated to pH 10,0 using dilute sodium hydroxide and an automatic titrimeter. This equilibration was carried out for 16 hours before draining and filtration under vacuum. Prior to use the carbon was oven dried at 110°C.

Preparation of carbon for kinetic testwork. Type G210-AS carbon (1,18 - 2,36 mm) was acid washed and equilibrated prior to oven drying as described above without prior milling.

Preparation of eluted carbon. 150 mL of the milled carbon as prepared above was placed in an elution column and eluted as previously described[1]. The carbon was initially acid washed at 90°C with 1 bed volume of 3 per cent (vol/vol) hydrochloric acid followed by pumping tap water for 30 minutes - both operations carried out at 5 bed volumes per hour. The temperature was then raised to 110°C and the carbon treated with 1 bed volume of 3 per cent sodium cyanide/0,5 per cent sodium hydroxide prior to eluting with 6 bed volumes of distilled water - both operations carried out at 2 bed volumes per hour. After removal of the carbon from the column it was dried at 110°C before further use.

Synthetic plant solution containing 10 g/t gold; 500 g/t calcium and 500 g/t sodium. 16,10 g $CaCl_2.H_2O$ was dissolved in 8 L of freshly distilled water (CO_2 free). 15,40 g Na_2SO_4 (anhydrous) was then dissolved in 500 mL of freshly distilled water (CO_2 free) and added to the above solution with stirring. 0,148 g $K\,Au(CN)_2$ was then added before diluting the composite solution to 10 L and adjusting the pH to 10,0 using a dilute NaOH solution. The final solution was stored in an aspirator under Ascarite to exclude CO_2.

Synthetic plant solution containing 200 g/t sodium cyanide. 200 g/t NaCN was added to the above synthetic solution as and when required.

Determination of equilibrium constants (k values). Six carbon samples, as prepared above were accurately weighed out in separate 250 mL conical flasks, in the mass range 0,01 - 0,08 g. Unless otherwise specified, 100 mL of synthetic solution was then pipetted into each flask and the complete batch of flasks agitated overnight in a shaking waterbath-the temperature having been preset to the required level. After 16-18 hours the samples were filtered under vacuum to minimise any change in temperature and a subsequent shift in equilibrium. This solution was then analysed for gold and from the concentration of gold in solution, the carbon loading for each sample mass was calculated. The equilibrium isotherm of the carbon was then determined using the Freundlich equation-plotting the gold loading of the carbon (kg Au/tC) against the concentration of gold in solution (g/t) on log-log graph paper. The equilibrium constant, k, was then obtained for the gold loading on the carbon in equilibrium with 1 g/t of gold in solution.

Gold adsorption rates. 1 L of synthetic plant solution was measured into a water-jacketed reaction vessel coupled to a thermostatically controlled water-bath via a refrigeration unit (if required). The solution was stirred by means of an anchor-shaped stirrer at 300 rpm while the pH was monitored by electrode. Carbon dioxide was excluded from the reaction vessel by purging the system with a slow flow of compressed air which was first scrubbed through a 1 N NaOH solution.

After allowing the solution to equilibrate at the desired temperature, oven dried carbon as prepared above (0,413 g) was added to the vessel and the time of addition noted. Samples of solution were then taken at specific time intervals and analysed for gold.

Gold analyses. Gold in solution was analyzed by atomic absorption spectrophotometry following solvent extraction.

THERMAL REGENERATION OF CARBON

Used Carbon A composite sample of "used" carbon type G210 was oven dried at 110°C for 20 hours prior to use. The composite was a mixture of eluted carbon resulting from the treatment of President Brand calcine material in the CIP mini-plant (see below). The eluted composite sample contained 35 g/t gold and 88 g/t silver. The virgin carbon used in the miniplant had a capacity constant for gold adsorption of 34 kg Au/tC when determined in standard borate buffer at pH 10.

Thermal Regeneration. Thermal regeneration was carried out in a stainless steel tube (500 mm long x 26 mm internal diameter) positioned in a vertical electrically heated tube furnace.

When the desired reaction temperature had been achieved, distilled water was pumped into the tube through a small gas inlet at the base of the tube at a rate of 45 mL/hour. This rate was maintained throughout the test. When the steam atmosphere was established in the reactor, a weighed sample (5 g) of the dry test carbon was poured into the tube and the time of addition noted. A glass cap with a vent tube was placed on the top of the reactor to inhibit diffusion of air into the reactor.

The above regeneration procedure was carried out in all tests although the subsequent sample cooling procedure was varied, as described below.

Cooling in air. A series of tests were carried out in the temperature range 300°C to 900°C at reaction times of 5, 10 and 15 minutes respectively. These tests were designed to evaluate the effects of temperature and time on the gold adsorption capacity of the regenerated carbon and the mass loss during a test.

At the end of the test period, steam generation was stopped and the reactor removed from the furnace. The carbon sample was then poured into a dry beaker and subsequently oven dried at 110°C for 20 hours before being cooled and weighed.

Cooling in steam. The regeneration procedure described above was followed except that, at the end of the heating period, the reactor was cooled in air and the steam atmosphere maintained for a period of four minutes prior to quenching the carbon in 25 mL of distilled water at ambient temperature. The

carbon was then filtered off and oven dried at 110°C for 20 hours before cooling and weighing. A reaction period of 10 minutes was used since experience gained in the previous tests had indicated that this regeneration period was satisfactory.

Cooling in nitrogen. In this test series, the same sample cooling method described above was followed except that at the end of the 4 minute cooling period in a steam atmosphere, the water supply to the reactor was stopped and a stream of nitrogen gas, at a flowrate of ca. 380 mL/minute passed into the reactor through the gas inlet for a period of 10 minutes: The carbon was then washed out of the reactor with distilled water into a beaker, the carbon filtered off, dried and weighed as described above.

Cooling in carbon dioxide. These tests were a repeat of those outlined above except that carbon dioxide gas was used instead of nitrogen in the cooling procedure.

Determination of equilibrium constants. Test samples of carbon were initially crushed to less than ca. 300 µm and oven dried at 110°C before the determination of the equilibrium capacity constant at pH 10 in borate buffer as previously described[2].

CHEMICAL REGENERATION OF CARBON

Used carbon. 0,84 - 1,20 mm type G210 carbon was placed in an adsorption column and loaded to 1,6 per cent gold using clarified gold plant pregnant solution originating from the Western Deep Levels gold mine as previously described. The loaded carbon was then pretreated with 1 bed volume of 5 per cent sodium cyanide/ 1 per cent sodium hydroxide and gold eluted at 91°C using 5 bed volumes of deionised water. Following this complete adsorption/elution cycle, the capacity constant of the carbon for gold adsorption measured in standard borate buffer[2] at pH 10 was shown to decrease from 37 to 25 kg Au/tC.

Chemical treatment. 10 g of oven dried used carbon was loaded into a water-jacketed glass column (300 mm long x 10 mm internal diameter) maintained at 91°C. The carbon was then treated with various reagents either sequentially or as mixtures. All such treatments were limited to pH levels below 6. When two different reagents were used sequentially, two bed volumes of water was pumped through the column in between the two individual treatments in order to minimise reagent mixing. Following this chemical treatment, the capacity constant of the carbon for gold adsorption at pH 10 was measured in standard borate buffer as before.

CIP MINIPLANT

Two seven stage carbon-in-pulp miniplants were constructed to assist in the design and commissioning of full scale plants within the Anglo American Corporation. The principal objective was to simulate plant operation as well as collate equilibrium and kinetic data under realistic conditions for theoretical analysis.

Pilot plant description. Each miniplant consisted of seven mechanically agitated contactors, each with a working volume of 10 litres. Each contactor had a 850 µm screen of area 100 cm^2 at the outflow which was cleaned by mechanically timed air and water blasts on the discharge side of the screen (air

was blasted for 10 seconds every 3 minutes and the water blasted for 5 seconds every 2 hours). The impellers consisted of three flat blades inclined at 45° to the shaft axis. The feed to the plant was kept constant by a Sigmamotor model T6-S variable speed finger pump. Each contactor contained a contact thermometer which activated a 500 watt heater to maintain the desired pulp temperature.

Pilot Plant operation. Cyanided pulp stored in a 3 m^3 mechanically agitated stock tank was fed to the plant at a flowrate of 10 L/hour. Under continuous operation carbon was sequentially transferred on a daily basis. The carbon and pulp in stage 1 was drained through a 710 μm screen, briefly washed and a small sample taken for analysis, the balance being fully loaded carbon. The carbon and pulp in stage 2 was similarly screened and then placed in stage 1 and this was repeated down to stage 7. The required volume of fresh carbon was then added to stage 7. The principal objective in running the plants was to simulate steady state operation and to this end each run was continued for several weeks.

RESULTS

FACTORS AFFECTING GOLD ADSORPTION

Reversibility of gold adsorption. Preliminary equilibrium experiments using a different batch of 75-100 μm type G210 carbon with 10 and 100 g/t gold solutions respectively, showed a considerable difference in k values (Figure 1) which could be indicative of an irreversible adsorption mechanism in borate buffer. In the presence of 200 g/t sodium cyanide the adsorption isotherms were identical at equilibrium concentrations above 1 g/t gold in solution. The presence of free cyanide would thus appear to be necessary in effecting a reversible adsorption mechanism.

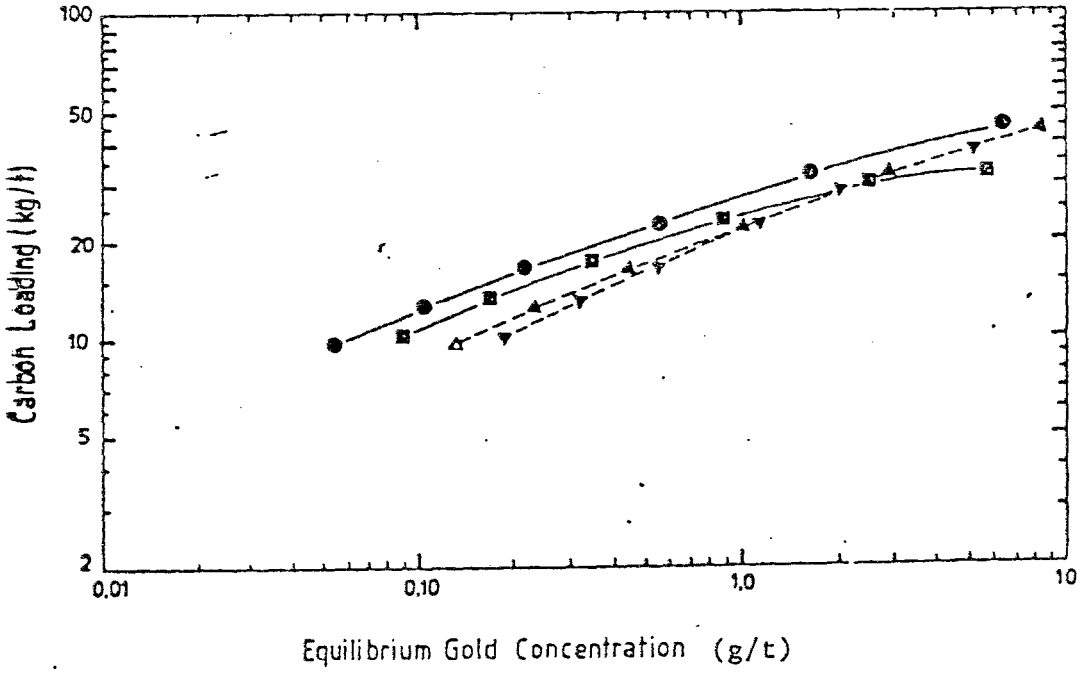

Figure 1 The Effect of Free Cyanide on Equilibrium Gold Loadings at 24°C in Standard Borate Buffer at pH 10.

(● 100 g/t Au initial concentration; ■ 10 g/t Au initial concentration; ▲ 100 g/t Au initial concentration +200 g/t NaCN; ▼ 10 g/t Au initial concentration + 200 g/t NaCN)

In order to simulate plant conditions, the borate buffer was replaced by a synthetic plant solution. Initially the reproducibility was poor compared with results obtained in borate buffer solutions. The pH of these solutions dropped markedly upon standing due to the absorption of carbon dioxide. When the air above the solution was replaced by oxygen or nitrogen, the drop in pH was only 0,1-0,2 units resulting in a considerable improvement in reproducibility due primarily to a tighter control of pH. Experiments initiated with 10, 100 and 200 g/t gold in synthetic plant solution containing 200 g/t sodium cyanide showed identical isotherms. The isotherms were linear between equilibrium concentrations of 100 and 0,1 g/t gold and slightly curved below 0,1 g/t gold. A k value of 36 Kg Au/tC was indicated. This value increased to 38 Kg Au/tC if the contact time was extended to 62 hours indicating a relatively slow approach to equilibrium conditions under the prescribed conditions.

Influence of pH. In these experiments the pH of the synthetic plant solution was adjusted to 9,5, 10,0 and 10,5 respectively before contacting the carbon. The pH drop during equilibrium was 0,2-0,4 pH units (Figure 2). As may be expected, an increase in adsorption capacity with decrease in pH may be noted.

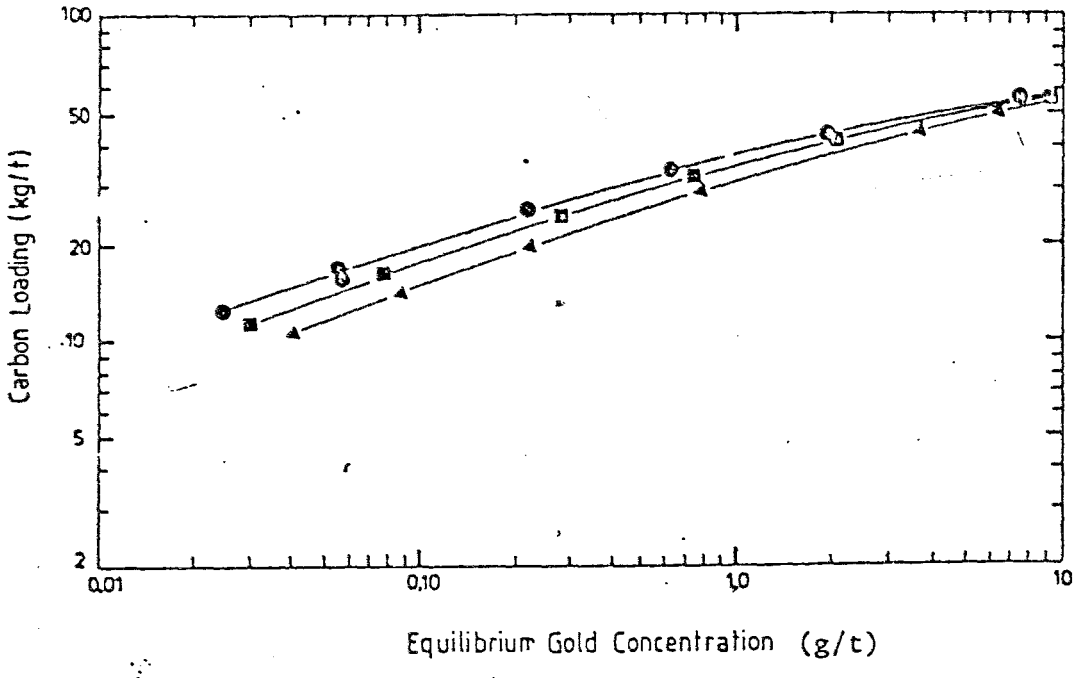

Figure 2 The Effect of pH on Equilibrium Gold Loadings in Synthetic Plant Solution at 25°C.

(-●- pH 9,5; -■- pH 10,0; -▲- pH 10,5)

Influence of atmosphere. When the atmosphere above the synthetic plant solution was replaced by oxygen, identical isotherms as with air resulted. Using nitrogen, the k value dropped to 28 compared with 36 obtained with air or oxygen (Figure 3). This evidence would suggest that oxygen is consumed in some way during gold adsorption.

Replacement of calcium by magnesium. Calcium in the synthetic plant solution was replaced by the same stoichiometric amount of magnesium (as $MgCl_2$). Virtually no change in the resulting isotherms was noted indicating that other than an ionic strength effect, calcium does not influence the adsorption capacity of the carbon under the prescribed conditions (i.e. in the virtual absence of CO_2).

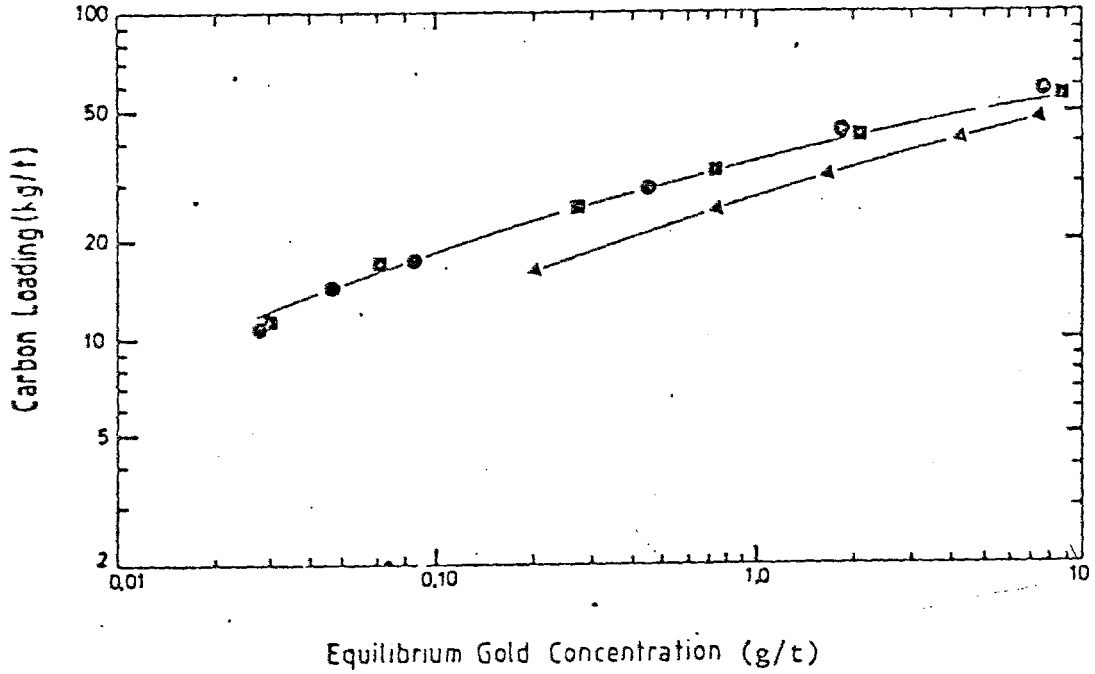

Figure 3 The Influence of the Atmosphere on Gold Adsorption from Synthetic Plant Solution at 25°C.

(-■- air-CO_2 free; -●- oxygen-CO_2 free; -▲- nitrogen)

Effect of cyanide concentration and temperature on adsorption equilibria. Equilibrium capacity data relating to the effect of temperature on gold adsorption in both the presence and absence of free cyanide is illustrated graphically in Figure 4. It is immediately evident that the presence of free cyanide (Series B) in the synthetic plant solution has a very considerable effect on gold loadings under the prescribed conditions when compared with the use of a similar solution containing no added cyanide (Series A). This detrimental effect of cyanide is seen to increase significantly with increasing temperature and a gold loading at 15°C of 41 kg Au/tC may be compared to 35 kg Au/tC, while at 40°C a loading of 28 kg Au/tC may be compared to 16 kg Au/tC. The decrease in gold loading at 15°C is therefore only 14 per cent but the loss at 40°C decreases further to 40 per cent. Likewise, the effect of temperature on the gold loading is a marked one. Within the prescribed temperature limits, the effect would appear to be linear both in the presence and absence of cyanide. The slope of the linear relationship increases from 0,50 kg Au/tC/°C in the absence of cyanide to 0,78 kg Au/tC/°C upon the addition of 200 g/t NaCN.

Although the cyanide concentration of the initial synthetic plant solution was 200 g/t NaCN for Series B, the final equilibrium cyanide concentration had dropped to 170-130 g/t NaCN after 16-18 hours depending on the mass of carbon added in each instance. These cyanide concentrations were not, however, monitored on a routine basis.

Equilibrium gold loadings (k values) were also determined at 25°C for 0,2, 5, 10, 25, 50, 100, 150, 200, 250, 400 and 500 g/t NaCN added to the synthetic plant solution. The cyanide was either weighed directly into conical flasks prior to addition of the synthetic plant solution or added using an Oxford sampler from a stock solution in the volume range of 0,1-0,5 cm^3. In all cases the pH was controlled after the addition of solution and where necessary a small volume of diluted sulphuric acid added to bring the pH back to 10,0. The addition of either the cyanide or sulphuric acid solution was negligible compared to the volume of synthetic plant solution. Equilibrium data are presented in Figure 5 where it can be seen that the addition of the first 10 g/t NaCN has a sharp effect on the lowering of the equilibrium constants. From 10 g/t to 50 g/t NaCN the decrease is steadier but after this point to the final one at 500 g/t NaCN the values of the equilibrium loading constant decrease slowly but consistently. These results suggest that once the carbon has become cyanide saturated a further excess of cyanide in the synthetic plant solution has relatively little effect on the adsorption of gold onto the carbon.

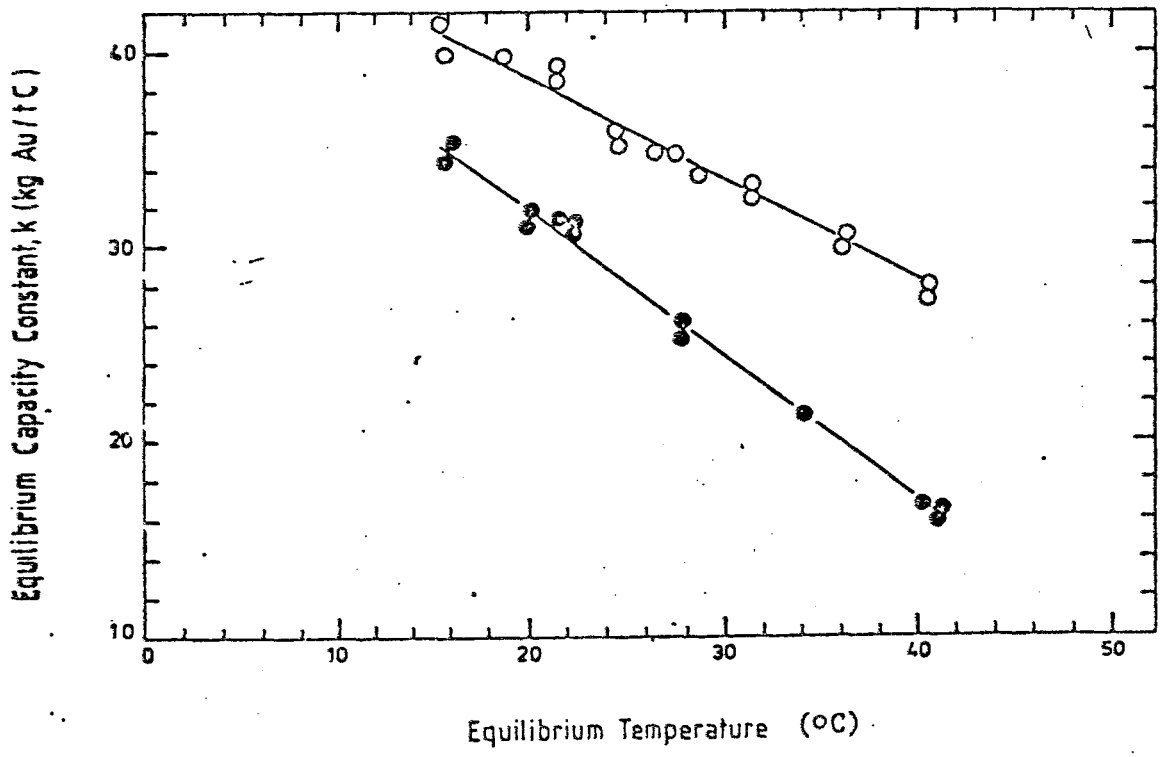

Figure 4 The Effect of Temperature on Equilibrium Gold Loadings in Synthetic Plant Solution at pH 10.

(-o- Series A - no cyanide present; -●- Series B - 200 g/t NaCN added)

Effect of temperature on adsorption rates. All kinetic studies were carried out in the presence of 200 g/t NaCN as it was felt that the results obtained would

be more meangingful and representative of the CIP process.

Adsorption rates were measured at 15, 20, 21, 23, 25, 28, 30, 35 and 40°C. After several modelling attempts, the Langmuir equation was found to depict the rate of adsorption of gold on carbon fairly well. A plot of time/gold loading versus time was found to be linear, fitting the expression:

$$y = \frac{t}{a + bt}$$

where 1/a = the rate of adsorption at zero time (a is the intercept on the y axis) (kg Au/tC/min)

1/b = the loading of the carbon at infinite time (b is the slope of the linear function) (kg Au/tC).

and t = reaction time (min)

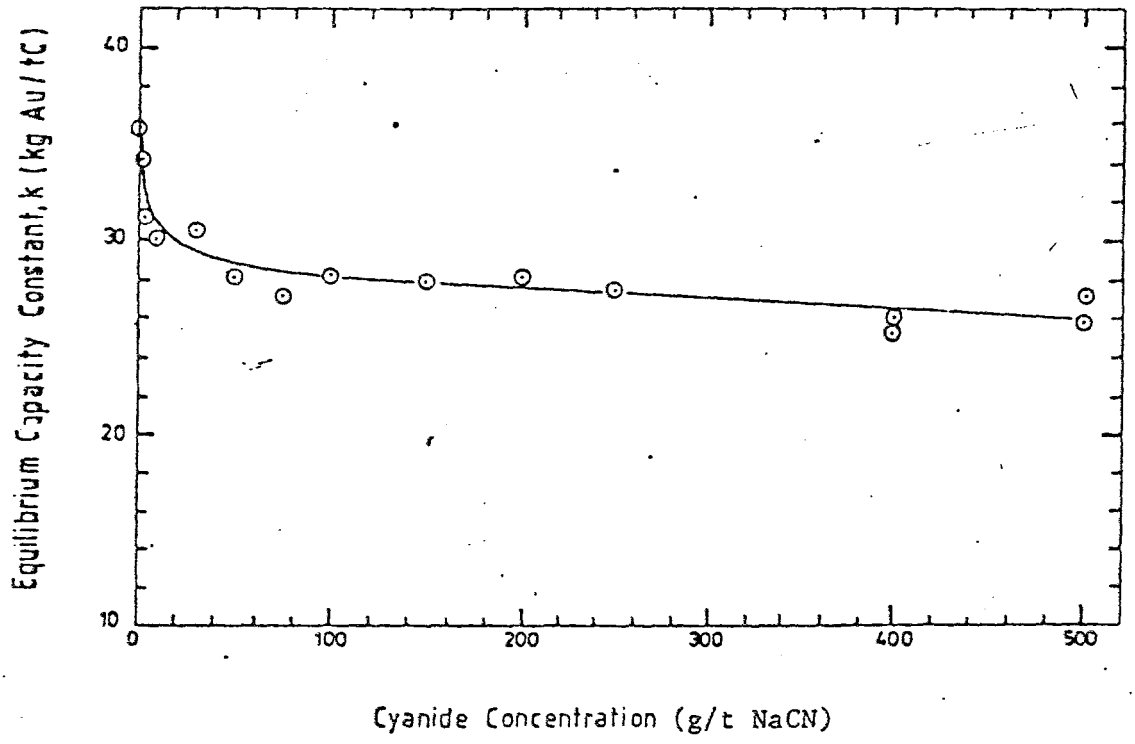

Figure 5 The Effect of Cyanide Concentration on Equilibrium Gold Loadings at 25°C in Synthetic Plant Solution at pH 10.

Values for the above constants as a function of temperature (15-40°C) are graphically illustrated in Figure 6. As may be expected, the data indicate that while initial gold adsorption rates increase significantly as the temperature rises, gold loadings approaching equilibrium are lowered with increasing temperature.

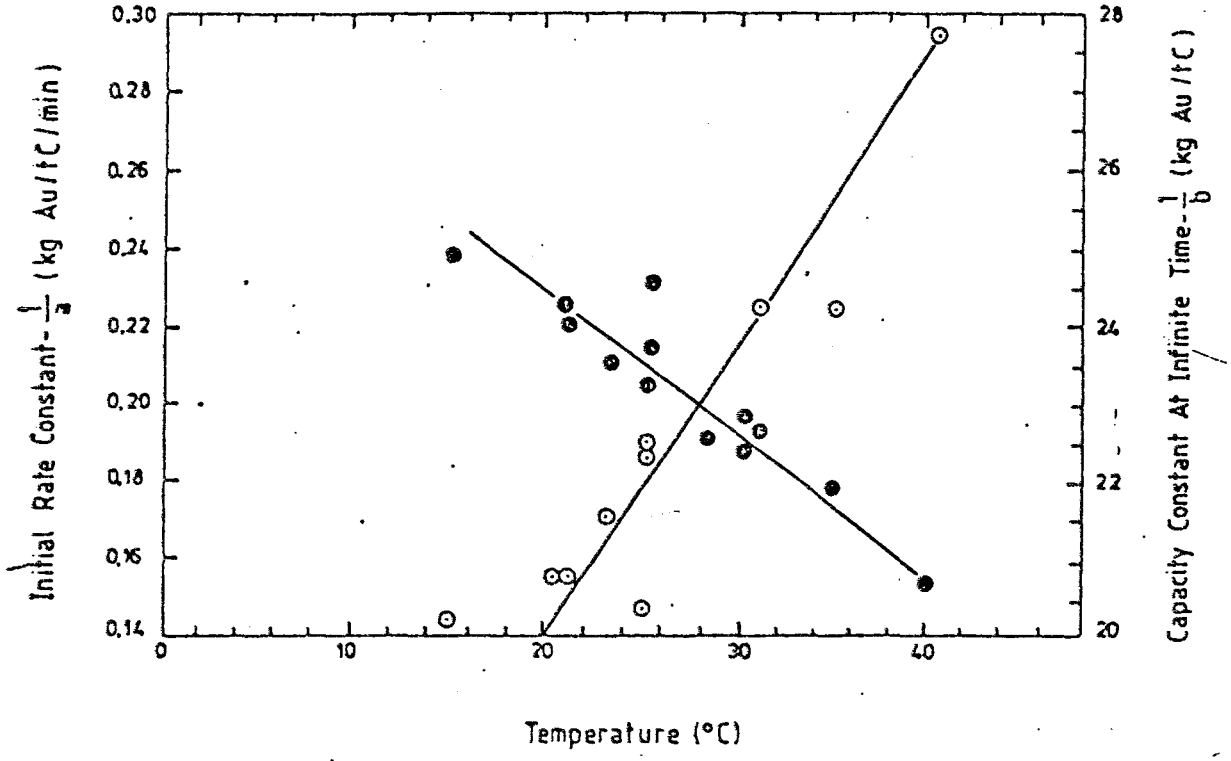

Figure 6 The Effect of Temperature on Gold Adsorption Rates in Synthetic Plant Solution at pH 10.

(-O- initial rate constant - 1/a; -●- capacity constant at infinite time - 1/b)

Equilibrium gold loadings on eluted carbon. As loaded carbon was eluted using a relatively strong cyanide pretreatment reagent (1 bed volume 3-5 per cent NaCN/1 per cent NaOH), the activity of virgin eluted carbon was compared with that of the virgin carbon. The determination of equilibrium loading constants for both carbons was carried out using a synthetic plant solution containing 200 g/t NaCN. The resulting equilibrium data indicated that the prescribed elution procedure had little or no effect on subsequent gold adsorption.

The effect of added ammonia on equilibrium gold loadings. It has been adequately demonstrated that the addition of 200 g/t NaCN to the synthetic plant solution had a most detrimental effect on equilibrium gold loadings in the temperature range 15-40°C. In order to define the species responsible for this effect i.e. CN^- or its subsequent oxidation product NH_4^+, NH_4Cl was added to the plant solution prior to the determination of adsorption constants at 25°C as previously described. The concentration of NH_4^+ added as NH_4Cl was equivalent to the NH_4^+ concentration resulting from the complete oxidation of 200 and 400 g/t NaCN respectively otherwise present in such a synthetic plant solution. The resulting equilibrium data clearly indicated that the addition of NH_4Cl had a small but significant effect on improving gold loadings. An increase in k values from 36 to 38-41 kg Au/tC was observed.

THERMAL REGENERATION OF USED CARBON

Cooling in air. Reaction times of 5, 10 and 15 minutes respectively, before the air cooling of samples, resulted in a sharp increase in the carbon mass loss between 600°C and 700°C, increasing rapidly to a maximum at 900°C. Carbon mass loss data following a 15 minute reaction time is presented in Figure 7. The increase in mass loss was less pronounced in the tests with a reaction time of 5 minutes and most significant in tests with a reaction time of 15 minutes. This suggested that partial combustion of the carbon may have occurred. On the other hand, the carbon mass loss may be partially due to the reaction of steam on the heated carbon followed by a combustion loss when the hot carbon is exposed to air (this would have to be substantiated by analysis of the reactor discharge gases).

An almost linear increase in the gold capacity constant with temperature to a maximum value of 35 kg Au/tC at 600°C in the tests with reaction times of 10 minutes and 15 minutes was observed. A more erratic increase with temperature occurred in the case of the tests with a reaction time of 5 minutes, reaching a maximum of 34 Kg Au/tC at 700°C. A minimum reaction time of 10 minutes would appear to be necessary to effect efficient carbon regeneration at 600-700°C under the prescribed experimental conditions. Equilibrium capacity data following a reaction time of 15 minutes is illustrated in Figure 7.

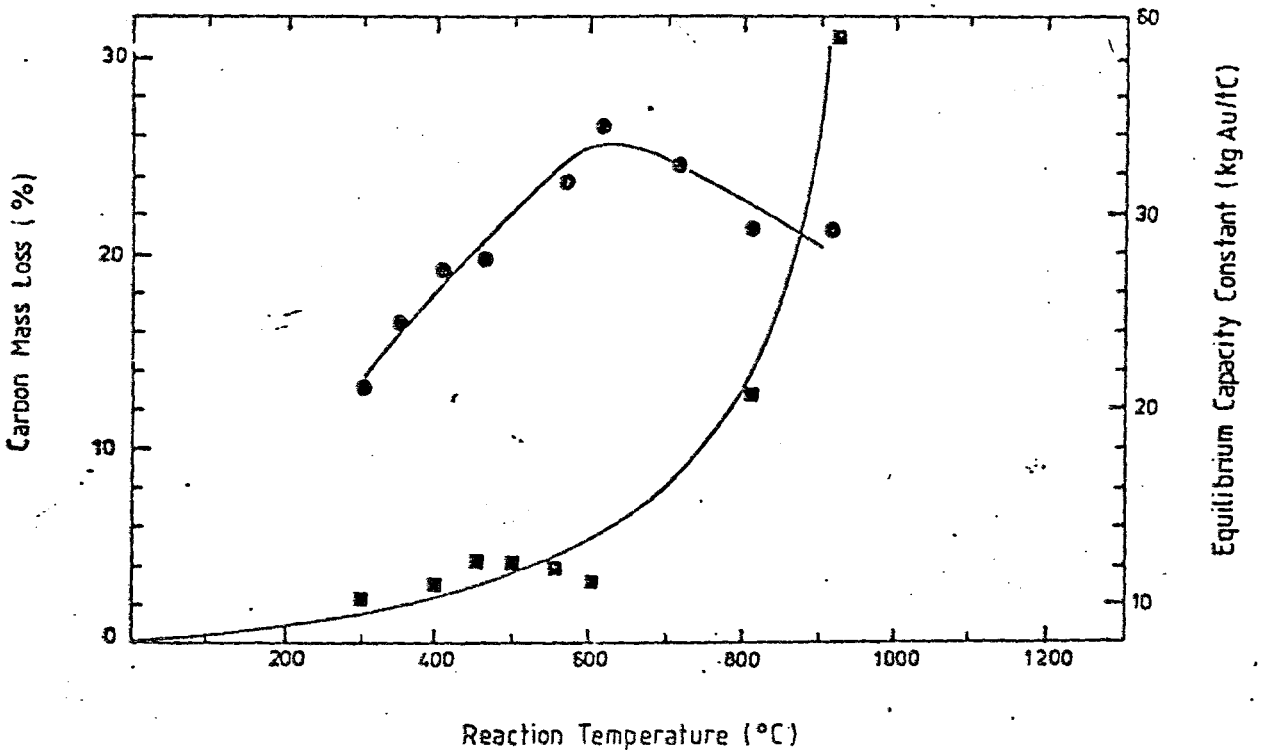

Figure 7 The Thermal Regeneration of Carbon - Samples Cooled in Air.

(-●- equilibrium capacity constant; -■- mass loss of sample)

Cooling in steam. The carbon mass losses when cooling in steam (Figure 8) were of the same order as those reported when air cooling was used except at a reaction temperature of 900°C a mass loss of ca 14 per cent may be noted compared with a mass loss of ca 20 per cent when the carbon was air cooled. This could indicate that at temperatures below 900°C, the most substantial part of the

mass loss occurred during the reaction period and not on exposure to cooling air and would support the hypothesis that carbon mass losses are partially or wholly due to the reaction of carbon with steam during the reaction period.

The resulting gold capacity constants in this instance (Figure 8) were greater in every case by ca 4 units when compared with air cooled tests where similar reaction parame ers were applied. A maximum value of 38 kg Au/tC was obtained at 600°C when cooling in steam which may be compared with a maximum of 35 kg Au/tC at 600°C obtained in the air cooled test series (10 minute reaction time). These data imply that regeneration at temperatures greater than 600°C in a steam medium will not improve the gold adsorption capacity of the carbon. Furthermore direct air cooling of the heated carbon may well detrimentally affect the gold adsorbing capacity of the carbon.

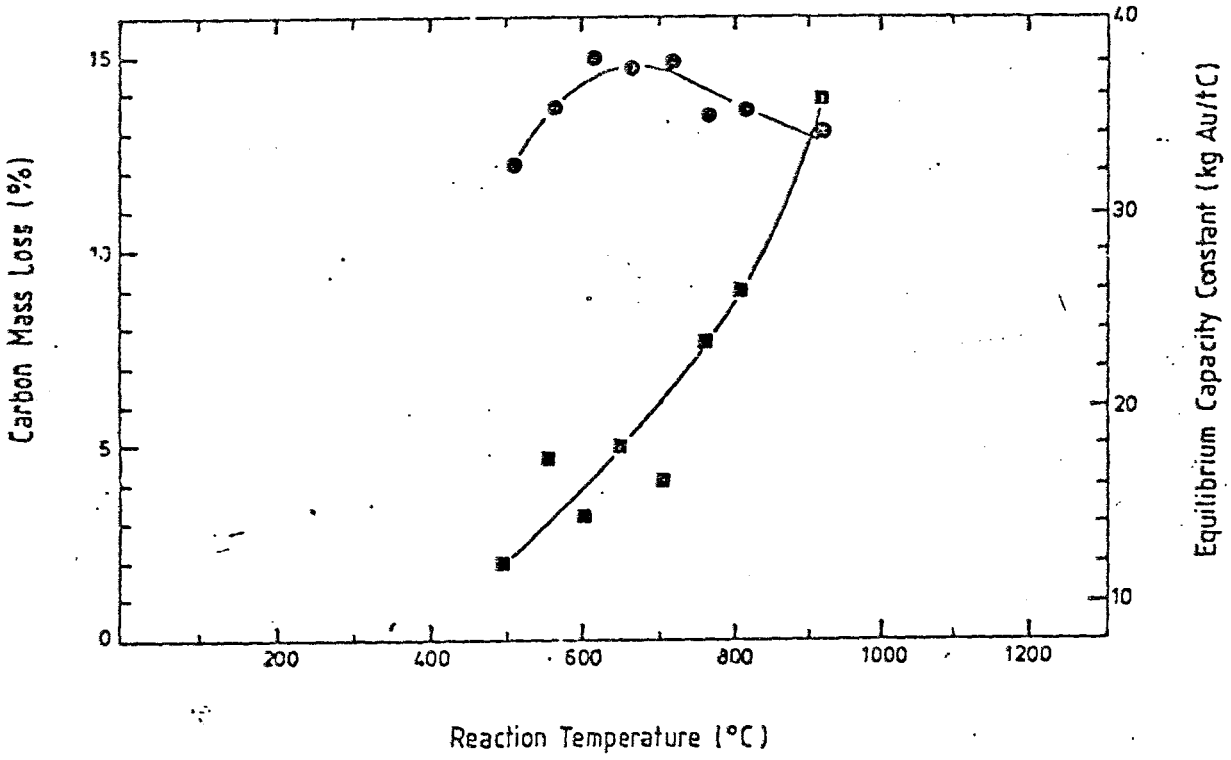

Figure 8 The Thermal Regeneration of Carbon - Samples Cooled in Steam.

(-●- equilibrium capacity constant; -■- mass loss of sample)

Cooling in nitrogen gas or in carbon dioxide gas. Although these tests were of an exploratory nature only and limited in number, it is interesting to note that the highest adsorption capacity values, in both the nitrogen series and the carbon dioxide series, were obtained at 700°C with a 10 minute reaction time; viz. 39 Kg Au/tC (N_2) and 41 kg Au/tC (CO_2) respectively. These values are considerably higher than the virgin carbon (34 Kg Au/tC) and a little higher than the maximum value of 38 Kg Au/tC obtained in the steam cooled test series. Further tests would be necessary in order to establish optimum test parameters since in both test series, capacity values increased with

temperature. Carbon mass losses, in both test series, peaked at 600°C with a ca 8 per cent mass loss (not all of the test data is presented).

The above evidence may indicate that the adsorption of carbon dioxide and/or nitrogen also play some role in the gold adsorption mechanism (see below).

CHEMICAL REGENERATION OF CARBON

From the data presented in Table I it is immediately evident that of the cyanide complexing reagents studied (nickel sulphate, zinc sulphate ferrous sulphate and copper sulphate), only solutions of copper sulphate had any noticeable effect on increasing the capacity of the used carbon for gold adsorption. An increase from 25 to 28 Kg Au/tC using a 2-3 per cent copper sulphate solution may be noted. Likewise, when the complexing reagent treatment of used carbon was followed by an additional treatment of the carbon with a solution of ammonium carbonate, only the use of a copper sulphate solution was effective in regenerating the carbon. In this instance, an increase in the capacity from 25 to 35-36 Kg Au/tC was observed. The use of solutions of 2-5 per cent ammonium carbonate in treating the carbon was shown to increase the capacity from 25 to 28 Kg Au/tC. The use of 2 per cent copper sulphate followed by 2-5 per cent sodium bicarbonate was shown to increase the capacity constant to 31-32 Kg Au/tC.

Optimum chemical regeneration of the used carbon was attained using either a mixture of 2-2,5 per cent copper sulphate in carbonated water or the sequential use of 2-2,5 per cent copper sulphate solution followed by the treatment of the carbon with carbonated water. In both instances an increase in capacity from 25 to 37-38 Kg Au/tC was observed indicated the virtual complete regeneration of the carbon (the virgin carbon had a capacity constant for gold adsorption of 37 Kg Au/tC.

The above data indicate that the adsorption of both copper and carbon dioxide significantly improve gold adsorption and as such play a role in the adsorption mechanism.

It has also been demonstrated at these laboratories that treatment of used carbons of low activity with various metal halides (H_2PtCl_6; VCl_2; $AuCl_3$; $FeCl_3$; $SnCl_3$; $TiCl_3$; AgCl) resulted in a considerable improvement in gold adsorption capacity. A significant increase in the zeta potential of the treated carbons was also noted.

CIP MINI-PLANTS

President Brand G.M. Calcines. Tables II and III give the mean final steady state results for each of the runs which are the basis for the discussion. The following parameters were varied: pulp pH, pulp density, free cyanide concentration in feed solution, carbon concentration, carbon flow rate, stirrer speed, temperature and type of carbon.

The mini-plant ran with a pulp pH of 8 in Run 1 and with a pulp pH of 10-11 thereafter. There are no strictly comparable data at pH 10, (the carbon parameters of Run 1 being an approximation to the design parameters of the President Brand plant) but the increase in pH from 8 to 10 did not appear to have any noticeable effect on the barren solution gold assay. The copper adsorption onto the carbon dropped from 34 930 g/t at pH 8 to 893 g/t at pH 10

(Table II). Calcium adsorption onto carbon did not increase significantly with the increase in pH.

For Run 6 the pulp density was reduced from 1,50 to 1,45. This gave an appreciably lower pulp viscosity and hence better mixing. At the same time the stirrer speed was reduced from 200 to 150 rpm and minor modifications were made to the impellers.

TABLE I The Chemical Regeneration of used Carbon

"Used" carbon was treated in a column at 91°C with various solutions (A) and an ammonium carbonate solution (B) either individually or sequentially as outlined below. Reagents were pumped at 2 bed volumes/h while 2 bed volumes of water was pumped at the same flowrate after each different reagent treatment.

Reagent A concentration/ treatment time (h)	Reagent B concentration/ treatment time (h)	Treatment Sequence	Equilibrium capacity constant after treatment (kg Au/tC)
NIL	NIL		25
0,5% $NiSO_4.6H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	25
1,0% $NiSO_4.6H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	27
2,0% $NiSO_4.6H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	26
	2% $(NH_4)_2CO_3$/2,5	B only	28
	3% $(NH_4)_2CO_3$/2,5	B only	28
0,5% $FeSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	25
1,0% $FeSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	27
2,0% $FeSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	26
0,5% $ZnSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	24
1,0% $ZnSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	24
2,0% $ZnSO_4.7H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	24
2,0% $CuSO_4.5H_2O$/1,0		A only	28
0,5% $CuSO_4.5H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	32
1,0% $CuSO_4.5H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	33
2,0% $CuSO_4.5H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	35
3,0% $CuSO_4.5H_2O$/0,5	2% $(NH_4)_2CO_3$/2,5	A + B	36
2,0% $CuSO_4.5H_2O$/1,0 saturated with CO_2		A only	37
3,0% $CuSO_4.5H_2O$/1,0 saturated with CO_2		A only	38

Previous runs had been at a sodium cyanide feed concentration of about 50 mg/L which was the mean value that the 200 mg/L concentration at the end of cyanidation fell to by the time the pulp entered the mini-plant. Since cyanide is known to inhibit gold adsorption on activated carbon it was decided that it would be more realistic to readjust the cyanide concentration in the feed daily to 200 mg/L free NaCN and this procedure was started in Run 6.

However, the free cyanide concentration still fell to about 50 mg/L in the mini-plant probably due to the catalytic effect of carbon on the oxidation of cyanide. Run 6 thus differed from Run 5 in three respects:- density, agitation and cyanide concentration. The change left calcium loading on carbon substantially unchanged but reduced copper loading from 3 000 to 900 g/t. The gold and silver analyses on the barren solutions increased from 0,006 and 0,024 to 0,013 and 0,034 mg/L respectively. Since the changes in density and agitation conditions were small, the drop in performance is tentatively attributed to the higher free cyanide cyanide concentration and/or the drop in copper loadings. The adsorption of copper onto carbon is very sensitive to both pH and cyanide concentration[4].

Carbon concentrations were varied from 30 g/L (Run 2), to 25 g/L (Run 3), and then to 15 g/L (Run 4) at constant carbon and pulp flow rates. The results indicated that lower carbon concentrations lead to lower gold and silver values in the tail solution. This is contrary to what would be expected on kinetic grounds, especially when it is appreciated that even in Run 2 the loadings were far short of equilibrium. One possible explanation is that owing to the short duration of Runs 2 and 3, steady state was not reached. Again, the high copper loadings effected in Run 1 may have benefited Runs 2 and 3.

Another aspect is that in Run 4 the whole stock of carbon in each cell was moved on each day giving an actual carbon residence time of one day per stage whereas in the other two runs only part of the carbon was moved on each day giving a distribution of carbon residence times. It is suggested that the system in which there is a distribution of carbon residence times is inherently less efficient than the plug flow situation where all particles have the same residence time, and this would help to explain the anomalous effect of carbon concentration on barren solution values.

No comparative test results in which carbon flow rate was altered at a constant carbon concentration are available. However, if the parameters are considered in another way, Runs 4 and 5 can be compared. From Run 4 to Run 5 the carbon concentration was reduced from 15 to 10 g/L and the carbon flow rate from 150 to 100 g/day i.e. the carbon residence time remained at an actual one day per stage. With the drop in carbon flow rate the tail solution gold and silver values rose from 0,002 and 0,003 to 0,006 and 0,024 mg/L respectively. The lower carbon flow rate must give higher gold loadings and this will lead to higher barren solution values, but the especially high rise in silver tail solution values is attributed to crowding out by gold loading on the carbon.

Except for the carbon concentrations which were those of the original process specifications for the President Brand Plant, the conditions of Run 9 approximated to the most severe conditions encountered on the President Brand Plant, with a pulp temperature of 40°C, a pH of 11 and a free sodium cyanide concentration of 400 mg/L. This gave barrens analysing 0,01 mg/L gold and 0,03 mg/L silver. These data are best compared with those of Run 1 which had the same carbon parameters but lower temperature, pH and free sodium cyanide

concentrations and somewhat different agitation conditions - Run 1 gave barrens of 0,003 mg/L gold and 0,002 mg/L silver. All three solution parameters (temperature, pH and free sodium cyanide concentration) will have acted in the same direction to give the higher barrens assays in Run 9, and the effect on silver is especially large. Although runs were not continued long enough to reach steady state, a comparison of the terminal barren solution assays indicated that with the Metsorb 107 type carbon a significant rise in gold barrens occurred (this carbon had a capacity constant of 20 kg Au/tC when measured in borate buffer[2]).

TABLE II Summary Data for CIP Treatment of President Brand C.M. Calcines

Run	Feed Solution		Barren Solution		Loaded Carbon				Pulp pH	Free NaCN In Feed Soln.	Carbon Concentration	Carbon Flow Rate	Duration of Run
	Gold mg/L	Silver mg/L	Gold mg/L	Silver mg/L	Gold g/t	Silver g/t	Calcium g/t	Copper g/t		mg/L	g/L	g/d	d
1	7,6	1,06	0,003	0,001	7510	943	8180	34830	8,0	50	24 in cells 1-4 32 in cells 5-7	190	24
2	7,7	1,43	0,004	0,013	9370	1567	7420	7570	10,0	50	30	150	3
3	7,5	1,26	0,005	0,002	8290	1480	7120	2390	10,0	50	25	150	5
4	6,8	1,20	0,002	0,003	10070	2133	7370	5630	10,0	50	15	150	10
5	8,0	1,46	0,006	0,024	13230	2188	9500	3040	10,0	50	10	100	19
6	7,3	1,37	0,013	0,034	13720	2167	6400	893	10,0	200	10	100	24
7	6,5	1,39	0,020	0,060	12360	2130			11,5	200	10	100	22
8*	6,9	1,20	0,093	0,205	6870	750			11,0	400	24 in cells 1-4 32 in cells 5-7	190	13
9	6,6	1,20	0,011	0,319	6940	400			11,0	400	24 in cells 1-4 32 in cells 5-7	190	16

*Metsorb 107 type carbon used (Davy Filtron (Pty) Ltd).

NOTE: Runs 1-7 were carried out at 20-25°C while Runs 8-9 at 40°C.
Runs 1-5 were carried out at a pulp density of 1,50 while Runs 6-9 at a pulp density of 1,45.

TABLE III Recovery and Loading data for CIP Treatment of President Brand G.M. Calcines

Run	Stage 1	Stage 2	Stage 3	Stage 4	Stage 5	Stage 6	Stage 7
	Gold in solution (mg/L) / Progressive solution Recovery (%)						
1	3,69 / 52,69	1,03 / 86,79	0,18 / 97,69	0,029 / 99,63	0,006 / 99,92	0,005 / 99,95	0,003 / 99,96
2	2,74 / 64,42	0,41 / 94,68	0,09 / 98,83	0,017 / 99,78	0,008 / 99,90	0,005 / 99,94	0,003 / 99,96
3	1,57 / 79,07	0,30 / 96,00	0,04 / 99,41	0,010 / 99,87	0,005 / 99,93	0,004 / 99,95	0,004 / 99,95
4	1,79 / 73,68	0,37 / 94,56	0,03 / 99,56	0,015 / 99,78	0,003 / 99,96	0,001 / 99,99	0,001 / 99,99
5	3,10 / 61,25	1,06 / 86,75	0,34 / 95,75	0,104 / 98,70	0,029 / 99,64	0,008 / 99,90	0,004 / 99,95
6	3,34 / 54,25	1,34 / 81,64	0,56 / 92,33	0,230 / 96,85	0,076 / 98,96	0,022 / 99,70	0,007 / 99,90
7	3,26 / 49,85	1,93 / 70,31	0,97 / 85,08	0,51 / 92,15	0,237 / 96,35	0,111 / 98,29	0,041 / 99,37
8	4,22 / 38,84	3,10 / 55,07	1,85 / 73,12	1,15 / 83,33	0,578 / 91,62	0,268 / 96,12	0,075 / 98,91
9	4,09 / 38,03	2,69 / 59,24	1,39 / 78,94	0,652 / 90,12	0,194 / 97,06	0,036 / 99,45	0,006 / 99,91
	Gold on Carbon (kg/t)						
1	7,470	3,050	0,910	0,190	0,055	0,030	0,030
2	8,640	2,400	0,500	0,140	0,045	0,025	0,020
3	8,400	2,180	0,450	0,120	0,030	0,030	0,010
4	7,830	3,240	1,180	0,110	0,035	0,020	0,010
5	13,400	5,300	1,870	0,710	0,260	0,130	0,060
6	14,180	7,240	3,830	1,450	0,650	0,210	0,070
7	11,376	6,617	3,327	1,733	0,867	0,393	0,167
8	6,430	4,850	3,490	2,470	1,500	0,850	0,320
9	7,140	5,240	3,650	2,400	1,150	0,420	0,110

DISCUSSION

The evidence presented in this study using both synthetic and plant solutions would indicate many potential benefits resulting from the use of marginally lower pH levels in a conventional CIP type operation. From a chemical point of view the advantages effected would be numerous and may be listed as follows:

(a) Fouling of carbon with calcium carbonate is considerably reduced with the formation of the more soluble calcium bicarbonate. At pH 8,3 $(CO_3^{=}) \approx (HCO_3^{-})$. Severe fouling has been encountered in several CIP operations in South Africa with calcium loadings approaching 50-60 kg/t in some instances.

(b) The detrimental effect of high cyanide concentrations as well as high pH levels on gold adsorption has been well established. High free cyanide concentrations are significantly reduced at lower pH levels. At pH 9,3 $(CN^{-}) \approx (HCN)$, while at pH 7 there is virtually no free cyanide.

(c) The oxidation of cyanide in the presence of activated carbon is enhanced at lower pH levels releasing carbon dioxide and ammonia into solution. Optimum oxidation is effected at pH 6-8 and is catalyzed by the presence of copper[5]. The presence of oxygen, carbon dioxide and ammonia have all been shown to benefit gold adsorption.

(d) Both the capacity and kinetic response of activated carbon is improved at lower pH levels probably as a result of (a), (b) and (c) above.

(e) At conventional pH levels and in the presence of cyanide, the adsorption mechanism is reversible. However, at lower pH levels (see (b) above) evidence presented suggests that the mechanism becomes irreversible.

(f) The displacement of silver by gold at high gold loadings is reduced at lower pH levels[3-4].

(g) The enhanced adsorption of copper at lower pH levels may well enhance gold adsorption. The ability of copper to increase the capacity of activated carbon ("chemical" regeneration) and to precipitate gold from acidified cyanide solutions would support this hypothesis. CIP miniplant data indicate improved gold recoveries concomittant with high copper loadings.

When this evidence is related to the gold adsorption mechanism as proposed by McDougall et al[6] strong support relating the specific character of gold adsorption with the solubility of the adsorbed species is evident. The formation of an insoluble sub-stoichiometric $AuCN_x$ surface species akin to the specific adsorption of an anion on a cathode surface is suggested. Such a mechanism would explain how decreases in temperature, cyanide concentration and pH together with increased copper loadings all tend to decrease the solubility of the reduced adsorbed species.

From a practical point of view it would thus appear that in a CIP type operation, pH is a most sensitive operational parameter requiring very careful and selective control. The optimum level of pH control will of course depend on many aspects related to a specific operation - grade, soluble gold/copper ratio, acid consumption and many other cost considerations, all of which will exert an influence. Depending on the circuit concerned, the benefits would appear to

involve either the lowering of the carbon concentration in the adsorption circuit (decrease bullion lock-up, minimise carbon attrition, improve screen duty) or to increase plant throughput; both being possible owing to improved adsorption rates. Both options would appear to make sound economic sense.

ACKNOWLEDGEMENTS

This paper is published with the permission of the management of the Anglo American Corporation. The authors are indebted to Miss T.P. Haas, Messrs D.A. Temple, J.D. Singer, V. Veronese and H. Weichselbaum for various experimental data. Thanks are also due to the analytical laboratory of Anglo American Research Laboratories for the chemical analyses. Funding of the project by the Gold Division of the Anglo American Corporation is also gratefully acknowledged.

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2.0 TREATMENT OF PLANT PULPS

Aspects related to the treatment of plant pulps in both conventional CIP contactors and in moving bed type contactors are presented. Innovations in interstage carbon screens are described as well as the development of the AAC-Delkor linear screen.

2.1 CARBON-IN-PULP

The effect of pulp alkalinity and cyanide concentration on plant performance currently receives little attention on existing CIP plants due primarily to poor control of cyanide and lime reagent additions. The effect of these parameters on plant performance was studied in a prolonged pilot investigation at the AARL (see appended paper) and clearly indicated the important role played by both cyanide concentration and pH in effecting optimum plant performance. With the recent development of automated monitoring and control equipment, improved metallurgical efficiency in existing plants is certainly possible while also effecting significant savings in reagent cyanide and lime.

DAVIDSON, R.J. A pilot plant study on the effect of cyanide concentration of the CIP process. GOLD 100. Proceedings of the International Conference on Gold. Volume 2: *Extractive Metallurgy of Gold.* Johannesburg, SAIMM, 1986

A Pilot Plant Study on the Effects of Cyanide Concentration on the CIP Process

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Two CIP pilot plants were used to investigate the effects of free cyanide concentration and pH on gold adsorption. Temperature, pH and cyanide concentration were all controlled and the plants allowed to approach steady state, in order to generate meaningful metallurgical data. In general, the investigation demonstrated the critical role played by both the cyanide concentration and the pulp alkalinity (pH) in the CIP process, and the need to control both parameters in effecting optimum metallurgical performance.

Introduction

Under plant conditions, the selective adsorption of gold onto activated carbon is thought to relate to the solubility of the ion paired species $Ca^{2+}\left[Au(CN)_2^-\right]_2$. Conditions favouring gold adsorption include low temperature, low cyanide concentrations and low alkalinity together with the adsorption medium displaying a high ionic strength and containing Ca^{2+} ions. It has been amply demonstrated[1,2] from a thermodynamic point of view that gold adsorption onto carbon is adversely affected by the presence of free cyanide in the adsorption medium and this effect becomes more pronounced with any increase in temperature. Also, the free cyanide concentration has a significant effect on the selectivity of carbon for gold when dealing with typical plant solutions or pulps. Under typical plant conditions activated carbon is most selective in adsorbing gold from such complex mixtures. Selective metal adsorption appears to take the form of a complex displacement sequence with gold followed by silver being the most strongly adsorbed metal species. No well defined sequence of metal displacement may be formulated as the adsorption of such metal values is sensitive to such processing parameters as pH, free cyanide concentration, temperature and of course the relative concentration of these metal values in a particular plant solution or pulp. In general, the selectivity of carbon for gold will increase with an increase in pH or free cyanide concentration.

While the presence of a divalent cation like calcium would appear to play a significant role in enhancing gold adsorption,[3,4] the simultaneous deposition of calcium carbonate which occurs on the carbon in a plant situation has a passivating effect on the rate of gold adsorption. Although this detrimental effect has not been satisfactorily documented and quantified, it would appear that calcium loadings of 10-20 kg/t have only a marginal effect on

adsorption rates while loadings of 50 kg/t result in a serious decline in adsorption rates. As it is well known that activated carbon readily adsorbs oxygen and acts as an oxidation catalyst for the oxidation of free cyanide[5] it may be presumed that the free cyanide and oxygen in a plant circuit are the main source of such $CaCO_3$ fouling. Based on solubility considerations, the deposition of calcium carbonate is most sensitive to pH. As the pH for the carbonate/bicarbonate transition is ca. 8,3, it is evident that the loading of calcium onto carbon becomes very pH dependent in the region pH 9-11 which is commonly found in gold plant circuits.

It is also well established[1,2] that gold adsorption onto carbon is strongly enhanced as the pH is reduced. As many investigations related to pH have been carried out in synthetic solutions it is perhaps not wise to assume a similar response when dealing with complex solutions or pulps. Both the presence of free cyanide and lime together with a heterogeneous mixture of metal cyanide complexes have a profound effect on gold adsorption. In the case of free cyanide, CN^- is protonated to the neutral HCN species at relatively high pH levels (pKa HCN = 9,5) thus lowering cyanide concentrations significantly when lowering pH below pH 10,3, while at the same time increasing the oxidation of free cyanide to $CO_3^=$ and NH_3. The lowering of pH below pH 10 will also have a significant effect on calcium carbonate fouling of the carbon.

As may be seen, the effect of free cyanide on gold adsorption is confounded to a large degree by other processing variables thus making any study on cyanide per se rather complex. While many earlier workers have demonstrated these effects using simple batch-type laboratory techniques, the effects of free cyanide concentrations on a counter current carbon-in-pulp (CIP) type operation treating typical Witwatersrand pulps have as yet received very little attention. The present investigation is a summarised account of such a study carried out at the Anglo American Research Laboratories during the period 1981-1983, in which the effects of free cyanide and pH on the efficiency of gold adsorption were studied. The study involved the operation and monitoring of two small CIP pilot plants under rather idealized conditions where cyanide concentration, pH levels and temperature were strictly controlled.

Experimental

The miniplants consisted of seven mechanically agitated contactors, each contactor having a working volume of 10 ℓ. Each contactor had an 850 µm screen (100 cm^2) at the outflow, the screens being cleaned intermittently using compressed air and water sprays on the discharge side of the screen. The impellers in each contactor were of the down draft type with blades inclined at 40° to the shaft axis so as to minimise short circuiting.

In order to simulate plant practice, a portion of the carbon was transferred daily, with the carbon and pulp from each stage being drained through a 700 µm screen. During carbon transfer feed to each miniplant was stopped, the carbon was washed and the required volume (a bulk density of 0,5 kg/ℓ was assumed) returned to the adjacent upstream contactor. Virgin carbon was added to the last stage (stage 6) and loaded carbon was collected from stage 1. The first agitated vessel in each miniplant contained no carbon and was used when required to adjust pH by the addition of 20% sulphuric acid using a Radiometer titrimeter. 500 W heating elements were used to maintain pulp temperature at 25°C in each miniplant. Two temperature controllers operated three elements, the first unit controlling the first three stages and the second controller operating in the last three stages.

G210 As coconut shell activated carbon (Le Carbone (Pty) Ltd) was used (1,12 to 2,35 mm). This carbon was initially conditioned by mechanically agitating 5 kg batches in 30 ℓ of water for 25 h. It was then wet screened at 1,0 mm to remove fines and air dried. The dried carbon was then rescreened and the oversize used as plant feed.

Unless otherwise specified, cyanided feed pulp to the two miniplants contained ca 50% solids and was pumped via a 3 m^3 mechanically agitated storage tank by means of adjustable finger pumps. Leached pulp was collected from the Western Deep Levels gold plant twice a week. As the cyanide content varied from batch to batch of pulp, the free cyanide concentration was adjusted daily to a specified level by the addition of reagent grade NaCN. The alkalinity in the storage tank was also adjusted to pH 10,5 by the addition of slaked lime or sulphuric acid as required. A relatively low rate of transfer of carbon was used to emphasise the difference in the efficiency of the two plants. In practice a faster rate of transfer and hence a lower gold loading on the carbon would be used to ensure good recovery.

The present investigation was carried out in three distinct phases, all concerned directly or indirectly with the influence of free cyanide concentration on CIP performance. Each phase involved minor operational changes which will be detailed below together with relevant results.

Results

Test 1: The effect of pH at high cyanide concentrations

In this investigation the cyanide tenor in the leached pulp was maintained at 300 g/t NaCN in solution and fed to both miniplants at a flowrate of 10 ℓ/h. A carbon concentration of 7,5 g/ℓ pulp was maintained in each plant, the one plant being operated at pH 10,5 and the other at pH 8,0. Carbon was transferred interstage at a rate of 40 g/d. Both plants were run continuously for 20 d with a common feed in order to allow for a direct comparison of results. However, an additional variable was air sparging of the contactors in the plant run at pH 8,0 to enhance cyanide destruction. Each of the contactors was fitted with a sintered glass diffusion tube through which the air flowrate was controlled at 0,70 ℓ/min. The only variation common to both plants during the period under review was the pulp grade which varied from 3,8-4,2 g/t gold in solution.

Test 1: Results

Gold and free cyanide solution data from the two miniplants, averaged so as to eliminate daily operational variations are presented in Tables 1 and 2 while progressive gold solution recoveries and progressive cyanide destruction data are illustrated in Figs. 1 and 2. As may be seen, the pH 8,0 plant using sparging in each contactor indicates a significantly better performance than the pH 10,5 plant (Figure 1). This improvement in gold recovery appears to be related to the free cyanide concentration profile (Figure 2) and to the loading of copper on the carbon (see below). In the pH 8,0 plant there is a very strong relationship between free cyanide in solution and gold in solution, particularly for the last three stages. This is well illustrated in Figure 3. It may also be noted that there is a far greater and more rapid destruction of cyanide in the pH 8,0 plant (79%) than in the pH 10,5 plant (25%).

The composition of carbon samples taken after 20 days in the two plants is presented in Table 3. The much steeper gold loading profile achieved at pH 8,0 may be noted, while the difference in loaded carbon values in the first stage indicates that steady state operation had not been achieved (by

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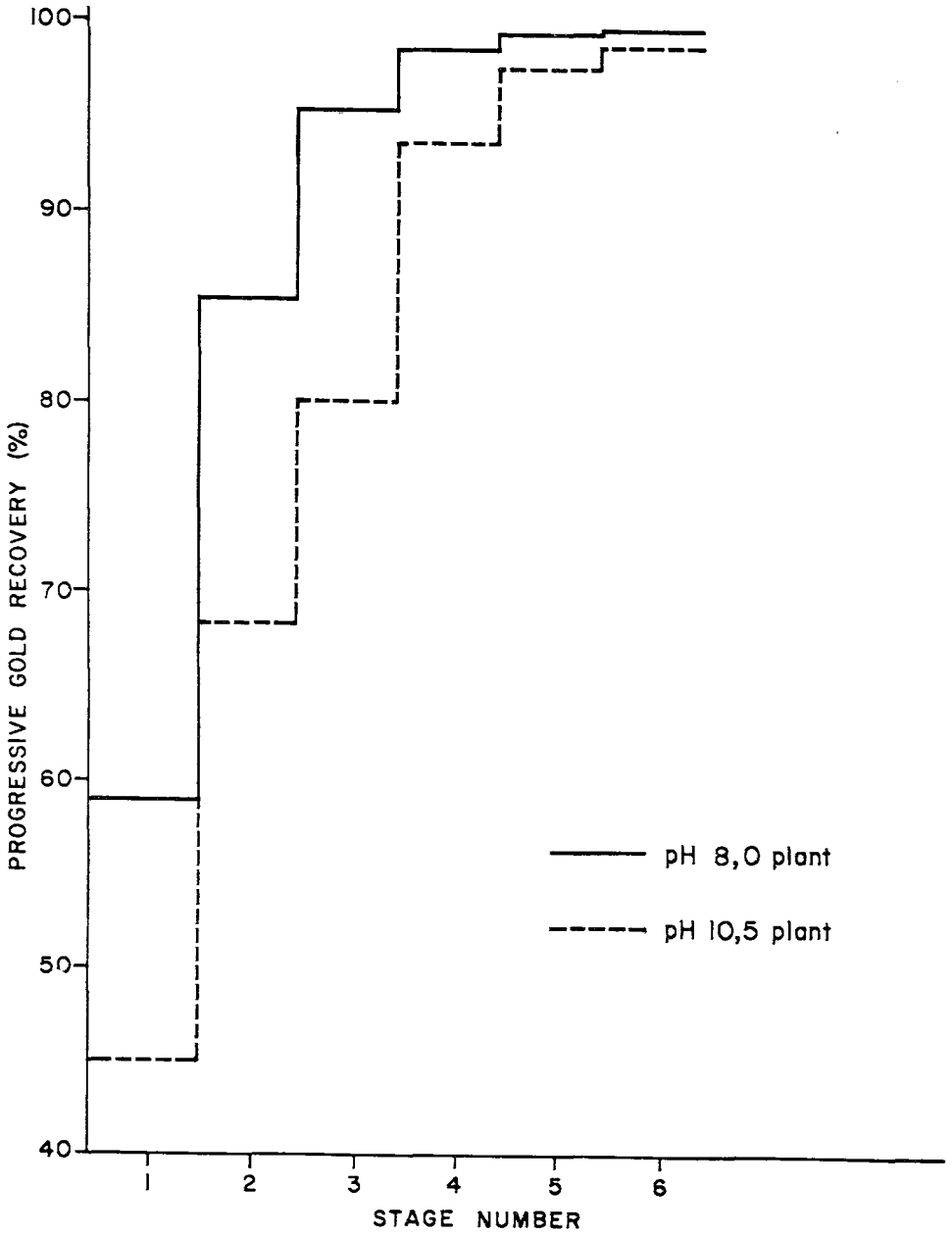

FIGURE 1. The effect of pH on the recovery of gold

TABLE 1. Gold solution data for test 1

Average gold concentrations in solution per stage (g/t)								
Time period (d)	Feed pH	Stage No.						
		Feed	1	2	3	4	5	6
6-10	8,0	3,796	1,530	0,640	1,198	0,064	0,030	0,014
	10,5	3,796	2,022	1,056	0,440	0,196	0,068	0,024
11-15	8,0	3,770	1,706	0,774	0,280	0,102	0,042	0,015
	10,5	3,770	2,528	1,786	0,948	0,434	0,166	0,082
16-20	8,0	4,158	1,700	0,600	0,196	0,058	0,023	0,012
	10,5	4,158	2,290	1,316	0,822	0,254	0,094	0,050

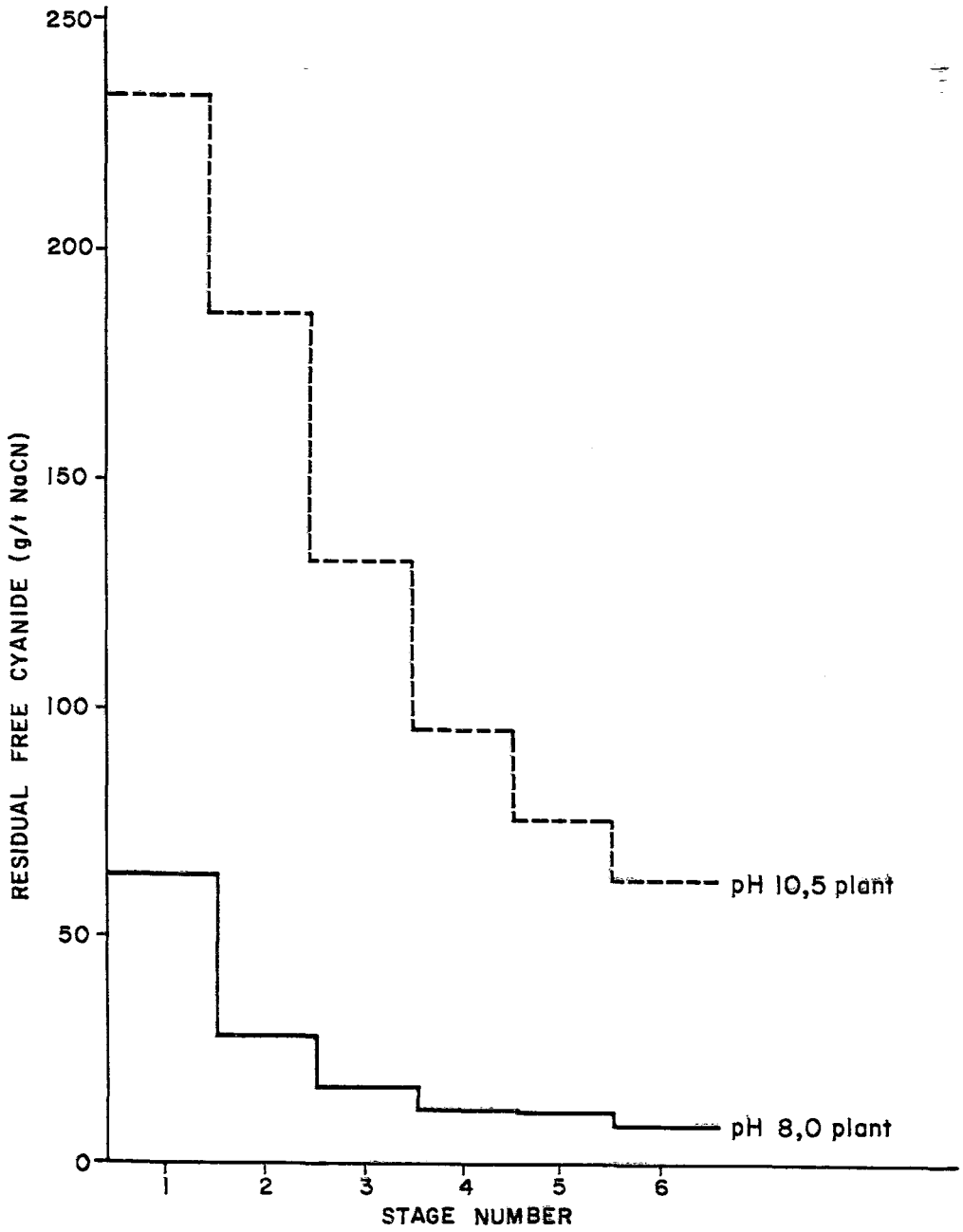

FIGURE 2. The effect of pH on cyanide destruction

TABLE 2. Cyanide solution data for test 1

Average free cyanide concentrations in solution (g/t NaCN)								
Time period (d)	Feed pH	Stage No.						
		Feed	1	2	3	4	5	6
6-10	8,0	291	66	33	26	18	15	13
	10,5		244	204	164	159	145	131
11-15	8,0	335	99	58	37	31	26	26
	10,5		290	264	246	223	198	190
16-20	8,0	312	64	28	17	12	12	8
	10,5		234	187	133	96	76	63

TABLE 3. Carbon loading profile data for test 1

Carbon composition	Feed pH	Stage No.					
		1	2	3	4	5	6
Au	8,0	20,2	8,73	4,41	1,93	0,73	0,23
(kg/t)	10,5	13,82	11,64	9,04	7,73	5,27	1,32
Ag	8,0	0,166	1,412	1,229	0,920	0,482	0,238
(kg/t)	10,5	1,033	0,834	0,977	1,107	1,264	0,763
Cu	8,0	1,62	1,95	3,50	6,01	9,46	10,37
(kg/t)	10,5	0,123	0,128	0,134	0,141	0,155	0,249
Ca	8,0	10,69	8,09	7,54	5,97	4,81	4,21
(kg/t)	10,5	34,7	32,8	29,8	18,14	9,33	3,26
SiO_2	8,0	5,7	6,2	5,6	5,0	6,1	4,9
(kg/t)	10,5	13,8	12,3	10,6	8,6	5,7	4,7

material balance the carbon gold loading for the pH 8,0 plant should be 19,1 kg/t after 6 stages and 18,8 kg/t for the pH 10,5 plant - the actual loadings of 20,2 kg/t and 13,8 kg/t respectively, indicate that the pH 10,5 operations has a much slower approach to steady state).

The greater destruction of cyanide at pH 8,0 may be explained in part at least by the increased loading of copper on the carbon (Table 3) as copper promotes the hydrolysis of cyanates to bicarbonate and ammonia.[5] With regard to silver adsorption, the rather flat profile indicated in the pH 10,5 operation may be compared with the steeper profile at pH 8,0, although silver displacement in stage 1 is evident due to the high gold loadings achieved. Lower loadings of both calcium and silica in the pH 8,0 operation are also evident.

Test 2: The effect of cyanide concentration at pH 10,5

This investigation followed Test 1. Leached pulp containing 300 g/t NaCN was fed to one miniplant at pH 10,5. Using similar operating conditions this operation was continued but for the cyanide concentration in the feed pulp which was progressively lowered from 300 to 145 to 93 to 50 and to 26 g/t NaCN during the course of the investigation. Each test at a controlled cyanidation concentration was run for at least 21 days in order to approach steady state. A minimum period of 7 days during which no sampling occurred, was allowed between each operational change to allow the plant to adjust to the change. Carbon profile samples were taken at the end of each such operational change.

Test 2: Results

From the averaged soluble gold concentration and gold recovery data presented in Tables 4 and 5, it may be seen that gold recovery over six adsorption stages was maximised using an intermediate cyanide concentration of ca 100-150 g/t NaCN. At both higher and lower cyanide concentration levels the efficiency was reduced.

Figure 4 presents data averaged over 5 days for the operations carried out at 300 and 145 g/t NaCN. From these data it is evident that low cyanide concentrations particularly

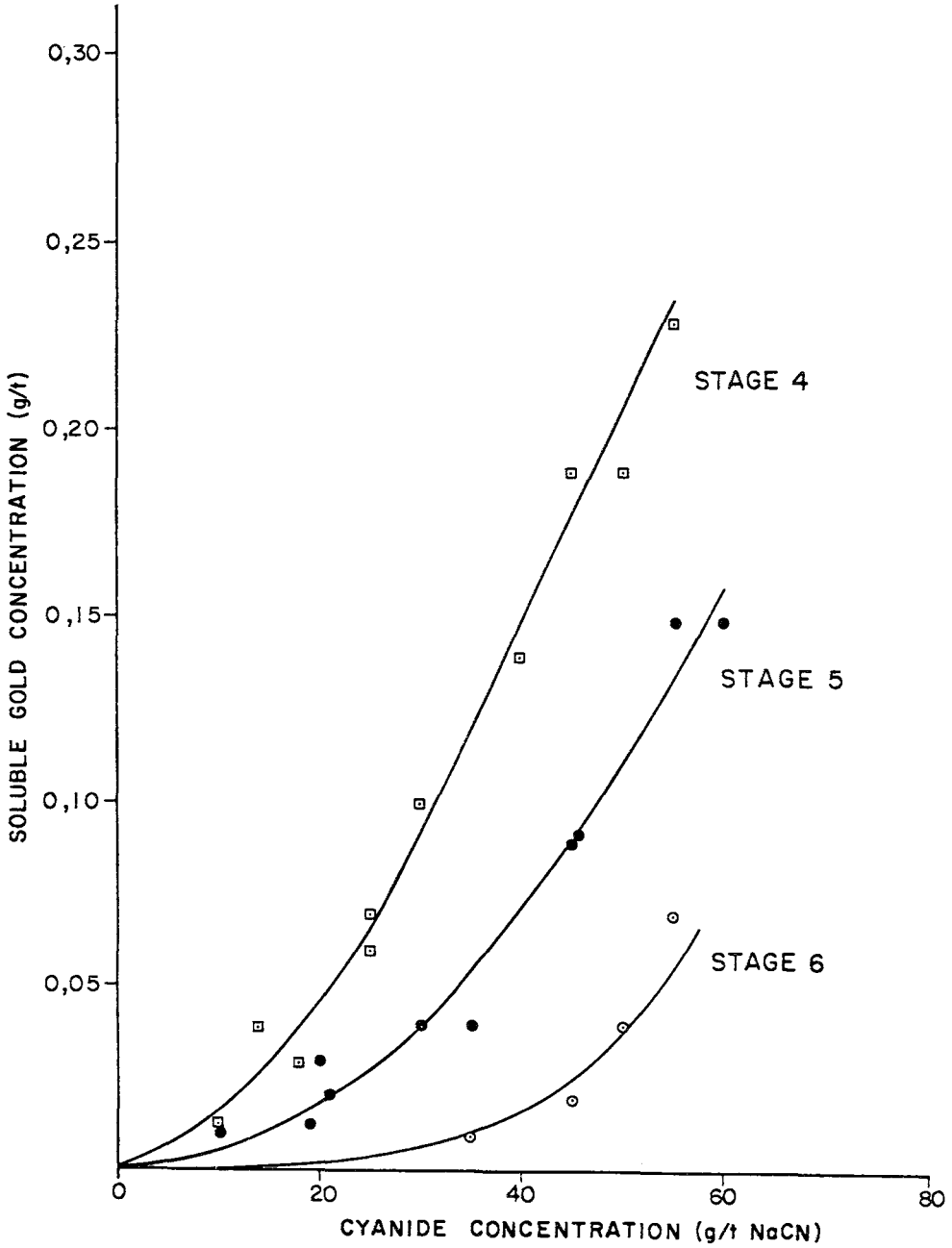

FIGURE 3. The correlation between residual gold concentration and residual cyanide concentration at pH 8,0. TEST 1

TABLE 4. Gold solution data for test 2

Free cyanide in feed pulp (g/t NaCN)	Average gold concentrations in solution (g/t)						
	Feed	Stage No.					
		1	2	3	4	5	6
300	3,99	2,27	1,42	0,76	0,31	,128	,064
145	5,87	2,54	1,37	0,59	0,24	,088	,035
93	3,50	1,70	1,07	0,50	0,21	,067	,017
50	5,42	3,46	2,36	1,54	0,68	,240	,058
26	3,93	2,16	1,35	0,82	0,39	,156	,037

TABLE 5. Average progressive percentage gold recovery data for test 2

Free cyanide in feed pulp (g/t NaCN)	Stage No.					
	1	2	3	4	5	6
300	43,0	64,5	80,9	92,1	96,8	98,4
145	56,8	76,6	89,9	95,9	98,5	99,4
93	51,4	69,5	85,7	94,1	98,1	99,5
50	36,6	56,8	72,0	87,5	95,6	98,9
26	45,0	65,6	79,2	90,1	96,1	99,1

TABLE 6. Carbon loading profile data for test 2

Carbon composition	Free cyanide in feed (g/t NaCN)	Stage No.					
		1	2	3	4	5	6
Au (kg/t)	300	13,8	11,64	9,04	7,73	5,27	1,32
	145	16,8	9,96	5,24	2,77	1,25	0,46
	93	18,9	14,00	10,63	7,52	3,70	1,16
	50	25,1	18,10	9,25	7,02	2,96	1,80
	26	15,3	9,01	5,28	3,12	1,46	0,19
Ag (kg/t)	300	1,03	0,83	0,98	1,11	1,26	0,76
	145	1,20	1,29	1,34	1,30	0,90	0,45
	93	0,95	0,75	1,04	1,57	1,50	1,02
	50	1,18	0,67	1,59	1,63	0,97	0,87
	26	0,28	0,16	0,16	0,14	0,14	0,06
Cu (kg/t)	300	0,12	0,13	0,13	0,14	0,16	0,25
	145	0,36	0,36	0,39	0,52	0,73	0,88
	93	0,54	0,70	0,79	1,86	3,33	4,36
	50	1,10	1,35	3,53	4,76	5,16	5,54
	26	11,40	13,01	15,89	18,19	15,70	7,79
Ca (kg/t)	300	34,7	32,8	29,8	18,14	9,33	3,26
	145	15,9	8,88	7,37	5,52	3,55	2,05
	93	19,8	17,2	17,1	16,0	9,0	6,9
	50	23,2	17,0	15,7	13,6	9,4	6,5
	26	16,3	12,7	14,7	14,3	8,1	7,8
SiO_2 (kg/t)	93	9,2	9,9	11,4	7,0	6,7	5,6
	50	10,8	8,9	8,0	7,3	6,7	6,2
	26	5,4	3,1	5,7	5,3	2,1	2,1

in the final stages of CIP are most beneficial to the lowering of the gold concentration in a particular stage. It is also noticeable that the progressive recoveries for cyanide concentrations of ca. 25-50 g/t NaCN are lower than the recoveries at ca 300 g/t NaCN up to stage 5 (Table 5). However, in stage 6 the situation is reversed. These results would seem to support the theory of low cyanide concentrations in the final stages being beneficial to gold recovery and high base metal and calcium loadings being detrimental (see below).

From the carbon loading profile data presented in Table 6 it is again evident that gold loadings in stage 1 are not consistent due to fluctuating feed values as well as the

long time periods required for the plant to reach steady state under the prescribed experimental conditions. Nonetheless, several macro-effects are evident which explain to some degree the poor gold adsorption observed at both high and low cyanide concentration feed levels. These effects may be discussed as follows:

a) The high calcium loadings resulting from the use of high cyanide concentrations (300 g/t NaCN) may explain in part at least the poor adsorption observed under such conditions.

b) The high copper loading resulting from the use of low cyanide concentrations is most noticeable. Copper and other base metals may compete with gold for adsorption sites on the carbon.

c) From the silver loading data presented in Table 6 it may be seen that silver loadings remain reasonably high and constant until low cyanide feed concentrations (ca 25 g/t NaCN) are reached. At such low cyanide concentrations significant silver losses may be noted. The precipitation of silver as either the chloride or the sulphate under such conditions may be speculated.

Test 3: The effect of cyanide concentration at pH 8,0

In this test leached pulp was fed at a flowrate of 20 ℓ/h to one miniplant operating at pH 8,0. The carbon concentration was maintained at 20 g/ℓ and transferred interstage at a rate of 75 g/d. This operation was continued but for the cyanide concentration in the feed which was progressively lowered from 100 to 50 g/t NaCN. Initially the plant was run for 12 days with no sampling so as to approach steady state. There followed a 10 day period of sampling for operation at a feed concentration of ca 100 g/t NaCN. The cyanide concentration in the feed pulp was then decreased to 50 g/t NaCN and a 7 day period of non sampling allowed to adjust to the change before sampling for a further 15 days. During the last 5 days of operation air sparging in each contactor using porous graphite blocks was introduced using an airflow of ca 0,3 ℓ/min per stage.

Test 3: Results

From the averaged gold solution profiles and gold recovery data presented in Tables 7-9 it is evident that there is a significant increase in gold recovery as the cyanide concentration in the feed pulp to the plant is decreased. Higher recoveries are achieved in each stage, thus minimising gold lock-up.

The averaged cyanide concentration profiles shown in Table 10 and the progressive cyanide destruction data presented in Table 11 show that at pH 8 the greater the destruction of free cyanide and the lower the actual concentration of cyanide, the better the resultant gold recovery. Figure 5 shows the clear relationship between gold concentration in solution and the free cyanide concentration for stages 4, 5 and 6, respectively. The data for this figure were obtained from results for the first three 5-day periods. The concave nature of these curves emphasises the strong effect of decreasing cyanide concentration (the accurate control of the cyanide concentration in the feed pulp is largely responsible for these distinct trends being apparent).

Both the progressive gold and cyanide destruction results for the fifth five-day period, during which air sparging was introduced, show optimum gold recoveries and the highest cyanide destruction values obtained (Tables 9 and 11). The aeration of the pulp and activated carbon ensures a plentiful supply of dissolved oxygen, necessary for the

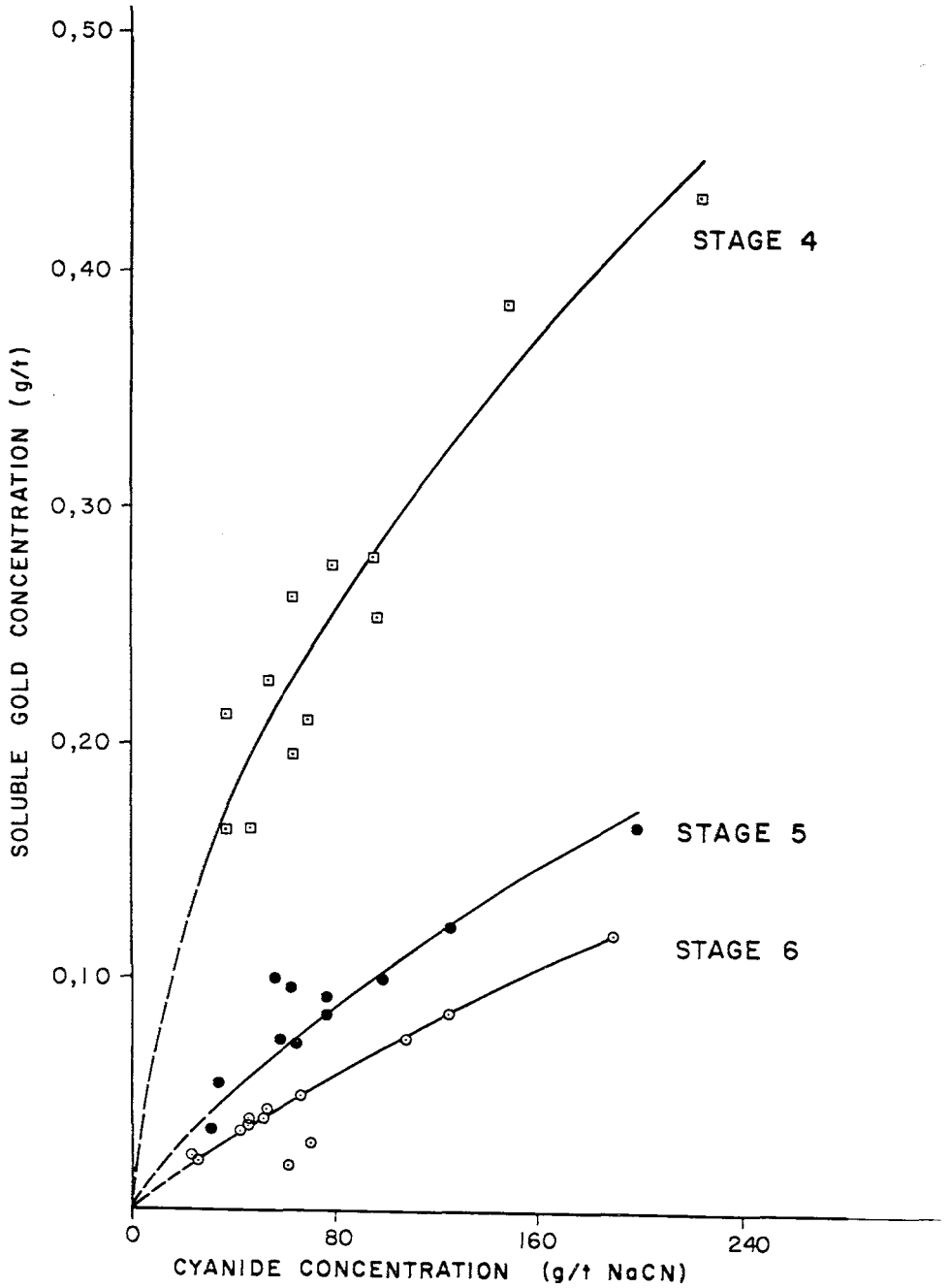

FIGURE 4. The correlation between residual gold concentration and residual cyanide concentration at pH 10,5. TEST 2

TABLE 7. Gold solution data for test 3

Average gold concentrations in solution (g/t)								
5 Day time periods	Free cyanide in feed (g/t NaCN)	Feed	Stage No.					
			1	2	3	4	5	6
1	102	3,79	1,79	0,74	0,40	0,28	0,132	0,041
2	80	3,42	1,92	0,68	0,37	0,176	0,055	0,018
3	56	2,86	1,01	0,38	0,154	0,072	0,020	0,010
4	48	2,86	0,91	0,36	0,178	0,079	0,014	0,005
5	53	3,55	0,77	0,26	0,103	0,052	0,007	0,003

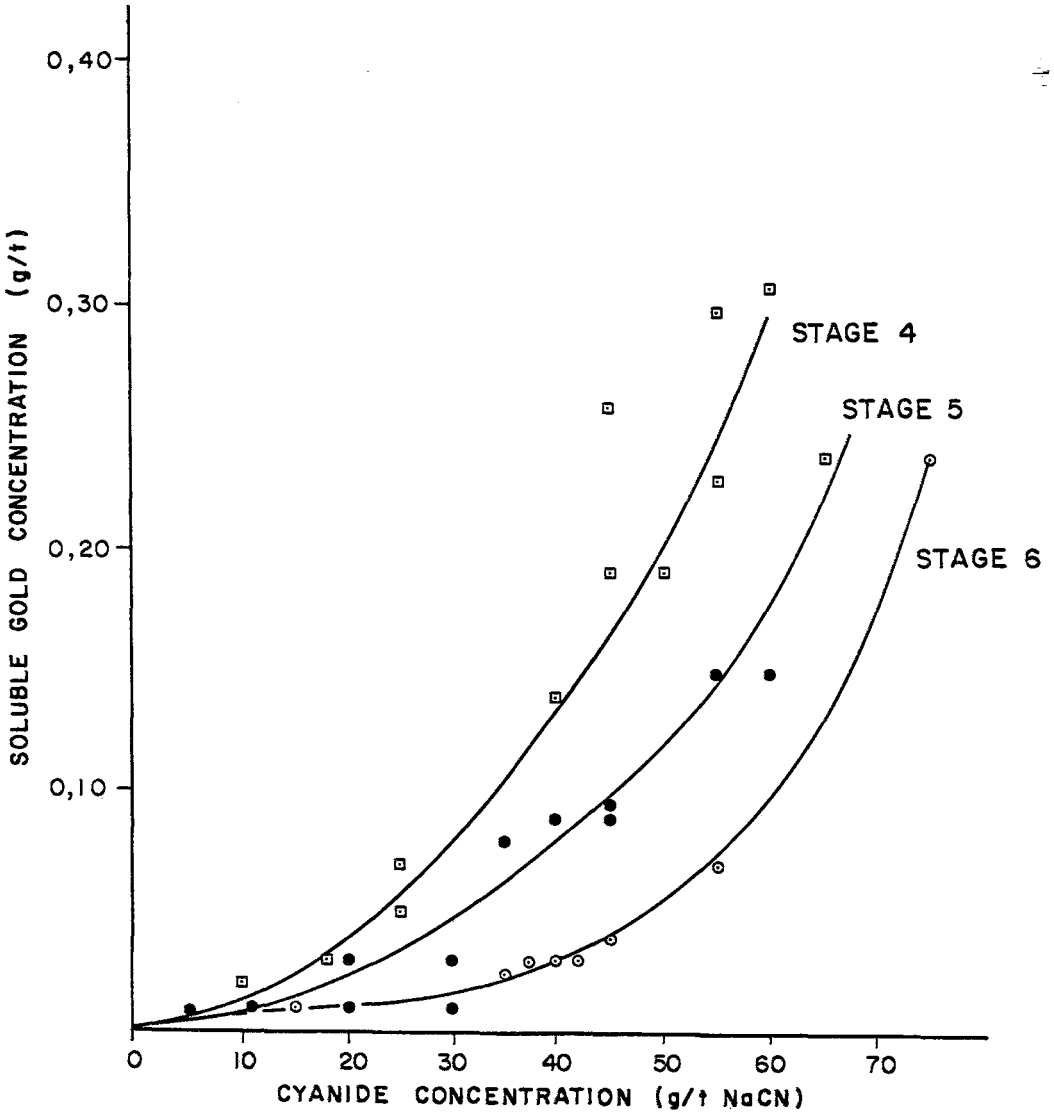

FIGURE 5. The correlation between residual gold concentration and residual cyanide concentration at pH 8,0. TEST 3

TABLE 8. Average percentage progressive gold recovery data for test 3

5 Day time period	Free cyanide in feed (g/t NaCN)	Stage No.					
		1	2	3	4	5	6
1	102	52,8	80,6	89,4	92,7	96,5	98,9
2	80	44,0	80,2	89,3	94,9	98,4	99,5
3	56	64,6	86,8	94,6	97,4	99,3	99,6
4	48	68,2	97,5	93,8	97,2	99,5	99,8
5	53	78,3	92,6	97,1	98,5	99,8	99,9

TABLE 9. Average percentage stage gold recovery data for test 3

5 Day time period	Free cyanide in feed (g/t NaCN)	Stage No.					
		1	2	3	4	5	6
1	102	52,8	58,7	45,9	30,0	52,9	69,0
2	80	44,0	64,6	45,6	52,4	68,8	67,3
3	56	64,6	62,4	59,5	53,2	72,2	50,0
4	48	68,2	60,4	50,6	55,6	82,3	64,3
5	53	78,3	66,2	60,4	49,6	86,5	57,1

TABLE 10. Cyanide solution data for test 3

Average free cyanide concentrations in solution (g/t NaCN)								
5 Day time periods	Feed	pH Conditioner	Stage No.					
			1	2	3	4	5	6
1	102	84	73	63	58	56	53	49
2	80	70	61	56	51	46	37	34
3	56	40	32	27	25	20	18	15
4	48	42	31	27	25	22	18	15
5	53	40	30	20	15	12	6	4

oxidation of cyanide.

Due to the improved gold recovery recorded at the ca 50 g/t NaCN concentration, compared with the recovery obtained at ca 100 g NaCN (Table 8), a steeper gold loading profile is expected and hence gold lock-up is reduced at the lower cyanide concentration.

Since only one gold loading profile for a cyanide concentration of 56 g/t was assayed (Table 12), the gold loading profiles for operation at 80 g/t and 48 g/t NaCN in the feed pulp were calculated by material balance. The mean gold loading for six stages of carbon for the 80 g/t NaCN operation was 4,97 kg/t and the mean loading over 6 stages for the 48 g/t NaCN operation was 3,89 kg/t. This is a significant decrease in gold lock-up and represents a large cost saving for a full-scale plant.

The gold, silver, copper, calcium and silica loading profiles for plant operation at 56 g/t NaCN in the feed, are listed in Table 12. The gold loading profile may be seen to be very steep, as predicted from a study of the solution gold profiles and gold recovery profiles. Furthermore, the high copper loadings may be noted. These high copper loadings are due to the lower cyanide concentration which reduces the solubility of the copper cyanide complexes. The steep gold profile and the satisfactory gold recoveries at pH 8 suggest that the loading of copper is not competing with gold loading. It is also possible that a more insoluble gold/copper cyanide complex may be precipitated under these conditions, thus actually aiding gold adsorption.[6]

A further beneficial effect of the high copper loading is the increased catalytic

TABLE 11. Cyanide destruction data for test 3

Average percentage progressive cyanide destruction							
5 Day time periods	Cyanide in feed (g/t NaCN)	Stage No.					
		1	2	3	4	5	6
1	102	28	38	43	45	48	52
2	80	24	30	36	42	54	58
3	56	43	52	55	64	68	73
4	48	35	44	48	54	62	69
5	53	43	62	72	77	89	92

TABLE 12. Carbon loading profile data for test 3 (56 g/t NaCN feed concentration

Carbon composition	Stage No.					
	1	2	3	4	5	6
Au (kg/t)	11,9	6,1	3,1	1,44	0,60	0,20
Ag (kg/t)	0,68	0,50	0,36	0,26	0,18	0,09
Cu (kg/t)	18,7	24,4	26,4	28,1	23,0	13,1
Ca (kg/t)	6,5	5,6	4,7	5,0	4,6	3,8
SiO_2 (kg/t)	4,5	4,4	3,9	3,5	3,4	3,5

oxidation of free cyanide. The destruction of cyanide at pH 8 is particularly beneficial since there is minimal precipitation of calcium carbonate. The calcium loadings (Table 12) are the lowest values obtained in CIP operation at AARL to date. Silica, which may also physically passivate carbon, was also loaded to a minimal extent at the low pH and cyanide concentrations used in the present investigation (Tables 3, 6 and 12).

The only detrimental effect of operation at pH 8 and low cyanide concentrations may be a resultant loss of silver. This loss may be caused by precipitation of the silver as a chloride or sulphate.

Discussion

From the results presented in this investigation, it is most evident that the cyanide concentration, the pulp alkalinity and the catalytic oxidation of free cyanide all play a vital role in the metallurgical performance of a CIP type operation. The careful control and balancing of these parameters would appear to be necessary in effecting optimum metallurgical efficiency. Poor metallurgical performance is indicated at high cyanide levels regardless of pulp alkalinity, while air sparging at conventional pH levels under such conditions would suggest severe passivation of carbon with calcium carbonate (earlier air agitated CIP circuits encountered this problem). On the other hand, poor performance is also encountered at low cyanide levels when operating at conventional pH levels due both to the loss of carbon selectivity for gold adsorption and to the relatively low capacity of the carbon under such conditions. This situation may change

dramatically if pH levels are reduced. While excellent efficiency was obtained, the effective use of air sparging indicated the important role played by copper in effectively destroying cyanide, with no passivation of the carbon with calcium carbonate being observed.

When the present findings are related to a large scale operation, several practical implications must be considered. In the case of the present study, very efficient mixing in the small pilot contactors limited to a large degree the diffusion control which may pertain to a large plant.[7] This may well decrease the magnitude of the effects illustrated in the present study considerably. Furthermore, while acidification with sulphuric acid was shown to have very beneficial effects, further investigation into the possible poisoning of carbon with calcium sulphate under simulated plant conditions would be required if such neutralisation were used. Also, the cost of the acid and its effect on the water quality in a particular circuit must be carefully considered. Losses of silver may also be encountered (particularly at low cyanide concentrations - compare Tests 1 and 3) if neutralisation is carried too far. Nonetheless, with the recent development of automated cyanide monitoring and control equipment, the effective control of cyanide (and lime) additions are now possible. Besides effecting significant savings in both reagent cyanide and lime, the control of free cyanide at say 100-150 g/t NaCN and the alkalinity in the region pH 10,3 would in themselves improve metallurgical efficiency considerably. Only when effective control has been established can a decrease in either cyanide concentration or pH be considered.

Acknowledgements

This paper is published with permission of the management of the Anglo American Corporation. The contribution of A.W.G. Reeve and A. Killat in operating the pilot plants is gratefully acknowledged, as is the funding of the project by the Gold Division of the Anglo American Corporation.

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2.2 AAC - DELKOR LINEAR SCREEN

Since the introduction of CIP units into run-of-mine circuits in South Africa, the presence of wood fibre in leached pulps has presented major operating difficulties. Besides the fouling of interstage screens, this material causes problems in the hydraulic transport of carbon, the blinding of elution column collectors and the liberation of fine carbon (active) when burning in the regeneration kilns.

Developed as a joint project with Delkor Technik with Anglo American retaining the patent rights, the AAC-Delkor linear screen has virtually removed the earlier problems associated with the presence of wood fibre in CIP circuits and is now in common use (more than 300 units have been marketed). The screen was first successfully demonstrated at Afrikander Leases as a means of removing wood fibre in the Pinned Bed prototype plant (2.4).

A. & A. Ref. No.

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FORM P7

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PATENT ATTORNEYS
MASADA BUILDING
PRETORIA

109709 AL

AS FILED 24 JUN 1985

REPUBLIC OF SOUTH AFRICA
Patents Act, 1978

COMPLETE SPECIFICATION

(Section 30 (1) — Regulation 28)

OFFICIAL APPLICATION NO.		LODGING DATE	
21	01 85/4746	22	24 JUNE 1985

INTERNATIONAL CLASSIFICATION	
51	B03B: B01D

FULL NAME(S) OF APPLICANT(S)	
71	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

FULL NAME(S) OF INVENTOR(S)	
72	1. IAN MACDONALD ARBUTHNOT 2. ERROL THOMAS AMM 3. RAYMOND JOHN DAVIDSON

TITLE OF INVENTION	
54	"AN APPARATUS AND A METHOD FOR SEPARATING ONE SOLID COMPONENT FROM ANOTHER SOLID COMPONENT IN SUSPENSION IN A LIQUID"

THIS INVENTION relates to an apparatus for and to a method of separating a first solid component from a mixture thereof with a liquid and a second solid component, the first and second solid components being in suspension in the liquid.

The Applicant is aware of a large number of apparatus and methods for separating one solid component from another. Screens and cyclones are examples of such apparatus. It is proposed in this specification that an endless travelling cloth be used as the separating medium, and the Applicant is not aware of an endless cloth ever having been used for such a purpose.

The Applicant is nevertheless aware of a large number of filters that utilise an endless travelling cloth. These cloths are mostly supported by belts. However, the philosophy of a filter is totally different

from that of the present invention. Thus, filters are intended to filter out solid particles that are in suspension in a liquid. With these prior filters the intention is to remove all the solid particles. With the present invention, the object is to separate the first solid component from the second solid component, with there being substantially more of the second solid component than there is of the first solid component, so that much more solid material passes through the apparatus than is removed by it.

According to the invention there is provided a separator for separating out a first solid component from a mixture thereof with a liquid and a second solid component which is of smaller average particle size than the first solid component, the first and second solid components being in suspension in the liquid and with the percentage by mass of the second solid component being substantially greater than that of the first solid component, which includes

an endless cloth which is foraminous and has foramen which are of a suitable size to permit the second solid component to pass therethrough and to impede the passage therethrough of the first solid component;

a feed means for feeding the liquid and the solid

components onto the cloth;

a liquid collecting means for collecting liquid and material that has passed through the cloth;

a first component removal and collecting means for removing material from the cloth and collecting it; and

a drive means for displacing the cloth past the feed means, the liquid collecting means and the first component removal and collecting means.

Further according to the invention there is provided a method of separating a first solid component from a mixture thereof with a liquid and a second solid component which is of smaller average particle size than the first solid component, the first and second solid components being in suspension in the liquid and with the percentage by mass of the second component being substantially greater than that of the first component, which includes

feeding the liquid and the solid components onto an endless cloth which has foramen of a suitable size to permit the passage therethrough of the second solid component and to impede the passage therethrough of the first solid component;

displacing the cloth;

collecting liquid and other material that has passed through the cloth; and

removing from the cloth first component particles and collecting them.

The material may pass through the cloth under the action only of gravity without utilising an artificially created vacuum.

It will be appreciated that the first solid component and the second solid component may be particles of different size, shape or configuration although they are of the same material. Thus, coarse particles may be removed from a slurry after a milling operation. Instead, they may be different materials. In particular, the first solid component may be wood chips and the second solid component may be a gold-bearing chemical composition. Thus, the cloth may be such that particles having a size of less than 200 micron may pass through and particles of a greater size are retained thereon. Thus, for this application, the cloth may have foramen of about 250 microns. It will also be appreciated that the first solid component that is retained on the cloth is subsequently removed by any suitable process which may be a mechanical process.

Referring to the cloth, it may be of a synthetic monofilament fabric. It may also be woven and may be of polypropylene, polyester or a polyamide material. Further, it may be of a single, double or triple layer monofilament fabric.

The cloth preferably extends over a region in a substantially horizontal manner, this being the operative separating region of the cloth. The liquid collecting means, in the form of a trough, is then located below the cloth in this region. The mixture may then be fed onto the operative separating region by means of a feed box located above the said region. The feed box may have a floor that extends over substantially the entire separating region and is spaced from the cloth. The floor then has feed apertures which are distributed over substantially its entire area such that the liquid with the components in suspension therein is fed onto substantially the entire operative separating region.

Jets may be provided, below the cloth in the operative separating region, in order to spray water onto and through the cloth in an attempt to repulp the solids and prevent the formation of a layer of solid material on the cloth, which would stop the second solid

component passing through the cloth, which is the stated objective.

The invention is now described, by way of an example, with reference to the accompanying drawings, in which:-

Figure 1 shows a side view of a separator in accordance with the invention;

Figure 2 shows a plan view of the separator;

Figure 3 shows an end view of the separator: and

Figure 4 shows a sectioned view of the separator along line IV-IV in Figure 1.

Referring to the drawings, a separator is shown therein, designated generally by reference numeral 10. The separator 10 has an endless cloth 12 which is supported on a number of support rollers 14 to define a horizontal separating region and also passes over a number of other rollers, including a drive roller 16. Beneath the support rollers 14 there is a collecting trough 18 which has an outlet 20. Material that is to be separated is fed onto the cloth 12 by means of a feed box 22. The feed box 22 has side walls 24, end walls 26 and a floor 28. An inlet 32 is provided in the rear end wall 26. The floor 28 has a number of circular apertures 34 that have a diameter of 1 to 2 cms and which are

distributed over substantially the entire floor 28. The floor 28 is spaced a few centimetres from the cloth 12, and extends over a substantial part of the operative separating region. Thus, the feed box 22 feeds material to be separated, in a distributed manner, onto substantially all of the operative separating region.

The cloth 12 is a fabric that is woven from a suitable synthetic mono-filament material. The material may be polypropylene, polyester or polyamide. The cloth is woven to have apertures of a suitable size depending on the application for which it is intended. Thus, in one application where it is desired to separate wood particles which are carried in a slurry of gold-bearing particles in suspension in water, in which the gold-bearing particles have a maximum particle size of about 200 microns, the cloth is provided with apertures of about 250 microns in size. Thus, when the material to be separated is fed onto the cloth 12 by means of the feed box 22 the slurry passes through the cloth 12 under the action of gravity to be collected in the trough 18 whereas the wood particles remain behind on the cloth 12. These are then washed off the cloth 12 to be collected in a trough 30.

Sprays 36 are located below the cloth 12 in the operative region, ie between support rollers 14, to spray water onto and through the cloth 12, to repulp any solid material tending to form a layer on the cloth 12.

The cloth 12 is driven at a speed of between 2 and 9 metres/minute if wood particles are to be removed from gold bearing slurry and between 5 and 30 metres/minute when particles that are to large are to be removed from a slurry after a milling operation.

By means of the invention, an apparatus and method are provided whereby, in general, larger particles that form a small part of the solid material in suspension in a liquid may be removed therefrom in a continuous manner, and in particular, wood particles may be easily removed from a gold bearing slurry.

CLAIMS:

1. A separator for separating out a first solid component from a mixture thereof with a liquid and a second solid component which is of smaller average particle size than the first solid component, the first and second solid components being in suspension in the liquid and with the percentage by mass of the second solid component being substantially greater than that of the first solid component, which includes

an endless cloth which is foraminous and has foramen which are of a suitable size to permit the second solid component to pass therethrough and to impede the passage therethrough of the first solid component;

a feed means for feeding the liquid and the solid components onto the cloth;

a liquid collecting means for collecting liquid and material that has passed through the cloth;

a first component removal and collecting means for removing material from the cloth and collecting it; and

a drive means for displacing the cloth past the feed means, the liquid collecting means and the first component removal and collecting means.

2. The separator claimed in Claim 1, in which the liquid collecting means is located beneath a portion of the cloth and the separation is effected by means of gravity only.

3. The separator as claimed in Claim 1 or 2, in which the cloth is of a synthetic monofilament fabric.

4. The separator as claimed in Claim 3, in which the cloth is of a woven fabric.

5. The separator as claimed in Claim 3, in which the cloth is of a single, double or triple layer monofilament fabric.

6. The separator as claimed in Claim 1, in which the cloth extends over a region in a substantially horizontal manner, this being the operative separating region of the cloth and with the liquid collecting means being located below the cloth that occupies this region, and with the feed means including a feed box that is located above this region and has a floor that extends over substantially the entire region and is spaced from the cloth; the floor having apertures distributed over substantially its entire area such that the liquid with the components in suspension therein is fed onto substan-

tially the entire operative separating region.

7. The separator as claimed in Claim 1, which includes a spray means for spraying liquid onto the underneath surface of the cloth that is located immediately beneath the feed means.

8. A separator as claimed in any one of the preceding claims, for removing wood chips from a mixture thereof with gold bearing particles, in which the cloth has foramen of about 250 microns.

9. A separator as claimed in any one of the preceding claims, for removing coarse particles from a slurry after a milling operation.

10. A method of separating out a first solid component from a mixture thereof with a liquid and a second solid component which is of smaller average particle size than the first solid component, the first and second solid components being in suspension in the liquid and with the percentage by mass of the second component being substantially greater than that of the first component, which includes

feeding the liquid and the solid components onto an endless cloth which has foramen of a suitable size to

permit the passage therethrough of the second solid component and to impede the passage therethrough of the first solid component;

displacing the cloth;

collecting liquid and other material that has passed through the cloth; and

removing from the cloth first component particles and collecting them.

11. The method claimed in Claim 10, in which the liquid and solid components are fed onto a portion of the cloth that is substantially horizontal.

12. The method claimed in Claim 11, in which the liquid and material that passes through the cloth are caused to do so under the action of gravity only.

13. The method claimed in any one of Claims 10 to 12, in which the cloth is displaced at a speed of between 2 and 30 metres/minute.

14. The method claimed in any one of Claims 10 to 12 which includes spraying liquid onto the underneath surface of that part of the cloth onto which material is being fed.

15. The method claimed in any one of Claims 10 to 14, in which wood chips are removed from a gold bearing slurry.

16. The method claimed in any one of Claims 10 to 14, in which fine carbon particles are removed from a slurry.

17. The method claimed in any one of Claims 10 to 14, in which coarse particles are removed from a slurry after a milling operation.

18. The method claimed in Claim 11, in which the liquid and solid components are fed onto substantially the entire operative separating region of the filter cloth.

19. A method of separating out a first solid component from a mixture thereof with a liquid and a second solid component which is of smaller average particle size than the first solid component, the first and second solid components being in suspension in the liquid, and with the percentage by mass of the first component being substantially greater than that of the second component substantially as described in the specification, with reference to the accompanying drawings.

20. A separator substantially as described in the specification with reference to the accompanying drawings.

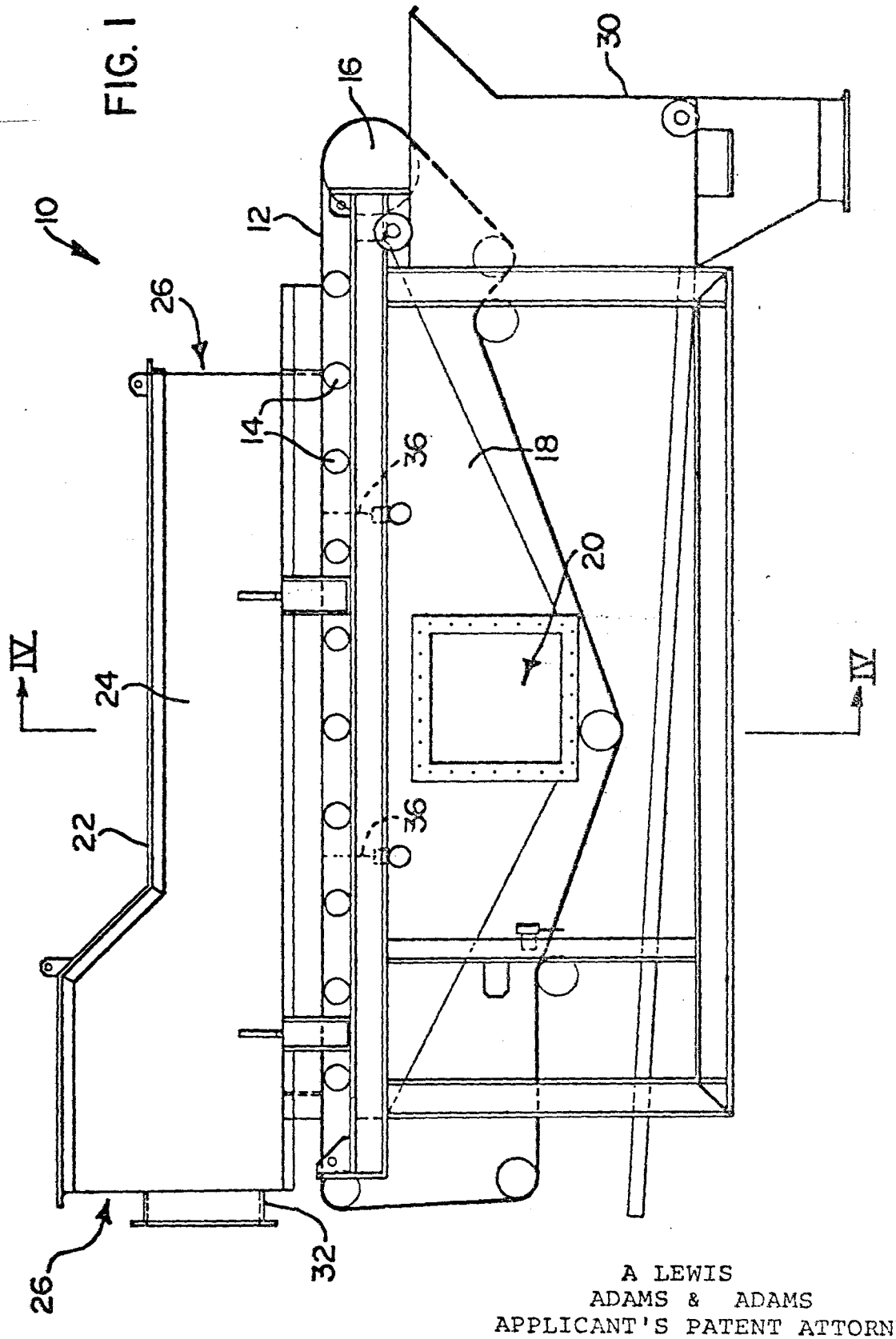
DATED THIS 24TH DAY OF JUNE 1985

A LEWIS
ADAMS & ADAMS
APPLICANT'S PATENT ATTORNEYS

ANGLO AMERICAN CORPORATION OF SA LTD

4 SHEETS
SHEET 1

FIG. 1



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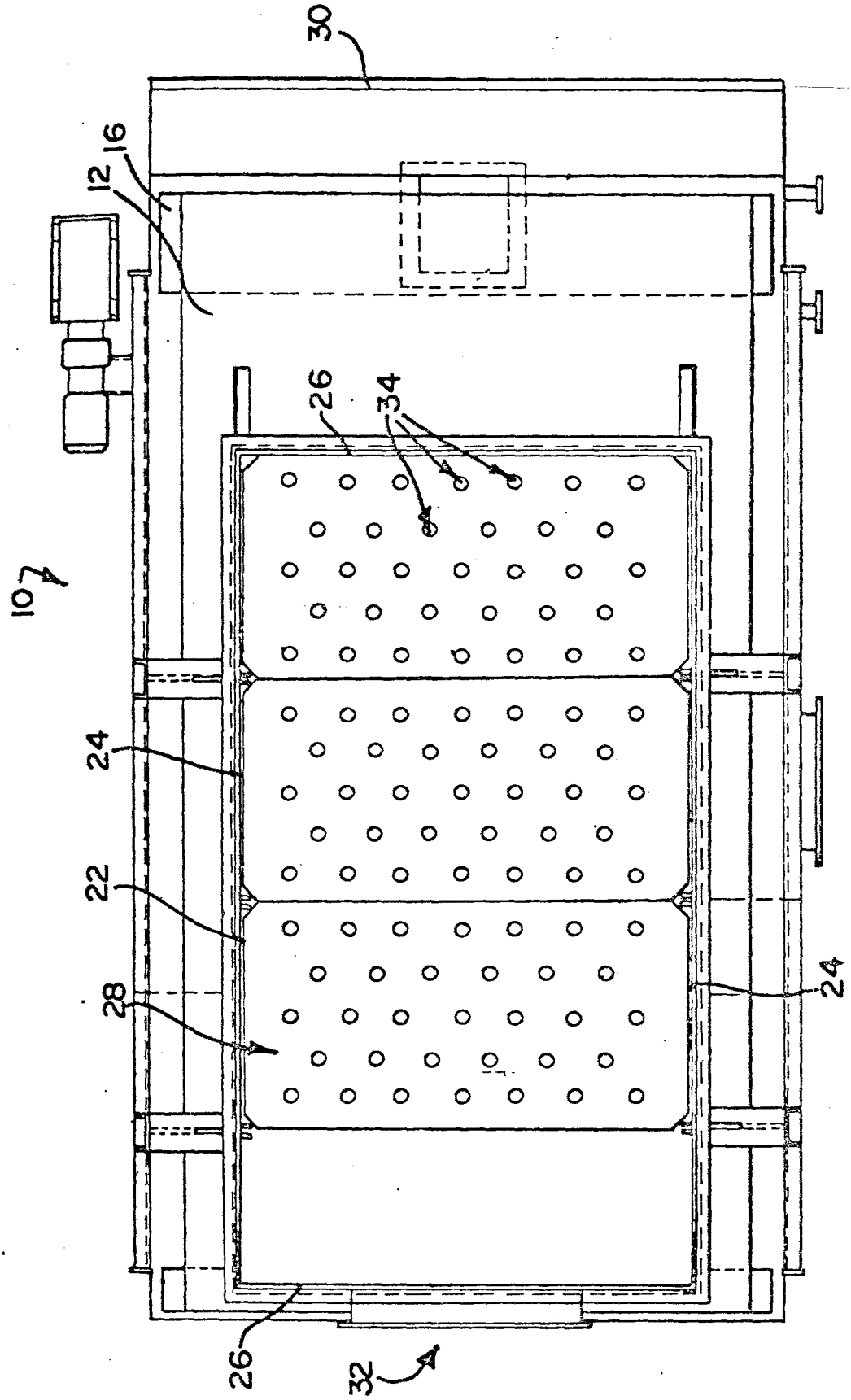

FIG. 2

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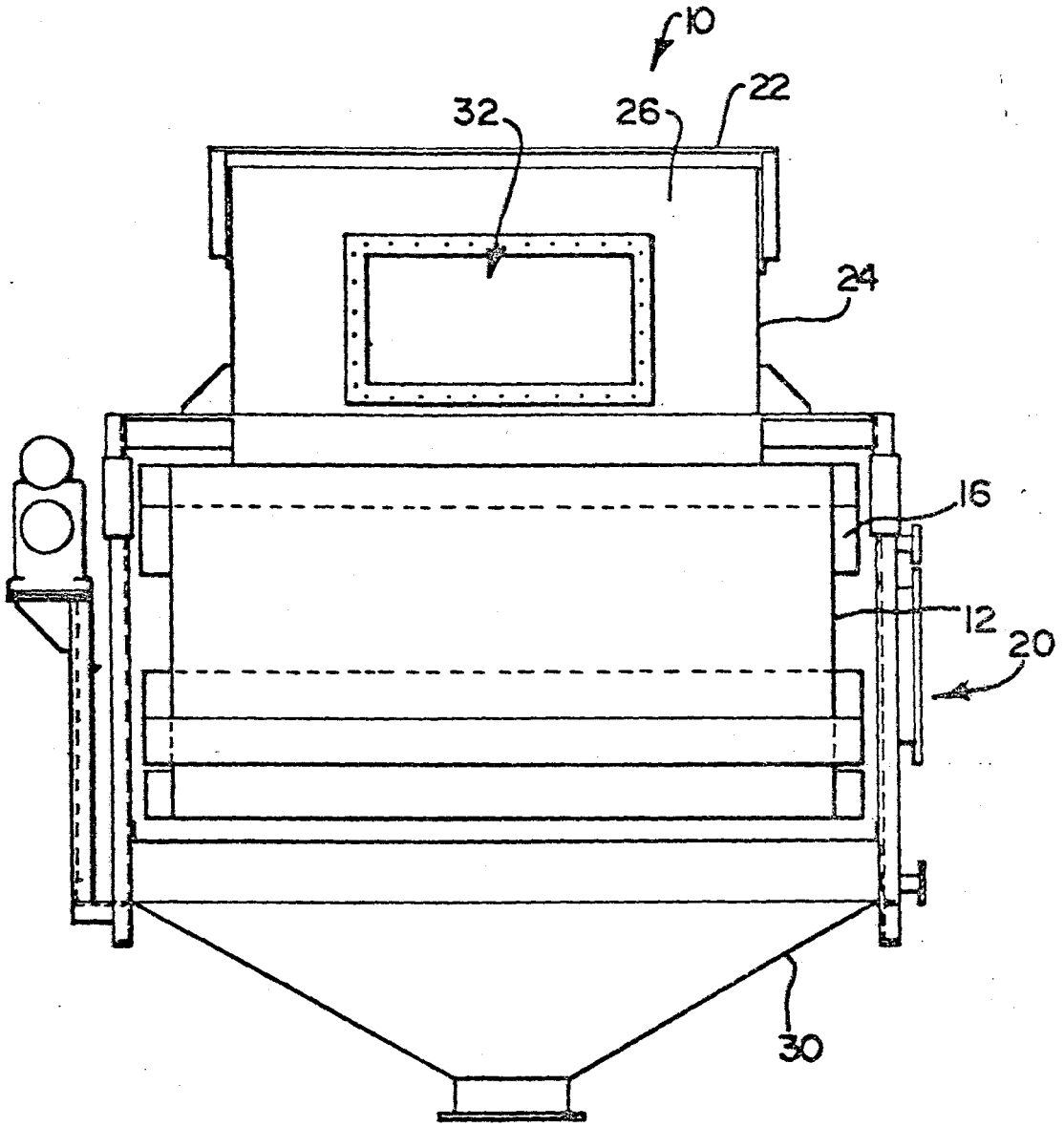

FIG. 3

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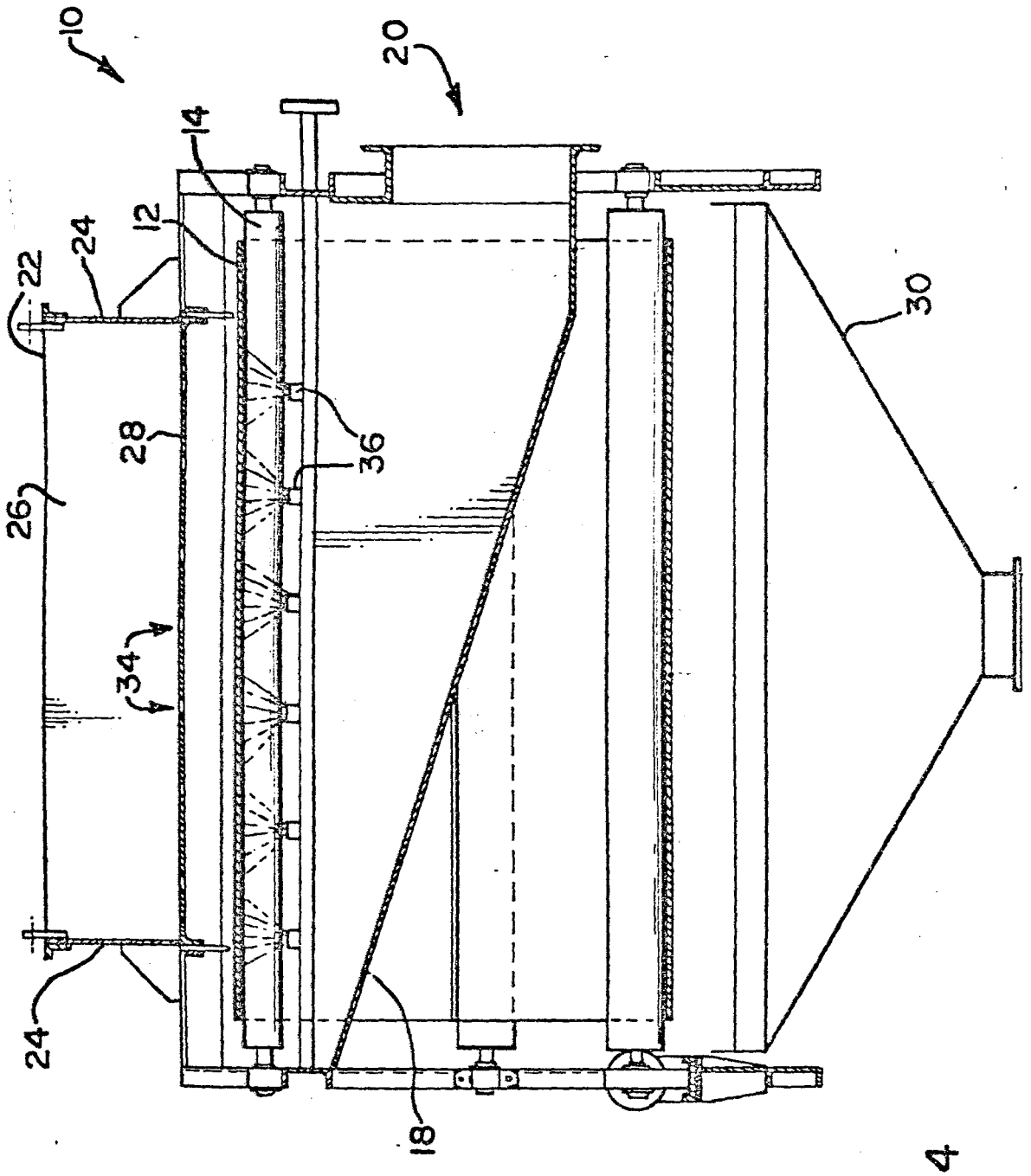

FIG 4

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The story of another South African achievement in the mining field which claims to solve an age-old problem in gold recovery.

Worldwide interest in South African linear wood fibre screens

By HENDRIK VORSTER

A new screening unit used in metallurgical processes has, over only one year, proved itself so successful that the company marketing and installing the equipment, Delkor Technik, has received orders for some 30 units to the value of about R2,5-million.

The Delkor linear screen (DLS) is a joint development between Anglo American and Delkor Technik.

With the use of the carbon in pulp system in gold recovery, wood chips and fibre derived from the timber supports used in mining proved to be difficult to remove efficiently using normal vibratory screens. Wood chips/fibre accumulate in the adsorption tanks, reducing the throughput rate.

Dr Ray Davidson at the Anglo American Reserch Laboratories, sought to solve the problem and at the end of 1984, in conjuction with Delkor, pioneered the use of a linear cloth screen to remove the wood fibre.

A prototype linear screen was installed at the Afrikander Lease Division of Vaal Reefs Gold Mine at Klerksdorp to remove wood fibre and other oversize material from the feed to the carbon adsorption plant.

"The screen solved a major problem. It was found to effectively remove more than 98 percent of the wood fibre without any sign of blinding in the fine apertures of the screen itself," Mr Phil Mostert, consulting metallurgist for the Anglo American Corporation, said.

"Without this screen we were experiencing problems with accumulation of wood chips/fibre in the absorption section of CIP plants."

Mr Mostert said that a Delkor linear screen unit would be installed on the new No 1 Shaft complex gold plant at the Western Deep Levels Gold Mine, Carletonville.

At Sallies, where a linear screen had been installed, it was found remarkably effective for removing the majority of the ash from the sand from which gold is being reclaimed.

DR RAY DAVIDSON

Ash has the disadvantage of absorping gold and reducing gold recovery.

Mr Mostert said the screen could also possibly improve the recovery of, say, tin in tin recovery plants.

It represented a breakthrough and there was no reason why other countries of the world should not also benefit by using such screens.

The Delkor Linear Screen comprises a coarse monofilament screen cloth supported on cloth rollers and driven by a head pulley coupled to a variable speed drive unit.

Slurry is fed through a distributor on to the moving monofilament cloth. The undersize material drains through the cloth by gravity and is collected in the underspan.

The wood fibre and oversize material retained on the screen is discharged at the drive pulley, using a belt wash spray.

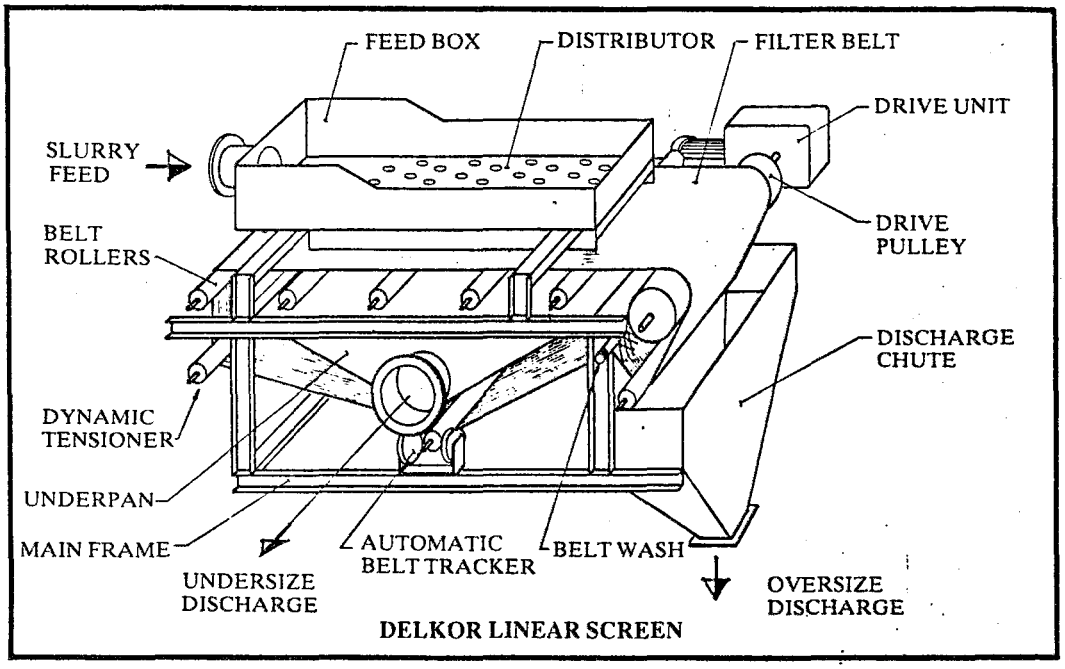
The belt wash spray ensures a continuous, self-cleaning screen, and by adjusting the weave configuration of the screen cloth, the cut point for oversize removal can be easily varied.

The prototype unit at Afrikander Lease Division of Vaal Reefs Gold Mine — a 6 sq m unit — has been operating practically trouble-free since the first week of January, 1985. It replaced a conventional vibrating screen having slotted apertures of 0,5mm.

Over a period of two weeks the concentration of wood fibre measured after the thickening section reduced from ,075 gm/l to ,002 gm/l wood fibre in the pulp.

Apart from achieving its objective as an efficient machine for the removal of woodchips /fibre, the DLS proved to have numerous further advantages:

- It operates without vibra-



DELKOR LINEAR SCREEN

tion, eliminating the need for expensive civil and structural construction common in conventional mill buildings;

- It eliminates common stress cracking problems encountered with vibratory screens;
- The noise level is markedly reduced due to the absence of vibration;
- Power consumption of the unit is a small fraction of that required for vibrating screens;
- Availabilities in excess of 97 percent have been recorded on most installations:
- High throughput rates per unit screen area are possible; and
- Simplicity of operation and open design have shown benefits in very low maintenance costs on the DLS
- The cloth is very simply and readily replaced.

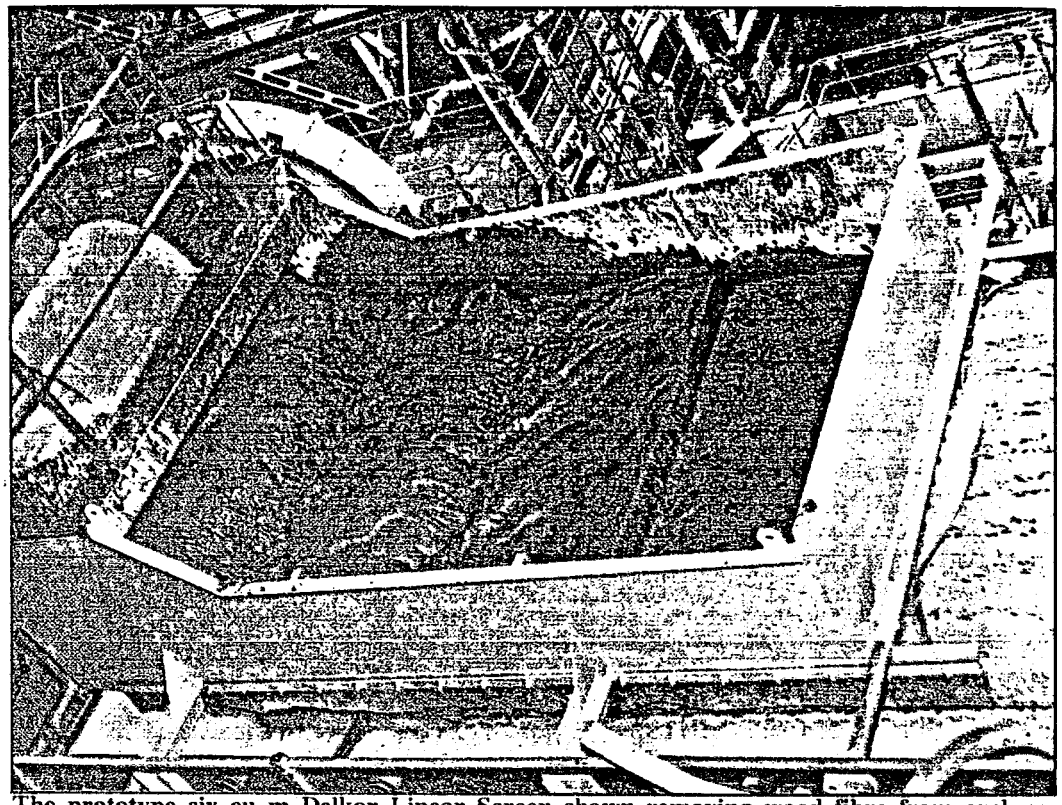
Mr Peter Sampson, sales manager of Delkor Technik, said that since the successful commissioning of the prototype at Afrikaner Lease, great interest had been shown in the unit by representatives in the mining industry and among major orders placed for new projects and replacement of conventional equipment were:

- East Daggafontein, 8x20sq m DLS
- Western Deep Levels, 2x12sq m DLS
- Harmony Gold Mine, 4x12sq m DLS
- Newmont Mining (Australia), 4x 6sq m DLS
- Ergo (Simmergo)

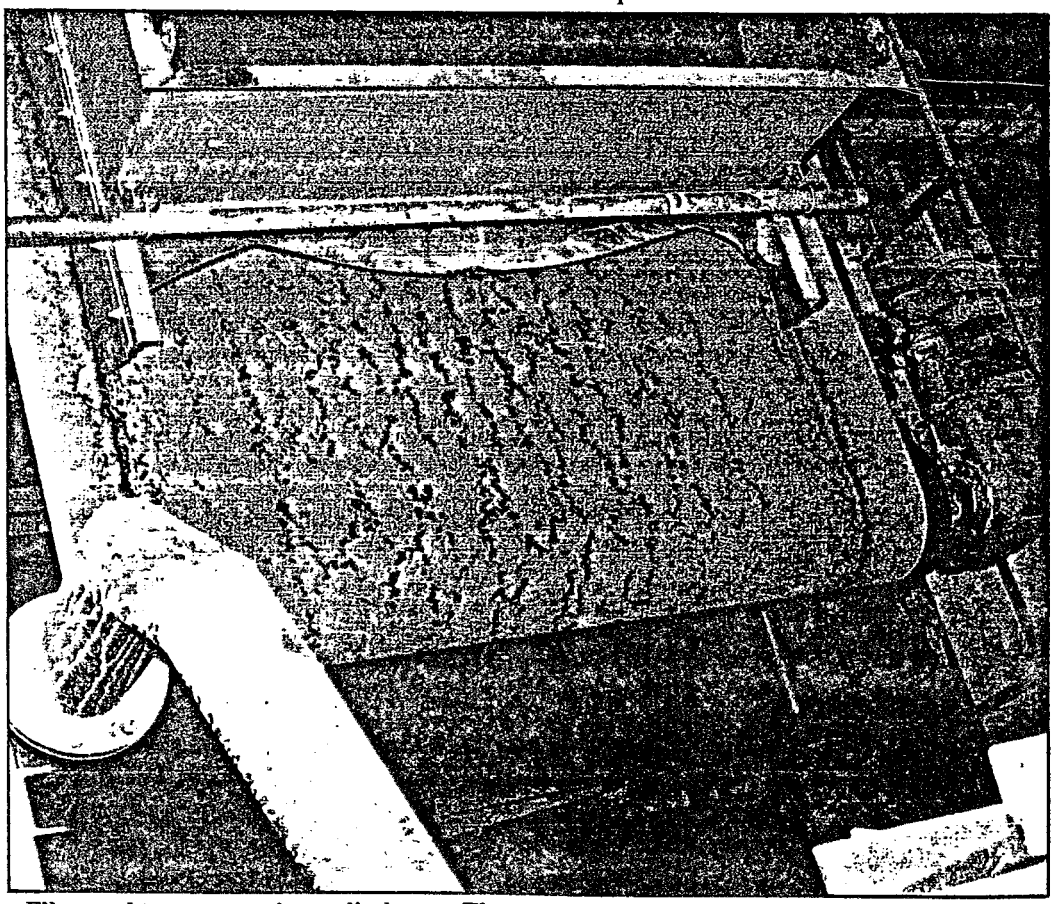
The DLS has been patented worldwide by Anglo American Corporation and is sold by Delkor Technik under licence on a worldwide basis. The standard product range is 0,5sq m, 1,5sq m, 3sq m, 6sq m, 12sq m and 20sq m.

The DLS has already proved to be extremely flexible with units being applied to cyclone overflow, thickener underflow, plant sands feed, carbon scavenging and resin bead screening.

Intensive research programmes are continuing between AAC and Delkor to investigate the use of the machine at cut points down to 100-mesh for use in milling circuits.



The prototype six cu m Delkor Linear Screen shown removing wood fibre from cyclone overflow during commissioning of the screen at Afrikander Leases Division of Vaal Reefs Gold Mine at Klerksdorp.



Fibre and tramp oversize at discharge. The screen removes 98 percent of oversize material.

Rogold's carbon-in-pulp gold treatment plant will be ready for full production next month

The 10 000 tons-a-month carbon-in-pulp gold treatment plant at Roodepoort Gold Holdings (Rogold) on the West Rand is expected to be commissioned in October.

Rogold was listed on the Johannesburg Stock Exchange in May and is due to build up to a first-phase production rate of 10 000 tons of ore a month by next January.

The plant is being built by Van Eck and Lurie, a member of the E L Bateman group.

Says Rogold mine manager, John Simpson: "After detailed studies and tests it was decided to split the plant into two phases of 10 000 tons of ore a month each. Phase Two will be operative early in 1989."

The mine's existing shafts are being rehabilitated and re-equipped. Workings on several levels of the Wilford shaft have been cleared and supported.

The headgear at Wilford shaft is complete. The shaft has been cleared down to seven level and completely re-equipped to four level, from where the initial production will come.

"The main winder was installed in July, accelerating the clearing and re-equipping process," says Mr Simpson.

The mine's management plans to start stoping this month to build up a stockpile for the initial mill feed.

This will considerably cut down on pre-production expenditure and conserve capital.

Says Mr Simpson: "We are building up our available ore reserves with about 500 m of re-equipped old reef drives and 70 m of prepared face for stoping."

Rogold holds claims covering more than 73 ha in the Roodepoort district on the West Rand, at the top of Durban Roodepoort Deep.

Rogold also has a tribute agreement to mine gold and silver from about 65 ha of adjacent Durban Roodepoort mining title.

So you think that we're living in a gold boom today?

With gold threatening to break through the $500 barrier on world markets and the stock exchanges in the grip of gold fever, one would expect records to tumble.

This is also the reason why several of the before-the-turn-of-the-century gold mines can again operate profitably.

But at a time when there was no fixed gold price — the days of the original Barberton gold fields — gold fever was perhaps even more rife than it is today.

A report from the Eastern Star of 100 years ago said: "In Venezuela, Central America, there exists a gold mine of such 'passing richness' that a nominal £440 share on which only £40 had been paid up, has been sold for the sum of £47 000.

"And why? Simply because each £440 share gives an annual income of some £9 000.

"In Australia and New Zealand, £1 shares have reached over £200."

Switchgear order

AEG has been awarded a contract by Gencor to supply 6,6 kV switchgear panels for two new mine sub-stations at St Helena and Beatrix mines in Welkom.

The contract requires AEG to supply a total of 41 type S2 double-bus bar panels, each rated at 6,6 kV 25 kA and vacuum circuit breakers, by October of Johannesburg.

Delkor linear screens bats past the ton

The order for the 100th Delkor linear screen has come from an Anglo American-administered mine, Vaal Reefs Exploration and Mining Company, No 9 shaft complex, where it will replace a vibrating screen.

This will be the third 6 sq m DLS at No 9 shaft.

The DLS was a joint development by Anglo American and Delkor Technik to solve the problem of woodchip and fibre removal from the carbon-in-pulp process used in gold recovery.

It is manufactured by Delkor under licence to Anglo American which has taken out patents worldwide.

Of the 100 units sold in a little over two years since the first prototype was installed, 16 have gone to Australia, four to North America and one to north Africa.

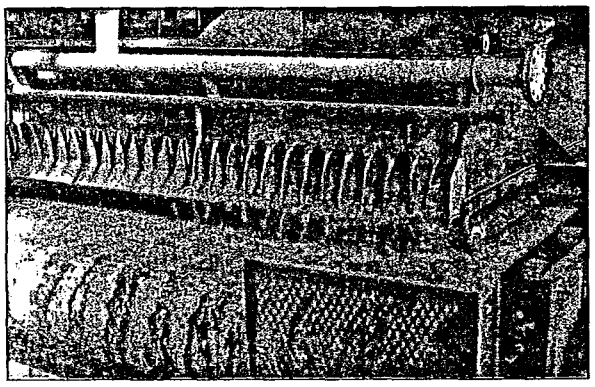
Dr Ray Davidson of Anglo American Research Laboratories, in collaboration with Delkor engineers and Vaal Reefs' metallurgical personnel, pioneered the use of a linear screen cloth to remove wood fibre.

The prototype unit was installed in 1985 at the Afrikander Lease Division of Vaal Reefs where it removed 98 percent of the wood fibre without blinding the fine apertures of the screen.

This unit is still in use at Afrikander Lease metallurgical plant.

All the major mining houses in South Africa are now using the DLS. Units in operation range from a 0,5 sq m screen at Nigel Gold Mine to eight 20 sq m screens at Ergo's Daggafontein Division.

The DLS has been applied successfully to fibre removal from cyclone overflow and thickener underflow, desanding plant feed and carbon scavenging.



Although most of the linear screens are installed at gold mines, success has been achieved in tramp oversize removal at a platinum mine and in a beach sand operation at Richards Bay.

The Delkor linear screen consists of a coarse monofilament screen cloth supported on cloth rollers and driven by a head pulley coupled to a drive unit.

Slurry is fed through a distributor on to the moving monofilament cloth.

The undersize material drains through the cloth by gravity and is collected in the underpan.

The wood fibre and oversize material retained on the screen are discharged at the drive pulley using a belt wash spray which ensures a continuous self-cleaning screen.

By adjusting the weave configuration of the screen cloth the cut point for oversize removal can easily be varied.

The DLS operates without vibration which, it is claimed, eliminates the need for costly support structures and uses far less power than vibrating screens.

It is virtually maintenance-free and the cloth is easily replaced, according to Delkor.

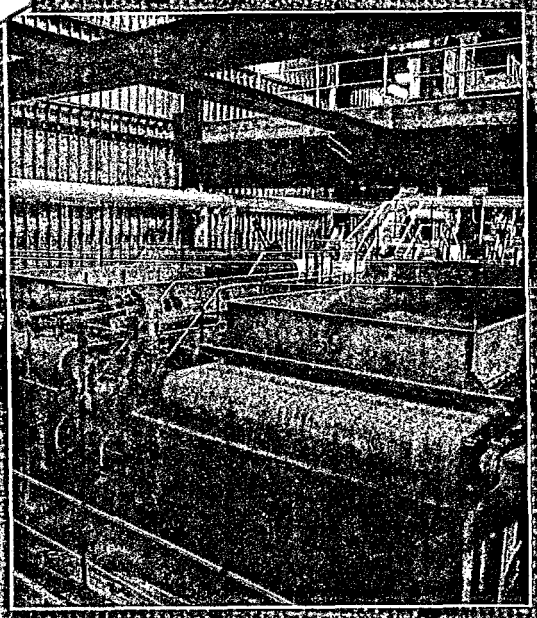
100 percent goal for separators

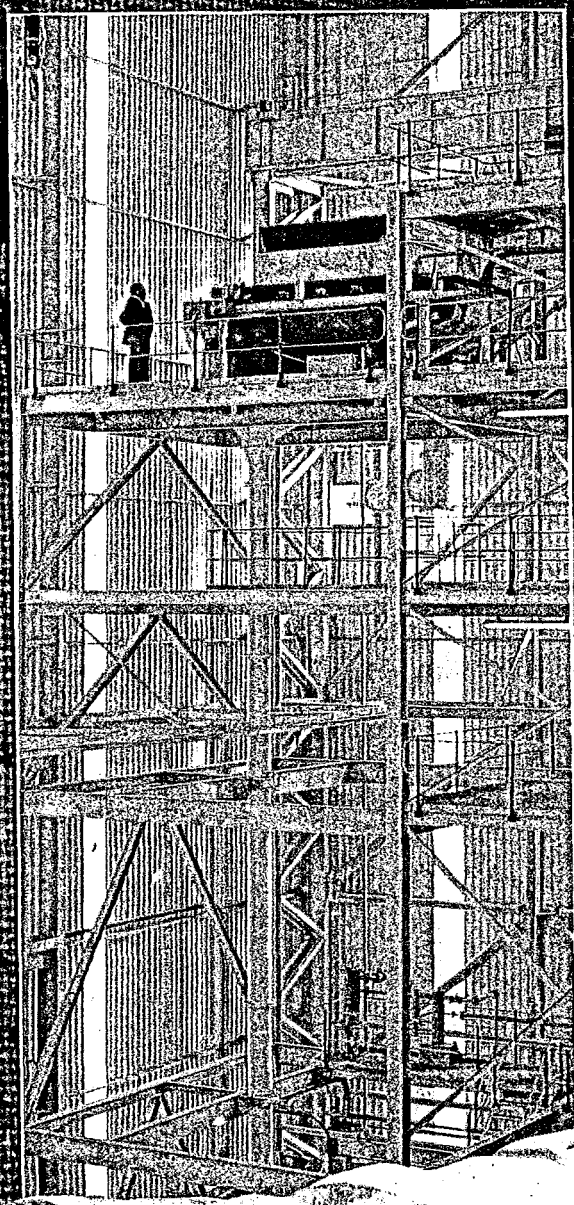
Lockers Engineers has increased the local content of Sweco Separators from 50 percent to 85 percent and plan to go up to 100 percent local content as soon as possible.

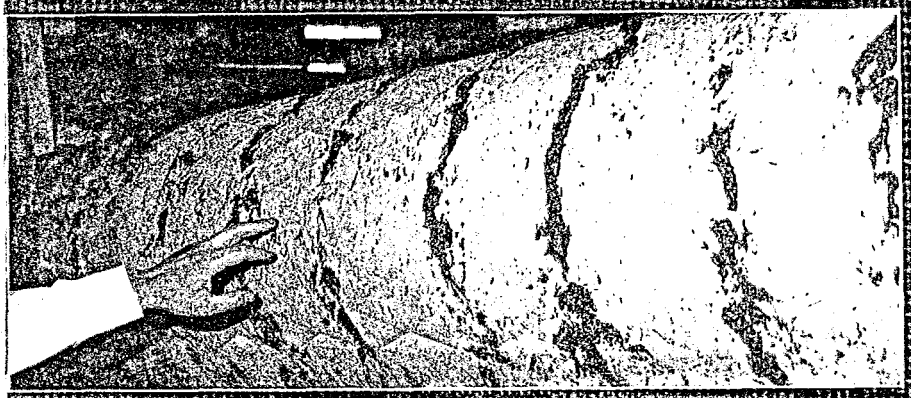
The end-user's cost of Sweco equipment has already been brought down by 20 percent as a result, and further cost reductions are possible as the programme achieves further increases in local content, Lockers claims.

Sweco Separators are widely used for the separation or recovery of solids from effluent or slurries in applications such as mining, the food industry and waste recycling.

DELKOR
LINEAR SCREEN







A breakthrough in fine screening technology

DELKOR

DEVELOPMENT

The Delkor Linear Screen (DLS) is a revolutionary screening system which has solved a major problem in the CIP process on gold plants. It was developed to remove wood fibre and tramp oversize from milled gold slurries and is capable of screening between 150 and 2 000 microns. The prototype DLS unit was built and installed by Delkor in January 1985. Since then numerous machines have performed extremely successfully on gold mines and gold dump reclamation plants. They have removed 98% of wood fibre and tramp oversize from slurries – compared with the 40 – 70% removal rate of other screening systems.

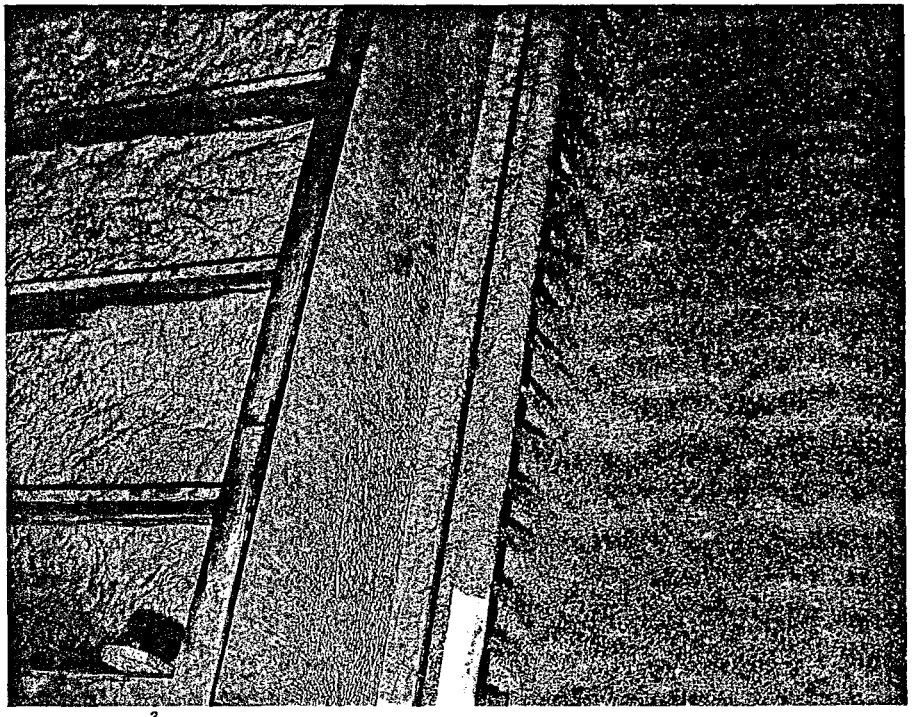
APPLICATIONS

The DLS removes wood fibre, carbon, vegetation, carbonaceous ash or any other tramp material from fine mineral or chemical slurries prior to:

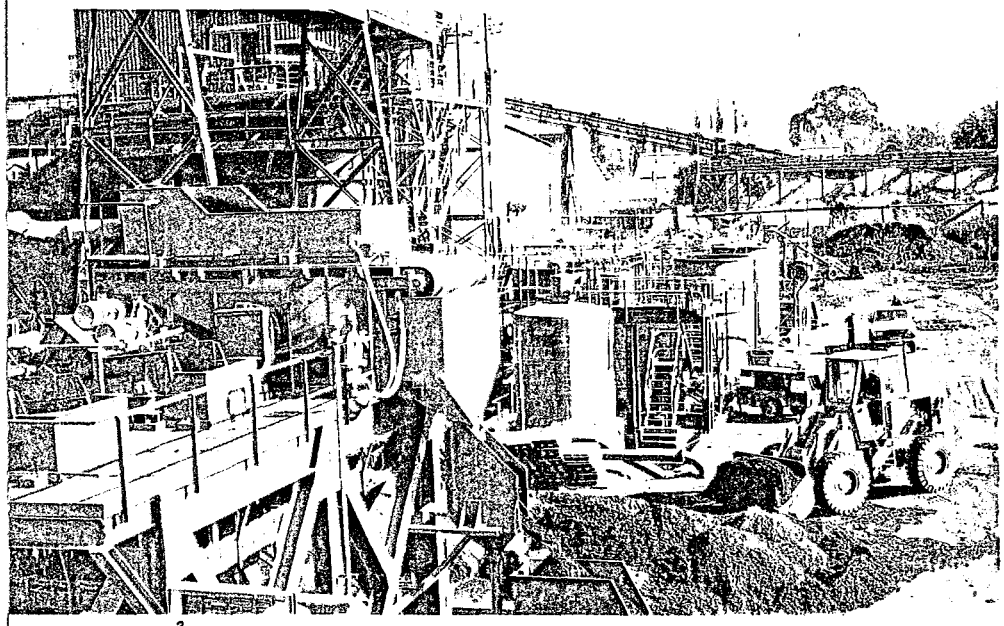
- Gold plant leach
- CIP adsorption
- Gravity concentration
- Flotation
- On stream analysis
- Dump retreatment processes

It is also suitable for:

- Removal of asbestos from liquid streams
- Removal of coarse boiler ash from repulped flyash dumps

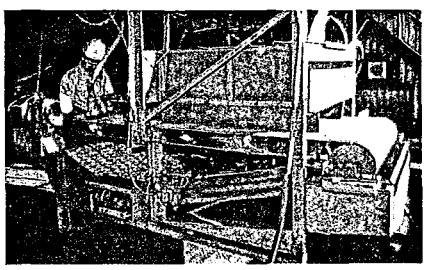


12m² DLS scalping reclaimed gold dump ash at 1,5mm from plant headfeed



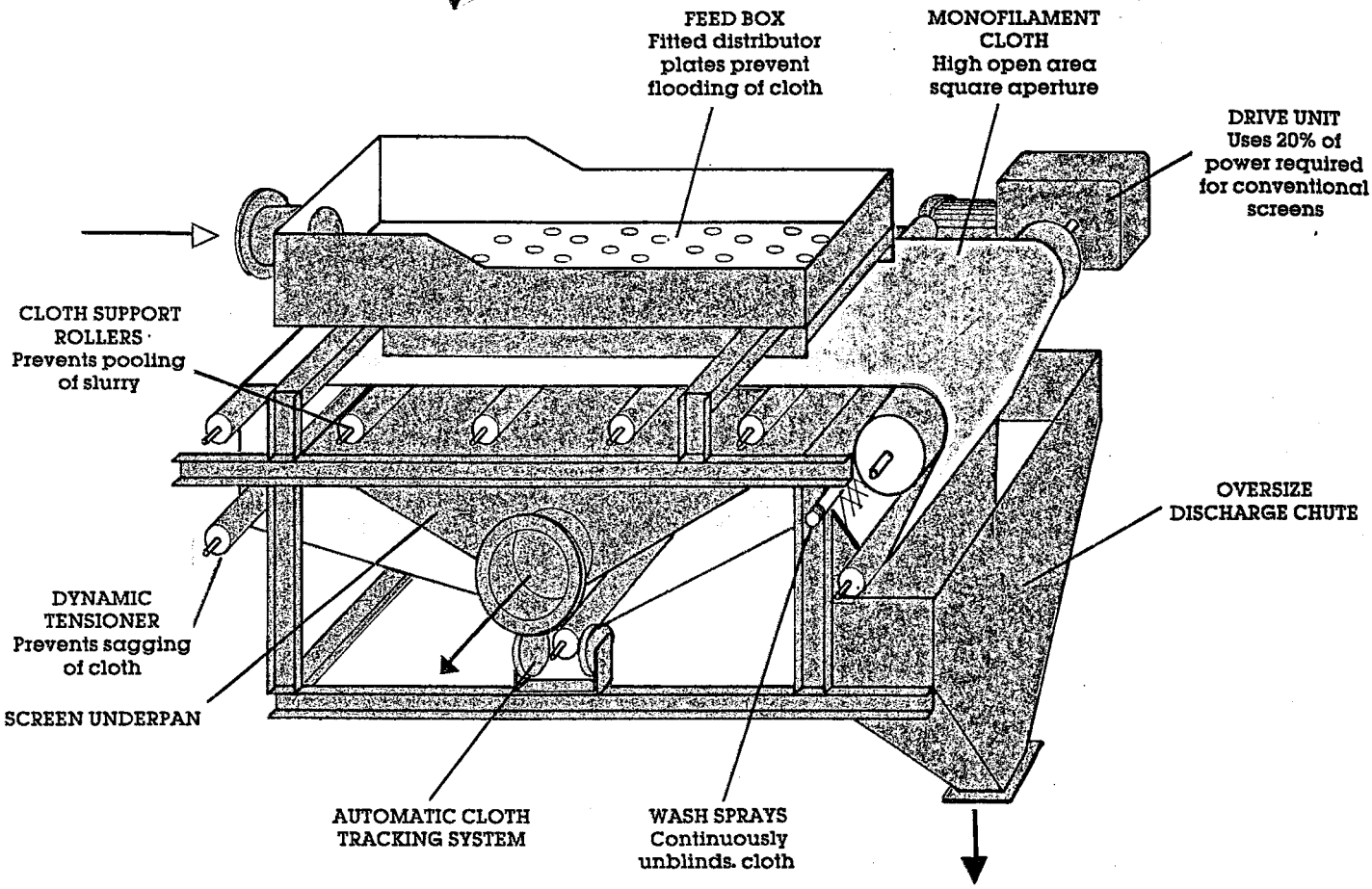
6m² DLS replacing two vibrating screens to remove carbonaceous ash from thickener underflow prior to gold leach

- Washing contaminated fine pulps in the paper industry
- Scavenging loaded carbon from gold plant tailings
- Fine screening of milling plant cyclone overflow
- Fine screening of feed to gravity concentration plants.



DLS pilot plant for fine screening at 180 micron in a milling circuit

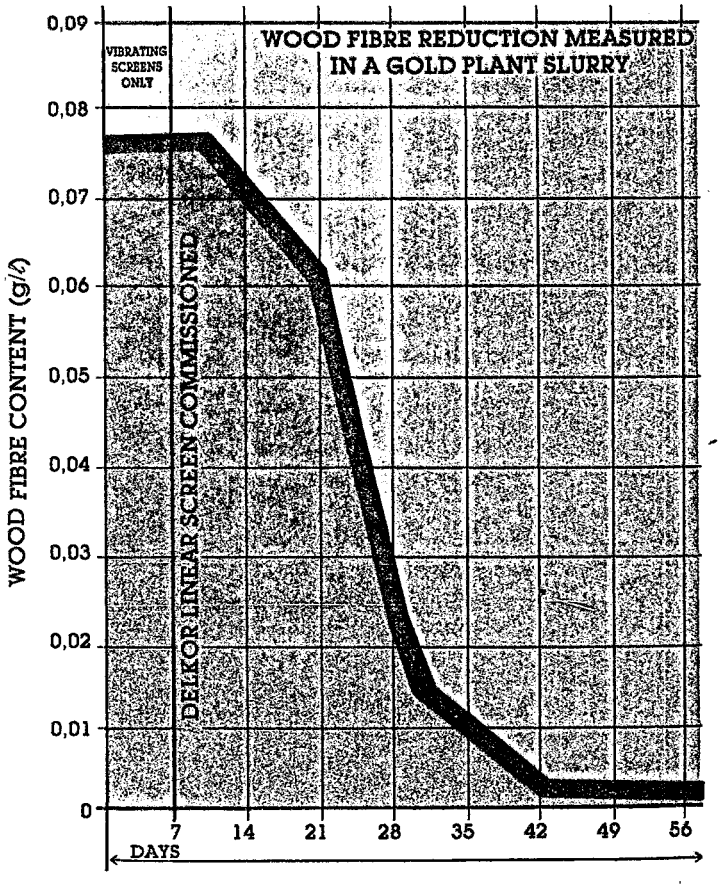
DELKOR



BENEFITS

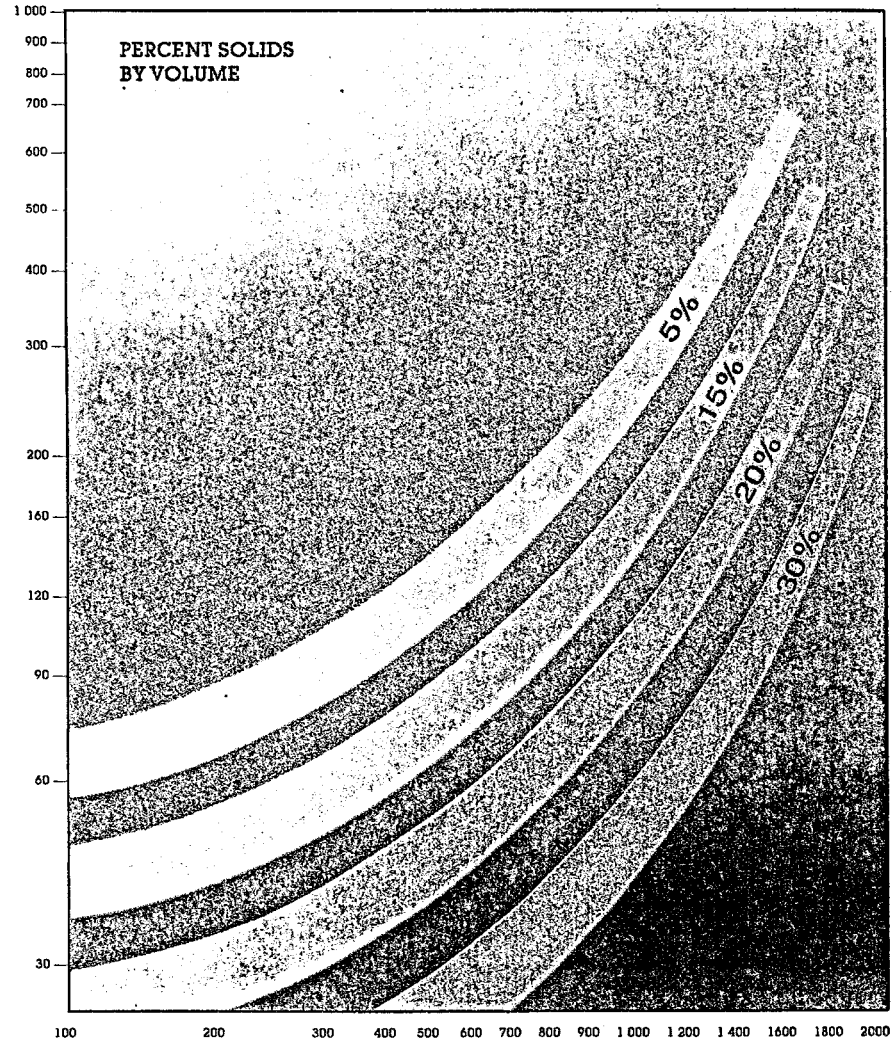
The DLS operates with no vibration. This prevents realignment of oversize material with the cloth aperture, ensuring retention of all oversize material. This resulted in 98,7% removal of wood fibre as measured on site by a mining group's research laboratory (see graph).

- Wide range of cloth apertures – from 150 μm to 2 000 μm
- Quick cloth changing time 20 – 30 minutes
- Cloth made of wear resistant polyester
- No vibration
- Low maintenance costs
- Low power consumption
- Low noise levels
- Low cost of supporting civils and structural steel
- Simple operation
- No high cost moving parts
- Availabilities of 99%
- No scaling of cloth from chemical precipitation



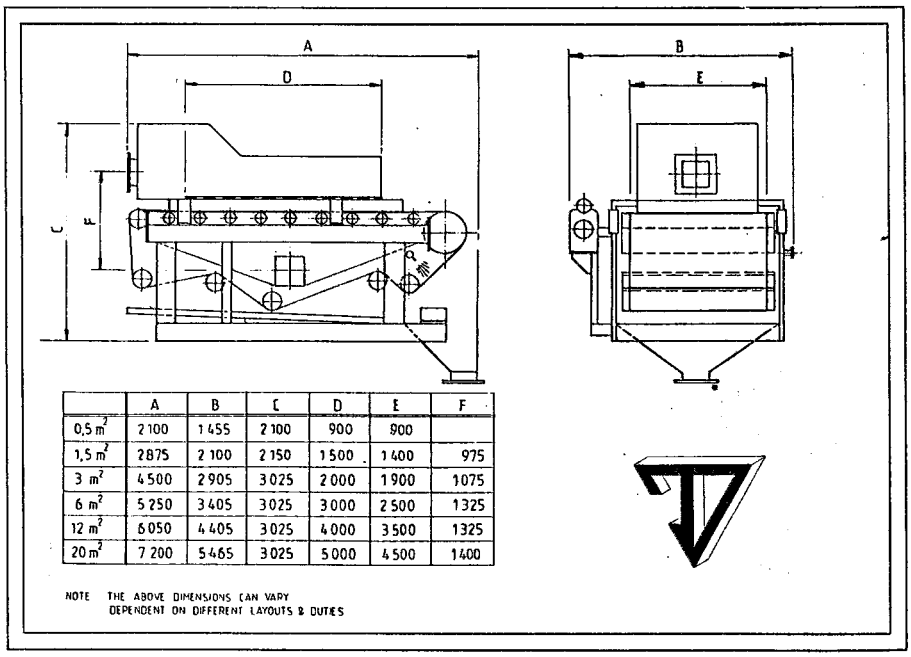
DELKOR

SCREEN CAPACITIES



NOTE:
These capacities are indicative and can only be verified by testwork

SCREEN DIMENSIONS



	A	B	C	D	E	F
0,5 m²	2100	1455	2100	900	900	
1,5 m²	2875	2100	2150	1500	1400	975
3 m²	4500	2905	3025	2000	1900	1075
6 m²	5250	3405	3025	3000	2500	1325
12 m²	6050	4405	3025	4000	3500	1325
20 m²	7200	5465	3025	5000	4500	1400

NOTE THE ABOVE DIMENSIONS CAN VARY DEPENDENT ON DIFFERENT LAYOUTS & DUTIES

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Tel. 802-8702

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Tel. 387-6900

DELKOR MINERALTECHNIK GmbH
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West Germany
Tel. 02361 12024 Telex 829107

MANUFACTURED UNDER LICENCE
TO ANGLO AMERICAN CORPORATION

DLS-REF # 141

DELKOR LINEAR SCREENS
REFERENCE LIST AS AT 25.2.91

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M^2 SIZE OF UNITS
AAC	Afrikander Leases	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Rand Mines	Harmony 4 Shaft	RSA	Tramp Oversize Removal, Cyclone Overflow	2	$12m^2$
Rand Mines	Harmony 4 Shaft	RSA	Tramp Oversize Removal, CIP Feed	1	$12m^2$
Rand Mines	Harmony 4 Shaft	RSA	Carbon Scavenging, CIP Residue	1	$12m^2$
Rand Mines	Virginia Section	RSA	Tramp Oversize Removal, Cyclone Overflow	2	$6m^2$
AAC	Simmergo	RSA	Tramp Oversize Removal, CIP Feed	1	$6m^2$
Rand Mines	RM3 Plant	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Mintek	-	RSA	Research purposes	3	$0,5m^2$
AAC	Western Deep Levels	RSA	Tramp Oversize Removal, Cyclone Overflow	2	$12m^2$
Nigel Town Council		RSA	Tramp and Root Matter Removal, Plant Feed	1	$3m^2$
AAC	East Daggafontein	RSA	Tramp and Ash Removal, Plant Feed	4	$20m^2$
AAC	East Daggafontein	RSA	Carbon Scavenging, CIL Residue	4	$20m^2$
AAC	S.A.Lands	RSA	Tramp and Ash Removal, Plant Feed	1	$12m^2$
Golden Dumps	Springs Dagga	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
Golden Dumps	Springs Dagga	RSA	Carbon Scavenging, CIP Residue	1	$1,5m^2$
AAC	Simmergo	RSA	Tramp and Ash Removal, Plant Feed	1	$12m^2$
AAC	FSG	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
			SUB-TOTAL	29	

DELKOR LINEAR SCREENS

REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M^2 SIZE OF UNITS
Gencor	Impala Platinum	RSA	Tramp Oversize Removal, Float Plant Feed	2	$1m^2$
Delkor		RSA	Research Units	2	$0,5m^2$
New Union Gold Mine		RSA	Tramp Oversize Removal, Plant Feed	1	$0,5m^2$
Newmont Mining	Telfer Gold Mine	AUSTRALIA	Tramp Oversize Removal, Plant Feed	2	$6m^2$
Newmont Mining	Telfer Gold Mine	AUSTRALIA	Carbon Scavenging, CIP Residue	2	$6m^2$
Kalgoorlie Mining Assoc.		AUSTRALIA	Tramp Oversize Removal	1	$3m^2$
Rand Leases Gold Mine		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Anglovaal	Sheba Gold Mine	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
Anglovaal	Hartebeestfontein	RSA	Tramp Oversize Removal, Cyclone Overflow	2	$12m^2$
JCI	Doornkop	RSA	Tramp Oversize Removal, Cyclone Overflow	6	$6m^2$
JCI	Doornkop	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Nevoria Gold Mine		AUSTRALIA	Tramp Oversize Removal	1	$1,5m^2$
Western Mining Co. (Kambalda)		AUSTRALIA	Carbon Scavenging, CIP Residue	1	$6m^2$
AAC	Vaal Reefs 9 Shaft	RSA	Tramp Oversize Removal, Cyclone Overflow	4	$6m^2$
Rand Mines	City Sands	RSA	Tramp and Ash Removal, Plant Feed	2	$12m^2$
Rand Mines	City Sands	RSA	Tramp Oversize Removal, CIP Feed	2	$12m^2$
Rand Mines	City Sands	RSA	Carbon Scavenging, CIP Residue	2	$12m^2$
			SUB-TOTAL	34	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M^2 SIZE OF UNITS
JCI	H J Joel	RSA	Tramp Oversize Removal, Cyclone Overflow	2	$6m^2$
JCI	H J Joel	RSA	Tramp Oversize Removal, CIP Feed	2	$6m^2$
JCI	H J Joel	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Sub-Nigel Gold Mining Co.		RSA	Tramp Oversize Removal, Cyclone Overflow	2	$1,5m^2$
Ashanti Goldfields Corp.		Ghana	Tramp Oversize Removal, CIP Feed	2	$3,0m^2$
Ashanti Goldfields Corp		Ghana	Tramp Oversize Removal, CIP Feed	1	$6,0m^2$
Bardoc		Australia	Tramp Oversize Removal	1	$1,5m^2$
Delkor		Australia	Research and Development	1	$1,5m^2$
Hannan's South Gold Mine		Australia	Tramp Oversize Removal	1	$1,5m^2$
Nigel Gold Mine No.1 Plant		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
Nigel Gold Mine No.1 Plant		RSA	Carbon Scavenging, CIP Residue	1	$0,5m^2$
Nigel Gold Mine No.2 Plant		RSA	Carbon Scavenging, CIP Residue	1	$1,5m^2$
Nigel Gold Mine No.2 Plant		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
Anglovaal	Hartebeesfontein	RSA	Tramp Oversize Removal, Float Plant Feed	1	$3m^2$
Campbell Red Lake Mines		CANADA	Tramp Oversize Removal, Cyclone Overflow	2	$1,5m^2$
Chimney Creek Mine		USA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
AAC	Western Deep Levels	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Anglovaal	Village Main	RSA	Tramp and Ash Removal, Plant Feed	1	$6m^2$
			SUB TOTAL	25	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M^2 SIZE OF UNITS
Mount Wilkinson		Australia	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
Worsley Alumina	Boddington	Australia	Carbon Scavenging, CIP Residue	4	$12m^2$
Worsley Alumina	Boddington	Australia	Tramp Oversize Removal, Cyclone Overflow	1	$12m^2$
Worsley Alumina	Boddington	Australia	Tramp Oversize Removal, Cyclone Overflow	2	$20m^2$
Newmont Gold	Carlin No. 2	USA	Tramp Oversize Removal, Cyclone Overflow	1	$12m^2$
Gencor	Impala Platinum	RSA	Tramp Oversize Removal, Spray Drier Feed	3	$3m^2$
Rand Mines	T.G.M.E.	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
Rand Mines	T.G.M.E.	RSA	Carbon Scavenging, CIP Residue	1	$1m^2$
Severin Mining	Eersteling	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1m^2$
Severin Mining	Eersteling	RSA	Carbon Scavenging, CIP Residue	1	$0,5m^2$
Gencor	Beatrix	RSA	Tramp Oversize Removal, Cyclone Overflow	3	$6m^2$
Gencor	Beatrix	RSA	Carbon Scavenging, CIL Residue	1	$12m^2$
AAC	Welkom G.M.	RSA	Tramp Oversize Removal,Cyclone Overflow	2	3 & $6m^2$
Anglovaal	E.T.C.	RSA	Carbon Scavenging, CIP Residue	3	$1m^2$
AAC	Vaal Reefs No 9	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
AAC	Vaal Reefs No 9	RSA	Tramp oversize removal, Thickener underflow	1	$6m^2$
Dome Mines	South Porcupine	Canada	Tramp Oversize Removal, Cyclone Overflow	2	$1m^2$
Roodepoort Gold Mine		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1m^2$
			SUB TOTAL	31	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M^2 SIZE OF UNITS
Croydon Gold Mine		Australia	Tramp Oversize Removal	2	$3m^2$
Richards Bay Minerals		RSA	Quartz Removal from Heavy Mineral Conc.	2	$3m^2$
Goldfields Venterspost		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
Rand Mines	E.R.P.M.	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Anglovaal	Lorraine	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$3m^2$
Ashanti Goldfields Corp.		Ghana	Tramp Oversize Removal, Float Plant Feed	2	$3m^2$
Paracale Gold Corp.		Phillipines	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
Gencor	Kinross	RSA	Tramp Oversize Removal, Cyclone Overflow	1	$20m^2$
Paddington Gold Mine		Australia	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
A.A.C.	President Steyn	RSA	Tramp Oversize Removal, Backfill Product	1	$6m^2$
A.A.C.	Ergo	RSA	Carbon Recovery	6	$1m^2$
ERG Resources	Timmins	Canada	Tramp Oversize Removal, Flotation Conc.	2	12 & $6m^2$
ERG Resources	Timmins	Canada	Carbon Scavenging, CIP Residue	1	$6m^2$
Gencor	Buffelsfontein	RSA	Woodfibre Recovery, Kamyr Process	1	$3m^2$
Martha's Hill		N'Zealand	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Mount Morgan		Australia	Deslime Vibrating Screen Oversize	1	$1,5m^2$
Rand Mines	RM3	RSA	Tramp and Ash Removal, Sand Plant feed	3	$412m^2$
			SUB TOTAL	28	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
Southgo	Gold Dam	RSA	Tramp and Ash Removal, Sand Plant feed	2	$12m^2$
Southgo	Gold Dam	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Rand Mines	T.G.M.E.	RSA	Tramp and Ash Removal, Plant feed	1	$6m^2$
A.A.C.	Simmergo	RSA	Carbon Scavenging, CIP Residue	1	$12m^2$
Gencor	Buffelsfontein	RSA	Calcine Plant, Tramp & Carbon Recovery	2	$1m^2$
Goldfields	Leeudorn	RSA	Tramp Oversize Removal, Cyclone Overflow	3	$6m^2$
Goldfields	Leeudorn	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
Goldfields	Leeudorn	RSA	Loaded Carbon Recovery, CIP	2	$1m^2$
Tongaat Hullets	Maidstone	RSA	Bagasse Screening, Sugar Mill Train	1	$12m^2$
Newmont Gold	Carlin No. 5	USA	Tramp Oversize Removal, Cyclone Overflow	4	$20m^2$
Newmont Gold	Carlin No. 5	USA	Carbon Scavenging, CIP Residue	3	$20m^2$
Cleveland Potash		UK	Test Unit	1	$0,5m^2$
Mocupia Tailings	Plant	Venezuela	Tramp Oversize removal, Cyclone Overflow	1	$3m^2$
Salsigne Gold		France	Tramp Oversize removal, Cyclone Overflow	2	$6m^2$
Salsigne Gold		France	Carbon Scavenging, CIP Residue	1	$1,5m^2$
			SUB TOTAL	28	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
AAC	Vaal Reefs No 8	RSA	Tramp Oversize removal, Thickener Underflow	2	$12m^2$
AAC	Vaal Reefs No 8	RSA	Carbon Scavenging, CIP Residue	2	$12m^2$
Rand Mines	Barbrook	RSA	Tramp Oversize removal, Cyclone Overflow	2	$1,5m^2$
Rand Mines	Barbrook	RSA	Carbon Scavenging, CIP residue	1	$0,5m^2$
Gazgold	Letaba	RSA	Tramp Oversize removal, Cyclone Overflow	3	$1\frac{1}{2}m^2$
De Beers	Premier	RSA	Tramp Removal, Process Water	1	$6m^2$
Lonrho	Shamva Mine	Zimbabwe	Tramp Oversize removal, CIP feed	3	$3m^2$
North Kalgurli Mines		Australia	Tramp Removal, Cyclone Overflow	2	$9m^2$
Mineral Hill		Australia	Tramp Removal, Cyclone Overflow	2	$1,5m^2$
Hemlo Mines	Giant Mine	Canada	Tramp Removal, Cyclone Overflow	1	$6m^2$
Rand Mines	RM3	RSA	Tramp and Ash Removal, Sand Plant Feed	2	$20m^2$
AAC	Navachab	Namibia	Tramp Removal, Cyclone Overflow	2	$6m^2$
AAC	Navachab	Namibia	Carbon Scavenging, CIP Residue	2	$3m^2$
Southgo	Knights	RSA	Tramp and Ash Removal, Sand Plant Feed	2	$12m^2$
Southgo	Knights	RSA	Carbon Scavenging, CIP Residue	2	$6m^2$
AAC	Ergo Division	RSA	Tramp and Ash Removal, CIL Plant Feed	4	$25m^2$
			SUB TOTAL	33	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
AAC	Ergo Divison	RSA	Dewatering DLS oversize	2	$3m^2$
Arcturus Mine		Zimbabwe	Tramp Removal, Cyclone Overflow	2	$1m^2$
Gencor	St Helena	RSA	Tramp Removal, Cyclone Overflow	3	$6m^2$
Gencor	St Helena	RSA	Tramp Removal, Cyclone Overflow	2	$12m^2$
Colosseum Mine		USA	Tramp removal, Cyclone Overflow	1	$6m^2$
Donna Lake Mine		Canada	Tramp Removal, Cyclone Overflow	1	$1,5m^2$
Newmont Gold	Carlin No. 4	USA	Tramp Removal, Cyclone Overflow	2	$12m^2$
Newmont Gold	Carlin No. 4	USA	Carbon Scavenging, CIP Residue	1	$12m^2$
Jerritt Canyon		USA	Carbon Scavenging, Cyclone Overflow	1	$20m^2$
Jerritt Canyon		USA	Tramp Removal, CIP Residue	1	$6m^2$
Bullfrog Mine		USA	Tramp Removal, Cyclone Overflow	2	$6m^2$
Burmine Operations-Copperhead		Australia	Tramp Removal, Cyclone Overflow	1	$3m^2$
Burmine Operations-Copperhead		Australia	Carbon Scavenging, CIP Residue	1	$1\frac{1}{2}m^2$
Reef & Basemetals		RSA	Tramp Removal, Float Plant Feed	1	$1m^2$
Rand Mines	Durban Deep	RSA	Tramp Removal, Cyclone Overflow	1	$6m^2$
			SUB TOTAL	22	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
Lonrho		Zimbabwe	Tramp Removal, Cyclone Overflow	1	$3m^2$
Rand Mines	Lefkochrysos	RSA	Tramp Removal, Float Plant Feed	2	$6m^2$
RTZ	Renko	Zimbabwe	Tramp Removal, Cyclone Overflow	1	$1\frac{1}{2}m^2$
AAC	Brakpan Sands	RSA	Tramp & Ash Removal, CIL Plant Feed	2	$12m^2$
JCI	Ammandlebult	RSA	Tramp Removal, Float Plant Feed	1	$1\frac{1}{2}m^2$
Lindum Reefs		RSA	Tramp Removal, Cyclone Overflow	2	$6m^2$
Lindum Reefs		RSA	Carbon Scavenging, CIP Residue	1	$3m^2$
Gencor	Winkelhaak	RSA	Tramp Removal, Cyclone Overflow	2	$6m^2$
Minprok	Kalgoorie	Australia	Tramp Oversize Removal, Cyclone Overflow	2	$12m^2$
Minprok	Kalgoorie	Australia	Carbon Scavenging, CIP Residue	2	$12m^2$
Atlas Mining	Kalgoorie	Phillipines	Tramp Oversize Removal, Cyclone Overflow	1	$9m^2$
Barrack Mining	Wiluna	Australia	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Freeport Mining	Nevada	USA	Tramp Oversize Removal, Cyclone Overflow	2	$6m^2$
Lega Dembi		Ethiopia	Tramp Oversize Removal, Cyclone Overflow	2	$6m^2$
Lega Dembi		Ethiopia	Carbon Scavenging, CIP Residue	1	$6m^2$
			SUB TOTAL	23	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
JCI	Cooke Plant	RSA	Tramp Oversize Removal, Cyclone Overflow	3	$12m^2$
JCI	Cooke Plant	RSA	Tramp Oversize Removal, Thickener Underflow	3	$12m^2$
Giant	Yellow Knife	Canada	Tramp Oversize Removal, Cyclone Overflow	2	$12m^2$
Cam & Motor		Zimbabwe	Tramp Oversize Removal, Cyclone Overflow	1	$6m^2$
Whim Creek	Iabouchere	Australia	Tramp Oversize and Carbon Recovery	2	$3m^2$
Rosario Gold Project		Dom Rep	Tramp Oversize Removal, Cyclone Overflow	2	$20m^2$
Rosario Gold Project		Dom Rep	Carbon Scavenging, CIP Residue	1	$3m^2$
Connary Minerals		Ireland	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
Mount Morgan		Australia	Carbon Scavenging, CIP Residue	1	$3m^2$
Placer Dome Porgera		N'Guinea	Tramp Removal, Flotation Concentration	2	$3m^2$
Placer Dome Porgera		N'Guinea	Carbon Scavenging, CIP Residue	2	$3m^2$
Rand Mines Merriespruit		RSA	Tramp Oversize Removal, Thickener Underflow	2	$12m^2$
Primrose G.M.		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1,5m^2$
African Gold		RSA	Tramp Oversize Removal, Cyclone Overflow	1	$1m^2$
Gencor	Oryx	RSA	Loaded Carbon Recovery,	2	$6m^2$
			SUB TOTAL	26	

DELKOR LINEAR SCREENS
REFERENCE LIST

CLIENT		COUNTRY	APPLICATION	NO. OF UNITS	M2 SIZE OF UNITS
Rand Mines	TGME	RSA	Woodchip removal	1	$3m^2$
Benoni GMCo	Benoni Sands	RSA	Woodchip removal	1	$6m^2$
				1	$12m^2$
Gecamines		RSA	Test unit	1	$0,5m^2$
Johannesburg	Municipality	RSA	Sludge dewatering	1	$8m^2$
Kaltails		Australia	Trash recovery dump retreatment	2	$12m^2$
Benguet		Phillipines	Trash recovery, dump retreatment	2	$6m^2$
Goldfields	Northam	RSA	Trash recovery, Float plant feed	2	$6m^2$
Ferrometals		RSA	Asbestos Recovery, blast furnace water	1	$6m^2$
			SUB TOTAL	12	
			TOTAL	313	

2.3 INTERSTAGE CARBON SCREENS

One of the keys to successful CIP operation is considered to be the interstage screening operation and significant developments have been made in the past ten years. These developments are summarised in the two appended papers. The paper entitled Screening Innovations in AAC Carbon-in-Pulp and Carbon-on-Leach Plants received a bronze medal award from the SA Mechanical Engineering Society.

While most of the screening innovations relate to the development of mechanically swept screens in the late 1980's, the author provisionally patented perhaps the first mechanically swept screen in July 1980 (provisional patent appended).

Screening innovations in AAC carbon-in-pulp and carbon-in-leach plants

by John Moore, Peter Smith and Dr Ray Davidson

The advent of the higher gold price in the 1970s coupled with a need to contain escalating capital and operating costs in the South African industry focused considerable attention on the carbon-in-pulp (CIP) process. The technology together with many innovative changes has been adopted to the virtual exclusion of alternative processing routes for the treatment of the more free-milling type ores. Since the commissioning in 1980 of the first large plant in South Africa at the President Brand Gold Mine, today nearly half of the tonnage treated and nearly one-third of the gold produced uses the CIP process.

The following report covers two areas where major advances have been made in CIP plant design, namely pre-screening of feed pulp and interstage screening of carbon.

Mr Moore is Consulting Metallurgist of the Gold and Uranium Division and Mr Smith Senior Divisional Metallurgist of the Technical Director's office at Anglo American Corporation. Dr Davidson is Senior Research Fellow at the Anglo American Research Laboratories.

Introduction

128.

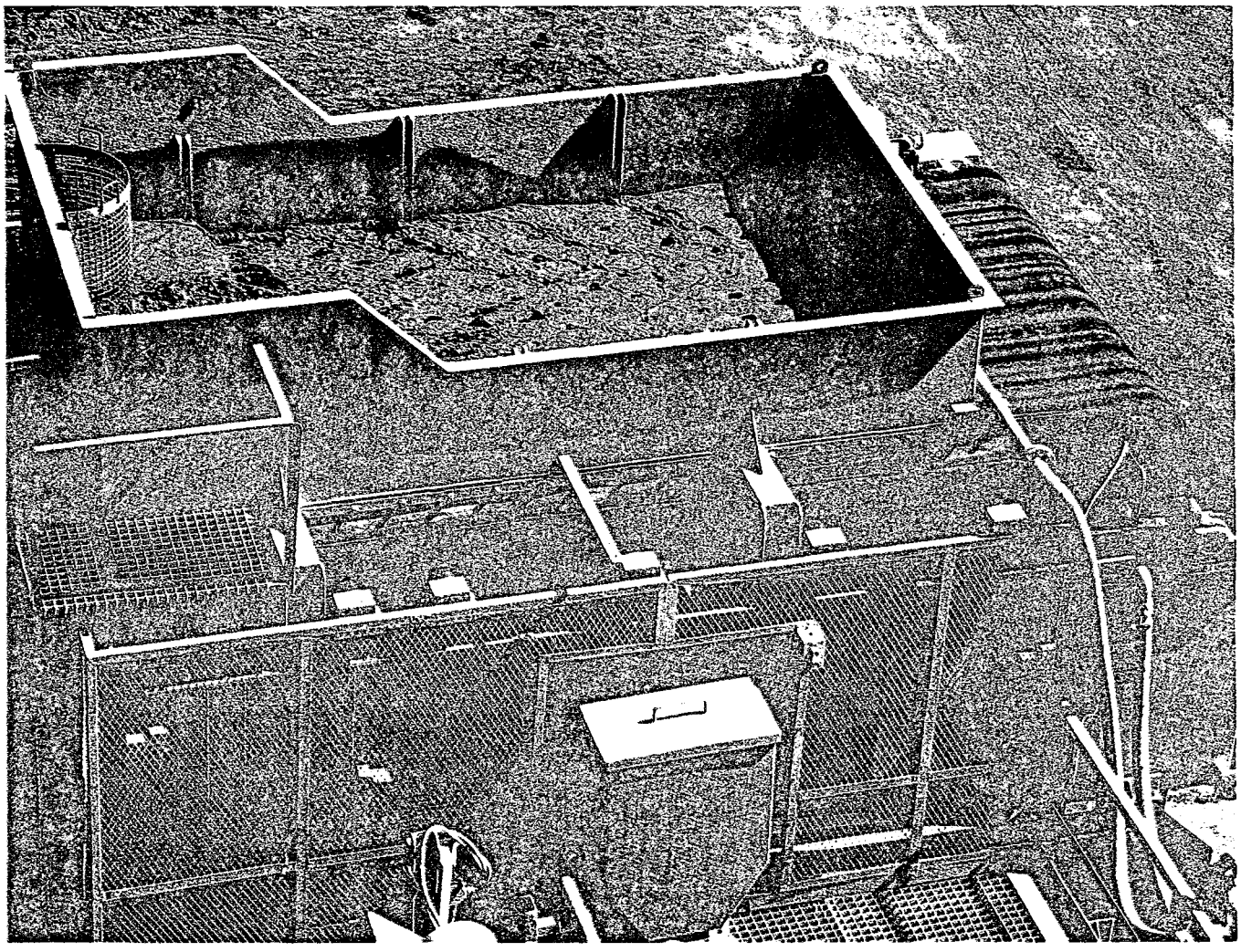
The advent of the higher gold price in the 1970s coupled with a need to contain escalating capital and operating costs in the South African industry focused considerable attention on the carbon-in-pulp (CIP) process. From 1975 the CIP process has developed explosively world-wide. The technology together with many innovative changes has been adopted to the virtual exclusion of alternative processing routes for the treatment of the more free milling type ores. Since the commissioning in 1980 of the first large plant in South Africa at the President Brand Gold Mine, today nearly half of the tonnage treated and nearly one-third of the gold produced uses the CIP process.

This report covers two areas where major advances have been made in CIP plant design, namely the pre-screening of feed pulp and the interstage screening of carbon.

Pre-screening of feed pulp

Due primarily to the very rapid development of the CIP process, the benefits

6m² Linear Screen installed at Simmergo in June 1985.



of effective feed preparation have until recently received little attention. In many of the 20 large plants and 10 small plants currently operating in South Africa, the presence of oversize materials in the feed pulp have presented major operating difficulties. As may be expected the particle size and pulp density of the feed pulp play a significant downstream role in any CIP operation. The presence of oversize material in the form of grit has obvious implications when considering interstage screen duties and sanding out problems in carbon contactors. Other contaminants in the form of woodchips (and woodfibre) and plastic insulating material originating from mining and milling operations also present considerable problems in some circuits.

Likewise, in some dump treatment operations the presence of coal ash and vegetable trash in the form of grass, roots, etc., may cause plant malfunction. The elimination of such materials in some sort of pre-screening circuit has been found to be most necessary in attaining satisfactory plant performance. These contaminants and their effects on the various CIP circuits will be discussed together with a review of the pre-screening techniques used to improve overall plant performance.

Woodfibre

The progressive build-up of woodfibre in various circuits in early CIP plants in South Africa highlighted an unforeseen problem of major magnitude. The concentration of woodfibre in carbon contactors caused blinding of interstage screens resulting in the co-current flow of carbon and loss of metallurgical efficiency. The transfer of loaded carbon containing woodfibre (up to 25 per cent by volume in some severe cases) to the elution and regeneration circuits also presented many problems. Physical choking of transfer lines, column distributors and collectors, carbon undersize screens and kiln screw feeders were some of the problems encountered. The volumetric dilution of carbon with this material also contributed to poor elution and regeneration efficiency, while the partial activation of woodchips in the regeneration kiln may well have some downstream disadvantages.

Due primarily to the physical problems associated with the presence of woodfibre in CIP plants treating run-of-mine material, the pre-screening of this material either in the milling circuit or prior to CIP has become necessary. This pre-screening of woodfibre has itself highlighted a further metallurgical problem related to the gold content of the collected material. Woodfibre may physically trap native gold particles in blasting and milling operations, and while quantitative data in this respect are lacking, loadings of up to 188 g/t fibre have been recorded.[1]

Furthermore, woodfibre may absorb soluble gold values in the leach circuit where gold loadings of 2 g/t in a low grade circuit and 18 g/t fibre in a higher grade circuit have been reported.[2] While many CIP plants are recovering significant quantities of woodfibre, no satisfactory means of treating the stockpiled material has been developed as yet. Assuming a fibre content of 180 g/t for a plant treating 250 000 t/m, this could account for nearly 1 kg of gold per month in a high grade circuit.[3]

Coarse grit

Besides causing sanding-out problems in both the leach and CIP circuits, the progressive build-up of grit in the adsorption circuit may result in poor plant performance. The build-up of grit in the carbon contactors would increase interstage screen duties considerably while also affecting adsorption efficiencies due to the slower movement rate of carbon. In a modelling exercise,[4] the transfer of carbon containing 14 per cent grit was shown to increase the soluble gold loss from 0.0138 to 0,0154 g/t.

Furthermore, the transfer of such contaminated carbon to the elution and regeneration circuits would incur further inefficiencies in these circuits. The presence of silica grit may also aggravate any slag formation in the regeneration kiln and thus reduce effective heat transfer and carbon regeneration.

Coal ash

The presence of coal ash in reclaimed dump material presents several problems in a CIP operation. If present in a leach circuit coal ash may adsorb significant amounts of gold with loadings approaching 80 g/t being recorded. The progressive build-up of coarse coal ash in the contacting circuit may also contribute to poor plant performance. As with coarse grit, an increase in interstage screen duties, poor adsorption efficiencies and increased elution and regeneration costs may be anticipated.

Unlike coarse grit, however, coal ash when present in the elution circuit is not easily separated from the loaded carbon using any sort of elutriation technique. Because of this, a large circulating load of coarse coal ash having a low activity may be returned to the adsorption circuit.

In many instances, a large proportion of the coal ash present in dump material is of a relatively coarse particle size and as such is amenable to recovery by simple screening operations.

Vegetable trash

As may be expected, vegetable trash in the form of grass, roots, twigs and leaves may present problems in any dump treatment operation. Besides the physical problems associated with such material (see woodfibre) significant gold losses caused by the presence of humic acids may be encountered.[5]

Plastic insulating material

The softening and melting of plastic material in the elution columns resulted in many earlier problems when blinding of column collectors occurred. The effective removal of this material by pre-screening has virtually eliminated such problems.

Equipment

Vibrating screens

In early testwork,[3] the Derrick screen — a vibrating screen inclined at 25–40° to the horizontal with a high frequency (3 000 rpm) and small amplitude (0,8 mm) — was shown to be acceptable for the screening of large volumes of pulp at 0,6 mm. As a result of this testwork, this type of screen was incorporated as a fine screening device into most of the initial designs for CIP plants in South Africa.

Due primarily to the limited life of square mesh steel screens, further testwork was carried out on the horizontal Velmet screen using polyurethane decks. These screens capable of duties of 19–60 $m^3/m^2.h$, generally gave better service and in 1985 were in common use in most CIP operations in South Africa.[3]

Many problems were, however, subsequently encountered with these vibrating screens due to such factors as:

- The very short life of metal screens — while polyurethane decks offered a very long service life they suffered the disadvantage of having a relatively small open area.
- The screening of soft woodfibre particles was not efficient due to the reorientation of such particles taking place on the vibrating screen deck.
- Any surges in woodfibre or pulp resulted in overflowing and by-passing of oversize material.
- The build-up of calcium carbonate on screen decks was found to restrict screen capacity.
- Because of vibration and resonance, expensive civil and structural construction is required.

Mintek tank

A further development was the use of the screen designed by the Council for

Carbon-in-Pulp

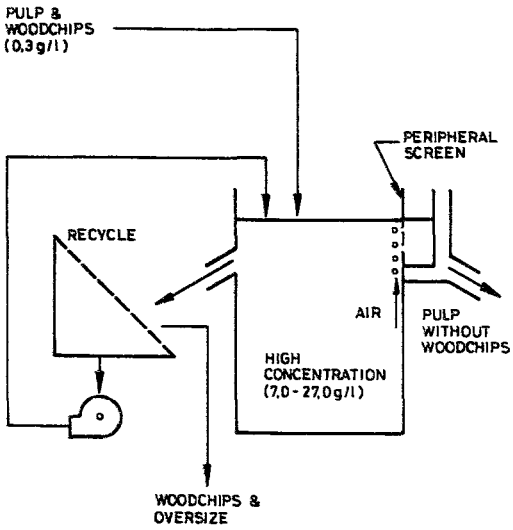


Figure 1
The Mintek tank.[6]

Mineral Technology (Mintek), the Mintek EPAC (equalized-pressure, air-cleaned screen) as a fine-screening device.[3,6] **Figure 1** shows how these screens are utilized in a pre-screening circuit. It was found that the curtain of air bubbles is effective in keeping woodchips and fibres and coarse particles from blinding the vertical 0,6 mm screens. This circuit was used on the Beisa, Afrikander Lease and Doornkop CIP plants.

Some of the advantages seen for this pre-screening procedure are:

- The tank can serve to absorb surges in the quantities of woodchips and in the volume of pulp.
- The volume of pulp to the vibrating screen is relatively small.
- The washing of the high concentration of woodchips on the vibrating screen is more efficient.
- There is more efficient screening by the vertical screens, as is shown by the very low content of oversize in the screened pulp.

AAC — Delkor linear screen

Following laboratory testwork in a joint project between Anglo American Research Laboratories and Delkor Technik (Pty) Ltd,[7,8] a 6 m² prototype Linear Screen was installed at Vaal Reefs Exploration and Mining Company. This unit was commissioned in January, 1985 at the Afrikander Lease CIP plant treating cyclone overflow in the milling circuit with the purpose of removing woodchips and woodfibre from the feed to CIP. Within five minutes of commissioning, the screen, was treating the full pulp flow of ca 600 m³/h at an RD of 1,1–1,2.

Figure 3 illustrates the decrease of woodfibre in the milling circuit from 75 g/m³ to 2 g/m³ over a period of five weeks. This time lag was due to the recirculating load present in the circuit. Due to the high efficiency of the Linear Screen (95–98 per cent woodfibre removal), the existing Velmet vibrating screens which were removing some 53 per cent woodfibre were shut down as they had become superfluous.

A schematic drawing of the screen unit is shown in **Figure 2**. It comprises a coarse monofilament cloth, supported on belt rollers, which is driven by a head pulley coupled to a variable speed drive unit. Slurry is fed into the feed box, passes through holes in a distributor plate and onto the moving monofilament cloth. The undersize material drains through the cloth by gravity and is collected in the underpan. The woodfibre and oversize material is discharged at the drive pulley using a cloth wash spray. The cloth returns via an automatic tracking system, a dynamic tensioner and is presented to the feed in a clean, unblinded condition. This action ensures a continuous, self-cleaning screen. By adjusting the weave configuration of the cloth, the cut point of the unit can be easily varied according to the required operating conditions.

The mechanism for high woodfibre recovery results primarily from three features: the high open area square aperture cloth; the feed distribution that prevents flooding of the cloth; and the passive cloth motion that minimises further agitation of the woodfibre.

Apart from achieving the objective of an efficient machine for the removal of oversize, the Linear Screen proved to have many further advantages:

- The unit operates without vibrating, eliminating the need for expensive civil and structural construction common in conventional mill buildings.
- The noise level is markedly reduced due to the elimination of vibration.
- The power consumption is a small fraction of that required for vibrating screens.

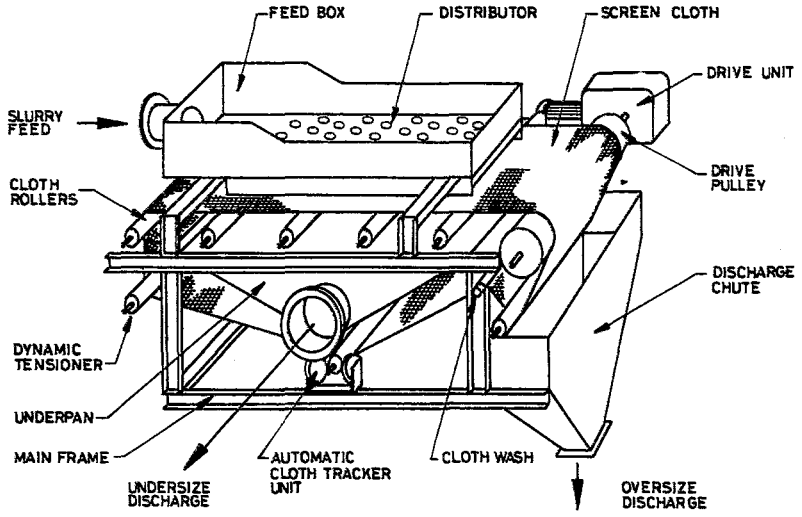


Figure 2
The AAC-Delkor Linear Screen.

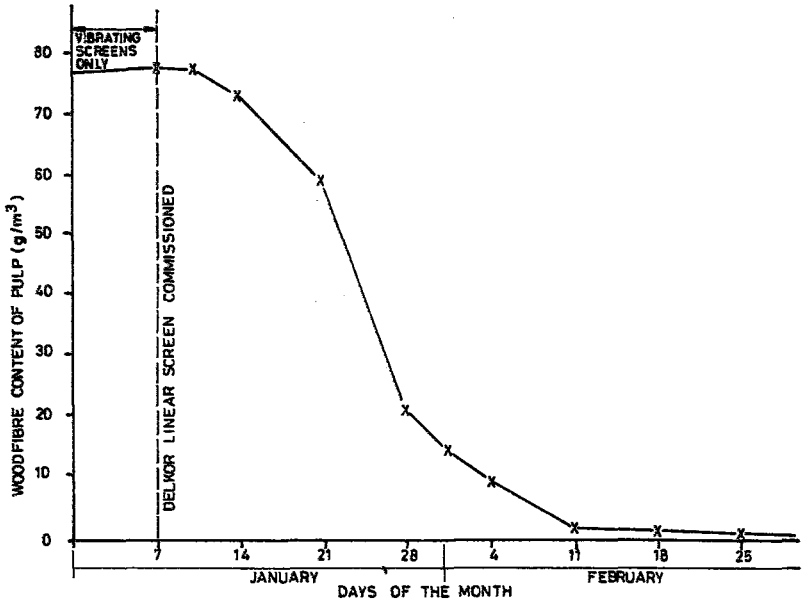


Figure 3
Removal of woodfibre using the Linear Screen.

- Availabilities in excess of 97 per cent have been recorded on all installations.
- Simplicity of operation and open design has shown benefits in very low maintenance costs.
- The unit is supplied with an integral feed box, underpan and discharge chute, reducing the required civil and structural work common for vibrating screen underpans.
- No calcium build-up on screen cloths has been recorded to date — this is felt to be due to the washing and flexing of the cloth.

The development of the Linear Screen has made a major contribution to the CIP process in the area of tramp oversize removal. This is evident in that 59 screens have been ordered by gold plants within 18 months of the first prototype being commissioned. Besides demonstrating the efficient removal of woodfibre, the Linear Screen has virtually eliminated the problems associated with the presence of coarse grit, coal ash and extraneous vegetable trash in the feed to CIP circuits.

Interstage screening

Since August 1980 the Gold and Uranium Division of the Anglo American Corporation (AAC) has commissioned seven carbon-in-pulp (CIP) and two carbon-in-leach (CIL) plants for gold recovery, as listed below:

Plant	*Type*	*kT/m*
President Brand	CIP (Calcine)	30
Afrikander Lease	CIP	60
Simmergo	CIP	180
Vaal Reefs	CIP	140
Vaal Reefs	CIP	250
President Brand	CIP	390
Western Deep Levels	CIP	120
Ergo	CIL	1 800
Daggafontein	CIL (in progress)	1 000

Over this period, considerable attention has been given to the interstage screens used for retaining carbon held in suspension in the adsorption (CIP) or leach/adsorption (CIL) tanks but enabling slurry to pass from stage to stage (normally between four and eight stages). The carbon is advanced on a regular basis, countercurrent to the slurry flow, for final screening and then gold recovery.

One of the keys to successful CIP/CIL plant operations is considered to be this interstage screening operation and significant developments have been made since the first AAC CIP plant was commissioned at President Brand seven years ago.[9] At this time, the Homestake plant, considered to be the first successful CIP plant commissioned in modern times, had eight years of useful experience in interstage screening and utilised external vibrating screens (600 mm × 1 300 mm Derrick screens with 20 mesh woven stainless steel cloth) for this purpose, pulp/carbon being airlifted to the screens on a continuous basis.[10] Carbon oversize was returned to the same adsorption tank, while the slurry undersize was advanced to the next adsorption stage. Carbon movement was effected by transferring carbon oversize by gravity on a regular basis back one stage.

President Brand Calcine CIP plant

At President Brand, peripheral woven wire 20 mesh stainless steel screens in 750 mm × 400 mm frames were installed into the full periphery of the 5,5 m dia × 5,5 m high adsorption tanks. Slurry flowed through the screens into sloping peripheral launders flowing to the next adsorption tank, the carbon granules (1,18 mm to 2,36 mm) being contained at the screen interface — **Figure 4**. The screen surfaces were kept clear by streams of compressed air emanating from fingers along the length of and inside each screen.

Screen duties of 12–15 $m^3/m^2.h$ were required and, after some initial teething problems, were obtained. At the low pulp throughput (80 m^3/h) screen duties of this level were quite feasible, particularly as the calcine grading was some 97 per cent minus 75 micrometre.

Afrikander Lease CIP plant

For significantly higher throughput plants, 150 m^3/h slurry, processing run-of-mine (ROM) ore milled to a comparatively coarser grind (70–75 per cent minus 75 micrometre), peripheral screens would be somewhat overloaded. This disadvantage was tackled for the next plant to be designed at Afrikander Lease by installing transverse rectangular launders across two chords of each adsorption tank with screens on each side of each launder allowing higher screen areas, easier by-passing facilities in the event of tank or screen maintenance and lower inter-tank height differential requirements. Again, compressed air-fingers were used to sweep carbon away from the upstream side of the screen panels. This design proved ineffective in keeping the screens clear as the presence of woodfibre in the feed pulp had become a major problem — at the President Brand Calcine CIP plant it was not a constituent of the feed.

The problem was finally resolved on an industry basis using a design proposed by Mintek with the equalised-pressure air-cleaned (EPAC) screen — **Figure 5** — where the pulp level on each side of the air-swept retaining screen was maintained as near equal as possible thus reducing significantly the differential pressure which caused the carbon and woodfibre to choke the screen surface.[11,12] EPAC units were subsequently installed at Afrikander Lease and relieved the problem with screen duties of ca 50 $m^3/m^2.h$ being obtained. Similar major problems on other CIP plants were also resolved.

There is no doubt that the EPAC concept was the first major breakthrough in interstage screening, not only in the SA gold mining industry, but also overseas. Local derivatives of the EPAC screen developed, each improving performance for particular plant requirements, but all used the EPAC concept.

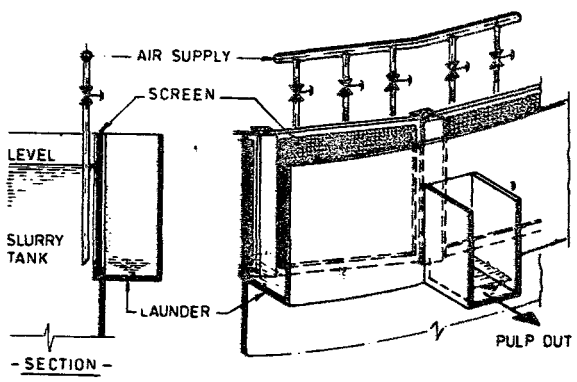


Figure 4
President Brand Calcine plant screens.

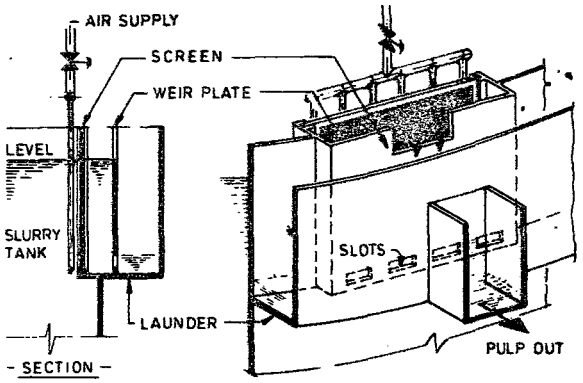


Figure 5
EPAC screen with weir plate and sand relief slots.

Simmergo CIP plant

The next AAC CIP plant to be commissioned was at Simmergo in January 1983 where standard EPAC screens were installed using a weir plate on the downstream side of the screen with sand gates cut into the bottom edges of the weir plates. Again, air fingers were used on the upstream side of the screen. By this time a design air flow rate of 1 m^3 air at NTP per m^2 screen area per hour had been formulated using air at a pressure of 250 kPa. However, screen operations proved erratic and extremely sensitive to variations in plant throughput, which were frequent. Choked screens led to tank overflows, carbon by-passing, flat carbon-loading profiles and poor adsorption efficiencies. These occurred because as plant volume throughput decreased the differential pressure across the screens increased resulting in carbon pinning to the screens. The resultant domino effect in the downstream tanks caused chaos. This led to the development by E L Bateman of the Lap Screen used at Doornkop where air-fingers were used on both sides of the screen in order to minimise any such pressure differential and thus all pulp overflowed over the downstream weir into the launder where pulp was transferred to the next adsorption stage—**Figure 6.**

The problem was partially resolved at Simmergo by installing an air sparge deflector plate above the air-fingers resulting in a curtain of air sweeping the upstream side of the screen rather than a series of air bubble streams—**Figure 7.**

Vaal Reefs CIP plants

In October 1984 and April 1985, two further CIP plants were commissioned at Vaal Reefs at the No 9 shaft (new grass roots plant) and No 8 shaft (filter plant replacement) plants respectively.

A new screen design, designated the Vaal Reefs design, was introduced to improve on some of the weaknesses in the original EPAC system consisting of nests of narrow box-like screens with screen cloth on each side of the box—**Figure 8.** The pulp flowed through the large submerged screen areas into the inside of the narrow box and over a weir across the narrow open end of the box into a peripheral tank launder. Air sparges between each box cleaned two screen faces. The nests of screen boxes were installed perpendicular to the tank circumference thus allowing significantly increased screen area than obtainable with peripheral or EPAC-type screens. Design throughput for each plant was easily obtained since with this increased screen area, the requirement was only of the order of 20 $m^3/m^2.h$.

Thus over some five years, considerable progress had been made in the design and operation of interstage screens by following two quite different routes. The first was the EPAC route where throughput rates of up to 120 $m^3/m^2.h$ could be achieved on an intermittent basis and about 50 per cent of this flow on an average basis and the second via the screen nest principle where significantly increased screen areas could be used lowering unit throughput.

The EPAC design still had its weaknesses—loss of screen area due to variations in pulp flow resulting from upstream operating problems while screen cloth life was low and air consumption quite high. The screen nests also required extensive labour to remove faulty units and replace with rehabilitated units.

Ergo CIL plant

With the advent of high tonnage low grade retreatment operations, the 1,8 million ton/month Ergo CIL plant being the first, the use of conventional EPAC screens for interstage screening was clearly impossible since the tank periphery was far too small, even on 13 m diameter tanks.

The Ergo project was approved in July 1983 and commissioned in April 1985. The time constraints on the project made it necessary to install the Vaal Reefs modular screen systems on a much larger scale. At first, problems related to relatively small design deficiencies in

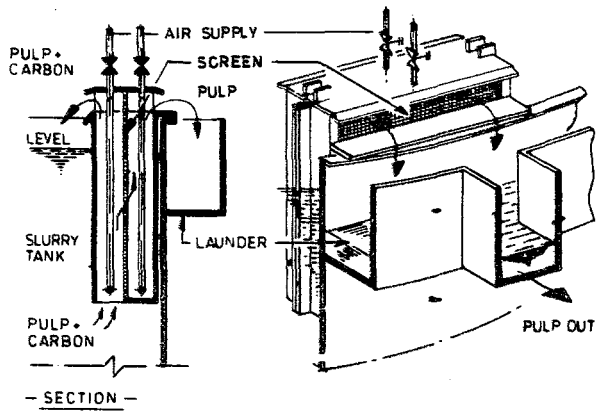


Figure 6
Bateman Lap Screen.

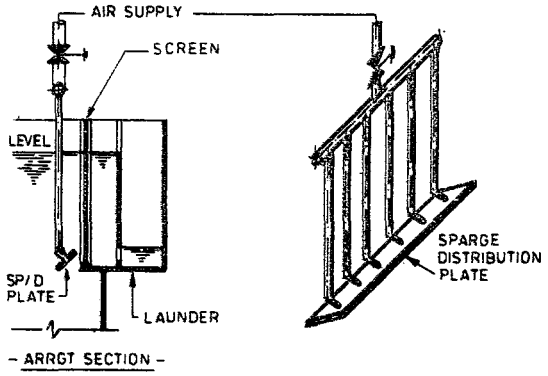


Figure 7
EPAC screen with air sparge distribution plate.

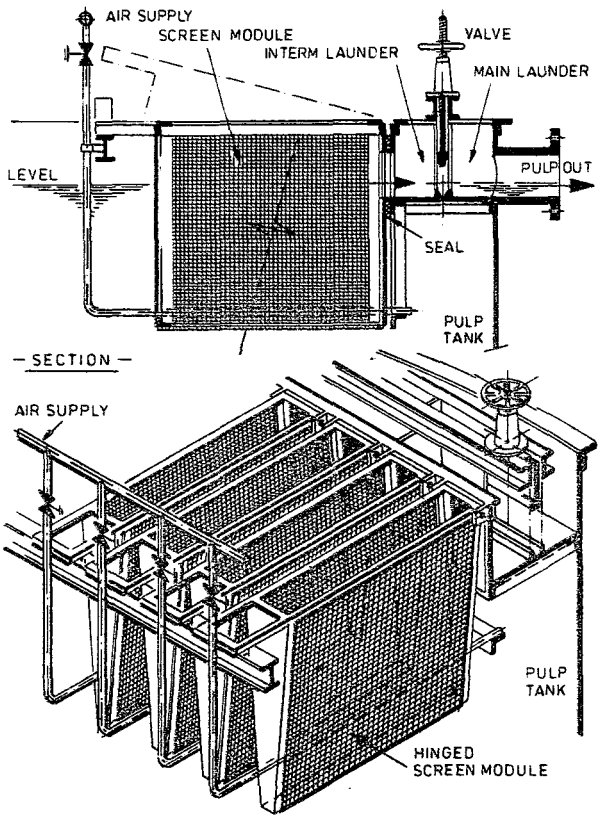


Figure 8
Vaal Reefs Nest Screen.

screen cloth tensioning and air distribution were experienced. In all, 1600 screening elements were involved and the subsequent modification programme took four months of intensive work to complete.

Screen life has improved from approximately two weeks to 25 weeks and the early problems of carbon by-passing have now been resolved. Plant performance is now satisfactory but this experience re-inforced the need for the development of an alternative high capacity low maintenance screening system.

New President Brand CIP plant and design work at Simmergo

The next plant to be designed was the 390 000 tons/month new gold plant for President Brand.[13] Not long before design had been frozen, the selected system was peripheral EPAC screens with air sparging for screen cleaning and air-lifting pulp from the compartment between the screen and the overflow weir. Then came the next major advance in interstage screening.

On a visit to the Kambalda plant in Western Australia[14] the project metallurgist noted a mechanically swept screen that seemed the answer to all the problems inherent in the EPAC and Vaal Reefs designs. The screen consisted of a fixed circular basket in the centre of the adsorption tank built around the central mixer shaft—**Figure 9**. The screen bottom consisted of panels of wedgewire with the flat side of the wedge facing down. The pulp flowed up through the wedgewire slots into the basket and flowed out by gravity along a launder leading to the next tank. Wiper blades constructed from angle iron were attached to the mixer shaft and rotated about 25 mm below the wedgewire and, with a pulsating action, effectively cleared carbon away from the screen interface. No air was used at all effecting a significant saving.

With full co-operation from Western Mining Corporation a 3 m diameter prototype was installed in one adsorption tank at Simmergo and proved to be an instant success with throughputs of 70–80 $m^3/m^2.h$ being achieved on a continuous basis. This throughput rate being slightly under plant throughput, a second screen was installed 4 m in diameter which reduced throughput rate to approx 40 $m^3/m^2.h$ and which easily handled the fluctuating pulp throughput.

Design improvements to the Mark II included:

- Increasing the size of the pulp exit pipes/launders to reduce back pressure on the screen.
- Installing a drop box at the point where the slurry leaves the screen surface and enters the pipes/launders, again to reduce back pressure on the screen and to reduce any damping effect on the pulsating action of the wiper blades.
- Adding an adsorption tank overflow to make it impossible for unscreened slurry to overflow into the screen basket, thereby eliminating carbon by-passing and also stopping coarse carbon particles becoming trapped in the "vees" of the wedgewire and causing chokes.

An important aspect of the design is the sealing of the wedgewire panels to the screen basket steelwork. At Simmergo, closed cell expanded polyethylene insertion was used and proved most effective. The distance between the wiper blades and the screen underside was found to be critical and the 25 mm used at Kambalda was found to be optimum at Simmergo. Any changes to this reduced throughput significantly. The wedgewire used was a ferritic stainless steel with an 0,6 mm gap and a 28 sb profile and panel lives of over a year have been achieved.

There is a limitation to the size of screen that can be installed in an adsorption tank or, more correctly to the size of wiper blade that can be used. This is because the power draw of an impeller (wiper blade) is proportional to the fifth power of the diameter, thus doubling the wiper blade diameter would increase the power draught by a factor of 32 as against a factor of 7,6 for a 50 per cent increase in wiper diameter.

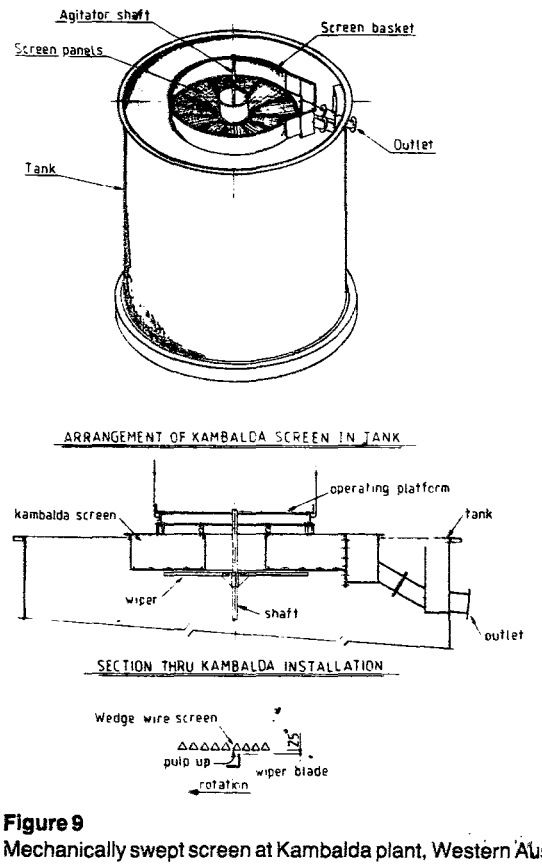


Figure 9
Mechanically swept screen at Kambalda plant, Western Australia.

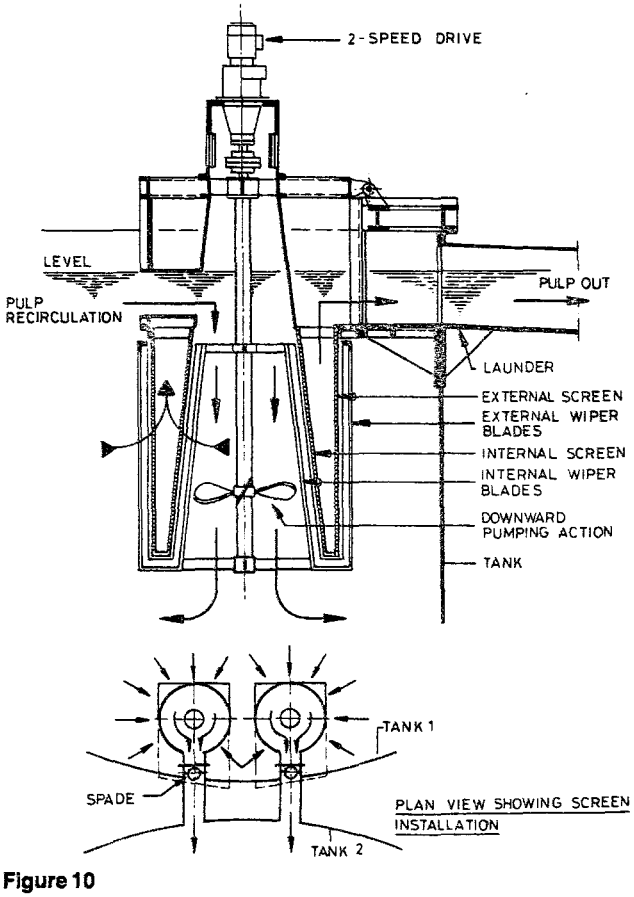


Figure 10
Modified NKM screen at Daggafontein plant.

The working cost cross-over point where the EPAC installation using low pressure air would be more economic than a Kambalda screen is approximately a four metre diameter screen.

The Kambalda screen proved so effective at Simmergo, where all adsorption tanks were converted, that the decision to change the design for the new President Brand plant was easy to take. Here the throughput through each of three lines of adsorption tanks was 130 000 tons/month or 200 tons/h at 90 per cent on-line. This was equivalent to 260 m^3/h slurry at 50 per cent solids and, for the 3,5 m dia screen basket, (6,6 m^2 available), a throughput rate of 40 m^3/m^2.h. In all, Kambalda screens were installed in 21 adsorption tanks (three lines of seven tanks each) and since the plant was commissioned in May 1986, operation of the interstage screens (0,5 mm wedgwire instead of square woven mesh cloth) has been most satisfactory. Carbon losses are considered low at about 25 g/t.

As a result of the success at Simmergo, the two Vaal Reefs plants are currently converting from EPAC nests to Kambalda screens. A design change has been included such that slurry from the screen flows into a peripheral drop box launder mounted around the inner edge of the Kambalda basket. This enables more rapid removal of slurry from the screen.

The Kambalda screen has also proved versatile in that it can be adapted successfully to operate in adsorption tanks utilising draft tubes in addition to the conventional mechanically agitated tank.

Western Deep Levels CIP plant

In May 1986, the new 120 000 tons/month Western Deep Levels No 1 shaft plant was successfully commissioned using the Kambalda design. Reinforcing the basket screen framework and increasing the number of wiper blades resolved the post-commissioning problem of low screen throughputs.

Daggafontein CIP plant

In January 1984, approval was given to proceed with the installation of a one million ton/month slimes retreatment CIL plant at Daggafontein for commissioning in 1987. Despite the success of the Kambalda screen on smaller tonnage plants, scale-up to a calculated 6,3 m diameter screen indicated that the power drawn by the wiper blades alone would be 30 kW, some 25 per cent higher than that required for the full tank agitator!

Considerable thought was given to ways and means of increasing significantly pulp flow through screen cloth and a small mechanically cleaned cylindrical (as opposed to the round flat Kambalda) screen was built and tested at the Anglo American Research Laboratories. The cylinder was 200 mm diameter and the submerged area 0,23 m^2 and this proved to be the next breakthrough in interstage screen development, since very high flowrates of 250 m^3/m^2.h were obtained. It was subsequently found that a screen of similar design was being used by North Kalgurli Mines (NKM) in Australia. Prior to continuing with the design and testwork an agreement was reached with NKM regarding the use and further development of the NKM screen.

Since time was of the essence, a full-scale model (scale-up factor in excess of 40) was designed and plant-scale testwork carried out at the Ergo CIL plant using one of the 2 000 m^3 contacting tanks. Design changes were made over a period of several months leading to the design selected for the Daggafontein plant shown in **Figure 10**.

Basically, the screen is in the form of a basket with wiper blades rotating around each screen surface. Pulp passes through both sides of the screen basket with carbon being retained to a small degree on the screen surfaces but dislodged due to the pulsating effect of the wipers thus enabling relatively unrestricted pulp flow through the screen. Pulp passes through the screen on to the next absorption tank, a spade in the transfer launder enabling a suitable back pressure which, unlike the Kambalda system, stabilises screen operation. A downthrust impeller is attached to the unit's central drive shaft which circulates pulp through the unit.

Currently the Daggafontein plant is in the early commissioning stages. Two NKM units are installed in each leach/adsorption tank, the screen diameter being 1 600 mm and height 1 800 mm to handle 1 200 m^3/h of slurry. The wiper blade shaft can rotate at 17 or 25 rpm and each unit has eight wiper blades, four internal and four external to the screen faces. The units will normally operate at the lower speed (power 3,5 kW) and, in the event of one being out of commission, one unit should temporarily handle the full flow (power 10 kW).

This last development is the culmination of years of investigation since CIP was first introduced to AAC gold plants. Slowly but surely the problems related to interstage screening have been reduced to a level where maintenance is no more intensive than that of normal plant equipment. The beneficial effect of the improved screen design on the metallurgical efficiency of AAC plants has been most noteworthy.

Acknowledgements

Acknowledgements are gratefully made to the managements and senior metallurgical staff from Vaal Reefs Exploration and Mining Company Limited, Afrikander Lease Limited, President Brand Gold Mining Company Limited, East Rand Gold and Uranium Company Limited, Western Deep Levels Limited, Anglo American Research Laboratories, the Gold and Uranium Division of Anglo American Corporation of South Africa Limited, the Council for Mineral Technology and Delkor Technik (Pty) Limited without whose support such significant improvements to CIP plant operations would not have been possible. Mention must also be made of the contributions made by both Western Mining Corporation at their Kambalda plant and North Kalgurli Mines towards the development of reliable interstage screening systems.

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Society of Chemical Industry

HYDROMETALLURGY '87

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INTRODUCTION

In August 1980, calcine from the roasters at the President Brand gold mine, situated near Welkom in the Orange Free State goldfields, was first introduced into the adsorption tanks of the newly constructed carbon in pulp (CIP) plant [1]. This event marked the first commitment by the Anglo American Corporation (AAC) to commercial application of CIP technology within South Africa, and was the culmination of a concerted research effort by the Anglo American Research Laboratories (AARL) and the government-sponsored Council for Mineral Technology (MINTEK) to introduce the benefits of in-pulp processing to the group's gold mining activities.

Despite several early setbacks, the calcine CIP plant at President Brand proved to be a major success, and set the pattern for a radical technology swing by the group towards carbon-based recovery methods. A listing of major gold recovery projects undertaken by AAC (Table 1) shows that, seven years on from commissioning at President Brand, approximately 87 tons of gold per annum will be recovered through the medium of CIP and, perhaps more significantly, this will be derived from the processing of over 4×10^6 tons of solids per month.

The pattern of early experience with the CIP process in South Africa has been well documented by others [2,3]. It is clear from these descriptions that a number of fundamental problems have attended the growth phase of the technology, and perhaps the greatest difficulties have arisen in the area of pulp screening – both preliminary fine screening to remove oversize material and interstage screening of activated carbon. Inadequacies in these key process areas greatly elevated the operational complexity of early CIP plants, and in many cases contributed to generally 'lower than expected' circuit performance, particularly during the first year of plant operation.

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The focus of the present paper is to highlight the importance of the interstage screening function in the CIP process, and to describe developments which have occurred within the gold plants of the AAC. The decision to embark upon a programme of screen development was motivated by a general dissatisfaction with the performance of early screen designs, and a parallel need to improve specific throughputs to satisfy the demands of new high-tonnage, dump reclamation projects at the East Rand Gold and Uranium Company (ERGO) and Daggafontein.

Key objectives in the search for an improved screen design were as follows:

(1) High unit capacities to reduce the number of individual screens.
(2) Reliable and effective screen performance, to minimize back-mixing of carbon and to maintain circuit driving forces.
(3) Robust construction to minimize operating and maintenance labour requirements.
(4) Capital and operating cost economies.

It will be shown that many of the screen designs used on AAC operating plants fall well short of these criteria, but through an evolutionary process involving collaboration with other mining groups, and in-house design initiatives, a screening concept has emerged which satisfies most of the objectives listed above, and is now being applied to the design of all new CIP plants with the group.

KEY FUNCTION OF INTERSTAGE SCREENING

The existence of a dynamic equilibrium between gold in solution and gold adsorbed on activated carbon dictates the use of countercurrent, multistage contacting of phases in the CIP process, to enhance gold loading on the carbon, and to achieve high overall gold recoveries. Phase separation relies upon the differing sizes of the solid particles in the pulp and the carbon, and typically this separation is effected by means of interstage screens, which may be of the external or internal tank type.

The importance of the interstage screening operation cannot be overestimated, since it not only provides the basis for efficient gold recovery, but also has a major cost implication, through its influence on activated carbon losses, and on 'sunk' costs associated with circuit carbon and gold inventories. To illustrate how screen 'inefficiency' can impair circuit operating performance, consider the following example, which uses a dynamic simulation technique based on the rate expression for carbon adsorption proposed by Menne [4] to examine the effect of carbon 'backmixing', corresponding to degrees of cocurrent, rather than countercurrent, carbon movement. This could arise as a result of a mismatch between carbon and screen aperture sizes, or through screen maloperation resulting in carbon leakage through, or around, the screen elements.

The rate expression used is as follows:

$$R = K_1C(y^* - y) - K_2y$$

where K_1, K_2 and y^* are values determined by experimental techniques and are particular to the system being investigated.

Other factors relevant to the example quoted are as follows:

Gold in feed solution	1,9 g Au m^{-3}
Number of stages	5
Solution hold-up per stage	300 m^3
Solution flowrate	300 m^3h^{-1}
Carbon hold-up per stage	12 tons
Fraction of carbon moved in 24 h	0,5
Gold on recycled carbon	nil

Two cases were investigated, corresponding to carbon 'leakage' rates of 10% and 20% of the net forward movement of carbon, and the influence of these screening inefficiencies on gold lock-up in carbon, and on gold soluble loss, is shown in Figure 1.

The effect of cocurrent carbon movement is to reduce the overall driving force for adsorption in the circuit, thereby raising both solution and carbon gold concentrations. The dynamic simulation predicts that a 10% leakage will result in a 21% increase in the terminal solution gold concentration, and a 24% increase in gold lock-up in the circuit; increasing the degree of back-mixing to 20% raises these values to 48% and 69% respectively.

As may be seen, the effect of carbon back-mixing is highly significant, and if one considers that, in the example quoted, a 10% leakage equates to a steady state carbon concentration in the interstage pulp of only 60 g m^{-3}, then it is clear that on many operating plants, particularly during the early phases of commissioning, such circumstances have arisen, and predictably have resulted in higher than expected soluble gold losses.

Although in essence the function of the interstage screen is to separate a relatively coarse granular carbon from a pulp containing finely divided solids, in practice the 'system' contains a variety of other species, which greatly compound the difficulties experienced in making the required separation.

The following are typical of the problems which have been experienced in operating plants:

(i) Partial blinding of the screening medium with 'near size' material, which may be grit resulting from inefficiencies in the pre-screening function, or the products of carbon attrition generated within the tanks themselves.
(ii) Screen blinding with fine woodchips, grass or plastic, which occur in the feed pulp as a normal consequence of mining or dump reclamation operations, and which have a tendency to pass through pre-screening owing to their acicular shape.
(iii) blinding of the screen due to the build-up of a calcium carbonate scale or, alternatively, a simple 'caking' on the screen medium.

It will be apparent from the foregoing comments that the designer faces a number of problems in attempting to quantify the dynamics of screen operation for an operating plant, and it is for this reason that most of the

developments described in the following section have resulted from 'in-plant' rather than laboratory or pilot scale testwork.

HISTORICAL PERSPECTIVE

The development of interstage screens in the AAC group may be broadly classified into two categories, namely:

(i) *Low tonnage – high grade primary treatment operations*
These plants are characterized by treatment rates of approximately 5×10^6 tons per annum or less and recoverable gold grades of 1 g ton^{-1} up to 10 g ton^{-1}.

(ii) *High tonnage – low grade retreatment operations*
These plants are characterized by treatment rates of more than 5×10^6 tons per annum and recoverable gold grades of less than 1 g ton^{-1}.

Within the group, ten plants of the first type have been commissioned and are in production, while one plant of the second type has been commissioned and a second, at the time of writing, is being commissioned.

The problems associated with interstage screening are, in general, magnified as the plant design throughput is increased, and it will be shown that acceptable solutions to screening operations in lower tonnage plants have often not been as successful when applied to larger tonnage plants.

At the design stage of the President Brand calcine plant prior to 1980, the only other commercial CIP plant in the world was the plant at Homestake commissioned in 1972. The latter utilised external vibrating screens (0,6 m × 1,3 m Derrick screens with 20 mesh woven stainless steel wire) whereby the forward moving slurry was airlifted onto the screens prior to flowing to the next adsorption stage, the carbon being returned as screen oversize to the tank of origin. Countercurrent movement of carbon between stages was achieved on a batch basis by merely redirecting external interstage screen oversize for a specified period each day.

Although consideration was given to the use of external screens during the design of the President Brand calcine plant, there was general concern over the carbon handling aspects of this approach. This concern, coupled with reported low screen lives, and high capital installation costs, influenced the decision towards the use of internal, air-swept screens, which were being evaluated at that time by MINTEK and others in South Africa (Fig. 2(a)) [5]. The screens, after some initial difficulties, operated quite satisfactorily, achieving a duty of between 12 and 15 m^3 m^{-2} h^{-1} which was low, but adequate for the application.

At the time of commissioning the above plant, the design for a second CIP plant with a throughput of 60 000 tons per month for Afrikander Lease was in progress. The design included 5,1 m diameter draft tube circulator tanks with a height to diameter ratio of approximately 2:1. Thus, approximately the same tank periphery as the President Brand calcine plant was available for treating double the flowrate of a slurry containing much coarser

particle sizes (75% – 75 μm). A system of transverse rectangular launders across two chords of the tank was adopted with air-swept woven wire screen panels incorporated into the sides of the launders. Perceived advantages were a lower intertank height differential, a greater submerged screen area, and an improved tank by-pass arrangement. This arrangement proved totally unsatisfactory, resulting in frequent blockages caused by carbon and woodfibre pinned to the screen. Attempted solutions at the time included the installation of electric vibrators on the screen panels, and the use of wedgewire to avoid the severe erosion of the screen cloth caused by localized action of the air spargers. Neither action manifestly improved the situation.

The plant at Afrikander Lease was the first of the AAC plants to demonstrate the major detrimental effect that woodfibre contaminants, derived from underground timber supports, had on the process of interstage screening. It was also at Afrikander Lease, four 'difficult' years later in January 1985, that AAC was eventually to achieve a radical breakthrough in solving the woodfibre problem, with the successful introduction of a prototype 6 m^2 linear screen [6], the end product of a joint project between AARL and Delkor Technik (Pty.) Ltd. The linear screen has since become a standard feature of AAC 'new generation' CIP plants, and has greatly eased the problems of interstage screening.

At the same time that Afrikander Lease was experiencing interstage screening problems, the CIP plant at Western Areas operated by the Johannesburg consolidated Investment Company was facing problems of a very similar nature. MINTEK were requested to provide 'trouble shooting' support to the plant, and it was at this stage that Laxen advanced the concept of the equalized-pressure air-cleaned (EPAC) screen (Fig. 2(b)) [7]. By introducing a backing plate, or weir, downstream of the screen element it was possible to equate pulp levels on both sides of the screen, thus substantially reducing the differential pressure which serves to pin the carbon to the screen, and increasing the area of screen available for pulp flow. Initial designs incorporated slots or holes at the base of the metal plate for 'sand relief' to prevent silting out between the screen and the weir plate.

The EPAC screen was introduced at Afrikander Lease resulting in a considerable improvement in adsorption performance and efficiency. Specific screen throughputs were increased substantially, with duties of 50 m^3 m^{-2} h^{-1} consistently achieved. The EPAC concept represented a major breakthrough in interstage screening, and the principle survives in a wide variety of proprietary forms used throughout the South African gold mining industry, e.g. Forbes, Kliptek, Kinross, Grootvlei, Vaal Reefs. Beatrix and LAP screens [5].

The next plant to be constructed by AAC was Simmergo which was commissioned in January 1983. Standard EPAC screens were installed and duties of up to 90 m^3 m^{-2} h^{-1} were eventually achieved. However, operation of the screens was erratic and sensitive to variations in plant throughput. Sudden decreases in pulp flowrate allowed the pulp between the screen interface and the weir plate to drain through the sand relief slots, causing the carbon to be pinned against the screen. The consequence of a blockage in one tank

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was to cause blockages in all the downstream tanks in a domino sequence. Elsewhere in the industry the weir plate slots were eliminated completely and air spargers were used to lift coarse sands over the weir plates. This is a feature of the Bateman lap screen used successfully at the Western Areas gold plant [8].

In the AAC group, the commissioning of two CIP plants at Vaal Reefs in October 1984 and April 1985 prompted the development of the 'Vaal Reefs' screen baskets. These consisted of 'nests' or clusters of narrow, box-like screens with screen cloth on both sides of the box. The pulp flows through the large submerged screen areas into the inside of the narrow box and over a weir into a peripheral tank launder. Air spargers between each screen box serve to clean adjacent screen faces. The 'nests' of narrow boxes were installed perpendicularly to the circumference of the tank, resulting in considerably more available screen area than that attainable with standard peripheral EPAC screens. Having this greater screen area, screen throughputs of only 15 to 20 $m^3\ m^{-2}\ h^{-1}$ were required, which were easily attained with this screen configuration. The Vaal Reefs system also facilitated easier maintenance of the screens, allowing a complete 'nest' of screens to be removed followed by immediate replacement with a refurbished set.

A further development occurred at Simmergo where a deflector plate was installed along the full length of the lower upstream edge of the screen. Instead of introducing sparge air via fingers, the air was directed to the underside of the deflector plate, resulting in a curtain of air sweeping the face of the screen and providing a more efficient cleaning action. This principle is used to good effect in the Davy McKee contractor, which employs a series of deflector plates in a louvre arrangement to distribute the air over the full face of the screen [9].

Notwithstanding the major improvements which had been made in the design of the air-agitated interstage screen, operating plants within AAC continued to experience problems with the use of this type of screen. The reasons were primarily the following:

(i) Screens oversensitive to variations in pulp flow, resulting in frequent screen blinding, because of loss of the EPAC condition.
(ii) Poor screen life owing to localized erosion of the screen cloth. Lives ranged from a few days to five months, and cloth replacement was costly and time consuming.
(iii) The proliferation of individual screens, each requiring effective sealing arrangements, and air distribution systems, required constant operator vigilance to prevent mishaps.
(iv) Air consumption was high. Figures quoted in the South African gold industry range from 0,6 m^3 air m^{-3} pulp at 80 kPa, to 1,7 m^3 air m^{-3} pulp at 500 kPa, depending upon screen design and process application.

In 1984, during the design phase of the 390 000 ton per month primary ore treatment plant at President Brand, a re-appraisal of screening possibilities was made, following a technical visit to a number of small gold plants in Western Australia.

EMERGENCE OF THE MECHANICALLY AGITATED SCREEN

A mechanically agitated interstage screen was seen on a visit to Kambalda in Western Australia [10], and this appeared to operate extremely well. The screen consists of a fixed circular basket in the centre of the tank sited around the central mixer shaft (Fig. 3) with the bottom of the screen made up of panels of wedgewire with the flat side facing down. The pump flows up through the wedgewire slots into the basket, and out by gravity through pipes leading from the basket to the next tank. Wiper blades constructed from 70 mm × 70 mm angle iron are attached to the central mixer shaft and rotate about 25 mm below the wedgewire, causing a pulsing action which clears carbon away from the screen surface.

In collaboration with Western Mining Corporation of Australia, AAC installed a 3 m diameter prototype in one adsorption tank at Simmergo. The unit was an immediate success, although marginally small for the required plant throughput. Screen duty was between 70 to 80 $m^3\ m^{-2}\ h^{-1}$. It is apparent that, for the Kambalda screen, the available screen area increases with the square of the diameter giving it a definite advantage over the normal peripheral screen when scaling up to higher plant throughputs.

A second 4 m diameter test unit was installed at Simmergo with a design screen duty of 40 $m^3\ m^{-2}\ h^{-1}$ and incorporating several improvements as a result of the trials with the prototype. These improvements were:

(i) Installation of an exit drop box, and enlarged discharge pipes to minimize the depth of slurry in the screen basket and to accentuate the pulsing action from the wiper blades.
(ii) Provision of a tank overflow to prevent slurry containing carbon overflowing into the screen basket and causing screen blockage.

The modified Kambalda screens at Simmergo have operated satisfactorily for over 18 months, and are well accepted by plant operating staff. The wedgewire used is a ferritic stainless steel with a 0,6 mm gap and 22% open area, and this has a lifetime in excess of a year. About once a week the tank level is lowered and the screen is brushed clean of any solids trapped in the slots.

The additional power required for the wiper blades must be provided by the main agitator shaft and, for the configuration at Simmergo, this additional power amounts to 8 kW, corresponding to an impeller power number of about 0,21. As will be shown later, the question of power draw becomes an important consideration for the Kambalda screen as the scale of operation is increased, but certainly for pulp flowrates of up to 600 $m^3\ h^{-1}$ the Kambalda screen is considered to offer a reliable and cost-effective alternative to the use of air-swept screens. The success of the Kambalda screen is considered to be due to the combination of induced vertical pulp flow, which gives rise to the 'pulse' which clear the screen of carbon, and a radial component which serves to 'pump' the carbon away from the screen interface. Because of the success of the screen testing at Simmergo, the entire plant was converted to Kambalda screens during 1986.

The tests also provided valuable screen design data for the new gold plant

at President Brand where Kambalda screens (3,5 m diameter) with 0,5 mm wedgewire were installed in 21 adsorption contactors arranged in three streams of seven contactors each [11]. This plant was commissioned in May 1986 and the operation of the interstage screen has been most satisfactory, with design adsorption throughputs and efficiencies being achieved within the first few weeks of operation. Fears of high carbon losses when using wedgewire instead of square woven mesh have proved unfounded, with carbon losses averaging about 25 g ton $^{-1}$.

As a further result of the success of the Simmergo testwork, the new Western Deep Levels plant commissioned in May 1986 treating 120 000 tons per month was designed with Kambalda screens, and both the No. 8 and the No. 9 shaft CIP plants at Vaal Reefs have decided to convert from their original screen 'baskets' to Kambalda screens.

ERGO CIL PLANT

ERGO recovers gold, uranium and sulphuric acid by the large scale retreatment of slimes dams in the eastern portion of the Witwatersrand goldfields. In July 1983, following a successful piloting campaign, a decision was taken to proceed with a 1,8 $\times$ 10^6 ton month^{-1} CIL plant, to recover additional gold from flotation tailings. The scale of the ERGO operation added a new dimension to the interstage screening problem, placing a heavy demand on screen flow capacity, as well as screening efficiency. Calculations showed that none of the interstage screens developed at the time and installed around the periphery of the tank could provide adequate screening capacity.

Time constraints on the project precluded innovative development and testing of any new screen design, and there was little alternative other than to employ the modular screen pack concept used on a much smaller scale at Vaal Reefs.

The ERGO CIL plant achieved design throughput in April 1985 shortly after commissioning, but from the outset problems were experienced with the interstage screens. Although derived mainly from relatively small design deficiencies related to cloth tensioning and air distribution, the need to effect these modifications to a total of over 1600 individual screening elements constituted an operating problem of major proportions. An intensive campaign of screen modification, mounted over a four month period, improved the average life of screen cloths from two to 15 weeks, which at present offers an acceptable condition for plant operation. The by-passing of carbon due to screening problems in the early stages of commissioning had a major detrimental effect on the efficiency of the plant. With progressive improvement in screen condition, there has been a slow but steady recovery of plant performance, but the experience at ERGO reinforced the need for development of an alternative high-capacity screen which could be relied upon to perform effectively from the outset of commissioning.

In January 1984, a decision was taken to proceed with a 1 $\times$ 10^6 ton month^{-1} slimes retreatment project at Daggafontein in the Far East Rand. The developments leading up to a final specification of screens for the Daggafontein plant are discussed below.

Review of alternatives

The success of the Kambalda screens tested at Simmergo had shown the major benefits arising from the use of a single-unit, mechanically cleaned screen. However, a quick calculation shoed that in order to process the design flow of pulp from one of the twin Daggafontein screens (1200 m^3 h^{-1}), the required Kambalda screen would need to be 6,3 m diameter, and since the rotational speed of the main agitator shaft is dicated by the smaller main tank impeller, the tip speed of the wiper blades would be 4,7 m s^{-1}, which would be excessive for an application of this type. Further, since impeller power draw is proportional to the fifth power of the diameter, the power requirement was calculated to be about 30 kW, which is 25% higher than that of the main tank agitator. Clearly this constituted an irrational design proposal.

Although a number of alternative configurations with the Kambalda screen were considered, notably separating the screen drive from the main agitator drive, the resulting designs were not considered appealing. The conclusion stemming from these studies was that whilst, in essence, mechanical agitation may be attractive, its implementation in the form of the Kambalda screen was not appropriate for the Daggafontein plant.

Laboratory testing of cylindrical screen

Leaning heavily on the principles of Kambalda screen operation, a mechanically cleaned cylindrical screen was proposed and tested at AARL. Using a 200 mm diameter screen with submerged area of 0,23 m^2, very high specific flows (200 – 300 m^3 m^2 h^{-1}), were obtained, which offered the first real possibility of developing a single screen capable of treating the full flow at the Daggafontein plant. It was subsequently discovered that a screen embodying similar principles was being used by North Kalgurli Mines (NKM), Australia [12], and therefore prior to continuing with the testwork, discussions were held and an agreement was reached between AAC and NKM regarding the use and further development of the 'NKM screen', as it is known.

Project design and construction deadlines precluded a pilot phase, and it was decided that an extrapolation would be made directly from the AARL laboratory screen to a 'prototype' suitable for plant scale testwork at the ERGO CIL plant. In order to achieve this major scale-up (scale factor in excess of 40 based on screen area), it was evident that an interpretation had to be put on the mechanism of the screening process, to permit a broad design specification for the plant prototype. Accordingly, a conceptual screening 'model' was postulated, which was used together with the AARL data to specify the design of the test unit.

The screening model

The basis of the screening process is the separation of granular carbon from a pulp using a cylindrical screening element which serves to retain the oversize carbon on the screen media, but allows the associated pulp to pass into the interior of the cylinder, and then on to the next contactor. Periodically, a blade rotating about a central axis passes in close proximity to the external surface of the screen, and the effect of the fluid forces generated by the blade is to cause the carbon particles to be dislodged from the screen and to be returned to the contents of the tank. The process is shown pictorially in Figure 4, which implies the build-up of a closely packed layer of carbon particles on the screen from one to three particle diameters thick. Calculations show that this is the situation which arises in practice when the screen is operating efficiently.

It is postulated that the process is analogous to filtration with an incompressible cake and a constant pressure differential resulting from the difference in slurry head between the inside and outside of the screen. The screen therefore functions as a multiplicity of individual 'elements', which are defined at any one time by the separation distance between successive blades. A characteristic of constant pressure filtration is the reduction in volume flow with time as the resistance to flow increases. This is shown graphically in Figure 4, which also presumes a short flow reversal or 'pulse', caused by the negative pressure gradient generated in the wake of the moving blade.

The net effect of the screening operation is that pulp will continue to flow through the screen, provided that the flowrate does not exceed the 'capacity' of the screen to re-distribute the associated carbon back into the tank.

The two essential features of the screen design are therefore:

(i) Provision of a periodic force capable of dislodging carbon particles retained on the screen.
(ii) Application of this force at a frequency which prevents the build-up of carbon to a point where the requirements of flow cannot be met.

The magnitude of the de-stabilizing force on the carbon particles will be a function of the shape and velocity of the blade, and the clearance between the blade and the screen. It was accepted that these were aspects which would need to be established independently for the production unit, and the prototype was designed to permit operation with a variety of blade types, operating over a range of blade tip speeds.

With regard to the relationship between screen throughput rate, carbon concentration, and the frequency of pulsing, it was assumed that this could be approximated by a simplification of the well-known Carmen Kozeny equation for filtration, which is as follows:

$$\frac{1}{A}\left(\frac{dV}{dt}\right) = \frac{(-\Delta P)\, g_c}{\mu\alpha W(V + V_e)/A} \qquad (1)$$

Performing the integration of this equation for incompressible cakes (α = constant) and for operation at constant $-\Delta P$ and, assuming as a first approximation that the resistance of the screen medium is small in comparison with that of the accumulating carbon bed (i.e. $V_e = 0$), then:

$$\frac{V^2}{2} = \frac{A^2 g_c(-\Delta P)}{\mu\alpha W} t$$

For pulps of similar rheological characteristics and carbons of similar size distribution and shape, μ and α are constant, and a further simplification is possible:

$$Q = \frac{V}{At} = \frac{K'}{t}\left(\frac{(-\Delta P)t}{W}\right)^{1/2} \qquad (2)$$

This is an expression which should be capable of predicting the flow characteristics of a screen at varying pressure drops, carbon concentrations, and pulsing frequencies. If it is assumed that the screen is operated at a fixed pressure differential, just below the level resulting in 'pinning' of the carbon to the screen, then this point will correspond to the maximum flow capacity of the screen, and will be given by the following relationship:

$$Q_{max} = K''\left(\frac{1}{Wt}\right)^{1/2}$$

and since the cycle period $t = 1/60\, nb$

$$Q_{max} = K''\left(\frac{60nb}{W}\right)^{1/2} \qquad (3)$$

where K'' is a constant which is specific to a particular set of pulp and screen medium conditions.

A limited programme of laboratory testwork was carried out in an attempt to measure Q_{max} for a range of operating conditions, and the results are shown in Table 2. Although there is scatter in the values of the constant K'', this partly reflects the experimental difficulty in properly identifying the point of maximum flowrate, and it is considered that the results give reasonable support to the flow relationship suggested above.

Using the mean value of K'' (20.2), a graph was constructed to show the implied relationship between cycle period (pulsing frequency), carbon concentration, and maximum screen flowrate (Figure 5). The curves were extrapolated beyond the region where experimental data were accumulated (shaded portion of graph), simply to give an appreciation of trends which might result from a change in operating conditions.

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The inference from the graph is that the screen will operate satisfactorily as long as the impressed flowrate is below that defined by the appropriate carbon concentration and cycle period (i.e. $Q_{actual} \leqslant Q_{max}$). The reaction time of the screen is extremely rapid, and should the flow limit be exceeded, even momentarily, such as with a flow surge for instance, then the screen will 'blind'.

Plant scale testing of cylindrical screen

Following indications from the screening model illustrated in Figure 5, the first plant prototype was specified to operate over the same range of cycle period (seconds per 'pulse') as the laboratory unit. A comparison of typical screening parameters for the laboratory and prototype screen is given in Table 3, from which it may be seen that, in order to meet this scale-up criterion without exceeding an imposed upper limit of 2 m s^{-1} on tip speed, it was necessary to increase the number of blades from two to four.

A programme of plant scale testwork and development was carried out in one of the 2000 m^3 capacity adsorption tanks at the ERGO CIL plant. Improvements to the design of the screen were made continuously over a nine month period leading up to the design finally selected for the Daggafontein plant which is shown in Figure 6. [13,14].

Although not discussed in detail, the following is a list of the main changes which occurred in the design of the screen over the development period:

(i) Original lower submerged bearing rendered unnecessary by more robust design of shaft and top bearing.
(ii) Internal agitation rakes replaced by a conical internal section designed to raise pulp velocities above that causing sedimentation.
(iii) Transition from woven wire mesh (0,8 m square) to wedgewire (0,63 mm slot), to improve screen life (increased from approximately four weeks to more than 130 days at the time of writing).
(iv) Use of wedgewire for internal cone together with internal cleaning blades, to increase screening area by 41%.
(v) Use of two-speed motors. When the screen requirement is lower, the slower speed is selected to minimize erosion of screening media.

During the evaluation period, an attempt was made to determine the maximum flow capability of the screen (Q_{max}) for a range of operating conditions. The results of testwork carried out using the screen design finally selected are shown in Figure 7, which shows the influence of rotational speed (n), number of blades (b) and carbon concentration (W) on the screen capacity. Also shown are the capacity relationships predicted by the original screening model, allowing for the fact that the constant K'' has been corrected to account for the difference in open area between the square mesh (46%) and the wedgewire (22%).

Despite considerable scatter, there are indications that, at least for rotational speed and carbon concentration, the essential characteristics of the model are being followed. There is also an indication that the present configuration of screen has an upper limit of about 160 m^3 m^{-2} h^{-1}, and that beyond this point the presumed relationships fail to apply.

The Daggafontein interstage screens

Design and operating data for the screens to be used at Daggafontein are summarized in Table 4. As indicated in Figure 6, the two screens are installed towards the periphery of each adsorption contactors, where their separate drive mechanisms allow them to be operated independently of the main tank agitator. Under normal circumstances both screens are used at the low speed to handle the design flow of 1200 m^3 h^{-1}, but if one screen is removed for maintenance, then the other has the capability to handle the design flow at the higher speed rating. Should an individual screen 'trip' unexpectedly, a simple control system switches the remaining screen to a higher speed to compensate for the loss in flow capacity.

The screen itself comprises a screening basket and an agitator system. The screening basket is made up of vertical, 'exterior', and inclined, 'interior', screening surfaces, which are contiguous with an upper pulp chamber receiving the screened pulp product. The agitator has a system of angled blades, which are adjusted to pass in close proximity to both exterior and interior screen surfaces, as well as having a downthrust impeller attached to the central shaft. A pipe opening provides a flow connection between the main tank and the interior chamber of the screen.

The combined action of the inclined wiper blades, and the central 'pump' impeller, is to provide a downward recirculation of pulp past the inner screening surface, to prevent localized high carbon concentrations in this region. The screen and agitator form a single modular unit, which is easily located and sealed to the exit launder of the tank using a system of locating pins and securing wedges. 'Spades' on the tank side of the connection are used to throttle the flow to some extent, since this has been found to give additional stability to screen operation. A level-equalizing valve has been included to permit rapid recovery of the screen in the event of losing the equalized pressure condition.

The cylindrical screen is energy efficient in comparison with other screening systems due to its high flow capacity and its shape. Figure 8 shows the calculated power requirements for a typical air-cleaned EPAC system (pulp flow 40 m^3 m^{-2} h^{-1}; air flow 1 m^3 air m^{-3} pulp), the Kambalda screen (pulp flow 40 m^3 m^{-2} h^{-1}; power number N_p = 0,21 approximately), and the Daggafontein screens (pulp flow 100 m^3 m^{-2} h^{-1}; N_p = 8,0 approximately).

The curves show that, even at low treatment rates, the two mechanical screens are more efficient than air agitation and, as pulp flow increases beyond about 600 m^3 h^{-1} the Kambalda screen power increases substantially, and the cylindrical screen becomes increasingly more attractive.

The installed capital cost of the Daggafontein interstage screens (14 operational 14 installed stand-by, 8 spares) amounted to R1,7 million or R0,14 per annual ton of slimes treated. Although this is about double the cost of an air-agitated EPAC system of the type installed at the ERGO CIL plant,

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it is expected that the cost difference will be more than offset by recovery benefits resulting from a reduction in the degree of carbon back-mixing.

CONCLUSIONS

The search by AAC for a reliable, high-capacity interstage screen, for 'in-house' application, has resulted in the acceptance of the 'unit' mechanically agitated screen as the preferred alternative. The appeal of the screen lies in its relative simplicity, reliability and efficiency.

It is believed that the success of the screen, following closely on the heels of the 'linear belt' development for pre-screening duties, effectively counters any threat which 'pulp screening' may have posed to the general application of CIP technology within the AAC group.

ACKNOWLEDGEMENTS

This paper is published by kind permission of the management of AAC and the Society of Chemical Engineering (England, U.K.). The contribution of W. Stange, AAC sponsored postgraduate student at the University of the Witwatersrand, in providing information on the dynamic simulation of the CIP circuits, is gratefully acknowledged, as is the development work carried out at ERGO by L. Potgieter, A. MacIntosh, S. McDougall and others.

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[14] Screening device, *AAC, RSA Patent Application no. 86/6324,* August 1986.

LIST OF SYMBOLS

A	area for screening, m^2
C	gold tenor of solution, g Au (ton solution)$^{-1}$
g_c	gravitational constant, m h^{-2}
K_1	adsorption rate constant, ton solution (g Au)$^{-1}$ h^{-1}
K_2	desorption rate constant, h^{-1}
ΔP	pressure drop across screen, kgf m^{-2}
R	rate of gold removal g Au h^{-1}
t	time, h
V	volume of pulp screened, m^3
V_e	volume of pulp equal to screen media resistance, m^3
W	mass of carbon in slurry, kg m^{-3}
y^*	number of sites for adsorbing gold g Au (ton C)$^{-1}$
y	gold loading on carbon, g Au (ton C)$^{-1}$
α	specific resistance of carbon layer, m kg^{-1}
μ	viscosity, kg m^{-1} h^{-1}

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TABLE 1
Summary of AA associated CIP-CIL plants (effective June 1987)

Operation	Date commissioned	Tonnage treated (tons month $^{-1}$)	No. of streams	Flow per stream ($m^3 h^{-1}$)	Gold production (kg per annum)
President Brand calcine plant (Freegold)	August 1980	30 000	1	75	2 660
Afrikander Lease GP (Vaal Reefs Exploration and Mining Co.)	November 1981	60 000	1	100	1 140
Mount Morgan (Australia)[a]	December 1982	275 000	1	400	1 800
Simmergo GP (ERGO)	April 1983	180 000	1	500	734
No. 8 shaft GP (Vaal Reefs Exploration and Mining Co.)	April 1985	250 000	1	800	30 640
No. 9 shaft GP (Vaal Reefs Exploration and Mining Co.)	October 1984	140 000		400	9 060
ERGO CIL plant	April 1985	1 850 000	2	2 000	2 100
Cuiaba calcine plant (Morro Velho, Brazil)	March 1986	9 000	1	20	780[b]
President Brand GP (Freegold)	May 1986	390 000	3	300	19 300[b]
No. 1 Shaft GP (Western Deep Levels gold mine)	May 1986	120 000	1	650	14 400[b]
Raposas GP (Morro Velho, Brazil)	February 1987	25 000	1	60	1 870[b]
Daggafontein GP (ERGO)	March 1987	1 000 000	2	1 150	3 230[b]
TOTAL	–	4 329 000	–	–	87 714

[a] 40% AAC participation with Peko Wallsend.
[b] Anticipated production levels.

TABLE 2
Experimental data from laboratory cylindrical screen

Rotational speed n (rev min^{-1})	Number of blades b	Carbon concentration W ($kg\ m^3$)	Maximum screening rate Q_{max} ($m^3 h^{-1} m^{-2}$)	$K'' = \frac{Q_{max}}{(60nb/W)^{½}}$
50	2	30	245	17,3
60	2	30	300	18,2
40	2	60	180	20,1
50	2	60	255	25,5
Mean value				20,2

TABLE 3
Comparison of screening parameters for laboratory and plant cylindrical Test Screens

Description	Units	Laboratory (typical)	Plant (typical)
Screen diameter	m	0,20	1,50
Screen length	m	0,37	2,00
Screen area	m^2	0,23	9,42
Screen mesh size	mm	0,8 × 0,8	0,8 × 0,8
Wiper assembly diameter	m	0,28	1,60
Number of blades	–	2	4
Screen-blade clearance	mm	20	30
Shaft speed	rev min^{-1}	60	24
Tip speed	$m\ s^{-1}$	0,88	2,0
Cycle period	s	0,5	0,625

TABLE 4
Screen design and operating data for the Daggafontein CIL plant

Parameter	Units	Value
Plant throughout (solids)	tons/month $^{-1}$	1×10^6
Number of streams	–	2
Design pulp flow (per stream)	$m^3 h^{-1}$	1 200
Carbon concentration	kg m^{-3}	15 – 25
Number of screens per tank	–	2
Screen area per unit	m^2	13
Rotational speed (two speed)	rev min^{-1}	17 and 25
Number of blades	–	8
Blade clearance	mm	40
Screen media	mm	0,63 woven wire slots
Screen diameter	mm	1 600
Screen height	mm	1 800
Maximum flow[a] (low speed)	$m^3 h^{-1}$	1 200 (approximately)
Maximum flow[a] (high speed)	$m^3 h^{-1}$	1850 (approximately)
Power (low speed)	kW	3,5
Power (high speed)	kW	10

[a] Maximum flowrates apply to newly installed screens. It is typical to lose up to 25% capacity over a two week period owing to screen blinding with grit, but thereafter performance remains fairly constant.

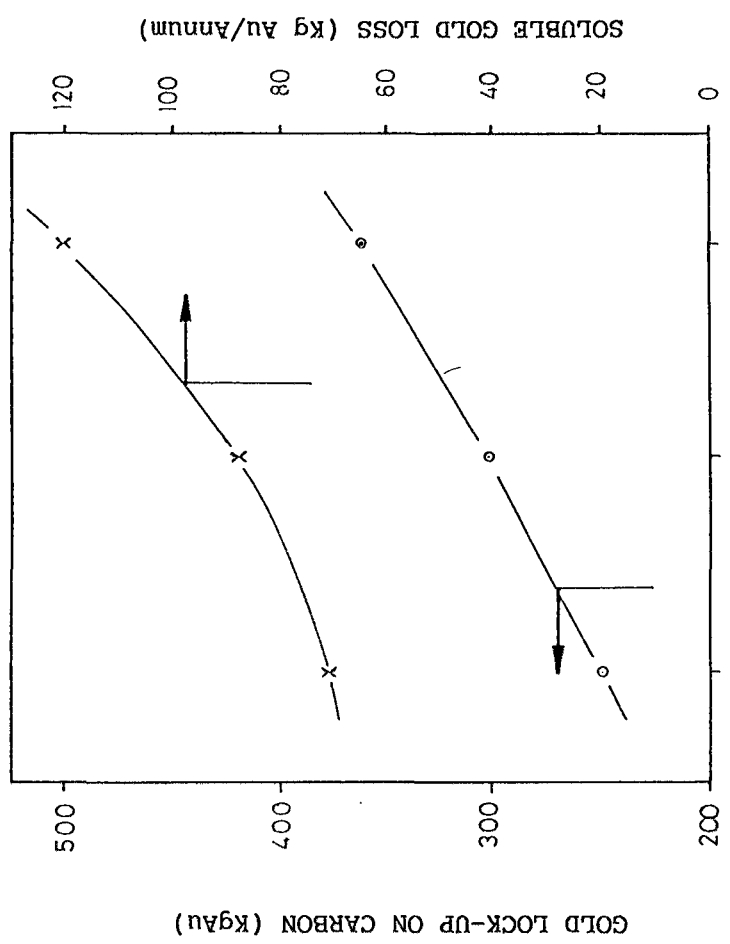

FIGURE 1
Effect of carbon 'leakage' on gold lock-up and soluble loss.

FIGURE 2
(a) Peripheral screen. (b) Epac screen

SECTION THRU KAMBALDA INSTALLATION

OPERATING DIAGRAM

FIGURE 3
Kambalda screen

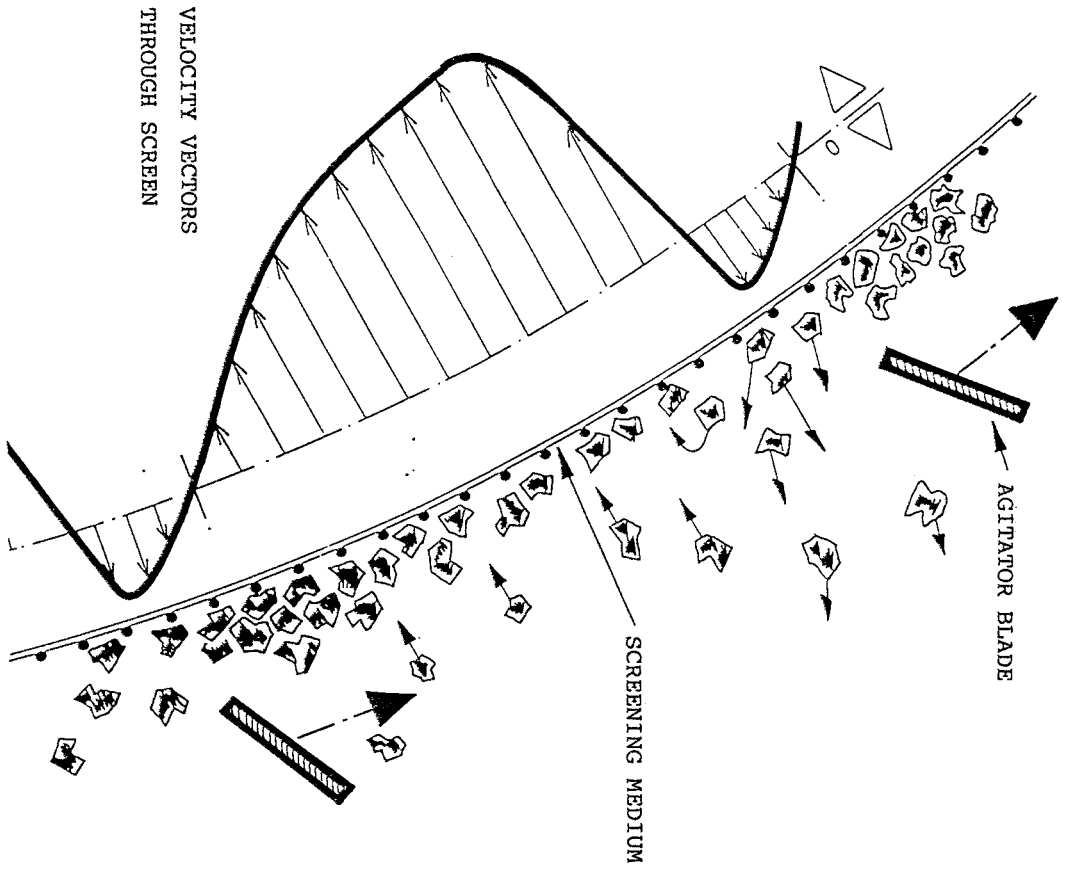

FIGURE 4
Diagram illustrating the operation of the mechanically cleaned cylindrical screen

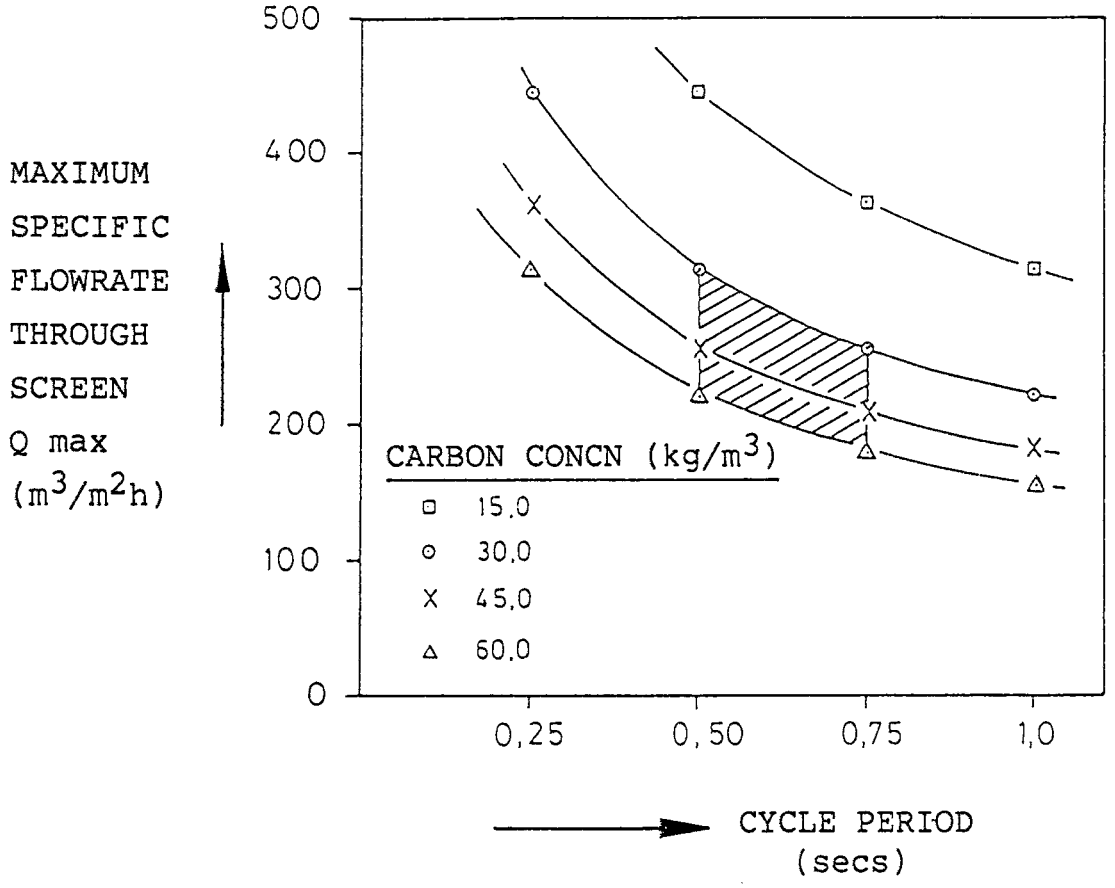

FIGURE 5
Prediction of flow characteristics using screen model

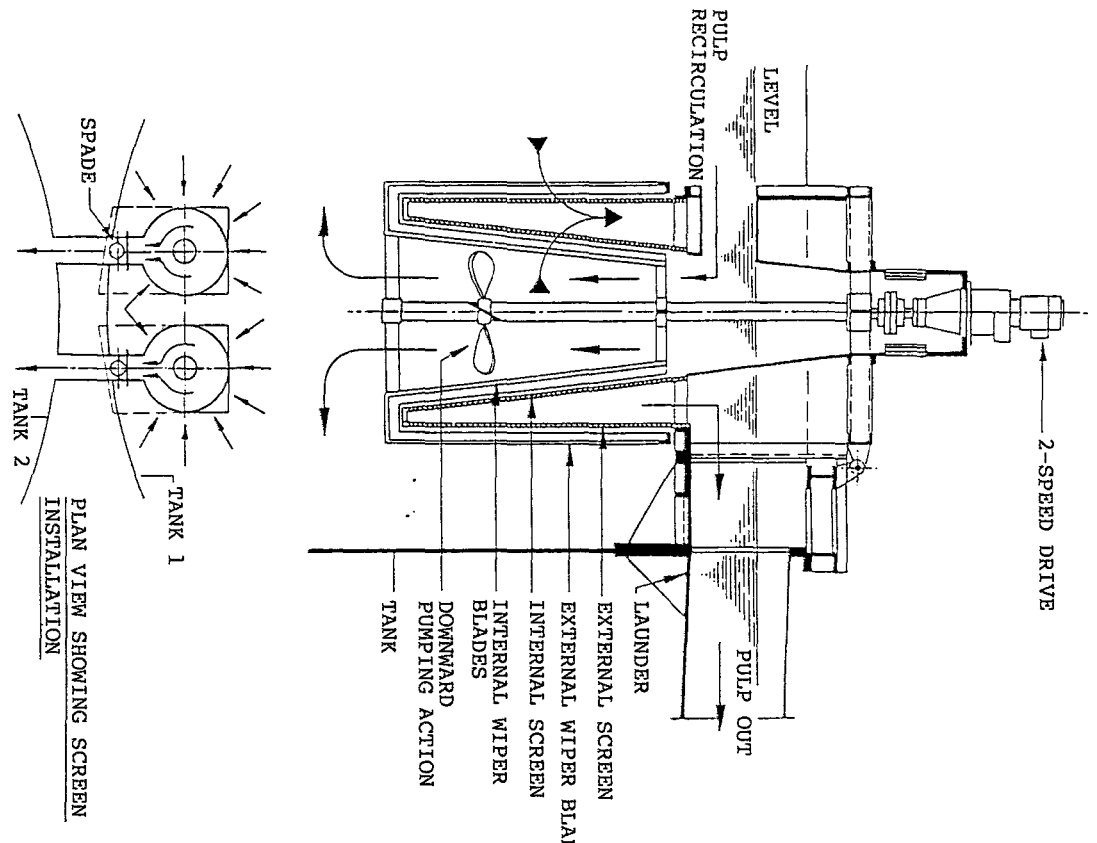

FIGURE 6
Modified NKM screen for Daggafontein plant

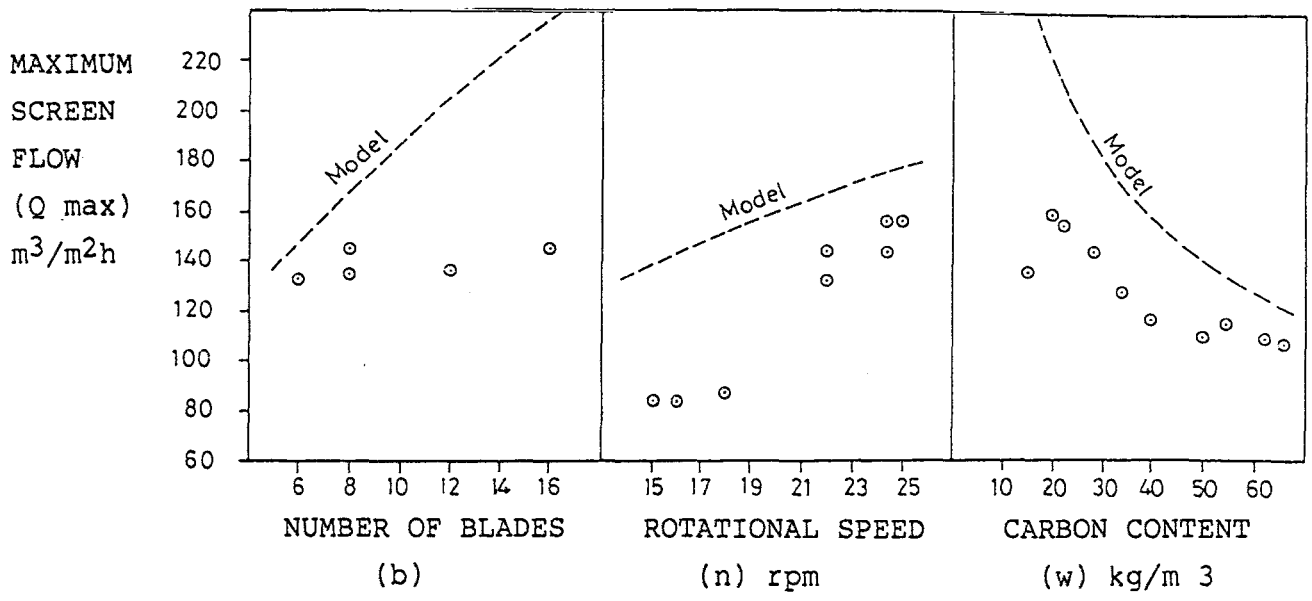

FIGURE 7
Flow characteristics of modified NKM screen for Daggafontein plant

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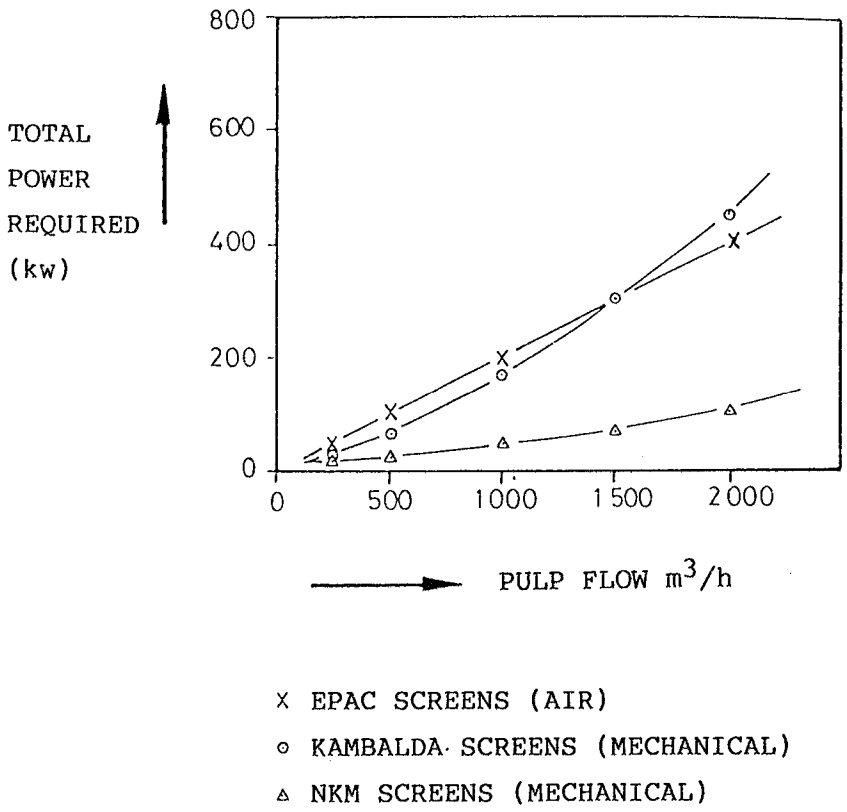

FIGURE 8
Effect of scale of plant operation on power requirements for various interstage screens (eight stages of adsorption)

The Optimisation and Control of Uranium Leaching from Pyrite Concentrate and Slime

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SYNOPSIS

Temperature and Sulphuric Acid concentration are two of the most important factors to be controlled in achieving optimum uranium leaching. A decision was made some years ago at the Metallurgical Scheme to maintain steam injection to the pyrite uranium leach only and to dispense with steam injection to the slime uranium leach previously heated to 60°C. The decision was taken after thorough investigation had revealed substantial economic benefits arising from reagent and steam generation cost savings.

It was recently observed that seasonal slime uranium recovery figures exhibited significant variations as ambient temperatures varied in the slime uranium plant, viz: 71% in winter and 78% in summer. Laboratory tests were set up to investigate and confirm the possible benefits of heating the slime uranium leach in winter to summer ambient temperatures as it appeared that recovery efficiency was very sensitive to temperature in the region of 30°C.

Increasing the leach temperature to a minimum of 33°C and controlling acid addition at 12 Kg/t resulted in an average increase of 16% in uranium dissolution in laboratory testwork. Decreasing the pyrite concentrate uranium leach temperature by 10°C from 65°C to 55°C had no deleterious effect on uranium dissolution or acid addition. These results were confirmed by subsequent full scale plant trials. Excess steam from the pyrite leach which was previously exhausted to atmosphere is now utilised in raising the temperature of the slime uranium leach by a few degrees in winter to achieve much improved uranium slime leach results.

INTRODUCTION

Uranium in South Africa is present as uraninite and is in association with the asphaltic mineral, thucholite, which is accompanied by considerable amounts of pyrite.

The uranium plant of the Metallurgical Scheme, sited at President Brand Mine in Welkom, treats two separate streams for the extraction of uranium, namely pyrite concentrate and slime. The pyrite is received from Flotation Plants sited at President Brand, President Steyn and Free State Geduld Mines. Slime is received from President Brand Flotation Plant and Welkom Gold

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978

PROVISIONAL SPECIFICATION

(Section 30 (1) – Regulation 27)

OFFICIAL APPLICATION NO.		LODGING DATE	
21	01	22	1 JULY 1980

FULL NAME(S) OF APPLICANT(S)

71	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

FULL NAME(S) OF INVENTOR(S)

72	RAYMOND JOHN DAVIDSON

TITLE OF INVENTION

54	"SEPARATION OF PARTICULATE SOLIDS FROM SLURRIES OR PULPS"

THIS invention relates to the separation of particulate solids from slurries or pulp (hereinafter referred to as "pulps").

One of the methods of recovering gold from cyanide pulps is to contact the gold-rich pulp with particulate activated carbon to absorb the gold values on to the carbon, separate the laden carbon from the lean pulp and then remove the gold values from the carbon. The gold-rich pulp will generally be produced in the standard cyanidation process, the gold values being in solution in the pulp in the form of a cyanide complex. The gold may be removed from the laden carbon in a number of ways, one of which is to contact the pre-treated laden carbon with deionised, distilled or otherwise softened water. This process of gold adsorption on to carbon is generally called a carbon-in-pulp process.

Contact between the gold-rich pulp and the carbon takes place in a series of tanks or carbon contactors as illustrated diagrammatically by Figure 1 of the attached drawings. Referring to this Figure, there are shown three tanks, 10, 12 and 14. Gold-rich solution and particulate carbon are introduced into tank 10. The pulp and carbon are agitated and the carbon adsorbs the gold-values from the solution. The pulp overflows into the surrounding launder 10b through screen 10a. Pulp from tank 10 passes into tank 12 where it is again contacted with particulate carbon. Further gold from the pulp is extracted and adsorbed on to the carbon. The pulp overflows through screen 12a, into launder 12b and then into tank 14 where it contacts more particulate carbon. The pulp, substantially stripped of gold, passes through the

screen 14a into launder 14b and then to further processing, if required.

While pulp is flowing in one direction under gravity there is an intermittent reverse flow of particulate carbon and pulp which is air-lifted from tank to tank. The particulate carbon passes from tank 14 to tank 12 and particulate carbon from tank 12 passes to tank 10. The gold-laden carbon is removed from tank 10 for further processing and fresh carbon is fed into tank 14. The process is thus continuous. Fresh carbon is fed to tank 14 and gold-rich pulp is fed to tank 10 while gold-laden carbon is removed from tank 10 and gold-stripped or lean pulp is removed from tank 14.

A problem with this process is the clogging up of the screens. These screens 10a, 12a and 14b are air purged, but nevertheless particulate carbon tends to collect on or in the screens and has to be removed mechanically. Periodically, the process has to be shut down for cleaning of the screens or individual screens have to be by-passed for cleaning purposes.

In a process of adsorbing values from solution or pulp on to a support in particulate form comprising contacting the solution or pulp with the support to produce a laden support and a lean solution or pulp and separating the laden support from the lean solution or pulp, the invention provides the improvement of separating the laden support from the lean solution or pulp by passing the support/solution or pulp mixture through a screen as hereinafter defined.

The screen which is used in the process of the invention presents to the incoming solution or pulp/solid to be separated a generally flat surface having a series of longitudinal slots or openings through which the solution can pass. The screen comprises a plurality of closed-spaced strips each having a tapered configuration in section, the strips being located adjacent each other so that their wide ends define the generally flat surface. The strips may be independent of each other or they may be individual windings of a continuous strip wound helically. In this latter case the screen has a cylindrical shape. These types of screens

are usually known as wedge-wire screens.

The invention further provides apparatus for use in the above-defined process, the apparatus comprising a container in which solution or pulp can be contacted with the particulate support, an outlet for removing solution or pulp from the container and a screen as defined above through which the solution or pulp must pass prior to exiting from the container through the outlet. Preferably means are provided to remove periodically or otherwise particulate support which collects on the upstream side of the screen.

The support may be particulate activated carbon or particulate resin and the process is preferably a gold-recovery process of the type described above. The process may also be a uranium recovery process in which the support is particulate resin.

An embodiment of the invention will now be described with reference to Figures 2 to 4 of the accompanying drawings. Figure 2 is a section through apparatus according to the invention, Figure 3 is a side elevation of a screen for use in the invention and Figure 4 is an enlarged fragmentary section of the ringed area of Figure 3.

Referring now to Figure 2, there is shown a tank 16 having a cylindrical screen 18 extending transversely across its interior. The screen is mounted for rotation, with one end being mounted in bearing 20 and the other end passing through the wall 22 of tank 16 and being mounted in sealed bearing 24. The open end 26 of the screen 18 provides the outlet for the tank.

The screen is illustrated by Figure 3 and 4. The screen comprises a strip 28 which is wedge-shaped in section and is wound into a helix. The winding is such that a thin gap 30 is provided between adjacent windings 32. The strip is so wound that the wide end 34 of the wedge is presented outwardly. Thus, the wider outer ends 34 together define a generally flat surface. The tapered end 36 of the wedge is presented

inwardly for each winding. The screen is provided with closed end 38 and open end 40.

In use, pulp and particulate support are placed in the container, the pulp filling the container in this example to a level above the upper surface 42 of the screen as illustrated by Figure 2. The screen in this example is rotated and pulp passes through the gaps in the screen and through outlet 26. Particulate support is trapped on the screen and may be removed therefrom periodically or otherwise. For example, a brush may be mounted to contact the outer surface of the screen and so continuously prevent any accumulation of particulate support thereon.

The apparatus has particular application to a gold recovery process of the type described above. The apparatus illustrated by Figures 2 and 3 can be used to replace the containers 10, 12 and 14 of the process described with reference to Figure 1. It has been calculated that for a throughput of 70 000 tons of pulp a month screens of internal diameter 10 mm are sufficient in each tank effectively to separate the gold-laden carbon from the lean pulp. Such a screen can easily be maintained free of accumulated carbon by means of a suitably located brush contacting the outer surface of the screen. The particulate carbon support is generally of a particle size in the range 1 to 2,4 mm and so the spacings or gaps between successive windings must be narrow enough to trap these particles.

DATED this 1st day of JULY 1980

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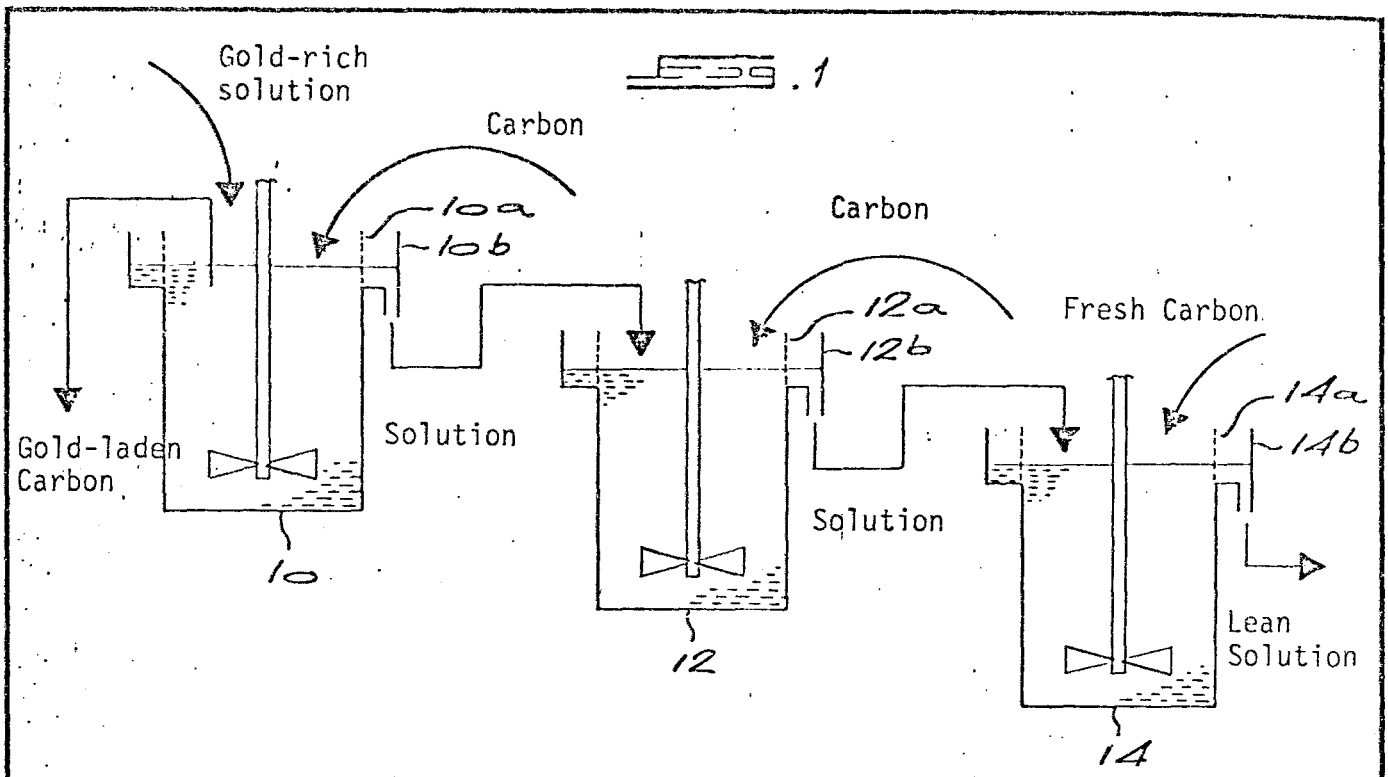

FIG. 1

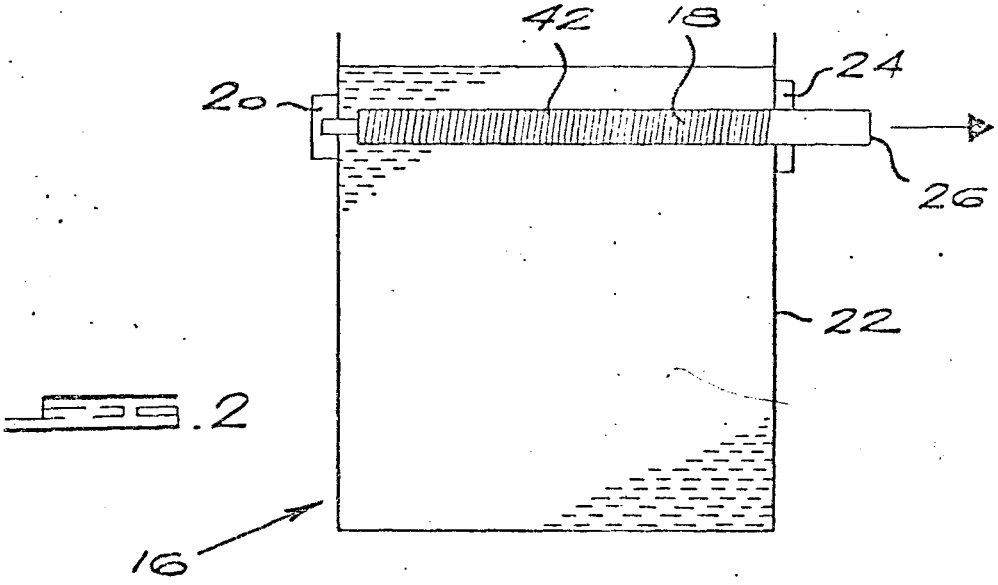

FIG. 2

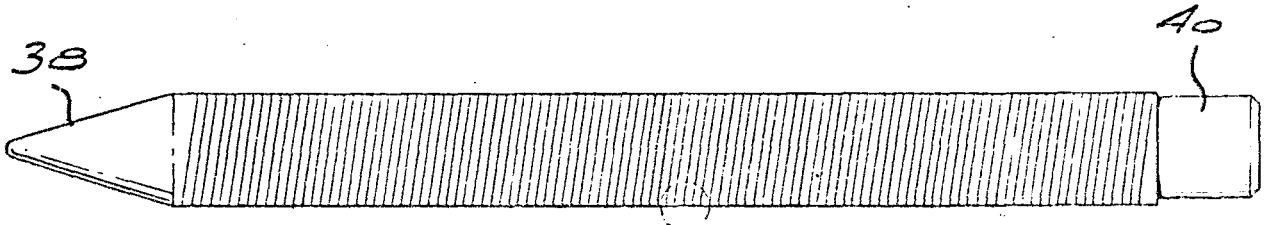

FIG. 3

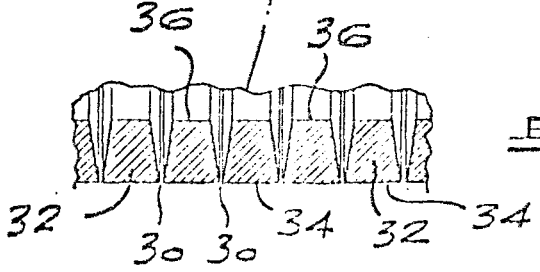

FIG. 4

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APPLICANT'S PATENT ATTORNEYS

2.4 THE PINNED BED

The concept of pumping cyanided pulp through a coarse bed of extruded activated carbon in a classical moving bed type of operation was patented in 1983 (SA Patent No 83/8659). Following the successful operation of a 200mm diameter pilot plant at Afrikander Leases where excellent mechanical and metallurgical performances were demonstrated, a full scale prototype plant was commissioned at Afrikander Leases (60 000t/m) at a cost of ca R2 million. Due to unforeseen mechanical problems the operation of the prototype plant was not successful and operation ceased in 1986.

Note: The presence of wood fibre proved a serious problem in the operation of the pinned bed and resulted in the development of the AAC-Delkor linear screen (2.2).

157.

R AND FISHER

FORM P.7
(To be lodged in duplicate)

Date of letter

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978

COMPLETE SPECIFICATION

(Section 30(1) – Regulation 28)

OFFICIAL APPLICATION NO.		LODGING DATE	
21 01	83/8659	22	21 NOVEMBER 1983

INTERNATIONAL CLASSIFICATION

51	C 22 B

FULL NAME(S) OF APPLICANT(S)

71	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

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TITLE OF INVENTION

54	RECOVERY OF METAL VALUES FROM A PULP CONTAINING SUCH VALUES IN DISSOLVED FORM

BACKGROUND OF THE INVENTION

This invention relates to the recovery of metal values from a pulp containing such values in dissolved form.

In mining operations, the mined ore is generally crushed and milled and is then subjected to a leaching step with a suitable chemical reagent in order to solubilise the metal values of interest in the ore, the resultant product being called a pulp. Such a pulp may contain from about 1 to about 60 percent by weight solids, but typically contains 30 to 50 percent be weight solids. Such pulps are commonly subjected to a liquid-solid separation stage prior to the recovery of the metal values from the resulting clarified or partially clarified solution. Typically such a liquid-solid separation stage may take the form of thickening, followed by filtration. Cycloning, centrifugation and counter-current decantation may also be used in such a liquid-solid separation stage.

Prior art methods of dealing with this clarified or partially clarified solution resulting from a liquid-solid separation stage as described above, cover a wide spectrum of metallurgical routes. Thus, for example, the use of ion exchange resins and solvents, adsorbents such as activated carbon, together with the chemical precipitation and electrowinning of metal values, are conventional techniques well known in the art of metal recovery.

The pulp may also be treated with ion exchange resins or adsorbents such as activated carbon in the so-called "in-pulp" processes, in which no prior liquid solid separation is carried out. One of the most successful methods currently being used is the carbon-in-pulp process for the recovery of gold and silver values. This method involves contacting the pulp with granular activated carbon. The gold and other valuable metal values are adsorbed on to the activated carbon from whence they may be recovered by a variety of known techniques. This method is usually carried out by contacting the leached pulp with the activated carbon in a counter-current fashion involving a series of contacting stages. In each stage the activated carbon is agitated in the leached pulp in a suitable vessel. Average pulp residence times in the process are commonly in the region of four to eight hours while average carbon residence times may be ten to thirty days. A high degree of interstage carbon mixing is common to the process by virtue of the methods used in transferring carbon from one stage to another. This carbon mixing has a detrimental effect on the metallurgical efficiency of the process.

While the carbon-in-pulp process achieves excellent recoveries of gold and silver values, it requires large contacting vessels and is energy intensive. The large contacting vessels are required because low carbon concentrations are used to minimise mechanical problems, e.g. inter-stage screening and breakdown of the carbon.

As the adsorption of gold and silver are critically affected by the carbon concentration, relatively slow carbon and pulp movement rates are common to the process. Such slow movement rates result in low gold loadings on the carbon going forward to the elution circuit as well as a high degree of metal value lock-up in the adsorption circuit. This lock-up of metal values in a carbon-in-pulp circuit may constitute a major indirect operating cost due to the value of the contained metals. Further, the interstage screening of carbon at each contacting stage of the process may also present a major mechanical problem particularly on large plants where relative screen areas are limited due to the large size of the contacting vessels.

SUMMARY OF THE INVENTION

According the present invention, there is provided a method of recovering metal values from a pulp containing such values in dissolved form including the steps of:

(a) forming a bed comprising a random mass of particles of a material capable of removing the metal values from the pulp, in a column having a feed end and a discharge end; and

(b) passing a stream of the pulp through the column from the feed end to the discharge end so that the particles of the bed are suspended in a substantially stationary non-fluidised state between the feed end and the discharge end of the column.

DETAILED DESCRIPTION OF THE INVENTION

The method of the invention will now be described in more detail.

Initially the particles in the bed are at rest. Once the stream of the pulp begins to be passed through the bed, the particles move towards the discharge end of the column and become suspended in a substantially stationary non-fluidised state between the feed end and the discharge end of the column. As the stream of the pulp passes through the bed, the metal values in the pulp are captured by the particles, and a concentration or loading gradient is established in the bed, with the particles closest to the feed end having higher metal value loadings than the particles near the discharge end. It is important that this concentration gradient be maintained in the bed so that periodically, the particles having the highest metal value loadings, may be removed from the bed, and replaced by fresh particles. This is achieved by ensuring that the particles are suspended in a substantially stationary non-fluidised state so that there is no mixing or agitation of the particles in the bed.

The column for use in the method of the invention is preferably inclined upwardly, with the feed end being below the discharge end, and more preferably, the column is substantially vertical. In this case, the stream of the pulp will be passed in an up-flow mode through the bed. The bed of the particles may be confined within the column by means of suitable screens which will allow the flow of pulp therethrough, but which will prevent the passage of the particles therethrough. When there is an upward flow of the pulp through the bed, the flow velocity and the viscosity of the pulp will generally be sufficient to lift and suspend the particles of the bed in a substantially stationary non-fluidised state.

The method of the invention preferably includes the step of periodically stopping the passing of the stream of the pulp through the column so that the particles of the bed move back towards the feed end of the column while remaining in a substantially non-fluidised state. This relaxation of the bed reduces any tendency for clogging or choking of the bed to occur.

The method of the invention preferably also includes the step of periodically removing the particles in a selected zone of the bed from the column. Preferably, the particles having the highest metal value loadings, i.e. the particles in the zone nearest the feed end of the column, are removed from the bed. The removed particles may then be further treated to recover the metal values therefrom.

Generally these removed particles will be replaced with a similar volume of fresh particles introduced into the column from the discharge end of the column.

The method of the invention may be applied to a plurality of beds in which each bed is discrete, or in which the beds form part of a continuous bed. Under normal operation, the passage of the pulp through the system is counter-current to that of the particles of the material forming the bed in order to maximise metallurgical efficiency. The movement of the particles of the bed in such an operation may be carried out either continuously or intermittently.

The material capable of removing the metal values from the pulp may be a suitable ion exchange resin, a suitable chelating resin, or an adsorbent such as activated carbon.

The rate of adsorption of metal values from solution must be such that the inventory of particles and the loading of metal values on the particles are within acceptable limits. Generally this implies the use of relatively small particles which have a relatively large surface area per unit volume.

Further, the particles of the material must be of a size and shape as to allow for the passage of the pulp through the bed of the particles. Thus, the particles may be elongate in shape and may be solid or may have an elongate hole extending partially or wholly therethrough. For example, when the material is activated carbon, the particles may comprise rod-like pieces having a substantially circular cross-section, with a diameter of 2 to 5 mm, which are made by an extrusion process. However, other shapes and sizes of particles are also suitable for the method of the invention, e.g. irregular shaped pieces with a maximum dimension of 2 to 20 mm.

Essential to the method of the invention is that the metal values are recovered from a pulp and not a clarified solution. The pulp, which may be formed by leaching a crushed and milled ore with a suitable chemical reagent, will generally have a solids content of 10 to 60 percent by weight, more typically 30 to 50 percent by weight. The solids in such pulps are usually in fine particulate form, e.g. having a typical particle size of less than 300 microns.

The method of the invention has particular application to the recovery of gold and silver values from a pulp containing such values in dissolved form. In this case, it is preferred to use particles of activated carbon in the bed.

Examples illustrating the method of the invention will now be given.

EXAMPLE 1

This example illustrates the treatment of a gold plant residue pulp taken from the President Steyn Gold Mine according to the method of the invention.

The pulp, resulting from a conventional gold leach operation, contained approximately 48 percent by weight solids. Approximately 80 percent of the solids had a particle size of less than 74 microns. As a considerable quantity of wood fibre was present in the pulp, this was screened off using a 425 micron vibrating screen prior to use. The pulp was contained in a 200 litre vessel fitted with a suitable stirrer from where it was circulated through a vertical column. The column was 2 metres long and had an internal diameter of 200 mm. A wedge wire screen with a 2 mm slit gap was positioned at the top of the column in order to confine the bed in the column. Approximately 30 kg of calcined coconut shell activated carbon, pre-screened to 3 to 15 mm in size, was contained in the column, leaving a freeboard depth of about 15 mm. The pulp was circulated through the column at a flow rate of 15 to 20 litres per minute. The gauge pressure drop across the column during this operation measured 80 to 90 kilopascals. No choking of the column with coarse solids was indicated following continuous operation lasting five hours.

EXAMPLE 2

This example illustrates the treatment of a gold plant residue pulp similar to that used in Example 1 according to the method of the invention.

The same column as used in Example 1 was used again, but the column contained approximately 30 kilograms of extruded activated carbon. The extruded activated carbon was tube-like in shape having a diameter of 3 mm with particle sizes varying from 5 to 15 mm long. The leached pulp was contained in a 200 litre vessel fitted with a suitable stirrer from where it was circulated through the column at a flow rate of 30 litres per minute. The gauge pressure drop across the column during this operation measured 80 kilopascals. No choking of the column with coarse solids was indicated following continuous operation lasting several days.

EXAMPLE 3

This example illustrates the gold recovery obtainable utilising the method of the invention.

Leached pulp containing 6 to 11 milligrams per litre of gold in solution was taken from the existing gold plant circuit at the Vaal Reefs Exploration and Mining Company (South Division) and pumped in an up-flow mode into a column containing activated carbon. The pulp resulting from a conventional gold leach using cyanide, contained approximately 40 to 50 percent solids by weight having an average particle size analysis as follows: 2,5 percent + 150 microns; 26,4 percent + 74 microns and 71,1 percent - 74 microns. The leached pulp also contained approximately 200 milligrams per litre cyanide as NaCN, and had a pH in the range 10 to 11,5. A considerable quantity of wood fibre was also present in the pulp and this was screened off using a 425 micron vibrating screen prior to use. The column was 10 metres long and had an internal diameter of 185 mm. A wedge wire screen with a 2 mm slit gap was positioned at the top of the column in order to confine the activated carbon in the column. Approximately 150 kilograms of 3 mm diameter extruded or molded activated carbon was contained in the column leaving a

free-board bed depth of about 300 mm. The pulp was pumped into the column at a flow rate of 15 litres per minute which corresponded to a superficial flow velocity in the column of 0,56 metre per minute.

The counter-current movement of carbon in the column was carried out on an eight hour shift basis. Loaded carbon was discharged from the bottom of the column, while a similar volume of virgin carbon was added to the top of the column. During normal column operation the line pressure to the column averaged 300 to 500 kilopascals (gauge pressure). The counter-current movement of carbon as described above had a most beneficial effect on reducing the line pressure to the column.

The column was run in the mode described for a period of 32 days with operational time including carbon movement approaching 97 percent. During this time the recovery of soluble gold in the pulp averaged 99 percent with the soluble feed to the column averaging 9,6 milligrams per litre of gold. When carbon was moved at a rate corresponding in volume to a bed depth of one third of a metre per eight hour shift, loaded carbon from the column contained 10 to 11 kilograms of gold per tonne of carbon. Likewise, when carbon was moved at twice this rate, the loaded carbon contained approximately 5 kilograms of gold per tonne of carbon.

EXAMPLE 4

Residue pulp containing 0,1 to 0,4 milligrams per litre of gold in solution was taken from the existing gold plant circuit at the Vaal Reefs Exploration and Mining Company (South Division) and pumped into a column as described in Example 3.

In the first test of 9 days duration, using a 10 metre long column, the gold in solution was reduced to 0,001 milligrams per litre. In a second test of 15 days duration, where the column was shortened to 5 metres, the gold in solution was reduced to 0,005 milligrams per litre. The respective gold recoveries were 99 percent and 95 percent. The pulp conditions in both these tests were similar to those described in Example 3. During normal operation on the 5 metre column, the line pressure to the column averaged 100 to 200 kilopascals (gauge pressure). In both tests the carbon movement rate was maintained at one third of a metre per 8 hour shift. The carbon removed was loaded with 150 to 200 grammes of gold per tonne of carbon. Any reduction in the carbon movement rate would result in higher gold loadings on the carbon.

The method of the invention has several advantages. Firstly, using the method of the invention, much lower pulp residence times in the column are obtainable, when compared with the carbon-in-pulp method. For example, pulp residence times for the method of the invention are a matter of minutes compared with residence times of several hours for the conventional carbon-in-pulp method. Secondly, it is possible to obtain higher metal value loadings on the metal-capturing particles.

CLAIMS

1.

A method of recovering metal values from a pulp containing such metal values in dissolved form includes the steps of:

(a) forming a bed comprising a random mass of particles of a material capable of removing the metal values from the pulp, in a column having a feed end and a discharge end; and

(b) passing a stream of the pulp through the column from the feed end to the discharge end so that the particles of the bed are suspended in a substantially stationary non-fluidised state between the feed end and the discharge end of the column.

2.

A method according to claim 1 wherein the column is upwardly inclined, with the feed end being below the discharge end.

3.

A method according to claim 2 wherein the column is substantially vertical.

4.

A method according to any one of claims 1 to 3 which includes the step of periodically stopping the passing of the stream of the pulp through the column so that the particles of the bed move back toward the feed end of the column while remaining in a substantially non-fluidised state.

A method according to any one of claims 1 to 4 which includes the step of periodically removing the particles in a selected zone of the bed from the column.

6.

A method according to claim 5 wherein the particles in the zone of the bed nearest the feed end of the column are removed from the bed.

7.

A method according to claim 5 or claim 6 wherein the particles removed from the bed are replaced with a similar volume of particles introduced into the column from the discharge end of the column.

8.

A method according to any one of the preceding claims wherein the particles have a relatively large surface area per unit volume.

9.

A method according to any one of the preceding claims wherein the material capable of removing the metal values from the pulp comprises a suitable ion exchange resin or a suitable chelating resin.

10.

A method according to any one of claims 1 to 8 wherein the material capable of removing the metal values from the pulp comprises activated carbon.

11.

A method according to any one of claims 1 to 10 wherein the particles are elongate in shape.

12.

A method according to any one of claims 1 to 11 wherein the pulp has a solids content of 10 to 60 percent by weight.

13.

A method according to claim 12 wherein the pulp has a solids content of 30 to 50 percent by weight.

14.

A method according to any one of the preceding claims for the recovery of gold and silver values from a pulp containing such values in dissolved form.

15.

A method of recovering metal values from a pulp containing such values in dissolved form includes the steps of:

(a) forming a bed comprising a random mass of particles of a material capable of removing the metal value from the pulp, in a column having a feed end and a discharge end;

(b) passing a stream of the pulp through the column from the feed end to the discharge end so that the particles of the bed are suspended in a substantially stationary non-fluidised state between the feed end and the discharge end of the column;

(c) periodically stopping the passing of the stream of the pulp through the column so that the particles of the bed move towards the feed end of the column while remaining in a substantially non-fluidised state; and

(d) periodically removing the particles in a selected zone of the bed from the column.

16.

A method according to claim 15 wherein the column is substantially vertical.

17.

A method according to claim 15 or claim 16 which includes the step of replacing the particles removed from the bed with a similar volume of particles introduced into the column from the discharge end of the column.

18.

A method according to any one of claims 15 to 17 for the recovery of gold and silver values from a pulp containing such values in dissolved form.

19.

A method of recovering metal values from a pulp containing such values in dissolved form substantially as herein described with reference to any one of Examples 1 to 4.

DATED this 21st day of NOVEMBER 1983

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SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS

2.5 THE TOPLESS MOVING BED

The development of a topless moving bed contactor for the recovery of soluble gold values from solutions and pulps is described. The present preliminary investigations aimed at the treatment of typical gold bearing pulps describes the physical deportment of 3mm extruded carbon in the contactor while recirculating residue pulp containing 30-40 per cent solids. Numerous conceptual aspects related to the operation of the contactor are discussed while further pilot testwork under plant conditions is recommended.

Because of the present depressed economy in the gold industry, funds for the further development of the topless moving bed contactor were not forthcoming. The presentation of the appended internal report is with the kind permission of the Anglo American Corporation.

DEVELOPMENT OF CARBON CONTACTOR

1.0 INTRODUCTION

While the concept of treating gold plant leached pulps using a coarse extruded carbon in a pinned bed type of operation appeared to have many metallurgical benefits, the physical performance of the circuit as introduced at Afrikander Leases had several operational short-comings. Suspected channel formation in the compacted carbon bed together with frequent choking of the system with wood fibre and/or coarse solids contributed to some of these problems. Difficulties associated with carbon movement into and from the circuit were also encountered.

The present investigation was aimed at maintaining the metallurgical efficiency of the moving carbon bed circuit while possibly improving the mechanical performance of such an operation. Rather than maintaining a packed carbon bed as in the pinned bed, the present concept was to allow for the expansion or compression of the carbon bed via the presence of a freeboard under the carbon bed while minimising any carbon mixing. The development of a simple "topless" peripheral screen also indicated significant practical benefits when compared with the screens used in the pinned bed at Afrikander Leases.

2.0 EXPERIMENTAL

2.1 Activated Carbon

Norit type RB4 extruded carbon having an apparent bulk density of 450 g/ℓ and pellet diameter of 2,9 mm was soaked in water overnight prior to use.

2.2 Moving Bed Contactor

The carbon contactor fabricated from clear PVC is represented diagrammatically in Figure 1. Carbon was contained in the vessel using a peripheral screen (2 mm square mesh) placed inside an overflow launder. A second 2 mm square mesh screen (not shown) was placed in the pulp feed inlet. The live or occupied volume of the contactor was 260 ℓ.

A paddle-type rake incorporating 5 horizontal paddles (445 mm x 50 mm) was positioned in the vessel in order to prevent any channel formation which may take place. The rake rotated at 1-2 rpm and was powered by a 30 Watt "sheep braai" motor (Bircraft SA (Pty) Ltd). Torque on the motor was monitored using an ameter which indirectly reflected the fluidity of the carbon bed when pumping pulp through the bed.

2.3 Plant Pulp

Residue pulp from Western Deep Levels No. 2 Shaft gold plant was used to simulate a plant feed to the carbon contactor. 350 ℓ of pulp was prepared at various pulp densities prior to use. Pulp was recirculated through the contactor using a type SP-40 Bredel pump. The maximum flowrate of the pump was 41 ℓ/min

2.4 Test Procedure

Pulp was recirculated through the contactor at various flowrates and pulp densities in order to observe the movement of the carbon bed while also monitoring any classification of the feed pulp. The effect of flowrates and pulp densities on effective screen areas and gauge pressures were also monitored during 10 days of continuous operation (effective screen area may be defined as the screen area actually in use during plant operation).

3.0 OBSERVATIONS

3.1 100 ℓ Carbon Addition

100 ℓ of extruded carbon was added into the open top of the contactor while circulating pulp containing 30 per cent solids. When pumping pulp at 14-41 ℓ/min, the top of the suspended carbon bed was contained by the peripheral screen while the carbon bed depth was seen to compress from 630 mm to 375 mm on increasing flowrates (Table I). A small increase in effective screen area was also observed as flowrates were increased. Due to the large freeboard present under the carbon bed (ca 150 ℓ), a significant degree of classification of the feed pulp at low flowrates was also indicated with the solids content of the exit pulp falling to 22 per cent. No noticeable increase in the gauge pressure was evident when flowrates were increased to 41 ℓ/min.

3.2 150 ℓ Carbon Addition

Following the addition of a further 50ℓ of carbon as before (3.1) pulp containing 32 per cent solids was circulated through the contactor. On increasing flowrates from 14 to 37 ℓ/min, the carbon bed depth was observed to decrease from 960 mm to 920 mm with the effective screen area increasing from 1335 cm^2 to 1492 cm^2 (Table I). On increasing the flowrate to 41 ℓ/min, raking of the carbon bed ceased due to the low power of the motor (30 Watt). This resulted in the carbon bed depth decreasing to 1178 cm^2. As before (3.1) no increase in the gauge pressure was evident when flowrates were increased while a smaller degree of pulp classification was evident due to the smaller freeboard present (ca 80 ℓ).

3.3 180 ℓ Carbon Addition

A further addition of 30 ℓ carbon resulted in the carbon bed occupying the entire cylindrical volume of the contactor leaving a freeboard of ca 60 ℓ in the cone section. Following the circulation of pulp containing 36 per cent and 40 per cent solids respectively

(Table I), it may be seen that at these higher pulp densities only a small degree of bed compression was observed as flowrates were increased while gauge pressures increased only marginally. Some classification of the pulp feed was again noted at the lower flowrates used. A carbon concentration of ca 360 g/ℓ in the carbon bed may be noted.

3.4 Mechanical Performance and Test Results

Following 10 days of continuous operation under the prescribed conditions (3.1-3.3) during which time gauge pressures did not exceed 50 kPa, the carbon bed remained mobile and was easily expanded or compressed by changing or even stopping pulp flowrates. No change in pulp coloration or the presence of floating carbon slime during this time indicated very little carbon abrasion.

The test results presented in Table I do no more than validate earlier findings related to the recirculation of plant pulp using the 190 mm diameter pinned bed pilot plant (AARL Project R/85/223, Report No. 1). The present investigation does however demonstrate the operation of a simpler more flexible carbon contactor in which some of the mechanical short comings encountered with the pinned bed may be overcome. Some of these conceptual advantages are presented below.

4.0 CONCEPTUAL ADVANTAGES

The present investigation although of a preliminary nature together with the conceptual flowsheet presented in Figure 2 would appear to indicate numerous merits which may be discussed as follows. As many of these aspects are at this stage conceptual, a certain degree of caution is required regarding any final conclusions.

1. The design and operation of the contactor is simple and robust. There are no moving parts but for the rakes which would only be used intermittently to monitor (torque gauge) and maintain the mobility of a channel-free carbon bed. Low capital costs when compared with existing carbon circuits may be expected.

2. The open peripheral screen at the top of the contactor affords the easy addition of circuit carbon and any maintenance services required. The screen could readily be replaced without disturbing production should cleaning be necessary. Any blinding of the screen would result in the gradual upward movement of the carbon bed which in turn would expose additional effective screen area and so balance the screen duty required. A level control device positioned at the top of the screen could activate a pulp feed bypass which would then collapse the entire carbon bed temporarily.

3. High mass transfer due to high intra-particle flow velocities in the carbon bed together with the minimum disruption of the loading profile are attainable when operating in the pinned bed mode (using no mechanical mixing). Both intermittent bed expansion (lowering pulp densities or flowrates or pulp bypassing) and the intermittent

use of the rakes minimise the vertical mixing of carbon. Where possible, the contactor would be operated in a pinned bed mode with no carbon movement. Further on-site testwork would be required to define such carbon mixing.

4. Low carbon losses in the contacting circuit may be expected due to the absence of mechanical mixers.

5. The volumetric discharge of loaded carbon from a side port in the contactor is simple. This port would also control the freeboard in the contactor (Figure 2).

6. The high ℓ/d aspect of the adsorption circuit (Figure 2) would complement the high ℓ/d aspect of a continuous elution circuit in which loaded carbon is conveyed via a vertical bucket elevator to the elution circuit (the hydraulic transport of carbon containing high gold loadings is not recommended due to the high abraded gold losses incurred).

7. Intermittent acid washing of the entire carbon inventory in the adsorption circuit would improve both the mechanical and metallurgical performance of the contactor while making any acid washing in the elution circuit unnecessary.

8. Coarse solids which would otherwise choke a pinned bed type of operation would collect and classify in the freeboard area of the contactor from where they may be easily removed. The recovery of this coarse material and/or pyrite may benefit overall gold recoveries following remilling or pyrite flotation stages.

9. Water flushing of the contacting circuit following a power breakdown would obviate any bed compaction and subsequent start-up problems.

10. Unlike the pinned bed operation where gauge pressures often exceeded 500 kPa, very much lower pressures in the contactor may be expected.

11. While the present investigation was aimed at the treatment of gold bearing pulps and residues, a similar contactor could also be used to treat unclarified gold plant pregnant and effluent solutions. This contactor, possibly using a smaller particle size carbon, would be more cost effective in terms of capital and ease of operation than the classical moving bed, the NIMCIX or the expanded bed cascade type of operation.

12. The concept may even apply to the continuous recovery of other metal ions using other adsorbants or ion exchange resins.

5.0 DISCUSSION

Many of the conceptual aspects discussed, require further investigation as the present study is of a very preliminary nature. The construction of a mobile moving bed pilot plant is recommended in order to generate more

meaningful metallurgical and mechanical performance data. Such an investigation would include the on-site treatment of high and low grade pulps, solutions and unclarified effluents.

If it may be assumed that if a superficial rising velocity of 1 m/min were used in a 100 000 t/m moving bed operation treating pulp averaging 40 per cent solids (velocities of 1-2 m/min were used in the pinned bed prototype plant at Afrikander Leases), this would require a column diameter of 2,36 m. The size and throughput of such a plant would appear to warrant such an investigation.

INVESTIGATED AND REPORTED BY:

R.J. Davidson

R.J. Davidson
HEAD: CARBON METALLURGY

RJD/CGS/lo'r

TABLE I

THE EFFECT OF FLOWRATES AND PULP DENSITIES ON CARBON DEPORTMENT

Influent Flowrate (ℓ/min)	Superficial Rising Velocity (cm/min)	Gauge Pressure (kPa)	Carbon Bed Depth (mm)	Carbon Conc. (g/ℓ)	Effective Screen Area (cm^2)	Effluent Pulp Density (% solids)
100 ℓ Carbon Addition Using Pulp Containing 30% Solids						
14	7,1	35	630	363	ND	22
25	12,7	35	640	357	ND	25
32	16,3	35	493	464	1335	27
37	18,8	35	450	511	1492	29
41	20,9	35	375	608	1570	30
150 ℓ Carbon Addition Using Pulp Containing 32% Solids						
14	7,1	38	960	359	1335	26
25	12,7	38	925	371	1570	28
32	16,3	38	920	373	1492	29
37	18,8	38	920	373	1492	32
41*	20,9	38	850	404	1178	32
180 ℓ Carbon Addition Using Pulp Containing 36% Solids						
14	7,1	40	1145	360	1367	28
25	12,7	40	1146	360	1414	29
32	16,3	42	1147	360	1398	36
180 ℓ Carbon Addition Using Pulp Containing 40% Solids						
14	7,1	42	1147	360	1382	35
25	12,7	42	1148	360	1450	35
32*	16,3	44	1150	358	1524	39
37*	18,8	48	1150	358	1649	40
41*	20,9	50	1150	358	1806	40

*Raking of bed discontinued and contactor operated in pinned bed mode.

Note: a) the measurement of carbon bed depths at low carbon additions (100 ℓ) and low pulp densities (30% solids) were difficult to define due to the expanded nature of the bed

b) the effective screen area particularly at low pulp flowrates was difficult to predict due to some silting taking place in the exit launder

c) both a) and b) do however depict relative trends.

FIGURE 1 CARBON CONTACTOR FOR THE TREATMENT OF GOLD PLANT CYANIDED PULPS AND RESIDUES

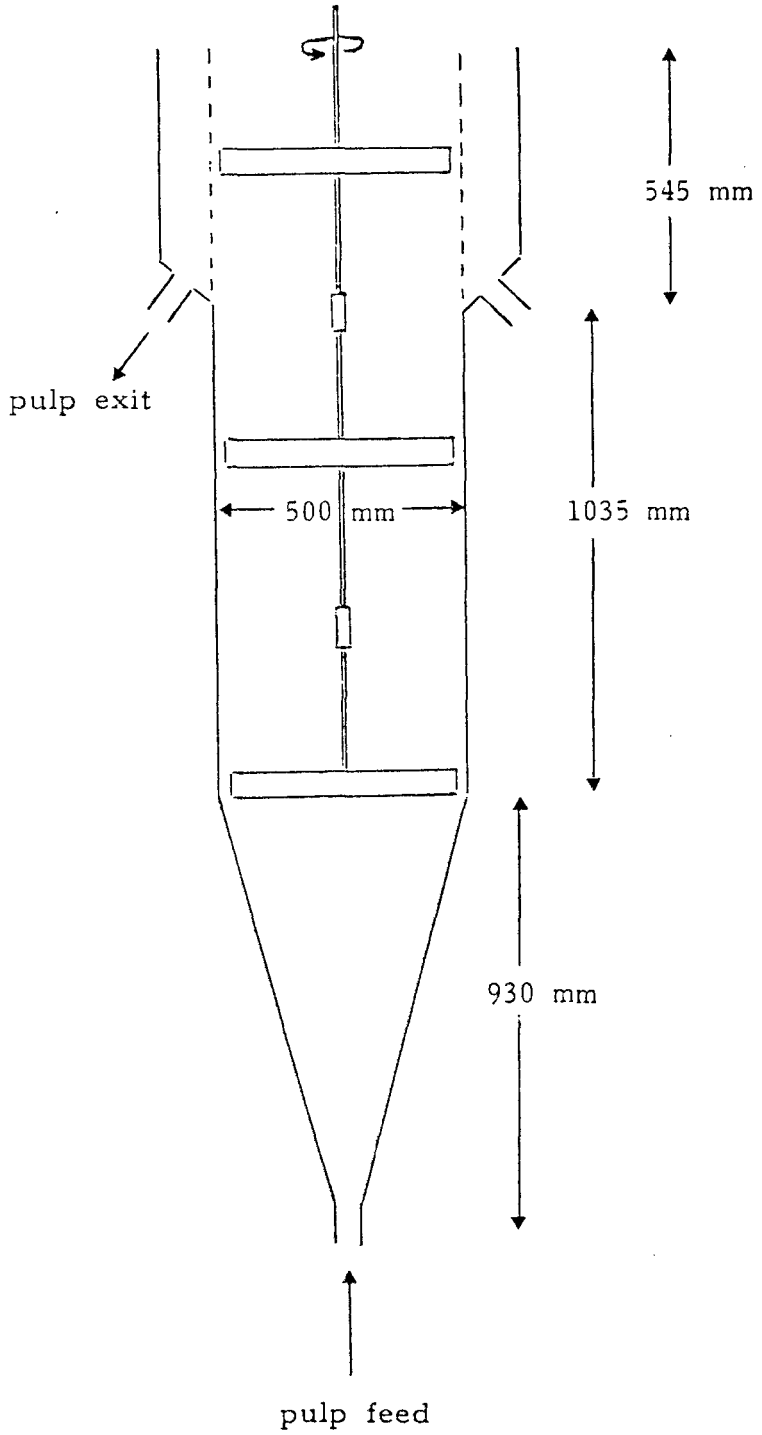
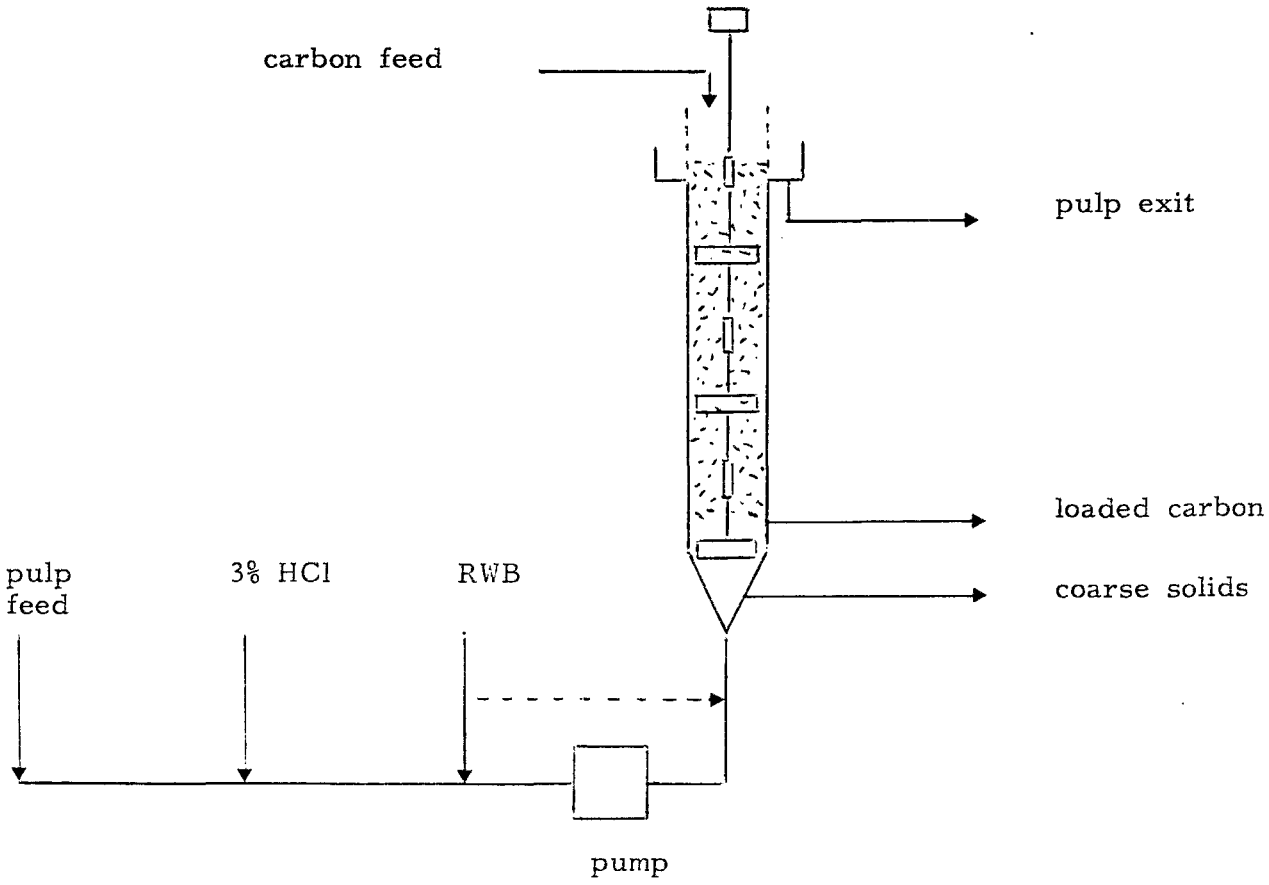


FIGURE 2 : CONCEPTUAL FLOWSHEET OF TREATMENT CIRCUIT



3.0 TREATMENT OF SOLUTIONS AND EFFLUENTS

Following investigations into the recovery of gold and silver from both synthetic and plant solutions, pilot plant operations were carried out using several different modes of carbon contacting.

3.1 FACTORS AFFECTING GOLD ADSORPTION

Early investigations into the parameters affecting gold adsorption onto activated carbon were aimed at the recovery of gold and silver values from synthetic as well as existing plant solutions and effluents. Both the lowering of pH (SA Patent No. 78/5147) and the intermittent acid washing of the carbon inventory (SA Patent No. 83/3952) were developed as a means of improving gold and silver recoveries and loadings. Both procedures are currently widely used in the gold industry for the treatment of plant effluents.

The use of activated carbon for the recovery of gold and silver from gold-plant solutions

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SYNOPSIS

When activated carbon is used for the recovery of gold and silver from Witwatersrand gold-plant solutions, severe fouling of the carbon by calcium carbonate is experienced. Also, difficulty is encountered in the retention of the adsorbed silver on the carbon because it is displaced by gold.

The acidification of plant pregnant solutions before they are contacted with carbon was therefore investigated as a means of overcoming these shortcomings. The results from batch laboratory experiments on the effect of pH on gold adsorption from plant solutions and from synthetic solutions of high ionic strength indicated reaction rate optima in the region of pH 5 and below pH 1. The effect of pH on the equilibrium capacity constant for gold adsorption onto activated carbon was also studied, using both a plant solution and a synthetic solution of high ionic strength. In the case of the plant solution, the results suggested only a marginal change in carbon capacity for gold in the pH range 12 to 6, with significant increases in capacity at lower pH values. In the case of the synthetic solution of high ionic strength, the capacity constant was found to increase steadily as the pH was lowered from 12 to 1. It was also found that the gold adsorption capacity of the carbon was significantly better for the synthetic solution. No precipitation of gold from the pregnant solution (gold content approximately 6 g/t) was noted as the pH was decreased to values as low as 1. However, about 25 per cent of the 0,6 g/t silver content was found to have precipitated when a pH value of 3,0 was attained. Further substantial precipitation of silver took place as the pH was lowered further.

The investigation was extended by the passing of suitably acidified gold-plant solutions through one or more columns containing activated carbon. Preliminary results showed that the particle size of the carbon had a significant effect on the rate of gold adsorption. No displacement of adsorbed silver by gold was observed to take place at an influent pH of 5. In a typical continuous 'merry-go-round' operation involving three columns in which gold was eluted from the lead column every 48 hours, effluent values averaging 0,001 g/t gold could be consistently achieved under simulated plant-operating conditions. The results were obtained at relatively high velocities of influent solution, resulting in very little hold-up of gold in the carbon-recovery circuit.

More than 99,9 per cent of the gold and more than 99,8 per cent of the silver was recovered in the adsorption circuit, while the elution data indicated an average elution recovery of 99,8 per cent of the adsorbed gold and 98,2 per cent of the adsorbed silver. Elution of the carbon was carried out at 91 °C. This involved pretreatment of the carbon with a sodium cyanide solution, followed by elution with deionized water. At a loading flowrate of 66 bed volumes per hour, the indicated take-up of metals on the carbon in the lead column was as follows: 53 kg/t gold, 6 kg/t silver, 57 kg/t copper, 14 kg/t nickel, and 1 kg/t zinc. Only a trace of cobalt was adsorbed. At an elution rate of 1,7 bed volumes per hour, the lead column was found to be completely stripped in under 5 hours, with concentrations of gold in the eluate reaching 6 to 7 g/l.

Besides the excellent adsorption and elution results obtained when acidified pregnant solution was used, no fouling of the carbon was indicated, and the pressure drop across the three columns was considered to be well within practical limits. When compared with conventional zinc precipitation, the carbon process would appear to have considerable economic merit, with strong indications that significant savings would be possible both in capital and operating costs. Further pilot-plant experiments, preferably on a gold plant, would be required before a meaningful cost comparison could be made with the existing process.

SAMEVATTING

Wanneer geaktiveerde koolstof gebruik word vir die herwinning van goud- en silwerwaardes uit Witwatersrand goudaanlegoplossings word besoedeling van die koolstof deur kalsiumkarbonaat ondervind. Verder is dit moeilik om geadsorbeerde silwer op die koolstof te behou weens die verplasing daarvan deur goud.

Die aansuring van dragtige aanleg oplossings voordat dit in aanraking met koolstof kom, is ondersoek as 'n moontlikheid om hierdie tekortkoming te oorkom. Gegewens van lotgewys laboratorium eksperimente op die effek van pH op goud adsorpsie van aanlegoplossings en sintetiese oplossings met 'n hoë ioonsterkte het gedui op reaksietempo optima in die omgewing van pH 5 en onder pH 1. Die invloed van pH op die ewewigskapasiteitskonstante vir die adsorpsie van goud op geaktiveerde koolstof, was ook ondersoek met beide aanlegoplossings en sintetiese oplossings van hoë ioonsterkte. In die geval van die aanlegoplossing was daar volgens die resultate slegs 'n geringe verandering in die goud kapasiteit van die koolstof in die pH gebied 12 tot 6, met betekenisvolle toenames by laer pH waardes. In die geval van die sintetiese oplossing met hoë ioonsterkte is gevind dat die kapasiteitskonstante voortdurend toeneem terwyl die pH verlaag word van 12 na 1. Dit is ook gevind dat die adsorpsiekapasiteit van koolstof vir goud betekenisvol beter is vir die sintetiese oplossing. Geen presipitering van goud uit die dragtige oplossing is opgemerk wanneer die pH tot selfs 1 verlaag is nie. Dit is egter gevind dat 25 persent van die silwer reeds neergeslaan het wanneer 'n pH van 3,0 bereik is. Met verdere verlaging van die pH het aansienlike verdere presipitering van silwer plaasgevind.

Verdere ondersoek was beperk tot die deurlating van geskikte aangesuurde goudaanlegoplossings deur een of meer kolomme wat geaktiveerde koolstof bevat. Voorlopige resultate het getoon dat die deeltjiegrootte van die koolstof 'n betekenisvolle uitwerking op die tempo van goudadsorpsie het. Geen verplasing van geadsorbeerde silwer deur goud is waargeneem by 'n invloei pH van 5 nie. In 'n tipiese ononderbroke 'mallemeule' opset met drie kolomme, en waarin goud elke 48 uur van die eerstekolom ge-elueer is, kon gemiddelde goudwaardes van 0,001 g/t bereik word onder gesimuleerde aanleg werktoestande. Die gegewens is by relatief hoë invloei oplossingstempo verkry met geringe ophou van goud in die koolstof herwinningsbaan.

Meer as 99,9 persent van die goud en meer as 99,8 persent van die silwer is herwin in die adsorpsiebaan, terwyl 'n gemiddelde elueringsherwinning van 99,8 persent van die geadsorbeerde goud en 98,2 van die geadsorbeerde silwer deur die elueringsgegewens aangedui is. Eluering van die koolstof is by 91 °C uitgevoer. Dit sluit vooraf

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behandeling van die koolstof met natriumsianiedoplossing in, gevolg deur eluering met gedeioniseerde water. By 'n ladingsvloeitempo van 66 bedvolumes per uur is die volgende aanduiding verkry vir opname van metale op die koolstof in die eerstekolom:- 53 kg/t goud, 6 kg/t silwer, 57 kg/t koper, 14 kg/t nikkel, 1 kg/t sink. Slegs spore van kobalt het geadsorbeer. Met 'n elueringstempo van 1,7 bedvolumes per uur is dit gevind dat die eerstekolom volledig gestroop kan word in minder as 5 uur. Goud konsentrasies in die eluaat het 6 tot 7 g/l bereik.

Behalwe vir die uitstekende adsorpsie en elueringsdata wat verkry is met aangesuurde dragtige oplossings, was daar geen aanduiding van besoedeling van die koolstof solank die drukval tussen die drie kolomme binne praktiese perke gehou is nie. Wanneer die proses met konvensionele sinkpresipitering vergelyk word, blyk dit dat die koolstofproses aansienlike ekonomiese voordele inhou, met sterk aanduidings van moontlike betekenisvolle besparings beide in kapitaal en bedryfskoste. Verdere proefnemings op toetsaanleg skaal, verkieslik op 'n goudaanleg, is nodig voordat 'n betekenisvolle kostevergelyking met die bestaande proses gemaak kan word.

Introduction

The zinc-cementation process for the recovery of gold and silver from cyanide solutions was introduced on the Witwatersrand in 1890[1]. With the addition of lead salts and the use of zinc dust rather than shavings, and with de-aeration of solutions, the process is one of great efficiency considering the low gold tenor of the solutions involved. Like cyanidation, cementation is a complex heterogeneous redox process. Zinc is dissolved at the anodic areas of a surface, and the electrons released serve to reduce the aurous ions at the cathodic areas.

It is generally accepted that cementation is sensitive to alkalinity and to the concentration of free cyanide in solution. Moreover, many of the more common constituents of gold-cyanidation solutions influence the cementation reaction. For example, sodium sulphide and the cyanide complexes of copper, arsenic, and antimony significantly decrease the recovery of gold when they are present in concentrations of more than 10^{-5} M. The extent of the decrease in gold recovery increases progressively with concentration until cementation ceases altogether.

The Merrill–Crowe process as currently practised in South Africa requires very careful control in order to yield barren solutions analysing less than 0,01 g/t gold. It is, at the same time, a relatively labour-intensive operation.

Activated carbon was introduced in about 1880 as a means of recovering gold from solution. It was a strong competitor for zinc precipitation during the early years of the cyanidation process. However, it rapidly lost ground as the technology of zinc cementation improved and because of certain limitations associated with its use, which are discussed below.

An important characteristic of activated carbon is its ability to function with unimpaired efficiency in foul cyanide solutions. It has been used for the recovery of residual gold from extremely dilute barren solutions and from the effluent from slimes dams. Carbon-in-pulp[2, 3] has been used to adsorb gold direct from cyanidation pulps, thus obviating any filtration stage.

The lack of a suitable means of stripping adsorbed gold from loaded carbon presented a serious disadvantage to the carbon process. In relatively recent times, processes have been developed that allow the carbon to be re-used. The gold is eluted with alkaline solutions of sodium sulphide and sodium sulphite[4], a boiling solution of sodium cyanide and sodium hydroxide[5], deionized water[6], or organic solvents[7], and the gold is then recovered from solution by electrolysis.

Investigations carried out at the Anglo American Research Laboratories (AARL) into the use of activated carbon for the recovery of gold from gold-plant effluents have resulted in the introduction of several important innovations in this field of gold technology. More efficient means of gold elution[6] and gold electrowinning[8], together with a relatively inexpensive method of thermal regeneration of the carbon, have all been demonstrated at pilot-plant level[9]. These developments in the recovery of gold from both solutions and pulps have undoubtedly made the use of activated carbon technically and economically more attractive. It was thus only a logical development to include the treatment of gold pregnant solutions as well as effluent solutions within the scope of the investigational work at these laboratories.

Materials

Test solutions

The adsorption of gold onto carbon from both synthetic and plant solutions was studied. Clarified pregnant solution obtained from the gold plant at Western Deep Levels Gold Mine was stored at the Laboratory in two 3,3 m^3 rubber-lined tanks. The composition of one such bulk solution, together with the effect of storage on its stability, is presented in Table I.

In view of the important role played by the ionic strength of the adsorption medium[10], two synthetic gold-bearing solutions, one of high ionic strength and the

TABLE I

COMPOSITION AND STABILITY OF GOLD-PLANT PREGNANT SOLUTIONS

Constituent	Concentration of solution as received (pH 10,6) g/t	Concentration of solution after storage for 1 month in sealed container (pH 9,1) g/t
Au	5,4	5,4
Ag	0,69	0,70
Cu	7	7
Ni	2	2
Co	< 1	< 1
Fe	1	< 1
Zn	15	3
Pb	< 1	< 1
Hg	0,01	0,01
Al	< 1	< 1
Ca	590	590
Ba	3	< 1
Mg	ND	< 1
Na	310	310
K	45	45
Si	20	12
CNS^-	90	100
$S_2O_3^=$	4	< 1
$SO_4^=$	1680	1660
Cl^-	240	250
NO_3^-	ND	190
As	ND	< 1
Free CN^-	92	19
Total CN^-	94	27
Suspended solids	5	ND

ND = Not determined.

other of low, having gold contents similar to the plant solution were also prepared. The synthetic solution of low ionic strength was made up from deionized water and CP-grade potassium aurocyanide, while the synthetic solution of high ionic strength was similarly prepared except that 1000 g/t calcium as calcium chloride was added to the solution.

Activated carbon

Type G210 and type G215 granular coconut-shell carbons supplied by Le Carbone (Pty) S.A. Ltd were used. Equilibrium capacity constants for gold were measured at pH 10 in standard borate buffer (see below). The values for the two grades of carbon were found to be 35 and 39 mg of gold per gram of carbon respectively.

Methods

Equilibrium capacity constants

The determination of equilibrium capacity constants for gold adsorption onto activiated carbon (type G210) at various pH levels was carried out as previously described[10]. The capacity constants were subsequently calculated on the assumption that the mechanism of gold adsorption was adequately described by the Freundlich equation over the entire pH range investigated. Any adjustment of solution pH was made with sulphuric acid. Prior to tests on the adsorption, samples of activated carbon were equlibrated at the desired pH level and then dried at 110 °C before use. The equilibration of the carbon samples was carried out in a closed circuit, the solution being circulated by means of a pump. The addition of the necessary amounts of sulphuric acid (10 per cent solution) was controlled by means of a titrimeter. This equilibration procedure was found to be necessary in order to prevent any upward drift in pH values following the addition of carbon to the test solutions.

Reaction rates

The effect of pH on the rate of gold adsorption from three different test solutions, all of which contained 6,2 g/t gold (see above), was compared. Gold adsorption as a function of time was monitored by the addition of 1 g of the equilibrated carbon (type G210) to 1 litre of test solution at the desired pH level while the solution was stirred constantly with the aid of a magnetic stirrer. Test samples of 5 ml each were withdrawn at frequent specified intervals (15, 30, 45, 60, 90, 130, 150, 180, 240, and 300 minutes), and the gold content was determined by atomic-absorption spectrophotometry.

Gold adsorption

Following acidification if required, gold-plant solution was pumped through a stainless-steel heat-exchange spiral immersed in a waterbath at 20 °C. The spiral was in series with three adsorption columns containing granular activated carbon (type G215). The influent solution was pumped at a constant flowrate using an MPL monopump metering pump. Unless otherwise specified, each column (16 mm internal diameter by 100 cm long) was packed with 38 g of a screened fraction of oven-dried granular activated carbon that had previously been equilibrated at the desired pH level. Flow through the columns was always in a downward direction. The effluent from each column was continuously monitored by means of an LKB Minirac fraction collector. Effluent from the first column was sampled every 45 minutes, samples of 20 ml being taken. Samples were taken every 30 minutes from columns 2 and 3 respectively. Gold values in the effluent solutions were analysed by solvent extraction followed by atomic-absorption spectrophotometry. A 10 ml aliquot was taken in the case of the first column, and the gold values were reported to two decimal figures; a composite 400 ml three-hour aliquot was taken from each of columns 2 and 3 respectively, and the gold values were reported to four decimal figures ($\pm$0,0005 g/t sensitivity).

A typical 'merry-go-round' operation was conducted, in which the lead column was taken off-line every 48 hours and eluted (see below) while an additional column containing virgin carbon was added in the last position. The initial loading of gold onto the lead column was subsequently calculated by measurement of the areas under the adsorption curves.

Gold elution

Gold-loaded carbon from the adsorption columns was transferred to a water-jacketed stainless-steel elution column (16 mm internal diameter by 75 cm long). Excess solution was allowed to drain from the column while the temperature of the column was raised to 91 °C. Unless otherwise specified, the carbon was then pretreated with one bed volume of a reagent containing 10 per cent sodium cyanide and 1 per cent sodium hydroxide, which was pumped into the column through a stainless-steel heat-exchange spiral also maintained at 91 °C. In the first continuous adsorption operation (see below), the pretreatment reagent was pumped at 2,83 bed volumes per hour. The pretreated carbon was allowed to 'soak' for 15 minutes, and the gold was eluted at 1,71 bed volumes per hour using deionized water[5]. In the second continuous adsorption experiment, the carbon was pretreated with only half a bed volume of the reagent of 10 per cent sodium cyanide and 1 per cent sodium hydroxide, and was eluted as before except that the 15-minute 'soaking' period was omitted. Eluate solutions containing relatively high concentrations of gold and silver were analysed by fire assay, while lower concentrations were analysed by solvent extraction followed by atomic-absorption spectrophotometry (reported to two decimal figures).

Results

Stability of gold-plant solutions

From the analytical data presented in Table I, it can be seen that the gold-plant solutions under study are not completely stable with respect to storage time. While no precipitation of gold or silver was indicated, the decrease in zinc, barium, total cyanide, and free cyanide, together with a drop in pH values, does indicate a certain degree of instability. The polymerization of free cyanide, together with the oxidation of cyanides to cyanates, may explain to some degree the reasons for such instability.

Because of the possible precipitation of aurous cyanide after the solution has been acidified, the precipitation of gold and silver from a plant solution was initially investigated as a function of pH. This was done by the addition of sulphuric acid to lower the pH of the test

solution to various levels between pH 1 and pH 3. The test samples were then allowed to stand for 24 hours before being filtered, and were analysed for gold and silver some 48 hours later. The results presented in Table II indicate no precipitation of gold even at a pH value as low as 1,0, but as much as 82 per cent of the contained silver was precipitated. The nature of the silver precipitate was not investigated, although it was most likely to be silver cyanide. At pH levels below approximately 4, the precipitation of a small amount of a pink slime was observed. Further analyses on an air-dried sample of the precipitate indicated the presence of 18 per cent copper, 8 per cent iron, 2,6 per cent silver, 1,1 per cent cobalt, 0,9 per cent nickel, and 26,3 per cent water. Examination of an X-ray diffractogram of the material showed that the precipitate was largely copper ferrocyanide, $Cu_2Fe(CN)_6$.

TABLE II

SOLUTION STABILITY OF GOLD AND SILVER IN CLARIFIED GOLD-PLANT SOLUTION

Solution pH	Acid addition ml H_2SO_4/l	Solution analysis* g/t Au	g/t Ag
9,8	as received	5,26	0,55
3,0	0,29	5,52	0,42
2,8	0,33	5,57	0,35
2,6	0,37	5,50	0,23
2,4	0,59	5,44	0,13
2,2	0,81	5,57	0,09
2,0	1,18	5,40	0,09
1,8	2,06	5,46	0,09
1,6	2,87	5,38	0,09
1,4	4,42	5,57	0,10
1,2	5,08	5,61	0,10
1,0	11,11	5,44	0,10

*Gold and silver determined by atomic-absorption specotrphotometry after solvent extraction.

Adsorption of gold onto carbon at conventional pH levels

Preliminary experiments were carried out in which both clarified pregnant solutions and partially clarified plant effluents were fed to a carbon column. This invariably resulted in excessive fouling of the carbon with calcium carbonate. The findings were similar in pilot-plant operations[9]. This carbonate fouling, besides adversely affecting the adsorption of gold, resulted in eventual choking of the carbon bed and necessitated frequent removal of the carbonate by treatment with dilute hydrochloric acid. The build-up of calcium carbonate was found to be erratic in that sometimes little or no carbon fouling occurred. Furthermore, fouling appeared to be most troublesome with fresh, rather than aged, plant solutions. These latter observations indicated that calcium fouling was most sensitive to the pH level of these solutions as the relatively small decrease in pH observed on storage (Table I) was sufficient to allay any fouling problem. Obviously, while the concentrations of both calcium and carbon dioxide play a significant role in this fouling mechanism, the judicious lowering of pH to between 8 and 9 (thus shifting the carbonate equilibrium to the bicarbonate side) would appear to overcome the problem to a large degree. Such a pH adjustment may have considerable merit if excessively alkaline plant solutions or pulps are to be treated with carbon, e.g. in a conventional carbon-in-pulp operation.

Apart from the problem of fouling by calcium carbon-

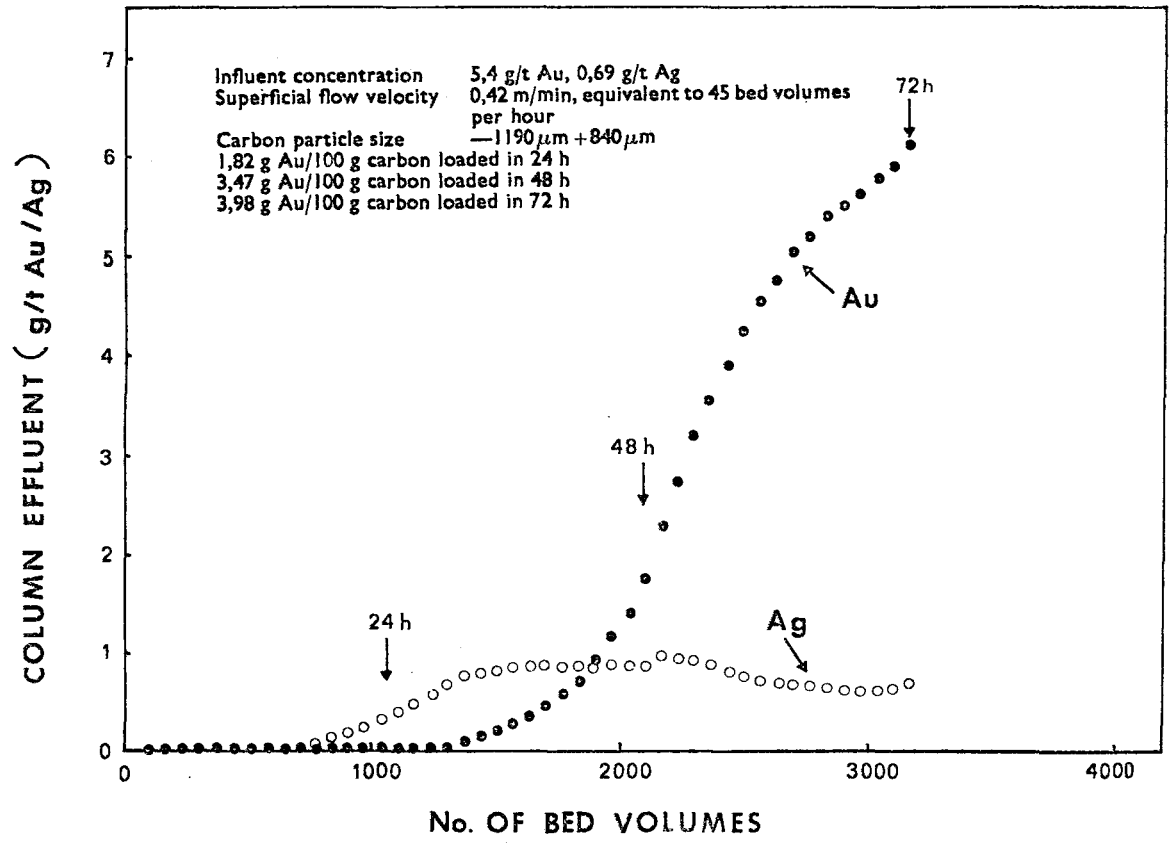

Fig. 1—The adsorption of gold and silver from a plant solution at pH 9,8

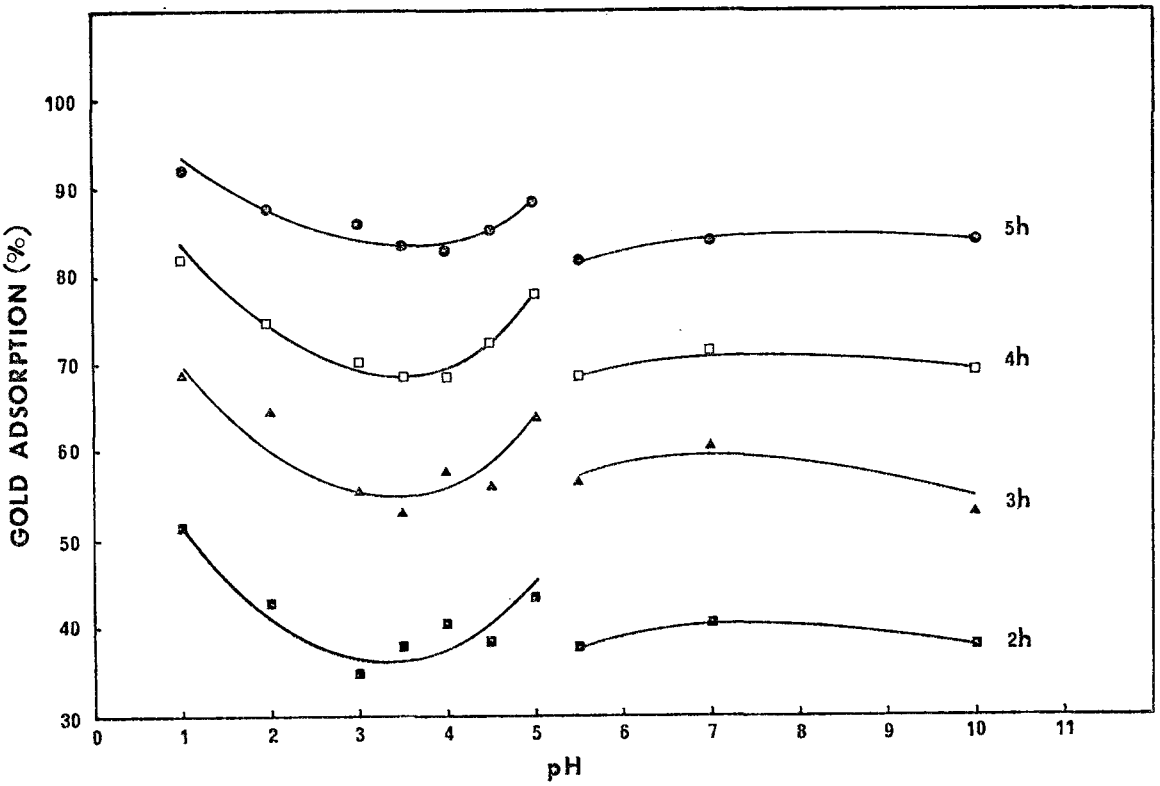

Fig. 2—The effect of pH on the rate of gold adsorption from a plant solution

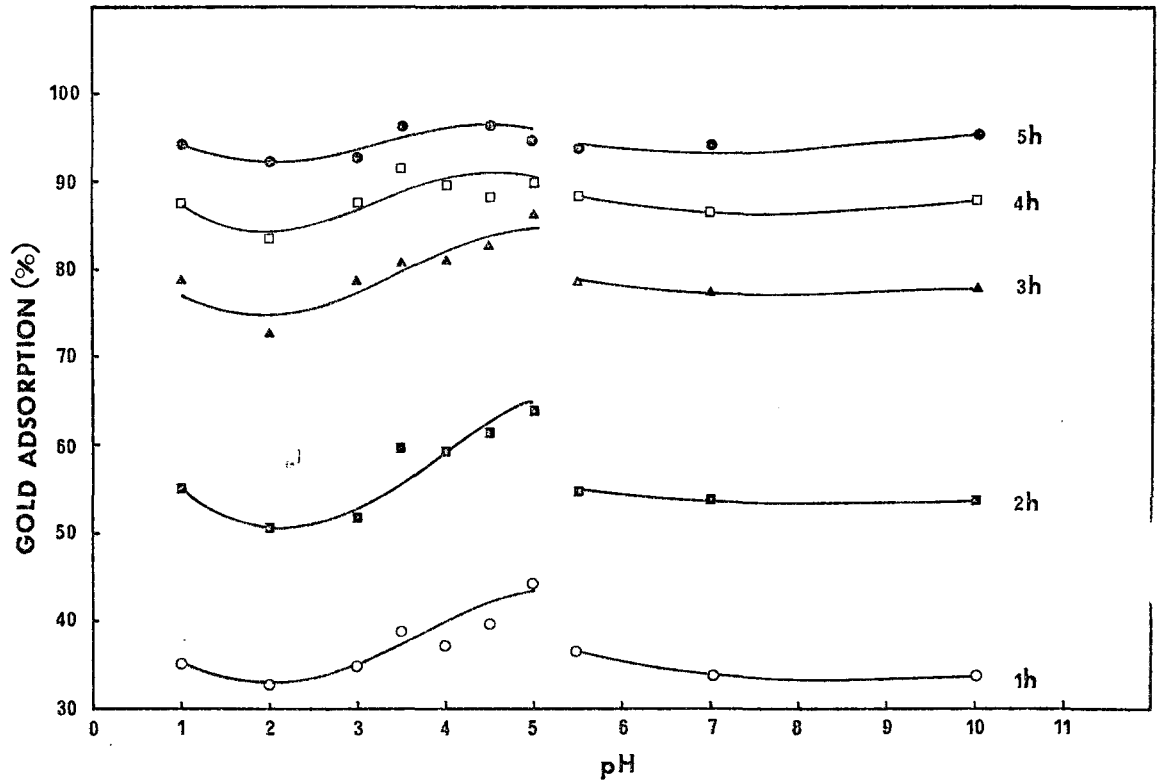

Fig. 3—The effect of pH on the rate of gold adsorption from a synthetic solution of high ionic strength

ate, the displacement of silver by influent gold values in a column operation presents problems in a merry-go-round system for the treatment of clarified pregnant solutions. Fig. 1 illustrates a typical breakthrough curve obtained when an 'aged' pregnant plant solution with a pH value of 9,8 was pumped through a column containing activated carbon. In this instance, fouling of the carbon by calcium carbonate was not observed. The early breakthrough of silver ahead of gold, together with the fact that the tenor of the silver in the effluent was higher than that in the influent, strongly suggests such a displacement mechanism. This displacement of silver, occurring at an average carbon loading of approximately 1 per cent gold in this instance, would result in a serious build-up of silver in a merry-go-round circuit, necessitating a more complex cycle of adsorption and elution if excessive losses of silver are to be avoided.

These two effects (i.e. the fouling of carbon with calcium carbonate and possible high silver losses) associated with the use of carbon at the pH levels occurring in a conventional plant were considered to be a sufficient inducement for further investigations to be conducted on the adsorption of gold at lower pH levels.

Effect of pH on the kinetics of gold adsorption

Adsorption data concerning the effect of pH on the rate of gold adsorption from the three test solutions studied are presented graphically in Figs. 2 to 4.

In the case of both the plant solution and the synthetic solution of high ionic strength, there are similarities in the rate of gold adsorption at various pH levels, although the adsorption is significantly better for the synthetic solution. In both cases, an optimum at pH 5 is indicated, together with a marked fall-off in the rate of gold adsorption above pH 5 (Figs. 2 and 3). However, gold adsorption from the synthetic solution of low ionic strength displays a very different kinetic picture (Fig. 4). In this instance, pH can be seen to have a far bigger effect on adsorption rates especially at high pH levels, with optima being indicated at pH levels of approximately 4 and below 1 respectively.

The reasons for the discrepancies noted between the rates of gold adsorption from the plant solutions and those from synthetic solutions are at this stage not easily defined. Both thiosulphate and thiocyanate anions were, however, shown to have a very detrimental effect on the rates of gold adsorption when added individually to a synthetic solution at the approximate concentration levels found in the plant solution (Table I). While the addition of 100 g/t thiocyanate as sodium thiocyanate was found to have a most detrimental effect at high pH levels, the presence of only 4 g/t thiosulphate added as sodium thiosulphate was found to have the most adverse effect on adsorption rates at low pH levels.

Effect of pH on equilibrium capacity constants

Capacity constants illustrating gold adsorption as a function of pH are presented graphically in Fig. 5. The data also allow comparison of gold adsorption from both a plant solution and a synthetic solution of high ionic strength. In the case of the plant solution, a very gradual increase in carbon capacity can be seen as the pH is decreased from 12 to 6. At lower pH levels, a sharp increase in capacity constant is indicated, with an optimum at approximately pH 2 to 2,5. The subsequent decline in gold capacity at still lower pH levels probably

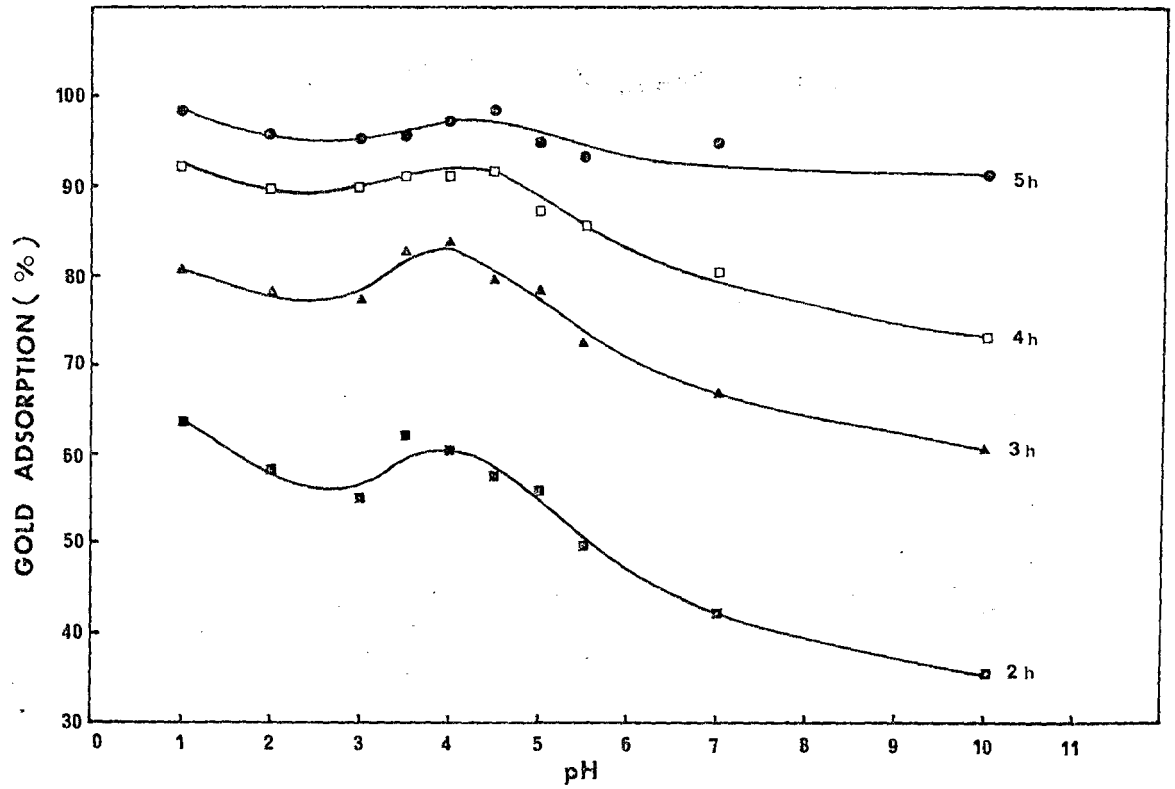

Fig. 4—The effect of pH on the rate of gold adsorption from a synthetic solution of low ionic strength

corresponds to the formation of aurous cyanide. The formation (and possible precipitation) of aurous cyanide would presumably interfere with any anionic adsorption mechanism.

The effect of pH on gold adsorption from the synthetic solution can immediately be seen to be similar to that from the plant solution. However, the capacity constants are generally very much better for synthetic than for plant solutions. Again, while the reasons for such differences are not easily explained, it was shown that the addition of both thiosulphate and thiocyanate to synthetic solutions had very detrimental effects on gold capacity constants. The addition of 100 g/t thiocyanate to a synthetic solution of high ionic strength at pH 5,0 was shown to reduce the capacity constant from 58 to 40 mg of gold per gram of carbon. Likewise, the addition of only 6 g/t thiosulphate reduced the capacity constant to 38 mg of gold per gram of carbon.

The effect of pH on the Freundlich constant[10], n, was also calculated from the above equilibrium adsorption

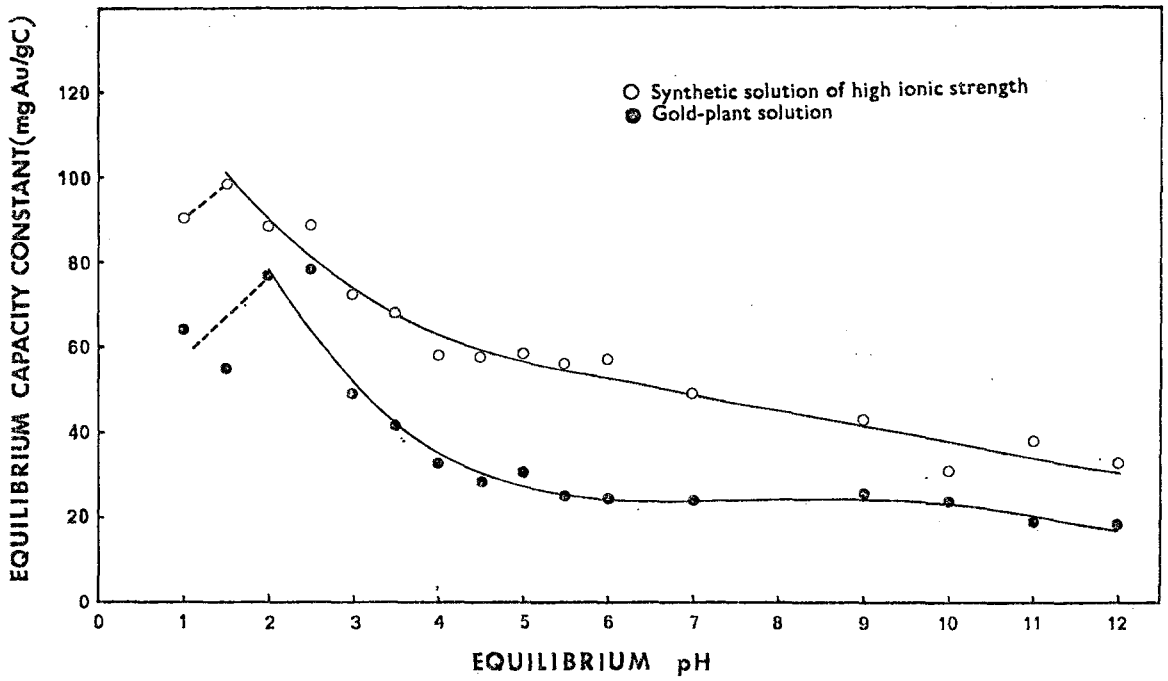

Fig. 5—The effect of pH on the Freundlich equilibrium capacity constant for gold adsorption onto activated carbon

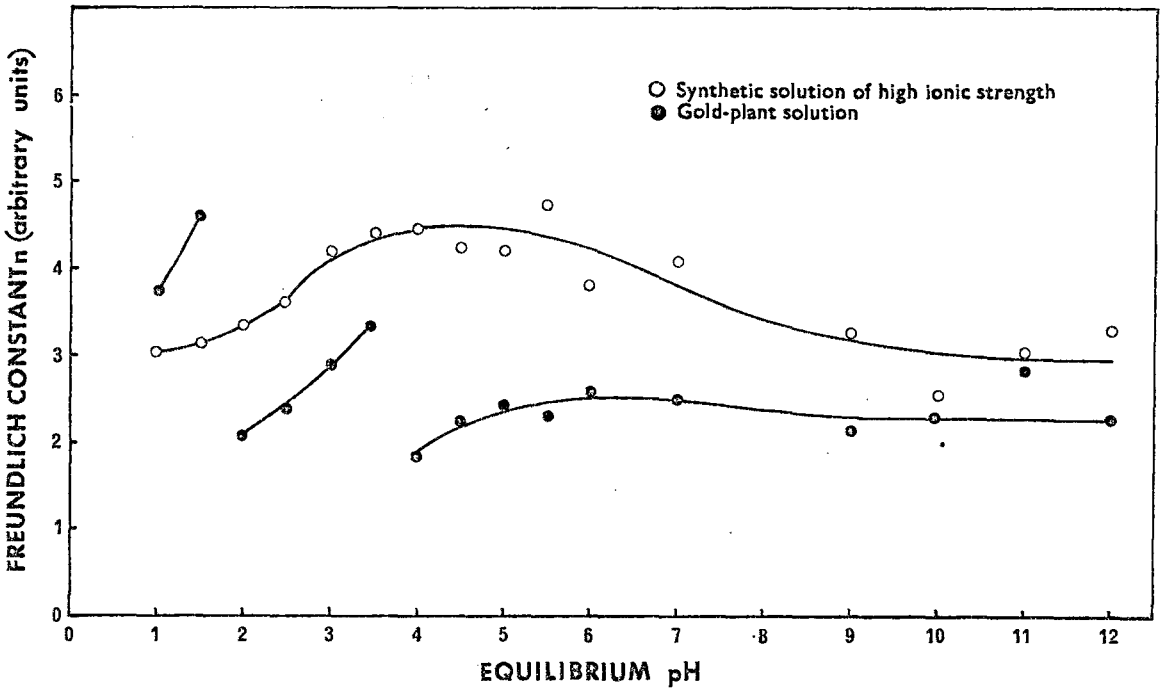

Fig. 6—The effect of pH on the Freundlich constant n

data (Fig. 6). While the behaviour of the plant solution is significantly different from that of the synthetic solution, indicating poorer gold adsorption, the breaks in continuity at pH levels below 4 and 2 are again evident. The optimum values of n were obtained when the synthetic solution was used at pH 4 to 5.

Gold adsorption at pH 5

Preliminary experiments were carried out in which acidified plant solutions at pH 2,5 were pumped through individual columns containing activated carbon. Acidification to this level entailed the addition of approximately 550 g/t sulphuric acid to the original plant solution. It was demonstrated that significantly higher gold loadings on carbon could be achieved at this influent pH. Loadings approaching 10 g of gold per 100 g of carbon were obtained when plant solution acidified to pH 2,5 was used. This compared with only 4 to 5 g of gold per 100 g of carbon achieved with a conventional plant solution (pH 10,2). It was also shown that higher influent flowrates could be used beneficially after acidification to pH 2,5. Furthermore, no displacement of silver by adsorbed gold values occurred at pH 2,5, although a significant amount of silver, together with copper ferrocyanide, was precipitated under such pH conditions (Table II). This precipitation resulted in a minor amount of carbon fouling, which increased the pressure drop across the carbon bed. The carbon fouling also presented operational problems when in-line acidification was carried out. In view of this latter problem, together with the earlier findings suggesting adsorption reaction rate optima in the region of pH 5 (see reaction rate data), further investigations were confined to the treatment of acidified plant solutions in this particular pH region. Acidification of plant solution to pH 5 entailed the addition of only 220 g/t sulphuric acid. While no precipitation of copper ferrocyanide or silver was observed at this pH level, a small amount of a white precipitate identified as zinc ferrocyanide was noted. Fig. 7 illustrates a typical partial breakthrough curve for gold and silver values, which was obtained when an acidified plant solution at pH 5 and containing 6,2 g/t gold was pumped through a single column as previously described. The sympathetic breakthrough of both gold and silver under such conditions can be compared with the premature displacement of silver that was observed when a non-acidified plant solution was used. (Fig. 1).

In view of the possible large hold-up of bullion in an envisaged carbon circuit, the parameters most likely to influence such a situation were also briefly considered. Fig. 8 illustrates the effect of the particle size of the carbon on the adsorption of gold when suitably acidified plant solution was pumped through individual columns containing two different screen fractions of type G215 carbon. The beneficial use of the finer fraction of carbon (that between 1190 and 840 μm) is most evident. In this instance, influent flowrates were maintained at approximately 45 bed volumes per hour while influent was pumped continuously for 72 hours. Even at such relatively high flow velocities, no excessive pressure drop across the carbon bed was observed in either instance (less than 20 kPa per metre of carbon bed). Testwork using a fraction between 840 and 297 μm of a similar grade of carbon, while indicating even better gold adsorption, showed a pressure drop of about 100 kPa per

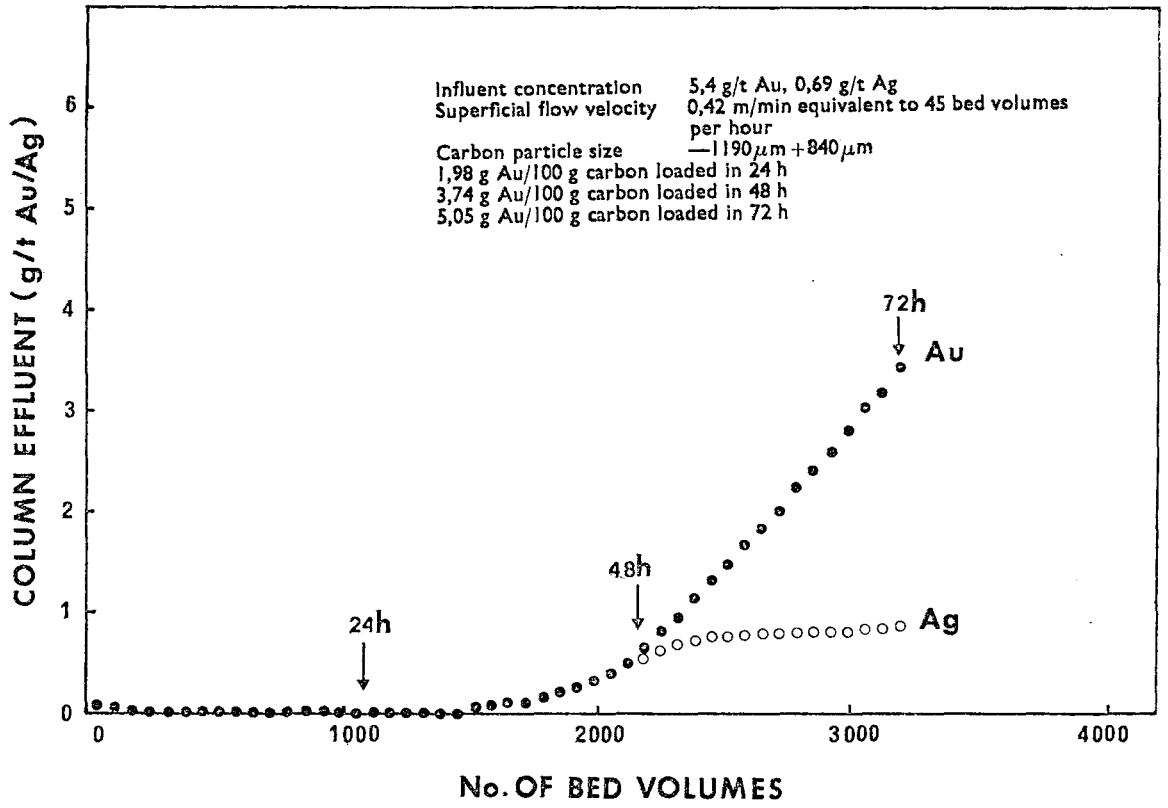

Fig. 7.—The adsorption of gold and silver from an acidified plant solution at pH 5,0

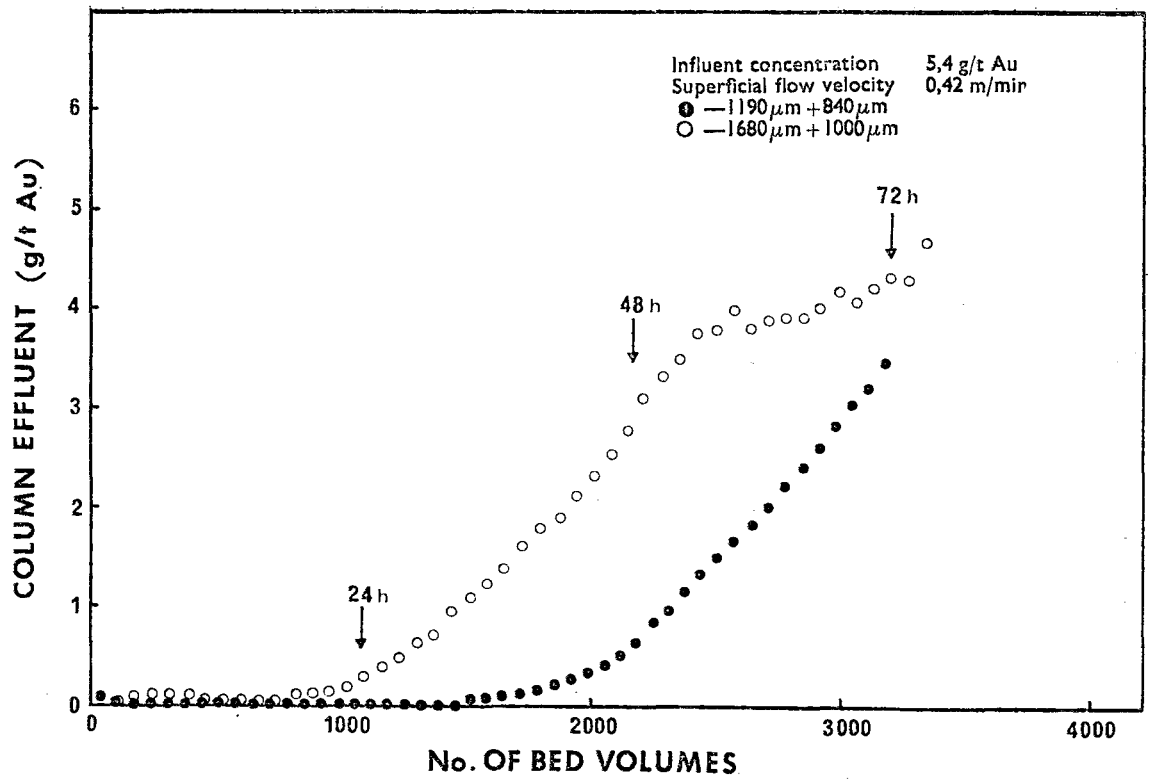

Fig. 8—The effect of carbon particle size on gold adsorption

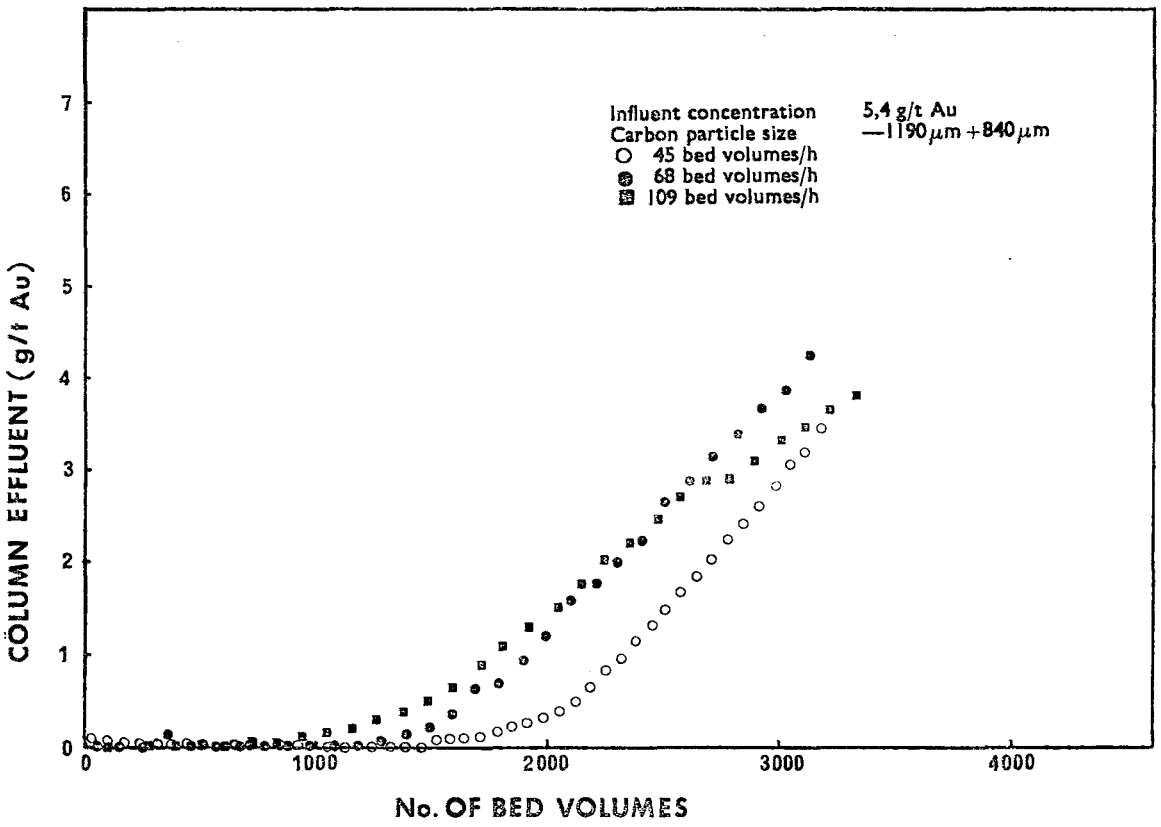

Fig. 9—The effect of influent flowrates on gold adsorption

metre of carbon bed across the carbon bed. This was considered to be excessive from the point of view of any possible scale-up in the size of the operation. Further testwork was thus confined to the use of a narrow size range (between 1190 and 840 μm) of type G215 carbon. The choice of a relatively narrow range of particle sizes of carbon has considerable operational advantages when related to packed-bed column operation. This is because it results in lower packing densities, and consequently lower pressure drops. The bulk density of the above selected carbon fraction was approximately 0,33 g/cm^3, which can be compared with packing densities of up to 0,5 g/cm^3 for carbon of a similar grade but wider particle-size range.

The effect of influent flowrates on gold adsorption was also investigated in the present context. As can be expected, the use of relatively high influent flowrates will minimize any hold-up of bullion in the carbon-adsorption circuit. Fig. 9 illustrates the effect of pumping suitably acidified plant solution through individual columns containing activated carbon. Superficial influent flow velocities varying from 0,42 to 1,05 m/min were employed. From the results presented, it is most evident that only small differences in the gold breakthrough curves resulted when flowrates were increased from 45 to 109 bed volumes per hour. These observations suggest that the rates of gold adsorption must be extremely high and the adsorption reaction zone very narrow. In the case where the influent was pumped at 109 bed volumes per hour, a loading of 4,16 g of gold per 100 g of carbon in 24 hours was noted. It was also observed that little, if any, displacement of silver by influent gold values occurred under the prescribed conditions.

Displacement of adsorbed metal values

Gold-plant solution having a pH of 9,3 and containing 5,4 g/t gold was pumped through a column at a superficial flow velocity of 0,42 m/min for 24 hours. Thereafter the entire column was carefully cut into sections and individual sections were analysed. A similar operation was carried out using the same gold-plant solution acidified to pH 4,5. Figs. 10 and 11 illustrate the values of adsorbed metal obtained at the two different pH levels. In the case of the conventional plant solution (Fig. 10), the displacement sequence gold–silver–nickel–copper was clearly evident, while the concentrations of calcium built up to nearly 1 per cent at the top of the column. Conversely, adsorption carried out under acid conditions (Fig. 11) indicated no such displacement sequence, although a major amount of copper was adsorbed at the top of the column. This adsorbed copper complex, which was thought to be $Cu(CN)_2^{-1}$, may explain to a large degree the disparity between acidified plant solutions and synthetic solutions under similar pH conditions (see above).

It has been suggested[11] that the increase in copper adsorption is related to the negative charge on the corresponding anionic complex. Thus, at conventional pH levels, where the $Cu(CN)_4^{3-}$ complex predominates, the adsorption of copper is poor and the copper is easily displaced by more strongly adsorbed metal species (see above). Conversely, at lower pH levels, decomposition of the above complex, yielding initially $Cu(CN)_3^{2-}$ and subsequently $Cu(CN)_2^{-1}$, results in a substantial increase in copper adsorption owing to the lower anionic charge on the copper complex. Furthermore, the observed precipitation of copper as a pink ferrocyanide complex at pH

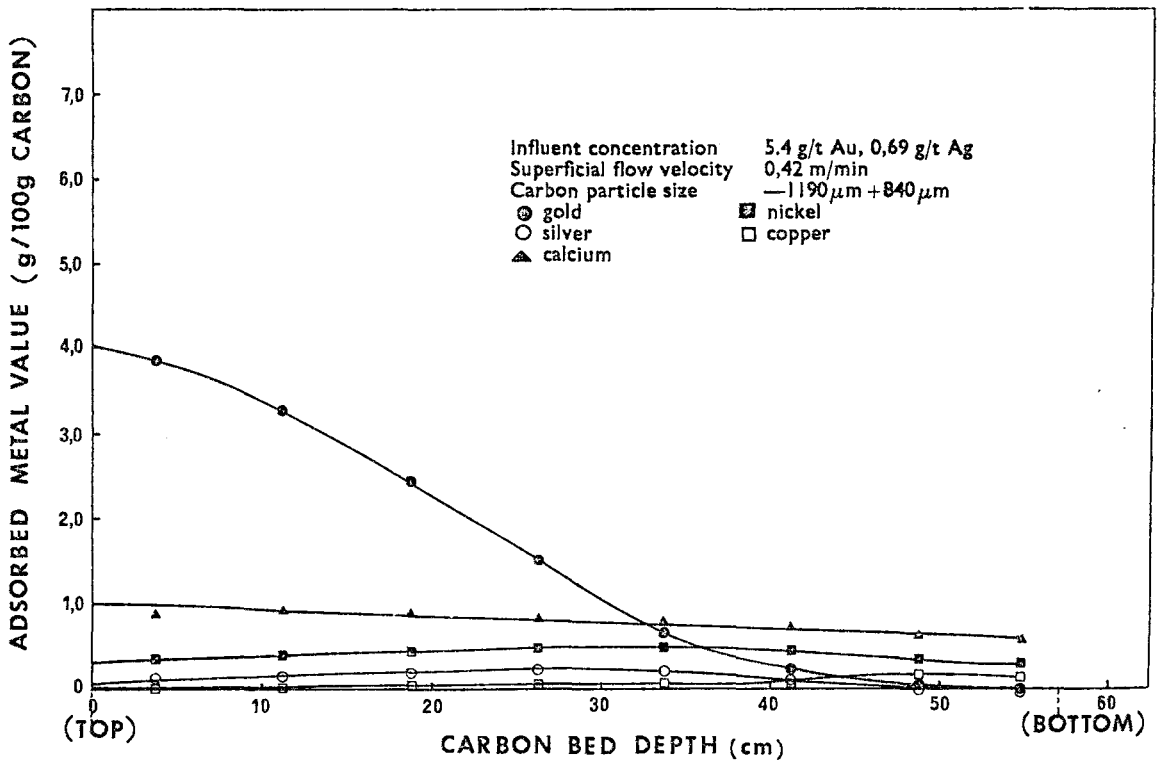

Fig. 10—The adsorption of metal values from a plant solution at pH 9,3

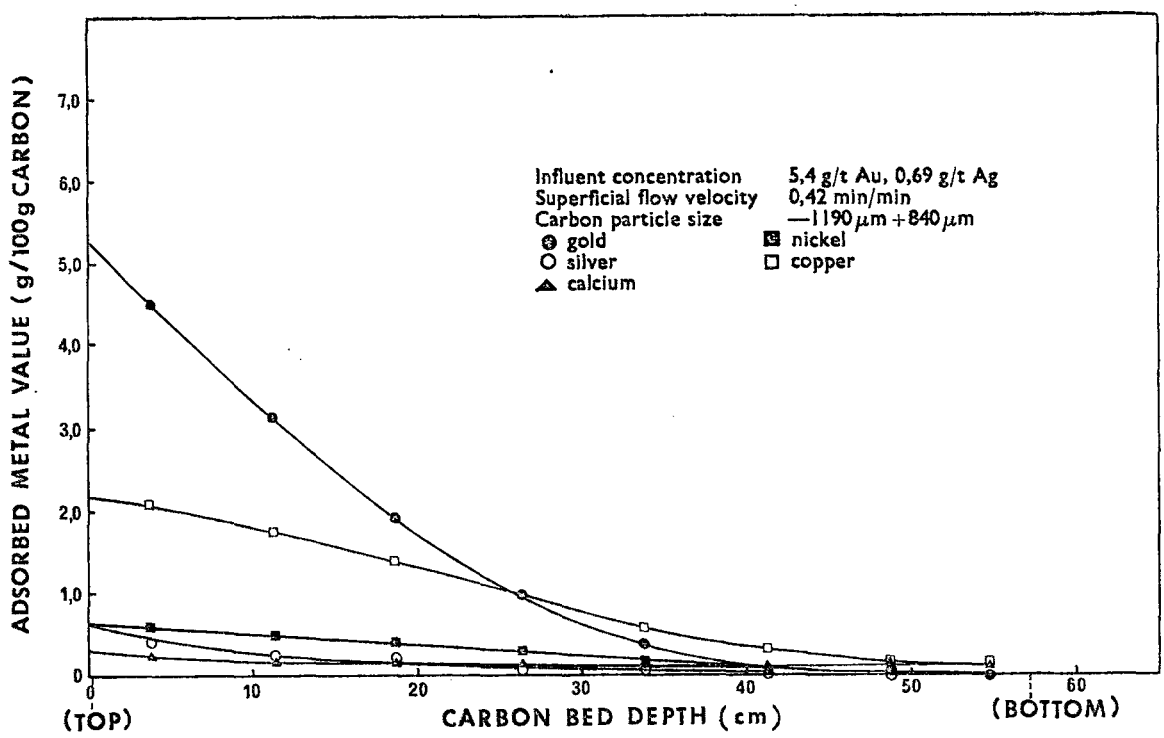

Fig. 11—The adsorption of metal values from an acidified plant solution at pH 4,5

TABLE III

OPERATION OF A 'MERRY-GO-ROUND' ACTIVATED-CARBON ADSORPTION CIRCUIT AT AN INFLUENT FLOWRATE OF 45 BED VOLUMES PER HOUR

(Superficial flow velocity = 0,42 m/min)

Column sequence	Total number of bed volumes passed/operational time, h	E uent from second column (daily average concentrations)		Effluent from last column (daily average concentrations)	
		g/t Au	g/t Ag	g/t Au	g/t Ag
A — B — C	1059/0–24	0,0018	< 0,001	0,0022	< 0,002
A — B — C	2117/24–48	0,0008	0,006	0,0008	< 0,001
B — C — D	3176/48–72	0,0013	< 0,001	0,0007	< 0,001
B — C — D	4234/72–96	0,0033	0,010	0,0040	< 0,001
C — D — E	5293/96–120	0,0010	< 0,001	0,0010	< 0,001
C — D — E	6351/120–144	0,0014	0,022	0,0019	< 0,001
Average effluent values for 6 days		0,0016	< 0,006	0,0018	< 0,001

levels below 4 may explain the discontinuities observed in gold adsorption rates and capacity constant data under such acidic conditions. Likewise, the discontinuities observed at pH 1 to 2 may coincide with the formation of aurous cyanide. Both of these aspects are well illustrated in Figs. 5 and 6.

Continuous gold adsorption at an influent flowrate of 45 bed volumes per hour

Acidified gold-plant solution at pH 5,0 containing 5,4 g/t gold was pumped at a flowrate of 45 bed volumes per hour through a simulated merry-go-round circuit for a period of 6 days (continuous pumping was suspended temporarily after 96 hours owing to a weekend shutdown). A summary of the results is presented in Table III, while the elution data for column B are shown in Table IV. Gold-adsorption data for the first column of the merry-go-round circuit are presented graphically in Fig. 12.

With regard to gold adsorption, average values of gold in the effluent over the 6 days of 0,0016 g/t and 0,0018 g/t from columns 2 and 3 respectively are most noteworthy. Such values also indicated that column C was in fact doing little or no work with respect to gold adsorption. A gold loading of 3,2 g per 100 g of carbon was indicated on the lead column (column A) using the adsorption data presented in Table III. However, elution data indicated progressive gold loadings of 3,3, 3,6, and 3,7 g per 100 g of carbon on columns A, B, and C respectively as these were brought into the lead position.

Fig. 13 illustrates a typical elution profile of gold and silver values from columns of loaded carbon. It shows that virtually complete gold elution was effected in 6 to 7 bed volumes (elution data taken from column C). Analyses of samples of the eluted carbon indicated that 99,9 per cent of the gold and 99,8 per cent of the silver was eluted under the specified conditions. The elution data also indicated the adsorption of considerable concentrations of copper (3,1 to 3,9 g per 100 g of carbon), zinc (1 to 1,3 g per 100 g of carbon), and nickel (0,9 to 1,1 g per 100 g of carbon). The elution of rela-

TABLE IV

ELUTION OF LOADED CARBON FOLLOWING A 48h ADSORPTION CYCLE AT 45 BED VOLUMES PER HOUR

ELUTION OF COLUMN B

Constituents	No. of bed volumes of eluate (1 bed volume = 166,6 ml)										Total mass mg	Carbon loading* g/110g C
	0,523	1,402	2,251	3,100	3,953	4,811	5,677	6,522	7,367	8,25		
Au (g/t)	348	894	4980	5196	2046	488	51	5,80	0,77	0,25		
Au (mg)	21,23	91,64	493,02	514,40	203,58	44,80	5,15	0,57	0,08	0,02	1374	3,62
Ag (g/t)	77	735	312	6	0	0	0,28	0,16	0,12	0,07		
Ag (mg)	4,70	75,34	30,89	0,59	0,00	0,00	0,03	0,02	0,01	0,01	111	0,29
Cu (g/t)	5100	11200	200	3	< 1	< 1	< 1	< 1	< 1	< 1		
Cu (mg)	311,1	1148,0	19,89	0,3	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 1479	3,89
Co (g/t)	3	2	< 1	< 1	< 1	< 1	< 1	< 1	< 1	< 1		
Co (mg)	0,2	0,2	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1		
Ni (g/t)	870	2660	860	19	< 1	< 1	< 1	< 1	< 1	< 1		
Ni (mg)	53,1	272,7	85,1	1,9	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	413	1,09
Fe (g/t)	22	44	6	< 1	< 1	< 1	< 1	< 1	< 1	< 1		
Fe (mg)	1,3	4,5	0,6	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1		
Zn (g/t)	1160	3940	95	1	< 1	< 1	< 1	< 1	< 1	< 1		
Zn (mg)	70,8	403,9	9,4	0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	484	1,27
Ca (g/t)	720	200	21	10	9	7	7	7	9	7		
Ca (mg)	43,9	20,5	2,1	1,0	0,9	0,7	0,7	0,7	0,9	0,7	72	0,19
SiO_2 (g/t)	35	22	320	210	120	90	95	75	80	75		
SiO_2 (mg)	2,1	2,3	31,7	20,8	11,9	9,0	9,6	7,4	7,9	7,7	100	> 0,29
S_2O_3 (g/t)	ND	1200	< 1	25	0,7	4,5	1,3	3,1	5,4	ND		
CNS^- (g/t)	ND	470	90	10	5,0	4,8	2,8	1,2	0,3	0,7		

ELUTION RECOVERY (COLUMN B)

Element	Carbon loading g/t	Eluted carbon g/t	Recovery %
Gold	36200	54	99,9
Silver	2900	81	97,2

*On the assumption of complete elution except for Au, Ag, and SiO_2

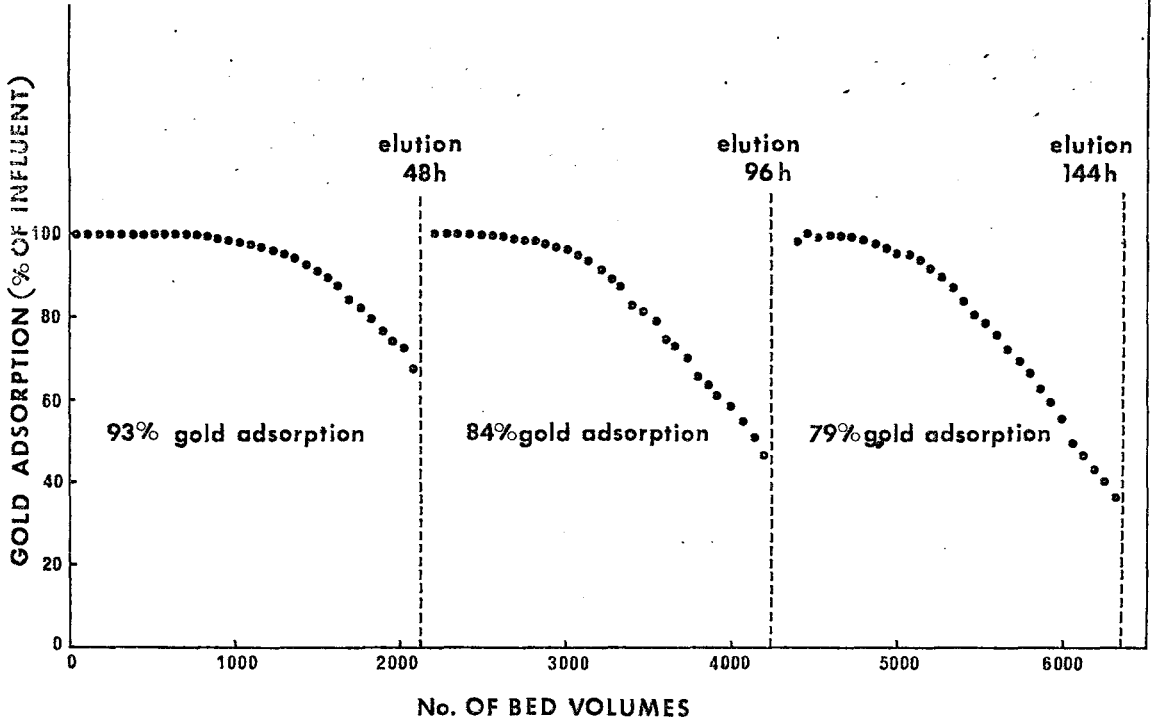

Fig. 12—The adsorption of gold in the first column of a merry-go-round circuit operating at 45 bed volumes per hour

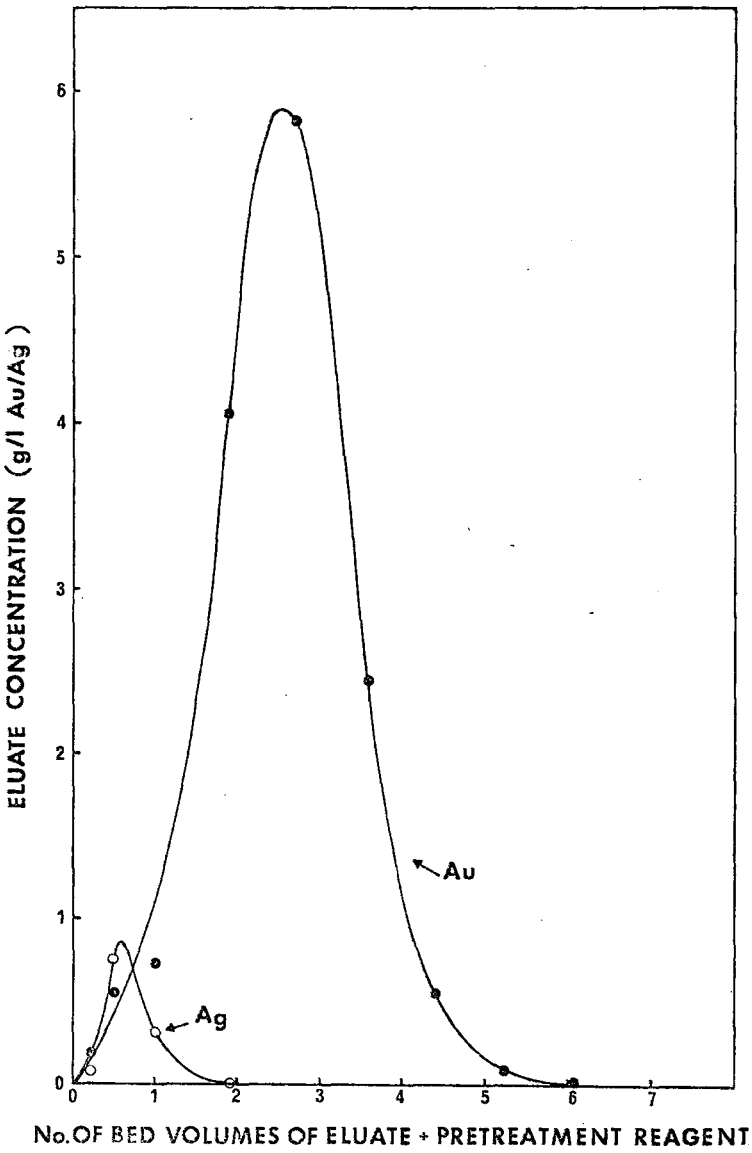

Fig. 13—The elution of gold and silver from loaded carbon using deionized water as an eluant

tively high concentrations of thiocyanate and thiosulphate complexes, together with low concentrations of iron and cobalt, are also noteworthy.

The increase in the pH of column effluent from an initial influent value of 5,0 to a final effluent value of approximately 6,0 was an interesting observation. The effluent from the first column was also found to have a pH of approximately 6,0, implying that the rise in pH occurred where the adsorption of gold was greatest. Ion exchange, or replacement of basic groups on the carbon surface by aurocyanide anions, is strongly suggested.

Continuous gold adsorption at an influent flowrate of 66 bed volumes per hour

In order to achieve even higher gold loadings on the lead column and at the same time to make more effective use of the last column, influent flowrates were increased from 45 to 66 bed volumes per hour and the previous operation (see above) continued. Also, in this operation the influent pH was lowered to a value of 4,3 because of the significant rise in pH observed on the lead column.

From the summary of the resulting adsorption data presented in Table V, average gold effluent values over the 6 days of operation of 0,0008 g/t again indicated a gold adsorption efficiency of more than 99,9 per cent. Likewise, the adsorption of silver was also exceptionally good (more than 99,9 per cent). The adsorption of gold in the first column of the circuit is graphically presented in Fig. 14, and typical elution data from column F are presented in Table VI. While average gold recoveries of 99,7 per cent and silver recoveries of 99,6 per cent were achieved in the present operation, carbon loadings approaching the following were also obtained: 53 kg/t gold, 6 kg/t silver, 57 kg/t copper, 14 kg/t nickel, and 1 kg/t zinc.

Again, the elution of considerable quantities of thiosulphate and thiocyanate complexes together with small amounts of mercury (0,002 g per 100 g of carbon loading)

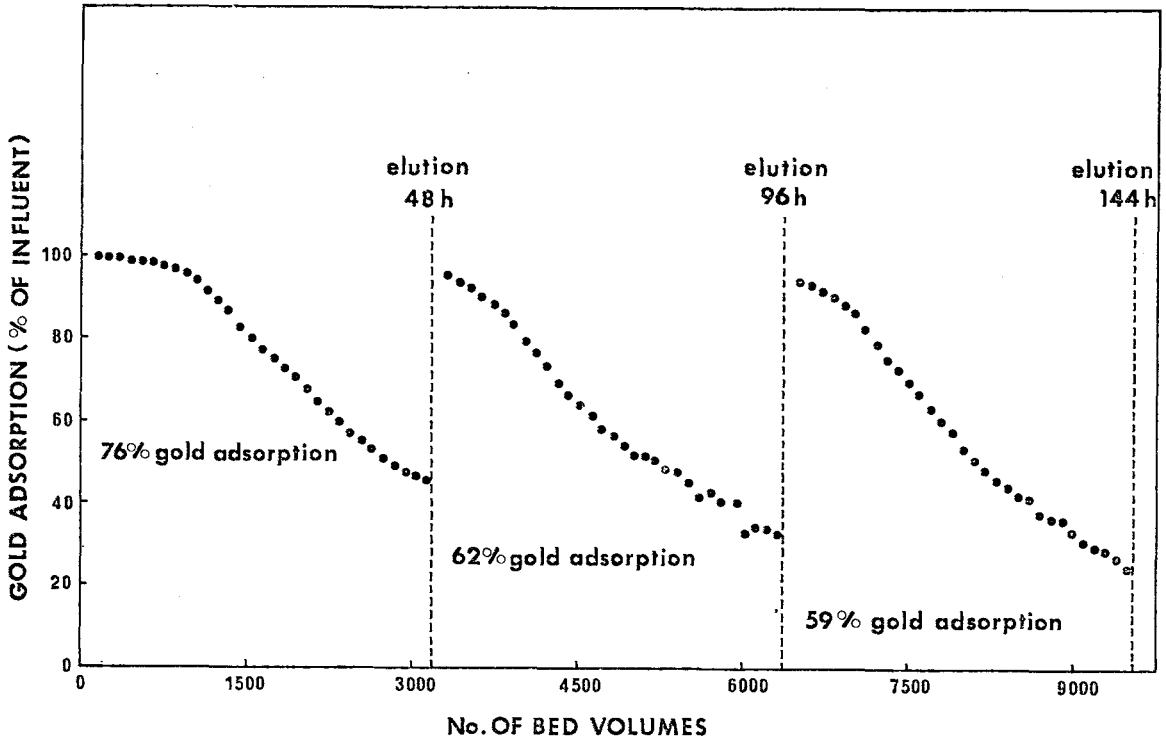

Fig. 14—The adsorption of gold in the first column of a merry-go-round circuit operating at 66 bed volumes per hour

TABLE V

THE OPERATION OF A MERRY-GO-ROUND ACTIVATED-CARBON ADSORPTION CIRCUIT AT AN INFLUENT FLOWRATE OF 66 BED VOLUMES PER HOUR

(Superficial flow velocity = 0,64 m/min)

Column sequence	Total number of bed volumes passed/operational time, h	E uent from second column (daily average concentration)		Effluent from last column (daily average concentration)	
		g/t Au	g/t Ag	g/t Au	g/t Ag
D — E — F	1590/0–24	0,0020	< 0,001	0,0015	< 0,001
D — E — F	3180/24–48	0,0220	0,014	0,0005	< 0,001
E — F — G	4770/48–72	0,0024	0,005	0,0005	< 0,001
E — F — G	6360/72–96	0,0709	0,032	0,0005	< 0,001
F — G — H	7950/96–120	0,0034	0,002	0,0008	< 0,001
Average eOuent values for 6 days		0,0279	0,016	0,0008	< 0,001

TABLE VI

THE ELUTION OF LOADED CARBON FOLLOWING A 48h ADSORPTION CYCLE AT 66 BED VOLUMES PER HOUR

ELUTION OF COLUMN D

Constituents	No. of bed volumes of eluate (1 bed volume = 166,6 ml)									Total mass mg	Carbon loading* g/100g C
	0,600	1,483	2,341	3,199	4,095	4,953	5,811	6,669	7,591		
Au (g/t)	308	1701	7134	6147	2415	764	194	33	5,70		
Au (mg)	21,56	175,20	713,40	614,70	252,37	76,40	19,40	3,30	0,61	1877	4,94
Ag (g/t)	79	1598	1104	64	69	10	17	0,12	0,10		
Ag (mg)	5,53	164,59	110,40	6,40	7,21	1,00	1,70	0,01	0,01	297	0,78
Cu (g/t)	1800	20200	130	11	2	1	1	1	< 1		
Cu (mg)	126,0	2080,6	13,0	1,1	0,2	0,1	0,1	0,1	< 0,1	2221	5,84
Ni (g/t)	1900	5300	150	1	< 1	< 1	< 1	< 1	1		
Ni (mg)	133,0	545,9	15,0	0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	694	1,83
Fe (g/t)	12	20	1	< 1	< 1	< 1	< 1	< 1	< 1		
Fe (mg)	0,8	2,1	0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	3	0,01
Zn (g/t)	9	420	10	1	< 1	< 1	< 1	< 1	< 1		
Zn (mg)	0,6	43,3	1,0	0,1	< 0,1	< 0,1	< 0,1	< 0,1	< 0,1	45	0,12
Ca (g/t)	340	110	10	2	2	2	2	4	5		
Ca (mg)	23,8	10,7	1,0	0,2	0,2	0,2	0,2	0,4	0,5	37	0,10
SiO_2 (g/t)	120	350	110	70	70	75	50	45	50		
SiO_2 (mg)	8,4	34,0	11,0	7,0	7,0	7,5	5,0	4,5	5,4	89	> 0,23
Hg (g/t)	0,43	5,46	< 0,01	< 0,01	< 0,01	< 0,01	< 0,01	0,03	0,02		
Hg (mg)	0,030	0,001	< 0,001	< 0,001	< 0,001	< 0,001	< 0,001	0,003	0,002	0,60	0,002
$S_2O^{2-}_3$ (g/t)	560	< 1	33	28	17	24	13	6,2	8,9		
CNS^- (g/t)	4650	4720	22	7,2	8,1	5,3	3,1	4,3	2,1		

ELUTION RECOVERY (COLUMN D)

Element	Carbon loading (g/t)	Eluted carbon (g/t)	Recovery (%)
Gold	49400	40	99,9
Silver	7800	33	99,6

*On the assumption of complete elution except for Au, Ag, and SiO_2

196.

is noteworthy. When these carbon loadings are compared with earlier loadings (see above), the marked decrease in zinc concentration (from 13 kg/t to 1 kg/t) would suggest that only zinc has been replaced on the carbon surface to any significant degree by other adsorbed metal species.

Carbon fouling

It had been established previously[9] (in experiments similar to those described above) that little or no build-up of metal values or silica on activated carbon results from the repeated use of the carbon in treating conventional plant solutions. However, there remains the possibility of incurring a build-up of calcium sulphate in the present instance. Calcium analyses of the eluted carbon samples are presented in Table VII. These, however, suggest that while a little fouling by calcium sulphate may take place when the pH of the influent solution is 5,0, little or no fouling takes place when the influent pH is lowered to 4,3. This marginal lowering of the influent pH from 5,0 to 4,3, resulting in a considerable decrease in the calcium content of the eluted carbon, may correspond to the formation of soluble calcium bisulphate, rather than the more insoluble calcium sulphate. This small build-up of calcium can be significant and would require careful monitoring in any future investigation.

TABLE VII

THE DETERMINATION OF CALCIUM ON SAMPLES OF ELUTED CARBON

Column sample	Influent pH	Calcium content, g/t
Original carbon equilibriated at pH 5,0		500
A	5,0	4180
B	5,0	4200
C	5,0	4580
D	4,3	1450
E	4,3	1630
F	4,3	2370

Discussion

The results presented indicate that the effect of pH on the rate of gold adsorption and on the capacity of activated carbon to adsorb gold is certainly more complex than originally expected. The similarity between the plant solution and the synthetic solution of high ionic strength may have been anticipated, but the rate optima displayed at pH 5 and below 1 are difficult to explain in terms of a single adsorption mechanism. Likewise, the capacity constant optima are difficult to explain, especially when it is seen that those pH values at which exceptionally high capacity constants were obtained (in the region of pH 3,0 to 3,5) correspond to the pH levels at which exceptionally low rates of gold adsorption were observed. The differences in reaction rates displayed by the plant solution and the synthetic solution of high ionic strength are also noteworthy. The reasons for such a disparity are at this stage only speculative, and further investigations in this direction would appear to be warranted. The increased adsorption of copper as a competing anion at a lower pH level may to some degree explain certain of these observed differences.

The present investigation has demonstrated that significant improvements can be made in the use of activated carbon for the recovery of gold and silver from gold-plant solutions (containing relatively low concentrations of copper). First, the carbon fouling that was observed both at high pH (due to calcium carbonate) and at low pH (due to copper ferrocyanide slime) is not evident when plant solutions are acidified to pH values between 4 and 5. This aspect, besides increasing the effectiveness of carbon with respect to gold adsorption, has significant operational advantages. Second, significant savings are effected in the sulphuric acid required to acidify to between pH 4 and 5 (and later the lime required to neutralize effluent to pH 8 or higher), when compared with acidification to lower pH values. Acidification to pH 5,3 was found to require the addition of 0,220 kg/t of sulphuric acid (0,55 cent per cubic metre of solution at a sulphuric acid cost of R25 per ton), while adjustment to pH 2,5 required 0,575 kg/t sulphuric acid (1,44 cents per cubic metre of solution). Such costs could be roughly doubled when the subsequent neutralization costs are included. These cost savings resulting from operation at pH 4 to 5 could be made without the sacrifice of gold-adsorption rates. Third, the efficient recovery of silver when adsorption takes place at pH 4 to 5 is a distinct advantage. The recovery of silver at current prices is worth approximately 10 cents per ton of ore treated, and it is therefore important for the silver to be recovered as completely as possible.

When compared with conventional zinc precipitation, the carbon process operated at pH 4 to 5 would appear to have considerable economic merit. If the gold in average plant effluents could be reduced from between 0,01 and 0,02 g/t (conventional circuit) to 0,001 g/t, a saving of 8 to 16 cents per ton of ore treated would result if the gold price is taken as R5000 per kilogram. This alone would correspond to between R16 000 and R32 000 per month on a mine treating 200 000 tons per month. The major operating costs in a carbon circuit of the type described would involve respectively the acidification and neutralization of the plant solutions before and after they were contacted with carbon. This cost is not likely to exceed 2 cents per ton of ore treated. Furthermore, both the lower capital investment for a carbon plant and the lower operating costs would appear to favour such an operation above conventional zinc precipitation. A preliminary costing exercise carried out at the AARL indicated a capital saving of approximately R380 000 for a mine treating 264 000 tons per month, while operating costs were reduced by R210 000 per annum.

From a practical point of view, the present findings indicate a very simple process for the recovery and electrowinning of gold from acidified gold-plant pregnant solutions. Whereas the adsorption of gold from alkaline solutions entails relatively slow adsorption kinetics, resulting in low carbon loadings and longer adsorption cycles, together with calcium carbonate fouling of the carbon, the reverse is true when adsorption is carried out at pH 4 to 5. Not only will very little hold-up of gold occur in such a process, but far more use is made of very much less carbon in a smaller carbon circuit.

Further investigations related to the re-use of the carbon in a continuous merry-go-round circuit, together with more exhaustive modelling exercises, would be required before an economically meaningful case could

be made for the use of carbon rather than zinc for recovering gold in an existing gold plant.

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Acknowledgements

The authors are grateful to the Management and to their colleagues at the Anglo American Research Laboratories for encouragement and numerous critical discussions. Particular indebtedness is acknowledged to the Analytical Laboratory for their very considerable and meticulous efforts, and to the Director of Research, Mr G. S. James, for his abiding interest in the whole project.

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Loraine Gold Mines Limited.
Marievale Consolidated Mines Limited.
Matte Smelters (Pty) Limited.
Natal Cambrian Collieries Limited.
Northern Lime Co. Limited.
O'okiep Copper Company Limited.
Otjihase Mining Co. (Pty) Limited.
Palabora Mining Co. Limited.
Photometric Sorters.
President Steyn G.M. Co. Limited.
Pretoria Portland Cement Co. Limited.
Prieska Copper Mines (Pty) Limited.
Rand Mines Limited.
R. J. Spargo Limited.
Rooiberg Minerals Development Co. Limited.
Rustenburg Platinum Mines Limited (Union Section).
Rustenburg Platinum Mines Limited (Rustenburg Section).
St. Helena Gold Mines Limited.
Shaft Sinkers (Pty) Limited.
S.A. Land Exploration Co. Limited.
Stilfontein G.M. Co. Limited.
The Griqualand Exploration and Finance Co. Limited.
The Messina (Transvaal) Development Co. Limited.
The Randfontein Estates Gold Mining Co. Witwatersrand Ltd.
The Robbins Co. (Africa) (Pty) Ltd.
The Steel Engineering Co. Ltd.
Trans-Natal Coal Corporation Limited.
Tvl Cons. Land & Exploration Co.
Tsumeb Corporation Limited.
Union Corporation Limited.
Vaal Reefs Exploration & Mining Co. Limited.
Venterspost G.M. Co. Limited.
Vergenoeg Mining Co. (Pty) Limited.
Vlakfontein G.M. Co. Limited.
Welkom Gold Mining Co. Limited.
West Driefontein G.M. Co. Limited.
Western Areas Gold Mining Co. Ltd.
Western Deep Levels Limited.
Western Holdings Limited.
Winkelhaak Mines Limited.

PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 78/5147

Date of Filing: 9 AUGUST 1979

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventor: 1. RAYMOND JOHN DAVIDSON
2. VITTORIO VERONESE

Title of Invention: THE RECOVERY OF GOLD AND SILVER VALUES

~~*Provisional*~~*/Complete Specification as Filed*

SPOOR AND FISHER
PATENT ATTORNEYS
JOHANNESBURG — PRETORIA — DURBAN

Bill Bros H 1490

Patents Form 3

200.

SPOOR AND FISHER
PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 78/5147

TITLE OF INVENTION: " THE RECOVERY OF GOLD AND SILVER VALUES "

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 MAIN STREET, JOHANNESBURG

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

BACKGROUND OF THE INVENTION

This invention relates to the recovery of gold and silver values from their ores.

Conventionally on the Witwatersrand gold and silver values are recovered from an ore by finely milling the ore and leaching the milled ore with a suitable cyanide solution. After filtration and clarification of the solution the metal values are precipitated from solution by means of zinc.

It has also been proposed to recover gold from solution by means of activated carbon. In this case the solution is sometimes not clarified and the solution is contacted with activated carbon while it is still a pulp. This is a carbon-in-pulp process. The present invention is applicable to the recovery of gold and silver values both from clarified solutions and in carbon-in-pulp processes.

Conventionally the solution used for contacting the activated carbon is prepared in the same manner as for zinc precipitation and is alkaline at a pH of about 10 or just below. It has been thought for various reasons that the pH should be kept at that level.

However, in the literature there are reports on the use of pH levels below 7 under experimental conditions for the recovery of gold from cyanide solutions by means of activated carbon.

In the recovery of gold from solution by means of activated carbon little attention has been paid to the recovery of silver. In fact with

conventional activated carbon processes less silver is recovered than with zinc precipitation. Obviously it would be of advantage to optimize the recovery of silver.

The present invention is based on the discovery that above a pH of 8 gold after a short contact time displaces silver on the activated carbon, while if the pH is lowered to below 8 this displacement is retarded while at the preferred pH values gold and silver break away virtually simultaneously.

SUMMARY OF THE INVENTION

According to the invention a method of recovering gold and silver values from a solution obtained as a result of the leaching of a gold and silver bearing ore with a cyanide solution and contacting the solution with activated carbon to adsorb the gold and silver values in solution, is characterised in that prior to the contacting step the pH of the solution is lowered to a suitable value below 8 and the contact time is such that substantially all the gold and silver values are adsorbed on to the activated carbon.

In the preferred form of the invention the pH is lowered to between 4 and 5.

The loaded carbon may be eluted in a variety of well known ways.

The advantages of adsorbtion at a pH of between 4 and 5 is firstly that fouling by calcium carbonate, arising from the calcium invariably added as lime during the processing of Witwatersrand ores, is minimized to a large extent. Secondly it is possible to recover substantially all of the silver present in solution since at a pH of between 4 and 5 very little adsorbed silver is displaced by gold.

BRIEF DESCRIPTION OF THE DRAWINGS

Figure 1 is a graphic illustration of gold and silver adsorption from a plant solution at pH 9,8; and
Figure 2 is a graphic illustration of adsorption from the same solution acidified to a pH of 5,0.

DESCRIPTION OF PREFERRED EMBODIMENTS

Experiment 1

A plant cyanide solution containing 5,4g/t Au and 0,69 g/t Ag was pumped at a superficial flow velocity of 0,42 m/min equivalent to 45 bed volumes/h through a bed of activated carbon which had previously been equilibrated at a pH of 9,8. The carbon particle size as - 1190 µm + 840 µm. The results are plotted in Figure 1.

The same solution was acidified to a pH of 5,0 and was treated in a similar bed of carbon equilibrated at a pH of 5,0 at the same flow rates. The results of this procedure are plotted in Figure 2.

A comparison of Figures 1 and 2 show that at a pH of 9,8 there is an early breakthrough of silver, i.e. the gold displaced the silver on the carbon at an early stage while at a pH of 5 gold and silver broke through at more or less the same time.

Experiment 2

In this experiment three adsorption columns at a time were linked in series and after a time the first column was taken out of service for elution while a fresh column was added as the third column in the series. In the tables below the columns are referred to by capital letters A to H.

The same solution as that used above and acidified to a pH of 5,0 was pumped at a flow rate of 45 bed volumes per hour through the columns for a period of six days (continuous pumping was temporarily suspended after 96 hours due to a week-end shut-down). A summary of the resulting adsorption data is given below in Table 1.

TABLE I

Column Sequence	Total number of bed volumes passed/operational time (h)	Effluent from Second Column (daily average concentrations)		Effluent from Last Column (daily average concentrations)	
		g/t Au	g/t Ag	g/t Au	g/t Ag
A - B - C	1059/0-24	0,0018	< 0,001	0,0022	< 0,002
A - B - C	2117/24-48	0,0008	0,006	0,0008	< 0,001
B - C - D	3176/48-72	0,0013	< 0,001	0,0007	< 0,001
B - C - D	4234/72-96	0,0033	0,010	0,004	< 0,001
C - D - E	5293/96-120	0,0010	< 0,001	0,0010	< 0,001
C - D - F	6351/120-144	0,0014	0,022	0,0019	< 0,001
Average effluent values for 6 days		0,0016	< 0,006	0,0018	< 0,001

The average values of gold in the effluent over the six days of 0,0016 and 0,0018 g/t from the third and second columns in Table I are particularly noteworthy.

It was found that from an initial influent value of 5,0 the pH of the effluent became approximately 6,0. Since the effluent from the first column was also at a pH of 6,0 it was inferred that the rise in the pH occurred where gold adsorption was greatest.

Experiment 3

In order to achieve higher gold loadings on the lead column and at the same time to make a more effective use of the last column influent flow rates were increased from 44 to 66 bed volumes per hour and the previous experiment continued. Also, in this experiment the influent pH was lowered to a value of 4,3 because of the significant rise in pH observed on the lead column.

The resulting adsorption data are summarised in Table II below. Again operation extended over six days and the gold effluent values of 0,0008 g/t indicated a gold adsorption efficiency of over 99,9 percent. Silver adsorption was exceptionally good and also over 99,9 percent.

TABLE II

Column Sequence	Total number of bed volumes passed/operational time (h)	Effluent from Second Column (daily average concentration)		Effluent from Last Column (daily average concentration)	
		g/t Au	g/t Ag	g/t Au	g/t Ag
D - E - F	1590/0-24	0,0020	< 0,001	0,0015	< 0,001
D - E - F	3180/24-48	0,0220	0,014	0,0005	< 0,001
E - F - G	4770/48-72	0,0024	0,005	0,0005	< 0,001
E - F - G	6360/72-96	0,0709	0,032	0,0005	< 0,001
F - G - H	7950/96-120	0,0034	0,002	0,0008	< 0,001
Average effluent values for 6 days		0,0279	0,016	0,0008	< 0,001

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED WE DECLARE THAT WHAT WE CLAIM IS:

1.

A method of recovering gold and silver values from a solution obtained as a result of the leaching of a gold and silver bearing ore with a cyanide solution and contacting the solution with activated carbon to adsorb the gold and silver values in solution, characterised in that prior to the contacting step the pH of the solution is lowered to a suitable value below 8 and the contact time is such that substantially all the gold and silver values are adsorbed on to the activated carbon.

2.

The method claimed in claim 1 in which the pH is lowered to between 4 and 5.

3.

The method claimed in either of the above claims in which the solution is contacted with at least three batches of activated carbon in series, periodically batch is removed at the end of the series and a fresh batch is introduced at the start of the series and batch removal and introduction takes place before the gold and silver content shows a marked displacement of silver by gold.

4.

The method claimed in claim 3 in which the solution is clarified before the contacting step and the batches are adsorption columns.

5.

A method according to claim 1 substantially as herein described in any one of experiments 1 to 3.

DATED THIS 9th DAY OF AUGUST 1979

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APPLICANT'S PATENT ATTORNEYS

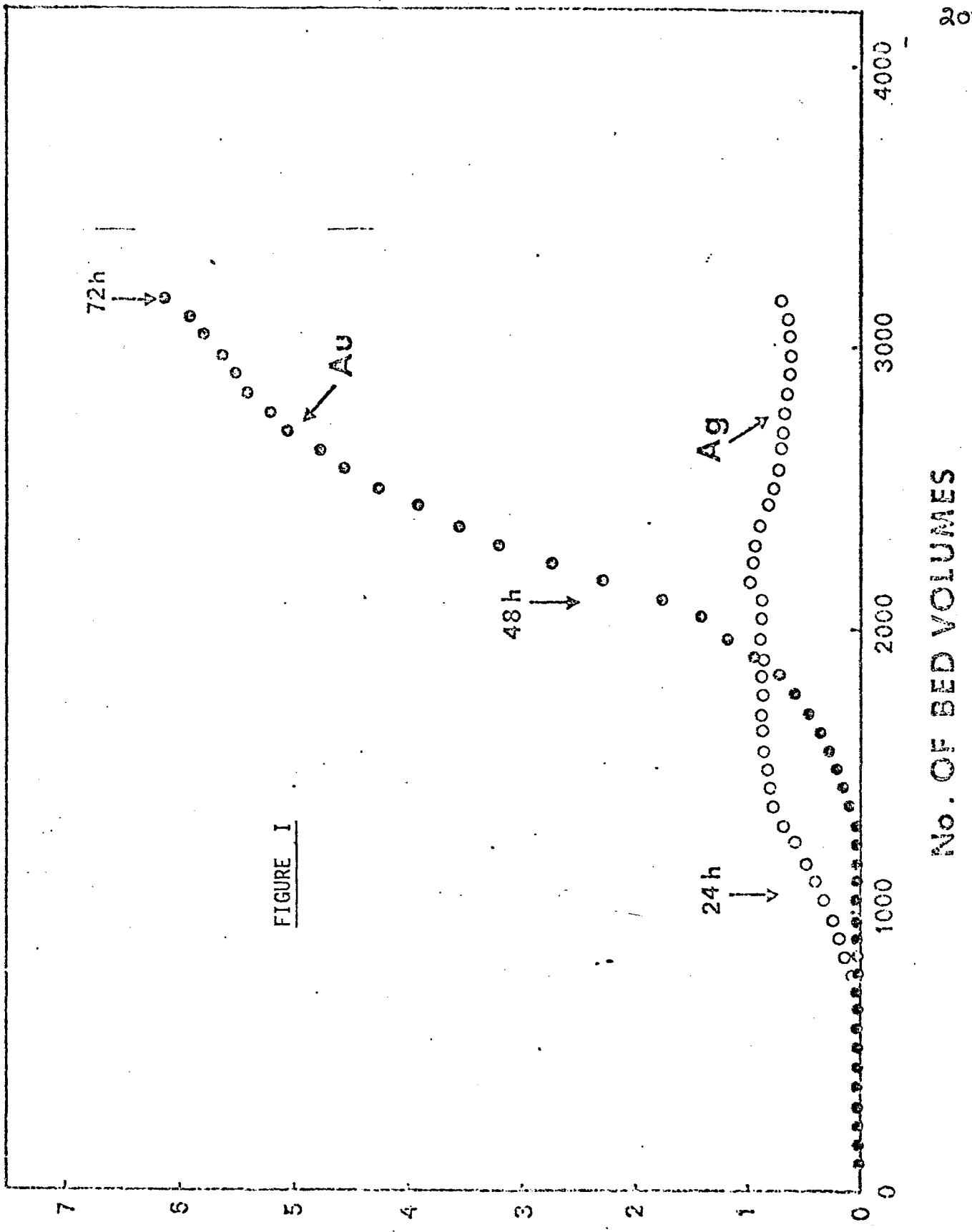
72h
48h
24h
Au
Ag
0
1000
2000
3000
4000
No. OF BED VOLUMES
0
1
2
3
4
5
6
7
COLUMN EFFLUENT (g/t Au/Ag)

FIGURE I

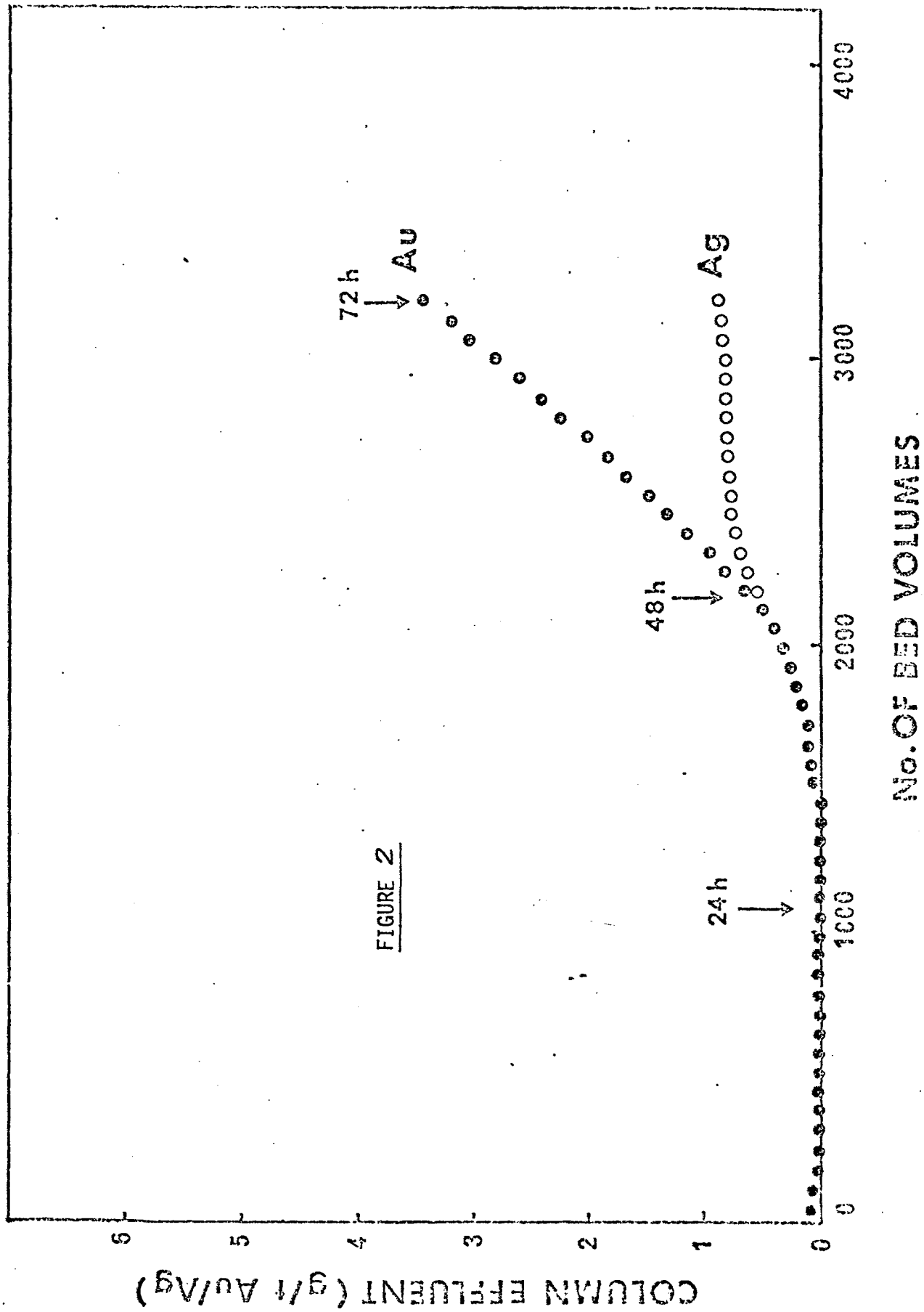
72 h
Au
Ag
48 h
24 h
0
1000
2000
3000
4000
No. OF BED VOLUMES
0
1
2
3
4
5
6
COLUMN EFFLUENT (g/t Au/Ag)

FIGURE 2

POOR AND FISHER

FORM P.7
(To be lodged in duplicate)

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978

COMPLETE SPECIFICATION

(Section 30(1) – Regulation 28)

OFFICIAL APPLICATION NO.		LODGING DATE	
21	01 833952	22	1 JUNE 1983

INTERNATIONAL CLASSIFICATION	
51	C 22 B

FULL NAME(S) OF APPLICANT(S)

71	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

FULL NAME(S) OF INVENTOR(S)

72	RAYMOND JOHN DAVIDSON

TITLE OF INVENTION

54	GOLD ADSORPTION

BACKGROUND OF THE INVENTION

This invention relates to the adsorption of gold values on to an activated carbon support.

Gold values are extracted from their ores using the well known cyanidation process. This process involves finely grinding the ore and then leaching the ore with a cyanide solution. The gold values are leached in the form of the aurocyanide ions and a gold-rich leach solution or pulp, known as a pregnant solution or pulp, is produced.

The gold may be recovered from the pregnant leach solution or pulp using various techniques such as adsorbing the gold values on to activated carbon. Adsorbing gold values on to activated carbon can be carried out in a fixed bed column or by slurrying the leach solution with the activated carbon (e.g. the so-called carbon-in-pulp process). The gold values adsorbed on to the activated carbon can be eluted therefrom using deionised water or other softened water.

The loaded activated carbon is subjected to a pre-treatment with a solution such as a sodium cyanide/sodium hydroxide solution, prior to the water elution. The solution which remains after this process is known as an effluent solution. The effluent solution generally still contains some gold values and a lower free cyanide concentration than the pregnant solution or pulp, and thus the effluent solution may be further treated, by contacting it with activated carbon to recover these residual gold values.

Effluent solutions are also obtained from other existing gold plant circuits using filtration and subsequent zinc precipitation.

The activated carbon which is used in these processes may be regenerated after use, so that it may once again be used for the adsorption of gold values from a solution or pulp. Various regeneration processes are known. For example, South African Patent No. 78/5864 describes and claims a process for regenerating activated carbon which has had its collecting capacity for gold adsorption reduced through use, which includes the step of contacting the carbon with an aqueous copper salt solution having a pH of less than 6.

SUMMARY OF THE INVENTION

According to the invention, there is provided a method of recovering gold values from a solution or a pulp containing dissolved gold values, dissolved copper values and free cyanide which includes the steps of:

(a) bringing the pH of the solution or the pulp to between 6 and 9;

(b) contacting the solution or the pulp with an activated carbon support so as to cause gold and copper values to be adsorbed on to the activated carbon;

(c) contacting the loaded activated carbon with an acid solution; and

(d) thereafter contacting the activated carbon with further of the solution or the pulp so as to cause further gold and copper values to be adsorbed on to the activated carbon.

It is believed that the treatment of the loaded activated carbon with an acid solution produces suitable conditions for the formation of a highly insoluble gold/copper cyanide complex thereby increasing the ability of the activated carbon to adsorb gold values.

The term "free cyanide" means cyanide which is uncomplexed and which is present as cyanide (CN^-) ions.

Steps (c) and (d) of the method of the invention are preferably repeated at least once, until a desired concentration of gold values is adsorbed on to the activated carbon. Preferably, steps (c) and (d) are repeated until the concentration of the gold values adsorbed on to the activated carbon relative to the concentration of the gold values in the solution or the pulp prior to treatment is increased 30 000 fold. Following the method of the invention, concentration factors of 40 000 to 60 000 can be achieved.

Step (c) of the above method is preferably carried out when the activated carbon is substantially fully loaded with the gold values. The activated carbon may be regarded as being fully loaded, when the concentration of gold values in the solution or the pulp prior to contact with the activated carbon is the same as or no more than 10 percent higher than the concentration of gold values in the solution or the pulp after it has been contacted with the activated carbon. In other words, when the activated carbon is fully loaded, it cannot adsorb further gold values without further treatment.

The method of the invention has particular application with solutions or pulps which have a low free cyanide ion concentration. The method of the invention appears particularly useful with solutions or

pulps in which the free cyanide ion concentration is less than 10 grams per metric ton, calculated as NaCN. Effluent solutions generally have such low free cyanide ion concentration, and thus the method of the invention is particularly useful with effluent solutions.

In step (c), the acid used is preferably a suitable mineral acid. Preferred acids are hydrochloric acid or sulphuric acid. Typically, the acid will be a dilute solution.

Preferably, the pH of the solution or the pulp in contact with the activated carbon during step (c), is brought to between 1 and 2 by the addition of the acid.

The method of the invention is preferably carried out at ambient temperature, i.e. between 10°C and 25°C, but it may be carried out at higher temperatures.

BRIEF DESCRIPTION OF THE DRAWINGS

Figure 1 is a schematic view of a fixed bed carbon pilot plant used in the method of the invention; and

Figure 2 and Figure 2 (continued) are graphs showing the effect of the addition of an acid solution periodically to the pilot plant of Figure 1, on the gold adsorption by the activated carbon in the pilot plant.

DESCRIPTION OF PREFERRED EMBODIMENTS

Embodiments of the invention will now be described both in relation to a fixed bed carbon plant and a carbon-in-pulp (CIP) plant.

FIXED BED CARBON PLANT

The method of the invention was tested in the fixed bed carbon pilot plant illustrated in Figure 1.

Materials and Method

1. Fixed Bed Carbon Pilot Plant

The fixed bed carbon pilot plant comprised three Perspex adsorption columns designated A, B and C, connected to each other in series (each column having an internal diameter of 50mm and a length of 400cm). Figure 1 illustrates the flowsheet of the circuit whereby a merry-go-round type of operation is carried out. In other words, the solution to be treated was first pumped through the first column known as the lead column, which at the start of the operation was Column A, and then through the second column (Column B), and finally through the third column (Column C). When the activated carbon in the lead column was fully loaded with gold values, it was removed from the circuit, the second column became the first column, the third column became the second column, and a new column was added as the third column. A similar procedure was effected each time the activated carbon in the lead column was fully loaded with gold values.

The influent solution was pumped at a constant flowrate using a model EKH2 Lewa diaphragm metering pump (Rhine Ruhr Pty. Ltd.). Each column was packed with activated carbon to a bed depth of 3m while solution sampling points were positioned at 1m intervals down the length of each column. 0,25mm resin strainers (Portals Water Treatment S.A. Pty. Ltd.) were placed at the top and bottom of each column to support and to confine the carbon bed during normal and backwashing cycles respectively. Carbon was

transferred to the columns by means of a water pressurised cone-shaped transfer vessel. Solution samples corresponding to each metre of carbon through the circuit were taken on a daily basis, while a daily composite drip sample was taken from each column.

2. Column Influent

Clarified primary U-plant filtrate (i.e. an effluent solution) was acidified using sulphuric acid and stored in a 500 m^3 storage tank. The initial bulk solution was adjusted to pH 8,0 and contained the following: 0,62 g/t Au; 0,007 g/t Ag; 0,8 g/t Cu; 4,7 g/t Ni; 0,3 g/t Zn; 771 g/t Ca; 3440 g/t Na; about 30 g/t suspended solids and less than 10 g/t CN^-. This solution was used as column influent for 66 days prior to refilling the bulk storage tank and adjusting pH. In the present investigation influent was pumped at a flowrate of 2,4 litres/min. Minor adjustments in pH were also made during the course of the period under review.

3. Activated Carbon

0,8 - 1,2mm type G215 carbon was initially in use in the plant. However, 0,6 - 1,7mm type G210-AS carbon was phased into use at the start of cycle 3A 3B 3C, i.e. columns 3C and 4A. The bulk density of this latter carbon was 0,55 g/cm^3.

4. Acid Washing

Acid washing involved the pulping of either 20 litres of 3 percent (vol/vol) HCl into the circuit at about 1 week intervals or the daily acid washing of the carbon with 3-5 litres of 3 percent (vol/vol) HCl. The acid was pumped into the circuit using the influent pump set at the same flowrate (i.e. 2,4 litres/min).

5. Gold Elution

Loaded carbon from the adsorption columns was pretreated, prior to elution, with 8 litres of a reagent containing 5 per cent NaCN and 2 per cent NaOH prior to elution with distilled water at 110°C. No acid washing of the loaded carbon was carried out in the elution process.

RESULTS

Gold Adsorption

Each day, the following concentrations were measured:

(a) the concentration of gold values in the influent solution prior to its entry into the pilot plant, designated C_o, and

(b) the concentration of gold values in the effluent solution after it had passed through each metre of activated carbon in each of the three columns, designated C_T, (sampling being carried out at the points marked A1, A2, A3, B1, B2, B3, C1, C2 and C3 in Figure 1).

The ratio of C_T to C_o was then calculated and plotted on the graphs shown in Figure 2 and Figure 2 (continued) for each day of operation, during the cycle designated 2C, 3A, 3B, in which Column C was the lead column, Column A the second column, and Column B the third column.

During this cycle two different acid washing procedures were used. Initially, the circuit was washed with 20 litres 3 per cent HCl intermittently, i.e. after 0, 5, 15, 22, 28 and 33 days respectively. The resulting breakthrough curves corresponding to each meter bed depth of carbon in the circuit relating gold adsorption at pH 8 indicate a very beneficial effect on gold adsorption with the column effluent averaging 0,002 g/t gold (see Figure 2). The acid washing procedure was then changed to the daily use of 3-5 litres 3 per cent HCl (see Figure 2 continued). This subsequent procedure may be seen to be even more beneficial to gold adsorption, besides being more economical with regard to the cost of the acid used as well as plant downtime (1-2 min/day was required to pump acid through the circuit). After 42 days of continuous operation carbon samples were taken from the lead column 2C indicating an average gold loading of 20 Kg/t. Further operation (42-54 days) continued to yield satisfactory results with calculated gold loadings approaching 30 Kg/t. Column effluent during cycle 2C 3A 3B averaged 0,005 g/t Au corresponding to a gold recovery of 99 per cent.

Gold Elution

Loaded carbon taken from the circuit during the period under review was pretreated with 8 litres 5 per cent NaCN/2 per cent NaOH and eluated at 110°C. No elution problems were encountered. In all cases, over 99 per cent of the contained gold and copper was recovered from the loaded carbon.

Gold Precipitation

The acidification of composite eluates to pH 1 using sulphuric acid resulted in the precipitation of 99 per cent of the contained gold and silver values. Subsequent calcining and smelting of the gold-copper rich precipitate with a borax flux resulted in bullion containing about 1 per cent base metal impurity.

CARBON-IN-PULP PLANT

The operation of a Carbon-in-Pulp plant is described below.

1. Cyanidation

A gold- and copper-bearing ore was ground and milled and the ore so treated formed into a pulp with a standard cyanide leach solution. Leaching was continued for 18 hours after which it was found the pulp had a free cyanide content of 150 grams per metric ton (g/t). The pH of the pulp was 10,5. The pH of the pulp was reduced to 8 by the addition of sulphuric acid and this had the effect of reducing the free cyanide content to 16 g/t.

2. Pilot Plant Description

Two seven stage carbon-in-pulp miniplants were used. Each miniplant consisted of seven mechanically agitated contactors, each with a working volume of 10 litres. Each contactor had an 850µm screen of area 100 cm^2 at the outflow which was cleaned by mechanically timed air and water blasts on the discharge side of the screen (air was blasted for 10 seconds every 3 minutes and the water blasted for 5 seconds every 2 hours). The impellers consisted of three flat blades inclined at 45° to the shaft axis. The feed to the plant was kept constant by a Sigmamotor model

T6-S variable speed finger pump. Each contactor contained a contact thermometer which activated a 500 watt heater to maintain the desired pulp temperature.

3. Pilot Plant Operation

Cyanided pulp stored in a 3 m^3 mechanically agitated stock tank was fed to the plant at a flowrate of 10 litres/hour. Under continuous operation, carbon was sequentially transferred on a daily basis. The carbon and pulp in stage 1 was drained through a 710 µm screen, briefly washed and a small sample taken for analysis, the balance being fully loaded carbon. The carbon and pulp in stage 2 was similarly screened and then placed in stage 1 and this was repeated down to stage 7. The required volume of fresh carbon was then added to stage 7. The principal objective in running the plants was to simulate steady state operation and to this end each run was continued for several weeks.

Cyanided pulp containing about 310 mg/litre NaCN with a relative pulp density of 1,45 was used as a common feed to the two miniplants operating simultaneously. In plant A the pulp was first neutralized to pH 8,0 using sulphuric acid (as described above) in the first mechanically agitated stage before being contacted with carbon in the remaining six stages. Each stage in this carbon circuit was sparged with air using sintered glass diffusers. In plant B, pulp at pH 10,5 was fed directly into six contacting stages. No air sparging was used in this instance. The pulp feed to each plant averaged 3,9 g/t soluble gold. In both plants, a carbon concentration of 7,5 g/litres was used, while carbon was moved at a rate of 56 g/d. Temperature was controlled in both plants at 25°C.

Summary data for plants A and B are presented in Table 1 following 20 days of continuous operation (recovery data was averaged over the last 5 days of operation in order to simulate steady state conditions). The recovery data show a significant improvement in gold recoveries in plant A operated at pH 8,0 with stage recoveries averaging 62 per cent compared with 51 per cent in plant B.

Likewise, an overall recovery of 99,7 per cent in plant A may be compared with a 98,8 per cent recovery in plant B. The very significant increase in cyanide destruction in plant A is also most noteworthy and is certainly responsible in part at least for the improved performance of this plant.

The carbon loading data (Table 1) also show very different profiles in the respective plants. The relatively high copper loadings in plant A and high loadings of calcium and silica in plant B are very evident. The large increase in copper loadings towards the tail of the circuit in plant A is almost certainly related to the decrease in cyanide concentration. Loaded carbon taken from the circuits after 20 days indicated loadings of 16,2 kg/t gold; 1,0 kg/t silver in plant A and 16,7 kg/t gold; 1,0 kg/t silver in plant B.

In the case of plant A it was found that the free cyanide concentration dropped markedly when the plant feed was adjusted to pH 8,0. From a feed concentration of about 310 g/t soluble NaCN at pH 10,5, the cyanide concentration decreased to about 120 g/t on lowering the pH to 8,0. This decrease in cyanide concentration was due largely to the increased reactivity of the pulp at pH 8,0.

Although the loaded activated carbon was not subjected to periodic acid washing there can be no doubt from an analysis of the data obtained and as set out in Table 1 such washings would improve further the concentration factors obtainable, provided the cyanide levels in the process are maintained at a suitably low level.

TABLE 1

RECOVERY AND LOADING DATA FOR CIP TREATMENT

Recovery Data	Plant	Stage Number					
		1	2	3	4	5	6
Gold in Solution (g/t Au)	A	1,700	0,600	0,196	0,058	0,023	0,012
	B	2,290	1,316	0.822	0,254	0,094	0,050
Progressive gold recovery (%)	A	59,0	85,50	95,30	98,60	99,50	99,70
	B	44,90	68,40	80,20	93,90	97,70	98,80
Stage Gold recovery (%)	A	59,0	64,6	67,5	70,4	61,2	46,7
	B	44,9	42,5	37,5	69,1	63,0	46,8
Free cyanide in solution (g/t NaCN)	A	64	28	17	12	12	8
	B	234	187	133	96	76	63
Carbon Loading Data* (kg/t)							
Au	A	20,2	8,7	4,4	1,9	0,7	0,2
	B	13,8	11,6	9,0	7,7	5,2	1,3
Ag	A	0,2	1,4	1,2	0,9	0,5	0,2
	B	1,0	0,8	1,0	1,1	1,3	0,8
Cu	A	1,6	1,9	3,5	6,0	9,5	10,4
	B	0,1	0,1	0,1	0,1	0,2	0,2
Ni	A	3,5	2,8	2,9	2,8	2,8	2,2
	B	1,4	1,3	1,3	1,5	1,5	1,2
Ca	A	10,7	8,1	7,5	6,0	4,8	4,2
	B	34,7	32,8	29,8	18,1	9,3	3,3
SiO	A	5,7	6,2	5,6	5,0	6,1	4,9
	B	13,8	12,3	10,6	8,6	5,7	4,7

* Carbon samples taken after 15 days.

FURTHER EXPERIMENTS

Experiments were performed in borate buffer at pH 7 with copper loaded carbon in both the absence and presence of cyanide. The copper was initially loaded on the carbon as a cyanide complex.

In the absence of cyanide the adsorbed copper had little influence on gold adsorption. Linear isotherms were obtained following a log-log plot compared with the slightly curved isotherms in borate buffer. There was little difference between carbon loaded with 13 and 32 kg Cu/tC.

In the presence of 200 mg/litre NaCN the gold loading was considerably lower employing the copper-loaded carbon. The k-values were 30,5 and 27,5 (kg Au/tC) for the carbon loaded with 13 and 32 kg Cu/tC respectively, compared with a k-value of 40 for the virgin carbon - the k-value is defined as the gold loading in equilibrium with 1 g/t gold in solution.

Optimum results were obtained with copper-loaded carbon (13 kg Cu/tC) after acid washing (3% HCl) and in the absence of cyanide. After acid washing the k-value was 70 compared with 42,5 for the virgin carbon. When the carbon was loaded with 13 kg/t Cu plus 21 kg/t Au and subsequently acid washed and re-equilibrated with gold solution, a k-value of 80 kg Au/tC was obtained. (In the presence of 200 mg/litre NaCN the k-values were 28,5 and 31 under the same conditions as above, compared with 40 for the virgin carbon).

The above data confirms in many respects the beneficial effects of periodic acid washing with dilute hydrochloric acid. The formation of a highly insoluble AuCN CuCN compound following such treatment would explain the irreversible character of the mechanism together with the abnormally high gold loadings achieved - concentration factors of 3000-5000 are commonly obtained in a conventional CIP operation which may be compared with a concentration factor of about 60000 obtained in the fixed bed operation.

CLAIMS

1.

A method of recovering gold values from a solution or a pulp containing dissolved gold values, dissolved copper values and free cyanide, includes the steps of:

(a) bringing the pH of the solution or the pulp to between 6 and 9;

(b) contacting the solution or the pulp with an activated carbon support so as to cause gold and copper values to be adsorbed on to the activated carbon;

(c) contacting the loaded activated carbon with an acid solution; and

(d) thereafter contacting the activated carbon with further of the solution or the pulp so as to cause further gold and copper values to be adsorbed on to the activated carbon.

2.

A method according to claim 1 wherein steps (c) and (d) are repeated, at least once, until a desired concentration of gold values is adsorbed on to the activated carbon.

3.

A method according to claim 2 wherein steps (c) and (d) are repeated until the concentration of the gold values adsorbed on to the activated carbon relative to the concentration of the gold values in the solution or the pulp prior to contact with the activated carbon is increased 30 000 fold.

4.
A method according to any one of the preceding claims wherein step (c) is carried out when the activated carbon is substantially fully loaded with the gold values.

5.
A method according to any one of the preceding claims wherein the concentration of free cyanide in the solution or the pulp, prior to step (a), is low.

6.
A method according to claim 5 wherein the concentration of free cyanide in the solution or the pulp, prior to step (a), is less than 10 grams per ton, calculated as NaCN.

7.
A method according to claim 5 wherein the solution is an effluent solution.

8.
A method according to any one of the preceding claims wherein in step (c), the acid used is a mineral acid.

9.
A method according to claim 8 wherein in step (c), the acid used is hydrochloric acid or sulphuric acid.

10.
A method according to any one of the preceding claims wherein the pH of the solution or the pulp in contact with the activated carbon during step (c) is brought to between 1 and 2.

11.

A method according to any one of the preceding claims wherein the method is carried out at ambient temperature.

12.

A method according to claim 11 wherein the method is carried out at a temperature between 10°C and 25°C.

13.

A method of recovering gold values from a solution or a pulp containing dissolved gold values, dissolved copper values, and free cyanide, substantially as herein described.

DATED this 1st day of JUNE 1983

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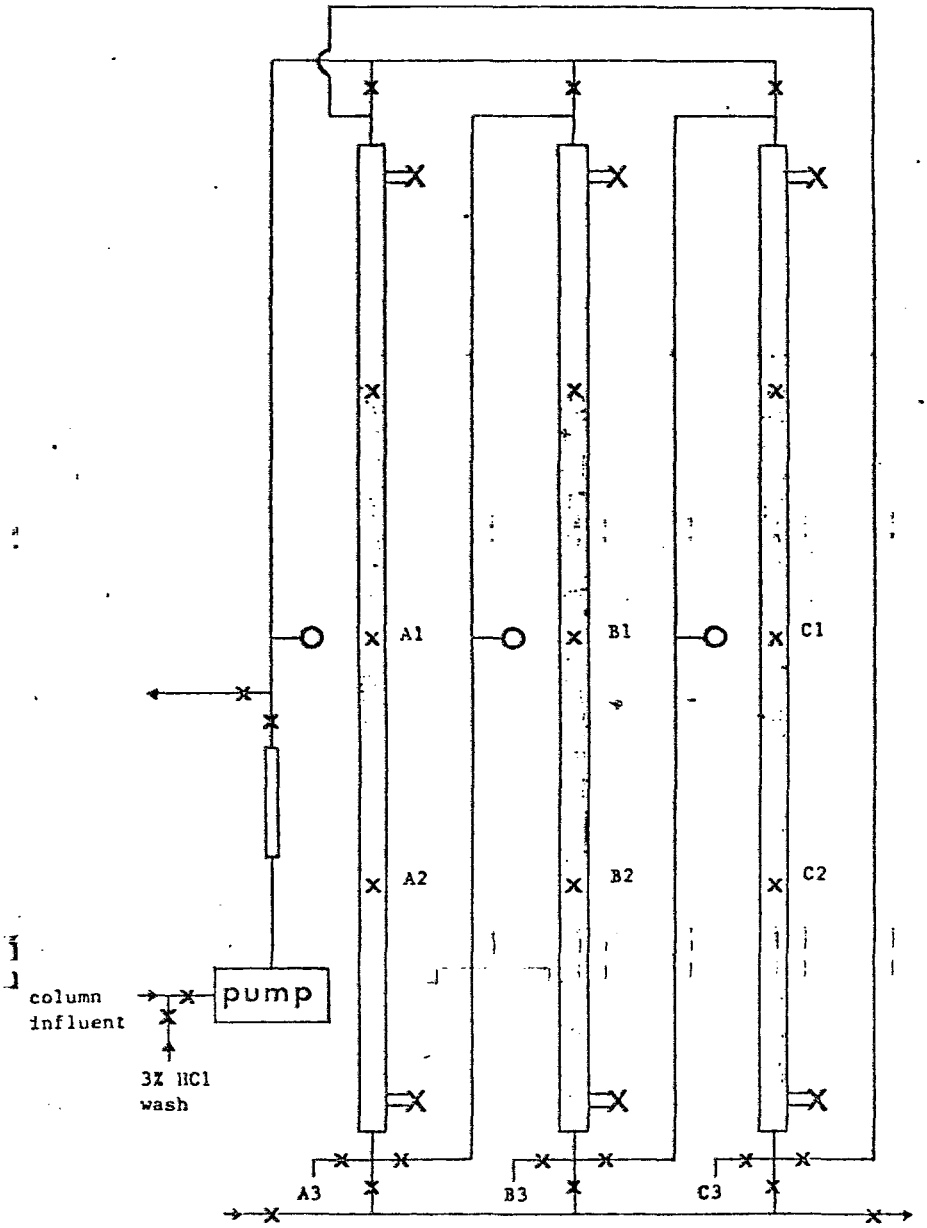
SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS

FIG 1.



A.A. Dyer

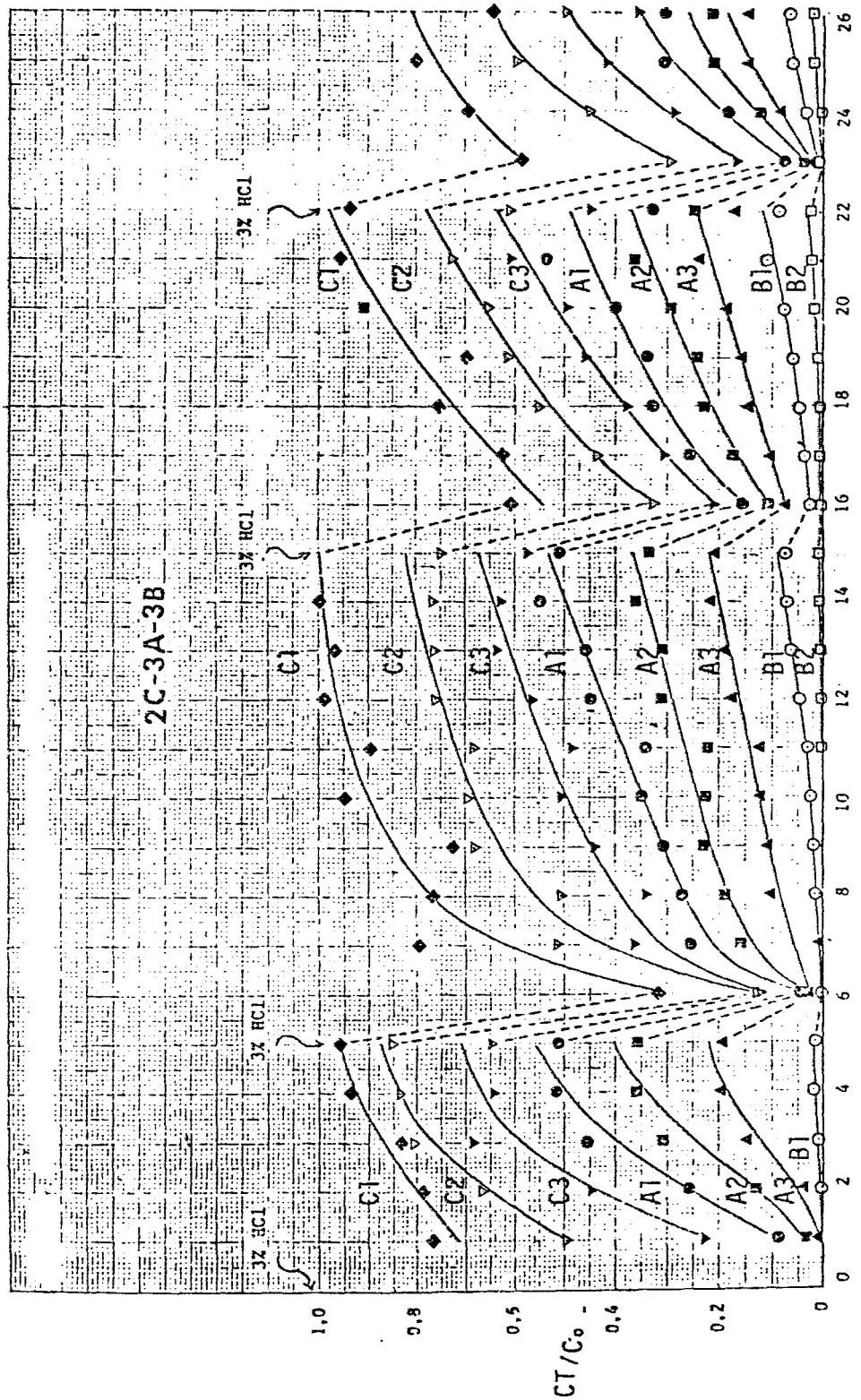
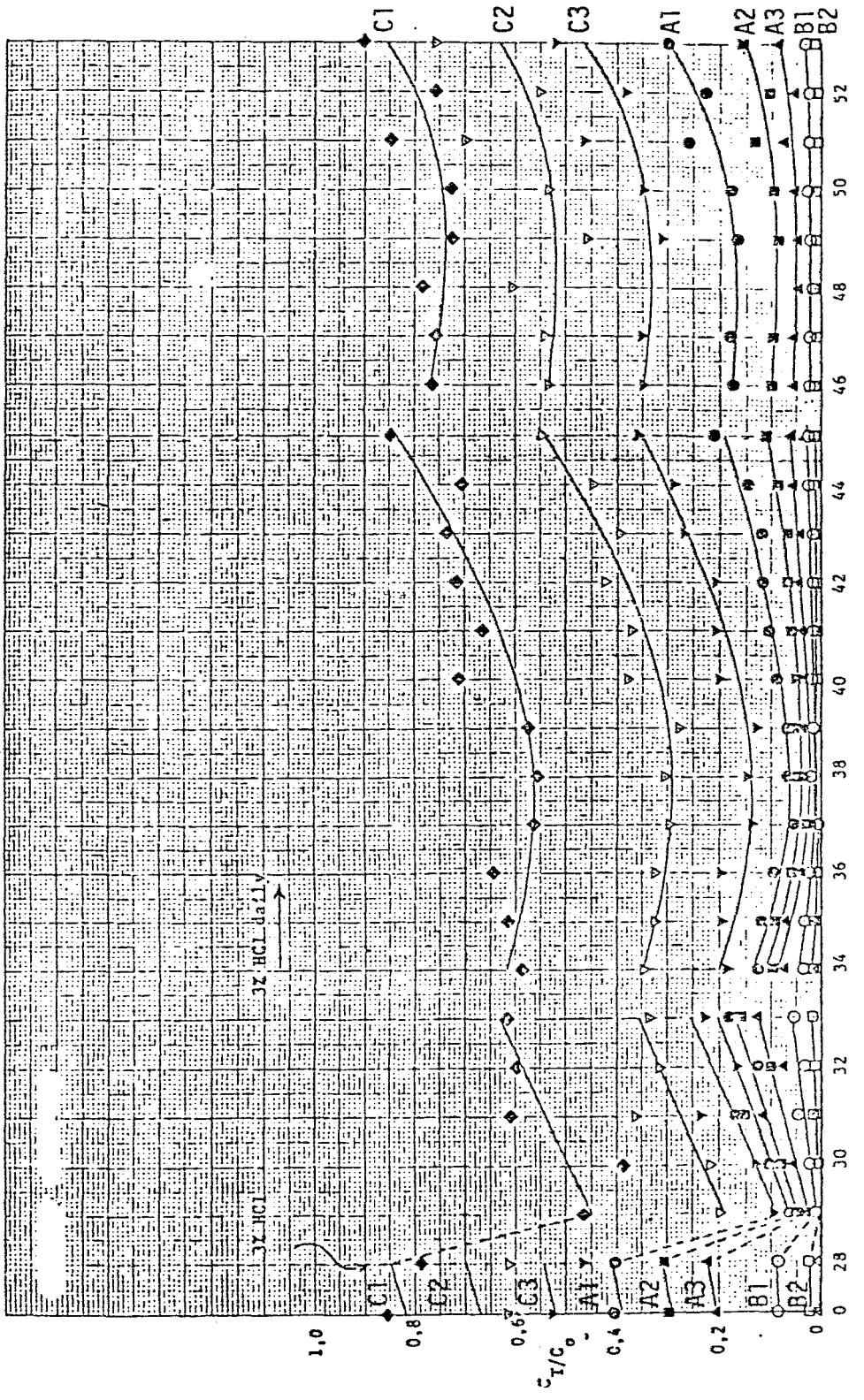

2C-3A-3B
3% HCl
C1
C2
C3
A1
A2
A3
B1
B2
CT
CT/Co -
TIME (days)
0
0,2
0,4
0,5
0,8
1,0
0
2
4
6
8
10
12
14
16
18
20
22
24
26

FIG. 2

FIG 3.2 (CONTINUED)



SPOOR and FISHER

3.2 FIXED BEDS

The on-site operation of a pilot plant for the recovery of gold from primary U-plant filtrate at the Vaal Reefs West division, resulted in the commissioning of a full scale operation treating 80 - 120 m^3/h at the West division in 1982. The capital cost of the plant using redundant ion-exchange columns was R210 000 which was recovered in 2 - 3 months. A similar full scale operation was later commissioned at the East division.

The recovery of gold from plant effluent by the use of activated carbon

by R. J. DAVIDSON*, Ph.D., M.Sc. and B. STRONG†, M.Sc., C.Chem., M.R.I.C.

SYNOPSIS

The piloting and eventual large-scale operation of fixed-bed carbon columns for the recovery of gold from plant effluents is described.

A fixed-bed carbon pilot plant comprising three adsorption columns was used for the recovery of gold from clarified plant effluent at the Vaal Reefs Exploration and Mining Company (West Division). The effluent ,which contained 0,1 to 0,5 g/t of gold and less than 10 g/t of free cyanide, was adjusted to pH 8 prior to use. Daily acid treatment of the carbon circuit with 3 per cent hydrochloric acid was shown to have a most significant effect on increasing the gold adsorption. The acid treatment involved a plant downtime of only 1 to 2 minutes per day. High loadings of gold (up to 30 kg/t) and copper (up to 75 kg/t) were obtained on the leading column, while the gold recoveries averaged 98 per cent over 17 weeks of continuous operation. The subsequent elution of loaded carbon in a pilot elution column indicated the use of abnormally high amounts of the pretreatment reagent (caustic soda and cyanide) used in gold elution. This was attributed to the high loading of copper on the carbon. However, the elution of both the gold and the copper at 110°C with demineralized water as the eluant was satisfactory.

In a similar large-scale operation, three fixed-bed carbon columns each containing 4,5 t of carbon were used to treat 80 to 120 t/h of plant effluent with a gold concentration of about 0,15 g/t. Gold loadings of 5 to 6 kg/t were obtained regularly, while the gold recovery averaged 97 per cent over a 12-month period. The capital cost of the plant was recovered in 2 to 3 months.

The evidence presented strongly suggests that copper plays a significant role in the adsorption and elution of gold under the prescribed conditions, and in the subsequent precipitation of gold following acidification of the eluates. The formation of a highly insoluble gold–copper complex produced under acid conditions appears to explain many of the observations made in the investigation.

SAMEVATTING

Die loods en uiteindelike grootskaalse bedryf van vastebedding-koolstofkolomme wat vir die nawinning van goud uit aanleguitvloeisels gebruik word, word beskryf.

'n Vastebedding-koolstofloodsaanleg bestaande uit drie adsorpsiekolomme is vir die winning van goud uit verhelderde aanleguitvloeisel by Vaal Reefs Exploration and Mining Company (Wes-afdeling) gebruik. Die uitvloeisel wat 0,1 tot 0,5 g/t goud en minder as 10 g/t vrye sianied bevat, is vóór gebruik tot pH 8 aangepas. Daar is bewys dat die daaglikse suurbehandeling van die koolstofbaan met soutsuur van 3 persent 'n uiters belangrike uitwerking op die verhoging van goudadsorpsie het. Met die suurbehandeling was 'n staantyd van die aanleg van slegs 1 tot 2 minute per dag nodig. Hoë ladings goud (tot 30 kg/t) en koper (tot 75 kg/t) is op die leierkolom verkry, terwyl goudwinning oor 'n ononderbroke tydperk van 17 weke gemiddeld 98 persent was. Die latere uitwassing van gelaaide koolstof in 'n loodsuitwassingskolom het die gebruik van buitengewoon hoë hoeveelhede van die byt–sianiedreagens, wat as voorbehandelingsmiddel by gouduitwassing gebruik word, getoon. Dit is toegeskryf aan die hoë lading van koper op die koolstof. Die uitwassing van goud sowel as koper by 110°C met die gebruik van gedemineraliseerde water as uitwasmiddel is egter bevredigend.

In 'n soorgelyke grootskaalse onderneming is drie vastebedding-koolstofkolomme wat elk 4,5 t koolstof bevat, gebruik om 80 tot 120 t/h aanleguitvloeisel te behandel wat ongeveer 0,15 g/t goud bevat. Goudafsettings van 5 tot 6 kg/t is gereeld verkry, terwyl goudwinning oor 'n tydperk van 12 maande gemiddeld 97 persent was. Die kapitaalkoste van die aanleg is binne 2 tot 3 maande verhaal.

Kragtens die bewys wat gelewer is, speel koper klaarblyklik 'n belangrike rol in die adsorpsie en uitwassing van goud onder die voorgeskrewe toestande asook met die latere presipitering van goud ná die aansuring van die uitwasmiddels. Die vorming van 'n hoogs onoplosbare goud–koperkompleks wat onder suurtoestaande gelewer word, is blykbaar 'n verklaring van baie van die waarnemings wat in die ondersoek gedoen is.

Introduction

Activated carbon is well known as a most efficient scavenger for gold from solutions containing less than 1 g/t of dissolved gold. In the South African gold-mining industry, there are a number of such process streams that lend themselves to the recovery of gold on activated carbon, yielding barren solutions with a gold content of less than 0,01 g/t. One such stream is the filtrate from the dewatering filters between gold and uranium plants. The filtrate from the West Uranium Plant, Vaal Reefs Exploration and Mining Co. Ltd, contains approximately 0,15 g/t of dissolved gold, and about 75 per cent of this solution reports to the slimes dams. No more than one-third of this liquor is returned to the plant.

This paper describes both the pilot operation and the first phase in the recovery of this dissolved gold from the dewatering filters that are at present in operation at the Vaal Reefs West Division.

Pilot Plant

To provide information for the design and operation of the production plant, a pilot operation was launched as a joint venture between Vaal Reefs Exploration and Mining Company Limited (V.R.E.M.) and Anglo American Research Laboratories (A.A.R.L.). The pilot plant comprised three columns operating in a conventional merry-go-round sequence. Influent through the circuit travelled downwards in order to promote a high shear situation that would be beneficial to the recovery of gold

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SA ISSN 0038 — 223X/$3.00 +0.00.

from very dilute solutions. Also, the use of higher flowrates of influent was made possible by the downwards flow provided that excessive pressure drops across the circuit were not incurred. Apart from high flowrates, this would depend largely on the clarity of the influent as well as on the careful selection of carbon particle size.

The pilot-plant unit was sited at the V.R.E.M. West Gold Plant, where the recovery of gold from plant effluent was being investigated. This investigation concerned the operation of this pilot plant over 17 weeks of continuous operation, together with the subsequent elution of loaded carbon from the pilot plant. The precipitation of gold from pregnant eluates using sulphuric acid was also briefly considered.

Adsorption of Gold

Clarified gold-plant effluent was pumped through three Perspex adsorption columns in series (each column being 50 mm in internal diameter and 4,0 m long). Fig. 1 illustrates the flowsheet of the circuit, showing that a merry-go-round type of operation was carried out. The influent was pumped at a constant flowrate by use of a diaphragm metering pump. Each column was packed with activated carbon to a bed depth of 3 m, and the sampling points were positioned at 1 m intervals down the length of each column. At the top and bottom of each column were 0,25 mm resin strainers to support and confine the carbon bed during normal and backwashing cycles respectively. The columns were designated A, B, and C. Carbon was transferred to the columns by means of a water-pressurized cone-shaped transfer vessel. Samples of solution corresponding to each metre of carbon through the circuit were taken on a daily basis, while a daily composite drip sample was taken from each column.

Backwashing of individual columns with regional water was carried out at intervals, as was acid washing of all three columns. Acid washing involved the pumping of 20 litres of 3 per cent (vol./vol.) hydrochloric acid into the circuit, and then allowing for a 30-minute soaking period before continuing normal operation. When daily acid washing was undertaken later, 3 to 5 litres of 3 per cent (vol./vol.) hydrochloric acid was pumped into the circuit and normal operation continued, i.e. no soaking period was allowed.

Use was made of 0,8 to 1,2 mm type G215 coconut-shell activated carbon (Le Carbone Pty Ltd). This relatively small particle size of granular carbon and its narrow range were selected[1] so that the kinetic adsorption of gold would be maximized while the pressure drop across the circuit during pumping at relatively high flowrates would be minimized. The bulk density of the wet-settled carbon was 0,42 g/cm^3.

Clarified plant effluent was neutralized with sulphuric acid and stored in a 500 m^3 storage tank. The column influent was pumped at a flowrate of 2,5 l/min, corresponding to a superficial flow velocity of 1,2 m/min and resulting in a pressure drop across the circuit of 267 to 310 kPa. For the initial 72 days of continuous operation (Phase I), the column influent adjusted to pH 7,4 contained the following: gold 0,15 g/t, silver 0,006 g/t, calcium 560 g/t, copper 8,3 g/t, nickel 5,9 g/t, iron 0,09 g/t, cyanide ions less than 10 g/t, and suspended solids about 20 g/t.

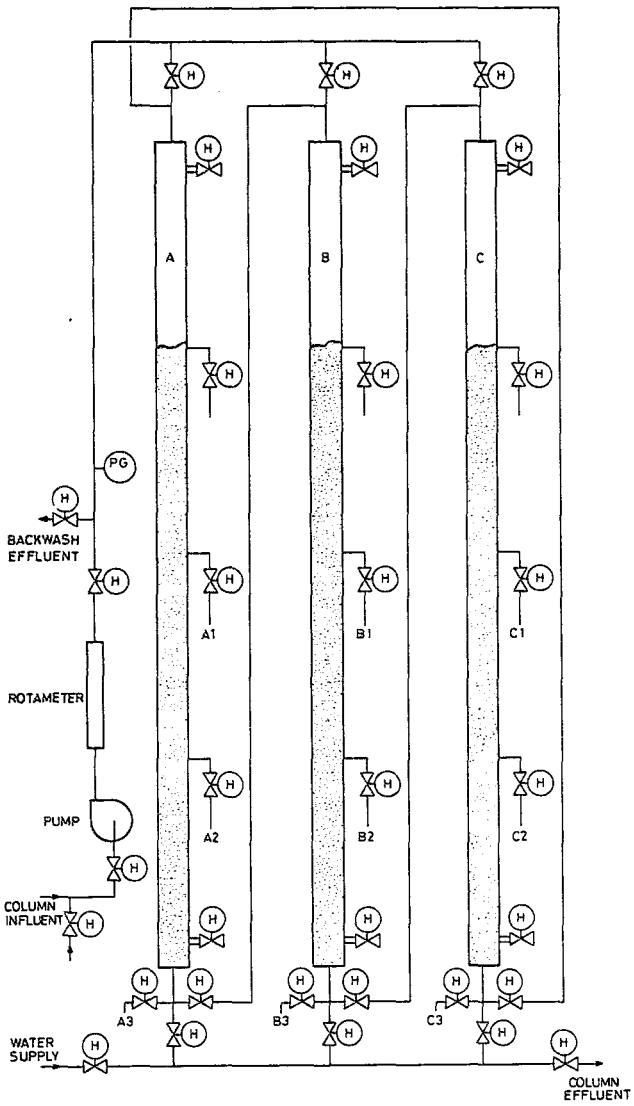

Fig. 1—The fixed-bed carbon pilot plant

The column influent for the subsequent 58 days of continuous operation (Phase II) adjusted to pH 8,0 contained the following: gold 0,49 g/t, silver 0,007 g/t, sodium 3440 g/t, calcium 771 g/t, copper 0,8 g/t, nickel 4,7 g/t, zinc 0,3 g/t, cyanide ions less than 10 g/t, and suspended solids about 30 g/t. This solution was used as column influent for 66 days prior to the refilling of the bulk storage tank and adjustment of pH (Phase III).

Elution of Gold

Loaded carbon from the lead adsorption column was removed in 1 m sections for analytical purposes before the three samples were composited and the gold eluted. Loaded carbon was transferred to an oil-jacketed stainless-steel elution column (50 mm in internal diameter and 4,0 m long) as illustrated in Fig. 2. The transfer of carbon was again effected by means of a water-pressurized cone-shaped transfer vessel. Excess solution was allowed to drain from the column before the tempera-

ture of the oil bath was raised to 112°C. Unless otherwise specified, the carbon was then pretreated with 6 litres of a reagent containing 5 per cent sodium cyanide and 2 per cent sodium hydroxide, which was pumped into the column through a stainless-steel heat-exchange spiral, also maintained at 112°C. The pretreatment reagent was pumped at a flowrate of 6 l/h, corresponding to a superficial velocity of 0,05 m/min. Gold was then eluted at the same flowrate by the use of distilled water[2], and the column eluate was collected in 2-litre fractions prior to analysis.

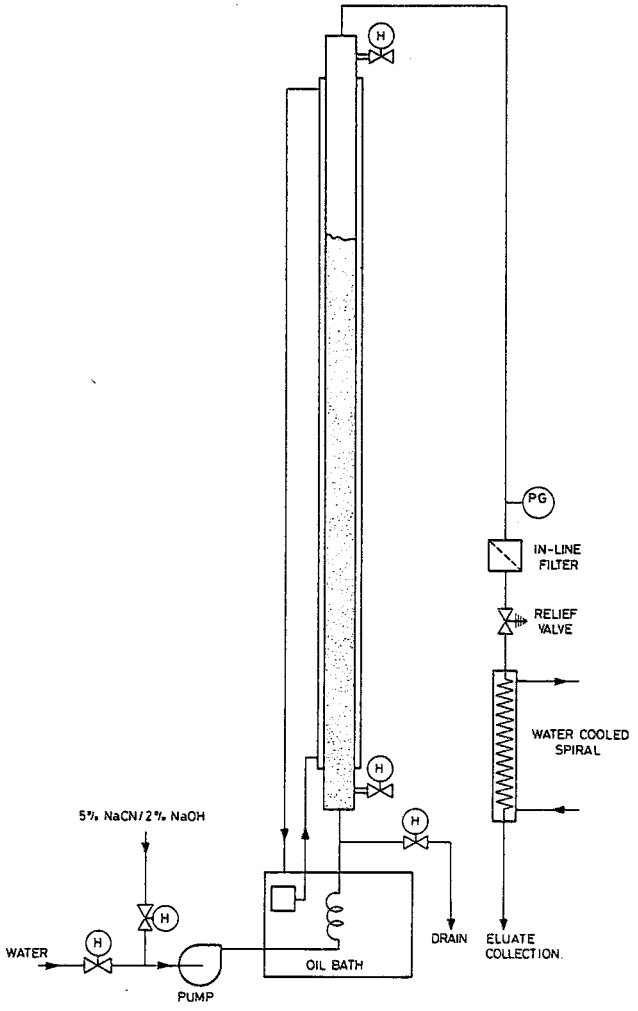

Fig. 2—The gold-elution pilot plant

The elution column together with the oil return-pipe line were well insulated with fibre glass, while pressure within the circuit was maintained at about 300 to 400 kPa by means of an adjustable relief valve. Under the above specified elution conditions, the temperature of the eluate from the top of the column was 109°C. Before being collected, the eluate was cooled by a water-cooled spiral.

Precipitation of Gold

Composite samples of eluate (1 litre each) from the elution column were acidified to a pH value of 1 by the use of sulphuric acid in a sealed 2-litre separating funnel agitated by a stirring mechanism. The resulting precipitate was allowed to settle before being collected for pressure filtration and analysis.

Pilot-plant Results

Adsorption of Gold, Phase I

Phase I entailed the treatment of a bulk solution with a gold content of 0,15 g/t. Table I gives a summary of the gold-adsorption results in the first five completed adsorption cycles during this period. The corresponding breakthrough curves for each metre of carbon through the circuit for the individual cycles are not presented.

TABLE I

OPERATION OF FIXED-BED CARBON PILOT PLANT

Adsorption cycle	Duration d	Average Au in daily effluent composites, g/t			Remarks
		1st column	2nd column	3rd column	
Phase I					
1A 1B 1C	15	0,054	0,011	0,003	
1B 1C 2A	11	0,098	0,050	0,005	
1C 2A 2B	7	0,125	0,063	0,003	
2A 2B 2C	10	0,116	0,038	0,002	
2B 2C 3A	29	0,094	0,040	0,004	20 litres of 3% HCl after 5, 10, 13, 23, and 39 days respectively
Phase II					
2C 3A 3B	0 to 33	0,184	0,055	0,002	20 litres of 3% HCl after 5, 15, 22, 28, and 33 days respectively
2C 3A 3B	34 to 45	0,073	0,020	0,004	3 litres of 3% HCl daily
2C 3A 3B	46 to 54	0,149	0,023	0,005	4 litres of 3% HCl daily
2C 3A 3B*	54 to 58	0,810	0,376	0,033	5 litres of 3% HCl daily
Phase III					
3A 3B 3C	12	0,541	0,137	0,009	4 litres of 3% HCl daily
3B 3C 4A	21	0,208	0,102	0,026	4 litres of 3% HCl daily

*Extremely hot weather on site with the effluent approaching 33 to 40°C

Apart from the duration of the individual cycles, only two operational changes were made. After the system was considered to be approaching equilibrium (illustrated by the considerable flattening of the breakthrough curves), the effects of backwashing and of acid treatment were investigated. In the first instance, during the cycle designated 2A 2B 2C, columns A and B were backwashed after 3 and 5 days respectively. The resulting effect on the gold adsorption was negligibly small, although a small decrease in pressure drop occurred across the circuit. In the second instance, acid washing of the columns during cycle 2B 2C 3A was carried out at

intervals of 5 to 8 days. The effect of the acid wash — 20 litres of 3 per cent (vol./vol.) hydrochloric acid — on the gold adsorption was seen to be most beneficial (see Phase II), with a consequent increase in the cycle time and gold loading. The average cycle time for the previous three cycles was about 9 days, whereas the cycle time for 2B 2C 3A was 29 days. Likewise, the average gold loading was about 1,6 kg/t, which subsequently increased to 3,0 kg/t (see below). The overall gold recovery during this period was more than 97 per cent, yielding effluent values with an average gold content of 0,004 g/t.

In the present investigation, use was made initially of the Bed Depth/Service Time (B.D.S.T.) model [3–5] in an attempt to define the rate of advance of the adsorption reaction front. A plot of service time versus depth of carbon bed for adsorption indicated a linear relationship. For cycle 1A 1B 1C, where a solution front of 0,015 g/t of gold, corresponding to a gold recovery of 90 per cent, was used to define the reaction front, a critical bed depth of 0,70 m with the reaction front moving at 0,46 m per day was obtained.

The B.D.S.T. model thus appeared to have considerable promise in describing gold adsorption in a classical fixed-bed operation. However, with the advent of intermittent acid washing and its obviously beneficial effect on gold loadings and gold adsorption (cycle 2B 2C 3A), the system became too perturbed to be modelled in a simple fashion. Further attempts to model the gold adsorption were thus suspended.

Adsorption of Gold, Phase II

Phase II initially entailed the treatment of a bulk solution containing 0,49 g/t of gold and adjusted to pH 8,0. During the first cycle (2C 3A 3B), four operational changes were made. The first change involved the intermittent acid washing of the circuit with 20 litres of 3 per cent hydrochloric acid on a 5 to 10 day basis during the initial 33 days of the cycle. This acid-washing procedure was then changed to a daily operation entailing 3 litres of 3 per cent hydrochloric acid (34 to 45 days), which was then increased to 4 litres of 3 per cent hydrochloric acid (46 to 53 days), and finally to 5 litres of 3 per cent hydrochloric acid (54 to 58 days). The resulting breakthrough curves corresponding to each metre of carbon-bed depth during this cycle (2C 3A 3B) are presented in Figs. 3 and 4, while a summary of the operating data is given in Table I. The effect of these acid-washing procedures on the gold adsorption can be seen to be most significant. The gold recovery during the period 0 to 54 days exceeded 99 per cent, with gold effluent values averaging 0,003 g/t. The loaded carbon from this cycle had a gold concentration of 21 kg/t. The gold loading actually approached 30 kg/t on the lead column after 53 days but, owing to the extreme temperature conditions encountered on the site during the 54 to 58 day period (the hottest summer temperatures in 53 years, with the column influent approaching 40°C), considerable elution of gold from the lead column to the second column took place. This effect is well illustrated in Fig. 4.

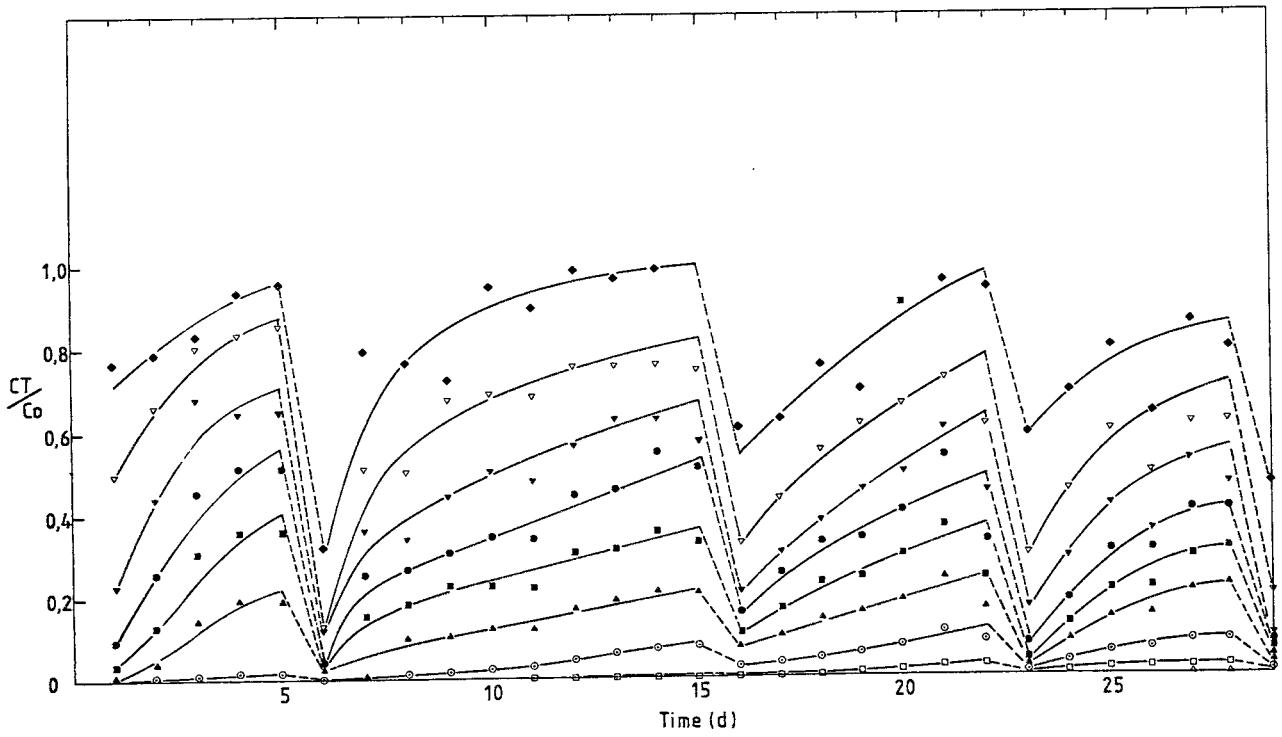

Fig. 3—The effect of periodic acid washing on the gold adsorption in a fixed-bed pilot plant
CT = gold concentration in effluent at specified depth of carbon bed; Co = gold concentration in column influent; breakthrough curves correspond to each metre of bed depth of carbon in the circuit as follows:
◆ = 1 m ▽ = 2 m ▼ = 3 m ● = 4 m
■ = 5 m ▲ = 6 m ○ = 7 m □ = 8 m △ = 9 m

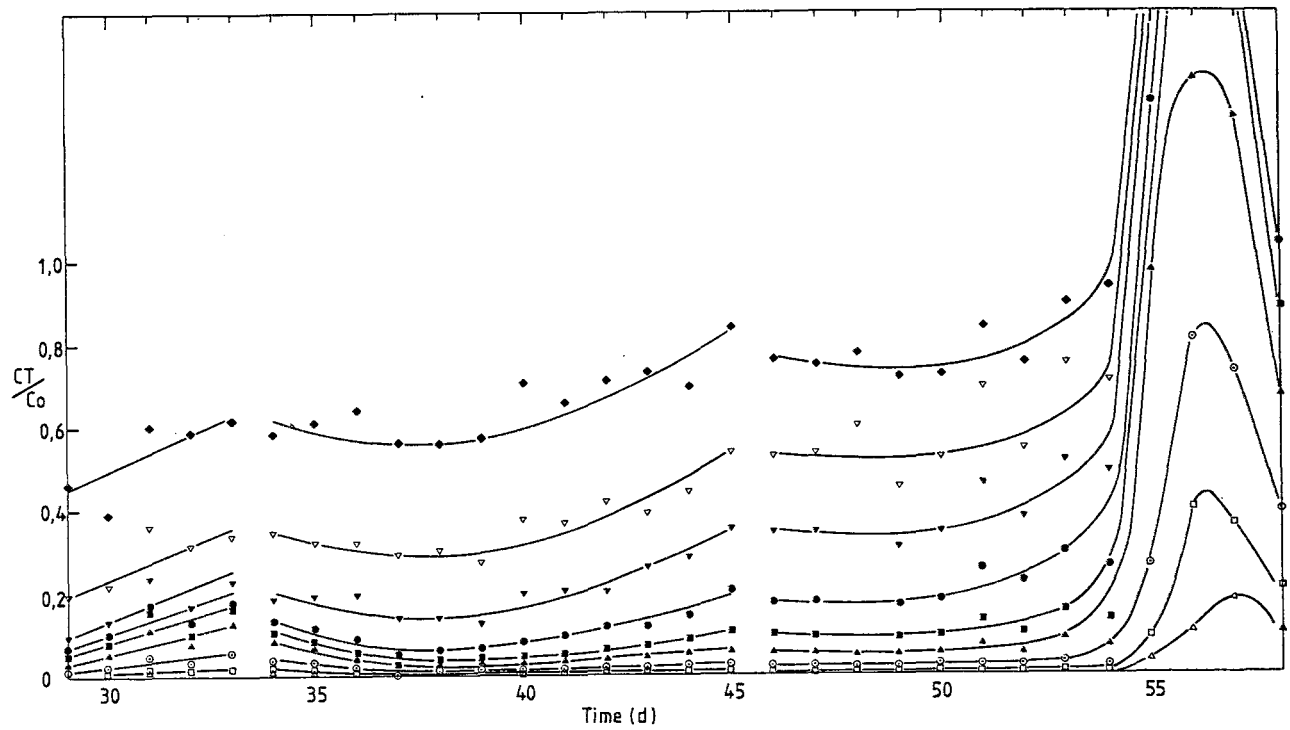

Fig. 4—The effect of daily acid wasing on the gold adsorption in a fixed-bed pilot plant
Designations to the individual breakthrough curves are similar to those in Fig. 3; breaks in the curves coincide with the refilling of the bulk storage tank

Adsorption of Gold, Phase III

The next adsorption cycle (3A 3B 3C of Phase III) was carried out at pH 8,5, using a daily acid wash of 4 litres of 3 per cent hydrochloric acid as described above. The resulting breakthrough curves indicated that the refilling of the bulk storage tank after 9 days of operation had a very serious adverse effect on the gold adsorption. This effect was almost certainly due to an increase in the pH value of the influent. The gold in the column effluent during this period averaged 0,009 g/t, corresponding to a gold recovery of 98 per cent, while the leading column (3A) averaged 16 kg/t of gold and 44 kg/t of copper (Table II).

The next cycle (3B 3C 4A) was carried out at pH 7,0 using a daily acid wash of 4 litres of 3 per cent hydrochloric acid as described above. The daily sampling of effluent from each metre depth of carbon bed was discontinued at this stage of the investigation. Compared with the earlier adsorption results, the gold recoveries were not good during this cycle — the gold in the column effluent averaged 0,026 g/t, corresponding to a gold recovery of only 91 per cent (Table I). Loaded carbon from column 3B averaged 18 kg/t of gold and 58 kg/t of copper. The excessive loading of copper at this lower influent pH may well be the cause of the poor adsorption response observed. (Although copper has been shown to irreversibly 'fix' gold under the prescribed acid-washing conditions, an excessive loading of copper may well physically passivate the carbon.)

Composition of Loaded Carbon

Table II lists the composition of the loaded carbon taken from the circuit during Phases I to III. The high loadings of copper (up to 53 kg/t), zinc (up to 21 kg/t), and nickel (up to 16 kg/t) are immediately evident under normal operation (i.e. no acid washing) and can be compared with the loadings achieved using a daily acid wash, when high loadings of gold (up to 22 kg/t) and copper (28 to 58 kg/t) and low loadings of nickel, zinc, and calcium are evident. There was no progressive build-up of either silica or calcium, while the loading of iron and cobalt was negligible.

Elution of Gold

Owing to the high loadings of base metals, copper in particular (Table II), the elution of the loaded carbon initially presented some problems. The excessive consumption of reagent, yielding low eluate pH levels (pH 8 to 9), resulted in the precipitation of calcium silicate in the eluate pipe lines while also adversely affecting the gold elution. The eventual use of larger volumes of the pretreatment reagent (5 per cent sodium cyanide and 2 per cent sodium hydroxide) was found to overcome these problems. Table III gives further elution results from this aspect of the investigation, while Fig. 5 illustrates the elution of gold and copper from the loaded carbon of 2C (adsorption cycle 2C 3A 3B).

Precipitation of Gold

The acidification of column eluates (see above) to pH 1 by the use of either hydrochloric acid or sulphuric acid resulted in the precipitation of more than 99 per cent of the contained gold from solution[6]. The resulting precipitate was a mixed gold–copper cyanide compound.

Adsorption Isotherms

As a result of the findings obtained from the fixed-bed miniplant, a series of equilibrium adsorption experiments was performed in borate buffer[7] at pH 7 with copper-loaded carbon in both the absence and presence of cyanide. The copper was initially loaded on the carbon as a cyanide complex.

In the absence of cyanide, the adsorbed copper had little influence on the gold adsorption. Linear isotherms were obtained following a log-log plot, compared with the slightly curved isotherms for borate buffer. There was little difference between carbon loaded with 13 kg of copper and that loaded with 32 kg of copper per ton of carbon.

In the presence of 200 mg/l of sodium cyanide, the gold loading was considerably lower for the copper-loaded carbon. The *k*-values were 30,5 and 27,5 kg of gold per ton of carbon for the carbon loaded with 13 and 32 kg of copper per ton of carbon respectively, compared with a *k*-value of 40 kg/t for the virgin carbon. The *k*-value is defined as the gold loading in equilibrium with 1 g/t of gold in solution.

TABLE II

THE COMPOSITION OF LOADED CARBON FROM THE FIXED-BED CARBON PILOT PLANT

Loaded carbon from lead column	Constituent, kg/t						
	Au	Ag	Cu	Ni	Zn	Ca	SiO_2
1A1	1,86	0,02	32,2	11,4	—	7,0	1,61
1A2	1,92	0,03	32,6	14,0	17,0	7,5	1,16
1A3	1,48	0,03	31,6	15,3	—	7,8	1,09
1B1	1,60	0,02	51,0	13,9	—	7,2	
1B2	1,46	0,03	50,3	14,6	—	7,6	
1B3	1,36	0,02	47,9	16,2	—	7,6	
1C1	1,69	0,03	53,4	14,9	20,5	7,3	1,45
1C2	1,56	0,03	48,3	13,6	18,6	7,9	1,22
1C3	1,45	0,03	42,9	12,7	16,6	7,3	1,22
2A1	1,68	0,04	74,6	16,3	0,77	0,20	1,23
2B1	3,41	0,04	55,7	22,2	0,23	0,21	1,31
2B2	2,80	0,04	52,3	17,0	0,17	0,33	1,40
2B3	2,66	0,04	48,3	17,1	0,17	0,47	0,58
2C1	20,10	0,16	28,3	0,82	0,02	0,58	1,25
2C2	21,70	0,18	35,0	1,03	0,03	0,27	0,58
2C3	20,59	0,02	37,6	0,10	0,01	0,32	0,68
3A1	15,19	0,17	45,3	7,44	1,15	5,62	1,55
3A2	16,30	0,19	43,4	6,28	0,88	4,17	1,22
3A3	16,11	0,23	43,4	5,47	0,66	3,45	1,10
3B (Composite)	17,59	0,33	57,8	2,28	0,21	0,56	0,91

Note: The loading of Fe and Co in column 1C averaged 60 and 80 g/t respectively

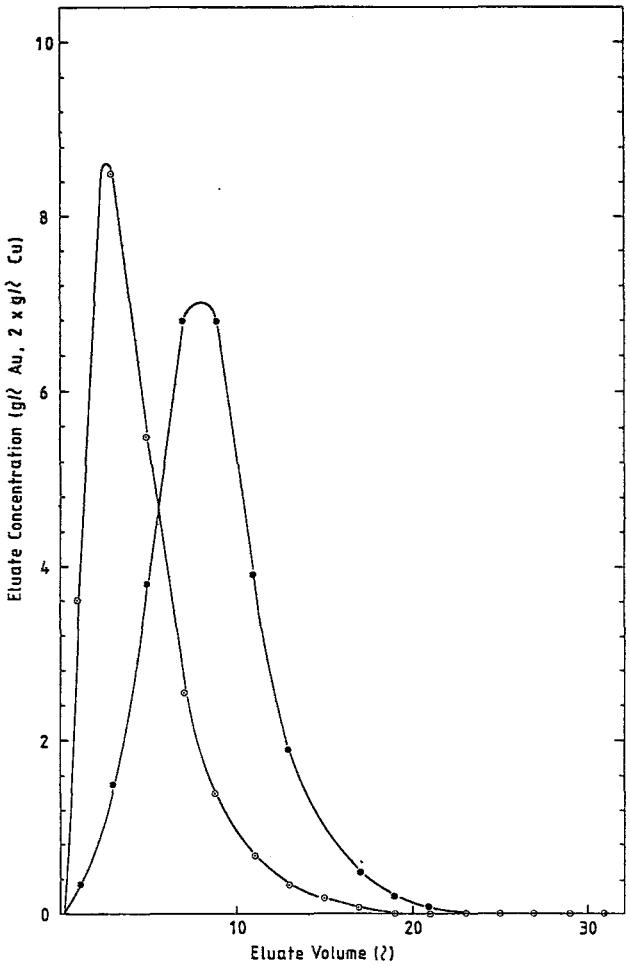

Fig. 5—The elution of gold (shaded) and copper (unshaded) in the elution pilot plant

TABLE III

THE COMPOSITION OF ELUTED CARBON FROM THE FIXED-BED CARBON PILOT PLANT

Eluted carbon from lead column	Pretreatment reagent		Constituent, g/t						
	Quantity, litre	Composition	Au	Ag	Cu	Ni	Zn	Ca	SiO_2
1A	5	5% NaCN–0,5% NaOH	9	1	7	13	626	4350	714
1B	5	5% NaCN–0,5% NaOH	10	1	75	20	49	4170	978
1C	5	5% NaCN–1,0% NaOH	58	5	1055	23	102	4540	1130
2A	6	5% NaCN–2,0% NaOH	116	22	4010	92	89	416	579
2B	6	5% NaCN–2,0% NaOH	300	51	3610	1168	73	354	540
2C	8	5% NaCN–2,0% NaOH	36	37	73	13	16	332	678
3A	8	5% NaCN–2,0% NaOH	127	29	48	—	—	—	—
3B	8	5% NaCN–2,0% NaOH	283	—	50	20	40	2400	1280

Optimum results were obtained with copper-loaded carbon (13 kg of copper per ton of carbon) after acid washing (3 per cent hydrochloric acid) and in the absence of cyanide. After acid washing, the *k*-value was 70 kg compared with 42,5 kg for the virgin carbon. When the carbon was loaded with 13 kg of copper plus 21 kg of gold per ton of carbon and subsequently acid-washed and re-equilibrated with gold solution, a *k*-value of 80 kg of gold per ton of carbon was obtained. (In the presence of 200 mg/l of sodium cyanide, the *k*-values were 28,5 and 31 kg under the same conditions as above, compared with 40 kg for the virgin carbon.)

In many respects, the above results confirm earlier findings[8] on the use of copper sulphate to chemically regenerate used carbon. The formation of a highly insoluble gold–copper cyanide compound following such treatment (see earlier) could explain the irreversible character of the mechanism, together with the abnormally high gold loadings achieved — concentration factors of 3000 to 5000 are often obtained in a conventional carbon-in-pulp operation, which can be compared with the concentration factor of about 60 000 obtained in the present work.

Full-scale Plant

Following the successful pilot-plant testwork, it was decided that a carbon-column plant should be built to treat approximately half the plant effluent from the West Uranium Plant, i.e. 120 kt of solution per month. On site at the plant were 24 obsolete columns that had originally been used in the ion-exchange plant for uranium extraction. Three of these columns were moved over to the gold plant, lined with rubber, and fitted with stainless-steel screens, top and bottom. The column dimensions were 3,70 m long by 2,14 m in diameter. After the screen installation, each column was filled with 4,5 t of G210 carbon (Le Carbone (Pty) Ltd) with a particle size of 0,6 to 1,7 mm. The bulk density of this carbon was 0,55 g/cm³.

Filtrate from the uranium plant was clarified in gold-plant leaf clarifiers and then pumped to 500 m³ holding tanks, where the pH value was adjusted to 7,5 as in the pilot-plant operation. These tanks were open to the sunlight, which would assist in destroying the cyanide. The pilot-plant testwork had shown that this was the optimum pH value for gold loadings, possibly because it reduced the cyanide content of the feed and increased the loading of copper. Several problems were experienced on the initial water commissioning owing to buckling of the screens and subsequent bypassing of carbon. Thus, the carbon had to be removed and more substantial screen supports installed with associated rubber sealing. After one week of water commissioning, gold-bearing filtrate was introduced to the columns in March 1982. Composite samples were taken daily from each column so that an accurate picture could be obtained of the gold loadings across the system.

Results for the first hundred days of operation are presented in Table IV. As can be seen, the progressive gold adsorption decreased steadily in the lead column (A) over the 100 days, but the barren solution from column C rose to 0,021 g/t of gold with an average plant-feed value of 0,145 g/t. A decision was made to keep the same operating sequence until a definite gold 'breakthrough' occurred in the lead column (A). This actually happened after 112 days of operation, when the samples of feed and barren solution for column A contained 0,094 and 0,208 g/t respectively. Whether this was due to self-elution was not known.

TABLE IV

OPERATION OF THE FULL-SCALE PLANT

Running time d	Progressive gold adsorption %	Gold loading, kg/t		
		Column A	Column B	Column C
1	99,23	0,1015	0,0023	—
10	98,79	0,8712	0,0207	0,0040
20	98,40	1,3609	0,1241	0,0263
30	98,50	2,0981	0,3736	0,0560
40	98,32	2,7781	0,6266	0,1391
50	97,38	3,3677	0,9258	0,2628
60	95,82	3,8477	1,3046	0,4257
70	95,60	4,4556	1,5450	0,6238
80	95,24	4,9174	1,8530	0,8833
90	93,95	5,0532	2,0210	1,1760
100	93,26	5,1571	2,2025	1,3039

The plant was closed down and backwashed with regional water, and the loaded carbon was pumped out of the lead column. After the column had been recharged with fresh carbon, the plant was restarted in the merry-go-round sequence. Since then, this self-elution or breakthrough phenomenon has occurred at regular intervals of 100 to 120 days, after which the loaded carbon in the lead column has been recharged. The recovery of gold from the plant effluent has, however, averaged 97 per cent over a 12-month period.

Loaded carbon from the columns was withdrawn, dried, and thoroughly mixed. The samples were analysed at V.R.E.M., A.A.R.L., and the Rand Refinery laboratories, and it is gratifying to note that the three gold loadings recorded were always within very close agreement. For example, the first batch of carbon analysed by the three laboratories gave gold loadings of 5,0 kg/t, 5,1 kg/t, and 5,1 kg/t. Up to now, three batches of carbon have been incinerated at the Rand Refinery, giving a total mass of gold recovered of 72 kg, i.e. approximately 24 kg of gold per loaded column. As the total capital cost of the plant including the first 17 t of carbon was R210 000, it can be seen that the first batch of loaded carbon paid for the plant. The operating costs were extremely low excluding the carbon costs.

At present, a second series of three columns is being erected and modified alongside the first full-scale plant, to treat all the effluent from the West Uranium Plant. One of the disappointing features of the full-scale operation was its inability to treat 120 kt of effluent per month, which corresponds to a feed rate of 166 t/h. The plant was able to run at only 120 t/h, which slowly decreased to 80 t/h. As a matter of routine, the columns were treated daily with 0,2 bed volume of 3 per cent (vol./vol.) hydrochloric acid, which resulted in a pressure drop across the circuit and thus maintained acceptable feed rates. When the feed rate was increased to more than 120 t/h, the stainless-steel support screens flexed, with a resultant

loss of carbon to the spillage area. The columns were not backwashed at regular intervals because of the absence of suitable screens positioned at the top of each column. Such backwashing would have improved the plant capacity. Thus, for the six-column plant, the screens will be six radial spirally wound distribution candles at the top and bottom of each column. These will not be able to flex, and it is hoped that they will promote higher throughputs and gold loadings because of better feed distribution. The six-column plant was scheduled for commissioning in April 1983.

As far as future work is concerned, an exercise is being carried out in conjunction with V.R.E.M. East Division. That division is planning to commission a similar plant in 1983, and a critical survey is being conducted of the economics of an elution–regeneration system for the loaded carbon from both divisions. No decision had been made at the time of writing.

Discussion

Vaal Reefs West Gold Plant, like many other South African plants, utilizes rotary-drum filters to dewater the leached slime prior to gold precipitation. This process obviously depends very much upon vacuum efficiency, filter-cloth and wash efficiency, etc. Thus, a finite amount of gold always leaves the plant as effluent gold and reports to the slimes dams. With the introduction of the carbon columns as a 'scavenger' circuit, this gold loss has been reduced substantially. The cost of the gold recovery, excluding carbon, is very low.

An added advantage is that the barren solution from the columns is extremely clear, and is bled straight into water tanks as a replacement for river water. Not only does this contribute directly to a saving of river water, but it also results in a cost saving seeing that river water is currently charged at about 4 cents per cubic metre. If the full-scale plant generates approximately 200 kt of solution per month, this represents a saving of R8000 per month to the company.

The evidence presented in this study indicates the very significant role played by copper in gold adsorption, elution, and precipitation. The formation of a highly insoluble copper–gold complex in all three processes is strongly suggested. When the evidence is related to the gold-adsorption mechanism proposed by McDougall *et al.*[9], it strongly supports the concept that the specific character of gold adsorption is related to the solubility of the adsorbed species.

When this mechanism is related to established observations, it becomes evident how decreases in temperature, cyanide concentration, and pH all tend to decrease the solubility of the adsorbed species. The subsequent acid-washing of the loaded carbon and the resultant formation of a highly insoluble gold–copper complex (see gold precipitation) would thus effectively regenerate the carbon and so promote further gold adsorption. Of course, such a mechanism would proceed only at low concentrations of cyanide and/or at low pH values owing to the solubility of the complex in alkaline cyanide solutions[6]. Such a mechanism can be expected to become very sensitive to changes in temperature, cyanide concentration, and pH at high gold loadings, which would explain the self-elution of gold that was observed in both the pilot-plant and the full-scale operations. It can thus be seen that the formation of this insoluble gold–copper complex appears to play a most significant role — not only in the adsorption of gold, where significant effects have been demonstrated — but in the elution of gold and its subsequent precipitation after acidification of the eluate. A simple reversal of the chemistry involved in each instance would explain many of the observations made in the present investigation.

Acknowledgements

This paper is published by permission of the management of the Anglo American Corporation. The fixed-bed pilot-plant operation was a collaborative project between Vaal Reefs Exploration and Mining Company and Anglo American Research Laboratories. The contribution of C. Robertse in operating both the pilot plant and the Phase I plant is gratefully acknowledged, as is the funding of the project by the Gold Division of the Anglo American Corporation.

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3.3 COUNTER CURRENT MOVING BED

The concept of recovering gold values from clarified plant solutions using a counter current moving carbon bed was first developed at the AARL. Following several pilot plant campaigns, a full scale moving bed operation treating ca 800 m^3/h of plant solution was commissioned at Western Deep Levels No 3 shaft for the treatment of waste-rock dump material and current gold plant residues. Gold recoveries from solution in this plant average more than 99% from a solution feed of ca 0,5 g/t gold.

The appended paper entitled 'The Recovery of Gold from Plant Solutions by use of a Counter-Current Moving Bed' was awarded a bronze medal by the SAIMM.

J. S. Afr. Inst. Min. Metall., vol. 85, no. 11.
Nov. 1985. pp. 387-394.

The recovery of gold from plant solutions by use of a countercurrent moving carbon bed

by R.J. DAVIDSON*, W.D. DOUGLAS†, and J.A. TUMILTY‡

SYNOPSIS

Two pilot plants operating on different types of carbon were used to generate steady-state adsorption data in the treatment of gold-plant solutions of high and low tenor.

A pilot plant comprising three 4 m long columns and using 3 mm extruded activated carbon was used to treat pregnant solutions with a gold concentration of 5 to 8 g/t at pH 10 to 11,5. Column influent was pumped at a superficial flow velocity of 0,5 m/min while the carbon was moved daily so as to effect gold loadings on the carbon of 5 to 10 kg/t. With a carbon-bed depth of 12 m, an adsorption efficiency of 99,9 per cent was obtained in a bed depth of 9 m, which was subsequently reduced to 8 m when the depth of the carbon bed was reduced to 8 m. This improvement in efficiency was attributed to the reduced residence time of the carbon from 12 to 8 days. A corresponding decrease in adsorbed silica indicated that some silica fouling of the carbon had occurred.

A second pilot plant, comprising a single column 6 m in length and using coconut-shell carbon of 1,0 to 2,4 mm, treated a solution with a gold concentration of 0,7 g/t at pH 8,0. Feed solution was pumped at a superficial flow velocity of 1,0 m/min while carbon was moved every 8 hours to give gold loadings of more than 5 kg/t. An adsorption efficiency of 99,5 per cent was indicated at steady state. Intermittent washing of the column with dilute hydrochloric acid was shown to have a beneficial effect on the gold adsorption and to reduce the pressure drop across the column. Intermittent washing with dilute sulphuric acid indicated that there had been some passivation of the carbon during the treatment of a solution with a gold concentration of only 0,2 g/t.

Compared with other adsorption systems, the countercurrent moving bed appears to offer significant advantages in terms of metallurgical efficiency and capital investment. The inability normally associated with moving-bed systems to treat solutions containing suspended solids may be overcome to a degree by the use of relatively coarse extruded carbon.

SAMEVATTING

Twee proefaanlegginge wat met verskillende soorte koolstof werk, is gebruik om bestendige adsorpsie data in verband met die behandeling van goudaanlegoplossings met 'n hoë en lae goudkonsentrasie te ontwikkel.

'n Proefaanleg, bestaande uit drie 4 m-lange kolomme wat van uitgeperste geaktiveerde koolstof van 3 mm gebruik maak, is gebruik om pregnante oplossings met 'n goudkonsentrasie van 5 tot 8 g/t by 'n pH van 10 tot 11,5 te behandel. Die kolominvloeisel is met 'n oppervlakvloeisnelheid van 0,5 m/min gepomp, terwyl die koolstof daagliks beweeg is ten einde goudladings van 5 tot 10 kg/t op die koolstof te kry. Met 'n koolstofbeddiepte van 12 m is 'n adsorpsiedoeltreffendheid van 99,9 persent in 'n beddiepte van 9 m verkry, en dit het later tot 8 m afgeneem toe die diepte van die koolstofbed tot 8 m verminder is. Hierdie verbetering in die doeltreffendheid is toegeskryf aan die verkorting van die koolstof se verblyftyd van 12 tot 8 dae. 'n Ooreenstemmende afname in die geadsorbeerde silika het aangedui dat daar 'n mate van versperring van die koolstof deur die silika plaasgevind het.

'n Tweede proefaanleg, bestaande uit 'n enkele kolom met 'n lengte van 6 m wat van kokosneutdopkoolstof van 1,0 tot 2,4 mm gebruik maak, het 'n oplossing met 'n goudkonsentrasie van 0,7 g/t by 'n pH van 8,0 behandel. Die toevoeroplossing is met 'n oppervlakvloeisnelheid van 1,0 m/min gepomp terwyl die koolstof elke 8 uur beweeg om goudladings van meer as 5 kg/t te gee. 'n Adsorpsiedoeltreffendheid van 99,5 persent is met 'n bestendige toestand verkry. Daar is getoon dat sporadiese was van die kolom met verdunde soutsuur 'n voordelige uitwerking op die goudadsorpsie het en die drukval oor die kolom verlaag. Sporadiese was met verdunde swaelsuur het getoon dat daar net tydens die behandeling van 'n oplossing met 'n goudkonsentrasie van 0,2 g/t 'n mate van passivering van die koolstof was.

Vergeleke met ander adsorpsiestelsels bied die teenstroom-bewegende bed blykbaar beduidende voordele in terme van metallurgiese doeltreffendheid en kapitaalbelegging. Die gewone onvermoë van bewegendebed-stelsels om oplossings met gesuspendeerde vaste stowwe te behandel, kan in 'n mate oorkom word deur die gebruik van betreklik growwe uitgeperste koolstof.

Introduction

With the advent of the carbon-in-pulp process in the South African gold-mining industry, considerable attention has been focused on the treatment of plant solutions and effluents with activated carbon. Such circuits as dam return water, uranium-plant primary filtrate[1], and dump leach liquors are currently receiving serious consideration. As the technology involved in the treatment of such solutions is currently very much in a state of development, numerous circuits have been considered.

Because of the complex confounding effects of numerous parameters on the adsorption of gold onto activated carbon from plant solutions and effluents, no simple formula can be used in the design of a recovery circuit. Very often a compromise must be made in terms of both the chemistry and the engineering of the system, while in most instances realistic design data can be obtained only after lengthy and costly pilot-plant operation. It is nevertheless necessary for the effectiveness of cer-

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SA ISSN 0038-223X/$3.00 + 0.00. Paper received 22nd October, 1984.

tain operational parameters to be assessed before a particular plant design can be chosen. The options include individual stages of fixed or fluidized beds of carbon arranged in a countercurrent circuit, or the use of fixed or fluidized stages contained in a single column.

Carbon contacting is normally carried out in a system of one or more columns. Three principal modes of operation can be used: expanded-bed upflow, fixed-bed downflow, and countercurrent moving bed. The design and mode of operation of a contacting circuit depend largely on the clarity of the solution or effluent, together with the gold content. Such characteristics define to some degree the particle size of the carbon, the contact time, the gold loadings, and the choice of a steady-state as opposed to a non-steady-state type of operation.

Expanded-bed Upflow

In this mode of operation, the pregnant effluent or solution enters the bottom of the column and leaves from the top. The carbon bed, in an expanded form, remains in the column until suitably loaded, at which time the carbon is removed for elution and regeneration. This mode of operation is especially useful for the treatment of unclarified effluents and when the clarity of the product water is of no concern. No significant filtration is effected, which may be a disadvantage if water reclamation is anticipated. The backwashing of stages is normally not required.

The specification of the number of stages in series is an integral part of the process design of an expanded-bed or fixed-bed mode of operation. As such staged systems cannot normally be operated to yield equilibrium gold loadings, there is an under-utilization of the capacity of the carbon. This, however, may be an important advantage if fluctuating gold values are encountered in that the second or third stages will contain a substantial amount of active carbon. Such adsorption circuits are normally operated in a typical carousel mode in order to promote the efficiency of the countercurrent adsorption.

The process design of both the expanded-bed upflow and fixed-bed downflow systems is most complex because of the non-steady-state modes of operation.

Fixed-bed Downflow

This mode differs from the expanded-bed mode only in that the process solution is passed from the top to the bottom through the column, and that the carbon bed is compacted. Accordingly, a substantial filtration effect is obtained. Furthermore, because of the static nature of the carbon bed, a strong chemical driving force is maintained in each stage, which is most desirable.

The fixed-bed mode of operation is therefore best suited to the treatment of clarified solutions or where water reclamation is anticipated.

Countercurrent Moving Bed

In this mode of operation, process water enters at the bottom of the column and leaves from the top. Fresh carbon is fed intermittently in slugs to the top of the bed, while loaded carbon is similarly discharged from the bottom. To minimize the mixing of carbon in the column, the carbon bed is normally not expanded. This system, applied to the recovery of gold rather than to the treatment of wastewaters, is the subject of the current investigation.

Pilot Plants

In the investigation described here, two pilot plants were used in which carbon was periodically slugged in and out of the system in a moving-bed type of operation. Pilot operations were carried out at the Western Deep Levels Gold Mine (WDL) and at Vaal Reefs Exploration & Mining Co. Ltd (VREM), in which different clarified plant solutions and types of carbon were used.

Plant A

The object in this investigation was to demonstrate efficient gold adsorption at relatively high flow velocities by the use of a coarse (3 mm) extruded carbon. The use of this carbon had demonstrated the feasibility of pumping a conventional gold-plant pulp containing 40 to 50 per cent solids through a single column in a 'pinned-bed' countercurrent moving-column type of operation[2]. This demonstration had suggested a similar treatment for unclarified gold-plant solutions and effluents, which would obviate clarification costs.

The pilot plant comprised three Perspex columns in series (each column being 50 mm in internal diameter and 4 m in length) in which slugs of carbon were moved countercurrently from one column to the other to simulate a single moving-bed column (Fig. 1). The carbon was moved intermittently by the volumetric discharging of a discrete bed depth of carbon from each column via a ball valve and the respective columns were refilled while maintaining the profile of a single adsorption column. The carbon used throughout the operation was 3,0 mm type RB3 extruded activated carbon (Norit Activated Carbon, Amersfoort, Netherlands).

Clarified pregnant solution was stored in a constant-head tank (1 m^3 capacity) and pumped into the columns by a diaphragm metering pump. Solution was fed to the plant at a rate of 1,0 l/min, corresponding to a superficial flow velocity of 0,5 m/min in the columns.

A profile of gold in solution was obtained daily from sample points placed at 1 m intervals down the bed. Samples were taken progressively down the column starting from the barren solution to the feed solution so that the solution profile would be disrupted as little as possible.

The column operated continuously save for brief interruptions while the carbon was moved. A defined downtime of 0,5 hour was allowed for moving the carbon. Each phase in the investigation involved at least 2 to 4 weeks so that steady-state adsorption data could be generated.

Plant B

The investigation on Plant B was used to generate design data for the proposed moving-bed column to treat clarified leach liquors from the No. 3 waste-rock dump project at WDL. The project entailed the conversion of the existing leaching and countercurrent-decantation circuits of the uranium plant to recover gold from milled waste rock at the No. 3 Shaft. The pilot operation was launched as a joint venture between WDL and Anglo American Research Laboratories.

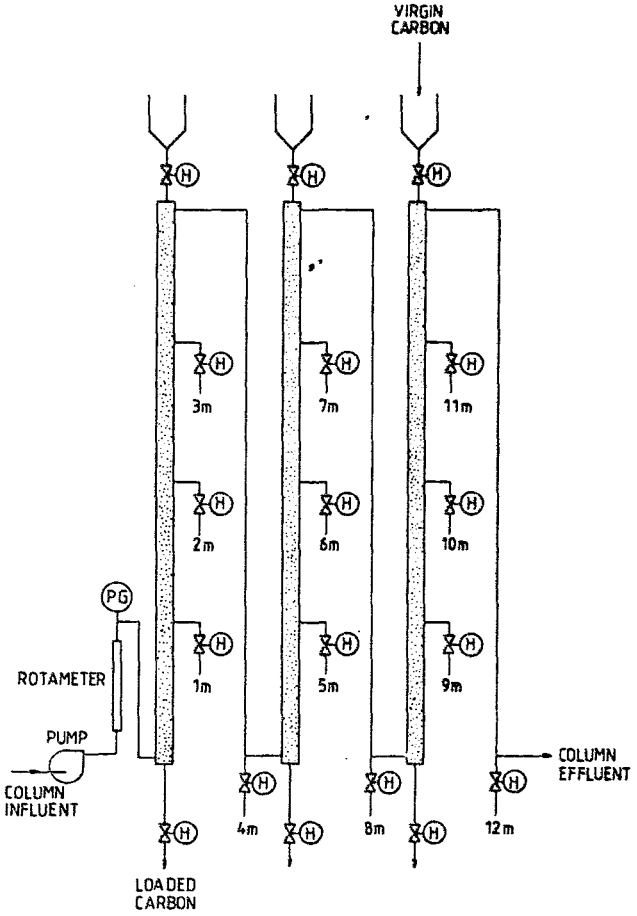

Fig. 1—Pilot plant A

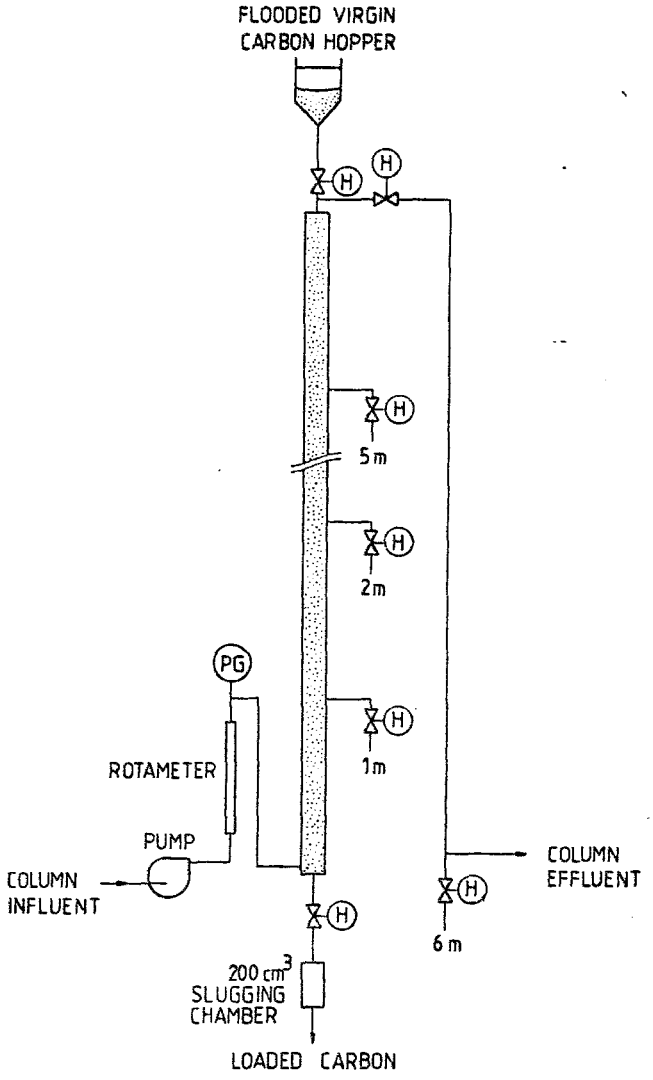

Fig. 2—Pilot plant B

The pilot plant consisted of a single Perspex column having an internal diameter of 50 mm and a length of 6 m. Carbon was fed into the top of the column via a flooded hopper, and was slugged periodically from the base of the column via a small 200 cm^3 slugging chamber fitted with an 850 μm screen. Fig. 2 illustrates the flowsheet of the circuit. Use was made of 1,0 to 2,4 mm Type G210-AS coconut-shell activated carbon (Le Carbone (Pty) Ltd) throughout the operation so that the particle size of the carbon in other proposed carbon circuits at WDL could be standardized.

Unless otherwise specified, clarified gold-plant pregnant solution was suitably diluted with Merrill barren solution so that its gold concentration was 0,7 g/t. This solution was stored at pH 10,5 in an 80 m^3 storage tank in which the content of free cyanide was adjusted periodically to a sodium cyanide concentration of 100 g/t. The pH value of this solution was then lowered to pH 8,0 by use of an automated titrimeter and 10 per cent sulphuric acid (by volume) in an in-line 200-litre agitated tank prior to being pumped to the column. A diaphragm metering pump was used to pump the solution at a flowrate of 1,96 l/min, which corresponds to a superficial flow velocity of 1,0 m/min in the adsorption column.

As before (Plant A), a profile of gold in solution was obtained daily from sample points corresponding to each metre of carbon in the column. As the movement of the carbon was effected in less than 10 minutes, the operational time was 98 to 100 per cent. Each phase in the investigation involved at least 2 to 3 weeks of continuous operation so that steady-state adsorption data would be obtained.

Only clarified solutions were used as column influents so that stable flowrates would be maintained by the diaphragm metering pumps. For this reason, the effect of suspended solids on the operation of the prescribed pilot operations could not be assessed.

Results Obtained on Plant A

Phase I

Gold-plant solution from WDL with gold and sodium cyanide concentrations of approximately 5 g/t and 150 g/t respectively (pH 10 to 11,5) was used as the column influent. The plant was operated for 5 days with no carbon movement so as to allow an initial adsorption profile to form. Subsequently, 2 litres of carbon, equivalent to a bed depth of 1 m, were moved daily.

Although the solution data are perhaps of more general interest, the adsorption data are presented in Table I in terms of progressive percentage recoveries so that they can be compared easily. From these data it can be seen

that the plant reached steady state in a relatively short time, maintaining a gold front of 0,005 g/t (99,9 per cent recovery) within 9 m for the initial 19,6 operating days. As the last 4 m of carbon in the circuit was shown to be adsorbing very little gold, the total bed depth of carbon in the circuit was reduced from 12 to 8 m (only 2 columns) in the subsequent 8,8 days of operation. This shortening of the circuit, which reduced the carbon residence time from 12 to 8 days, appeared to have a beneficial effect since the 0,005 g/t gold front was subsequently maintained in the 8 m long circuit. (During that time, the adsorption data presented in Table I indicate a slight decrease in gold adsorption in the first 6 m carbon, the improved performance of the remaining 2 m of freshly added carbon being clearly evident.)

Carbon loadings for the period under review are presented in Table II, which gives gold loadings of 6 to 9 kg/t, silver of 1 kg/t, copper of 1 kg/t, calcium of 9 to 12 kg/t, and silica of 13 to 18 kg/t. While the rather inconsistent gold loadings are due partly to fluctuations in the gold content of the feed solution (Table I), the decrease in silica loadings from approximately 18 to 13 kg/t towards the end of the operation (days 24 to 28) may well be significant. This decrease in silica may explain to some degree the improved gold adsorption resulting from the shorter carbon residence time after 19,6 operating days.

Phase II

Pregnant solution from the gold plant at VREM South Division was used as column influent. It had a gold concentration of approximately 8 g/t and a sodium cyanide concentration of 200 g/t (pH 10 to 11,5). Based on the experience gained in the WDL exercise (Phase I), the depth of carbon bed was 8 m (only 2 columns). So that steady state would be approached in a shorter time, this phase was started with the same carbon profile in the column as that at the completion of Phase I. As before, 2 litres of carbon, equivalent to a bed depth of 1 m, were moved daily through the circuit.

The daily recoveries are shown in Table III, from which it can be seen that relatively good steady-state data were achieved while the gold front was maintained at 0,003 g/t (99,96 per cent) within 7 m. Since it took at least 8 days for a complete change in carbon inventory, this is the minimum operating time before the column can be

TABLE I
PROGRESSIVE PERCENTAGE GOLD RECOVERY FROM PLANT A AT WDL (PHASE I)

Operating time d	Gold in column influent g/t	Depth of carbon bed, m									
		1,0	2,0	3,0	4,0	5,0	6,0	7,0	8,0	9,0	10,0
4,0	3,78	38,62	70,11	90,90	90,06	98,94	99,68	99,87	99,97	99,97	99,97
5,0	5,02	46,02	74,66	93,57	97,85	99,49	99,85	99,93	99,98	99,98	99,98
5,9	5,14	41,73	70,77	88,56	95,75	98,48	99,51	99,78	99,93	99,97	99,98
6,9	5,50	35,62	64,94	82,89	92,51	97,04	99,10	99,66	99,89	99,96	99,98
7,9	5,44	41,04	69,56	87,06	93,79	97,72	99,36	99,81	99,93	99,97	99,98
8,9	5,40	41,61	74,63	89,94	94,90	98,31	99,55	99,85	99,96	99,98	99,98
9,7	5,48	41,54	75,09	88,93	94,64	98,03	99,50	99,81	99,94	99,98	99,98
10,8	4,80	41,73	72,86	86,63	94,61	97,66	99,24	99,76	99,90	99,97	99,98
11,7	4,78	41,02	75,57	85,34	93,38	96,76	98,65	99,51	99,78	99,77	99,98
12,7	4,34	41,45	73,68	83,77	90,04	94,95	97,55	98,95	99,54	99,85	99,95
13,7	4,38	37,39	65,71	80,28	88,39	93,81	96,72	98,47	99,35	99,74	99,91
14,6	4,42	36,36	66,02	81,93	90,95	95,34	97,78	98,97	99,58	99,86	99,94
15,6	4,38	31,36	66,36	82,73	90,45	95,39	98,28	99,23	99,72	99,90	99,97
16,6	4,96	31,91	64,88	81,05	89,06	94,99	98,00	99,16	99,73	99,89	99,96
17,6	5,10	38,37	65,61	81,81	90,46	95,73	98,21	99,32	99,77	99,92	99,97
18,6	4,98	38,89	67,56	84,52	91,87	96,23	98,51	99,41	99,75	99,92	99,98
19,6	5,12	39,01	67,82	85,45	92,57	96,44	98,65	99,50	99,81	99,95	99,98
Average	4,88	39,04	69,76	85,61	92,78	96,78	98,72	99,47	99,80	99,93	99,97
Depth of carbon bed reduced from 12 to 8 m											
20,6	4,98	39,90	69,80	86,24	93,56	97,28	99,02	99,66	99,89		
21,6	5,08	43,14	72,56	86,66	93,94	97,89	99,39	99,88	99,98		
22,5	5,12	44,71	72,45	85,78	93,24	97,51	99,43	99,90	99,98		
23,5	5,30	40,21	69,77	84,84	93,41	97,76	99,62	99,94	99,98		
24,3	5,54	28,69	56,27	71,22	82,87	91,75	96,95	99,30	99,88		
25,4	5,26	20,37	49,91	61,02	78,98	88,54	95,68	98,94	99,78		
26,4	5,30	25,95	61,84	75,76	91,42	95,38	98,45	99,57	99,88		
28.3	5,38	30,99	65,07	85,67	96,10	99,00	99,76	99,93	99,98		
Average	5,25	34,24	64,71	79,68	90,44	95,64	98,54	99,64	99,92		

TABLE II
CARBON LOADING DATA FROM PLANT A AT WDL (PHASE I)

Operating time d	Carbon loading, kg/t				
	Au	Ag	Cu	Ca	SiO_2
5,0	12,3	0,8	1,0	10,1	19,7
6,9	14,7	0,8	0,9	10,1	19,9
7,9	11,9	1,3	1,0	9,0	18,4
8,9	10,2	1,2	1,1	8,8	18,8
9,7	9,4	1,0	1,1	8,7	18,4
10,8	10,2	1,2	1,0	9,0	18,4
11,7	9,3	1,2	1,0	9,5	18,8
12,7	9,4	1,2	1,0	7,4	14,3
13,7	8,4	1,2	1,1	9,7	18,8
14,6	7,5	0,7			
15,6	8,0	1,1	1,0	9,8	17,5
16,6	7,4	1,2	1,1	10,7	17,2
17,6	7,2	1,0	1,0	11,2	17,9
18,6	7,2	0,9	0,9	10,8	17,5
19,6	7,6	1,0	1,0	12,9	17,5
Depth of carbon bed reduced from 12 to 8 m					
20,6	7,8	1,0	1,0	19,6	17,9
21,6	7,1	1,1	1,1	14,9	17,5
22,5	5,1	0,8	0,8	9,7	17,9
23,5	6,8	1,0	1,1	12,5	17,2
24,3	6,8	1,0	1,1	11,6	14,8
25,4	5,7	0,8	1,0	10,3	13,4
26,4	6,5	0,8	1,0	12,3	14,3
28,3	11,1	1,1	1,2	12,2	13,6

TABLE IV
CARBON LOADING DATA FROM PLANT A AT VREM (PHASE II)

Operating time d	Gold loading kg/t
1,0	10,5
2,0	9,2
2,9	9,7
3,9	12,2
4,9	10,8
5,9	9,7
6,9	10,4
7,8	11,2
8,8	11,2
9,8	11,8
10,8	11,4
11,8	15,4
12,7	15,0
13,7	12,5
14,7	10,2

TABLE III
PROGRESSIVE PERCENTAGE GOLD RECOVERY FROM PLANT A AT VREM (PHASE II)

Operating time d	Gold in column influent g/t	Depth of carbon bed, m							
		1,0	2,0	3,0	4,0	5,0	6,0	7,0	8,0
1,0	8,38	58,23	84,49	97,46	99,16	99,98	99,89	99,98	99,99
2,0	7,35	42,28	70,76	91,97	99,14	99,84	99,94	99,97	99,99
2,9	8,67	38,58	69,29	90,26	98,63	99,71	99,88	99,97	99,99
3,9	8,27	47,87	76,98	92,84	97,58	99,55	99,85	99,98	99,99
4,9	6,98	43,08	75,61	91,72	97,02	99,50	99,86	99,97	99,99
5,9	6,65	41,60	76,60	91,49	97,30	99,61	99,89	99,96	99,99
6,9	7,07	43,73	73,91	90,38	97,33	99,59	99,93	99,96	99,99
7,8	8,33	45,00	75,32	91,30	97,13	99,54	99,91	99,96	99,98
8,8	6,60	45,55	76,96	92,57	97,49	99,56	99,88	99,96	99,98
9,8	7,86	45,99	77,59	92,46	97,65	99,35	99,83	99,94	99,99
10,8	8,19	45,79	76,20	92,21	97,62	99,40	99,84	99,95	99,99
11,8	7,60	44,14	72,45	91,07	97,77	99,61	99,91	99,97	99,99
12,7	7,13	40,94	65,51	87,78	97,49	99,52	99,88	99,97	99,98
13,7	7,27	39,17	61,88	84,72	97,33	99,45	99,87	99,96	99,97
14,7	7,66	39,92	64,17	84,66	96,42	99,17	99,78	99.92	99,95
Average	7,60	44,12	73,18	90,86	97,67	99,55	99,88	99,96	99,98

considered to have reached steady state. Hence, the data from days 9 to 15 can be regarded as representing steady-state conditions. During this period, consistent gold loadings of 10 to 11 kg/t were obtained (Table IV), while the gold-loading profile based on solution-adsorption data also indicated a very high chemical driving force: loaded carbon = 10,9 kg/t, 1 m = 6,1 kg/t, 2 m = 3,2 kg/t, 3 m = 1,2 kg/t, 4 m = 0,29 kg/t, 5 m = 0,06 kg/t, 6 m = 0,01 kg/t, 7 m = 0,002 kg/t, and 8 m = <0,001 kg/t.

Results Obtained on Plant B

During the course of the investigation, there were two tenors of gold in the bulk solution fed to the column so that gold loadings on the carbon of more than 5 kg/t could be demonstrated. While a gold concentration in

solution of 0,7 g/t was considered to be the expected average feed value for the large-scale operation, a gold concentration of only 0,2 g/t was considered to represent the lowest gold value likely to be encountered. Likewise, while the concentration of free cyanide in the column influent of the large-scale operation was difficult to forecast because considerable decomposition of cyanide was likely to take place in the countercurrent-decantation circuit, a free-cyanide level of 100 g of sodium cyanide per ton was considered to represent the highest value likely to occur.

Phase I

A bulk solution with a gold concentration of 0,7 g/t and a sodium cyanide concentration of 100 g/t was used as column influent. At regular intervals during each 8 hour shift, i.e. 3 times daily, 200 cm^3 of carbon (equivalent to a bed depth of 0,1 m) was slugged in and out of the column. The resulting adsorption data corresponding to each metre of carbon in the column, together with the daily composite carbon loadings for the period under review, are presented in Fig. 3 and Table V respectively.

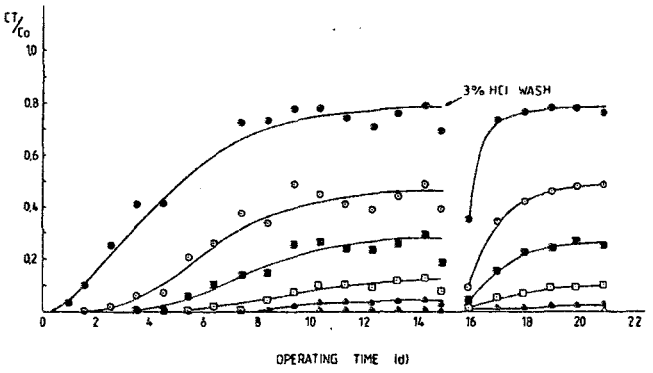

Fig. 3—Gold-adsorption data from pilot plant B (Phase I). C_T = gold concentration in effluent at specified depth of carbon bed; C_o = gold concentration in column influent; adsorption curves correspond to each metre of carbon in the circuit as follows:

● = 1 m o = 2 m ■ = 3 m
□ = 4 m ▲ = 5 m △ = 6 m

After 15 days of continuous operation with the plant approaching steady state, the following carbon loadings were obtained: gold 5,1 kg/t, silver 0,5 kg/t, copper 17 kg/t, nickel 3,3 kg/t, zinc 21 kg/t, and calcium 5 kg/t (Table V). At that stage, the circuit was treated with acid, 3 per cent hydrochloric acid (by volume) being pumped for 10 minutes to alleviate a blockage in the circuit. (A build-up of calcium carbonate precipitate in the storage tank was thought to be responsible.) The acid treatment had a marked effect in lowering the pressure drop across the column and in improving the gold adsorption, the gold loadings increasing to 7 kg/t for the remaining 6 days of the test period. This effect, which obviously upset the steady state of the circuit and the loading profile, is well illustrated in Fig. 3. The increase in gold loading was also accompanied by a decrease in the loading of copper (13 kg/t), zinc (13 kg/t), and calcium (1,6 kg/t).

TABLE V
CARBON LOADING DATA FROM PLANT B AT WDL (PHASE I)

Operating time	Carbon loading, kg/t					
d	Au	Ag	Cu	Ni	Zn	Ca
1,6	3,91	0,38	4,2	2,6	8,0	2,7
3,5	5,07	0,50	7,4	3,2	10,0	3,4
5,5	4,87	0,50	12,0	3,1	12,9	3,1
7,5	5,14	0,51	17,0	3,5	15,0	4,2
9,4	4,89	0,48	16,8	3,5	22,1	4,0
11,4	5,21	0,50	17,0	3,3	22,1	4,0
13,3	5,32	0,50	17,1	3,2	20,8	3,8
14,3	5,79	0,53	13,5	3,1	20,5	4,5
15,0	6,41	0,53	10,4	3,1	9,8	3,4
3% HCl wash						
15,9	7,04	0,57	11,8	3,1	8,6	1,9
17,0	7,15	0,58	13,6	3,4	11,8	1,7
18,0	7,22	0,51	13,7	3,2	14,9	1,6
18,9	7,09	0,55	14,0	2,9	16,5	1,9
19,9	7,35	0,58	11,5	2,7	14,9	1,9
20,9	7,71	0,59	12,1	2,8	9,4	0,8

Gold recoveries during this test period (an operating time of 21 days) exceeded 99,5 per cent, the gold barren values of 0,002 to 0,003 g/t being at steady state.

Phase II

After Phase I had been completed, a bulk solution with a gold concentration of 0,2 g/t and a sodium cyanide concentration of 100 g/t was used as the column influent. As before (Plant A), in order to approach steady state in a shorter period of time, this phase was started with the same carbon profile in the column as that at the completion of Phase I. Carbon equivalent to a bed depth of 0,1 m (200 cm^3) was slugged in and out of the column at regular intervals (only every 24 hours) to effect a gold loading on the carbon of approximately 5 kg/t. The column was washed with acid every 24 hours. For the first 9 days of operation, the acid wash entailed the pumping of 3 per cent hydrochloric acid (by volume) for 3 minutes, followed by a 5-minute soaking period prior to normal operation. This was subsequently changed to the pumping of the same strength of sulphuric acid for 3 minutes, followed by a soaking period of 5 minutes for a further 9 operating days. The adsorption data and daily carbon loadings are shown in Fig. 4 and Table VI respectively.

From the adsorption data obtained for the first 8,4 operating days, a gradual decrease in overall gold recovery can be noted as the system approached steady state. Although the adsorption curves (Fig. 4) are difficult to extrapolate, a gold recovery in the region of 95 per cent and yielding a barren solution with a gold concentration of approximately 0,01 g/t is suggested at steady state. A slow fall-off in gold loadings could be expected since the carbon was being moved at a slower rate in order to effect a loading of 5 kg/t (Table VI). There was an increase in the copper loading to about 20 kg/t, together with marginal increases in the nickel and zinc loadings.

While the effect of the stronger acid addition during the following 9 days in improving the gold recovery was immediately evident (Fig. 4), the longer-term effect of the

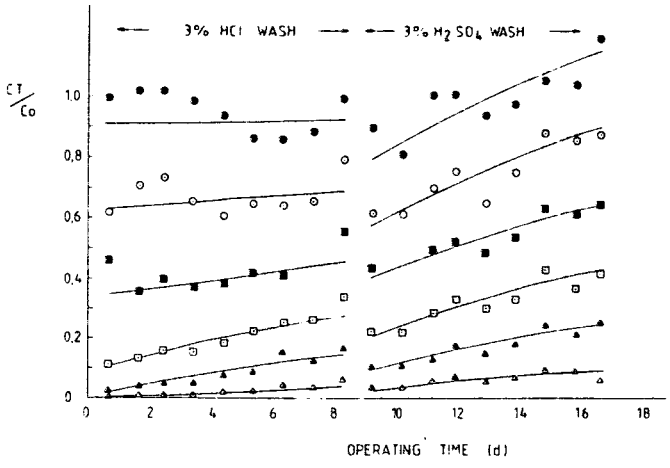

Fig. 4—Gold adsorption data from pilot Plant B (Phase II). The designations of the individual adsorption curves are similar to those in Fig. 3

TABLE VI
CARBON LOADING DATA FROM PLANT B AT WDL (PHASE II)

Operating time	Carbon loading, kg/t					
d	Au	Ag	Cu	Ni	Zn	Ca
Daily 3% HCl wash						
0,7	7,56	0,60	14,2	3,3	9,7	2,0
1,7	7,49	0,46	11,0	3,0	9,3	3,6
2,5	7,11	0,41	10,8	3,3	8,1	3,1
3,5	7,06	0,51	12,0	3,6	8,4	3,5
4,4	7,19	0,50	12,9	3,8	9,8	3,8
5,4	6,28	0,38	15,8	4,2	13,6	4,5
6,4	6,26	0,43	19,2	4,8	13,1	4,5
7,4	6,48	0,40	20,3	4,7	12,6	4,0
8,4	6,92	0,69	13,7	4,3	11,6	3,3
Change to daily 3% H_2SO_4 wash						
9,27	6,42	0,70	22,9	5,1	15,5	4,6
10,25	5,86	0,63	21,8	4,9	7,8	4,8
11,23	5,94	0,64	23,6	4,9	7,3	4,6
11,96	6,09	0,65	24,9	5,1	6,3	4,4
12,94	6,01	0,65	22,5	4,5	6,6	4,2
13,92	5,83	0,66	27,0	5,3	8,1	13,8
14,90	5,71	0,66	26,4	5,2	6,4	11,5
15,88	5,57	0,65	21,3	5,1	5,4	4,9
16,72	5,29	0,62	18,0	4,8	3,2	4,9
16,85	5,70	0,65	17,9	4,9	2,1	8,0
17,41	6,05	0,69	21,1	4,8	1,1	36,8
17,93	5,07	0,57	15,2	4,9	1,5	9,5
18,43	5,05	0,61	12,9	4,4	1,4	14,1

sulphuric acid treatment indicated some degree of carbon passivation. This was accompanied by significant increases in the calcium loadings, together with decreases in the zinc and copper loadings (Table VI). The very significant increase in the silver loading, possibly indicating a high recovery, should be noted. This is almost certainly due to the insoluble character of silver sulphate.

Discussion

In general, the results obtained with the moving carbon bed indicate several important aspects of the recovery of gold from plant solutions. First and foremost is the exceptional metallurgical efficiency that was obtained with relatively high flowrates of influent: the barren values were low, and the gold loadings were high. (The solution contact time in the 6 m column was only 2,4 minutes on the assumption of a carbon voidage of 40 per cent, while that in the 8 m column was 8 minutes on the assumption of a voidage of 50 per cent.) These features are significant in reducing both the size and the capital and operating costs in any proposed carbon circuit. The approach of the system to a true countercurrent operation with a steep loading profile through the circuit as the result of a strong chemical driving force is undoubtedly responsible for the efficiency observed. The hold-up of bullion, which may constitute a considerable indirect cost in a circuit where carbon mixing takes place, or where long carbon residence times are required, can also be expected to be reduced in a moving-bed adsorption circuit.

The critical nature of the carbon residence time in a gold-adsorption circuit is particularly interesting. Owing to the heterogeneous nature of the gold adsorption from plant solutions (and pulps) and the considerable carbon fouling that occurs progressively through the circuit, it can be expected that carbon residence times are critical. This was admirably demonstrated in Plant A, where the depth of the carbon bed was reduced from 12 to 8 m, with a resultant improvement in metallurgical efficiency. This aspect is often overlooked in plant designs that use one or two stages more than are required. In many instances, these extra stages may well be counterproductive. This investigation also indicated that adsorbed silica may be partly responsible for the carbon fouling that was observed. Such fouling by silica is particularly difficult to monitor in a pulp circuit for obvious reasons, and can be expected to increase significantly at high pH levels.

As very rapid kinetics are generally common to such adsorption circuits, it can be expected that the particle size of the carbon plays a most important role. However, because of such high adsorption rates, relatively high flowrates of influent are possible, thus making the selection of particle size a compromise between the chemical and the physical restraints of the system. This may be particularly important in the processing of unclarified plant solutions, or of effluents in which the clarity of the influent and the rate and regularity at which the carbon is moved play an important role. A comparison of the data from Plants A and B, which treated plant solutions of different tenors and used carbons at different particle sizes and flowrates, shows that good metallurgical efficiency was achieved in both instances. Plant A used a coarse (3 mm) extruded carbon and a slow flow velocity (0,5 m/min), which an be expected to successfully treat an unclarified solution of high tenor such as gold-plant primary filtrate. On the other hand, where further clarification is expected, the use of a finer carbon and high flow velocities (1 m/min) as in Plant B are suggested.

In most carbon-in-pulp operations in which the loading of gold is not allowed to approach equilibrium, film diffusion predominates in limiting the rate of gold extraction. Under such conditions, physical aspects such as mix-

ing, pulp density, carbon particle size, and carbon concentration play a most significant role[3], and are likely to overshadow any chemical considerations. Conversely, in an adsorption circuit such as the moving bed, where the loading of gold may well approach equilibrium, the rate of extraction decreases, and therefore any factor that influences the loading capacity has a bearing on plant performance under these conditions. In a plant situation, such variables as pH, cyanide concentration, and even temperature are likely to play an important role. The high gold loadings obtained in Plant B in the treatment of plant solutions of low tenor at pH 8, and the large effect of the acid wash, indicate to some degree the sensitivity of the system to such changes.

From a practical point of view, both the lowering of influent pH by the use of sulphuric acid and the intermittent acid washing of the column with dilute hydrochloric acid must be carefully considered. While both procedures undoubtedly improve the metallurgical efficiency, such improvements must be off-set in terms of the cost of the acid. This assessment is not easy since both procedures also have a considerable effect on the physical operation of the system owing to the build-up of calcium carbonate in the column. In Plant A, where no acid was added and where a coarse extruded carbon was used, the build-up of calcium carbonate was often visually excessive but appeared to have little effect on either the metallurgical efficiency or the physical operation of the plant. Conversely, earlier column operations using finer coconut-shell carbons[4] had indicated the occurrence of severe calcium passivation of the carbon, and even physical cementing of the carbon, in the columns during the treatment of conventional plant solutions. These observations suggest that the relatively large pore-size distribution of the extruded carbon is not nearly as susceptible to fouling by calcium as is the coconut-shell carbon with its fine pores. Of course, the higher pressure drop across a column containing the finer coconut carbon would also induce more calcium carbonate to precipitate.

Of adsorption systems for the treatment of plant solutions, the greater metallurgical efficiency and lower carbon dosage (bullion lock-up) offered by the moving-bed circuit are the most significant of its advantages. Other advantages offered by the moving bed include the following.

(a) It usually requires a lower capital investment.
(b) It requires less space, which is a key consideration when a system is to be added to an existing gold plant.
(c) It can be tightly controlled so that the quality of the effluent is close to the specification value, which is a most important consideration in the treatment of gold-plant solutions.
(d) The short time required to approach steady state in the moving-bed system makes the piloting and design relatively simple.

The greatest disadvantage of the moving-bed absorber is that it is not emptied in normal operation, and inspection and/or repairs can therefore be carried out only by the taking of an adsorber off-line and emptying the entire carbon content into temporary containers. The other disadvantage normally associated with moving-bed systems treating wastewaters is their inability to treat influents containing suspended solids, and some prefiltration is generally required. However, in gold adsorption, where the fast kinetics allow for high rates of carbon movement, the frequent slugging of carbon (and collected solids) from the circuit, together with the use of a relatively coarse carbon, may well circumvent this problem.

Acknowledgements

This paper is published by permission of the management of the Anglo American Corporation. The contributions of Messrs J.W. Chandler, M.P. Hallewell, G. Nicholson, A.W.G. Reeve, and B.S. Tatterson in the pilot-plant operations are gratefully acknowledged. The single 6 m pilot-plant operation (Plant B) was a collaborative project between Western Deep Levels Gold Mine and Anglo American Research Laboratories. The contribution and co-operation of the Vaal Reefs Exploration & Mining Company (South Division) for the Plant A operation is gratefully acknowledged, as is the funding of the project by the Gold Division of the Anglo American Corporation.

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3.4 EXPANDED BEDS TREATING DAM RETURN EFFLUENT

Pilot testwork at Ergo on the treatment of dam return effluent containing less than 1 g/m^3 free cyanide indicated that weekly acid washing of the carbon inventory (SA Patent No. 78/5147) negated the necessity to maintain a high loading profile in a single expanded bed type of operation. Although the gold lock-up in such an operation may be higher than in a staged circuit, this is not of major concern on treating very low grade effluents. This testwork resulted in the commissioning of 4 expanded bed columns at Ergo treating 1 000 m^3/h at a cost of R1,25 m. Gold recoveries in 1985 - 1986 in this circuit averaged 83% from an influent containing only 0,03 g/m^3 gold while producing loaded carbon containing 450 g/t gold. This may be equated with a concentration factor of x 15 000 which significantly reduced elution costs when compared with CIP where concentration factors of 2 000 - 3 000 are achieved.

Similar acid washing routines are carried out at Vaal Reefs using staged NIMCIX type contactors treating dam return effluent containing 0,1 g/m^3 gold.

4.0 GOLD ELUTION

Due to the lengthy elution times required in the Zadra elution (ca. 72h) used in the early Homestake Gold Mine operation (1973), initial investigations at AARL were directed towards the development of a more cost effective process using water as an eluant. Several patents and papers were published in this regard culminating in the development of AARL batch elution process which is now in common use. Typically the process involves an elution time of 3-4 hours at 110-120°C with many circuits carrying out three elutions per day.

Other aspects related to the elution of gold which are presented include the recent development of the continuous AARL elution process, the use of organic solvents and the separate elution of gold and silver values. The cost effective recovery of free cyanide from copper-rich eluates is also described.

4.1 AARL ELUTION

Numerous developments and modifications to the AARL process (SA Patent Nos. 73/89/39 and 76/4202) for the elution of gold from activated carbon have taken place since the earlier publications in 1974-79. Following successful piloting, first at the Vaal Reefs Mining and Exploration Company (1975) and then at the Fairview Gold Mine (1978), the process was first used on a large scale at the President Brand Gold Mine (1980) and is now in common use in South Africa and Australia with several circuits operating in North America.

A review of the AARL process for the elution of gold from activated carbon is also appended. This review was presented in part as Chapter 15 in the S.A.I.M.M. Carbon Schools held in 1985 and 1987, as well as a paper published at the 13th C.M.M.I. Conference in Singapore (1986). The review also forms part of a reference package supplied to licencees and users of the patent.

PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 73/8939 (Cognating 74/1182 and 74/2482)

Date of Filing: 23rd October, 1974

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventor: RAYMOND JOHN DAVIDSON

Title of Invention: "METAL, PARTICULARLY GOLD, RECOVERY"

~~Provisional~~/Complete Specification as Filed

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Patents Form 3

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251.

FILING DATE:

REPUBLIC OF SOUTH AFRICA

THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 73/8939 (Cognating 74/1182 and 74/2482)

TITLE OF INVENTION: "METAL, PARTICULARLY GOLD, RECOVERY"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street, JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

Brill Bros. F 8812

THIS invention relates to metal recovery, particularly gold recovery.

The cyanidation process is a well known process for extracting gold values from gold-bearing ores. The process involves finely grinding the ore and then leaching the ore with a sodium cyanide solution. The gold values are leached in the form of aurocyanide ($Au(CN)_2$) ions according to the following reaction:

$$2\,Au + 4\,NaCN + O_2 + H_2O = 2\,NaAu(CN)_2 + 2NaOH$$

The gold may be recovered from the leach solution using conventional techniques such as electrowinning. Once the bulk of the gold values has been recovered from the leach solution, the problem remains to recover the residual amounts of gold. In the past this has been achieved by, for example, adsorbing the gold values on to an activated charcoal support and then desorbing or solubilising the values from the support using one of the following methods:

(1) The activated charcoal is contacted with a boiling mixture of sodium cyanide and sodium hydroxide. The desorbed gold can be recovered from the resulting solution using conventional techniques such as electrowinning or zinc precipitation.

(2) The activated charcoal is contacted with anhydrous liquid ammonia. The ammonia can be flashed off the resulting solution to recover the gold.

(3) The activated charcoal is contacted with a solution of sodium sulphide and sodium hydroxide. The gold can be recovered from the resulting solution by electrowinning.

The above three methods have the disadvantage that they are expensive and employ unpleasant reagents.

According to the invention, a method of recovering a transition or noble metal values from a support having adsorbed thereon the metal values in the form of an ionic complex including the step of contacting the support with water of low metal cation concentration to desorb metal values.

The water preferably has a low multi-charged cation, e.g. alkaline earth metal cation, concentration. In particular, waters having a metal cation concentration of less than 100 ppm, preferably less than 50 ppm, are preferred. Suitable waters are distilled, deionised and softened waters.

The invention has particular application to metal values in the form of cyanide ionic complexes. In such complexes the metal values will form part of the anionic portion of the complex. In the case of gold, the anionic portion will be the aurocyanide anionic complex, $Au(CN)_2^-$.

It has been found that desorption of the metal values by the low cation water is more effective at elevated temperatures. If higher than atmospheric pressures are used, temperatures as high as 150°C can be used. Furthermore, when the support forms part of a desorption column it is preferable that the length to diameter ratio is high and slow rates of elution are employed.

Any support suitable for adsorbing the metal values in their ionic complex form may be used. A particularly suitable support is charcoal, particularly activated charcoal.

It has been found that when the cation of the complex is an alkaline earth metal, the complex is adsorbed on to the support very much more strongly than when the cation is an alkali metal. This is particularly so in the case of aurocyanide complexes. In fact the degree of adsorption has been found to follow the series Ca>Mg>H>Li>Na>K. Consequently, if the cation of the complex is an alkaline earth metal, particularly calcium, it is preferable to subject the support to a pre-treatment with an alkali metal salt solution prior to water desorption. Such pre-treatment will effectively result in an exchange reaction between the alkali metal and the alkaline earth metal. Such exchange reaction will be enhanced by providing a salt whose anion forms an insoluble or substantially insoluble salt with the alkaline earth metal so that the alkaline earth metal is effectively removed from the system as exchange takes place. Suitable salts are carbonates, oxalates, sulfites and fluorides.

It has further been found that the pre-treatment is rendered more efficient if the solution is basic.

The alkali metal is preferably sodium, potassium or lithium. A preferred pre-treatment solution is a potassium carbonate/potassium hydroxide solution having a basic pH.

The invention finds particular application in the recovery of gold in the cyanidation process. In one form, the cyanidation process involves slurrying the gold bearing ore with a sodium cyanide/calcium hydroxide leach solution and granulated charcoal. The gold is leached in the form of aurocyanide ions and in this form is adsorbed on to the charcoal. The charcoal is screened off and the gold recovered from the charcoal using one of the methods mentioned above. Instead of these prior art methods, the

gold may be recovered by elution with low metal cation water according to the invention. Alternatively, the method of the invention may be used to recover residual gold values from the effluents of prior art gold recovery methods.

In an example of the invention, a gold plant effluent was adsorbed on to a commercially available coconut shell granulated charcoal (0,5 - 2mm) support. Adsorption was effected by mixing the effluent with the charcoal followed by drying. The analysis of the loaded charcoal was as given below:

ELEMENT	Ppm ON LOADED CHARCOAL
Au	3300
Ag	45
Fe	1500
Ni	2500
Cu	240
Co	2
Zn	380

The loaded charcoal was placed in an elution column of 0,9 cm.internal diameter and 40 cm length. The loaded charcoal was subjected to a pre-treatment by a 5 weight percent potassium carbonate solution at 50°C, followed by elution with deionised water at 90°C. The results are given in the following Table I :

TABLE 1

Elution volume (ml)	Au (ppm)	Ag (ppm)	Fe (ppm)	Ni (ppm)	Cu (ppm)	Co (ppm)	Zn (ppm)
19	1,3	<1	5	9	2	<1	3
39	1,5	↓	3	11	1	↓	5
60	1,6		2	14	2		5
80	1,8		2	17	3		5
100	2,0		2	18	4		5
121	28	-	2	-	-	-	-
141	96	12	<1	860	35	<1	5
160	144	16	↓	200	12	↓	2
179	185	13		60	6		2
198	213	9		20	5		1
217	235	6		11	3		<1
236	228	3		7	2		↓
255	209	2		4	2		
275	192	1		3	2		
293	170	1		2	1		
312	145	1		2	1		
332	121	1		1	1		
349	92	<1		1	1		
370	82	↓		1	1		
389	65			1	1		
440	52			1	1		
423	39			1	<1		
442	29			1	↓		
461	20			1			
480	13,3			1			
498	9,1			1			
517	7,0			<1			
536	5,8			↓			
555	5,1						
575	4,8	↓	↓		↓	↓	↓

Note: 1. Fractions 0 - 100 ml pretreatment
2. Fractions 100 - 575 ml - deionised water elution
3. 25 ml = 1 bed volume

Using materials and methods described above, the effect of various pretreatments was investigated. Pretreatments were carried out with a 10 weight percent potassium carbonate solution, a 1,5 weight percent lithium carbonate solution, a ten weight percent sodium carbonate solution, a 10 weight percent ammonium carbonate solution and a 10 weight percent sodium chloride solution. Subsequent elution with deionised

water resulted in the following recoveries:

Potassium carbonate	84 percent
Lithium carbonate	68 percent
Sodium carbonate	60 percent
Ammonium carbonate	18 percent
Sodium chloride	10 percent

From the above it will be noted that potassium carbonate pretreatment proved the most effective. Similar experiments were carried out using various potassium carbonate pretreatments and the results of these experiments can be found in the following table II:

TABLE II

THE OPTIMISATION OF THE CARBONATE PRETREATMENT OF ACTIVATED CHARCOAL CONTAINING 0,3 PER CENT Au LOADED USING GOLD PLANT EFFLUENT

Pretreatment	Flow velocity (cm/min)	No. of bed volumes (carbonate pretreatment)	Temp. of pretreatment (°C)	Temp. of elution (°C)	Gold recovery
1% K_2CO_3	2,37	4	50	90	70% in 24 bed vol
5% "	0,74	8	50	90	91% in 19 " "
10% "	2,37	8	50	90	48% in 20 " "
0,5% K_2CO_3 + 0,5% KOH	0,74	ca. ½	90	90	35% in 24 " "
1% K_2CO_3 + 1% KOH	0,74	ca. ½	90	90	73% in 16 " " 82% in 24 " "
3% K_2CO_3 + 1% KOH	0,74	ca. ½	90	90	88% in 16 " " 93% in 24 " "
5% K_2CO_3 + 3% KOH	0,74	ca. ½	90	90	96% in 16 " " 98% in 25 " "
5% K_2CO_3 + 3% KOH	0,74	1	90	90	98% in 16 " " 99% in 25 " "
10% K_2CO_3 + 5% KOH	0,74	1	90	90	94% in 7 " " 98% in 12 " " 99% in 24 " "
5% K_2CO_3 + 10% KOH	0,74	1	90	90	94% in 7 " " 99% in 12 " " 100% in 22 " "

NOTE: All percentages are by weight.

In order to demonstrate the effect of increased temperature on desorption efficiency, 20 bed volumes of deionised water at 90°C and at 70°C was used to desorb gold values adsorbed on to an activated charcoal support in the manner described above. In each case a pretreatment with a half bed volume solution of 5% potassium carbonate/3% potassium hydroxide solution was carried out. It was found that elution with the 90°C water resulted in 98% gold recovery as against a recovery of only 73% using the 70°C water.

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

1.

A method of recovering transition or noble metal values from a support having adsorbed thereon the metal values in the form of an ionic complex including the step of contacting the support with water of low metal cation concentration to desorb metal values.

2.

A method according to claim 1 wherein the water has a low concentration of multi-charged cations.

3.

A method according to claim 1 or claim 2 wherein the water has a metal cation concentration of less then 100 ppm.

4.

A method according to claim 1 or claim 2 wherein the water has a metal cation concentration of less than 50 ppm.

5.

A method according to any one of the preceding claims wherein the water is deionised water or softened water.

6.

A method according to any one of the preceding claims wherein the complex is a cyanide ionic complex, the metal values forming part of the anionic portion thereof.

7.
A method according to any one of the preceding claims wherein the metal values are selected from gold and silver.

8.
A method according to any one of the preceding claims wherein the support is activated charcoal.

9.
A method according to any one of the preceding claims wherein the desorbing is effected at elevated temperatures.

10.
A method according to any one of the preceding claims wherein the cation of the complex is sodium, potassium or lithium.

11.
A method according to any one of the preceding claims wherein the cation of the complex is an alkaline earth metal cation and the support is contacted with an alkali metal salt solution prior to the desorption step.

12.
A method according to claim 11 wherein the alkali metal salt solution is a sodium, potassium or lithium salt solution.

13.
A method according to claim 11 or claim 12 wherein the anion of the alkali metal salt solution is one which forms an insoluble or substantially insoluble salt with the alkaline earth metal cation.

14.
A method according to any one of claims 11 to 13 wherein the alkali metal salt solution is a sodium, potassium or lithium carbonate solution.

15.
A method according to any one of the preceding claims wherein the alkali metal salt solution has a basic pH.

16.
A method according to claim 11 wherein the alkali metal salt solution is a potassium carbonate/potassium hydroxide solution having a basic pH.

17.
A method of recovering gold or silver values from a support having adsorbed thereon the metal values in the form of an alkaline earth metal cyanide ionic complex including the steps of contacting the support with alkali metal salt solution, the anion of the salt being capable of forming an insoluble or substantially insoluble salt with the alkaline earth metal of the complex, and desorbing the metal values from the support with deionised or softened water.

18.
A method according to claim 17 wherein the alkali metal salt solution is potassium, sodium or lithium carbonate solution having a basic pH.

19.
A method according to claim 17 or claim 18 wherein the support is activated charcoal.

20.
A method according to claim 1 or claim 17 substantially as herein described.

21.
Metal values recovered by a method according to any one of the preceding claims.

DATED THIS 23rd DAY OF OCTOBER, 1974.

SPOOR AND FISHER,
APPLICANT'S PATENT ATTORNEYS.

PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 76/4204

Date of Filing: 28th June, 1977

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventors: 1. RAYMOND JOHN DAVIDSON
2. DAVID DUNCANSON

Title of Invention: "RECOVERY OF METAL VALUES"

~~Provisional~~/*Complete Specification as Filed*

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HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 76/4204

TITLE OF INVENTION: "RECOVERY OF METAL VALUES"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

THIS invention relates to metal recovery, in particular to the recovery of gold, silver, nickel or copper metal values.

A method of recovering the above metals from supports on which they are adsorbed as complexes is covered by South African Patent No. 73/8939. According to this method, the metal values are desorbed by contacting the support with water of low metal cation concentration. The water preferably has a low-multi-charged cation (e.g. alkaline earth metal cation) concentration. Concentrations less than 300 ppm, preferably 50 ppm, are preferred. In other words the water is relatively pure and has a low ionic strength. Suitable waters are for example distilled, deionised and softened waters.

The above method has particular application to metal values in the form of cyanide ionic complexes, the metal value forming the anionic portion of the complex. Any suitable adsorbent support may be used, carbon, particularly activated carbon, being preferred.

It is also disclosed that when the cation of the complex is an alkaline earth metal, the complex is adsorbed on to

the support very much more strongly than when the cation is an alkali metal, particularly when the complex is an aurocyanide one. Accordingly, if the cation of the complex is an alkaline earth metal, particularly calcium, the support is preferably pre-treated with an alkali metal salt solution prior to water desorption.

This pretreatment results in an exchange reaction between the alkali metal and the alkaline earth metal. The exchange reaction is enhanced by providing a salt the anion of which forms an insoluble or substantially insoluble salt with the alkaline earth metal so that the latter is effectively removed from the system as exchange takes place. Suitable salts for this purpose are carbonates, oxalates, sulphites, and fluorides, the alkali metal preferably being sodium, potassium, or lithium.

A preferred pretreatment solution according to the above patent is a potassium carbonate/potassium hydroxide solution having a basic pH.

We have now discovered that improved results are obtained with other pretreatment solutions.

According to the invention there is provided a method of

recovering metal values selected from gold, silver, copper and nickel from a support having one or more of these metal values adsorbed thereon in the form of an alkaline earth metal ionic complex, the metal value forming part of the anionic portion thereof including the steps of contacting the support with a solution selected from the group of an alkali metal cyanide solution, an alkali metal hydroxide solution and a mixture thereof, followed by desorbing the metal values from the support with water having a low concentration of metal cations.

The solution may be one containing alkali metal cyanide of concentration in the range 1 to 10 percent by weight and alkali metal hydroxide of concentration in the range 1 to 20 percent by weight. It is preferred that the concentration of cyanide ions exceeds the concentration of the hydroxyl ions in solution. The preferred mixture is sodium cyanide and sodium hydroxide.

One particularly suitable pretreatment solution contains an alkali metal cyanide concentration of about 10 percent by weight and an alkali metal hydroxide concentration of about 1 percent by weight.

Another suitable pretreatment solution is a sodium hydroxide

solution. It has been found that concentrations not exceeding 10 percent by weight provide satisfactory results.

Initially the metal values are adsorbed on to the support in the form of an alkaline earth metal ionic complex. The preferred support is activated carbon. The desorbing, after treatment of the support with the selected solution, is carried out using the conditions and waters as described above and in the above mentioned patent. It is important to note that the process of desorbing involves only physical adsorption/desorption action and does not involve ion exchange.

The alkaline earth metal will generally be calcium and the complex will generally be a cyanide ionic complex. The invention has particular application to the recovery of gold and other metal values from solutions containing these values obtained in the cyanidation of gold bearing ores with, for example, a sodium cyanide/calcium hydroxide leach solution as described in the above mentioned patent.

The invention is illustrated in the following Examples:

Example I :

Type G210 granular activated carbon (Le Carbone (Pty) Ltd)

was laden using gold plant cyanidation effluent. The laden charcoal analysed as follows:

Gold	2,9 kg/t
Silver	64 g/t
Nickel	2,2 kg/t
Copper	90 g/t

These metals were present on the charcoal in the form of calcium cyanide ionic complexes, e.g. $Ca(Au(CN)_2)_2$.

The original capacity constant of the virgin charcoal was 22 mg gold per gram of charcoal.

Various runs were conducted using different pretreatment solutions and operating conditions. Except where otherwise indicated, in each case 16 g of charcoal were laden using a vibrator into a small glass elution column, to provide a charcoal bed of 1 cm internal diameter and 25 cm in length.

The bed was pretreated by pumping a half-bedvolume of pretreatment solution at a rate of approximately 1 bedvolume per hour (apparent flow velocity $9,2 \times 10^{-3}$ cm/sec) through the bed, after first draining the column of any surplus water.

The pretreated bed was then eluted using deionised water at 90°C, at elution rates in the range 0,5 to 2,0 bed-volumes per hour.

10 bedvolumes were then collected and analysed.
Finally the charcoal was removed from the column, ovendried, and analysed and activity tested.

In all the experiments conducted, the pretreatment solution comprised a mixture of sodium cyanide and sodium hydroxide.

(a) Runs 1 to 11

In a first series of experiments, the effect of variations in the relative proportions of these two components were studied. The results of these experiments are set out in Tables I and II hereunder.

TABLE I

EFFECT OF CHANGE IN SODIUM HYDROXIDE CONCENTRATION USED AS A PRETREATMENT REAGENT

No.	Pretreatment reagent/conditions	Elution peak concentration (g/t)			Charcoal residues (g/t)				Residual charcoal activity (mg Au/g charcoal)
		Au	Ag	Ni	Au	Ag	Cu	Ni	
1	10% NaCN at 90°C	940	40	1600	16	<7	4	64	13
2	" /1% NaOH "	862	37	1920	15	<7	7	100	12
3	" /5% NaOH "	825	25	1430	23	3	8	57	12
4	" /14% NaOH "	820	25	1350	18	5	1	70	13
5	" /20% NaOH "	770	23	1260	19	11	-	40	12

TABLE II

EFFECT OF CHANGE IN SODIUM CYANIDE CONCENTRATION USED AS A PRETREATMENT REAGENT

No.	Pretreatment reagent/conditions	Elution peak concentration (g/t)			Charcoal residues (g/t)				Residual charcoal activity (mg Au/g charcoal)
		Au	Ag	Ni	Au	Ag	Cu	Ni	
6	1% NaOH at 90°C	550	11	-	42	22	17	300	13
7	" /1% NaCN "	630	21	1250	17	5	-	90	13
8	" /3% NaCN "	705	28	1560	26	33	18	88	13,5
9	" /5% NaCN "	670	25	1360	18	11	14	64	13
10	" /8% NaCN "	740	24	1560	18	9	6	30	13
11	" /10% NaCN "	820	25	1350	18	5	1	70	13

It is seen that as the sodium hydroxide concentration was decreased, and the sodium cyanide concentration increased, the recovery of metal values improved, the best results being obtained using a pretreatment solution comprising 10 percent by weight sodium cyanide and 1 percent by weight sodium hydroxide.

(b) Runs 12 to 14

In a second series of experiments (Runs 12 to 14) the efficiency of the process in regard to high gold loadings was studied. The results are set out in Table III.

G 215 activated charcoal was laden to 4-6 percent gold using

clarified gold plant pregnant solutions. The metal values in the solutions were in the form of cyanide ionic complexes. The laden charcoal was treated with various sodium cyanide/ sodium hydroxide reagents, and eluted at 90°C with deionised water.

In Run 12 the analysis of the laden charcoal was:

Gold	4 percent
Silver	600 g/t
Nickel	2 600 g/t

The pretreatment solution comprised 10 percent sodium cyanide/14 percent sodium hydroxide

Recoveries of 98,6% gold, 83% silver, and over 99% nickel were obtained.

After 6 bedvolumes of deionised water (12 hours) the gold eluate contained only 6 g/t of gold.

In Run 13 the analysis of the laden charcoal was:

Gold	6 percent
Silver	2 000 g/t
Nickel	6 000 g/t

The pretreatment solution comprised 12 percent sodium cyanide/1 percent sodium hydroxide.

Recoveries of 99,9 percent gold, 97,1 percent silver, and

99,9 percent nickel were obtained. The lower hydroxide concentration is seen to provide better results.

Similar results were obtained in Run 14, wherein the charcoal analysis was 4 percent gold, 1 000 g/t silver, and 1 700 g/t nickel.

TABLE III

ELUTION DATA AT HIGHER GOLD LOADINGS

No.	Pretreatment conditions	Original loading (g/t)			Elution peak (g/t)			Charcoal residue (g/t)			% Recovery			Residual charcoal activity (mg Au/g charcoal)
		Au	Ag	Ni	Au	Ag	Ni	Au	Ag	Ni	Au	Ag	Ni	
12	½ Bedvolume 14% NaOH/10% NaCN at ½ Bedvolume/h elution Temperature 90°C	4,0	0,06	0,26	3500	140	970	550	125	9	98,8	83	99,6	21
13	½ Bedvolume 12% NaCN/15 NaOH at ½ Bedvolume/h elution Temperature 90°C	6	0,2	0,6	14000	1000	3500	77	36	8	99,9	97,1	99,7	16
14	As above	4	0,1	0,17	8800	470	770	78	93	12	99,8	91,0	99,3	17

(c) Runs 15 to 17

In a third series of experiments the process of the invention was compared with two other processes, namely potassium carbonate pretreatment followed by elution with deionised water at 90°C, and elution at 65°C with sodium sulphide/ sodium hydroxide solution.

The results are set out in Table IV hereunder and graphically illustrated in Figures 1 (Gold), 2 (Silver) and 3(Nickel).

TABLE IV

COMPARISON OF DIFFERENT METHODS USED FOR ELUTION OF GOLD FROM LOADED CHARCOAL

No.	Pretreatment/elution reagent/conditions	Elution peak concentration (g/t)			Charcoal residue (g/t)				Residual charcoal activity (mg Au/g charcoal)
		Au	Ag	Ni	Au	Ag	Cu	Ni	
15	1 Bedvolume 10% K_2CO_3/5% KOH at ½ Bedvolume/h followed by deionised water at 90°C	478	12,6	-	27	7	10	340	13,5
16	10 Bedvolumes of 3% Na_2S/3% NaOH/1% Na_2SO_3 elution at 65°C at 2 Bedvolumes/h followed by 10 Bedvolumes H_2O Followed by 4 Bedvolumes of 5% HNO_3 Followed by 10 Bedvolumes of H_2O	1050	0,2	300	81	40	50	85	10
17	½ Bedvolume 10% NaCN/1% NaOH Followed by deionised water at ½ Bedvolume/h at 90°C	862	37	1930	15	<2	7	100	12

h = hour

The process employing the sodium cyanide/sodium hydroxide pretreatment is seen to be superior to both the other processes.

In the case of gold recovery, the Na_2S/NaOH provided marginally better recovery than NaCN/NaOH, but the unpleasantness of handling Na_2S/NaOH solutions and the costs thereof make it unacceptable as a pretreatment reagent.

There is no precipitation of by-products on the charcoal; close to 100 percent recoveries are possible. Costs of reagents are relatively low since the regenerated sodium

cyanide and sodium hydroxide solution may be reused in other parts of the gold circuit; elution is flexible in that it may be effected at a rate between one-half and one bedvolume per hour in 5 to 7 bedvolumes of eluate at 90°C, or 4 to 5 bedvolumes at 125°C (3 hours elution time); and the higher the gold loading on the charcoal, the more efficient becomes the elution.

EXAMPLE II

In this example the effect of using sodium hydroxide pre-treatment solutions was investigated. The procedures and materials used were similar to those used in Example I. The metal rich solution, as in Example I, was a cyanidation effluent from a gold recovery process.

Various sodium hydroxide solutions were investigated and the results of these investigations can be found in Table V. From these results it can be seen that there is no appreciable advantage to be gained by using sodium hydroxide solutions of concentration greater than 10 percent by weight. It should also be noted that in all of runs 18 to 21, the loaded charcoal was acid washed with hydrochloric acid to remove calcium carbonate from the carbon prior to treatment with sodium hydroxide.

TABLE V

No.	Pretreatment conditions	Elution peak concentration (g/t)				Carbon Residue (g/t)					Recovery (%)					
		Au	Ag	Ni	Cu	Au	Ag	Ni	Cu	Ca	Au	(Au)*	Ag	Ni	Cu	Ca
18	½ bedvolume 10% NaOH. Eluted for 12 hours	247	9	140	146	68	42	240	488	462	96,4	(89,0)	58	91,3	49,2	97,
19	½ bedvolume 20% NaOH. Eluted for 12 hours	301	10	17	192	86	18	238	443	490	95,4	(91,3)	82	91,2	53,4	97,
20	½ bedvolume 30% NaOH. Eluted for 12 hours	327	5	24	197	90	36	316	468	530	95,2	(91,5)	64	88,6	51,2	97,
21	½ bedvolume 10% NaOH. In contact for 16 h. Eluted for 8 hours	310	5	69	210	132	50	232	430	442	93	(93)	50	91,6	55,2	98,

( ) * Calculated Au recovery after 8 bedvolumes

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

1.

A method of recovering metal values selected from gold, silver, copper and nickel from a support having one or more of these values adsorbed thereon in the form of an alkaline earth metal ionic complex, the metal value forming part of the anionic portion thereof, including the steps of contacting the support with a solution selected from the group of an alkali metal cyanide solution, an alkali metal hydroxide solution and a mixture thereof, followed by desorbing the metal values from the support with water having a low concentration of metal cations.

2.

A method according to claim 1 wherein the solution contains alkali metal cyanide of concentration in the range 1 to 10 percent by weight and alkali metal hydroxide of concentration in the range 1 to 20 percent by weight.

3.

A method according to claim 2 wherein the concentration of cyanide ions exceeds the concentration of hydroxyl ions.

4.

A method according to claim 2 or claim 3 wherein the solution contains sodium cyanide and sodium hydroxide.

5.

A method according to claim 1 wherein the solution is a sodium hydroxide solution of concentration not exceeding 10 percent by weight.

6.

A method according to any one of the preceding claims wherein the water has a metal cation concentration of less than 300 ppm.

7.

A method according to any one of the preceding claims wherein the water is deionised water or softened water.

8.

A method according to any one of the preceding claims wherein the support is activated carbon.

9.

A method according to any one of the preceding claims wherein the complex is a cyanide ionic complex.

10.

A method according to claim 1 substantially as herein described with reference to the Examples.

11.

Metal values recovered by a method according to any one of the preceding claims.

DATED THIS 28th DAY OF JUNE, 1977.

SPOOR AND FISHER

APPLICANT'S PATENT ATTORNEYS

Fig - 1

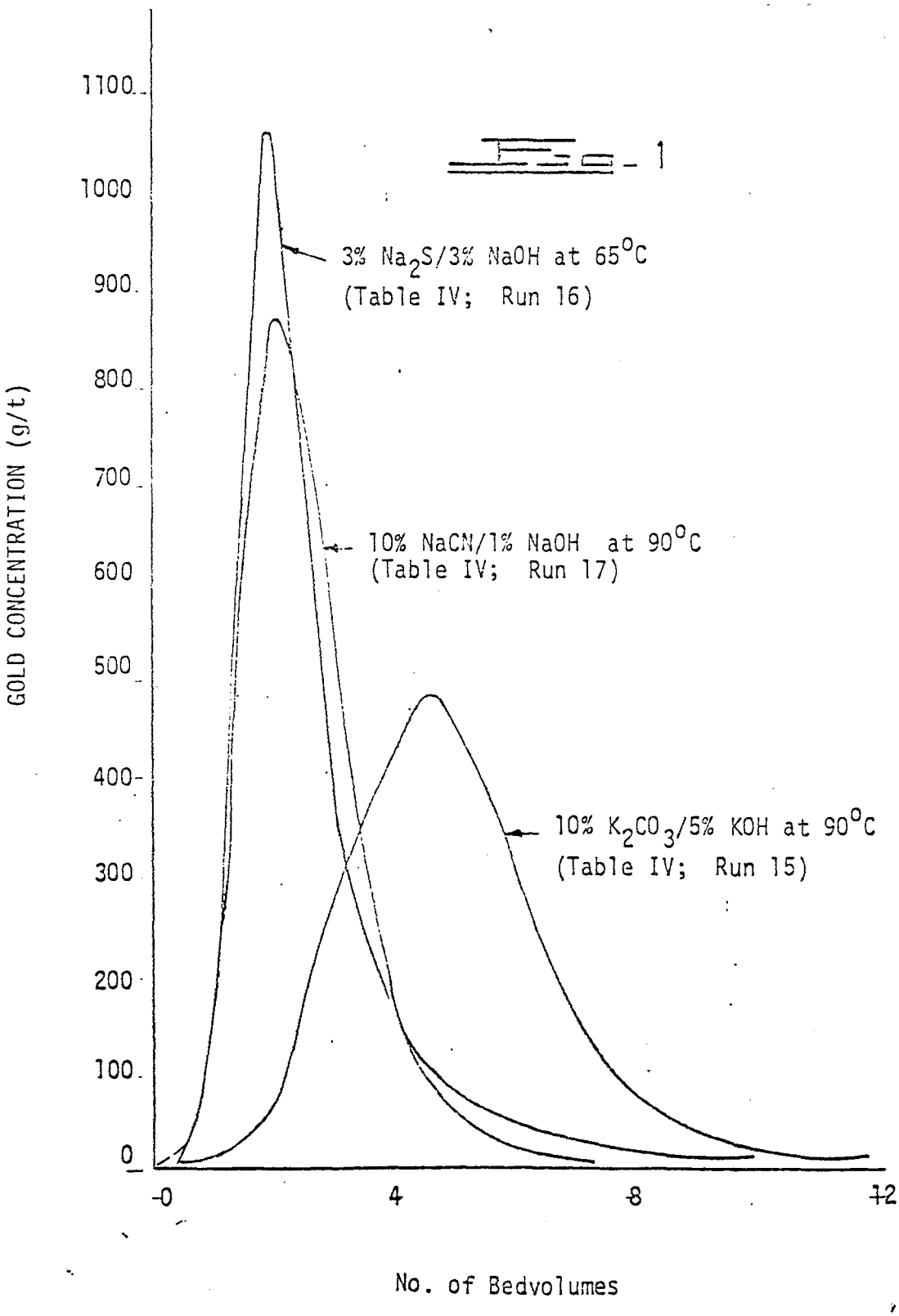

1100
1000
900
800
700
600
500
400
300
200
100
0
GOLD CONCENTRATION (g/t)
3% Na_2S/3% NaOH at 65°C
(Table IV; Run 16)
10% NaCN/1% NaOH at 90°C
(Table IV; Run 17)
10% K_2CO_3/5% KOH at 90°C
(Table IV; Run 15)
0
4
8
12
No. of Bedvolumes

Fig - 2

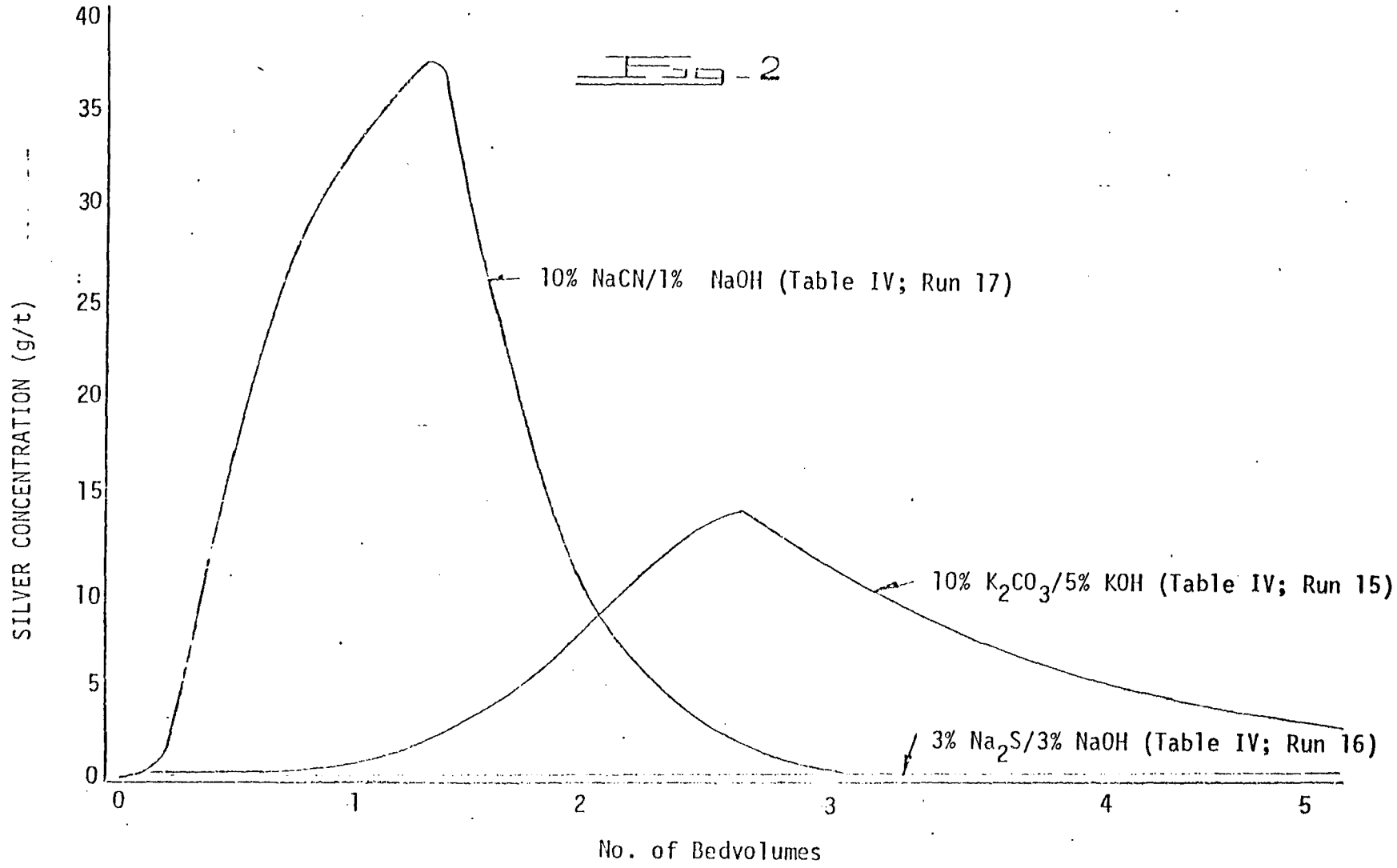
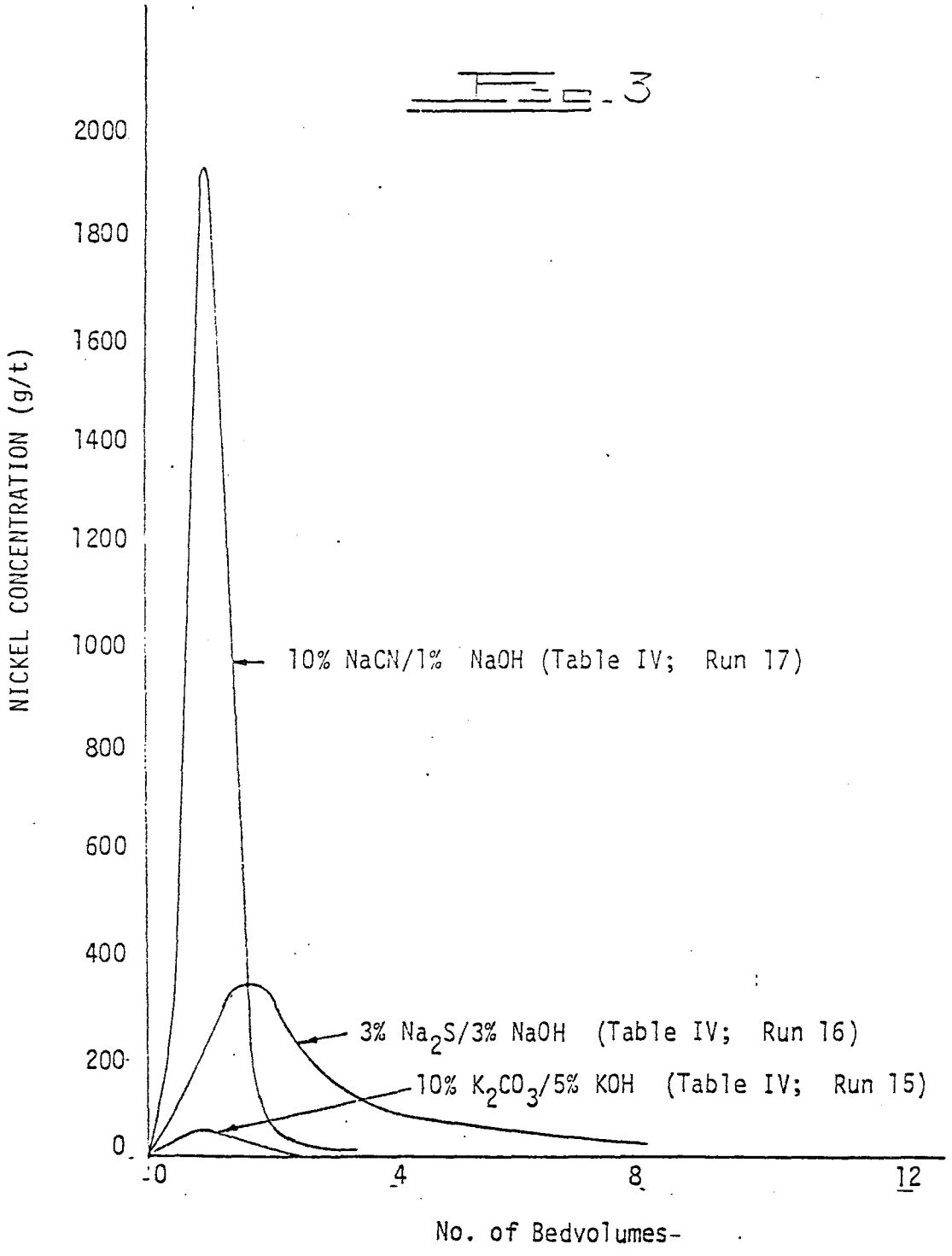

SILVER CONCENTRATION (g/t)
40
35
30
25
20
15
10
5
0
0
1
2
3
4
5
No. of Bedvolumes
10% NaCN/1% NaOH (Table IV; Run 17)
10% K_2CO_3/5% KOH (Table IV; Run 15)
3% Na_2S/3% NaOH (Table IV; Run 16)

FIG. 3



SPROGER & FISCHER

The elution of gold from activated carbon using deionized water

by R. J. DAVIDSON*, M.Sc., Ph.D. (Rhodes) (Visitor) and
D. DUNCANSON*, H.N.C., B.Sc. Hons. (Strathclyde) (Visitor)

SYNOPSIS

The pretreatment of gold-loaded activated carbon with a relatively strong sodium cyanide solution prior to elution with deionized water at 90 to 125°C has been shown to have a most beneficial effect on the elution of gold. The findings emphasize the important role played by the calcium cation and the hydroxyl anion in effecting efficient elution of gold. The elution of the relatively unstable complexes of silver and base metals with cyanide was also significantly affected by the ratio of the sodium cyanide to sodium hydroxide in the reagents used in the pretreatment stage. It was demonstrated that the use of 0,5 bed volume of a pretreatment reagent comprising 10 per cent sodium cyanide stabilized with 1 per cent sodium hydroxide for safety reasons resulted in a virtually 100 per cent recovery of adsorbed gold, silver, and base metals on subsequent elution with deionized water. The elution procedure takes a relatively short time, about 14 hours at 90°C (7 bed volumes) or 3 hours if pressure elution at 125°C (4 to 5 bed volumes) is used. Because the mechanism of elution after pretreatment entails the shifting of a physical rather than a chemical equilibrium, the higher the gold loading on the activated carbon, the more efficient is the elution procedure. When compared with other elution procedures, the proposed method appears to have several economic advantages.

SAMEVATTING

Daar is aangetoon dat die voorafbehandeling van goudbelaaide geaktiveerde koolstof met 'n betreklike sterk natriumsianied oplossing voor stroping met gedeioniseerde water teen 90-125°C 'n besonder voordelige uitwerking het op die herwinning van goud. Die bevindings onderstreep die belangrike rol wat deur die kalsium katioon en die hidroksiel anioon gespeel word by die verkryging en daarstelling van doeltreffende goudstroping, terwyl die herwinning van die betreklik onstabiele silwer- en basismetaal sianiedkomplekse aansienlik aangetas is deur die verhouding van natriumsianied/natriumhidroksied reagense wat gedurende die voorafbehandelingsstadium gebruik is. Dit is aangetoon dat die aanwending van 'n voorafbehandelingsreagens saamgestel uit 'n 0,5 bedvolume 10 persent natriumsianied gestabiliseer met 1 persent natriumhidroksied om veiligheidsredes, ten gevolge gehad het dat feitlik 100 persent van die opgeloste goud, silwer, en basis-metaalwaardes teruggevind is. Die stropingsprosedure kan in 'n betreklike kort tyd verrig word t.w. ca 14 uur teen 90°C (7 bedvolumes) of in drie uur as drukherwinning teen 135°C (4 tot 5 bedvolumes) aangewend word. Omdat die meganisme van stroping na voorafbehandeling eerder 'n fisiese as 'n chemiese ewewigsverskuiwing behels, sal bevind word dat hoe hoër die goudbelading op die geaktiveerde koolstof, hoe doeltreffender is die herwinningsprosedure. Wanneer die voorgestelde metode met ander stropingsmetodes vergelyk word, blyk dit dat die voorgestelde metode 'n hele aantal ekonomiese voordele inhou.

Introduction

It has been amply demonstrated that, even on a commercial scale, activated carbon is a most efficient and selective scavenger for gold from gold-plant leach liquors, pregnant slurries, and effluents, the major problem to date being the recovery of the gold from the loaded carbon to allow re-use of the latter. Methods involving expensive and hazardous reagents such as sodium sulphide–sodium sulphite–sodium hydroxide[1], boiling sodium cyanide–sodium hydroxide[2], pressure leaching at 150°C with sodium cyanide–sodium hydroxide[3], and anhydrous ammonia[4], have all been used but have not proved very attractive. This has perhaps inhibited the wider use of activated carbon as a means for the economic concentration of gold from dilute cyanide solutions. Later work[5] related to the mechanism of gold adsorption onto activated carbon has shown that both the kinetics and the equilibrium capacity of gold adsorption are dependent on the ionic strength and the presence of the calcium cation, as well as on the hydroxyl anion in the adsorbent medium. It was also demonstrated that, provided the adsorbed gold that initially concentrated on the activated carbon as the calcium aurocyanide complex was converted to either the sodium or potassium complex by a pretreatment stage involving an alkaline metal carbonate–hydroxide reagent, such complexes could be successfully eluted from the carbon with hot deionized water. The quality of the water used as eluant was shown to have a very definite effect on the efficiency of elution.

As the pretreatment of gold-loaded activated carbon with an alkaline metal carbonate prior to elution with deionized water involved the precipitation of calcium carbonate to effect the elution of either the potassium or sodium aurocyanide complex, the build-up of calcium carbonate on the carbon was at that stage a distinct disadvantage. The poor elution characteristics of the adsorbed silver and base metals due to the instability of such complexes under high pH pretreatment conditions were also indicated. The present investigation was thus carried out in order to optimize the elution of all such adsorbed values from activated carbon using deionized water as the eluant. The pretreatment of loaded activated carbon with a relatively strong solution of sodium cyanide and sodium hydroxide before elution was shown to have considerable promise in this latter regard.

Materials

Type-G210 granulated coconut-shell carbon (0,5 to 2 mm) supplied by Le Carbone South Africa (Pty) Limited was loaded with approximately 0,3 per cent gold from a gold-plant effluent containing about 0,1 g/t gold. The charcoal was then washed with dilute hydrochloric acid to remove any calcium carbonate that had built up on the charcoal. The chemical compositions of the virgin,

*Anglo American Research Laboratories, Johannesburg.

TABLE I
THE COMPOSITION OF ACTIVATED CARBON

Element	Virgin carbon g/t	Loaded carbon g/t	Acid-washed loaded carbon g/t
Na	18 500	4 500	—
K	10 400	1 900	—
Ca	640	14 800	4 100
Mg	300	100	—
Au	0	2 700	2 900
Ag	< 10	45	64
Fe	1 500	1 500	1 600
Ni	10	2 500	2 200
Cu	20	240	90
Co	4	2	—
Zn	120	380	80

loaded, and acid-washed carbon are shown in Table I.

All the other chemicals used in the present investigation were of C.P. grade.

Method

With the aid of a vibrator, 15 g of air-dried, acid-washed carbon (moisture content about 22 per cent) was loaded as a wet slurry at 90 °C into a water-jacketed chromatography column (0,9 cm internal diameter by 40 cm long) in series with an LKB 17 000 Minirac fraction collector. Unless otherwise specified, the pretreatment solutions of sodium hydroxide and sodium cyanide were pumped at about 1 bed volume per hour (apparent flow velocity $9{,}2 \times 10^{-3}$ cm/s; 1 bed volume=25 ml) after the column had been drained of surplus water, while the elution flow rates varied from ½ to 2 bed volumes per hour. Up to 12 bed volumes of eluate were collected. Finally, the carbon was removed from the column and, after being oven-dried at 110 °C, was analysed and the adsorption activity determined.

The capacity of the eluted carbon for adsorbing gold was evaluated as follows. A given mass of oven-dried carbon was shaken for 17 hours under ambient conditions with 100 ml of approximately 100 g/t gold solution added as potassium aurocyanide in a standard borate buffer of pH 10,0. The slurry was then filtered through Whatman No. 540 filter paper and the filtrate analysed for gold. The degree of gold adsorption occurring at equilibrium was then measured by use of the empirical Freundlich isotherm as previously described[5].

Results

Pretreatment with Sodium Hydroxide

Fig. 1 illustrates the elution of gold, silver, and nickel from activated carbon pretreated with potassium carbonate–potassium hydroxide as outlined in an earlier study[5]. The recovery of gold was 99,1 per cent in 12 bed volumes. This pretreatment reagent was then compared with the use of an equimolar sodium carbonate–sodium hydroxide reagent, which was shown to have very similar elution characteristics for gold, silver, copper, and nickel (Table II). The use of sodium salts rather than potassium salts would have obvious economic advantages.

As it had already been shown that potassium hydroxide had a markedly beneficial effect on the elution of gold when added to a potassium carbonate pretreatment reagent[5], it was questionable whether the

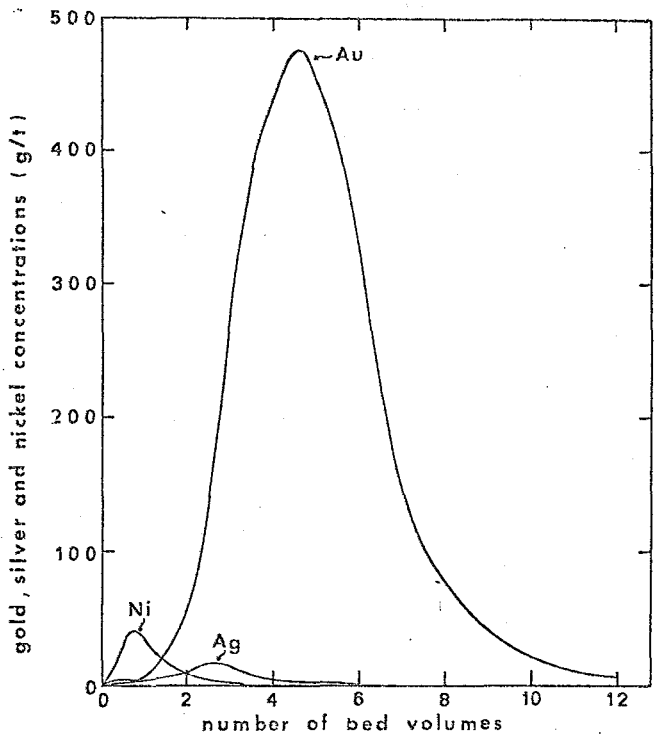


Fig. 1—The elution of gold, silver and nickel from activated carbon after pretreatment with 10 per cent potassium carbonate–5 per cent potassium hydroxide

carbonate or the higher pH (hydroxide) was individually responsible for the efficient elution, or whether a strong hydroxide pretreatment in the absence of added carbonate would suffice.

Fig. 2 (Table II) presents the gold-elution curves for the strong equimolar hydroxide pretreatment reagents of potassium hydroxide, sodium hydroxide, and lithium hydroxide. From these results, it can be seen that, while efficient gold elution can be carried out in the absence of added carbonate, the use of potassium hydroxide is only marginally better than the use of sodium hydroxide, which is significantly better than the use of lithium hydroxide. The elution recovery of gold was 98,6 per cent in 12 bed volumes after pretreatment with 14 per cent sodium hydroxide, but the recoveries of silver, copper, and nickel were poor (Table II).

At this stage in the present investigation, the poor elution characteristics of silver, copper, and nickel were thought to be due, in part at least, to the instability of such cyanide complexes under the prescribed high pH pretreatment conditions. This instability would be reduced considerably by the pretreatment of the carbon with sodium cyanide. Furthermore, such a pretreatment would promote the transformation of the calcium aurocyanide to the sodium form more efficiently than would a carbonate pretreatment, because calcium cyanide is more soluble than either calcium carbonate or calcium hydroxide and would be eluted from the carbon rather than remain as an insoluble precipitate. This early elution of soluble calcium from the system would surely have a very beneficial effect.

Pretreatment with Sodium Cyanide

Figs. 3 to 5 show the corresponding curves for the

elution of gold, silver, and nickel respectively when the loaded activated carbon was pretreated with 10 per cent sodium cyanide containing up to 20 per cent sodium hydroxide. It can be seen that, as the concentration of sodium hydroxide in the pretreatment reagent was decreased, the elution of gold, silver, and nickel all improved, with sharper, more symmetrical peaks and earlier elution profiles in all cases. This improvement in the elution efficiency of all three species as the concentration of sodium hydroxide in the pretreatment reagent decreased supports the view that, for silver, copper, and nickel (Table II), the stability of the respective cyanide complexes is lower at high pH values. The elution recovery was 99,5 per cent gold, 98,0 per cent silver, 92,3 per cent copper, and 95,5 per cent nickel in 12 bed volumes after the carbon had been treated with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide. A very similar effect was noted when the concentration of the sodium cyanide was increased from 1 to 10 per cent in the presence of 14 per cent sodium hydroxide (Table II).

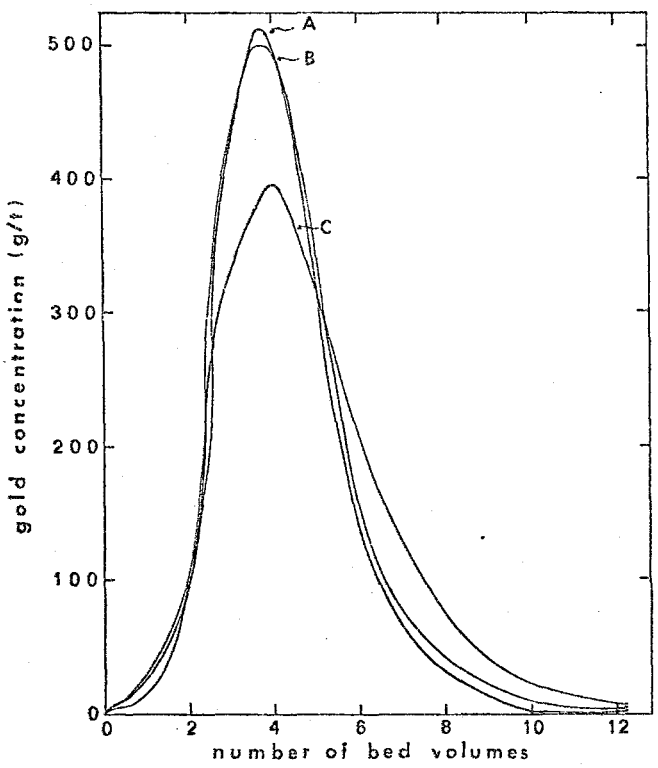


Fig. 2—The elution of gold from activated carbon after pretreatment with equimolar hydroxide reagents
A 10 per cent potassium hydroxide,
B 7 per cent sodium hydroxide
C 4,3 per cent lithium hydroxide

Effect of Flow Velocity

From Table III it can be seen that the elution of gold, silver, and nickel from activated carbon pretreated with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide is not seriously affected by elution rates within the prescribed experimental parameters. The extraction of gold after seven hours of elution at 1 bed volume per hour was more than 97 per cent, while the recovery of gold after 3,5 hours of elution at 2 bed volumes per hour was 90,8 per cent.

Elution at High Gold Loadings

Type-G215 activated carbon supplied by Le Carbone (Pty) Limited was loaded with between 4 and 6 per cent gold from clarified gold-plant pregnant solutions averaging 8 g/t gold. The loaded carbon was then treated with sodium cyanide–sodium hydroxide reagents and eluted at 90°C as previously described.

Elution curve A (Fig. 6) shows the effect of eluting a loaded carbon (4,0 per cent gold, 600 g/t silver, and 2600 g/t nickel) after pretreatment with 10 per cent sodium cyanide–14 per cent sodium hydroxide. The gold concentration at peak elution was 8500 g/t, with recoveries of 98,4 per cent of the gold, 83,0 per cent of the silver, and over 99,6 per cent of the nickel in 6 bed volumes (Table IV). The recovery of silver was poor owing to the high concentration of sodium hydroxide used (see Table II).

TABLE II

THE EFFECT OF VARIOUS PRETREATMENT PROCEDURES ON THE ELUTION OF GOLD FROM ACTIVATED CARBON

Eluant: Deionized water at 0,5 bed volume per hour and 90°C

Pretreatment reagent	Elution peak concentration, g/t			Recovery after 12 bed volumes, %				Residual carbon activity mg Au/g C
	Au	Ag	Ni	Au	Ag	Cu	Ni	
1 bed volume 10% K_2CO_3–5% KOH	478	13	43	99,1	80,0	89,0	ca 40	14
1 bed volume 10% KOH	510	8	—	98,9	62,5	78,9	85,3	14
1 bed volume 7% NaOH (equivalent to 10% KOH)	500	10	—	98,1	89,1	86,7	87,8	8
1 bed volume 4,3% LiOH (equivalent to 10% KOH)	390	9	—	97,2	92,2	52,3	86,2	13
1 bed volume 14% NaOH	580	11	—	98,6	65,7	81,2	89,7	13
0,5 bed volume 10% NaCN	940	40	1 600	99,5	98,0	95,6	97,1	13
0,5 bed volume 1% NaOH–10% NaCN	862	37	1 930	99,5	98,0	92,3	95,5	12
0,5 bed volume 5% NaOH–10% NaCN	825	25	1 430	99,1	95,4	91,2	97,4	12
0,5 bed volume 14% NaOH–10% NaCN	820	25	1 350	99,4	92,2	98,9	96,8	13
0,5 bed volume 20% NaOH–10% NaCN	770	23	1 260	99,4	82,9	—	98,2	12
1 bed volume 14% NaOH	580	11	—	98,6	65,7	81,2	89,7	13
0,5 bed volume 1% NaCN–14% NaOH	630	21	1 250	99,4	92,2	—	95,9	13
0,5 bed volume 3% NaCN–14% NaOH	705	28	1 560	99,1	94,9	80,0	96,0	14
0,5 bed volume 5% NaCN–14% NaOH	670	25	1 360	99,4	82,9	84,5	97,1	13
0,5 bed volume 8% NaCN–14% NaOH	740	24	1 560	99,4	86,0	93,4	98,6	13
0,5 bed volume 10% NaCN–14% NaOH	820	25	1 350	99,4	92,2	98,9	96,8	13

Elution curve B (Fig. 6) illustrates the elution of loaded carbon (5,9 per cent gold, 1900 g/t silver, and 6000 g/t nickel) after being pretreated with 12 per cent sodium cyanide–1 per cent sodium hydroxide. The gold concentration at peak elution was 14 400 g/t, with the recovery of 99,8 per cent of the gold, 97,1 of the silver, and 99,9 per cent of the nickel in 6 bed volumes (Table IV). The recovery of silver was significantly better than that shown in curve A because of the lower concentration of sodium hydroxide. Similar elution data are plotted in curve C, which illustrates the elution of gold from activated carbon loaded with 4,1 per cent gold, 1200 g/t silver, and 1700 g/t nickel (Fig. 6). In this series of elution investigations at high gold loadings, the carbon was not washed with acid before the elution, but this omission appeared to have little or no effect on the efficiency of the gold, silver, and nickel elution. (More recent investigations in these laboratories[7] have shown that the acid washing of loaded carbon has a significantly beneficial effect on the elution of gold and silver.)

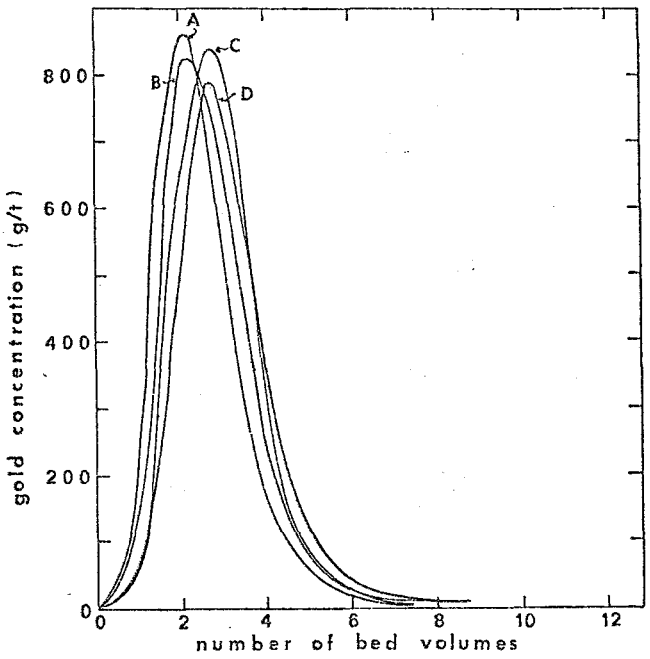


Fig. 3—The elution of gold from activated carbon as a function of the concentration of sodium hydroxide in the pretreatment reagent
- A 1,0 per cent sodium hydroxide–10 per cent sodium cyanide
- B 5 per cent sodium hydroxide–10 per cent sodium cyanide
- C 14 per cent sodium hydroxide–10 per cent sodium cyanide
- D 20 per cent sodium hydroxide–10 per cent sodium cyanide

As the elution of gold with deionized water relies on the displacement of a physical rather than a chemical equilibrium, the higher the initial gold loading on the carbon, the more efficient in terms of gold recovery the elution procedure becomes. This is demonstrated by a comparison of the above results with earlier results at lower gold loadings.

Elution at Higher Temperatures

Earlier findings[6] had shown that the gold–cyanide complex and particularly the silver–cyanide complex are susceptible to decomposition at high elution temperatures and pressures when potassium carbonate–potassium hydroxide is used in the pretreatment. In the present investigation, loaded carbon was pretreated

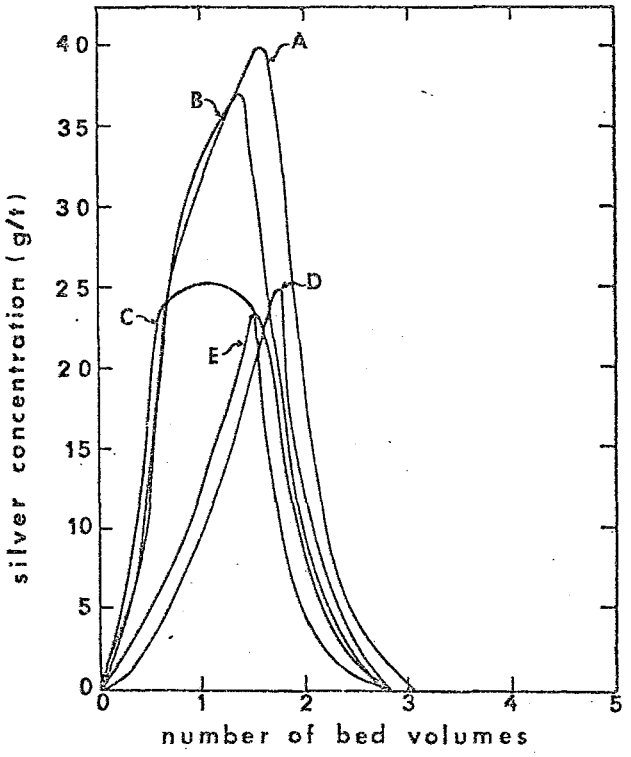


Fig. 4—The elution of silver from activated carbon as a function of the concentration of the sodium hydroxide in the pretreatment reagent
- A 0 per cent sodium hydroxide–10 per cent sodium cyanide
- B 1,0 per cent sodium hydroxide–10 per cent sodium cyanide
- C 5 per cent sodium hydroxide–10 per cent sodium cyanide
- D 14 per cent sodium hydroxide–10 per cent sodium cyanide
- E 20 per cent sodium hydroxide–10 per cent sodium cyanide

TABLE III

EFFECT OF APPARENT FLOW VELOCITIES ON THE ELUTION OF GOLD FROM ACTIVATED CARBON

Pretreatment: 0,5 bed volume of 10 per cent sodium cyanide–1 per cent sodium hydroxide

Elution rates at 90°C	Concentration at elution peak, g/t			Recovery after 7 bed volumes, %				Residual carbon activity, mg Au/g C
	Au	Ag	Ni	Au	Ag	Cu	Ni	
0,5 bed volume/h ($4,9\times10^{-3}$ cm/s)	862	37	1 930	99,0	98,0	92,3	95,5	12
1 bed volume/h ($9,2\times10^{-3}$ cm/s)	820	34	1 540	97,3	95,4	91,2	96,7	—
2 bed volumes/h ($17,0\times10^{-3}$ cm/s)	625	26	1 190	90,8	95,4	83,4	95,2	—

with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide and eluted under pressure at 125°C as previously described[5]. At an elution rate of 1,7 bed volumes per hour involving 4,7 bed volumes of eluate (Table V), 99,2 per cent of the gold and 95,3 per cent of the silver were recovered in an elution time of 3 hours. This can be compared with 14 hours for elution at 0,5 bed volume per hour at 90°C (Table III). From these data, it can be seen that, if time is of major importance, the use of increased elution temperatures can have significant advantages.

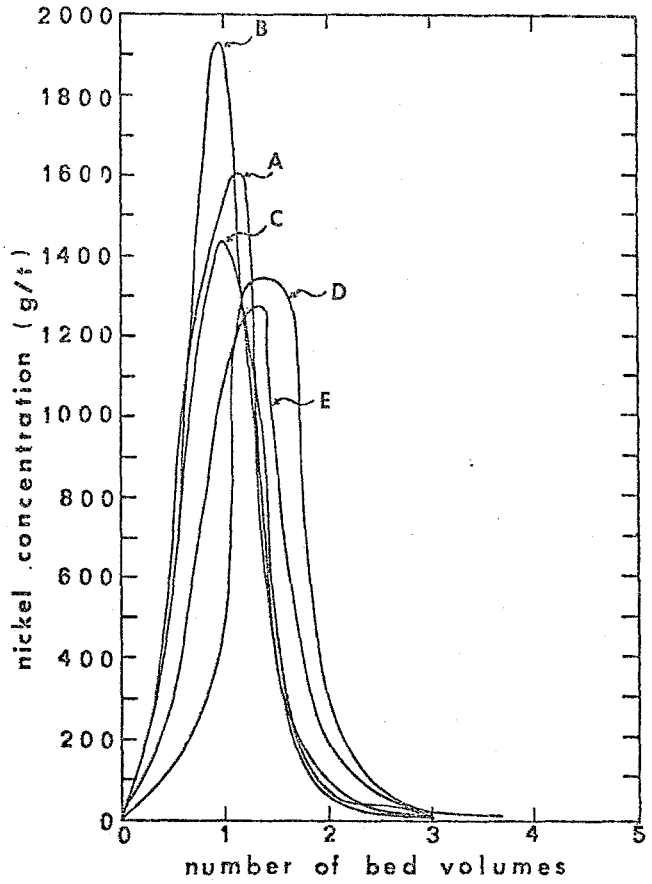


Fig. 5—The elution of nickel from activated carbon as a function of the concentration of the sodium hydroxide in the pretreatment reagent

- A 0 per cent sodium hydroxide–10 per cent sodium cyanide
- B 1,0 per cent sodium hydroxide–10 per cent sodium cyanide
- C 5 per cent sodium hydroxide–10 per cent sodium cyanide
- D 14 per cent sodium hydroxide–10 per cent sodium cyanide
- E 20 per cent sodium hydroxide–10 per cent sodium cyanide

Other Elution Conditions

Figs. 7 to 10 illustrate the elution of gold, calcium, silver, and nickel from loaded carbon by three different sets of conditions: pretreatment with sodium cyanide and sodium hydroxide followed by elution with deionized water at 90°C (condition A), pretreatment with potassium carbonate–potassium hydroxide followed by elution with deionized water at 90°C (condition B), and elution with sodium sulphide–sodium hydroxide[8] at 65°C (condition C). Further elution details are given in Table VI.

Fig. 7 shows that condition A while being comparable with condition C provides a more efficient elution mechanism (a recovery of 99,0 per cent of the gold in 7 bed volumes) than that provided by condition B. This is due, in part at least, to the mass action effect involved in the replacement of calcium with sodium in the cyanide complexes as previously described. In the cyanide

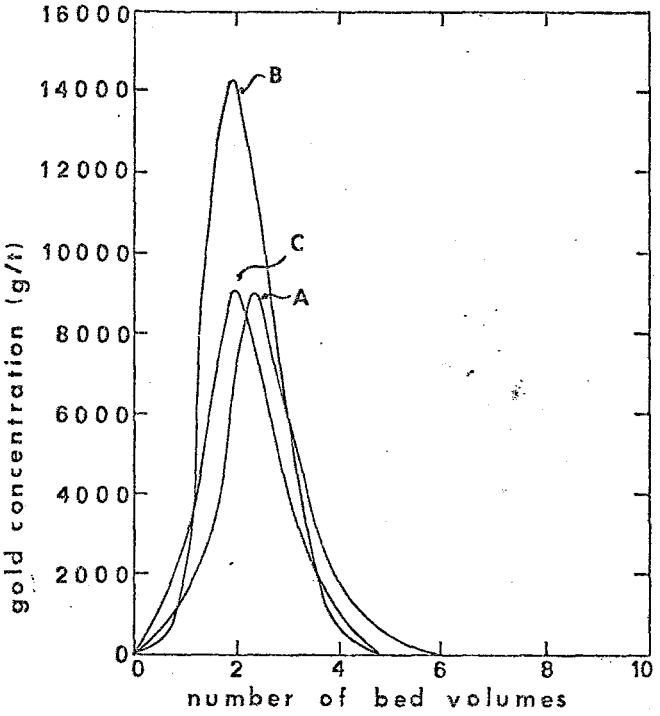


Fig. 6—The elution of gold from activated carbon at high gold loadings

- A 4,0 per cent gold loading, pretreatment with 14 per cent sodium hydroxide–10 per cent sodium cyanide
- B 5,9 per cent gold loading, pretreatment with 1,0 per cent sodium hydroxide–12 per cent sodium cyanide
- C 4,1 per cent gold loading, pretreatment with 1,0 per cent sodium hydroxide–12 per cent sodium cyanide

TABLE IV

ELUTION AT HIGHER GOLD LOADINGS

Eluant: Deionized water at 0,5 bed volume per hour and 90°C

Pretreatment reagent	Original loading, %			Concentration at elution peak, g/t			Recovery after 6 bed volumes, %		
	Au	Ag	Ni	Au	Ag	Ni	Au	Ag	Ni
0,5 bed volume 10% NaCN–14% NaOH ..	4,0	0,06	0,26	8 500	140	970	98,4	83,0	99,6
0,5 bed volume 12%NaCN–1% NaOH	5,9	0,19	0,60	14 000	1 000	3 500	99,8	97,1	99,9
0,5 bed volume 12% NaCN–1% NaOH	4,1	0,12	0,17	8 800	470	720	99,5	91,0	99,3

pretreatment, the calcium is removed efficiently as a soluble product, whereas, in the pretreatment with potassium carbonate, the product is a precipitate of calcium carbonate. The virtually complete removal of calcium as soluble calcium cyanide would appear to have a more beneficial effect on the subsequent elution mechanism. This is well illustrated in Fig. 8, which shows the elution of calcium after the two pretreatments.

Fig. 9 illustrates the elution of silver under the same three sets of conditions as shown in Fig. 7. Again, the pretreatment with sodium cyanide–sodium hydroxide (A) yields significantly better recoveries of silver than does pretreatment B. The poor recoveries of silver after pretreatment B are undoubtedly due to the high concentration of potassium hydroxide used, while the very poor recovery of silver from the elution with sodium sulphide–sodium hydroxide (C) is due to the precipitation of silver sulphide on the carbon. Similarly, Fig. 10 illustrates the elution of nickel under these three sets of conditions and, again, A (pretreatment with sodium cyanide–sodium hydroxide) has very distinct advantages over the other two sets of conditions.

Activity of Eluted Carbon

The virgin activated carbon used in the present investigation had an equilibrium capacity constant, k, of 22 mg of gold per gram of carbon. As can be seen from the data presented (Tables II, III, V, VI), the capacity constants of the used carbon after elution was about 8 to 14 mg. Thermogravimetric analysis was carried out on various samples of loaded, virgin, and eluted carbons to determine the losses in mass resulting from the volatile organic constituents present in the material. It was thought that the organic constituents originating from mine waters were primarily responsible for reducing the activity of the used carbon.

The largest losses in mass in the range 180 to 540°C were recorded for the eluted and loaded carbons. The virgin carbon showed only a small loss in mass in this temperature range, suggesting the presence of organic poisons in the used carbons that are firmly bound to the carbon and that are not easily removed during elution. This strongly suggests that the used carbon should be regenerated by thermal means.

TABLE V

THE PRESSURE ELUTION OF GOLD FROM ACTIVATED CARBON AT 125°C and 200 kPa

Eluant: Deionized water at 1,7 bed volumes per hour

Pretreatment reagent	Recovery after 4,7 bed volumes, %			Residual carbon activity mg Au/g C
	Au	Ag	Ni	
1 bed volume 5%NaCN—10% NaOH	98,8	93,3	98,4	12
1 bed volume 14%NaCN—1% NaOH	99,2	95,3	—	14

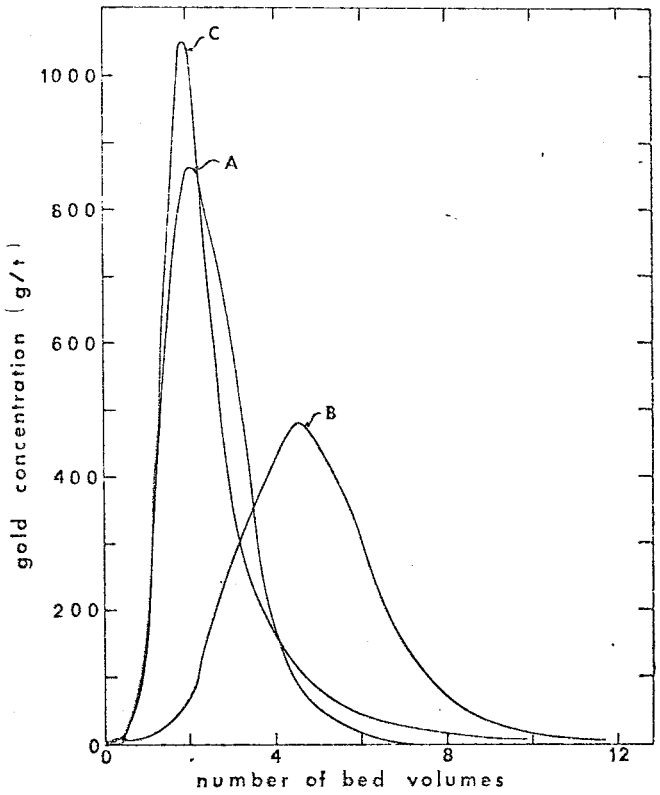


Fig. 7—The elution of gold from activated carbon under three sets of conditions

- A pretreatment with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide, elution with deionized water at 90°C
- B pretreatment with 10 per cent potassium carbonate–5 per cent potassium hydroxide, elution with deionized water at 90°C
- C elution with 3 per cent sodium sulphide–3 per cent sodium hydroxide at 65°C

TABLE VI

COMPARISON OF THREE DIFFERENT SETS OF ELUTION CONDITIONS

	Conditions	Concentration at elution peak, g/t			Recovery after 7 bed volumes, %				Residual carbon activity mg Au/g C
		Au	Ag	Ni	Au	Ag	Cu	Ni	
A	0,5 bed volume 10% NaCN–1% NaOH, followed by deionized water at 0,5 bed volume per hour and 90°C	862	37,0	1 930	99,0	98,5	92,0	95,5	12
B	1 bed volume 10% k_2CO_3–5% KOH, followed by deionized water at 9,5 bed volumes per hour and 90°C	478	13,0	43	85,0	95,5	89,0	ca 40	14
C	10 bed volumes of 3% Na_2S–3% NaOH–½% Na_2SO_3 at 65°C at 2 bed volumes per hour, followed by 10 bed volumes water, followed by 4 bed volumes 5% HNO_3, followed by 10 bed volumes water	1 050	0,2	300	94,3	< 10,0	44,5	65,4	10

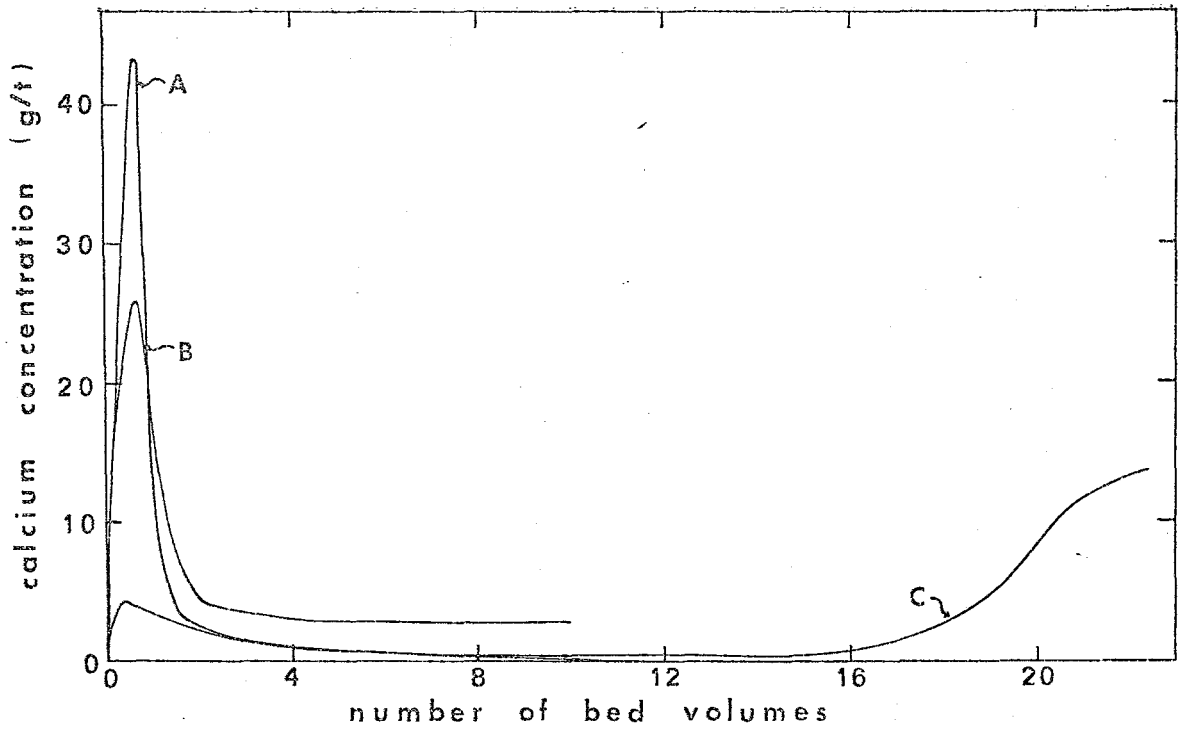


Fig. 8—The elution of calcium from activated carbon after three types of pretreatment

A pretreatment with 10 per cent sodium cyanide, elution with deionized water at 90°C

B pretreatment with 10 per cent sodium cyanide–14 per cent sodium hydroxide, elution with deionized water at 90°C

C pretreatment with 10 per cent potassium carbonate–5 per cent potassium hydroxide, elution with deionized water at 90°C

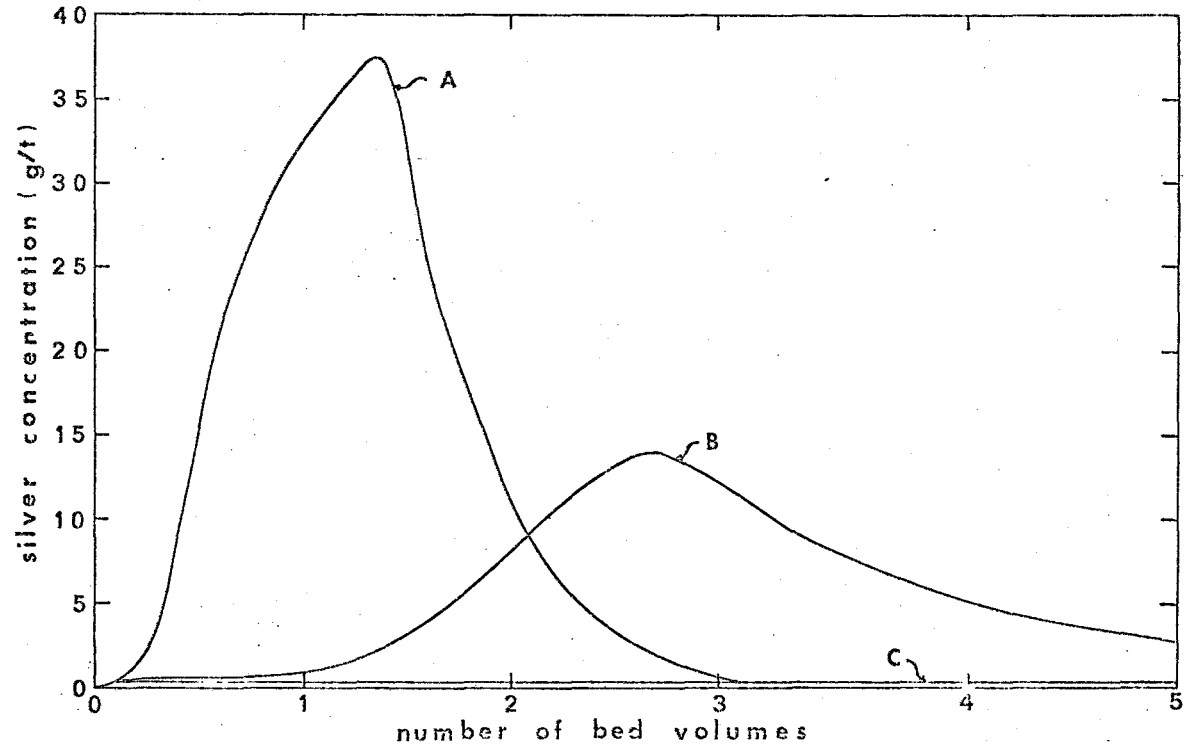


Fig. 9—The elution of silver from activated carbon under three sets of conditions

A pretreatment with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide, elution with water at 90°C

B pretreatment with 10 per cent potassium carbonate–5 per cent potassium hydroxide, elution with deionized water at 90°C

C elution with 3 per cent sodium sulphide–3 per cent sodium hydroxide at 65°C

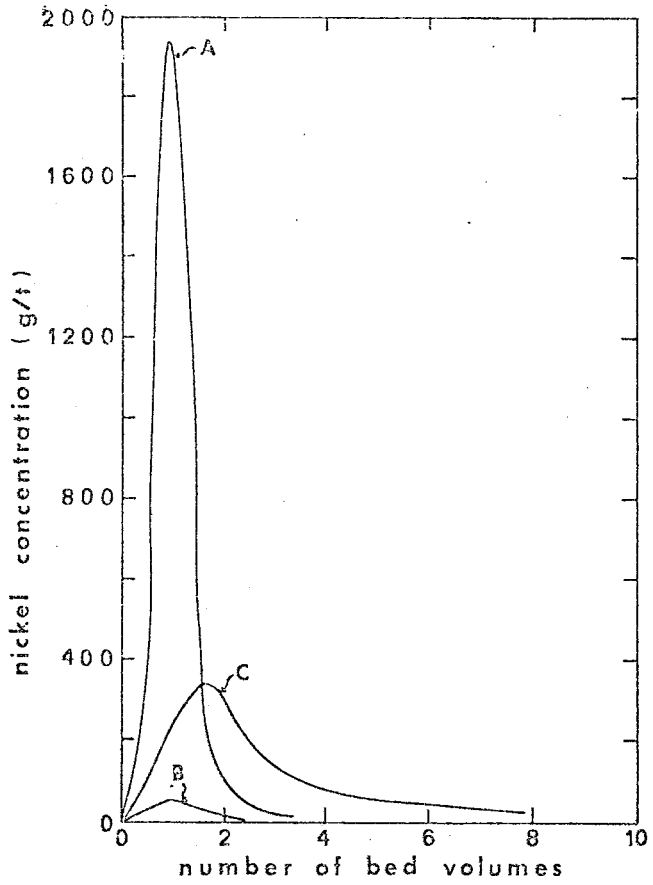


Fig. 10—The elution of nickel from activated carbon under three sets of conditions

- A pretreatment with 10 per cent sodium cyanide–1,0 per cent sodium hydroxide, elution with deionized water at 90°C
- B pretreatment with 10 per cent potassium carbonate–5 per cent sodium hydroxide, elution with deionized water at 90°C
- C elution with 3 per cent sodium sulphide–3 per cent sodium hydroxide at 65°C

Discussion

The present findings regarding the elution of gold, silver, and base metals from activated carbon using deionized water emphasize the important role played by the calcium cation in effecting efficient extraction of gold. Pretreatment of the loaded carbon with relatively strong sodium cyanide has a most beneficial effect on the elution of the otherwise unstable silver and base-metal complexes. As was demonstrated, the use of 0,5 bed volume of pretreatment reagent containing 10 per cent sodium cyanide that had been further stabilized with 1,0 per cent sodium hydroxide for safety reasons, followed by elution with deionized water, results in a virtually 100 per cent recovery of the adsorbed gold, silver, and base-metal values. The elution procedure takes a relatively short time, about 14 hours at 90°C (7 bed volumes) or 3 hours if pressure elution at 125°C (4 to 5 bed volumes) is used. This can be compared with the pressure stripping of gold from carbon[3] in 2 to 6 hours at 150 to 160°C using 12 bed volumes of 0,4 per cent sodium hydroxide–0,03 per cent sodium cyanide, or with the 24 to 30 hours required for atmospheric-pressure elution as described by Zadra[2].

When compared with other elution procedures, the proposed method appears to have several economic advantages. Firstly, there is no precipitation of calcium carbonate or of silver and base-metal sulphides. Such precipitates tend to build up on the carbon with extended use, resulting in carbon poisoning if they are not removed. With regard to reagent costs, the re-use of both regenerated cyanide in an eluate-electrowinning stage and caustic soda in other parts of the gold plant appear to be distinct possibilities. While the cost of deionized water or steam condensate would incur a small additional expenditure, other gold-elution procedures are unlikely to compete in this latter regard. Lastly, because the mechanism of elution after pretreatment entails the shifting of a physical rather than a chemical equilibrium, the efficiency of the elution procedure increases as the amount of loaded gold is increased.

Acknowledgement

The authors are indebted to T. Szendrei for the thermogravimetric analyses, C. G. Schmidt for the pressure-elution data, and the analytical section of Anglo American Research Laboratories for the chemical analyses.

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Training programmes

The Federation of Societies of Professional Engineers has announced that the following four programmes for engineers-in-training have been accepted in principle by the S.A. Council of Professional Engineers:

UKOR (Electrical Engineers)
City of Johannesburg (City Engineers, Civil)
SAPPI (Chemical Engineers)
Dept of Water Affairs (Civil Engineers).

Further studies on the elution of gold from activated carbon using water as the eluant

by R. J. DAVIDSON*, Ph.D. (Rhodes) and V. VERONESE†, Dip. Chem. Analyst

SYNOPSIS

Some of the parameters affecting the elution of gold from freshly loaded activated carbon using pure water as an eluant were studied. The procedure involved pretreatment of the gold-loaded activated carbon with a strong reagent of sodium cyanide and sodium hydroxide, followed by elution of the gold with water. Gold-loaded carbon obtained from three different sources was studied.

It was shown that the elution temperature was the most critical parameter, and column operation at 110°C (gauge pressure of 50 to 100 kPa) is suggested.

The effects of variations in the concentration of sodium cyanide required for the pretreatment of loaded carbon and of variations in elution flowrate were also investigated. Both these parameters were found to have only a marginal effect on the efficiency of gold elution. The investigation also indicated that gold-elution rates were affected to a considerable degree by the particle size of the carbon concerned. Higher elution flowrates could be used when carbons of finer particle size were eluted. A modified eluate-collection procedure was also developed during the course of the investigation. Oven-drying of the loaded carbon was shown to have little or no effect on the efficiency of the subsequent gold elution.

The calcium and silica contents of the various carbon samples were also determined. It was immediately evident that the calcium content was very considerable in the case of loaded carbons originating from carbon-in-pulp operations. Calcium concentration was particularly high where air agitation had been employed in the carbon-contacting circuit. To some degree, the calcium build-up explains the serious carbon fouling taking place, resulting in a decreased efficiency of gold adsorption. The fouling problem was found to be less severe when mechanical agitation was used. No poisoning by calcium carbonate was observed where the pregnant solution had been acidified before it was contacted with the carbon.

SAMEVATTING

Sommige van die parameters wat 'n invloed uitoefen op die eluering van goud uit vars gelaaide geaktiveerde koolstof, deur suiwer water as uitwasmiddel te gebruik, was bestudeer. Die prosedure het die voorbehandeling van goudgelaaide geaktiveerde koolstof met 'n sterk natriumsianied/natriumhidroksied reagens gevolg deur eluering van die goud met water, behels. Goudgelaaide koolstof, verkry van drie verskillende bronne, was bestudeer.

Dit was aangetoon dat die elueringstemperatuur die mees kritieke operasionele parameter was en die werking van die kolom teen 110°C (meterdruk van 50 tot 100 kPa) word aanbeveel.

Die effek van variasies in die natriumsianied konsentrasie benodig vir die voorbehandeling van die gelaaide koolstof sowel as variasies in die eluering vloeitempo was ook ondersoek. Beide hierdie parameters het slegs 'n geringe effek op die doeltreffendheid van die goudeluering gehad. Die ondersoek het ook aangedui dat goudelueringstempos tot 'n noemenswaardige mate geaffekteer is deur die partikelgrootte van die betrokke koolstof. Hoër elueringsvloeitempos kon gebruik word by die eluering van fyner partikelgrootte koolstowwe. 'n Veranderde eluering versamel prosedure was ook ontwikkel gedurende die verloop van die ondersoek. Dit was aangetoon dat oonddroging van die gelaaide koolstof min tot geen effek gehad het op die doeltreffendheid van die daaropvolgende goudeluering nie.

Die kalsium en silika inhoud van die verskillende koolstof monsters was ook bepaal. Dit was onmiddellik duidelik dat die kalsium inhoud aansienlik was in die geval van die gelaaide koolstowwe afkomstig van die koolstof-in-pulp tipe operasies. Die kalsiumkonsentrasie was besonder hoog in die geval waar lugroering gebruik was in die koolstofkontakbaan. Tot 'n seker mate verduidelik die opbou van kalsium die ernstige koolstofbesoedeling wat plaasgevind het, met 'n gevolglike afname in goudabsorpsiedoeltreffendheid. Die besoedelingsprobieem was minder ernstig wanneer meganiese-roering gebruik was. Geen kalsiumkarbonaat vergiftiging was waargeneem in die geval waar die dragtige oplossing aangesuur was voor kontak met die koolstof nie.

Introduction

Because of the recent increase in the price of gold, the treatment of lower-grade ores, as well as waste-rock dump material and slimes-dam residues, is currently receiving considerable world-wide attention. Carbon-in-pulp type operations for the recovery of gold from such materials offer numerous advantages in the form of lower capital and operating costs[1]. Consequently, considerable emphasis, both at the Anglo American Research Laboratories (AARL) and the National Institute for Metallurgy (NIM), has been placed on the optimization of this field of gold technology. Pilot-plant studies and research-and-development efforts have been directed towards improving the carbon-contacting circuits as well as other ancillary aspects. Investigations carried out at AARL have demonstrated more efficient means of gold elution[2] and gold electrowinning[3, 4], together with relatively inexpensive means of regenerating used carbon thermally[4] and chemically[5].

A carbon-in-pulp plant is currently being commissioned at Golden Dumps (Pty) Ltd[6], where crushed waste-rock fines are being treated for the recovery of gold. At the Fairview Gold Mine[7], flotation concentrates produced from calcined cyanidation residues are being processed. A carbon-in-pulp pilot plant has already been satisfactorily operated at Durban Deep Gold Mine[8] on low-grade run-of-mine material, while a pilot operation using activated carbon in a countercurrent column for the recovery of gold from unclarified gold-plant effluent was successfully run at the Vaal Reefs Gold Mine[4].

The effectiveness of all the above processes depends to a large degree on the development of an efficient means of stripping the gold from the loaded carbon. Stripping with a hot dilute solution of sodium hydroxide and sodium cyanide at atmospheric pressure, followed

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by electrowinning, has been employed[9]. However, this method has the disadvantage that the procedure involving simultaneous stripping and electrowinning is slow. Elevated temperatures and pressure have also been employed,[10] thereby accelerating the stripping rate, but this technique requires the use of a pressurized reactor, which makes the process more complicated. The use of mixtures consisting of an organic solvent, sodium hydroxide, and water has also been considered[11, 12]. Apart from relatively expensive reagents, the use of such solvents in a gold plant would constitute a considerable fire hazard. Furthermore, investigations carried out at AARL[12] have indicated the inefficient elution of silver when such organic-solvent mixtures are used. A method involving the pretreatment of the loaded carbon with a relatively strong reagent consisting of sodium cyanide and sodium hydroxide and elution with water has also been suggested[2]. This method has many advantages arising from the efficiency, speed, and cost of the gold elution[4]. The recovery and regeneration of cyanide in the subsequent electrowinning operation may also render the process more economically viable.

In view of the likely developments in the use of activated carbon for the recovery of gold and the elution procedure developed at AARL, further elution studies were undertaken. These studies were designed so that certain operational parameters of an elution plant could be optimized, and so that an assessment could be made of the elution response of loaded carbons from three different sources.

Materials

Sample A

A sample of activated type SCII coconut-shell carbon (minus 2380 μm plus 1000 μm, Davy Ashmore S.A. (Pty) Ltd) loaded with gold (16 900 g/t) and silver (300 g/t) was obtained from the Fairview Gold Mine carbon-in-pulp operation.

Sample B

Two samples of loaded type SCII coconut-shell carbon (minus 2380 μm plus 1000 μm) from Golden Dumps (Pty) Ltd were obtained during the early stages of plant commissioning. Sample B, taken after the first week of operation, contained 1604 g/t gold and 182 g/t silver. Sample B1, taken after the second week of commissioning, contained 3140 g/t gold and 365 g/t silver.

Sample C

Pregnant gold-plant solution from the Western Deep Levels Gold Mine, which had been acidified to a pH value between 4 and 5, was pumped through a stainless-

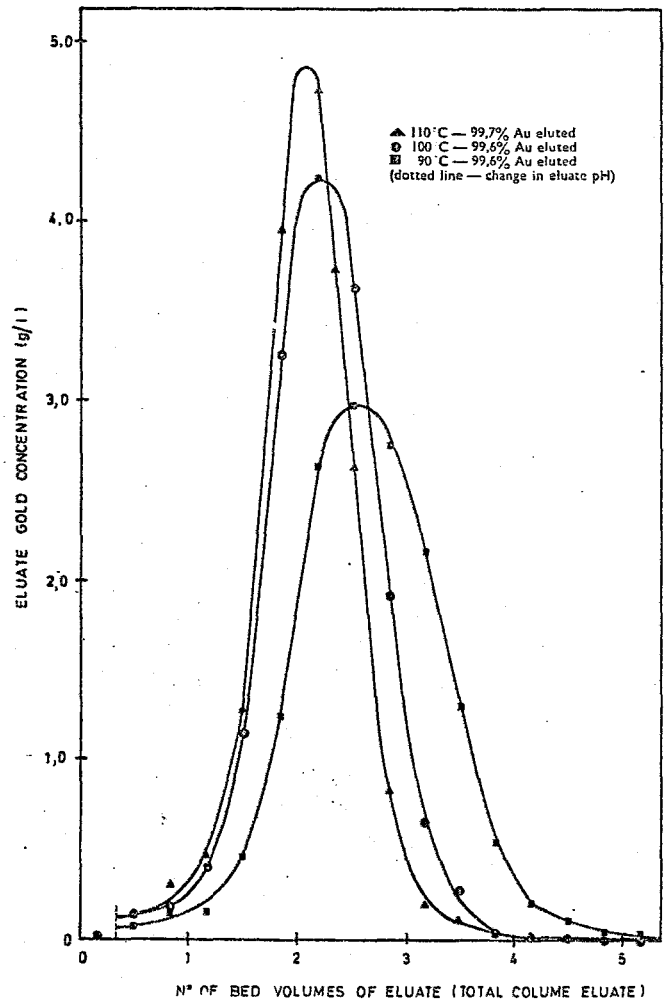

Fig. 1—The effect of temperature on the elution of carbon sample C

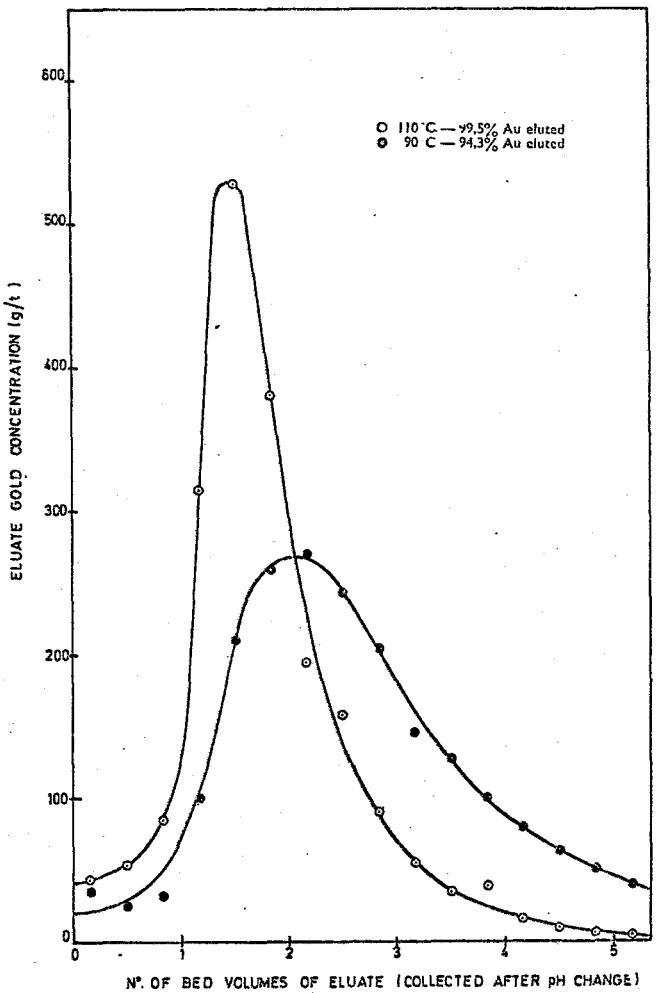

Fig. 2—The effect of temperature on the elution of carbon sample B

steel heat-exchange spiral immersed in a water-bath at 20 °C. The spiral was in series with an adsorption column containing granular activated carbon (minus 1190 μm plus 840 μm of type G215 coconut-shell type carbon supplied by Le Carbone (Pty) Ltd). The influent solution was pumped at a constant flowrate by an MPL Monopump metering pump. The adsorption column (16 mm internal diameter and 100 cm long) was packed with up to 100 g of oven-dried granular activated carbon that had previously been equilibrated at a pH value between 4 and 5. The superficial flow velocity in the column was maintained at 0,8 m/min, and pumping continued until a loading of approximately 2 per cent gold on the carbon had been achieved (ca 24 h). Flow through the column was always in a downward direction. The piloting of a similar process using a countercurrent ion-exchange type of column operation is due to commence at Durban Deep Gold Mine in the near future[4].

Method

The method of gold elution used in the investigation was as follows: 150 cm³ of wet-settled gold-loaded carbon was placed in an oil-jacketed stainless-steel elution column of 0,9 cm internal diameter and 75 cm long; oil was heated in a separate vessel and then circulated through the jacket to maintain the contents of the column at the desired elution temperature; before entering the elution column, the eluting solution was heated to elution temperature by being passed through a stainless-steel spiral placed inside the oil-jacket; the elution solution was pumped through the column by an MPL Monopump metering pump. Unless otherwise specified, the elution was carried out at 110 °C at a gauge pressure of from 80 to 120 kPa. Pressure within the system was maintained by means of a stainless steel blow-off valve and a manually operated stainless-steel needle valve situated on the effluent line of the column. Column eluate was cooled in a water-cooled glass reflux condenser before being collected in 50 ml volumetric flasks. The elution was stopped after 900 ml of eluate (6 bed volumes) had been collected. When the elution was carried out at 90 °C, the system was not pressurized.

The elution procedure used in the present investigation included the pretreatment of the gold-loaded carbon in the column with a reagent consisting of 3 to 10 per cent sodium cyanide and 1 per cent sodium hydroxide. This was followed by the elution of the gold using deionized water[2]. For the A and B samples of carbon (carbon-in-pulp operations), the carbon was first washed at a rate of 5 bed volumes per hour with 1 to 5 per cent hydrochloric acid. No gold or silver was found to be eluted from the carbon during this acid pretreatment. Acid-washing was continued until the column effluent was acid (i.e. a pH value of less than 3), after which 2 bed volumes of tap water at the same flowrate was pumped through the column. This removed excess acid, and the pretreatment of the carbon with the cyanide reagent could then be started. Sample C was not acid-washed since no build-up of calcium carbonate occurred under the particular loading conditions used. After the acid-washing (if used), the loaded carbon was treated with 1 bed volume of the cyanide reagent and the gold was then eluted with water. Because samples A and B were of different particle size from sample C, the first two carbons were allowed to 'soak' in the cyanide pretreatment reagent for 30 minutes before the elution, whereas sample C was 'soaked' for only 15 minutes. The technique for eluate collection was also modified during the course of the investigation. The procedure for the elution of sample C involved draining of the carbon bed before the cyanide pretreatment, followed by collection of the total column eluate. As already mentioned, no gold or silver was eluted during the initial acid-washing of the column, and the collection of eluate from samples A and B was begun only after the column effluent had become alkaline. A sharp change in the effluent pH defined the alkaline pretreatment front.

Results

Effect of Temperature on Gold Elution

Figs. 1 and 2 illustrate graphically the effect of temperature on the elution of gold from carbon samples C and B respectively. The elution was carried out at a rate of three bed volumes per hour in both instances. For sample B, the carbon was pre-washed with approximately 2 bed volumes of 2 per cent hydrochloric acid.

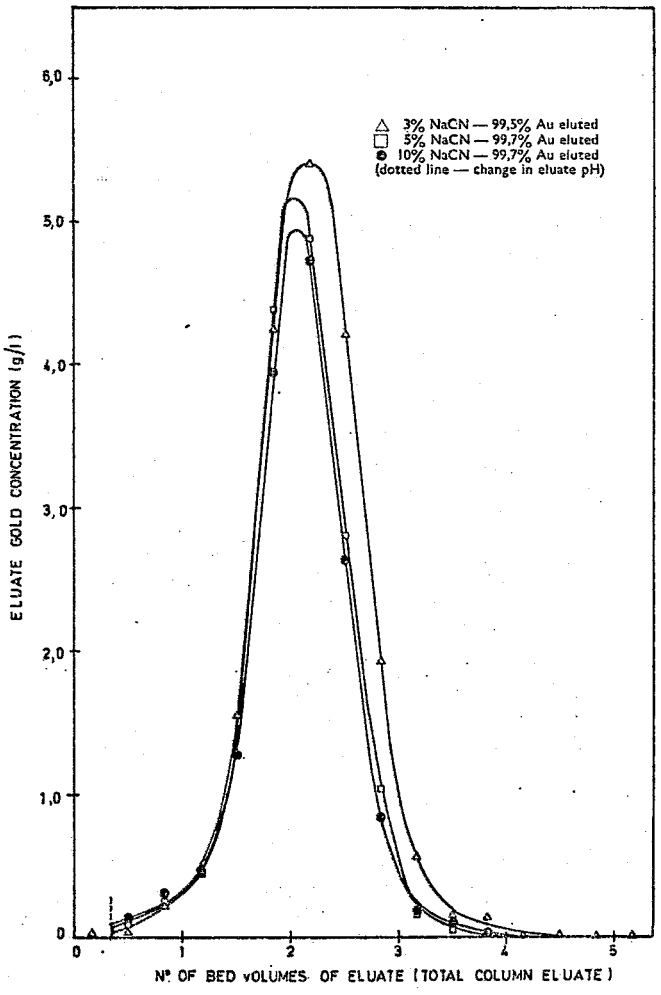

Fig. 3—The effect of cyanide concentration in the pretreatment and elution of carbon sample C at 110 °C

In each of these instances there is a significant beneficial effect if the elution temperature is increased from 90 to 110°C. Thus, the elution at 110°C of 97,1 per cent of the gold from carbon C after 3 bed volumes (99,7 per cent after 6 bed volumes) can be compared with the elution at 90°C of 97,3 per cent of the gold after 4 bed volumes (99,6 per cent after 6 bed volumes). Likewise, in the case of carbon B, the elution of 97,4 per cent gold after 4 bed volumes (99,5 per cent after 6 bed volumes) at 110°C can be compared with an 88,8 per cent gold recovery after 4 bed volumes (94,3 per cent after 6 bed volumes) at 90°C. While the elution results from these two carbon samples are not strictly comparable (owing to the different methods of eluate collection used), the more symmetrical elution curve resulting from the treatment of carbon C (minus 1190 μm plus 840 μm) is noteworthy. It is undoubtedly due to the finer particle size of this carbon when compared with carbon B (minus 2380 μm plus 1000 μm).

In view of the above findings, unless otherwise specified, further elution testwork in the present investigation was confined to the elution of gold at 110°C.

Effect of Cyanide Concentration

Figs. 3 and 4 illustrate the elution of gold at 110°C from samples C and A respectively when sodium cyanide

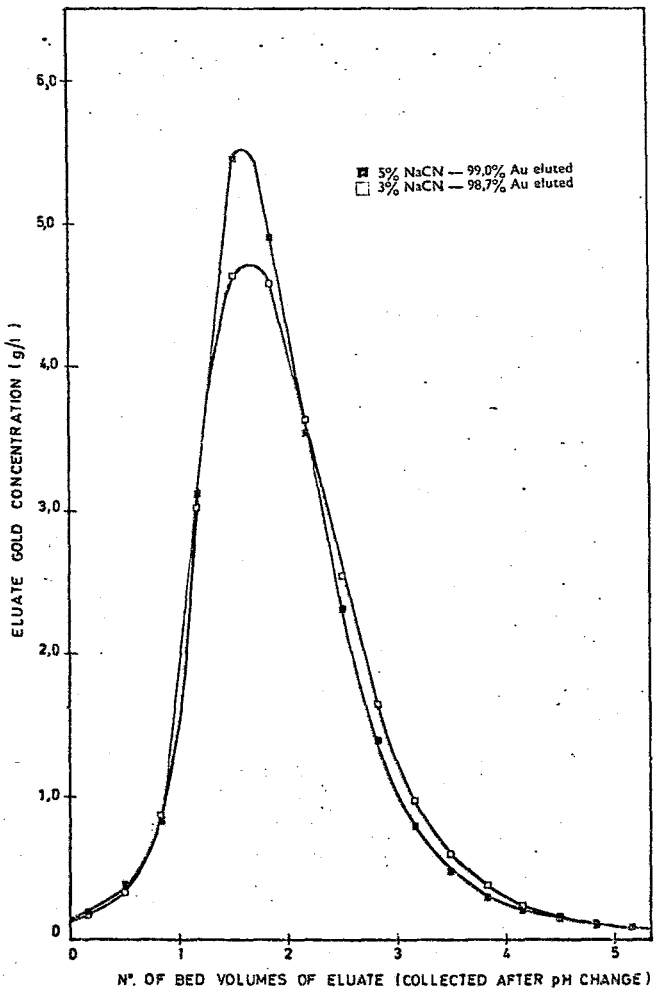

Fig. 4—The effect of cyanide concentration in the pretreatment and elution of carbon sample A at 110°C

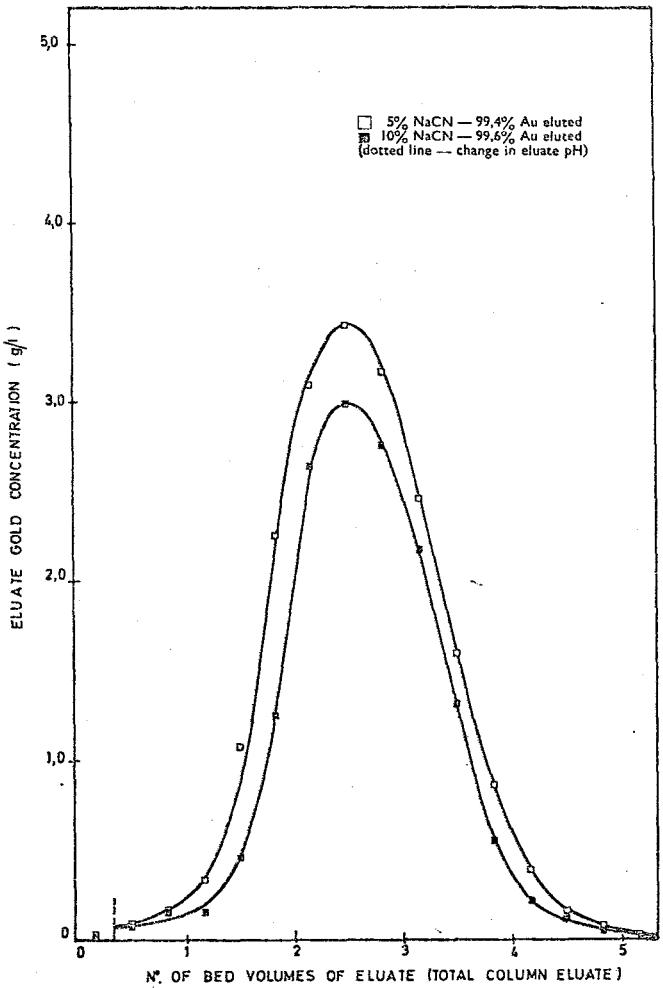

Fig. 5—The effect of cyanide concentration in the pretreatment and elution of carbon sample C at 90°C

of various concentrations was used as a pretreatment reagent. Elution was carried out at a rate of 3 bed volumes per hour in all instances, sample A being prewashed with approximately 1,6 bed volumes of 3 per cent hydrochloric acid before the elution. The results from the elution of sample C (Fig. 3) would suggest that concentrations of sodium cyanide as low as 3 per cent can be used with little or no adverse effects. The results from the elution of sample A indicate a small decrease in elution efficiency from the use of such a low concentration of cyanide. Similar testwork was carried out on sample C at an elution temperature of 90°C (Fig. 5). Here again it can be noted that the use of a sodium cyanide concentration of 5 per cent was equally as effective as the 10 per cent concentration. Unless otherwise specified, further testwork was therefore confined to the use of a pretreatment reagent consisting of 5 per cent sodium cyanide and 1 per cent sodium hydroxide.

Later experiments showed that the volume of the cyanide pretreatment reagent could be reduced from the 1 bed volume previously used to only 0,5 bed volume without seriously affecting the elution efficiency. Fig. 6 illustrates the use of smaller volumes of reagent (5 per cent sodium cyanide and 1 per cent sodium hydroxide) prior to the elution of gold at 110°C from sample C. A

more concentrated eluate was collected when only 0,5 bed volume of this reagent was used.

Effect of Eluate Flowrates on Gold Elution

Figs. 7, 8, and 9 illustrate gold elution at 110°C and various elution flowrates from carbon samples A, B, and C respectively. Samples A and B were pre-washed with acid as previously described. The differences in the elution efficiencies at 110°C are in all cases only marginal. However, the results would suggest that the coarser carbons (samples A and B) are best eluted at a flowrate of 3 bed volumes per hour, while a faster rate of 5 bed volumes per hour can be used for the elution of the more finely sized carbon (sample C). Similar testwork carried out on sample C at 90°C (Fig. 10) would indicate that a flowrate of only 3 bed volumes per hour should be used for the more finely sized carbon when operating at the lower elution temperature.

Effect of Acid-washing

Because of the physical choking of the carbon due to a build-up of calcium carbonate, which is generally observed when conventional plant solutions are treated, the effect of acid washing on the elution of gold from sample A was briefly investigated. Fig. 11 illustrates graphically the considerably improved gold elution at 110°C following acid-washing with 3 per cent hydrochloric acid at 90°C when compared with gold elution carried out on the sample as received. Gold was eluted at a rate of 3 bed volumes per hour in both instances. These findings agree with earlier conclusions[4].

Drying of the Carbon

The effect of air-drying or oven-drying of the loaded activated carbon on the subsequent elution of gold had to date always been rather speculative. Fig. 12 illustrates the effect on the gold elution of the oven-drying of sample A (110°C for 17 hours). This can be compared with the elution of the sample as received. Following acid-washing with 3 per cent hydrochloric acid, gold was eluted at 110°C at a flowrate of 3 bed volumes per hour. The elution results presented suggest that oven-drying, or even air-drying, of loaded carbon prior to elution has little or no effect on the subsequent elution of gold.

Collection of Column Eluate

The draining of the carbon bed following acid-washing and the pretreatment and elution of gold had been previously advocated[1] in order to produce a more concentrated eluate for electrolysis. As it had been established during the course of the present investigation that

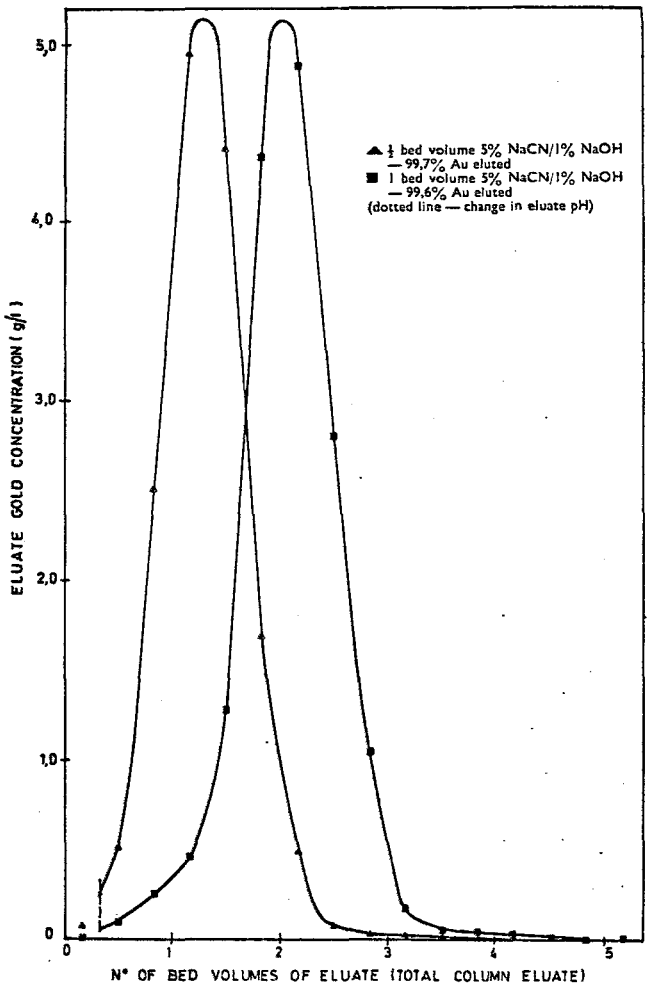

Fig. 6—The use of smaller quantities of pretreatment reagent for the elution of gold from carbon sample C at 110°C

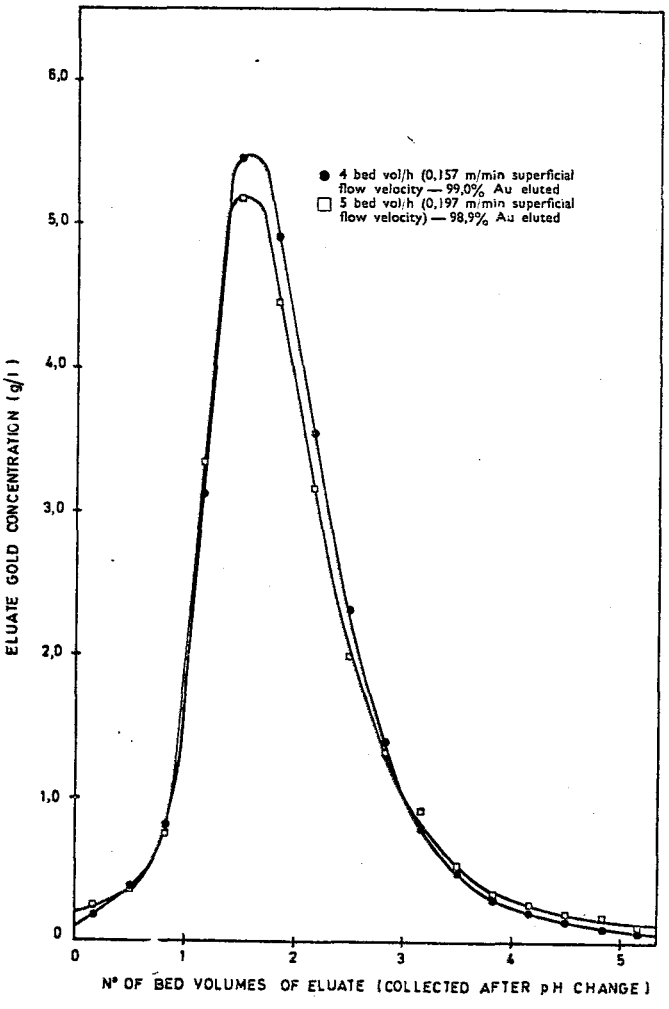

Fig. 7—The effect of eluate flowrates on gold elution from carbon sample A at 110°C

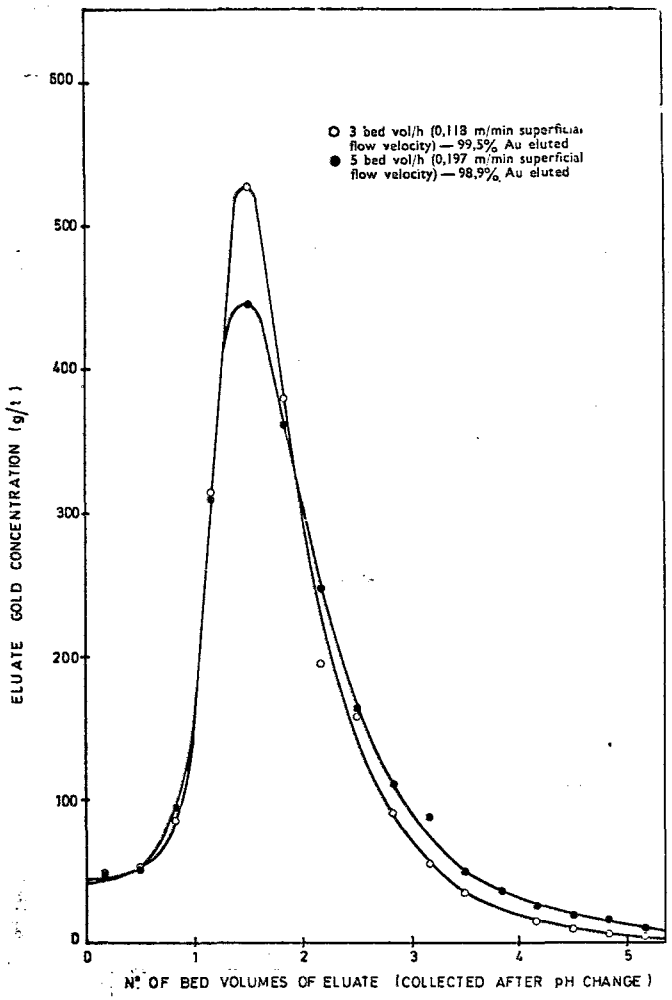

Fig. 8—The effect of eluate flowrates on gold elution from carbon sample B at 110°C

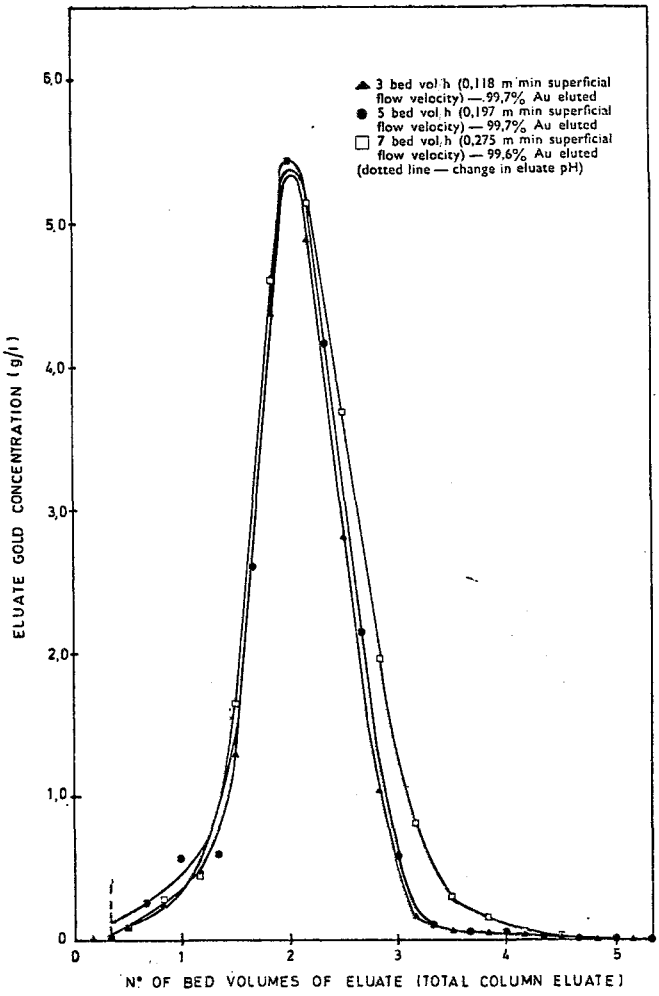

Fig. 9—The effect of eluate flowrates on gold elution from carbon sample C at 110°C

little or no gold was eluted under acid conditions, and that gold commenced eluting only under alkaline conditions, the collection of column eluate following a change in eluate pH afforded a simpler operational procedure. Fig. 13 illustrates the above eluate-collection procedures, indicating little or no change in the resulting gold-elution profiles. Gold was eluted in both instances at a flowrate of 3 bed volumes per hour.

Elution of Silver

In general, the accurate monitoring of the elution of silver in the present investigation was not satisfactory owing to several reasons.

(a) Silver was found to elute ahead of gold when the prescribed method[2] was used, and the resulting solutions were found to be photosensitive, resulting in the precipitation of silver with time. Visible darkening of the silver-rich eluate fractions was always observed. This phenomenon was undoubtedly due to the instability of the silver cyanide complex at the high pH values of the elution fractions taken at the early stages of gold elution. Either the immediate analysis of eluate fractions (which was not always possible), or the addition of 2 ml of 10 per cent sodium cyanide to each fraction (50 ml), was found to overcome the precipitation problem.

(b) The analysis of carbon residues by conventional fire assay does not yield accurate silver values because of parting difficulties associated with the analysis of such relatively gold-rich material.

In the present investigation, both gold and silver were monitored, but it was found that very poor material balances resulted for silver owing in part to the difficulties outlined above. For this reason, together with the very minor economic role played by silver, the silver results are not presented in any great detail. However, the elution results did in general indicate the elution of 80 to 100 per cent of the contained silver from the carbon. As an example, Fig. 14 illustrates the effect of temperature on the elution of silver from sample C at a flowrate of 3 bed volumes per hour.

The Removal of Calcium and Silica

The average calcium and silica contents of various carbon samples treated in the investigation are presented in Table I. It is immediately evident that the build-up of calcium is very considerable in the case of the loaded carbon originating from the carbon-in-pulp operations, namely Fairview and Golden Dumps. In the case of the Fairview carbon, a calcium content of 1,5 per cent and a gold loading of 16 900 g/t can be compared with a 3,4 per cent calcium content for the Golden Dumps

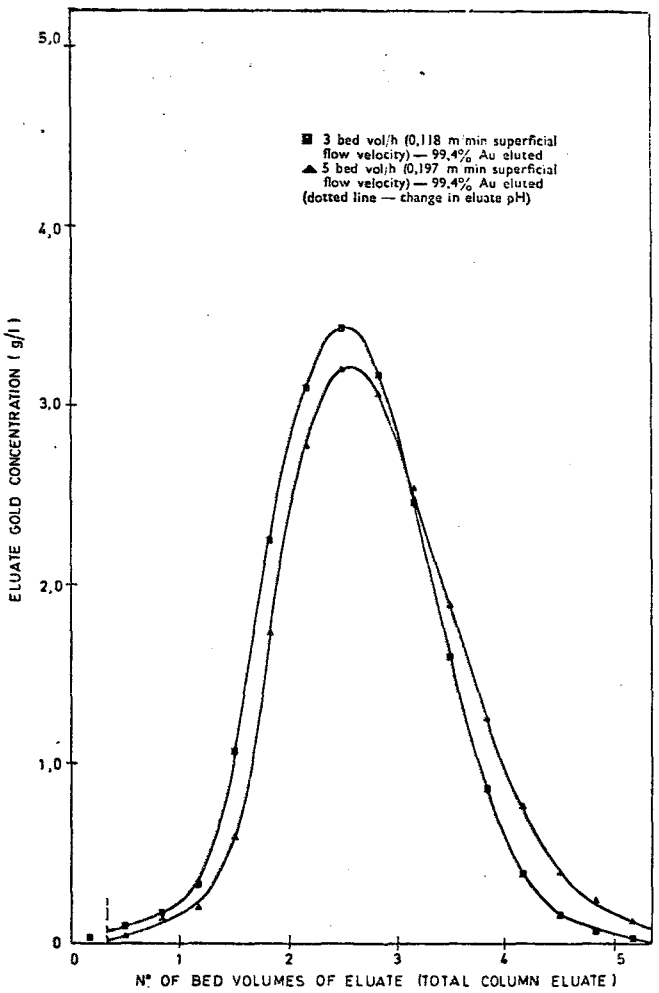

Fig. 10—The effect of eluate flowrates on gold elution from carbon sample C at 90°C

carbon and a gold loading of only 3140 g/t. The use of air agitation in the carbon-contacting circuit at Golden Dumps could to a large degree explain the fouling of the carbon by calcium. Mechanical agitation is used in the carbon contactors of the Fairview Gold Mine.

Fouling of the carbon by calcium can be explained on the basis of the following reactions. Clearly, the presence of carbon dioxide is also needed since calcium occurs on the carbon in the form of calcium carbonate.

(1) $CaO+CO_2 \rightarrow CaCO_3$
(This reaction could take place in the carbon contactors, but not necessarily on the carbon surface. The carbon dioxide is derived from the air, where it is present to the extent of 320 p.p.m.)

(2) $CN^- + \frac{1}{2}O_2 \rightarrow CNO^-$
$CNO^- + 2H_2O \rightarrow NH_4^+ + CO_3^{2-}$
(The above reactions are catalysed by activated carbon[13]. A strong smell of ammonia was observed in the vicinity of the carbon contactors at Golden Dumps (Pty) Ltd, suggesting that reactions such as those given may occur in practice.)

(3) $Au(CN)_2^- \underset{\uparrow}{\downarrow} Au^+ + 2CN^-$ + virgin carbon + $H_2O + \frac{1}{2}O_2 \rightarrow$ loaded carbon + $HCO_3^- + OH^-$
(This is the reaction mechanism proposed for the adsorption of gold by carbon[14].)

TABLE I

THE REMOVAL OF CALCIUM AND SILICA FROM VARIOUS LOADED SAMPLES OF ACTIVATED CARBON

Nature of carbon (as received/eluted)	Composition of carbon, g/t			
	Ca	Si	Au	Ag
Sample A as received (ex Fairview)	14 800	1 160	16 900	300
Eluted sample A (average values from 5 different elution tests)	1 794	901	194	47
Sample B as received (ex Golden Dumps after 1st week of operation)	13 900	2 670	1 604	182
Eluted Sample B (average values from 4 different elution tests)	910	1 348	88	41
Sample B1 as received (ex Golden Dumps after 2nd week of operation)	35 250	—	3 140	365
Eluted sample B1 (only 1 elution test)	965	1 750	223	66
Sample C as received (carbon loaded from acidified gold-plant solution)	870	860	18 850	1 750
Eluted sample C (average values from 5 different elution tests)	640	263	97	44

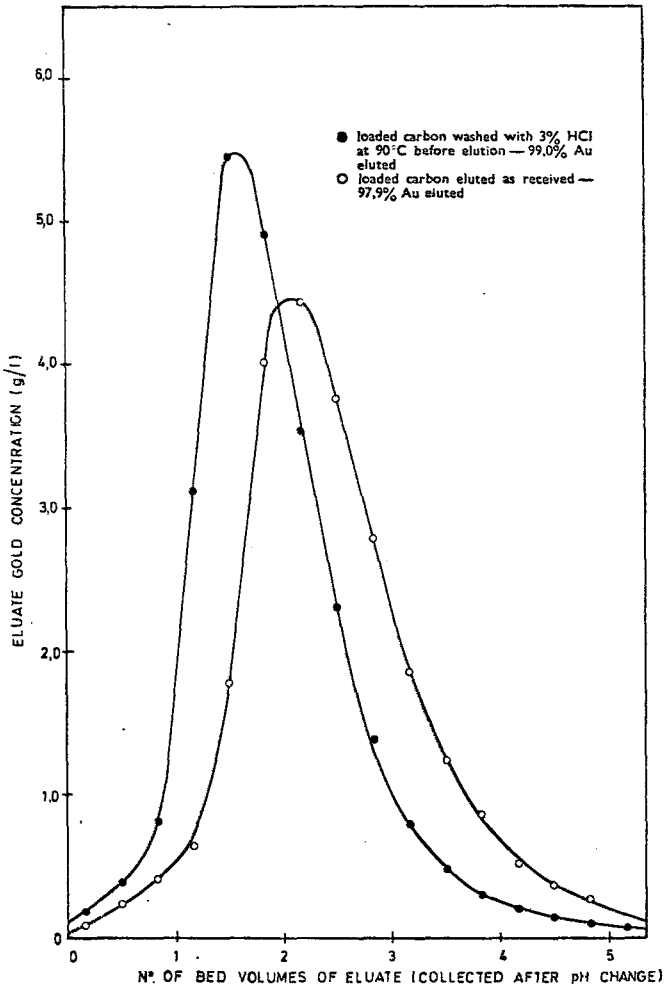

Fig. 11—The effect of acid-washing on gold elution from carbon sample A at 110°C

No poisoning by calcium carbonate was observed in the carbon loaded from acidified pregnant plant solutions. Here a calcium content of only 0,09 per cent was noted (the calcium content of the virgin coconut-shell carbon was 0,1 per cent). While the build-up of silica on the loaded carbon samples is not considered to be excessive, there was nevertheless a substantial increase. Thus, the presence of up to 0,27 per cent silica in Golden Dumps sample B can be compared with 0,09 per cent silica found to be present on sample C (the silica content of the virgin carbon was 0,1 per cent).

The results indicate that the removal of calcium and silica following acid-washing and elution of carbon samples A and B is effective in removing virtually all the calcium and silica. In the case of sample C, where no acid washing was considered necessary, much lower levels of calcium and silica were found to be present following gold elution (0,06 per cent calcium and 0,03 per cent silica).

It should be noted that prussic acid is evolved during acid-washing, and due precautions must be taken.

Discussion

The most significant factors affecting the elution of gold from activated carbon are temperature and acid-washing. The other gold-elution parameters studied have only marginal effects on the efficiency of gold elution. Accordingly, elution parameters other than elution temperature and acid-washing can be modified if convenient to do so without adverse effect on the elution procedure. These findings have a direct bearing on the current operations at the carbon recovery plants of both the Fairview Gold Mine[7] and Golden Dumps (Pty) Ltd[8].

As concluded in the present investigation, as well as in earlier elution studies[2], temperature is perhaps the most important operational variable in gold elution. Consequently, the elution of gold from carbon at 100 to 110°C (gauge pressure of 50 to 100 kPa) is strongly suggested. Problems associated with possible boiling in the elution column when the process is operated close to the boiling point of water at ambient pressures would support pressurization of the elution column. This problem of boiling was encountered during pilot-plant elution of carbon loaded at the Fairview Gold Mine[7]. Operationally, elution at such low gauge pressure is facilitated simply by the addition of a suitable pressure-relief valve positioned on the eluate line from the elution column. From the safety aspect, any pressurized system must, of course, comply with the provisions of the Factories Act.

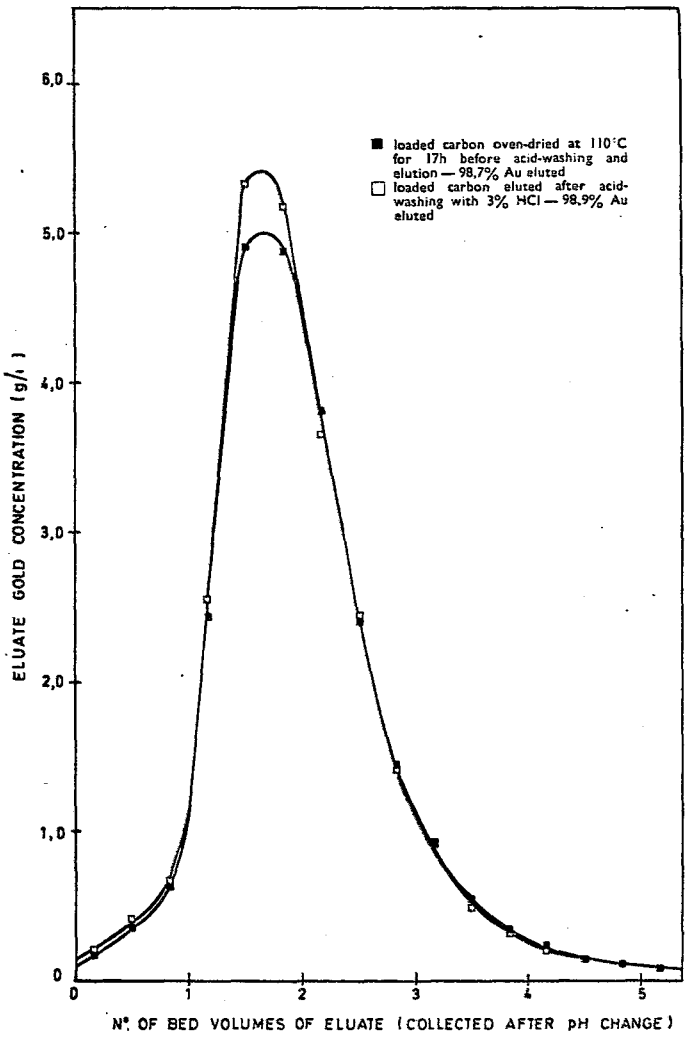

Fig. 12—The effect of oven drying of carbon sample A on gold elution at 110°C

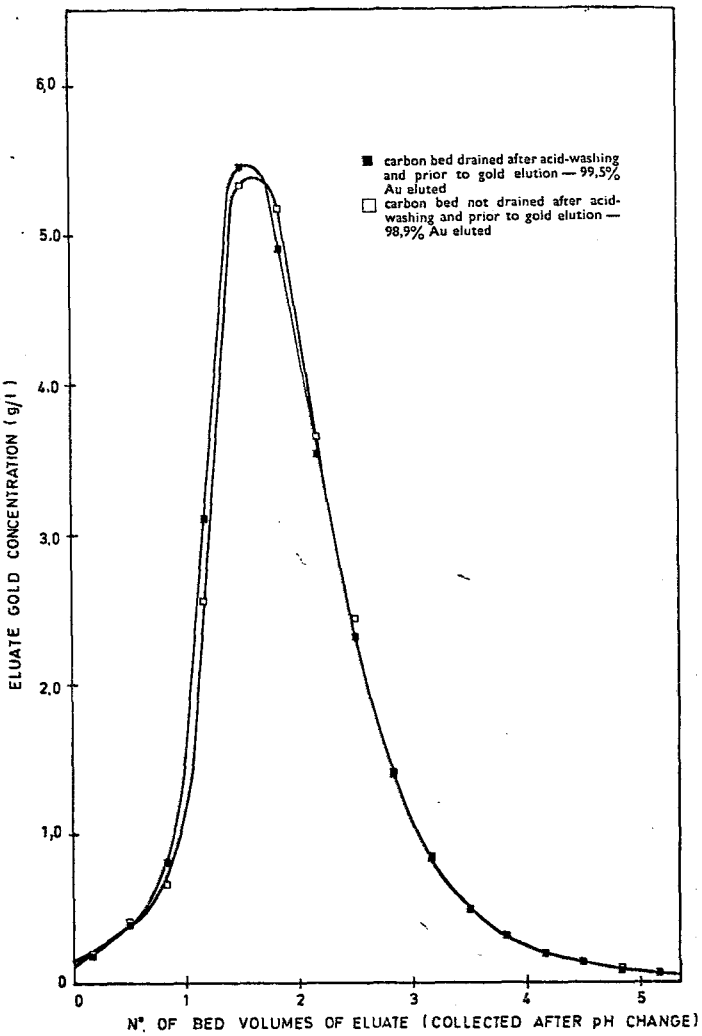

Fig. 13—Two different eluate-collection procedures used in the elution of gold from carbon sample A at 110°C

Several advantages are gained from operation at the elution temperatures of 100 to 110°C. Not only can elution be carried out in shorter times (2 to 3 h) owing to the use of higher elution flowrates, but more-concentrated gold eluates are produced for the subsequent electrowinning of gold. The use of shorter 'soaking periods' with the sodium cyanide–sodium hydroxide reagent and lower additions of pretreatment cyanide can also be expected. Of course, the particle size of the loaded carbon is to some degree also significant, and must be taken into account in the selection of operating parameters. As can be expected, the choice of a carbon of finer particle size is advocated from an elution point of view. A further operational modification suggested as a result of the findings of this investigation is that only the alkaline column eluate should be collected so that a more concentrated eluate suitable for subsequent electrolysis can be produced.

With regard to the elution response of the three different carbon samples studied in the present investigation, few problems were encountered. While minor differences in the elution profiles of the respective carbons were obtained following acid-washing if required, such differences were small and undoubtedly related to the particle size of the sample concerned. When this consideration is taken into account, the results presented indicate that the prescribed method for the elution of gold from activated carbon is both reproducible and efficient.

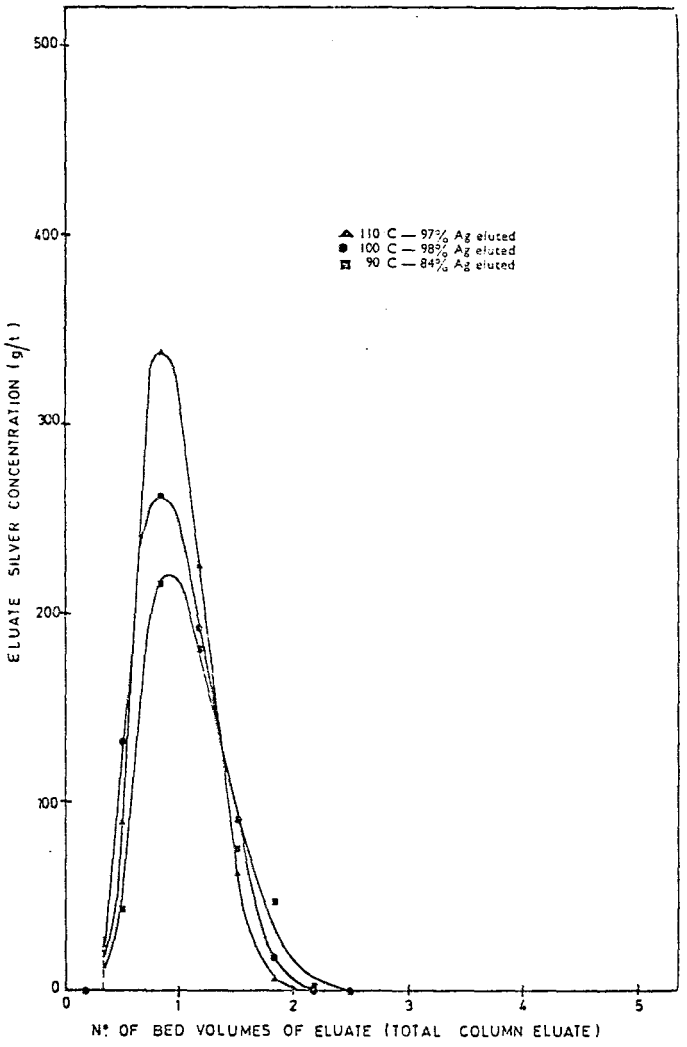

Fig. 14—The effect of temperature on the elution of silver from carbon sample C

Acknowledgements

The authors are indebted to the Analytical Laboratory of Anglo American Research Laboratories for the chemical analyses, and to the management of Anglo American Research Laboratories for permission to publish these findings.

The authors also thank both the National Institute for Metallurgy and the sponsors of the Chamber of Mines' projects involved in the Fairview Gold Mine and Golden Dumps carbon operations for their kind co-operation. Thanks are also due to the management of the Fairview Gold Mine and to Golden Dumps (Pty) Ltd for the supply of the loaded-carbon samples used in the present study.

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A REVIEW OF THE AARL PROCESS FOR THE ELUTION OF GOLD FROM ACTIVATED CARBON

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by

R.J. Davidson
Senior Research Fellow

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SUMMARY

Parameters affecting gold elution using the AARL process are reviewed and critically analyzed in an attempt to optimise metallurgical efficiency and minimise costs in both current and future carbon adsorption operations. Such factors as acid washing, column geometry, elution temperatures and pretreatment reagents are discussed together with eluant water quality and eluate flowrates.

Hot acid washing of carbon prior to gold elution has many benefits. Besides improving elution efficiency when eluting with regional water, carbon pretreatment and elution may be initiated at or near operational temperatures. Laboratory data indicated that hot acid washing reduces the mass loss of carbon during thermal regeneration. The eluate going to either zinc precipitation or electrowinning is also relatively free of troublesome impurities (organic contaminants and ferrocyanides in particular).

Because of the chromatographic character of the AARL elution where an ionic strength gradient is used to inhibit the readsorption of gold, column geometry plays a significant role. While practical considerations and capital costs may restrain the length of the elution column, high length/diameter ratios are advocated in order to promote the plug flow of column eluate. The effect of column geometry on eluant requirements is presented below.

Elution temperature (^{o}C)	No. of bed volumes required		
	L/D = 3	L/D = 6	L/D = 12
90	19	14	12
100	14	11	9
110	10	7	6
120	8	6	5
130	7	6	5
140	6	5	4

From/...

From both a metallurgical and costing viewpoint, the two significant factors affecting gold elution using the process under review are temperature and reagent cyanide additions. While high temperatures and high reagent additions within prescribed limits will invariably improve metallurgical efficiencies, a balance between energy and reagent costs is required in order to minimise overall elution costs. Such factors will of course become more critical when eluting carbon containing low gold loadings.

As pretreatment reagents constitute a considerable portion of the total elution costs, both the addition and consumption of sodium cyanide must be carefully considered. Although lower cyanide additions may be effective in certain instances, the more practical use of 0,6 bed volume of a 2-3 per cent NaCN reagent containing 2 per cent NaOH is advocated. This is equivalent to an addition of 20-30 kg NaCN and 20 kg NaOH per ton carbon treated. The effects of cyanide additions and temperature on total elution costs are presented as follows:

Elution temperature (°C)	Total elution cost (R/t carbon)		
	10 kg/t NaCN addition	20 kg/t NaCN addition	30 kg/t NaCN addition
100	136,04	129,09	141,84
110	117,91	115,12	128,92
120	115,29	123,31	134,91
130	134,17	122,31	138,06

The most cost effective conditions indicated an elution temperature of 110°C using 20 kg/t NaCN and 20 kg/t NaOH as a pretreatment reagent. Under these conditions the following cost structure may be approximated:

TABLE/...

	Operating Cost (R/t carbon)	Total Cost (%)
Reagent acid	17,35	15,1
Column preheating	25,17	21,9
Pretreatment reagents	42,60	37,0
Eluant heating	30,00	26,0
	115,12	100,0

When related to gold production, elution costs become most sensitive to carbon loadings. When eluting a carbon containing 5 kg/t gold, the cost of elution may be as little as R20/kg gold produced, whereas on eluting a carbon containing only 0,5 kg/t gold such costs may exceed R200/kg gold.

Optimum silver elution is obtained using low cyanide reagent additions (10 kg/t NaCN) and high elution temperatures (130^{o}C).

1.0 INTRODUCTION

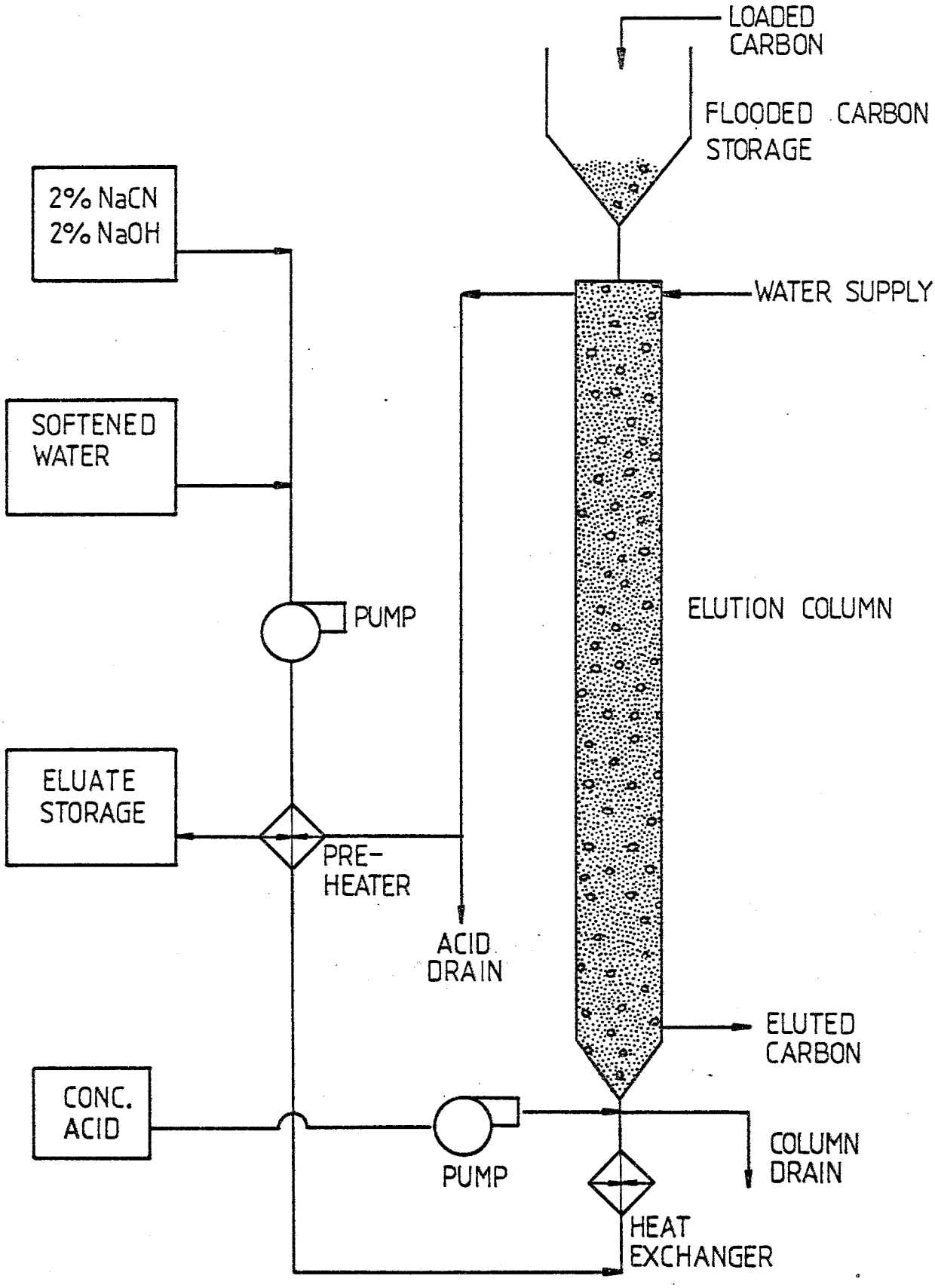
Numerous developments and modifications to the AARL process (S.A. Patent Nos. 73/8939 and 76/4202) for the elution of gold from activated carbon have taken place since the earlier publications in 1974-79. Following successful piloting, first at the Vaal Reefs Mining and Exploration Company (1975) and then at the Fairview Gold Mine (1978), the process was first used on a large scale at the President Brand Gold Mine (1980) and is now in common use in South Africa. In the light of such rapid development together with the on-going investigations carried out at the Anglo American Research Laboratories (AARL) it is perhaps necessary to review and possibly emphasize such related findings which have particular practical significance.

In a plant situation, the selective adsorption of gold onto activated carbon is thought to relate to the solubility of the ion paired adsorbed species $Ca^{2+}[Au(CN)_2^-]_2$. Hence, conditions favouring gold adsorption include low temperatures, low cyanide concentrations and low alkalinity together with the adsorption medium displaying a high ionic strength and containing Ca^{2+} ions. A simple reversal of such conditions may be expected to apply to the elution of gold from activated carbon. As conditions favouring high cyanide concentrations and low ionic strength levels are not compatable, the pretreatment of carbon at high temperature with a strong sodium cyanide reagent followed by elution with a hot high quality water is used to effect the rapid and efficient elution of gold. This staged process in which gold is first solubilised and then eluted with water is the significant difference between the AARL process and the classical Zadra procedure in which a cyanide reagent is circulated in a closed circuit containing an electrowinning cell.

The AARL elution is essentially a chromatographic mechanism whereby the metal cyanocomplexes are sequentially desorbed from the pretreated carbon using an ionic strength gradient as a means of effecting elution. Because of this mechanism, both the plug flow

FIGURE 1

AARL GOLD ELUTION FLOWSHEET



of eluate through the system and the geometry of the column play an important role. Furthermore, while temperature plays perhaps the most significant role in gold elution, the parameters affecting elution are confounded to some degree thus making their evaluation and definition somewhat complex.

The present review will critically evaluate the AARL elution procedure in an attempt to optimise metallurgical efficiency and minimise costs in both current and future carbon adsorption operations. Pertinent subject matter is indexed so as to afford information retrieval.

2.0 AARL ELUTION PROCESS

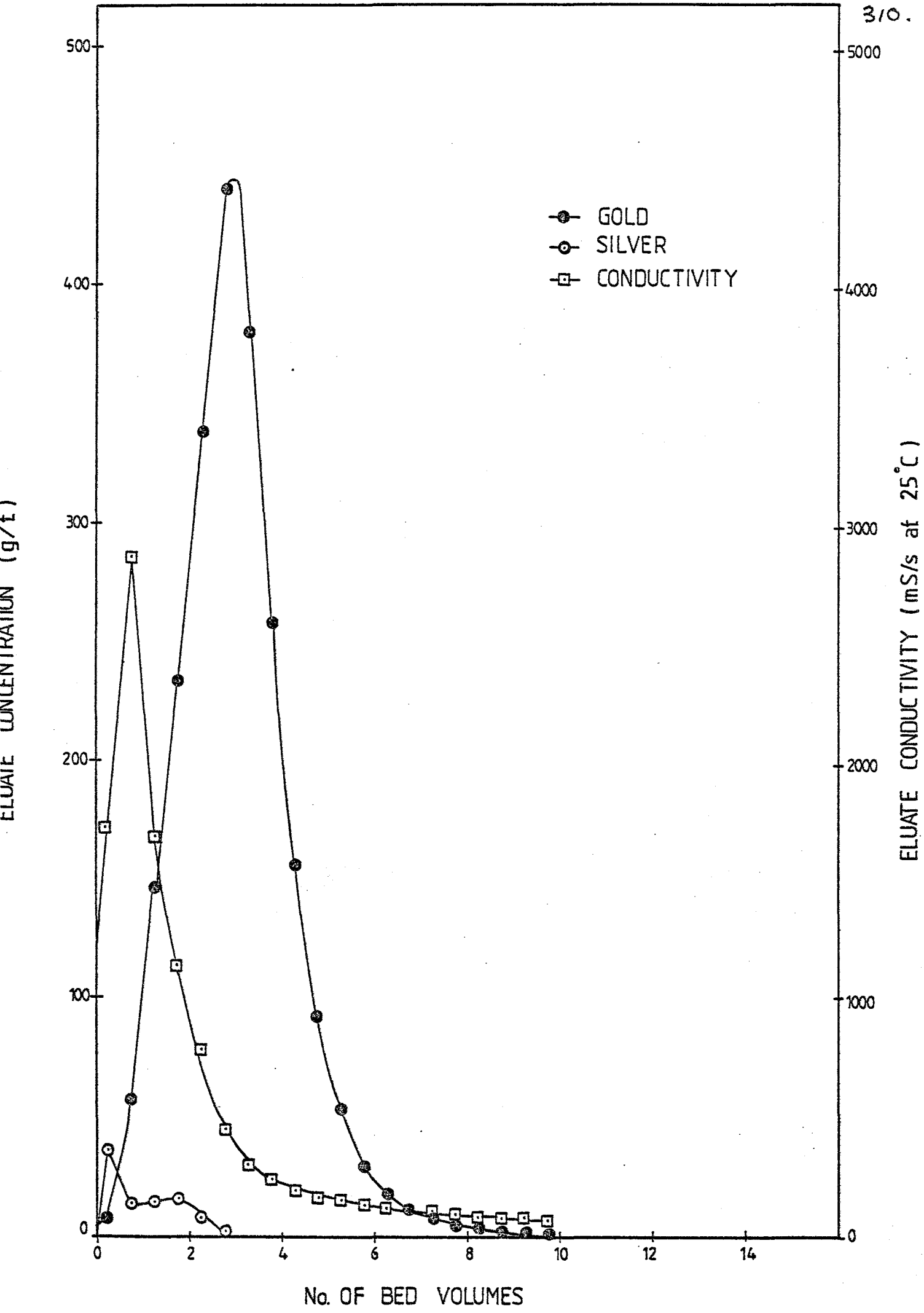
2.1 Flowsheet

Figure 1 represents a simplified flowsheet of the elution circuit with the elution column having a length/diameter ratio of greater than 6.

Loaded carbon preferably free of pulp, wood fibre and plastic material is first loaded into the elution column via gravity from the flooded carbon storage hopper (if carbon eduction is required, loaded carbon is best transferred from a blowcase in order to minimise attritional losses - transfer velocities in pipelines should not exceed 2 m/s). Pulp slime, wood fibre and plastic may be effectively removed from the carbon using a separate elutriating vessel operating with an upflow velocity of ca 1m/min. Plastic material in particular may soften at elution temperatures and choke column collectors, while wood fibre may well activate in the thermal regeneration stage. The presence of excess pulp will also result in excessive caustic consumption due to sodium silicate formation at elution temperatures (see below).

Following/...

FIG. 2 TYPICAL GOLD, SILVER AND CONDUCTIVITY PROFILES USING AARL ELUTION



Following carbon loading into the elution column in which the column is completely filled, the carbon is first acid washed with ca 1,0 bed volume of 3 vol. per cent hydrochloric acid followed by 1-2 bed volumes of water (the bed volume relates to the volume of the empty column). During this period the system is raised to the required elution temperature (110-120°C) and pressure (200-300 kPa gauge) while discharging the acid effluent to the adsorption circuit. The acid washed carbon is then pretreated with ca 0,6 bed volume of a strong caustic cyanide reagent (2-3% NaCN + 2% NaOH) and the carbon allowed to 'soak' in this reagent for ca 30 minutes (0,6 bed volume of reagent is used so as to ensure the complete displacement of wash water from the elution column).

Gold is then eluted from the carbon at a flowrate equivalent to 2 bed volumes/h using 5-6 bed volumes of a high quality softened water preferably containing less than 300 g/t sodium. Figure 2 illustrates a typical gold and silver elution profile together with a conductivity profile showing the dramatic change in the ionic strength of the eluate. Each batch elution may be carried out in an 8 hour operating shift, with the sequencing of valves conveniently automated using a time control unit.

Column eluate, which is only taken following the sharp change in pH effected by the pretreatment reagents, is then subjected to either electrowinning or zinc cementation for the eventual recovery and smelting of gold and silver values.

2.2 Acid Washing

The hot acid washing of carbon with dilute hydrochloric acid prior to gold elution using the AARL procedure is most preferable if overall efficiency in a particular plant is to be maintained. Beside activated carbon's ability to adsorb a substantial amount of acid, the requirements for acid washing will also

depend on the calcium carbonate content of the carbon (see below). In most instances the use of approximately 1 bed volume of 3 vol. per cent hydrochloric acid (31-33%) has been found to effectively remove calcium from the loaded carbon. While cold dilute acid will effectively remove calcium and zinc from carbon, hot acid (ca 90°C) will remove calcium, zinc and nickel together with a large proportion of contained iron and silica. Gold, silver and copper are not eluted from carbon using either hot or cold acid. Acid washing of used carbon is also a form of chemical regeneration commonly used in water treatment circuits. In the case of gold loaded carbons, cold acid does not effectively remove organic contaminants. On the other hand, hot acid treatment has been shown to remove a substantial proportion of organic poisons from the carbon and may well deminish the necessity to thermally regenerate used carbon.

Note: The acid treatment of loaded carbon results in the considerable evolution of hydrocyanic acid (prussic acid) which is extremely toxic and due precautions must be taken. Also, acid washing with nitric acid is not advisable as partial deactivation of the carbon and even explosive reaction may take place.

Aspects which have a bearing on both the need for as well as the sequencing of the acid wash and the effect on the each unit process will be discussed.

2.21 Gold adsorption. Poor gold adsorption may be encountered due to the passivation of the carbon surface with calcium carbonate. The origin of the calcium carbonate is due to the catalytic oxidation of free cyanide in the contacting circuit with the formation of ammonia and carbon dioxide (equation 1).

$$CN^- + \tfrac{1}{2} O_2 \xrightleftharpoons{\text{activated carbon}} CNO^- \quad \ldots \text{equ}^n\ 1$$

$$CNO^- + 2H_2O \xrightleftharpoons{\text{hydrolysis}} NH_3 + HCO_3^-$$

Owing to the solubility differences of calcium carbonate and calcium bicarbonate, this carbon passivation is most sensitive to the pH of the pulp. While calcium loadings below 20 kg/t carbon may not have a serious effect on the kinetic response of carbon in the adsorption circuit, loadings of 40-60 kg/t have a detrimental effect on gold adsorption.

The reaction of calcium carbonate with dilute hydrochloric acid is a simple acid base reaction (equation 2) evolving carbon dioxide. The reaction is most favourable even at ambient temperature and is virtually complete at acid levels below pH 2.

$$CaCO_3 + HCl \rightleftharpoons \underset{\text{(soluble)}}{CaCl_2} + CO_2 \qquad \text{....equ}^n\ 2$$

2.22 Gold elution. When related to elution in particular, the hot acid washing of carbon prior to elution serves several purposes.

a) Under acid conditions the elution system can be brought to operational temperatures using preheated water while not eluting any gold or silver values from the carbon. This preheating of the elution circuit to the required operating temperature ensures the subsequent rapid and efficient elution of gold in a well defined elution profile (Figure 2).

b) In the case of severe calcium passivation acid washing has a very beneficial effect on subsequent gold elution.

c) When excessive carbon loadings of nickel, iron, zinc and silica are encountered as in some effluent circuits operating at reduced pH levels, the acid washing of the carbon will reduce the reagent

FIG. 3

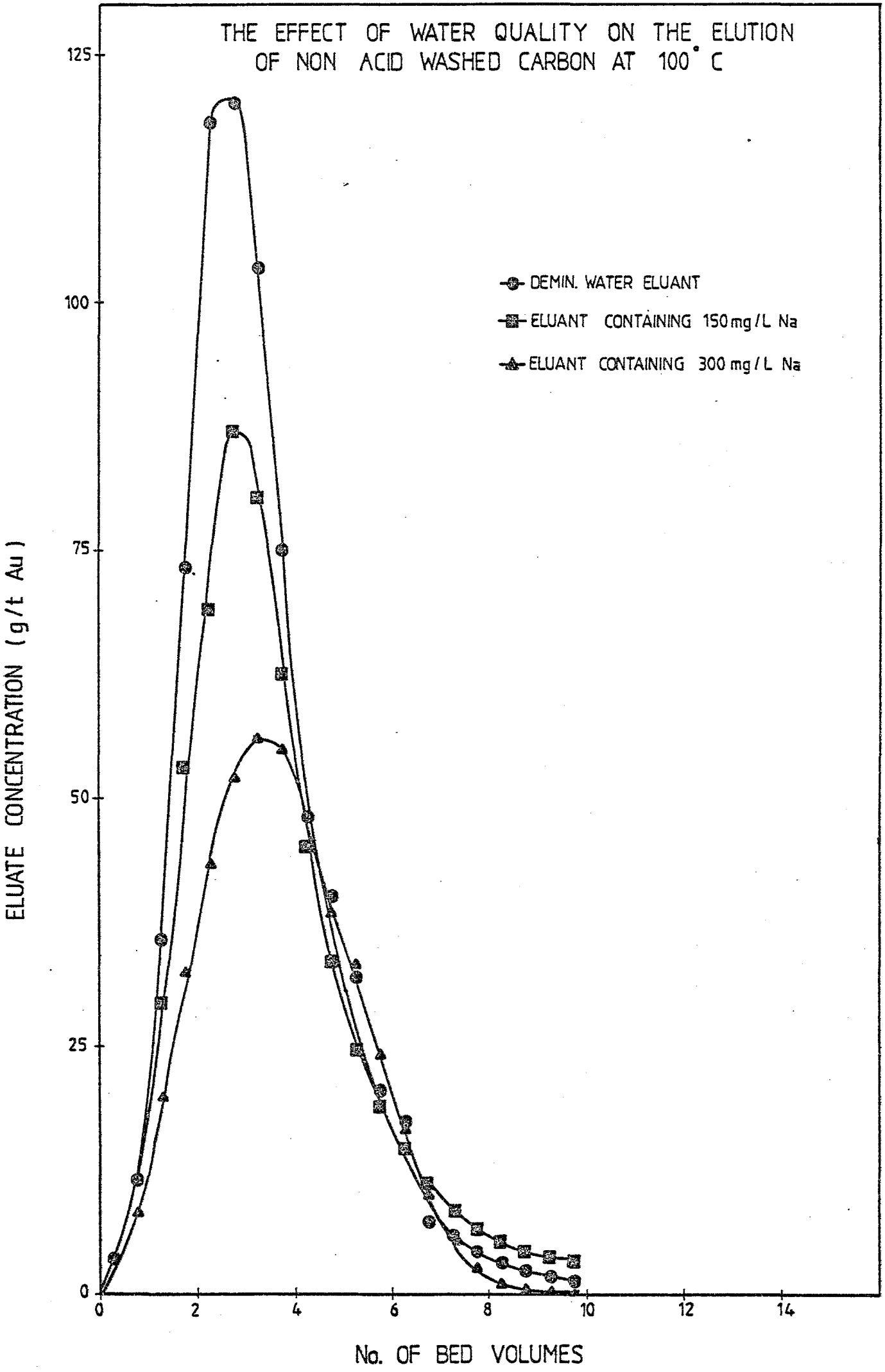
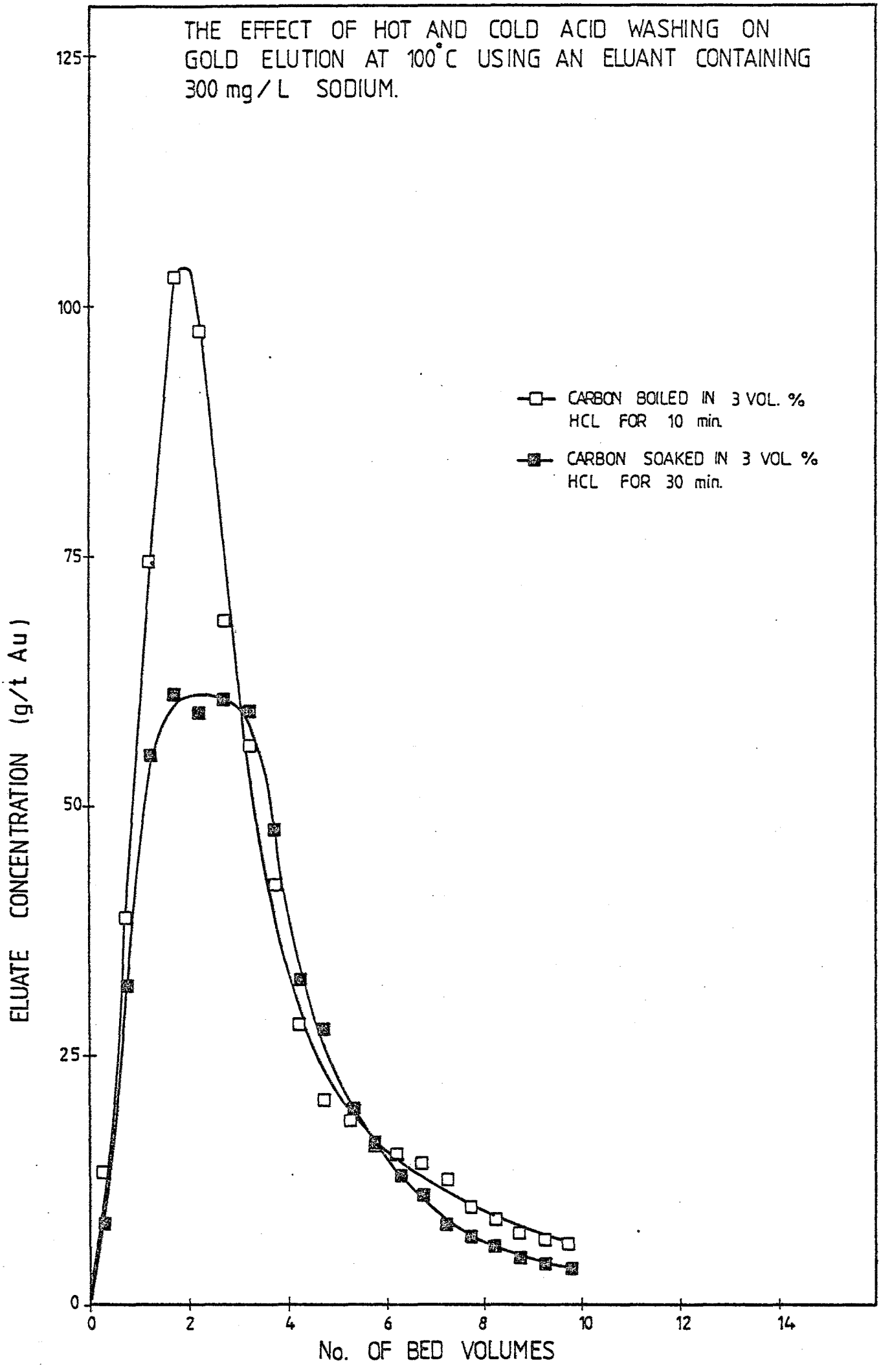

THE EFFECT OF WATER QUALITY ON THE ELUTION OF NON ACID WASHED CARBON AT 100° C
DEMIN. WATER ELUANT
ELUANT CONTAINING 150 mg/L Na
ELUANT CONTAINING 300 mg/L Na
ELUATE CONCENTRATION (g/t Au)
No. OF BED VOLUMES
0
25
50
75
100
125
0
2
4
6
8
10
12
14

FIGURE 4


THE EFFECT OF HOT AND COLD ACID WASHING ON GOLD ELUTION AT 100°C USING AN ELUANT CONTAINING 300 mg/L SODIUM.
CARBON BOILED IN 3 VOL. % HCL FOR 10 min.
CARBON SOAKED IN 3 VOL % HCL FOR 30 min.
ELUATE CONCENTRATION (g/t Au)
No. OF BED VOLUMES
0
25
50
75
100
125
0
2
4
6
8
10
12
14

requirements needed to effect efficient elution. The possible selective removal of these elements may also benefit both the adsorption and eluate treatment circuits (2.24).

d) In adsorption circuits where organic fouling such as float reagents are excessive hot acid washing may remove a large proportion of such contaminants which may be discarded rather than returned to the adsorption circuit. These organic contaminants may also give problems in the eluate treatment circuit (2.24).

e) Possibly the most significant effect of acid washing prior to gold elution relates to the quality of water used for gold elution. In the case of eluting gold from non acid washed carbon, it has been adequately demonstrated that elution efficiency is most sensitive to water quality. This effect is well illustrated in Figure 3 where a significant ionic strength effect may be noted. Rather surprisingly, however, the hot acid washing of carbon produces a carbon which is not as sensitive to water quality, thus allowing the efficient elution of gold using good quality regional waters. A plausible explanation for this effect may relate to the character of the adsorbed gold species following hot acid washing (see mechanism below).

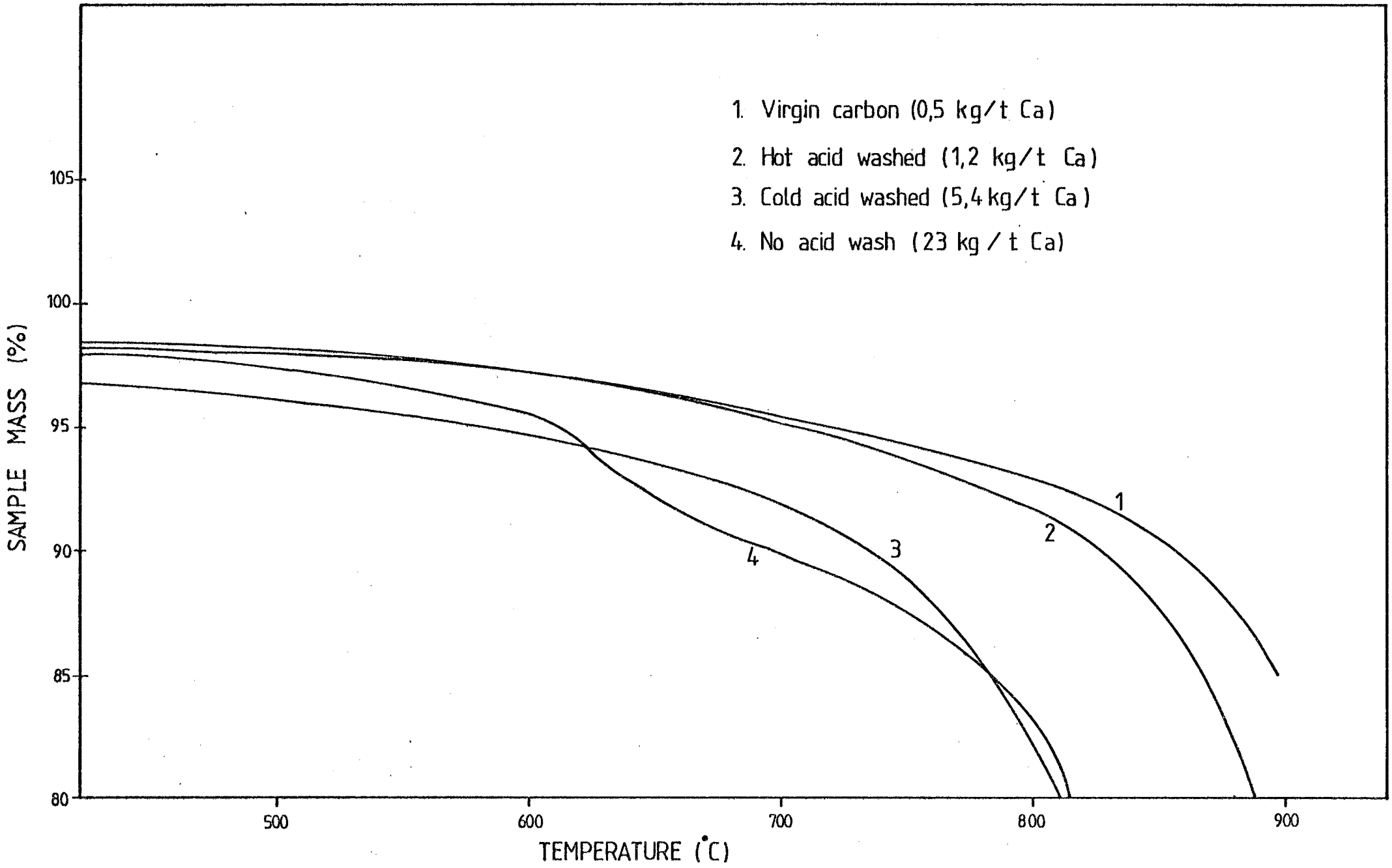
$$\underset{\text{(strongly adsorbed)}}{Ca\left[Au(CN)_2\right]_2} \xrightarrow{\text{hot acid}} \underset{\text{(very strongly adsorbed)}}{AuCN} \xrightarrow{\text{pretreatment}} \underset{\text{(moderately adsorbed)}}{NaAu(CN)_2}$$

As the ion pair $Ca\left[Au(CN)_2\right]_2$ is more strongly adsorbed onto carbon than is the $Na\,Au(CN)_2$ species, it may be expected to be more sensitive to such small changes in the ionic strength (water quality) of the water eluant.

Cold/...

FIG. 5

THERMOGRAVIMETRIC ANALYSIS OF ELUTED CARBON SAMPLES ORIGINATING FROM A CARBON-IN-PULP PLANT



Cold acid washing has been shown to be ineffective under the prescribed elution conditions. Comparative elution profiles using hot and cold acid are shown in Figure 4 using a synthetic regional water eluate containing 300 g/t sodium added as NaCl. A significantly different elution profile following cold acid washing may be noted (S.A. Patent No. 84/2020 specifies the use of hot acid for the formation of the aurocyanide compound AuCN). While hot acid washing resulted in an elution recovery of 92 per cent, cold acid washing only effected a recovery of 73 per cent (eluting at 100°C using 10 bed volumes).

f) Little or no elution of calcium is effected on eluting a non acid washed carbon.

2.23 Thermal regeneration. Acid washing prior to the thermal regeneration of carbon may also serve several important functions.

a) As calcium, magnesium and iron are well known catalytic oxidants for the water gas reaction, acid washing prior to thermal regeneration will thus minimise carbon losses and at the same time maintain physical carbon hardness. Thermogravimetric evidence supporting this conclusion is well illustrated in Figure 5, where a most significant effect may be noted following the hot acid washing of an eluted carbon not previously acid washed. The possible use of higher reaction temperatures and longer residence times, resulting in more efficient carbon regeneration will also apply following acid washing.

b) The partial removal of silica effected by hot acid washing followed by gold elution will prevent to some degree, the formation of refractory crystalline or sintered silicates on the carbon surface. Severe

silicate fouling of carbon has occurred on several CIP operations where only the treatment of the fouled carbon with hydrofluoric acid was shown to restore carbon activity.

c) Acid washing of carbon is a recognised form of chemical regeneration which may well deminish the frequency of thermal regeneration in many instances. Carefully monitored carbon in pulp pilot operations at AARL indicated that non-regenerated carbon may be re-used at least 5-6 times with little fall-off in carbon activity being observed when using a hot acid wash ($93^{o}C$) followed by the AARL elution.

d) Acid washing of carbon following gold elution is not advisable as severe chloride corrosion in the furnace may be encountered.

2.24 Eluate treatment. As ferrocyanides in particular may well present problems in both the electrowinning of gold and in zinc cementation, the selective removal of iron from carbon using a hot acid wash will produce a gold eluate relatively free of iron (and zinc and nickel). Although not well documented, float reagents and organic contaminants may also present problems in electrowinning or zinc cementation operations.

2.25 Seperate acid washing facility. Due to corrosion related problems, carbon may also be hot acid washed and then neutralised in a separate facility (usually made from a fibre glass/polyester laminate) ahead of actual gold elution. This modification to the recommended procedure (2.1) may not be cost effective.

a) The transport of carbon into and from such a facility, particularly if highly loaded, will always incur an additional gold loss due to carbon attrition.

b) The additional cost of such an acid washing facility may be greater than the cost of lining the elution column with a non corrosive material.

c) The cooling of carbon following a separate hot acid wash will incur additional energy costs.

d) Gold elution will be adversely affected as some preheating of the carbon up to the required temperature will be necessary.

2.3 Column Geometry

In the absence of acid washing, copper, nickel, zinc, silver and then gold are eluted sequentially from loaded carbon using the AARL procedure (2.1). Silver and gold are eluted as well resolved profiles strongly supporting the chromatographic mechanism of the process (Figure 2). The resolution of the gold elution profile may thus be expected to be most dependent on the length/diameter ratio of the elution column, which in turn may have a significant effect on the energy requirements for efficient gold elution (3.0).

Elution data from earlier investigations at AARL were collated in an effort to quantify the effects of column geometry and elution temperature on elution energy requirements. This data expressed in terms of the number of bed volumes required to elute at least 95 per cent gold from a loaded carbon containing more than 1 kg/t gold is presented in Table I.

TABLE I/...

TABLE 1

THE EFFECT OF COLUMN GEOMETRY ON GOLD ELUTION

Elution temperature (°C)	No. of bed volumes required L/D = 3	L/D = 6	L/D = 12
90	19	14	12*
100	14	11	9*
110	10	7	6*
120	8*	6*	5*
130	7	6	5*
140	6	5	4*

*Actual laboratory data obtained on eluting at 2 bed vol./h (the remaining data are approximations made for costing purposes - 3,0).

The significant effect of column geometry on the elution effiency is most evident suggesting that particular attention be given to the design of column distributors so as to ensure the plug flow of column influent.

While the practical restraints on the geometry of an elution column are perhaps difficult to predict, the laboratory results presented would indicate the use of high length/diameter ratios if energy requirements are to be minimised (3.0).

2.4 Elution Temperatures

The effects of temperature, ionic strength gradient and cyanide addition (2.5) on gold elution efficiency are confounded to a degree thus making it difficult to isolate each parameter per se. At relatively high cyanide additions (60 kg NaCN/t carbon), temperature would appear to be the single most important parameter, particularly below 110°C (Table I). The selection of an operating temperature will depend on the

elution costs related to a particular adsorption operation. Because energy and reagent costs together constitute the major operating costs, such costs will define both parameters. These costs which are discussed in more detail later (3.0) would indicate the use of an elution temperature in the region of 110°C. While elution at high cyanide reagent additions is undoubtedly improved by increasing elution temperatures, the reverse may be true at low cyanide additions due to increased cyanide destruction. These effects will be discussed below (2.5).

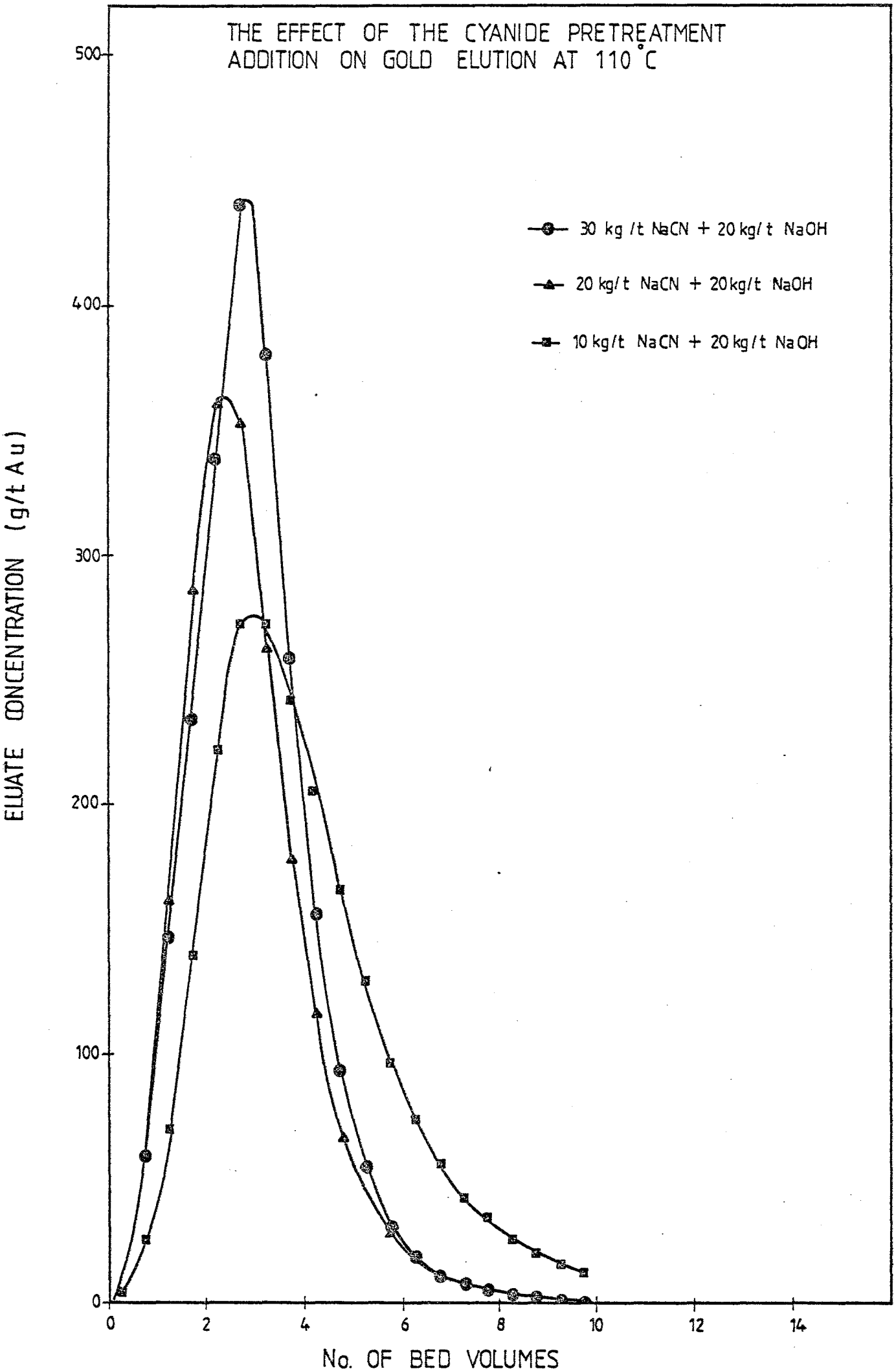
The selection of a suitable operating temperature will also depend on the choice of column lining. If columns are lined with butyl rubber, care should be made in adequately vulcanising the coatings, while operation at temperatures exceeding 110°C is not recommended. Other linings which may be considered include Hastalloy C and PDVF (polydivinylidine fluoride).

2.5 Pretreatment Reagents

Loaded carbon is pretreated with a relatively concentrated caustic cyanide reagent in order to initially solubilise contained gold and silver values before inhibiting their readsorption onto the carbon by drastically reducing the ionic strength of the medium using water as the eluant. Sodium cyanide is required for solubilising the adsorbed metal complexes while sodium hydroxide is used to neutralise the acid washed carbon and to maintain high alkalinity during subsequent elution (calcium cyanide is not effective when used as a pretreatment reagent). A considerable amount of sodium hydroxide may also react with silica at elution temperatures (see below).

In most operating plants the carbon is pretreated at temperature with approximately 1 bed volume of a 3 per cent NaCN/1 per cent NaOH solution (i.e. 60 kg NaCN + 20 kg NaOH per t carbon

FIGURE 6


THE EFFECT OF THE CYANIDE PRETREATMENT ADDITION ON GOLD ELUTION AT 110 °C
30 kg /t NaCN + 20 kg/ t NaOH
20 kg/ t NaCN + 20 kg/t NaOH
10 kg/t NaCN + 20 kg/ t NaOH
ELUATE CONCENTRATION (g/t Au)
No. OF BED VOLUMES
0
100
200
300
400
500
0
2
4
6
8
10
12
14

treated) which is pumped into the column and then allowed to 'soak' for 30 minutes. Softened regional water is then used to elute gold. In many instances, the 'soaking' of the carbon is not found to have any significant effect on subsequent gold elution. Using a pretreatment volume equivalent to 1 bed volume, it has been shown that little beneficial effect on gold elution is effected when increasing the cyanide concentration from 3 to 10 per cent NaCN. This effect may be due to the greater ionic strength of the more concentrated cyanide reagents having a counteracting effect on elution efficiency in the subsequent elution with water. The use of such high cyanide concentrations would thus not be cost effective.

When using cyanide concentrations of less than 3 per cent NaCN, the concentration of the reagent has a pronounced effect on gold elution. From the collated laboratory data presented in Tables II and III, it may be seen that cyanide concentrations of 2-3 per cent NaCN may be effectively used, while in most instances the volume of reagent may be reduced to approximately 0,6 bed volume (0,6 bed volume may be adjudged to be the minimum volume required to effectively displace the wash water from the carbon column). A cyanide addition equivalent to 20-30 kg/t carbon treated would appear necessary for efficient gold elution. From the data presented in Table III it may be seen that elution efficiency in terms of the number of bed volumes required for elution is improved significantly by increasing temperature from 100 to 110°C, but only marginally by increasing temperature within the limits 110-130°C. Conversely, the reagent cyanide addition may be seen to have a profound effect both on elution efficiency and on the peak height (maximum eluate concentration) of the eluate profile. This effect is most pronounced at low cyanide additions viz. 10 kg/t NaCN and is related in part at least to the increase in cyanide destruction taking place at the higher elution temperatures (see below). The effect of cyanide addition on

gold elution at 110°C is illustrated in Figure 6. While it is perhaps difficult to define cyanide requirements in general terms, significant cost savings in reagent cyanide would appear possible when related to current plant practises. The reagent cyanide requirement does not appear to be sensitive to the gold loading on the carbon.

TABLE II

REAGENT CYANIDE REQUIREMENTS FOR GOLD ELUTION

Carbon source	Elution temperature (°C)	Carbon pretreatment		Elution recovery (%/no. of bed volumes)
		bed volumes/ % NaCN	NaCN addition (kg/t carbon)	
A (7,1 kg/t Au)	110	0,5/1,0	10	95,6/7,5
	110	0,5/2,0	20	99,1/7,5
	110	0,5/3,0	30	99,3/7,5
	120	0,5/3,0	30	99,7/7,5
B (0,5 kg/t Au)	100	1,0/3,0	60	94,0/10,0
	110	0,6/5,0	60	93,8/7,5
	110	0,6/4,0	48	94,3/7,5
	110	0,6/3,0	36	93,4/7,5
	110	0,6/2,0	24	80,1/7,5
C (0,6 kg/t Au)	120	1,0/2,0	40	93,1/5,0 hot + 2 cold
	120	0,5/4,0	40	93,0/4,5 hot + 2 cold
	140	1,0/2,0	40	99,1/5,0 hot + 2 cold

TABLE III

THE EFFECT OF CYANIDE ADDITION AND TEMPERATURE ON ELUTION EFFICIENCY OF CARBON CONTAINING 1,9 kg/t GOLD

Elution temperature (°C)	No. of bed volumes required to elute 98% of contained gold (maximum eluate concentration - g/t Au)		
	10 kg/t NaCN addition	20 kg/t NaCN addition	30 kg/t NaCN addition
100	16,8 (165)	11,4 (255)	10,8 (287)
110	10,4 (278)	6,5 (362)	6,2 (440)
120	8,3 (295)	6,9 (328)	6,2 (407)
130	10,4 (258)	5,6 (366)	5,7 (369)

Significant/...

Significant quantities of reagent cyanide are either oxidised or polymerised during elution producing ammonia and sodium carbonate (equation 1), together with a distinctive yellow-brown coloured eluate thought to be azulmic acid (a complex cyanide polymer). If caustic additions are not adequate, a bluish eluate becomes evident which invariably indicates poor gold elution.

Because sodium cyanide constitutes a considerable portion of the total elution costs, both its addition and consumption must be considered if elution costs are to be minimised. At the present time, only limited laboratory date are available concerning cyanide consumption during elution. These data are presented in Table IV and graphically illustrated in Figure 7, where it may be seen that cyanide destruction is particularly sensitive to the cyanide addition (concentration) and to elution temperatures above 120°C. Whereas on average only 46 per cent cyanide was consumed at 100°C, 75 per cent was consumed on eluting at 130°C (Table IV). While the addition and recovery of cyanide may be significant when related to overall elution costs, it does define to some degree the selection of an optimum elution temperature. These costs will be discussed later (3.0).

TABLE IV/...

FIGURE 7

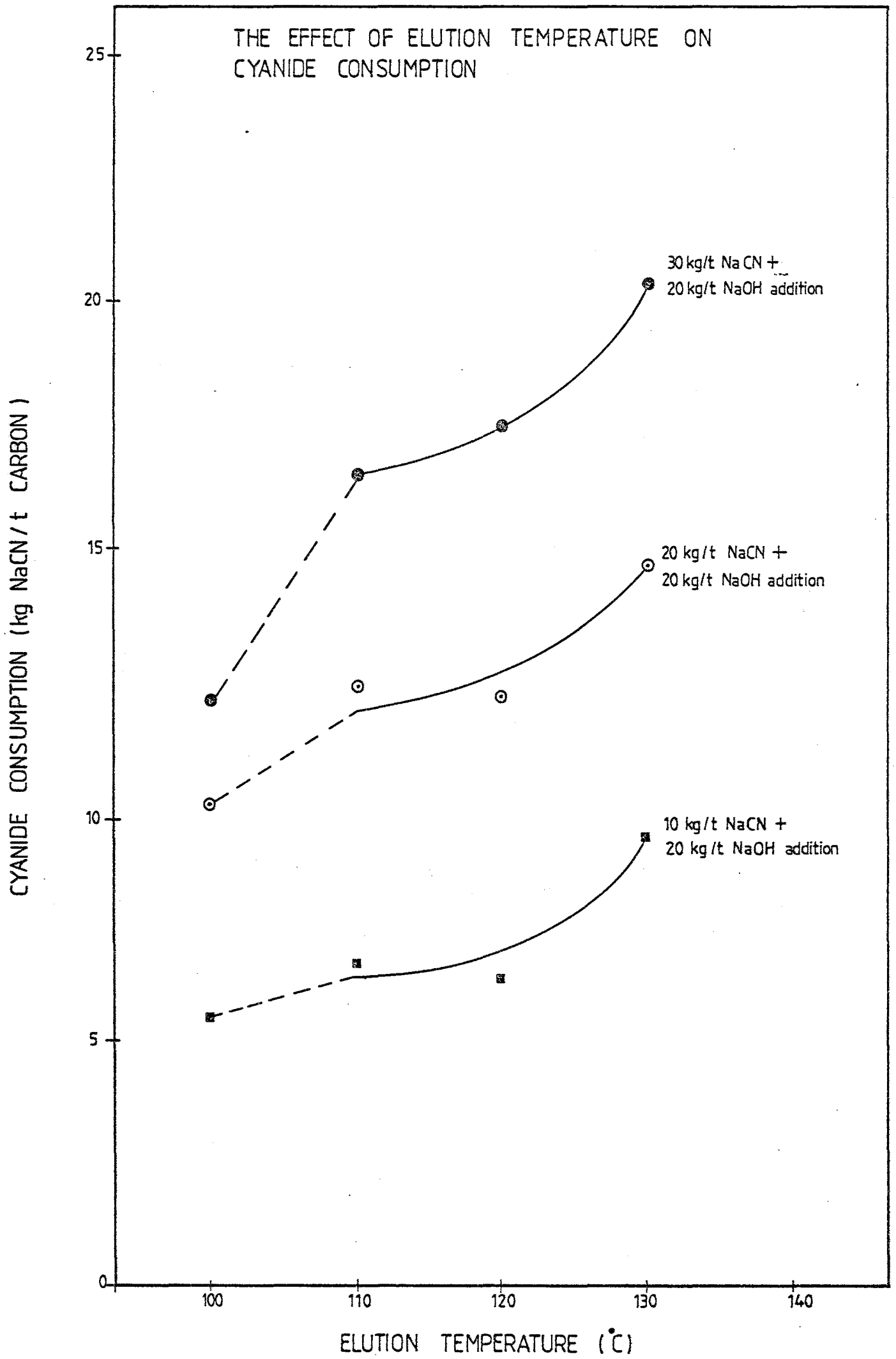

THE EFFECT OF ELUTION TEMPERATURE ON CYANIDE CONSUMPTION
30 kg/t Na CN + 20 kg/t NaOH addition
20 kg/t NaCN + 20 kg/t NaOH addition
10 kg/t NaCN + 20 kg/t NaOH addition
CYANIDE CONSUMPTION (kg NaCN/t CARBON)
ELUTION TEMPERATURE (°C)
0
5
10
15
20
25
100
110
120
130
140

TABLE IV

THE EFFECT OF CYANIDE ADDITION AND TEMPERATURE ON CYANIDE CONSUMPTION

Elution temperature (°C)	Cyanide balance (kg NaCN/t carbon)		
	Cyanide addition	Cyanide consumption	Cyanide recovered
100	10,6	5,5	5,1
100	20,6	9,8	10,8
100	31,2	11,9	19,3
110	10,7	6,6	4,1
110	20,2	12,2	8,0
110	32,4	16,5	15,9
120	10,7	6,4	4,4
120	20,4	12,0	8,4
120	30,4	17,5	12,9
130	10,5	9,2	1,3
130	20,9	14,7	6,2
130	30,8	20,4	10,4

While the pretreatment reagents may be pumped through the circuit in the prescribed sequence (2.1), both reagent recirculation as well as the draining of reagents from the column may be carried out. Due to higher reaction rates (2.4), reagent recirculation at 110°C may not significantly improve subsequent gold elution, whereas reagent recirculation at 90°C for 6-8 hours was shown to substantially improve elution recoveries. In the first instance, preheating of the column to operating temperature may be required, while draining of reagents from the carbon bed does afford the reuse of a concentrated cyanide reagent while producing eluate with a relatively low cyanide content. Both modifications may have merit in certain instances. The recirculation of column eluate as a means of reducing energy requirements is not recommended as elution efficiency is adversely affected.

In the/...

In the case where high copper loadings are encountered, excessive cyanide additions may be required yielding copper rich eluates which may in turn present problems on subsequent electrowinning or zinc precipitation. In such instances, copper may be selectively stripped from the carbon prior to gold elution. This removal of copper may be effected at ambient temperature by treating the carbon with an excess of sodium cyanide (equation 3).

$$Cu(CN)^{-}_{2} \quad \underset{\rightleftharpoons}{NaCN} \quad Cu(CN)^{=}_{3} \;\&\; Cu(CN)^{\equiv}_{4} \;\ldots\; equ^{n}\ 3$$

(strongly adsorbed) (weakly adsorbed)

At low gold loadings (less than 1 kg/t Au) more than 90 per cent of the contained copper may be removed while eluting less than 2 per cent of the contained gold (laboratory data).

Apart from maintaining the highly alkaline conditions necessary for gold elution, higher sodium hydroxide additions may be required in certain instances.

a) If excessive corrosion of anodes is encountered on the electrowinning of gold from eluates, further additions of sodium hydroxide may be required to maintain the eluate alkalinity above pH 12.

b) Higher caustic additions may be required if excessive silica fouling of the carbon is encountered due to the formation of soluble silicates (equation 4).

$$SiO_2 + 2NaOH \rightleftharpoons NaO_2.SiO_2.H_2O \;\ldots\; equ^{n}\ 4.$$

On cooling of the eluate much of this silica may precipitate as a calcium alumino silicate which may present problems when zinc cementation is practised.

Note:/...

Note: If caustic additions are excessive, poor elution of silver and base metals will be encountered particularly when operating at high temperatures. The precipitation of silver and base metals as hydroxides under such conditions would explain these observations. The elution of gold is not seriously affected due to the stability of the $Au(CN)_2^-$ complex.

2.6 Gold Elution

2.61 Water quality. Gold is normally eluted with a high quality softened regional water. Softening of the water is most desirable if choking of primary heat exchangers and poor thermal efficiency is to be avoided. While water quality undoubtedly plays a role in efficient gold elution, the use of softened regional water containing ca 300 g/t sodium has been shown to be satisfactory provided elution is preceeded by a hot acid washing stage (2.22). A marginal improvement in elution efficiency may be expected if demineralised water is used but the additional cost involved (ca R0,35/m^3) may not be cost effective.

2.62 Eluate flowrates. Assuming a column geometry length/diameter ratio of approximately 10, the elution of gold is virtually independent of flowrate in the region 1-5 bed volumes/h. Little change in the resulting elution profiles and on overall elution efficiency in this region would suggest that the elution mechanism is not intraparticle controlled under these conditions.

The selection of a suitable eluate flowrate is rather difficult to predict in terms of capital and operating costs as a practical compromise between peak power demand and elution time must be considered. Obviously, if several elution columns are operated in sequence, such peak power demands may be further decreased. If a complete

elution cycle is to be carried out in an 8 hour shift, flowrates of 2-3 bed volumes/h have been found to be satisfactory. At such slow flow velocities (20-30 m/h superficial) no significant pressure drop across the column may be expected.

As significant gas evolution may be encountered on acid washing carbon, eluate flow is preferably in an upflow direction. Also, if eluate collectors become choked, they are easily serviced from the top of the column (upflow) rather than at the base of the column (downflow).

2.63 Split elution. When elution costs become critical due to low gold loadings, several modificaitons to the elution procedure (2.1) may be made.

a) After normal water elution for 4-5 bed volumes the following 3-4 bed volumes of warm low conductivity eluate are used as the initial eluate for a subsequent batch elution. This modification may effect energy savings while only marginally impairing elution efficiency. The eluate tenor going forward to electrowinning or zinc precipitation may also be virtually doubled using this procedure.

b) Towards the conclusion of a normal elution, cold water may be used to displace the last 1 or 2 bed volumes of eluate thus effecting further energy savings.

2.7 Silver Elution

Because of the low content of silver in Witwatersrand type ores, little attention has been devoted to the elution of silver at the AARL. While routine tests using the AARL process have indicated elution recoveries of 60-80 per cent from

loaded carbons originating from South African operations, the elution of carbons containing high silver loadings have not been investigated. In a detailed factorial investigation on gold elution (2,5; Table III) eluted carbon samples were also analyzed for silver in order to establish the effects of temperature and pretreatment reagent additions on the elution of silver. The resulting recovery data are presented in Table V, where optimum silver elution at high temperatures (130°C) and low reagent cyanide additions (10 kg/t NaCN) may be noted. While these findings are perhaps rather unexpected, the data presented clearly indicate such a trend.

TABLE V

THE EFFECT OF TEMPERATURE AND REAGENT CYANIDE ADDITIONS ON THE ELUTION OF SILVER

Elution temperature (°C)	Elution Recovery (%)			
	10 kg/t NaCN addition	20 kg/t NaCN addition	30 kg/t NaCN addition	Average
100	59	69	66	65
110	83	79	63	75
120	86	79	72	79
130	93	86	72	84
Average	80	78	68	

3.0 ELUTION COSTS

In the absence of plant data related to the effect of various operational changes on gold elution, the present costing analysis will be based on collated evidence from bench tests carried out at AARL. Initially the analysis will relate to the cost of eluting 1

ton carbon/...

ton carbon and this cost then related to the eventual elution cost per kg of gold produced. As energy and reagent costs together comprise perhaps 90 per cent of such operating costs, other incidental costs will not be considered.

3.1 Acid Washing and Column Preheating

If it is assumed that 3 bed volumes are required to acid treat and wash 1 t carbon while preheating the system to the desired operating temperature ($t^{o}C$) and that this is carried out at a flowrate equivalent to 2 bed volumes/h, the following energy requirements may be made.

$$\text{Heat input} = \frac{2 \times 3 \times 10^{6} \times 1{,}0\ (t - 20) \times 4{,}2}{10^{6}} \quad \text{MJ/t}$$

$$= 2 \times 3 \times (t - 20) \times 4{,}2 \times 0{,}0111 \quad \text{R/t}$$

(bulk density of carbon 0,5 t/m^{3}; specific heat water 1,0 cal./g; ambient temperature $20^{o}C$. As the acid wash is not routed through a heat exchanger because of corrosion problems no heat recovery is made; assume an energy cost of R0,0111/MJ-R0,04/K watt h).

Assuming an acid requirement of 1 bed volume 3 vol. % HCl (32%) for acid washing carbon (0,03 x 2 000 x 1,18 kg) the cost of the acid would be R17,35/t carbon when costing acid at R245/t.

3.2 Pretreatment Reagents

Assume a cost of R1,52/kg for sodium cyanide and R0,61/kg for sodium hydroxide.

3.3 Elution Energy Requirements

The energy required to elute 1t carbon at temperature t°C may be calculated assuming a 50 per cent heat recovery from the eluate and a radiant heat loss of 5 per cent/h when eluting at 2 bed volumes/h (3.1).

Total heat = Heat input + Radiant loss

$$\text{Heat input} = \frac{2 \times \text{No. of bed vols.} \times 10^6 \times 1{,}0\ (t-20) \times 4{,}2}{2 \times 10^6} \quad \text{MJ/t}$$

$$= \text{No. of bed vols.}\ (t-20) \times 4{,}2 \times 0{,}0111 \quad \text{R/t}$$

Radiant loss = 0,05 x heat input x 3 MJ/t

= 0,1 x heat input x 0,0111 R/t

(the radiant loss of 5%/h was obtained from E.L. Bateman and relates to the President Brand elution circuit).

3.4 Total Elution Costs

Elution costs may be either related to the cost of eluting a given quantity of carbon or more specifically related to gold production. Both costing procedures have certain merits and will be dealt with separately.

a) Costs related to carbon treatment based on cyanide additions.

In order to identify high operating cost areas in a particular elution operation, such costs may be most conveniently related to a given quantity of carbon.

Using the data presented in Table III, total elution costs (3.1-3.3) relating to a column with a length/diameter ratio of 12 may be calculated for each prescribed set of operating conditions. These costs are presented in Table VI where it may be seen that elution at 110°C using a cyanide addition of 20 kg/t carbon treated is the most cost effective.

TABLE VI/...

TABLE VI

TOTAL ELUTION COSTS BASED ON CYANIDE ADDITIONS

Elution temperature (°C)	Total Elution Cost (R/t carbon)			
	10 kg/t NaCN + 20 kg/t NaOH	20 kg/t NaCN + 20 kg/t NaOH	30 kg/t NaCN + 20 kg/t NaOH	Average
100	136,04	129,09	141,84	135,65
110	117,91	115,13	128,92	120,65
120	115,29	123,31	134,91	124,50
130	134,17	122,31	138,06	131,51
Average	125,85	122,46	135,93	

Under these optimum conditions the following cost structure may be approximated:

	Operating cost (R/t carbon)	Total cost (%)
Reagent acid	17,35	15,1
Column preheating	25,17	21,9
Pretreatment reagents	42,60	37,0
Eluant heating	30,00	26,0
	R115,12	100,0

The very significant reduction in reagent cyanide from the ca 60 kg/t NaCN (R91,2) commonly in use in many elution circuits to only 20 kg/t (R30,4) may be noted in this particular exercise.

b) Costs related to carbon treatment based on cyanide consumptions.

Using/...

Using the cyanide consumption data presented in Table IV, total elution costs may be similarly calculated for each prescribed set of conditions. These costs are presented in Table VII where once more elution carried out at 110°C using 20 kg/t cyanide may be seen to be most cost effective.

TABLE VII

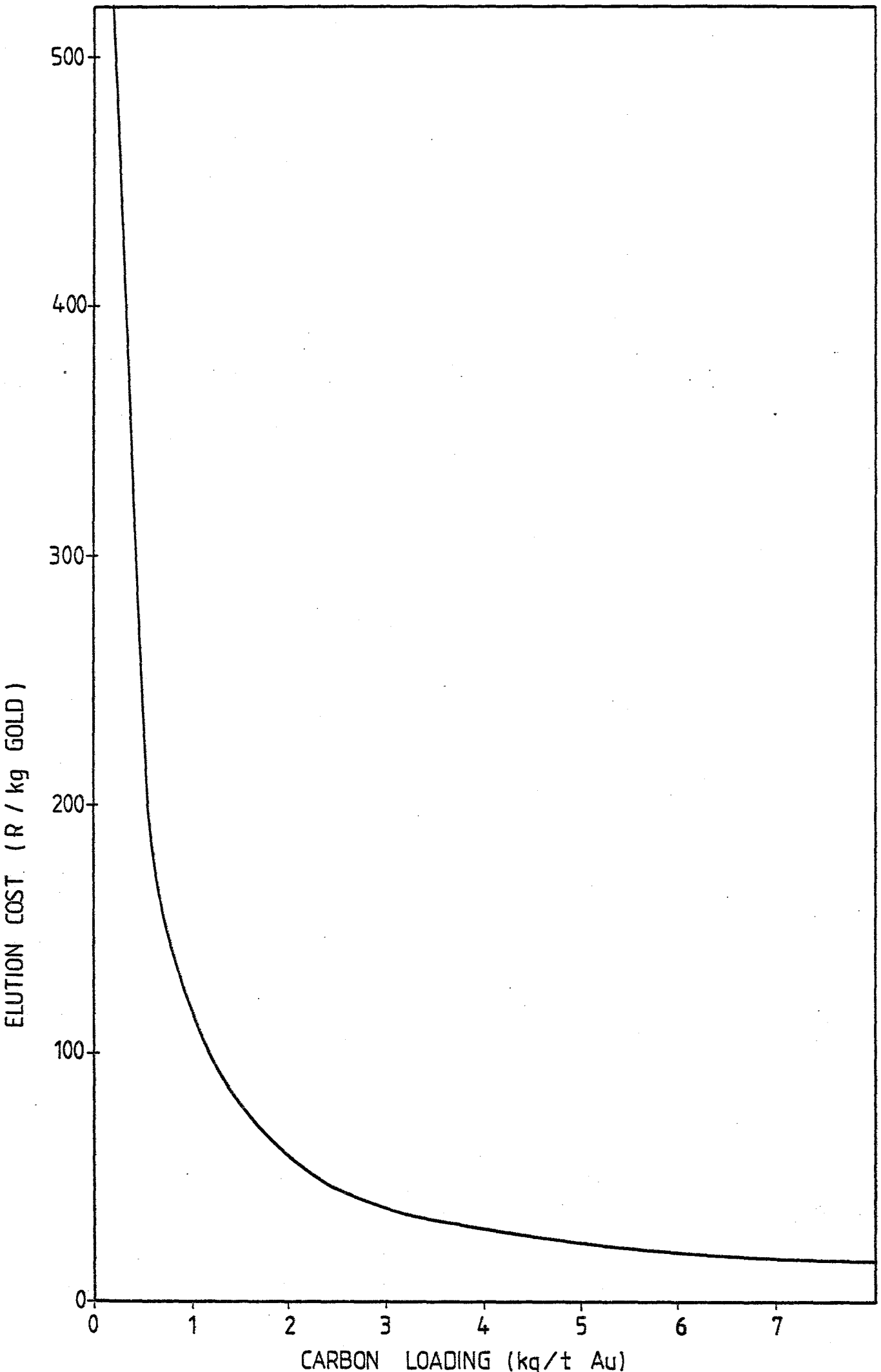
TOTAL ELUTION COSTS BASED ON CYANIDE CONSUMPTIONS

Elution temperature (°C)	Total Elution Cost (R/t carbon)			
	10 kg/t NaCN + 20 kg/t NaOH	20 kg/t NaCN + 20 kg/t NaOH	30 kg/t NaCN + 20 kg/t NaOH	Average
100	129,20	113,59	114,33	119,04
110	112,74	103,27	108,40	108,14
120	109,82	111,15	115,91	112,29
130	132,95	114,25	123,47	123,56
Average	121,18	110,57	115,52	

c) Costs related to gold production.

As may be expected, the elution cost becomes more meaningful when related to actual gold production. However, such costs are particularly sensitive to carbon loadings so that elution costs cannot be assessed in isolation as much will depend on the loadings achieved in the adsorption circuit. This sensitivity to carbon loading is well illustrated in Figure 8 where the effect of carbon loadings on elution costs have been considered while operating a plant as outlined in a) above under the most cost effective conditions. As may be seen from this data, elution costs may be only R10-20 per kg gold produced when eluting carbon containing high gold loadings but these costs may exceed R400 per kg gold produced at low carbon loadings.

FIG. 8 THE EFFECT OF CARBON LOADINGS OF ELUTION COSTS



3.5 Column Geometry

Elution energy requirements will also depend to a large degree on the column geometry and its effect on elution efficiency (2.3). Assuming the data presented in Table I, the effect of column geometry on elution costs may be calculated as before (3.1-3.3). While high length/diameter ratios will undoubtedly improve metallurgical efficiencies, practical and capital cost restraints may limit the length of the elution column. Current operations using columns having length/diameter ratios of 8/1-10/1 have been found to operate most successfully.

3.6 Elution Cost Savings

The present costing exercise (3.1-3.5) has been based on a generalised elution procedure which may well be modified in certain situations where costs become critical or where the assumptions made do not apply. These aspects will be discussed.

a) The costing exercise is based on the cost of reagent additions and not on reagent consumptions. Significant reagent cost savings may be made depending on the subsequent procedure used for eluate treatment. If the eluate were treated by zinc cementation and the barren eluate returned to the leach, little reagent cyanide would be lost as the zinc complex $Zn(CN)_4^{=}$ is an efficient lixiviant for gold. Alternatively, if the eluate was routed to electrowinning in a Mintek type cell, a large proportion of the free cyanide present would be anodically destroyed. This oxidation of cyanide would not occur in an AARL type cell containing a cation exchange membrane. Furthermore, in the case of electrowinning, higher caustic additions may be required to avoid anode corrosion.

b) In a situation where low gold loadings result in critical energy costs, further energy savings may be made by using a split elution (2.63). The use of a cold water displacement wash at the termination of the elution would also afford

some/...

some saving. These latter modifications were not considered in the costing exercise but may well result in the saving of 1-3 bed volumes of hot eluate. This energy saving may be equated to ca R6-18/t carbon treated when elution is carried out as prescribed above (3.4).

c) Provided elution is preceeded by a hot acid wash in which all but the gold, silver and copper values are removed from the carbon, little change in the prescribed method of elution should be required unless abnormally high copper loadings are encountered (2.5).

4.0 DISCUSSION

From both a metallurgical and costing viewpoint, the two significant factors affecting gold elution using the process under review are temperature and reagent cyanide additions. As the effects of temperature, reagent cyanide addition and ionic strength gradient are undoubtedly confounded to a degree, it is not possible to quantify each parameter in isolation. At high cyanide additions where a high ionic strength gradient is operative, temperature would appear to play the dominant role in effecting gold elution. Conversely, at low cyanide additions where a relatively low ionic strength gradient is involved, cyanide addition rather than temperature is the more significant operational parameter. While high temperatures and high reagent additions within the prescribed limits will invariably improve metallurgical efficiencies, a balance between energy and reagent costs is required in order to minimise overall elution costs. Such factors will of course become more critical when eluting carbon containing low gold loadings.

Other considerations which affect gold elution include the column geometry and the acid washing of carbon. Column geometry may well have both practical and capital cost restraints limiting the length of the elution column, while the many benefits effected by hot acid washing prior to elution are difficult to quantify in simple terms.

A further most important consideration which is often overlooked concerns the need for carbon regeneration. In many current operations inefficient gold elution produces eluted carbons containing as much as 50-150 g/t gold which if returned to the last stage of an adsorption circuit will not adsorb gold due to equilibrium rather than kinetic restraints. However, if this carbon is heated above ca 350°C in a regeneration kiln this adsorbed gold is reduced to the metallic state so that when returned to the adsorption circuit it responds much the same as a virgin carbon. This aspect is most important if the thermal regeneration of carbon following each adsorption cycle is not carried out. In such circumstances, the efficient elution of gold producing carbon containing 10-20 g/t gold is necessary in order to scavenge gold at the 0,01-0,03 g/t level.

The intention in a review of this nature is to provide an in-depth up-dated metallurgical view on the AARL elution process per se while not comparing the merits or otherwise of other elution processes. While considerable attention at the AARL has been devoted to studying both the Zadra elution and the elution of carbon using polar organic solvents, neither of these processes would appear to compete favourably in terms of operating costs. In the longer term, however, the development of the AARL elution from a batch to a continuous counter current process (S.A. Patent Application No. 84/2343) would appear to offer significant energy and reagent cost benefits together with improved metallurgical efficiencies.

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4.2 CONTINUOUS ELUTION

Following the early development and piloting of continuous gold elution by the Anglo American Research Laboratories (SA Patent No. 85/2404), a 1 t/d prototype continuous elution plant was installed by General Mining Metals and Minerals at the Grootvlei Gold Mine in December 1989.

Initial metallurgical testwork aimed at reducing reagent costs has shown that elution efficiencies of 98-99% could be achieved using reagent additions as low as 16 kg/t NaCN and 28 kg/t NaOH while consistently producing eluted carbon containing less than 50 g/t gold.

Significant savings in elution reagents and power were indicated when compared with the traditional batch AARL, Zadra and Micron alcohol elution procudures.

Further optimisation testwork is continuing while these early findings were presented at Randol Gold Forum 90, Colorado, U.S.A.

SPOOR AND FISHER

FORM P.7
(To be lodged in duplicate)

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978

COMPLETE SPECIFICATION

(Section 30(1) – Regulation 28)

OFFICIAL APPLICATION NO.		LODGING DATE	
21 01	85/2404	22	29 MARCH 1985

INTERNATIONAL CLASSIFICATION	
51	C 22 B

FULL NAME(S) OF APPLICANT(S)	
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TITLE OF INVENTION	
54	METAL RECOVERY

BACKGROUND OF THE INVENTION

This invention relates to a method of recovering metal values selected from gold, silver, copper and nickel from a support having one or more of these values captured thereon.

Generally, metal values are extracted from their ores by finely grinding the ore, and then leaching the ore with a suitable leaching solution. For example, gold values are extracted from their ores by leaching finely ground gold-bearing ore with a cyanide solution. The product of the leaching step is a metal-rich leach solution or pulp, known as a pregnant solution or pulp. The metal values may then be recovered from the pregnant leach solution or pulp using various techniques. One widely used technique is to capture the metal values on a suitable support such as an activated carbon. Thereafter, the loaded support is placed in a column and is eluted with a suitable eluant until substantially all the metal values have been eluted from the support, the metal values reporting in the eluate. Once the metal values have been eluted from the support, the support is removed from the column and is either regenerated and then re-used or simply re-used. The elution of the metal values from the support takes a considerable time, and requires a considerable volume of eluant.

SUMMARY OF THE INVENTION

According to the present invention there is provided a method of recovering metal values selected from gold, silver, copper and nickel from a support comprising activated carbon having one or more of these metal values captured thereon, including the steps of:

(a) placing an amount of the support with one or more of the metal values captured thereon in a column having a first end and a second end,

(b) introducing an eluant into the column at or near the second end of the column to elute the metal values captured on the support,

(c) collecting the eluate containing the metal values eluted from the support at or near the first end of the column,

(d) while continuing the flow of eluant through the column, periodically removing a portion of the support from which substantially all the metal values have been eluted, from the second end of the column, and

(e) replacing the removed portion of the support with a further amount of the support having metal values captured thereon, the further amount of the support being introduced at the first end of the column.

The method of the invention is a batch-continuous process in that the support having the metal values captured thereon is passed in the form of discrete slugs through the column and the eluant is continuously passed through the column countercurrent to the support.

The column is preferably substantially vertical with the first end above the second end.

The support having the metal values captured thereon is preferably pre-treated with a suitable pre-treatment reagent prior to being introduced into the column. The suitable pre-treatment reagent may be selected from an alkali metal cyanide solution, an alkali metal hydroxide solution, mixtures of these two solutions, and solutions comprising a mixture of an organic solvent and a solution selected from the group of an alkali metal cyanide solution, an alkali metal hydroxide solution and mixtures thereof. The preferred pre-treatment solution is a solution containing 1 to 10 percent by weight of sodium hydroxide and 1 to 20 percent by weight of sodium cyanide.

The eluant is preferably water having a low concentration of cations. The water preferably has a concentration of metal cations of less than 300ppm, preferably less than 50ppm. The water may be, for example, deionised or distilled water.

The method of the invention is preferably carried out at a temperature of between 90 and 120°C. This means that the column itself is heated, e.g. by means of hot oil pumped through a jacket located around the outside of the column, to a temperature between 90 and 120°C, that the eluant is heated to a similar temperature prior to being introduced into the column, and in that the pre-treatment step is carried out at a similar temperature. The method

may be carried out at a temperature of 90 to 100°C at ambient pressure or at a temperature of 100 to 120°C at an elevated pressure, e.g. a pressure of between 300 and 400kPa (gauge).

The amount of the support removed from the column and the frequency of removal, depend on the size of the column, the quantity of metal values captured on the support, the rate of flow of the eluant through the column, the nature of the support, and various other factors. Clearly, the amount of the support removed and the frequency of removal must be such that substantially all the metal values have been eluted from the support before it is removed from the column. Equally, to ensure that the method of the invention is efficient, those portions of the support from which the metal values have been substantially eluted, must not be left in the column.

BRIEF DESCRIPTION OF THE DRAWING

The drawing is a schematic view of apparatus for use in carrying out the method of the invention.

DESCRIPTION OF PREFERRED EMBODIMENTS

Referring to the drawing, apparatus for carring out the method of the invention comprises a column 10, a pre-treatment vessel 12 connected to the top of the column 10 via a line 14 and a valve 16, and a slugging chamber 18 connected to the bottom of the column 10 via a line 20 and a valve 22. The pre-treatment vessel 12 includes an inlet 24 for the loaded support, and is conected via lines 26 and 28 to a pre-treatment reagent tank 30. The slugging chamber 18 includes an inlet for water 32, and an outlet 34 for the eluted support. The column 10 includes an inlet 36 for the eluant, the

inlet 36 being connected by a line 38 to an eluant storage tank 40 via a pump 42 and an oil bath 44. The column 10 also includes an outlet 46 for the eluate.

The method of the invention will now be described with reference to the elution of gold values from activated carbon. The activated carbon having gold values adsorbed thereon (loaded activated carbon), after a pre-treatment as described hereinafter, is placed in the column 10. The column 10 is heated to a temperature between 90 and 120°C for example by means of hot oil which is circulated around the column 10 from the oil bath 44. The eluant, which is preferably deionised water, is fed via the pump 42 from the eluant storage tank 40 through the oil bath 44 to heat it to a temperature between 90 and 120°C, and then via the inlet 36 into the column 10. The eluant moves up through the column 10 in the direction of the arrow A and desorbs the gold values from the activated carbon. The eluate containing the gold values, exits the column 10 through the outlet 46 and is collected. Periodically, the valve 22 in the line 20 is opened so that a portion of the activated carbon from which the metal values have been desorbed, flows through the line 20 into the slugging chamber 18. The carbon flow from the column 10 into the slugging chamber 18 may be assisted by the utilisation of a high frequency water pulse. When the slugging chamber 18 is full, the valve 22 is closed again. Thereafter the valve 16 in the line 14 is opened and a fresh amount of activated carbon flows from the pre-treatment vessel 12 into the column 10. The eluted activated carbon is removed from the slugging chamber 18 by passing a flow of water via the inlet 32 through the slugging chamber 18 so that the eluted carbon exits the slugging chamber 18 through the outlet 34. The slugging chamber 18 is now ready to receive a further portion of activated carbon from which the gold values have been desorbed from the column 10.

Before the loaded activated carbon is fed into the column 10, it is treated in the pre-treatment vessel 12 with a suitable pre-treatment reagent at a temperature of 90 to 120°C, fed to vessel 12 from the storage tank 30 via line 26. Once the loaded activated carbon has been pre-treated with the pre-treatment reagent, the pre-treatment reagent is bled out of the pre-treatment vessel 12 and flows via the line 28 back into the tank 30 for re-use.

The step of removing a portion of the activated carbon from which the gold values have been desorbed from the column is repeated periodically while the elution proceeds continuously. It can be seen that there is a countercurrent flow of the activated carbon and the eluant through the column 10.

A comparison between a known elution process and the method of the invention will now be given. Both the known elution process and the method of the invention were carried out in apparatus such as that illustrated in the drawing.

MATERIALS AND METHOD

Loaded Activated Carbon

The experiment was carried out using loaded activated carbon taken from the Ergo CIL pilot plant. The gold loading of the carbon was about 300 g/t.

Elution Column

The elution column consisted of a 3m long stainless steel column of 52,5mm internal diameter with a pre-treatment vessel positioned at the top of the column and a slugging chamber positioned at the base of the column. The slugging chamber and pre-treatment vessel which

were pre-heated by column eluate allowed for the plug flow of carbon through the circuit while operating at 120°C and 300-400 kPa (gauge). The volume of the elution column was 6,5ℓ and the volume of the slugging chamber was 1,0ℓ. Heating of the circuit was provided by hot oil which was pumped through a jacket surrounding the elution column and which pre-heated the column eluant.

Carbon Pre-treatment

About one litre quantities of air dried loaded carbon were treated with 500mℓ of 6 percent NaCN/2 percent NaOH reagent and placed in a loaded carbon storage hopper above the pre-treatment vessel at regular intervals so as to maintain a slight excess of loaded carbon in the hopper.

Eluant

The eluant used was deionised water.

PRIOR ART BATCH PROCESS

In terms of the prior art batch process, the column was filled with an amount of the loaded activated carbon. The gold values were eluted from the activated carbon with the eluant comprising deionised water. Approximately 12 litres of eluant were used per kilogram of carbon. The process was carried out at a temperature of 120°C with a carbon residence time in the column of 6 hours. The results of this process are given in Table I.

METHOD OF THE INVENTION

Before the start of continuous operation, the column was filled with 6,5 ℓ of eluted carbon. 3 ℓ of air-dried loaded carbon and 1,5 ℓ of pre-treatment reagent were then added to the pre-treatment vessel and the loaded-carbon hopper while the bottom slugging chamber and the elution circuit were purged of air. After the circuit had been heated to the desired temperature (120°C), the column eluant pump was switched on and the slugging of carbon at regular intervals commenced.

During continuous operation, carbon was moved through the circuit at regular time intervals using the following sequence:

(a) Carbon was allowed to flow into the slugging chamber using a high frequency water pulse (2,4 s^{-1}) to assist the carbon flow, while the warm water displaced from the slugging chamber entered the elution column. When this slugging chamber was filled with carbon (1-2 min), it was isolated and the carbon evacuated using tap water at ca 500 kPa. Each slugging cycle removed about 500 g of carbon from the elution column.

(b) Carbon from the pre-treatment vessel which was otherwise kept filled with carbon, was allowed to flow into the top of the elution column. A water pulse (2,4 s^{-1}) was again used to assist carbon flow, while the hot eluate displaced from the elution column entered the pre-treatment vessel.

(c) Following the above carbon slugging operations, the pre-treatment vessel was filled with carbon from the storage hopper. When necessary, excess pre-treatment reagent which collected in the hopper due to carbon displacement was periodically drained from the hopper.

During operation, deionised water was pumped continuously, at a preset rate, into the elution column.

The results of three test runs carried out as described above are given in Table I, as Tests Nos. 2, 3 and 4. Test No. 1 is the prior art batch process test.

TABLE I

Test No.	1	2	3	4
Elution Temp.°C	120	120	120	120
Carbon Residence time in elution (hr)	6	3	3	6
Solution usage (litres per kg of carbon)	12,00	7,25	4,00	4,78
Steady State Conditions: Eluate Au Conc. (mg/ℓ)	-	48,0	63,8	45,3
Eluted carbon Au loading (g/t)	-	10	61	14
Gold Recovery %	95,3	96,7	79,7	95,3

It can be seen that the percentage gold recovery achieved utilising the method of the present invention, at least for runs 2 and 4, is as good as the percentage gold recovery achieved using the prior art process. Further, for all three runs of the method of the present invention, the usage of eluant was far less than the usage in the prior art process. It appears that solution usages of between 4 and 7,25 ℓ per kilogram of carbon can be achieved. In addition, residence times of less than 6 hours can also be achieved using the method of the present invention. The results of test No. 3 are not particularly good. This, however, can be attributed to operational problems which were encountered during this test. These included the incorrect opening and closing of valves in the circuit which resulted in a sudden loss of pressure and hence a temperature drop, and the loss of carbon and solution from the base of the column also causing a loss of pressure and temperature.

The method of the invention has several advantages over the prior art batch process. Firstly, the method of the invention requires the use of much less eluant than the prior art batch process. Secondly, carbon residence times in the column are substantially reduced when the method of the present invention is utilised. Thirdly, the method of the invention leads to substantial energy savings as there is no need to stop the process, cool down all the apparatus, remove the eluted support from the column, and then fill the column with a fresh amount of loaded support which then has to be re-heated. Fourthly, less power is required and the peak power demand is very much lower. Fifthly, using the method of the invention, the pre-treatment reagent is not pumped through the bed of the support in the column. This means that the pre-treatment reagent may be recovered for re-use, and the elution of the metal values from the support is improved. Finally, improved metallurgical efficiency is achieved.

The method of the invention has particular application to the elution of activated carbon with low loadings of Au and Ag values.

1.

A method of recovering metal values selected from gold, silver, copper and nickel from a support comprising activated carbon having one or more of these metal values captured thereon includes the steps of:

(a) placing an amount of the support with one or more of the metal values captured thereon in a column having a first end and a second end;

(b) introducing an eluant into the column at or near the second end of the column to elute the metal values captured on the support;

(c) collecting the eluate containing the metal values eluted from the support at or near the first end of the column;

(d) while continuing the flow of eluant through the column, periodically removing a portion of the support from which substantially all the metal values have been eluted from the second end of the column, and

(e) replacing the removed portion of the support with a further amount of the support having metal values captured thereon, the further amount of the support being introduced at the first end of the column.

2.

A method according to claim 1 wherein the column is substantially vertical with the first end above the second end.

3.

A method according to claim 1 or claim 2 wherein the support having the metal values captured thereon is pre-treated with a suitable pre-treatment reagent prior to being introduced into the column.

4.

A method according to claim 3 wherein the pre-treatment reagent is selected from an alkali metal cyanide solution, an alkali metal hydroxide solution and mixtures thereof.

5.

A method according to any one of claims 1 to 4 wherein the eluant comprises water having a low concentration of metal cations.

6.

A method according to claim 5 wherein the eluant comprises deionised or distilled water.

7.

A method according to any one of the preceding claims wherein the method is carried out at a temperature of between 90 and 120°C.

8.

A method according to any one of claims 1 to 6 wherein the method is carried out at a temperature of 90 to 100°C at ambient pressure.

9.

A method according to any one of claims 1 to 6 wherein the method is carried out at a temperature of 100 to 120°C at an elevated pressure.

10.

A method according to claim 1 substantially as herein described with reference to the accompanying drawing.

DATED this 29th day of MARCH 1985.

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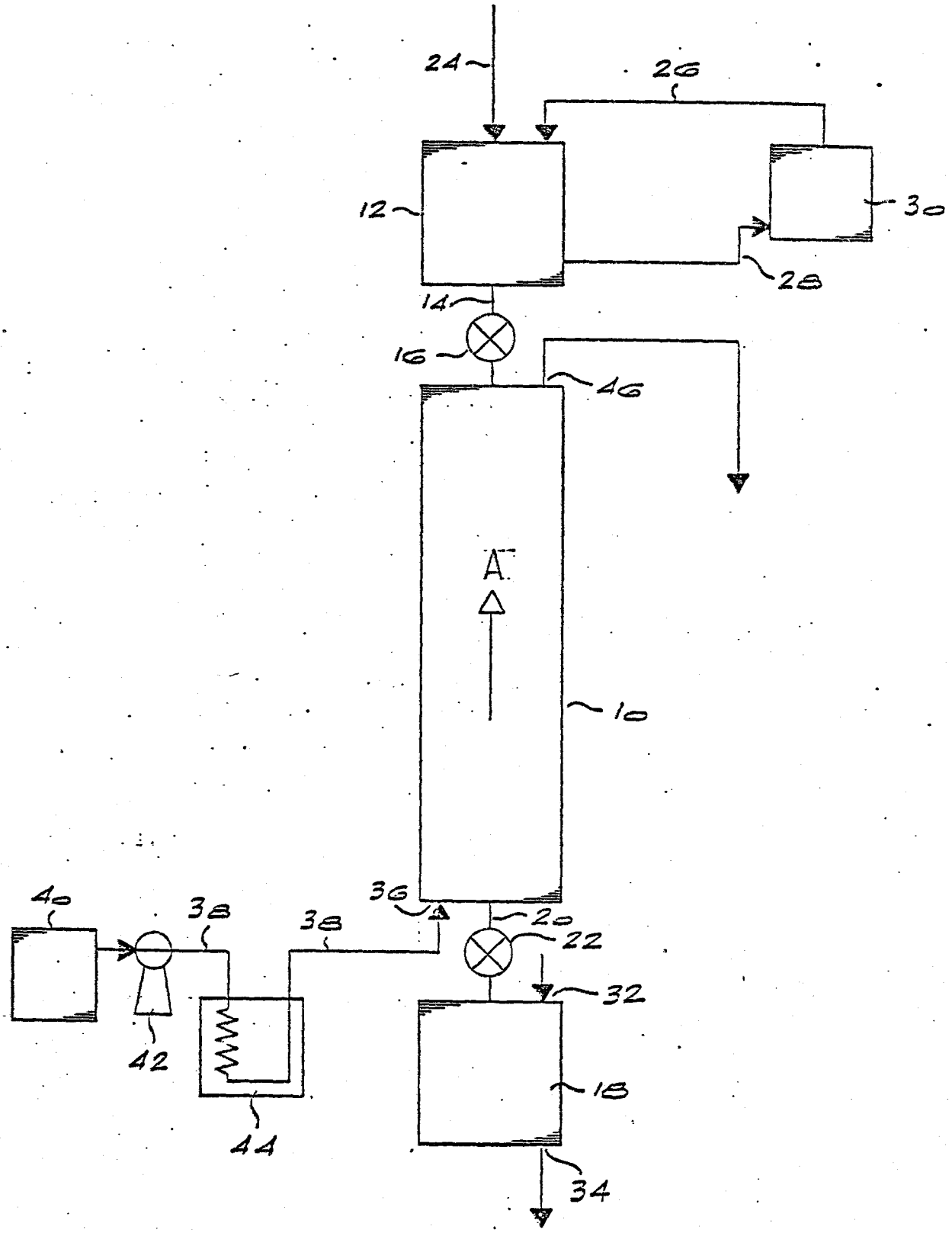
SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS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CONTINUOUS ELUTION OF GOLD FROM ACTIVATED CARBON

BY

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SUMMARY

A 1 t/d continuous elution pilot plant is being operated by General Mining at their Grootvlei Gold Mine using the technology developed at the Anglo American Research Laboratories. Flowsheet design and plant erection was carried out by Van Eck & Lurie and completed in December, 1989.

Following the conclusion of 6 test runs designed to optimise reagent additions, it was shown that elution efficiencies of 98-99% could be achieved using reagent additions as low as 16 kg/ton of carbon NaCN and 28 kg/ton NaOH while consistently producing eluted carbon containing less than 50 g/t gold.

Significant savings in elution reagents and power were indicated when compared with the traditional batch AARL, Zadra and Micron alcohol elution procedures.

Further optimization testwork is continuing.

CONTINUOUS ELUTION OF GOLD FROM ACTIVATED CARBON

1.0 INTRODUCTION

Following the early development and piloting of continuous gold elution by the Anglo American Research Laboratories Ref. 1, a 1 t/d prototype continuous elution plant was installed by General Mining Metals and Minerals at the Grootvlei Gold Mine in December 1989.

Design flowsheets for the plant were prepared by Van Eck & Lurie with assistance from AARL. Erection of the plant by Van Eck & Lurie commenced in November, 1989 and was completed in December, 1989. The pilot plant was incorporated into the existing Grootvlei plant using existing steelwork to support the columns and for access. The cost of the hardware for the pilot plant including erection was approximately R260 000.

1.0/continued

Following plant commissioning, preliminary metallurgical testwork was started in February, 1990. The present report concerns the design, engineering and commissioning of the prototype plant at Grootvlei with a summary of the metallurgical testwork carried out to date.

2.0 PLANT DESIGN

Figure 1 illustrates a simplified flowsheet of the elution circuit where gold is continuously eluted in a moving bed type mode using deoinised water as the eluant.

The elution circuit comprises 3 basic functional units:

A. Plant Section

Loaded carbon is initially treated with a caustic cyanide solution which is continuously circulated in the pretreatment section at ca 90°C. Solution levels in this section are controlled via the hydrostatic level of the eluate discharge. As the volume of the drained carbon fed to this section approximates the volume of the carbon being discharged from it, reagent losses from the circuit are minimised. The carbon residence time for carbon pretreatment when feeding 1 t/d is 7 h.

B. Elution Column

The elution column made from type 316 stainless steel has solution sampling points spaced at 2 metre intervals with the option of adding/removing insulating lagging for controlling temperature profiles. The carbon residence time in this section when feeding 1 t/d carbon is 3,5 h. The superficial flow velocity in the column when pumping 200 l/h eluate is 106 mm/min to prevent bed expansion at these flowrates and temperatures.

B./continued

200 l/h of eluate may be equated with eluting carbon using 3,3 bed volumes when treating 1 t/d of loaded carbon.

C. Slugging chamber

Carbon is moved out of the elution column via the slugging chamber using a pulsing pump to aid carbon movement. The water in the slugging chambers is preheated prior to carbon transfer and is displaced into the column during transfer. Carbon movement is controlled by three automatic valves sequenced from a time controller. The volume of carbon moved at each sequence is 37 l, thus carbon is moved every 27 min when treating 1 t/d carbon. The slugging chamber may be calibrated by physically drying and weighing several carbon discharges, and thus a more exact estimating of treatment tonnages is possible - bulk volumetric measurements of carbon as a means of estimating carbon tonnages are notoriously inaccurate.

Following plant erection, the plant was first commissioned with cold and then hot water whereby all pumps and the slugging chamber were calibrated. Several problems were encountered with carbon movement due to the presence of oversize tramp material in the carbon feed which were overcome by screening all carbon at 5 mm (square mesh) entering the circuit. Various problems related to the calibrating of temperture probes were also encountered and rectified.

3.0 PLANT OPERATION

Metallurgical testing commenced on 12th March, 1990 aimed at investigating the effects of various pretreatment reagent cyanide and caustic additions on gold elution efficiences. These reagents were controlled at specific concentrations in the pretreatment section where solution was recirculated at 500 l/h while maintaining a temperature of 90°C. Unless otherwise specified, eluant flowrate to the elution column was maintained at 210 l/h while carbon was moved from the column at 80 l/h, equivalent to 986 kg/d. Drained, acid washed, loaded carbon containing 35 - 40% moisture was used as feed to the elution plant. At least 2 days of continuous operation was assumed to be necessary in order to approach steady state plant operations using any set of prescribed operational parameters.

Test runs were carried out in the present phase of the investigation using pretreatment cyanide concentrations of 0,25-3,0% NaCN and caustic concentrations of 0,5-2,0% NaOH. These reagents were added continuously as concentrated solutions into the pretreatment section by means of micropumps while checking reagent concentrations in the pretreatment section at regular 2 hour intervals. Each test run involved the continuous operation of the plant for at least 6 days in order to generate meaningful metallurgical data.

Averaged steady state metallurgical data for these 6 test runs are presented below (4.0).

4.0 METALLURGICAL RESULTS

4.1 Gold Elution

Elution data for the 6 test runs carried out are summarised in Table I, where it may be seen that 98-99% elution efficiency was obtained using cyanide pre-treatment concentrations greater than 0,25%

4.0 /continued

and caustic pretreatment concentrations greater than 1,0%. The effective use of 0,25% NaCN/1,0% NaOH (test run No. 5) involving a cyanide addition of 16,5 kg/t NaCN and a caustic addition of 28,2 kg/t NaOH may be noted. Relatively high eluate concentrations (230-442 g/m^3 Au) are also noteworthy.

4.2 Energy Requirements

Using the water balance data for test runs 3-6 presented in Table II where eluant temperatures average 118°C and eluate temperature average 88°C, heat balance calculations (Appendix I) indicate an energy requirement of 630 kWh/t carbon treated. (These calculations assume that 50% of the heat in the eluate may be recovered in a secondary heat exchanger on a commercial plant).

4.3 Further Evaluation

Testwork is continuing to evaluate the effect of feed temperature and temperature profile across the column. The use of lower eluant flowrates as well as the effect of water quality is also to be investigated for the eluant.

5.0 DISCUSSION

When compared with other batch elution circuits operating in South Africa (Ref. 2), the present findings indicate significant savings in terms of the reagents associated with gold elution from loaded carbon. The reagent cyanide addition of 16 kg/t required for continuous elution may be compared with the 50-80 kg/t used in batch AARL type circuits or the 35-50 kg/t used in Zadra circuits, although 50-70% of the reagent cyanide is recovered in AARL elution circuits. Likewise, a reagent caustic

5.0 /continued

addition of 28 kg/t may be compared with 20-40 kg/t used in AARL circuits or the 50-100 kg/t used in Zadra circuits. The savings in reagent cyanide in continuous elution are undoubtedly related in part to the low cyanide concentrations and temperatures used in carbon pretreatment as the hydrolysis of cyanide increases considerably at temperatures above 110-120°C. The containment of these reagents in the pretreatment section also minimises reagent losses.

The present investigation although not optimised in terms of energy requirements (elution temperatures or flowrates) would also indicate savings in power costs. The calculated energy requirement of 630 kWh/t for continuous elution may be compared with ca 1050 kWh/t for the AARL elution and 3530 kWh/t for Zadra elution.

Continuous elution would also exclude any peak power demand surcharge which may be incurred using the AARL batch elution and which may be excessive in some instances.

Other advantages which may be offered by continuous elution include high metallurgical efficiency together with savings in labour and maintenance as the system is well suited to on-line control and operation. Design considerations are less stringent as the elution column is not classified as a pressure vessel as is the case with the batch AARL elution and pressure Zadra process.

Based on the results obtained to date, considerable interest has been shown in the continuous elution design and this has resulted in the design being started on a 3 t/d continuous acid wash and elution circuit to be incorporated on a new plant. In addition, investigations have been undertaken to modify existing batch

5.0/continued

circuits to achieve the advantages shown by the continuous elution pilot plant. Circuits being studied range up to 25 t/d.

DR R.J. DAVIDSON

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ACKNOWLEDGEMENTS

The contributions made by General Mining Metals and Minerals, the Anglo American Corporation and by Van Eck & Lurie (Pty) Ltd as well as operating staff at the Grootvlei Gold Mine are gratefully acknowledged.

TABLE I

THE EFFECT OF PRETREATMENT REAGENT CONCENTRATIONS ON GOLD ELUTION EFFICIENCIES

Test run No.	1	2	3	4	5	6
Cyanide/Caustic Conc. (%NaCN/% NaOH)	3,0/2,0	1,0/2,0	0,5/2,0	0,5/1,0	0,25/1,0	0,25/0,5
Eluant temperature (°C		106	115	117	118	119
Eluate temperature (°C		82	88	88	88	88
Loaded carbon (g/t Au)	2090	1950	2080	1515	2493	3041
Eluted carbon (g/t Au)	28	35	18	21	43	140
Eluate Conc.(g/m^3 Au)	276	281	344	230	442	416
Elution recovery (%)	98,7	98,2	99,1	98,6	98,3	95,4
Cyanide addition (kg/t carbon as 100% NaCN)	70	46,7	20,4	27,9	16,5	15,7
Caustic addition (kg/t carbon as 100% NaOH)		51,7	40,3	34,9	28,2	23,9

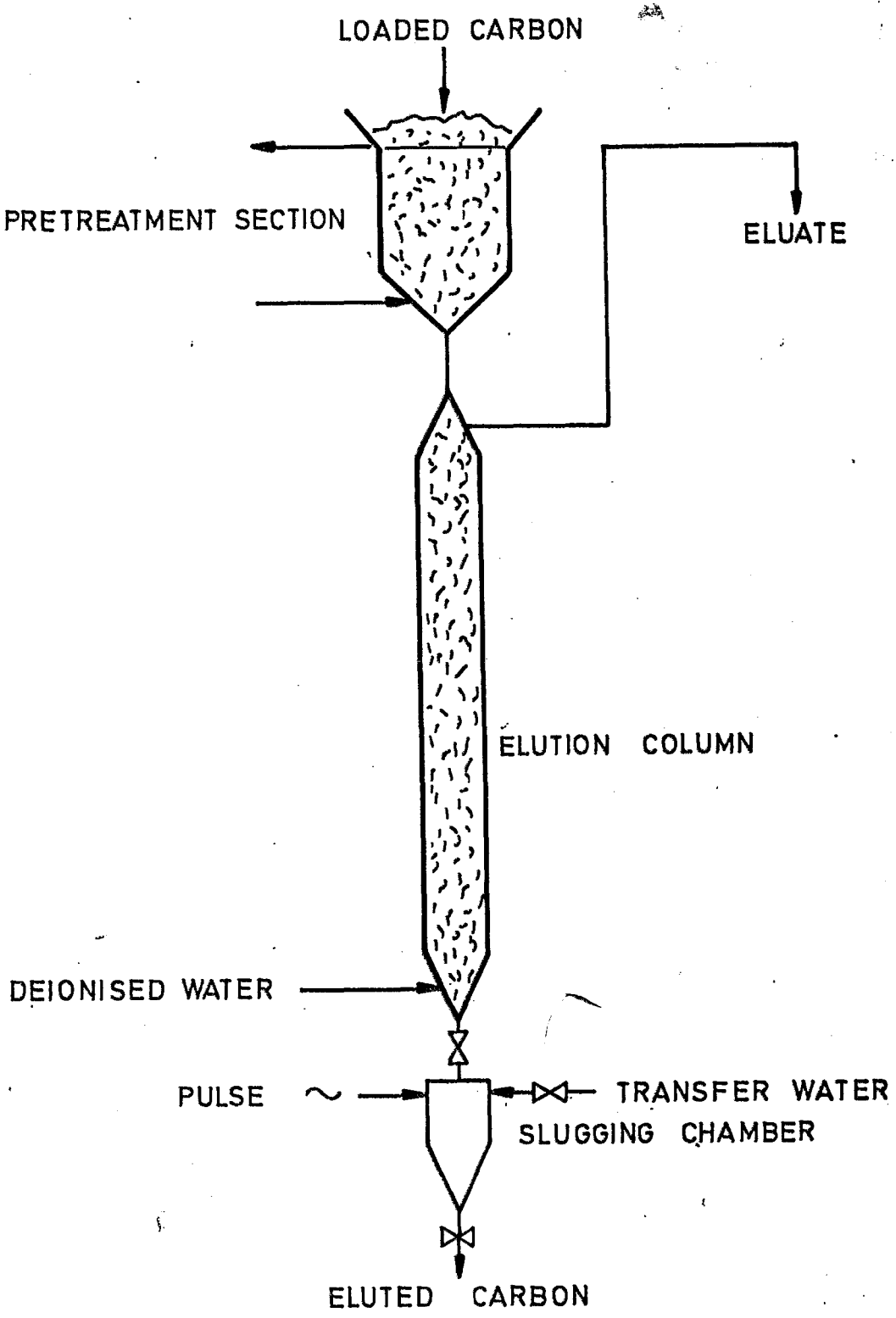
TABLE II

WATER BALANCE DATA FOR THE PILOT PLANT AT GROOTVLEI

Testrun No.	2	3	4	5	6
SOLUTION ENTERING CIRCUIT (l/h):					
Eluant	154	213	213	210	210
Moisture on carbon	29,7	29,7	29,7	22,5	22,5
Displacement from Slugging Chamber	64,1	64,1	64,1	64,1	64,1
Solution from pretreatment	15,3	0,0	0,0	0,0	0,0
Solution from reagent cyanide	2,9	2,4	2,4	1,9	2,2
Solution from reagent caustic	4,6	1,6	1,6	3,1	2,9
TOTAL	270,6	310,8	310,8	301,6	301,7
SOLUTION LEAVING CIRCUIT (l/h):					
Lost as vapour	3,1	3,1	3,1	3,1	3,1
Eluate	207	264	264	250	220
Moisture on carbon*	44,8	44,8	44,8	44,8	44,8
TOTAL	254,9	311,8	311,8	297,8	267,8
DIFFERENCE	15,7	1,0	1,0	3,8	33,9

*50% moisture assumed

FIGURE 1. CONTINUOUS ELUTION FLOWSHEET



APPENDIX I

HEAT BALANCE CALCULATIONS FOR TESTRUNS 3 - 6

Solutions Entering Circuit (l/h)

Eluant	210
Slugging Chamber	64
In Carbon	30
Reagents	5
	309

Solution Leaving Circuit (l/h)

Eluate	261
In Carbon	45
Vapour loss	3
	309

1. HEAT BALANCE FOR ELUTION SECTION

Assuming recovery of 50% of sensible heat in eluate.

Heat lost by eluate = Heat gained by eluant

$0,5 \times 261 \times 4,2\ (88-0) = 210 \times 4,2 \times (TH - 20) = 210 \times 4,2 \times (74,7-20)$

ie. TH = 74,7°C

∴ Eluant heat input required $= 210 \times 4,2 \times (118 - 74,7)$
$= 38\ 191$ kJ/h
$= 38,2$ MJ/h

2. HEAT BALANCE FOR PRETREATMENT SECTION

Heat input = Heat gained by recirculating solution +
Heat gained by reagent solution +
Heat gained by wet carbon.

$= 500 \times 4,2 \times (88 - 83)^*$
$+ 5 \times 4,2 \times (88 - 20)$
$+41 \times 1,05 \times (88 - 20)$
$+30 \times 4,2 \times (88 - 20)$

$= 10\ 500 + 1\ 428 + 2\ 927 + 8\ 568$
$= 23\ 423$ kJ/h
$= 23,4$ MJ/h

* Assume 5° radiation loss

APPENDIX I

HEAT BALANCE CALCULATIONS FOR TESTRUNS 3 - 6

/continued

3. HEAT BALANCE FOR SLUGGING SECTION

Heat input = Heat gained by displacement solution
= 80 x 4,2 x (118 - 20)
= 32 928 kJ/h
= 32,9 MJ/h

4. TOTAL HEAT DEMAND

Total heat demand = Het input to elution section
+ Heat input to pretreatment section
+ Heat input to slugging section

= 38,2 + 23,4 + 32,9 MJ/h
= 94,5 MJ/h
= 2268 MJ/d
= 630 kWh/d

4.3 OTHER ELUTION ASPECTS

Modifications to the AARL batch elution process include the use of organic solvent additions to the carbon pre-treatment stage (SA Patent No. 78/5463) and the specific elution of silver prior to gold elution (SA Patent No. 78/5464). These modifications have as yet not been exploited commercially, although the addition of alcohol in the continuous elution circuit at Grootvlei (4.2) will be investigated shortly.

PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	78/5463
Date of Filing:	August 21, 1979
Name of Applicant:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
Name of Inventors:	1. RAYMOND JOHN DAVIDSON 2. VITTORIO VERONESE
Title of Invention:	"METAL RECOVERY"

~~*Provisional*~~/*Complete Specification as Filed*

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JOHANNESBURG — PRETORIA — DURBAN

Bell Bros H 1690

Patents Form 3

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 78/5463

TITLE OF INVENTION: "METAL RECOVERY"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

THIS invention relates to metal value recovery, in particular to the recovery of gold, silver, nickel or copper metal values from solution.

A method of recovering the above metals from supports on which they are adsorbed as complexes is covered by U.S. patent No 3,970,737. According to this method, the metal values are desorbed after pre-treatment by contacting the support with water of low metal cation concentration. The water preferably has a low-multi-charged cation (e.g. alkaline earth metal cation) concentration. Concentrations less than 300 ppm, preferably less than 100 ppm, are preferred. In other words the water is relatively pure and has a low ionic strength. Suitable waters are for example distilled, deionised and otherwise softened waters.

The above method has particular application to metal values in the form of cyanide ionic complexes, the metal value forming the anionic portion of the complex. Any suitable adsorbent support may be used, carbon, particularly activated carbon, being preferred.

It is also disclosed that when the cation associated with the complex is an alkaline earth metal, the complex is adsorbed on to the support very much more strongly than when the cation is an alkali metal, particularly when the complex is a gold cyanide one. Accordingly, if the cation of the complex is an alkaline earth metal, particularly calcium, the support is preferably pre-treated with an alkali metal salt solution prior to water desorption.

Our S.A. Patent No 76/4204 describes and claims an improved pre-treatment solution selected from the group of an alkali metal cyanide solution,

an alkali metal hydroxide solution and a mixture thereof.

According to the present invention, a pre-treatment reagent for an elution process as described above comprises a mixture of an organic solvent and an alkali metal salt and/or hydroxide solution.

Thus, the invention provides a method of recovering metal values selected from gold, silver, copper and nickel from a carbon support having one or more of these values adsorbed thereon in the form of an alkaline earth metal ionic complex, the metal value forming part of the anionic portion thereof, including the steps of contacting the support with a pre-treatment reagent, followed by desorbing the metal values from the support with water having a low concentration of metal cations characterised in that the pre-treatment reagent is a mixture of:-

(a) an organic solvent, and

(b) a solution selected from the group of an alkali metal cyanide solution, an alkali metal hydroxide solution and a mixture thereof.

The organic solvent may be an alcohol such as isopropyl alcohol or a ketone such as acetone.

Generally, the organic solvent will be present in an amount of from 1 to 10, typically 5, percent by volume of the pre-treatment reagent.

An example of a suitable solution for the reagent is one containing an alkali metal cyanide concentration in the range 1 to 10 percent by weight and an alkali metal hydroxide concentration in the range 0,1 to 20 percent by weight. An example of another suitable solution for the reagent is an alkali metal hydroxide solution of concentration 0,1 to 20 percent by weight.

The pre-treatment will generally be about a half to one bed volume.

The pre-treatment is followed, as mentioned above, by the water elution in the manner described in the above mentioned patents.

It has been found that metal values, particularly gold and silver

values, can be eluted from activated carbon absorbents at high elution rates and using relatively small volumes of eluant. Consequently, the eluant contains a high metal concentration which is advantageous, particularly for subsequent electrowinning processes.

A number of experiments have been carried out using the pre-treatment reagent of the invention. The results thereof are illustrated graphically by the attached Figures 1 to 3:

Figure 1: This graph illustrates the elution of gold and silver from an activated carbon. The carbon was loaded with gold and silver from a cyanide leach solution which had a high base metal content. The loaded carbon contained 1,9 percent by weight gold and 0,5 percent by weight silver. The pre-treatment reagent contained 95 volume percent cyanide solution (5% NaCN/0,5%NaOH) and 5 volume percent isopropyl alcohol. The extent of pre-treatment was one bed volume. The eluant was deionised water and the elution rate was 2 bed volumes per hour.

Figures 2 and 3: These graphs illustrate the elution of gold and silver from an activated carbon which was pre-treated with a number of pre-treatment reagents. The gold and silver was adsorbed on to the carbon from a cyanide leach solution which contained a low concentration of base metals.

The loaded carbon contained about 4 percent by weight gold and about 0,5 percent by weight silver. The pre-treatment was one bed volume. The eluant was deionised water and the elution rate was 2 bed volumes per hour.

On these graphs the lines A,B and C represent the results obtained with the following pre-treatment reagents:

A : 95 vol % of a solution containing 0,5% NaOH and 5% NaCN and 5 vol % of isopropyl alcohol.

B : A solution containing 10%NaCN and 1% NaOH.

C : 95 vol % of a solution containing 2,5% NaCN and 0,25% NaOH and 5 vol % isopropyl alcohol.

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED WE DECLARE THAT WHAT WE CLAIM IS:

1.

A method of recovering metal values selected from gold, silver, copper and nickel from a carbon support having one or more of these values adsorbed thereon in the form of an alkaline earth metal ionic complex, the metal value forming part of the anionic portion thereof, including the steps of contacting the support with a pre-treatment reagent, followed by desorbing the metal values from the support with water having a low concentration of metal cations characterised in that the pre-treatment reagent is a mixture of:-

(a) an organic solvent, and

(b) a solution selected from the group of an alkali metal cyanide solution, an alkali metal hydroxide solution and a mixture thereof.

2.

A method of claim 1 wherein the organic solvent is present in the mixture in an amount of 1 to 10 volume percent.

3.

A method of claim 1 wherein the organic solvent is present in the mixture in an amount of 5 percent by volume.

4.

A method of any one of the preceding claims wherein the organic solvent is an alcohol or a ketone.

5.

A method of claim 4 wherein the alcohol is isopropyl alcohol and the ketone is acetone.

6.
A method according to any one of the preceding claims wherein the solution contains an alkali metal cyanide of concentration in the range 1 to 10 percent by weight and an alkali metal hydroxide of concentration in the range 0,1 to 20 percent by weight.

7.
A method according to any one of claims 1 to 5 wherein the solution is an alkali metal hydroxide solution of concentration 0,1 to 20 percent by weight.

8.
A method according to any one of the preceding claims wherein the carbon support is activated carbon.

9.
A method according to any one of the preceding claims wherein the water having a low concentration of metal cations is deionised, distilled or otherwise softened water.

10.
A method according to claim 1 substantially as herein described with reference to the accompanying graphs.

DATED THIS DAY OF

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APPLICANT'S PATENT ATTORNEYS

Figure 1: Elution of Gold and Silver

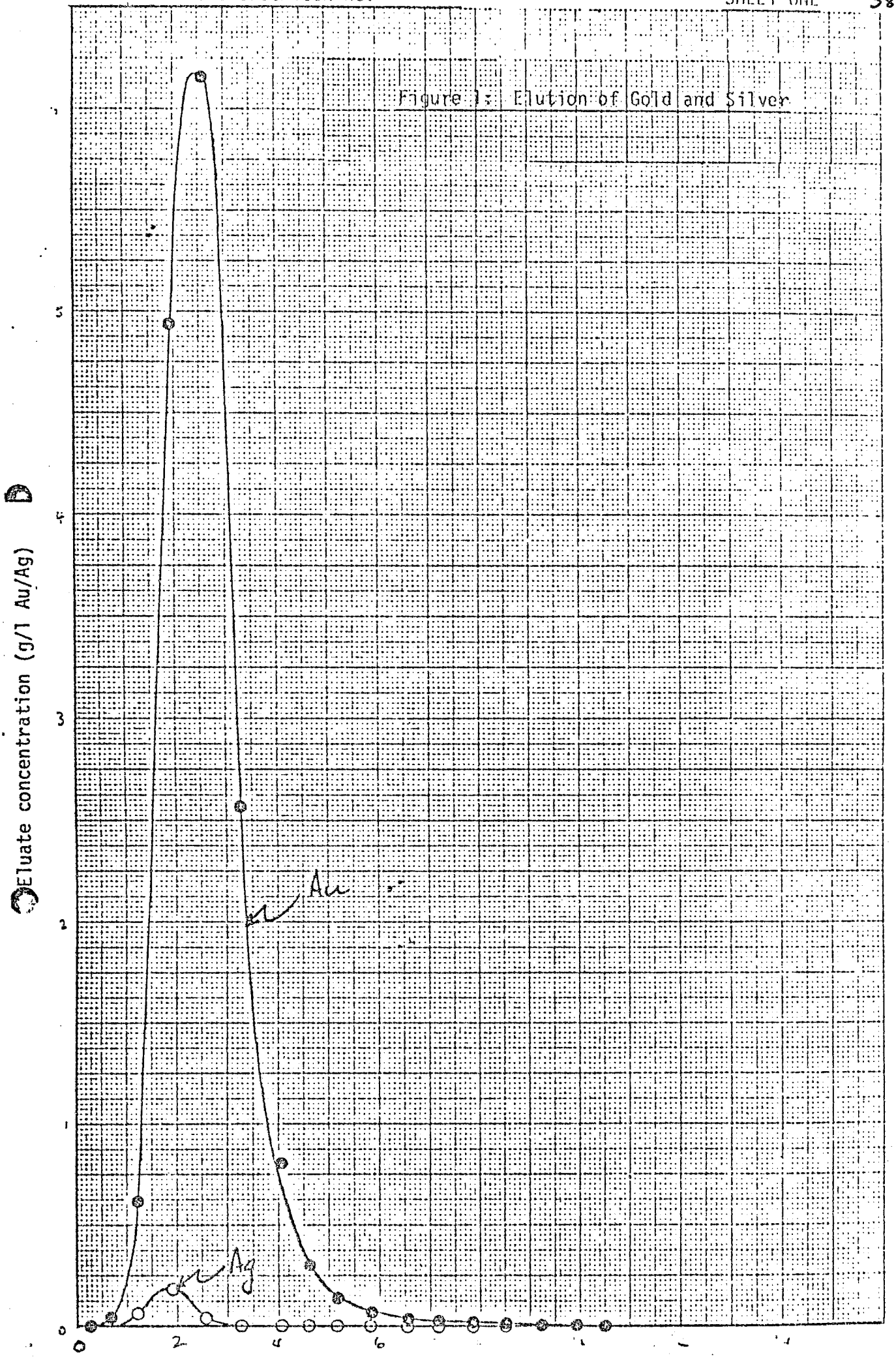


FIGURE 2. The Elution of Gold using additions of Isopropyl Alcohol to the Cyanide Pretreatment Reagent

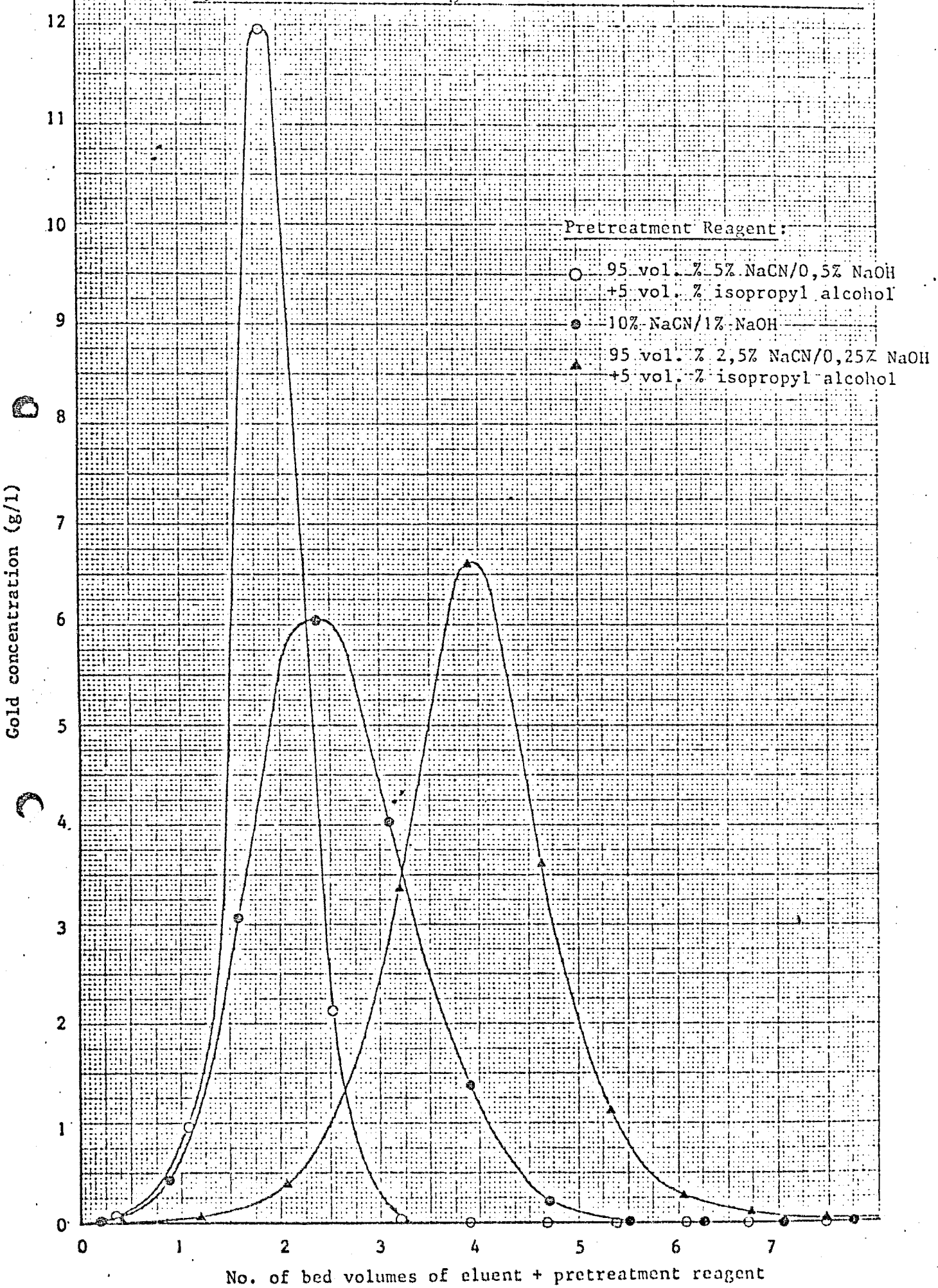
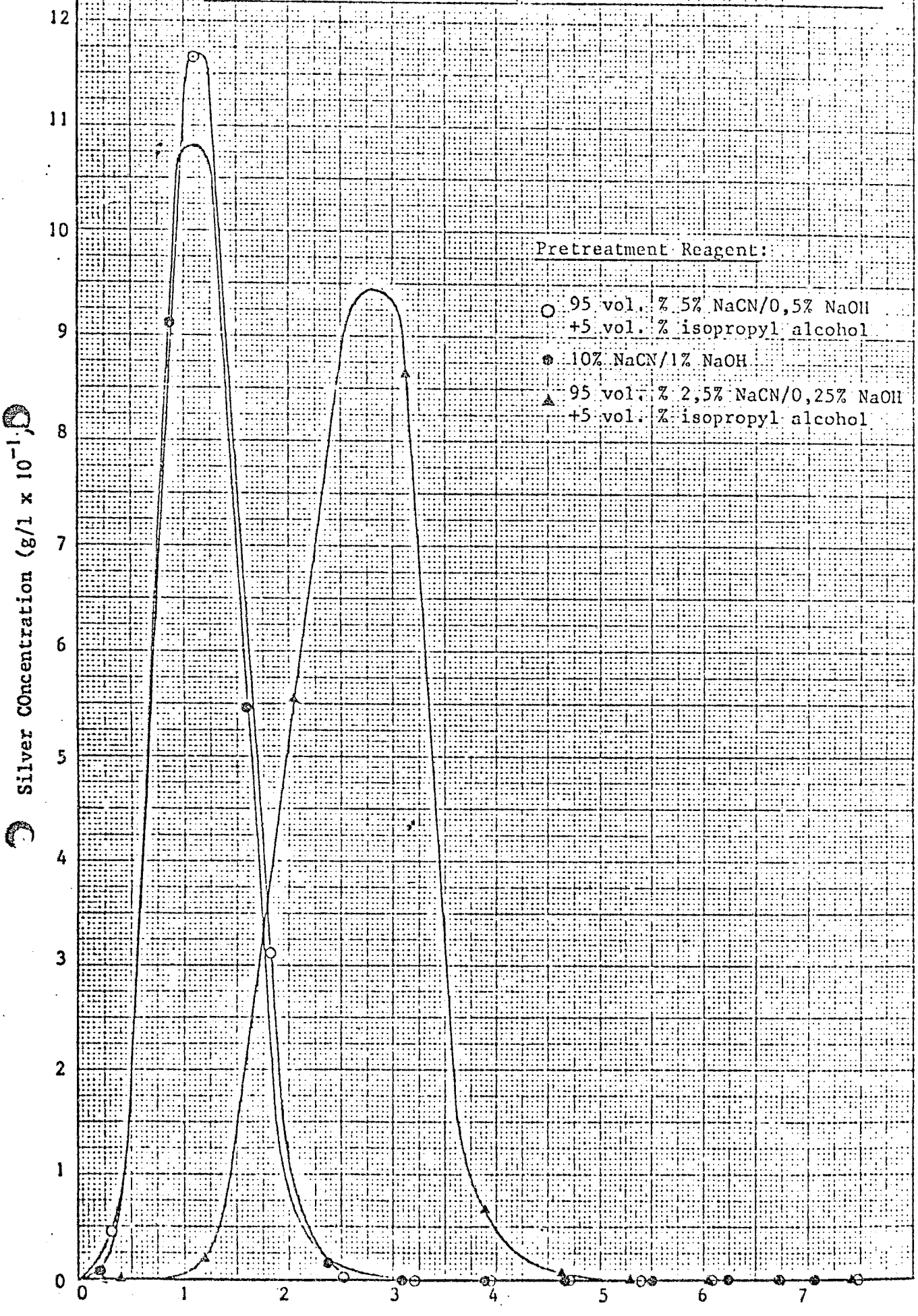


FIGURE 3. The Elution of Silver using additions of Isopropyl Alcohol to the Cyanide Pretreatment Reagent



PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	78/5464
Date of Filing:	AUGUST 21, 1979
Name of Applicant:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
Name of Inventor:s:	1. RAYMOND JOHN DAVIDSON 2. VITTORIO VERONESE
Title of Invention:	"NOBLE METAL RECOVERY"

~~Provisional~~/*Complete Specification as Filed*

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JOHANNESBURG — PRETORIA — DURBAN

SAIB 01 M1490

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 78/5464

TITLE OF INVENTION: "NOBLE METAL RECOVERY"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

XXWE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

THIS invention relates to the preferential separation and recovery of silver and gold from solutions containing these noble metals.

The cyanidation process is a well known process for extracting gold and silver values from gold and silver bearing ores. The process involves finely grinding the ore and then leaching the ore with a cyanide solution. The gold and silver may be recovered from the leach solutions by adsorbing such values on activated carbon and then desorbing or stripping the values from the adsorbent using one of several known methods, for example:

The loaded activated carbon may be pretreated with a hot solution of sodium cyanide and sodium hydroxide and the gold then eluted using pure deionised, distilled or otherwise softened water (R.J. Davidson and D. Duncanson, Journal of the South African Institute of Mining and Metallurgy, volume 77, 1977, page 25). The desorbed gold and silver can then be recovered from the resulting eluate using any conventional technique such as electrowinning.

The above methods have a possible disadvantage that the gold and silver are mostly recovered simultaneously and require further refining in order to separate the contained gold from the silver.

According to the present invention, a method of removing silver from activated carbon having adsorbed thereon gold and silver values includes the step of contacting the carbon with a basic thiosulphate solution. Contact of the laden carbon results in the preferential removal of silver values therefrom leaving substantially all the gold values on the carbon. The gold remaining on the carbon may then be recovered using any of several known methods, for example, the procedure described in the above publication.

The thiosulphate solution which is preferably a sodium thiosulphate solution may have a concentration of 5 - 20 percent by weight, typically 10 percent by weight. The pH of the solution is basic, e.g. pH 10. The pH may be adjusted to a basic pH by addition of a suitable hydroxide such as sodium hydroxide.

It has been found that the removal of silver using the above procedure is best carried out at ambient rather than elevated temperatures. Where other metal values are also adsorbed on the carbon it has been found that the thiosulphate solution simultaneously elutes such values, in particular copper and calcium.

Generally, the laden activated carbon will be in a column and the thiosulphate solution passed through the column to elute the silver values.

The gold and silver values will generally be adsorbed on the carbon in the form of cyanide ionic complexes.

In an example of the invention, granular coconut shell activated carbon was loaded to 3,9 percent gold and 0,3 percent silver by passing a gold plant pregnant solution (i.e. a cyanide solution) through a column containing the activated carbon. The loaded carbon was placed in a water-jacketed elution column of 0,9 cm internal diameter and 40 cm long. The loaded carbon was then treated with 3 bedvolumes of a solution containing 10 weight percent of $Na_2S_2O_3.5H_2O$ and adjusted to pH 10 using sodium hydroxide. Elution was carried out at ambient temperature at a flowrate of ½ bedvolume per hour. The preferential elution of silver is graphically illustrated in Figure 1 where it may be noted that 94,5 percent of the contained silver on the carbon was recovered while 99,7 percent of the contained gold remained on the carbon. To recover the gold the elution column containing the loaded carbon was then raised to 91°C and the carbon pretreated with 1 bedvolume of a solution containing 10 weight percent sodium cyanide and 1 weight percent sodium hydroxide. The carbon was then eluted with deionized water at the same temperature at a flowrate of 2 bedvolumes per hour. The elution of gold and silver from the carbon is graphically illustrated in Figure 2 where it may be noted that >99,9 percent of the remaining gold was recovered. The eluted carbon following the two elution procedures described analysed 20 g/t gold

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

390.

1.
A method of removing silver from activated carbon having adsorbed thereon gold and silver values including the step of contacting the carbon with a basic thiosulphate solution.

2.
A method according to claim 1 wherein the thiosulphate solution has a concentration in the range 5 to 20 percent by weight.

3.
A method according to claim 1 wherein the thiosulphate solution has a concentration of 10 percent by weight.

4.
A method according to any one of the preceding claims wherein the thiosulphate solution is a sodium thiosulphate solution.

5.
A method according to any one of the preceding claims wherein the pH of the thiosulphate solution is 10.

6.
A method according to any one of the preceding claims wherein the silver removal takes place at ambient temperature.

7.
A method according to any one of the preceding claims wherein the gold and silver values are adsorbed on the carbon in the form of cyanide ionic complexes.

8.
A method according to any one of the preceding claims wherein the gold remaining on the activated carbon after silver removal is removed from the carbon by contacting the carbon with deionised or otherwise softened water.

9.

A method of claim 1 substantially as herein described with reference to the attached graphs.

DATED THIS 21ST DAY OF AUGUST, 1979.

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APPLICANT'S PATENT ATTORNEYS

FIGURE 1 THE ELUTION OF SILVER

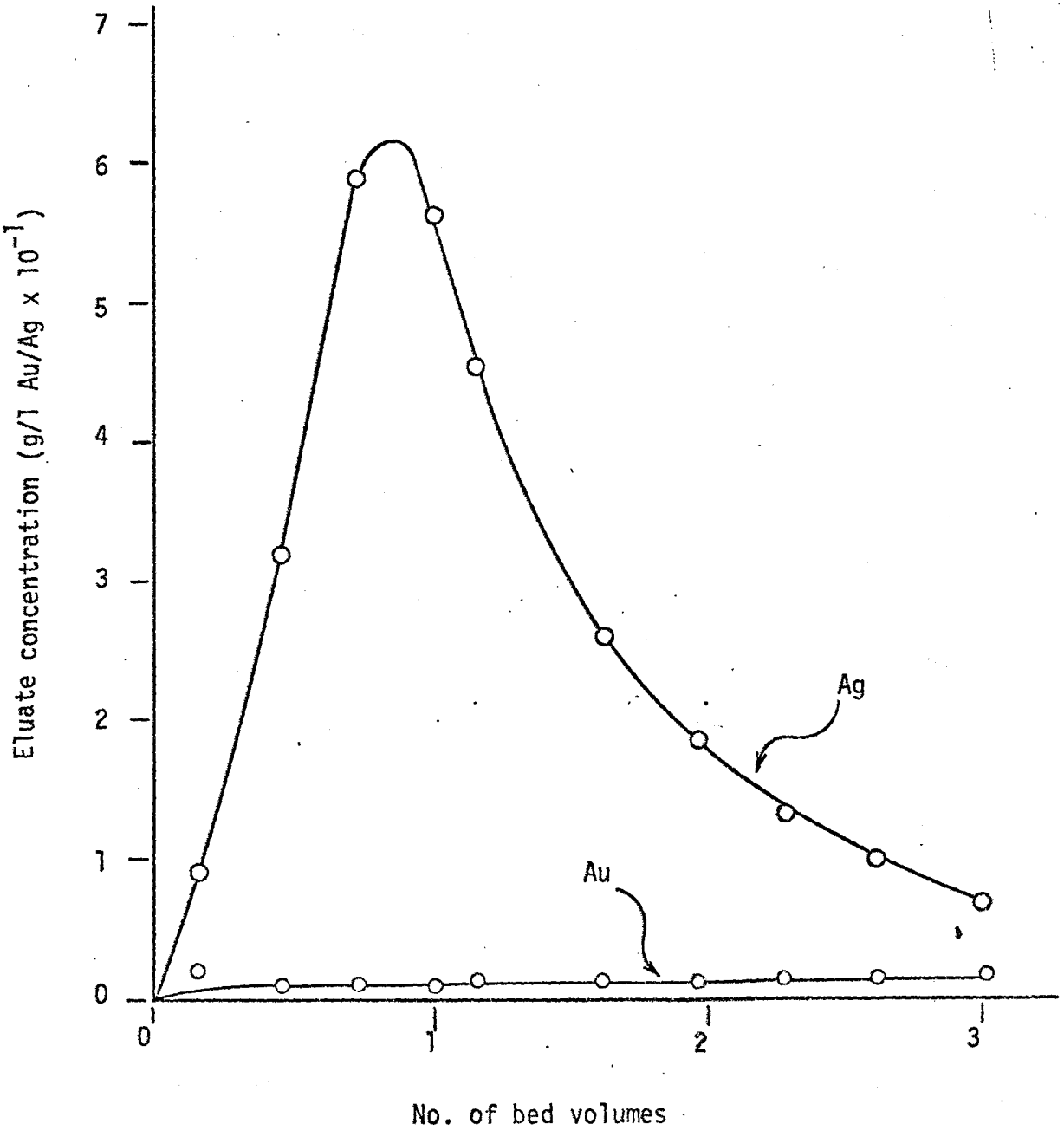

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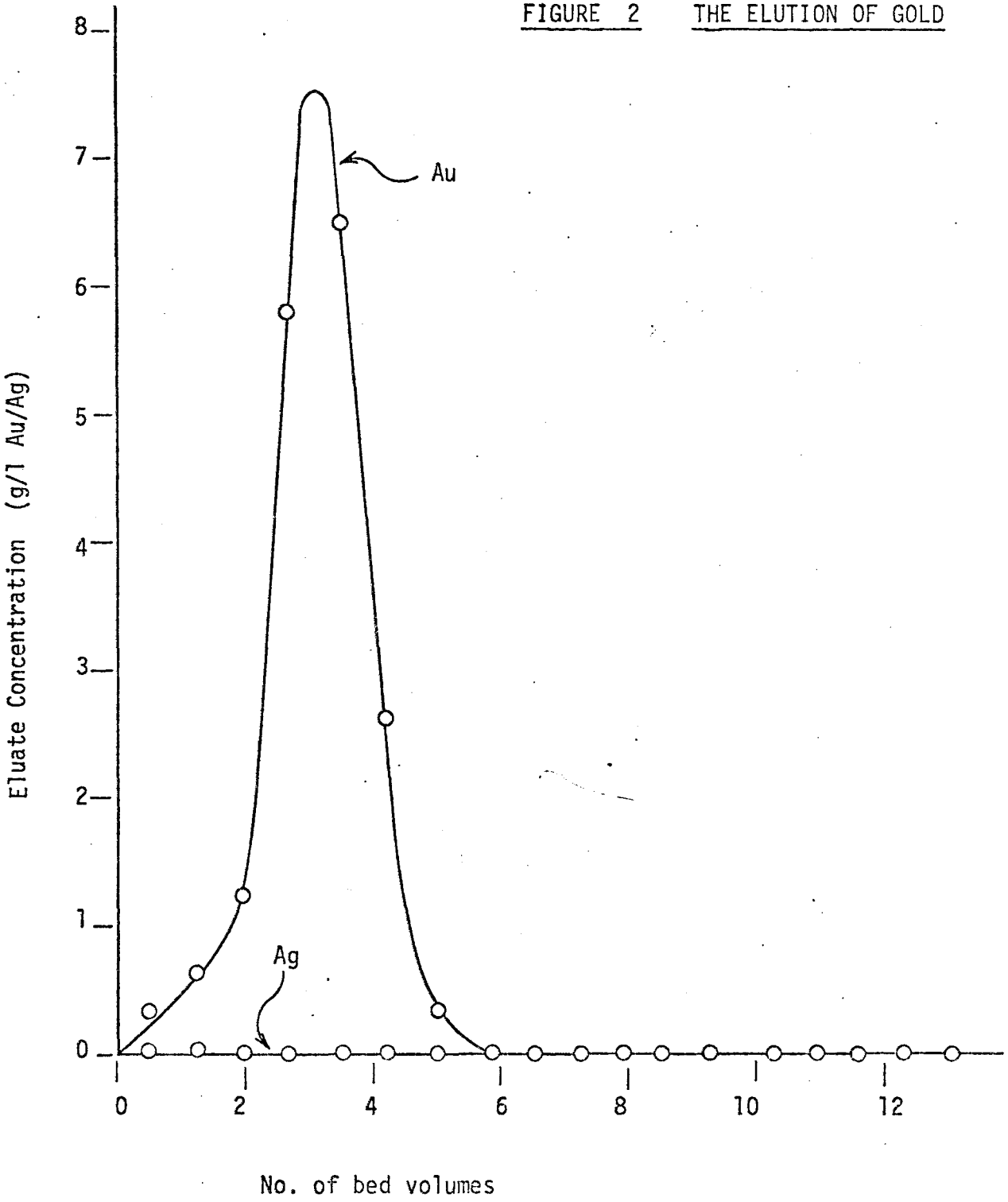
FIGURE 2 THE ELUTION OF GOLD



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4.4 TREATMENT OF ELUATES

The acidification of eluates was investigated as a means of recovering gold (SA Provisional Patent Application No. 81/8894) as well as the recovery of free cyanide and base metal values (SA Patent No. 90/8081). These investigations have establised a processing route which is currently being piloted at ERGO where the cost effective recovery of free cyanide and copper sulphate have been demonstrated.

REPUBLIC OF SOUTH AFRICA
PATENTS ACT, 1978

PROVISIONAL SPECIFICATION

(Section 30 (1) – Regulation 27)

OFFICIAL APPLICATION NO.		LODGING DATE	
21 01	81/8894	22	23 DECEMBER 1981

FULL NAME(S) OF APPLICANT(S)

71	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

FULL NAME(S) OF INVENTOR(S)

72	1. PAUL FREDERICK SORENSEN 2. RAYMOND JOHN DAVIDSON

TITLE OF INVENTION

54	"GOLD RECOVERY"

BACKGROUND TO THE INVENTION

THIS invention relates to gold recovery.

The most commonly used method of solubilising gold from its ores is the cyanidation process. This process involves finely grinding the ore and then leaching the ore with a sodium cyanide solution. The gold may be recovered from the leach solutions using a variety of methods, e.g. zinc cementation, mercury amalgamation.

Where the gold concentration in solution is commonly low as above, the gold can be concentrated using activated carbon. This process involves adsorbing the low gold values on to activated carbon and then desorbing the gold therefrom using sodium cyanide solution followed by washing with deionised or otherwise softened water. Typically, the gold is recovered from gold-rich eluants so obtained using electrowinning techniques. Contact between the gold solution and the carbon may take place in an adsorption column or in pulp medium (carbon-in-pulp).

While electrowinning gold from gold-rich solutions is satisfactory, it does have a number of drawbacks. First, it is a complex method. Second, it is time consuming and has to be run at night which gives rise to security problems.

SUMMARY OF THE INVENTION

According to the present invention, there is provided a method of recovering gold values from a non-acidic gold-bearing solution including the step of reducing the pH of the solution to below 6 in the presence of a base metal ion and free cyanide ion (CN^-). This causes the gold to report in a precipitate which may be removed from the supernatant liquor.

DESCRIPTION OF THE PREFERRED EMBODIMENTS

The gold-bearing solution typically has a pH of 9 or higher and free cyanide concentration sufficient to complex with the metal values. Such solutions are typically those which emanate from intensive cyanidation processes, carbon-in-pulp plants, and carbon desorption columns.

The concentration of gold in the gold-bearing solution may be high or low. Typically, it will be in the range 10 mg/l to 2500 mg/l.

The pH of the solution will generally be lowered to below 5, typically 2 or 1, to cause the gold-bearing precipitate to form. The pH may be lowered by adding a suitable mineral acid such as sulphuric acid or hydrochloric acid.

A base metal ion must be present to enable the precipitate to form. The preferred base metal is copper. Base metal ions are frequently present in the gold bearing solutions. If they are not present, they may be added in the form of their soluble salts.

The precipitation may be carried out at ambient or elevated temperature. Generally, the temperature will not exceed 75°C.

The settling of the precipitate may be improved by adding a suitable flocculating agent. Typically, the flocculant will be a non-ionic one such as that sold under the trade name Cyanamid N 300. Typically, such flocculant may be added in an amount of 0 - 1 mg/l.

The precipitate may be removed from the supernatant liquor using methods well known in the art. An example of a suitable method is the use of a settling cone followed by centrifuging.

The gold -rich precipitate may be dissolved in a suitable medium in smaller volume than that of the original solution and by so doing, improve the gold concentration by up to five times. The gold may then be recovered using conventional electrowinning techniques on a proportionally reduced scale.

Alternatively, and preferably, the gold is recovered from the precipitate using the smelting method described hereinafter.

The smelting method includes the steps of smelting the precipitate in the presence of an oxidant to produce a gold-rich metal phase and a slag phase and separating the metal phase from the slag phase. Prior to smelting the precipitate may be calcined in the presence of a suitable fluxing agent such as borax or separately, the borax being added after calcining. Calcining must take place below the melting temperature of the precipitate/fluxing agent combination. Calcining typically takes place in the presence of oxygen (in the form of air, for example) at a temperature in the range 150 to 500°C. Where smelting takes place in the presence of a strong oxidising agent such as a nitrate calcining has been found not to be necessary.

Two examples of the invention will now be described.

EXAMPLE 1

A gold-rich solution from a cyanidation plant was found to contain the following amounts of dissolved gold, silver and copper and free cyanide.

Gold	-	2200 mg/l
Silver	-	224 mg/l
Copper	-	2500 mg/l
Free Cyanide as CN^-	-	5400 mg/l

The pH of the solution was 10.

The solution was put into a settling cone and concentrated sulphuric acid added thereto to reduce the pH to 1. When the pH of the solution reached about 5 a precipitate began to form. The solution was maintained at a temperature of about 50°C during the sulphuric acid addition. At this temperature it was found that an easily separable precipitate formed. At lower temperatures, the addition of a flocculating agent is preferred to ensure that an easily separable precipitate forms.

The precipitate settled to the bottom of the cone and this fraction amounting to some 20 percent by volume of the original volume was separated from the supernatant liquor by tapping from the bottom of the cone. This fraction was separated further in a centrifuge, pressure or vacuum filter and the supernatant liquor recycled to the settling cone.

Borax was added to the precipitate in the proportion of 3 mass units of borax for every mass unit of dried precipitate. A paste resulted which was calcined at 200°C for a period of one hour in air.

The calcined product was smelted in a furnace at a temperature above 1063°C to produce a gold-rich metal phase and a slag phase. The slag phase was removed from the gold-rich metal phase which was found to have the following analysis (on a mass basis):

Gold	-	94 percent
Silver	-	4 percent
Copper	-	2 percent

The supernatant liquor from the settling cone may be contacted with activated carbon to remove any residual gold, silver and base metal values which did not report in the precipitate.

EXAMPLE 2

The same procedure as in example 1 was followed except potassium nitrate was added to the gold-rich precipitate in an amount of one mass unit to every one mass unit of precipitate. Smelting of the combination took place without prior calcining. A gold-rich metal phase and a slag phase again formed. The gold-rich metal phase had the same analysis as in Example 1.

DATED this 23rd day of DECEMBER 1981

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APPLICANT'S PATENT ATTORNEYS

SPOOR AND FISHER

JOHANNESBURG

PATENT SPECIFICATION

COUNTRY:	SOUTH AFRICA
APPLICATION NUMBER:	90/8081
DATE OF FILING:	9 OCTOBER 1990
NAME OF APPLICANT:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
NAMES OF INVENTORS:	RAYMOND JOHN DAVIDSON DAVID ERNEST MILLINGTON
TITLE OF INVENTION:	"METHOD FOR THE RECOVERY OF METAL METAL VALUES"

COMPLETE SPECIFICATION AS FILED

FILE REF: JP/A 663
DATE: 24 OCTOBER 1990

BACKGROUND OF THE INVENTION

This invention relates to a method of recovering copper and cyanide values from an eluate containing complexed copper values and complexed cyanide values.

It is a common process to recover gold values by adsorbing the gold values on to an activated carbon or a resin and then eluting the gold values therefrom. In this process, copper values may also be adsorbed and eluted. There is a need for a process for the recovery of such copper values and optionally also for the recovery of cyanide values, as well as the recovery of the gold values.

Various processes for the recovery of the copper and/or cyanide values are known and these are reviewed in an article entitled "Cyanide destruction versus cyanide regeneration - evaluation of the processes for optimum mill effluent treatment", Piret N.L. and Schippers H.J., Extraction Metallurgy '89 Int. Symp., organised by the I.M.M., London, July 1989, pp 1041-1081.

SUMMARY OF THE INVENTION

According to the invention there is provided a method of recovering copper and cyanide values from an eluate containing complexed copper values and complexed cyanide values including the steps of:

(a) acidifying the eluate with a suitable acid to produce a first precipitate comprising a major portion of the copper values as a copper cyanide compound and a first liquid phase containing a major portion of the cyanide values in soluble form;

(b) treating the first precipitate with a suitable agent to convert the copper cyanide compound to a copper oxide or a copper hydroxide or a mixture thereof to produce a second precipitate comprising the copper oxide or the copper hydroxide or the mixture thereof and a second liquid phase containing a copper cyanide complex; and

(c) treating the second precipitate with sulphuric acid to produce copper sulphate.

The eluate may also contain gold values. In this case, in step (a) the acidification of the eluate with a suitable acid produces a first precipitate comprising a major portion of the copper values and a major portion of the gold values as copper cyanide compounds. Then in step (b) the major portion of the gold values report in the second liquid phase containing the copper cyanide complex. The gold values may then be recovered in a known manner from the second liquid phase.

Preferably, the method of the invention includes the step of:

(d) recycling the second liquid phase from step (b) to step (a).

Preferably, the method of the invention also includes the step of:

(e) treating the first liquid phase with a neutralising agent to recover the free cyanide values.

Preferably, the method of the invention also includes the step of:

(f) recycling the liquid phase from step (c) after removal of the copper sulphate to step (b), to ensure complete recovery of the copper values and the gold values if present.

In step (a), the suitable acid may be for example sulphuric acid or hydrochloric acid. Preferably, the eluate is acidified to a pH of about 2.

In step (b), the suitable agent to convert the copper cyanide salt to a copper oxide, e.g. CuO (tenorite) or a copper hydroxide, e.g. $Cu(OH)_2$ (spertiniite) may be caustic soda. Preferably, during the course of step (b) the mixture is sparged with air or oxygen or mixtures of sulphur dioxide.

In step (e), the neutralising agent may be for example caustic soda or lime.

The eluate may be an eluate from the elution of an activated carbon or a resin loaded with copper values and optionally also gold values. Thus, the eluate may contain substantially only copper values or may contain copper and gold values.

BRIEF DESCRIPTION OF THE DRAWING

Figure 1 is a flow diagram of an embodiment of the method of the invention.

DESCRIPTION OF EMBODIMENTS

This invention relates to a method for the recovery of copper as copper sulphate and for the recovery of cyanide as free cyanide from an eluate containing complexed copper values and complexed cyanide values. The copper values and free cyanide values are recovered as valuable by-products.

The method of the invention is described in more detail below in relation to certain experiments and the results thereof.

EXPERIMENTS AND RESULTS

1. Eluate preparation

Copper-rich eluates were prepared in order to investigate the subsequent treatment of a cuprocyanide precipitate obtained following acidification of the eluates.

1.1 CIL barren eluate

CIL pregnant eluate was pumped through a carbon moving bed pilot plant at a gold plant in order selectively to remove the contained gold values while not adsorbing copper values. A bulk sample of the column effluent containing 1,2 g/ℓ Cu; 4,4 g/ℓ total NaCN and pH 10,2 was used for subsequent acidification testwork.

1.2 Copper elution at ambient temperature

Loaded carbon from a CIL plant containing 25 kg/t Cu was initially acid washed at ambient temperature with equal volumes of 3 vol. % hydrochloric acid. After decant washing, the carbon was adjusted to pH 12 using caustic soda before eluting with sodium cyanide as follows:

1 ℓ samples of wet settled carbon (640 g/ℓ bulk density) were contacted with 1 ℓ 3% sodium cyanide for 18 hours in a rolling bottle. The copper-rich eluate was then drained from the carbon before preparing a 13 ℓ bulk eluate composite. The bulk composite eluate containing 8,6 g/ℓ Cu; 22,0 g/ℓ

total NaCN and pH 10,5 was used for subsequent acidification and cuprocyanide precipitate preparation.

2. Eluate acidification and cuprocyanide precipitate preparation

5 ℓ aliquots of eluate were acidified to pH 2 with 50% sulphuric acid. Following acidification, stirring was continued for 5 minutes before adding 7,5 mℓ 0,1% N101 Superfloc i.e. 1,5 g/m^3. After allowing for flocculation for 3 minutes the cuprocyanide precipitate was allowed to settle for 10 minutes while monitoring settling rates. The clear supernatant solution (ca. 4 ℓ) was then pumped out of the acidification vessel and made alkaline using caustic soda. ca.4 ℓ wash water was then added to the precipitate which was again stirred and settled (3 minutes) before pumping out the wash solution as before. Following vacuum filtration and washing, the precipitate was oven dried and weighed.

The acidification of the CIL barren eluate containing 1,2 g/ℓ Cu and 4,4 g/ℓ total NaCN required 8,1 kg/m^3 sulphuric acid and produced 2,70 kg/m^3 of dried precipitate. Analysis of the clear supernatant solution following acidification and neutralising as described above indicated the precipitation of 97,0% of the contained copper, while 68% of the total cyanide in the eluate was recovered in the supernatant solution as free cyanide.

In the case of the composite eluate prepared at ambient temperature which contained 8,6 g/ℓ Cu and 22,0 g/ℓ total NaCN, acidification required 19,6 kg/m^3 sulphuric acid and produced 12,02 kg/m^3 of dried precipitate. Analysis of the neutralised supernatant solution indicated the precipitation of 98,8% of the contained copper, while 68% of the total cyanide in the eluate was recovered in the

supernatant solution as free cyanide.

3. Treatment of CuCN Precipitate

Using pure CuCN (71% Cu; 29% CN) as the starting material, it may be seen that 1 tonne could produce 2790 kg $CuSO_4.5H_2O$ and 522 kg NaCN (100%). In the case of the CIL barren eluate the precipitate analysed 62% Cu and 24% CN indicating a CuCN content of ca. 85%. Analysis of the precipitate prepared from the ambient temperature eluate was 72% Cu and 26% CN indicating a relatively pure compound.

CIL barren eluate precipitate

Samples of precipitate prepared following acidification of the eluate and containing 62% Cu and 24% CN were treated with caustic soda as follows:

(a) 2 g precipitate and 2 g NaOH in 40 mℓ water was boiled for 1 hour while sparging with excess oxygen. After making up to volume (200 mℓ), analysis indicated the dissolution of 33% Cu, while the black precipitate remaining contained 62% Cu and 0,87% CN. This precipitate was identified by X-ray diffraction as being tenorite (CuO) and was soluble in dilute sulphuric acid.

(b) 10 g precipitate and 5 g NaOH in 40 mℓ water was agitated for 2 hours at ambient temperature while sparging with excess oxygen. After making up to volume analysis indicated the dissolution of 32% Cu, while the green precipitate was identified as spertiniite $Cu(OH)_2$.

With regard to the above investigation, the rather consistent dissolution of ca. 33% copper as a cyanide complex suggests the following reactions:

(a) $3CuCN + 2NaOH\ (hot) + \frac{1}{2}O_2 \rightarrow Na_2Cu(CN)_3 + 2CuO$ (tenorite)

(b) $3CuCN + 2NaOH\ (cold) + \frac{1}{2}O_2 + H_2O \rightarrow Na_2Cu(CN)_3 + 2Cu(OH)_2$ (spertiniite)

Ambient temperature eluate precipitate

5 g samples of the CuCN precipitate and 2,5 g NaOH in 200 mℓ water were agitated for various time periods while sparging the mixture with air. Following vacuum filtration, solution samples were taken to assess copper dissolution and the oven dried washed precipitate remaining analysed for total cyanide content.

From the results presented in Table 1, it may be again seen that following a reaction time of 3 hours, 32-33% of the total copper content containing 97-99% of the total NaCN content is present in the filtrate. The balance of the copper was contained in the hydroxide precipitate which was soluble in dilute sulphuric acid and contained less than 1% CN and analysed 66% copper ($Cu(OH)_2$ contains 65,1% Cu). These findings would support the above proposed reaction mechanism.

TABLE 1

THE CAUSTIC DIGESTION/OXIDATION OF THE CuCN PRECIPITATE AT ca 20°C

Reaction Time (h)	Weight Loss (%)	Copper Dissolution (%)	CN Dissolution (%)	CN Content of precipitate (%)
0	0	0	0	26
0,5	40	28	77	5,9
1	39	29	88	3,1
2	39	28	95	1,2
3	36	34	99	0,3
4	32	32	97	0,8
5	36	31	99	0,3

4. Acidification of caustic filtrate

In order to further substantiate the proposed reaction mechanism:

$$3CuCN + 2NaOH + \tfrac{1}{2}O_2 + H_2O \rightarrow \underset{\text{(filtrate)}}{Na_2Cu(CN)_3} + 2Cu(OH)_2$$

acidification of the filtrate obtained following the caustic digestion of the cuprocyanide precipitate was carried out as follows:

35 g of cuprocyanide precipitate prepared from the ambient temperature eluate and containing 72% Cu and 25% CN was digested with 13 g caustic soda in 3 ℓ water for 4 hours while sparging with air. Following vacuum filtration and washing, the filtrate which contained 2,5 g/ℓ Cu and 2,9 g/ℓ total CN was acidified to pH 2 as previously described. This acidification yielded 11,8 g of precipitate (theoretical yield 10,6 g CuCN) which analysed 71% Cu and 27% CN (CuCN contains 72% Cu and 29% CN). Analysis of the supernatant solution following acification indicated that ca. 60% of the total cyanide was liberated in solution as free cyanide, while precipitating ca. 100% of the contained copper from the original filtrate. The material balance of copper and cyanide (total and free) confirm the reaction:

$$\underset{\text{(filtrate)}}{Na_2Cu(CN)_3} + H_2SO_4 \rightarrow CuCN + 2HCN + Na_2SO_4$$

Referring to the drawing, there is illustrated one embodiment of the method of the invention in which gold, copper and cyanide values are recovered from an eluate containing the values.

There is fed to stage I the eluate containing complexed copper, gold and cyanide values which eluate is acidified with sulphuric acid. There is also fed in to stage I the recycled second liquid phase from the stage II reaction, containing the complexed gold values and the undigested complexed copper values. The material is fed from stage I to a decant - wash stage where a first liquid

phase containing a major portion of the cyanide values in soluble form is separated from the solid phase. The solid phase or precipitate containing the copper values and the gold values as copper cyanide compounds is fed to stage II and treated with a suitable agent such as caustic soda to convert the copper cyanide compounds to copper oxide compounds. This mixture is filtered with the solid precipitate containing the copper oxide being fed to the stage III reaction and with the second liquid phase containing the complexed gold values and some complexed copper values being recycled to stage I. In stage III, the second precipitate is treated with sulphuric acid to produce copper sulphate which is recovered, and a further liquid phase containing undigested copper cyanide/gold cyanide values which is recycled to stage II. In a neutralization stage the first liquid phase from stage I containing the cyanide values is treated with caustic soda to recover the cyanide values. The gold values may be recovered from time to time by a conventional process such as adsorption on to activated carbon or a resin.

From the above experimental procedures and results, it can be determined that the method of the invention preferably includes the following steps:

1. An eluate containing complexed copper values and complexed cyanide values is acidified with a suitable acid, preferably sulphuric acid to produce a first precipitate comprising a major portion of the copper values, preferably up to 99% of the copper values, as a copper cyanide salt, CuCN, and a first liquid phase containing a major portion, preferably up to 70%, of the cyanide values in soluble form as HCN.
2. Following decant washing, the first precipitate is oxidised/digested with a suitable agent, preferably caustic soda, to produce a second precipitate comprising a copper oxide or a copper hydroxide or a mixture thereof and a second liquid phase containing a copper cyanide

complex, $Cu(CN)_3$.

3. The second liquid phase containing the copper cyanide complex is recirculated to step 1 so allowing for the complete recovery of both free cyanide and copper sulphate.
4. The second precipitate is treated with sulphuric acid to produce copper sulphate which is recovered.
5. The first liquid phase is treated with a neutralising agent such as lime or caustic soda to produce NaCN which is recovered.

The method of the invention provides a simple but cost effective means of recovering free cyanide while also producing copper sulphate as a profitable by-product.

CLAIMS

1.

A method of recovering copper and cyanide values from an eluate containing complexed copper values and complexed cyanide values includes the steps of:

(a) acidifying the eluate with a suitable acid to produce a first precipitate comprising a major portion of the copper values as a copper cyanide compound and a first liquid phase containing a major portion of the cyanide values in soluble form;

(b) treating the first precipitate with a suitable agent to convert the copper cyanide compound to a copper oxide or a copper hydroxide or a mixture thereof to produce a second precipitate comprising the copper oxide or the copper hydroxide or the mixture thereof and a second liquid phase containing a copper cyanide complex; and

(c) treating the second precipitate with sulphuric acid to produce copper sulphate.

2.

A method according to claim 1 wherein the eluate also contains gold values and wherein in step (a) the acidification of the eluate with a suitable acid produces a first precipitate comprising a major portion of the copper values and a major portion of the gold values as copper cyanide compounds and a first liquid phase containing a major portion of the cyanide values in soluble form, and wherein in step (b) when the first precipitate is treated with a suitable agent to convert the copper cyanide compounds to a copper oxide or a copper hydroxide or a mixture thereof to produce a second precipitate comprising the copper oxide or the copper hydroxide or the mixture thereof and a second liquid phase, a major portion of the gold values report in the second liquid phase.

3.
A method according to claim 1 or claim 2 which includes the step of:

(d) recycling the second liquid phase from step (b) to step (a).

4.
A method according to any of claims 1 to 3 which includes the step of:

(e) treating the first liquid phase with a neutralizing agent to recover the free cyanide values.

5.
A method according to any one of claims 1 to 4 which includes the step of:

(f) recycling the liquid phase from step (c) after removal of the copper sulphate to step (b).

6.
A method according to any one of claims 1 to 5 wherein in step (a) the suitable acid is selected from sulphuric acid and hydrochloric acid.

7.
A method according to claim 6 wherein in step (a) the eluate is acidified to a pH of about 2.

8.
A method according to any one of claims 1 to 7 wherein in step (b) the suitable agent to convert the copper cyanide salt to a copper oxide or a copper hydroxide is caustic soda.

9.
A method according to claim 8 wherein in step (b) the mixture is sparged with air or oxygen or mixtures of sulphur dioxide.

10.
A method according to any one of claims 1 to 9 wherein in step (e) the neutralizing agent is selected from caustic soda or lime.

11.
A method according to claim 1 substantially as herein described.

DATED this 9th day of OCTOBER 1990

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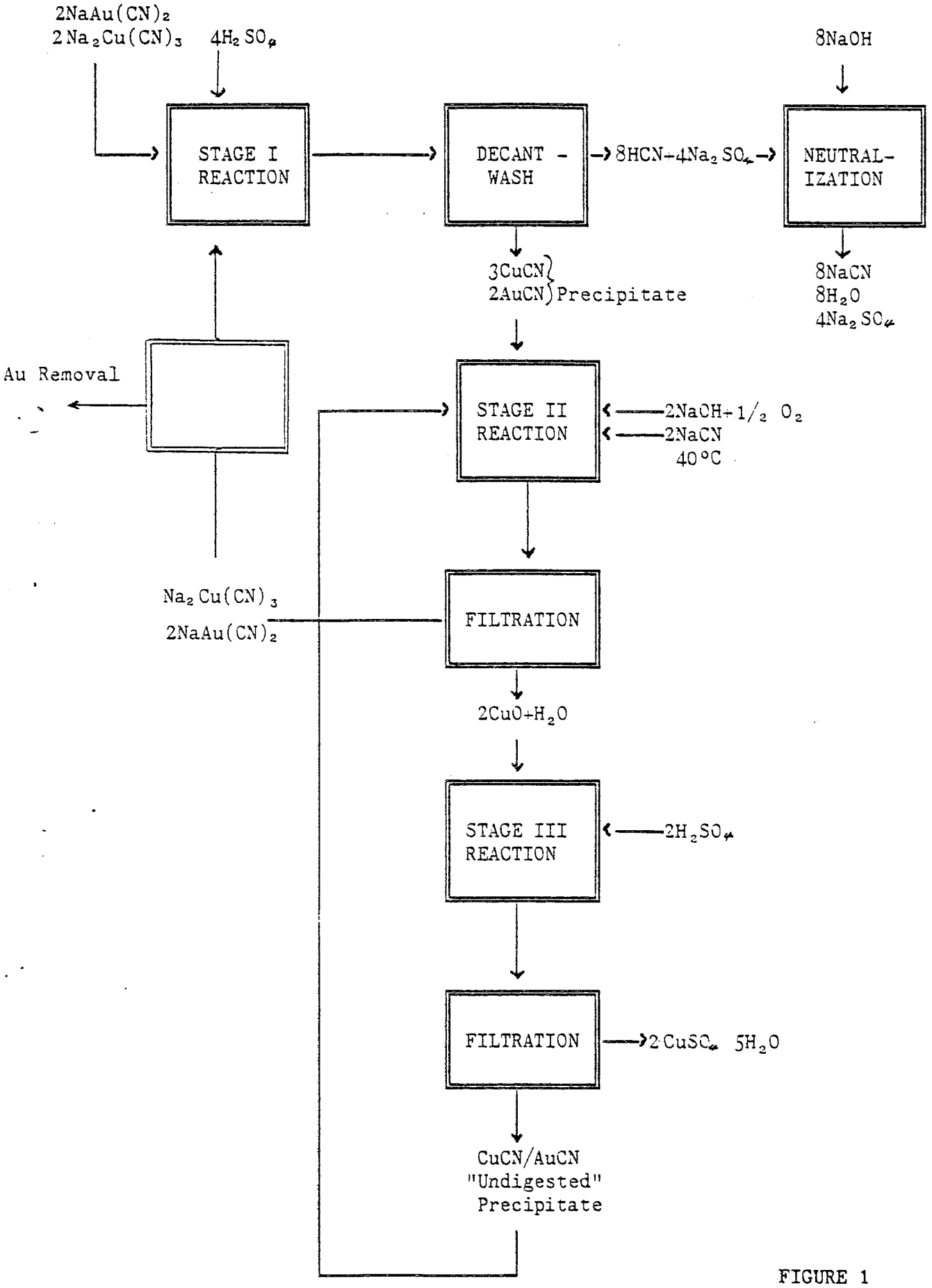
2NaAu(CN)2
2Na2Cu(CN)3
4H2SO4
8NaOH
STAGE I REACTION
DECANT - WASH
8HCN+4Na2SO4
NEUTRAL-IZATION
3CuCN
2AuCN
Precipitate
8NaCN
8H2O
4Na2SO4
Au Removal
STAGE II REACTION
2NaOH+1/2 O2
2NaCN
40°C
Na2Cu(CN)3
2NaAu(CN)2
FILTRATION
2CuO+H2O
STAGE III REACTION
2H2SO4
FILTRATION
2CuSO4 5H2O
CuCN/AuCN
"Undigested"
Precipitate

FIGURE 1

5.0 CARBON ACTIVATION/REGENERATION

Because carbon activity and carbon regeneration play such an important role in carbon based circuits, investigations into activation/regeneration other than by the normal means was studied at the AARL. As a result of these investigations several patents were issued.

a) Anodic activation/regeneration (SA Patent Nos. 77/4285 and 77/4287) appeared to have some potential but as yet has not been commercially exploited.

b) While the use of copper sulphate under acid conditions (SA Patent No. 78/5864) was shown to regenerate used carbon, the claims made at this early stage were not correctly interpreted as the chemistry involved was not understood. Later plant investigations (3.0) have however shown that intermittent acid washing (SA Patent No. 78/5147) results in the formation of the insoluble AuCNCuCN complex which effectively allows for further gold adorption well above any equilibrium constraints eg treatment of dam return effluent (3.4).

c) A provisional patent on the use of ammonium chloride as a solvent for adsorbed organic material was allowed to lapse.

PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 77/4285

Date of Filing: 8TH JUNE, 1978

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventors:
1. RAYMOND JOHN DAVIDSON
2. VITTORIO VERONESE

Title of Invention: "ACTIVATION OF CARBON"

~~Provisional~~/*Complete Specification as Filed*

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 77/4285

TITLE OF INVENTION: "ACTIVATION OF CARBON"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

X I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

FIELD OF THE INVENTION

This invention relates to a process for increasing the adsorption activity of carbon. The process of the invention is concerned not only with activating virgin carbon but also with increasing the activity of an activated carbon.

DESCRIPTION OF PRIOR ART

The adsorption capabilities of activated carbon have wide application in industry. Activated carbon is used for purifying many industrial effluents and is also used extensively in the sugar industry. In the field of metallurgy, activated carbon is used for efficiently extracting valuable metals such as gold and silver from solution. During use, activated carbon loses its effective activity. Deactivated carbon is usually reactivated thermally at temperatures of the order of 500 to 900°C. Thermal treatment is also used to produce activated carbon from virgin carbon.

It is an object of the invention to provide a process which is capable of not only increasing the activity of an activated carbon but also of producing an activated carbon from virgin carbon.

SUMMARY OF THE INVENTION

According to the present invention there is provided a process for increasing the adsorption activity (hereinafter referred

to simply as "activity") of carbon comprising the steps of providing an acidic electrolyte having an inert anode and cathode immersed therein, locating the carbon in the electrolyte in the region of the anode, and applying an effective potential across the electrodes.

PREFERRED EMBODIMENTS OF THE INVENTION

The carbon will be provided in powdered or granulated form. The carbon may be provided in the region of the anode in the form of a static bed or a fluidised bed.

The anode, at least, must be inert which means that it must be substantially non-corrodible under the conditions of electrolysis. The cathode should also preferably be inert. Suitable inert anode electrodes are platinum electrodes, platinised-titanium electrodes and graphite electrodes. In one typical example the anode is a platinum electrode and the cathode is a platinum or copper electrode.

In order to achieve effective activation of the carbon the electrolyte must be acidic. The electrolyte typically has a pH in the range 2-3, preferably 2,5. Examples of suitable electrolytes are mineral acids such as sulphuric, sulphurous or phosphoric acids. The electrolyte may also contain other ionic compounds such as copper sulphate and an alkali metal iodide such as potassium iodide. Indeed,

it has been found that the activation of an activated carbon is improved when the electrolyte contains an alkali metal iodide.

The potential which is applied across the electrodes must be sufficient to produce electrolysis. Typically the voltage which is applied across the electrodes is in the range 10 to 50 volts. The current which is applied will vary according to the nature of the carbon being treated. In the case of the activation of virgin carbon the current is preferably less than 1 amp. In the case of the activation of an activated carbon currents as high as 3 amps may be employed. In general, however, the current in all cases is less than 1 amp.

Any suitable electrolysis cell known in the art may be used for carrying out the process of the invention. For example the anode may be a platinum mesh anode wherein the carbon for activation may be accommodated.

The invention also extends to an activated carbon produced by a process as set out above.

The invention is discussed further with reference to the following examples which illustrate some specific embodiments of processes according to the invention. Examples 1 to 4

illustrate the activation of virgin charcoal, examples 5 to 8 the activation to a higher degree of already activated carbon.

EXAMPLE 1:

A sample of granulated virgin charcoal was prepared by crushing commercial charcoal briquettes to a size in the range -10 to + 20 mesh.

The sample was placed in the anode compartment of an electrolytic cell having platinum electrodes, and containing an electrolyte solution of pH 2,5 and consisting of dilute sulphuric acid in distilled water. An initial voltage of 30 volts (DC) and a constant current of 25 mA were applied across the electrodes to produce electrolysis. The temperature was kept in the range 25 to 35°C.

After the process had continued for 24 hours the charcoal, which in its virgin state had an activity constant (K value) of zero, was found to have a K value of 13. After 92 hours the K value had increased to 17.

The activity constant of a carbon is a measure of its adsorption capabilities and is determined, for example, by the method set out in the article "The Mechanism of Gold Adsorption on Activated Carbon" by R.J. Davidson, Journal of the S.A. Institute of Mining and Metallurgy", 75 (4)

November, 1974 at page 67.

EXAMPLE 2:

Under similar conditions as in Example 1, but applying an initial voltage of 24 volts and constant current of 50mA, virgin charcoal was found after 72 hours' treatment to have attained a K value of 16.

EXAMPLE 3:

Granulated virgin charcoal of the type used in Example 1 was subjected to electrolysis as above in an electrolytic cell having a platinum anode and copper cathode. The electrolyte was a solution of copper sulphate in water (10g/I), adjusted to pH 2 with dilute sulphuric acid. An initial voltage of 50 volts was applied, and the current kept constant at 0,7 amps.

The charcoal was found after only 2 hours to have attained a K value of 11.

EXAMPLE 4:

A sample of granulated virgin charcoal was subjected to electrolysis as described above in an electrolytic cell having a platinum anode and copper cathode, and containing an electrolyte comprising a sulphurous acid solution, prepared by forming a saturated solution in distilled

water of SO_2(5 to 6%).

A constant voltage of 50 volts and initial current of 0.7 amps was applied.

After only four hours a K value of 10 was attained.

EXAMPLE 5:

The process conditions were the same as for Example 1 except that instead of virgin charcoal, a sample of Le Carbone type G215 activated carbon (K value 27) was treated. After 24 hours, the K value was found to have increased to 35.

EXAMPLE 6:

A sample of the same activated carbon as Example 5, was treated as above using an electrolyte comprising a 0,1 m KI solution adjusted to pH 2,5 with dilute sulphuric acid. A constant current of 2 amps, initial voltage 30 volts, was applied.

The K value improved from 27 to 31 in only half an hour.

EXAMPLE 7:

Proceeding similarly to Example 6, but applying a constant current of 1 amp and initial voltage of 14 volts, the K value was found to increase from 27 to 33 in only one hour.

EXAMPLE 8:

Using a constant current of 3 amps and initial voltage of 17,5 volts, and an electrolyte comprising a 0,1 m KI/0,1 m Na_3PO_4 solution, and proceeding as above, a sample of the activated charcoal of Example 5 was found after only half an hour to have attained a K value of 33.

The process of the invention is accordingly seen to have important utility. Not only does it provide a simple method of producing activated carbon from virgin charcoal, but it also provides an effective method for upgrading the activity of commercial activated carbon.

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DELCARE THAT WHAT WE CLAIM IS:

1.

A process for increasing the activity of carbon comprising the steps of providing an acidic electrolyte having an inert anode and cathode immersed therein, locating the carbon in the electrolyte in the region of the anode, and applying an effective potential across the electrodes.

2.

A process according to claim 1 wherein the carbon is virgin carbon.

3.

A process according to claim 1 wherein the carbon is an activated carbon.

4.

A process according to any one of claims 1 to 3 wherein the pH of the electrolyte is in the range 2 to 3.

5.

A process according to any one of claims 1 to 3 wherein the pH of the electrolyte is 2,5.

6.

A process according to any one of the preceding claims wherein the electrolyte is a mineral acid.

7.

A process according to claim 6 wherein the mineral acid is selected from sulphuric acid, sulphurous acid and phosphoric acid.

8.

A process according to any one of claims 1 to 7 wherein the voltage applied across the cathode and anode is in the range 10 to 40 volts.

9.

A process according to any one of claims 1 to 8 wherein the current generated between the cathode and anode is less than 1 amp.

10.

An activated carbon produced by a process of any one of the preceding claims.

DATED THIS 8TH DAY OF JUNE, 1978.

SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS

PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	77/4287
Date of Filing:	8th June, 1978
Name of Applicant:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
Name of Inventors:	1. RAYMOND JOHN DAVIDSON 2. VITTORIO VERONESE
Title of Invention:	"REGENERATION OF USED ACTIVATED CARBON"

~~Provisional~~/Complete Specification as Filed

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PATENT ATTORNEYS
JOHANNESBURG — PRETORIA — DURBAN

Cat. Form. B 1490

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 77/4287

TITLE OF INVENTION: "REGENERATION OF USED ACTIVATED CARBON"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

FIELD OF THE INVENTION

THIS invention relates to a process for regenerating an activated carbon, the activity of which has been reduced through use.

DESCRIPTION OF PRIOR ART

The adsorption capabilities of activated carbon have wide application in industry. Activated carbon is used for purifying many industrial effluents and is also used extensively in the sugar industry. In the field of metallurgy, activated carbon is used for efficiently extracting valuable metals such as gold and silver from solution. Particular examples of recovering gold, silver, copper and nickel values from solution using activated carbon are described in the complete specifications of S.A. patent No. 73/8939 and its patent of addition No. 76/4204.

During use, activated carbon loses its effective activity. Deactivated carbon is usually reactivated thermally at temperatures of the order of 500 to 900^{o}C.

It is an object of the invention to provide a process which is capable of regenerating a spent activated carbon, i.e. a carbon whose activity has been reduced through use.

SUMMARY OF THE INVENTION

According to the present invention there is provided a process for regenerating an activated carbon, the activity of which has been reduced through use, comprising the steps of providing a basic electrolyte having an inert anode and cathode immersed therein, locating the carbon in the electrolyte in the region of the anode, and applying an effective potential across the electrodes.

PREFERRED EMBODIMENTS OF THE INVENTION

The carbon will be provided in powdered or granulated form. The carbon may be provided in the region of the electrode in the form of a static bed or a fluidised bed.

The anode, at least, must be inert which means that it must be substantially non-corrodible under the conditions of electrolysis. ~~The cathode should also preferably be inert.~~ Suitable inert anode electrodes are platinum electrodes, platinised-titanium electrodes and graphite electrodes.

In order to achieve effective activation of the carbon the electrolyte must be basic. The electrolyte typically has a pH in the range 8 to 11, preferably 10. Examples of suitable basic electrolytes are hydroxides, preferably alkali metal hydroxides such as sodium hydroxide.

The potential which is applied across the electrodes must be sufficient to produce electrolysis. Typically the voltage which is applied across the electrodes is in the range 10 to 50 volts. The current which is applied will vary according to the nature of the carbon being treated, but is preferably less than 1 amp.

Any suitable electrolysis cell known in the art may be used for carrying out the process of the invention. For example the anode may be a platinum mesh anode wherein the carbon for activation may be accommodated.

The invention has particular application to the regeneration of an activated carbon which has been used for recovering gold, silver, copper or nickel in the manner described and claimed in the specifications of the abovementioned S.A. patent and its patent of addition.

The invention also extends to an activated carbon regenerated by a process as set out above.

The invention is discussed further with reference to the following example.

Example:

A sample of granulated activated charcoal obtained from a gold extraction process as described in the complete specification of S.A. patent No. 73/8939 was found to have an activity constant (K value) of 11 (as compared with a value before use of 22).

The spent charcoal was located in the anode compartment of an electrolytic cell having platinum electrodes and containing a sodium hydroxide electrolyte solution of pH 10.

An initial voltage of 30 volts was applied across the electrodes, a constant current of 25 mA was maintained, and after 24 hours the activated charcoal was found to have attained a K value of 33.

Not only did this process therefore effectively regenerate the used catalyst, but its activity was in fact after regeneration significantly higher than it had been before use.

The activity constant of a carbon is a measure of its adsorption capabilities and is determined, for example, by the method set out in the article "The Mechanism of Gold Adsorption on Activated Carbon" by R.J. Davidson, Journal of the S.A. Institute of Mining and Metallurgy, 74 (4).

1.

A process for regenerating an activated carbon, the activity of which has been reduced through use, comprising the steps of providing a basic electrolyte having an inert anode and cathode immersed therein, locating the carbon in the electrolyte in the region of the anode, and applying an effective potential across the electrodes.

2.

A process according to Claim 1 wherein the pH of the electrolyte is in the range 8 to 11.

3.

A process according to claim 1 or claim 2 wherein the pH of the electrolyte is 10.

4.

A process according to any one of the preceding claims wherein the electrolyte is a hydroxide.

5.

A process according to claim 4 wherein the hydroxide is an alkali metal hydroxide.

6.

A process according to any one of claims 1 to 5 wherein the voltage applied across the cathode and anode is in the range 10 to 50 volts.

7.

A process according to any one of claims 1 to 6 wherein the current generated between the cathode and anode is less than 1 amp.

8.

An activated carbon produced by a process of any one of the preceding claims.

PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 78/5864

Date of Filing: AUGUST 21, 1979

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventors: 1. RAYMOND JOHN DAVIDSON
2. VITTORIO VERONESE

Title of Invention: "REGENERATION OF ACTIVATED CARBON"

~~Provisional~~/Complete Specification as Filed

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PATENT ATTORNEYS
JOHANNESBURG — PRETORIA — DURBAN

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 78/5864

TITLE OF INVENTION: "REGENERATION OF ACTIVATED CARBON"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street,
JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

THIS invention relates to the regeneration of activated carbon.

According to the invention, a method of regenerating an activated carbon which has had its collecting capacity for gold adsorption reduced through use, includes the step of contacting the carbon with an aqueous copper salt solution having a pH of less than 6. The aqueous solution may be a copper chloride, a copper nitrate, or preferably a copper sulphate solution. Such solutions will generally contain up to 5 percent by weight of the copper salt.

The collecting capacity for gold adsorption or equilibrium capacity constant of activated carbon may be determined by standard methods as, for example, that described in the article "The mechanism of gold adsorption on activated carbon" by R.J. Davidson, Journal of the S.A. Institute of Mining and Metallurgy, 75 (4), November, 1974, at page 67.

The mildly acidic copper salt solution preferably also contains carbonic acid. The carbonic acid may be produced in situ by bubbling carbon dioxide through the solution, or by the addition of a suitable carbonate salt, such as sodium carbonate. Preferably, in the case of bubbling CO_2 through the copper salt solution the solution is saturated with carbon dioxide and this may take place in some instances at elevated pressures.

According to another aspect of the invention, the carbon is contacted with the copper salt solution (without carbonic acid or other carbonate additions) and is thereafter contacted with a solution of a suitable carbonate salt such as ammonium carbonate. Ammonium carbonate is the preferred carbonate salt due to its ability to complex any excess adsorbed copper which will thereafter be eluted from the carbon.

The carbonate solution, whether added to the copper salt solution or used separately will generally contain up to 5 percent by weight carbonate.

Contact between the carbon and the copper salt solution may be achieved by agitating the carbon in the solution. Preferably, the carbon to be regenerated is placed in a column and the regeneration solution or solutions then passed through the column, either individually or sequentially.

The copper salt solution may be contacted with the carbon at ambient temperature or preferably at an elevated temperature such as 90 to 110°C.

In particular, the method of the invention has application to the regeneration of activated carbon, the equilibrium constant of which has been reduced from use in the method described and claimed in U.S. Patent No. 3,970,737.

Examples of the invention will now be described.

EXAMPLE 1

An activated carbon had its collecting capacity for gold adsorption reduced from 37 mg Au/gC to 25 by use in a known gold recovery process. This carbon was placed in an elution column of 25 mℓ volume. A copper sulphate solution of concentration 2 to 3% by mass was prepared. Carbon dioxide was bubbled through the copper sulphate solution at the rate of 1 ℓ/m at ambient temperature and pressure until the solution was saturated with carbon dioxide. At this point the pH of the solution was 3,5. The copper sulphate solution, at room temperature, saturated with carbon dioxide, was then pumped into the elution column which was maintained at 91°C. 5 bed volumes (125 ml) of the heated copper sulphate solution were passed through the elution column at the rate of 5 bed volumes per hour. The carbon was removed from the column and was found to have had its collecting capacity increased to 37 mg Au/gC again.

The kinetic response of the chemically regenerated carbon was also tested. This was done by monitoring the rate of movement of the gold

reaction front in a typical column operation as described by V. Greenblau, University of Cape Town, Dept. of Civil Engineering, Project Report 5.18, 1970. Pregnant gold plant solution containing 6,7 g/t Au was pumped through an adsorption column containing granular activated carbon at a superficial flow velocity of 1m /min. The gold reaction front was defined by effluent having a concentration of 0,01 g/t Au. The resulting test data involved plotting graphically the service time of the column against the relevant carbon bed depth whereby the slope of the plot indicated the rate of movement of the gold reaction front as defined above.

Figure 1 graphically illustrates the results obtained. The plot of the "first cycle" is that obtained using fresh activated carbon for a total service time of 48 hrs. The plot of "second cycle" is that obtained using the same activated carbon which had been regenerated by the method of the invention for a similar length of time. It can be seen that the rates for the two graphs are the same, viz 4,3 cm/hr, thereby showing that the kinetics of gold adsorption using the regenerated activated carbon were very similar to the kinetics observed using the virgin carbon.

Figure 2 illustrates the case where no chemical regeneration of the carbon was carried out. In this instance a gold plant solution containing 7,079/t Au was used as a column influent.

EXAMPLE 2

Type G215 coconut shell carbon having a collecting capacity for gold adsorption of 37 mg Au/gC at pH 10 was used to adsorb gold from an acidified gold plant pregnant solution at pH 4-5. After loading 2-4% Au, the carbon was eluted at 91^{o}C using deionized water. The eluted carbon had a capacity constant of 25 mg Au/gC.

The eluted carbon was maintained at 91^{o}C and treated with solutions of copper sulphate, ammonium carbonate and soda water either individually, sequentially or as mixtures as set out in Table 1 below:

TABLE 1

A-1 bed volume $CuSO_4$ solution (% $CuSO_4.5H_2O$/ hours treatment time)	B-5 bed volumes $(NH_4)_2CO_3$ solution (% $(NH_4)_2CO_3$/ hours treatment time)	Treatment sequence	Collecting capacity of used carbon (mg Au/g C)
Nil	Nil		25
2,0/1		A only	28
2,5/1		"	28
3,0/1		"	28
	2,0/2,5	B only	28
	3,0/2,5	"	28
	5,0/2,5	"	
0,5/½	2,0/2,5	A + B	32
1,0/½	2,0/2,5	"	33
2,0/½	2,0/2,5	"	35
2,5/½	2,0/2,5	"	36
3,0/½	2,0/2,5	"	36
2,0/1 but saturated with CO_2		A only	37
2,5/1 but saturated with CO_2		"	38
3,0/1 but saturated with CO_2			

EXAMPLE 3

A similar series of regeneration experiments to those of Example 2 were carried out using a variety of copper sulphate/ammonium carbonate contact solutions and comparisons made with other metal salt solutions such as nickel sulphate, zinc sulphate and iron sulphate solutions. The results of these experiments are set out in Table 2 below. The significant regeneration of activated carbon using treatment solutions of the invention will be seen from these results.

In Table 2, all percentages given are percentages by weight and BV/h stands for bed volumes/hour.

TABLE 2

Virgin Carbon 37 mg Au/gC

The charcoal was loaded with WOL Au pregnant solution at pH 10 for 48 h. The charcoal was eluted using 5% NaCN and 1% NaOH (1BV) in contact for 30 min. follow by 5 BV of deionized water - at 91°C - The charcoal has an initial capacity of 25 mg Au/gC after elution.

Test No.	CHEMICAL REAGENT				Contact Reagent (1)	Contact Reagent (2)	FLOW RATE			Bed Volume of Water	Capacity after Regeneration	ppmCu in the charcoal
	Reagent (1)	Bed Volume	Reagent (2)	Bed Volume	Min.	Min.	Reagent (1) BV/h	Reagent (2) BV/h	Water after each Reag. BV/h		mgAu/gC	
CR1	0,5% $NiSO_4.6H_2O$	1	2% $(NH_4)_2CO_3$	5	30	passed	2	2	2	2	25	
CR2	1,0% $NiSO_4.6H_2O$	1	"	5	30	"	2	2	2	2	25	
CR3	2,0% $NiSO_4.6H_2O$	1	"	5	30	"	2	2	2	2	27	
CR4	2,0% $NiSO_4.6H_2O$	1	-	-	30	-	2	-	2	2	26	
CR5	-	-	2% $(NH_4)_2CO_3$	5	-	passed	-	2	2	2	26	
CR6	2% $NiSO_4.6H_2O$	1	2% $(NH_4)_2CO_3$	5	60	"	2	2	2	2	25	
CR7	"	1	CO_2 gas	1ℓ/min	30	"	2	30 min	2	2	26	
CR8	3% $NiSO_4.6H_2O$	1	2% $(NH_4)_2CO_3$	5	30	"	2	2	2	2	26	

CR9	0,5% $ZnSO_4 \cdot 7H_2O$	1	2% $(NH_4)_2CO_3$	5	30	passed	2	2	2	2	24	
CR10	1,0% $ZnSO_4 \cdot 7H_2O$	1	"	5	30	"	2	2	2	2	24	
CR11	2,0% $ZnSO_4 \cdot 7H_2O$	1	"	5	30	"	2	2	2	2	24	
CR12	3,0% $ZnSO_4 \cdot 7H_2O$	1	"	5	30	"	2	2	2	2	24	
CR13	2% $ZnSO_4 \cdot 7H_2O$	1	CO_2 gas	1ℓ/min	30	"	2	30 min	2	2	24	
CR14	2% $ZnSO_4 \cdot 7H_2O$	1	2% $(NH_4/_2CO_3$	5	60	"	2	2	2	2	25	
CR15	0,5% $FeSO_4 \cdot 7H_2O$	1	2% $(NH_4)_2CO_3$	5	30	"	2	2	2	2	25	
CR16	1,0% $FeSO_4 \cdot 7H_2O$	1	"	"	30	"	2	2	2	2	27	
CR17	2,0% $FeSO_4 \cdot 7H_2O$	1	"	5	30	"	2	2	2	2	26	
CR18	0,5% $CuSO_4$	1	"	5	30	"	2	2	2	2	32	3070
CR19	1,0% $CuSO_4$	1	"	5	30	"	2	2	2	2	33	4600
CR20	2,0% $CuSO_4$	1	"	5	30	"	2	2	2	2	35	5500
CR21	"	1	"	5	30	"	2	2	2	2	35	7137
CR22	2,5% $CuSO_4$	1	"	5	30	"	2	2	2	2	36	8800

CR23	$CuSo_4$	1	$(NH_4)_2CO_3$	5	30	passed	2	2	2	2	36	11162
CR24	2,5% $CuSO_4$	1	5%$(NH_4)_2CO_3$	5	30	"	2	2	2	2	35	5700
CR25	3,0% $CuSO_4$	1	"	5	30	"	2	2	2	2	35	6225
CR26	2,0% $CuSO_4$	1	CO_2 gas	1ℓ/min	30	-	2	-	2	2	37	12150
CR27	2,5% $CuSO_4$	1	into the		30	-	2	-	2	2	38	11562
CR28	3,0% $CuSO_4$	1	$CuSO_4$		30	-	2	-	2	2	38	13975
CR29	2,0% "	1	CO_2 gas	1ℓ/min	30	-	2	60 min	2	2	37	12600
CR30	2,5% $CuSO_4$	1	only		30	-	2	60 min	2	2	38	14050
CR31	3,0% "	1			30	-	2	60 min	2	2	38	14475
CR32	2,0% "	1	-	-	60	-	2	-	2	2	28	15650
CR33	2,5% "	1	-	-	60	-	2	-	2	2	28	16175
CR34	3,0% "	1	-	-	60	-	2	-	2	2	28	16750
CR35	-	-	2%$(NH_4)_2CO_3$	5	-	passed	-	2	2	2	28	-
CR36	-	-	3%$(NH_4)_2CO_3$	5	-	"	-	2	2	2	28	-
CR37	-	-	5%$(NH_4)_2CO_3$	5	-	"	-	2	2	2	28	-
CR41	2% $CuSO_4$	1	2% $NaHCO_3$	5	30	passed	2	2	2	2	32	9900
CR42	"	1	5% "	1	30	"	2	2	2	2	31	10862
CR43	"	1	5% "	1	30	30 min	2	2	2	2	32	10400
CR44	"	1	2%$(NH_4)_2CO_3$	1	30	30 min	2	2	2	2	32	8781
CR45	"	1	5%$(NH_4)_2CO_3$	1	30	30 min	2	2	2	2	33	7025

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

1.
A method of regenerating an activated carbon which has had its collecting capacity for gold adsorption reduced through use, includes the step of contacting the carbon with an aqueous copper salt solution having a pH of less than 6.

2.
A method of claim 1 wherein the copper salt solution is a copper chloride or copper nitrate solution.

3.
A method of claim 1 wherein the copper salt solution is a copper sulphate solution.

4.
A method of any one of the preceding claims wherein the copper salt solution contains up to 5 percent by weight of the copper salt.

5.
A method of any one of the preceding claims wherein the copper salt solution also contains carbonic acid.

6.
A method of claim 5 wherein the carbonic acid is produced in situ by bubbling carbon dioxide through the solution.

7.
A method of claim 6 wherein the solution is saturated with carbonic acid.

8.
A method of claim 5 wherein the carbonic acid is produced in situ by adding a suitable carbonate salt to the solution.

9.
A method of claim 8 wherein the carbonate salt is added in an amount of up to five percent by weight of the solution.

10.
A method of any one of claims 1 to 4 wherein contact between the copper salt solution and the carbon is followed by contacting the carbon with a suitable carbonate solution.

11.
A method of claim 10 wherein the carbonate is ammonium carbonate.

12.
A method of claim 10 or claim 11 wherein the carbonate solution contains up to five percent by weight of the carbonate.

13.
A method of any one of the preceding claims wherein contact between the carbon and the solution or solutions is achieved by agitating the carbon in the solution or solutions.

14.
A method of any one of claims 1 to 12 wherein contact between the carbon and the solution or solutions is achieved by placing the carbon in a column and passing the solution or solutions through the column.

15.
A method of any one of the preceding claims wherein contact between the carbon and the solution or solutions takes place at a temperature in the range 90 to 110°C.

16.
A method of claim 1 substantially as herein described with reference to Examples 1 to 3.

DATED THIS 21ST DAY OF AUGUST, 1979.

SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS

FIGURE 1

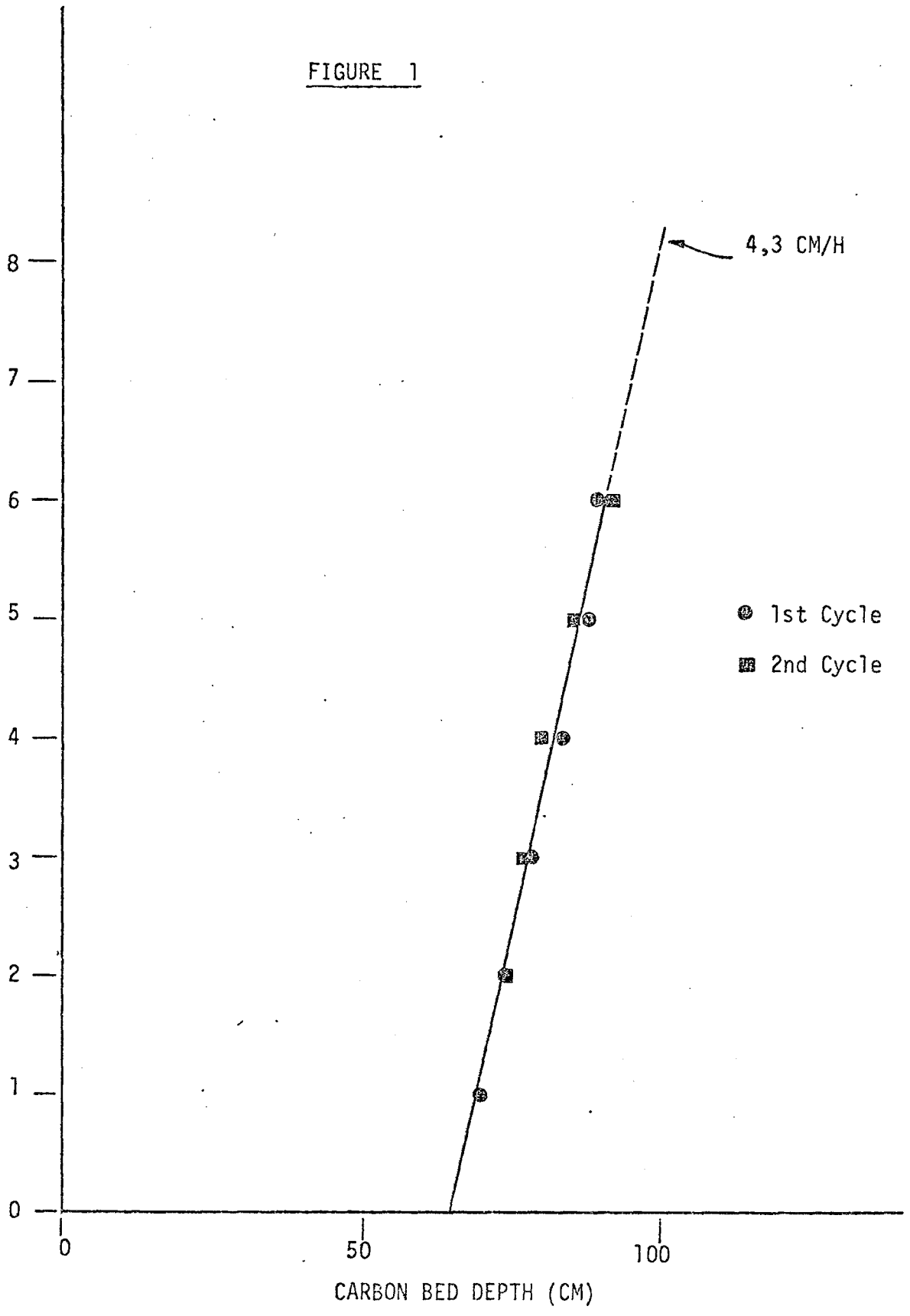
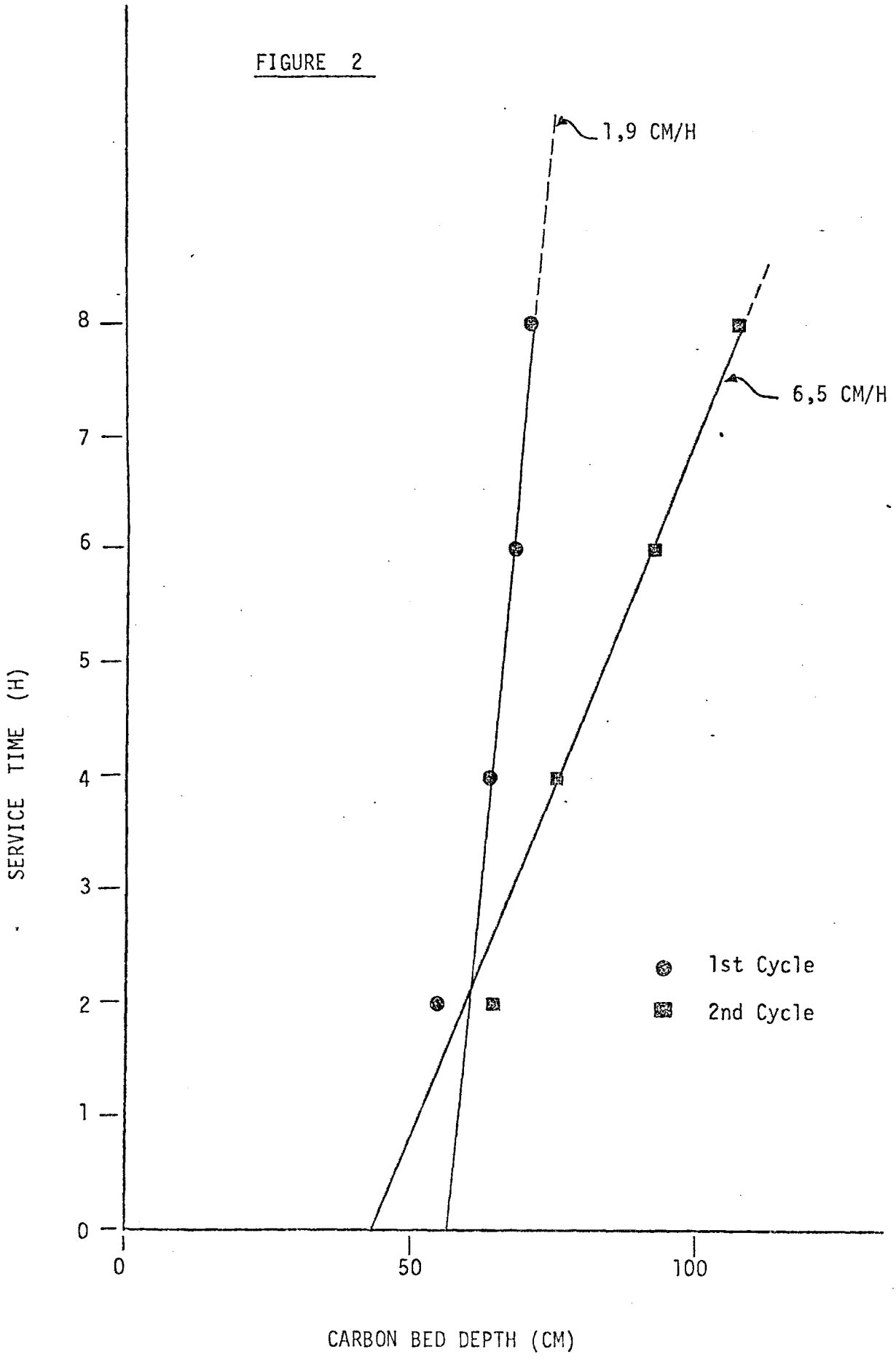

4,3 CM/H
8
7
6
5
4
3
2
1
0
1st Cycle
2nd Cycle
0
50
100
CARBON BED DEPTH (CM)

FIGURE 2



SPOOR and FISHER
APPLICANT'S PATENT ATTORNEYS

6.0 CARBON MANAGEMENT AND SELECTION

Three papers related to carbon selection and management are appended.

Following a review of the parameters affecting gold adsorption (Lecture 10 of the 1985 Carbon School) as well as an attempt to interrelate various properties of carbon (paper presented at the Australasian Institute of Mining and Metallurgy), test specifications were defined for the purchasing of suitable coconut shell carbons - see AARL test specifications.

Finally, a paper entitled Carbon Management (Chapter XX) was prepared for the intended Carbon Monograph to be published by the S.A.I.M.M. which has, however, been postponed. This paper will nevertheless be published shortly by the S.A.I.M.M. (April/May 1991).

LECTURE 10

PROPERTIES OF CARBON FOR GOLD ADSORPTION AND THEIR MEASUREMENT

by

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Anglo American Research Laboratories

SUMMARY

Factors affecting gold adsorption onto activated carbon are reviewed in order to emphasise the parameters which require control when assessing carbon selection or carbon quality control. Carbon activity measured in terms of equilibrium capacity data was shown to be most sensitive to the ionic strength of the adsorption medium as well as to the cyanide concentration and temperature. While carbon activity measured in terms of kinetic response was also sensitive to these parameters, the particle size of the carbon was perhaps more significant. The effects of calcium carbonate and organic fouling are also discussed together with the effects of other plant constituents.

Various test procedures including the measurement of equilibrium adsorption capacity, gold adsorption rate and attrition resistance are outlined. With regard to carbon selection, the carbon activity together with physical hardness are the most important considerations apart from the commercial price of the carbon. In this instance testing in which the physical and chemical properties and prices of commercial products are compared is most empirical, although the "weighting" of the various properties is likely to be most complex.

Carbon quality while being the single most important aspect in any carbon adsorption operation, is very often the most neglected aspect. Emphasis must be placed on the reproducibility of test procedures if meaningful control is to be achieved. With regard to the quality control of carbon in existing plants, no general test procedures are currently in use. Each plant would appear to make use of some carbon testing although in many instances no quality control is exercised. In view of this situation,

it is not easy to comment on quality control per se, suffice to say that in general it is sadly lacking in most operations.

INTRODUCTION

Activated carbons are made by heating organic materials in a mildly oxidizing atmosphere. In the course of activation the carbon develops a cellular structure with an extended surface area. In many ways activated carbon is a unique material with properties not possessed by any other class of organic substance. Since activated carbon usually contains considerable quantities of adsorbed oxygen and hydrogen the term carbon is perhaps a misnomer. It is essential to regard activated carbon as a complex organic polymer and not merely as an amorphous form of the element carbon. Activated carbon differs from other organic polymers in having an exceptionally large surface area and in being a good conductor of electricity. Many of its intriguing characteristics stem from the latter fact, as well as being related to the various functional groups present on the carbon surface.

When related to the adsorption of gold and the carbon-in-pulp process, both the selection of suitable granular carbons together with carbon quality control presents a complex and sometimes confusing picture. Not only is the mechanism of gold adsorption and the functional groups taking part in the reaction on the carbon surface not well understood at the present time, but little selection other than by trial and error is documented in the literature. Furthermore, the technology itself is presently very much in a state of development. Perhaps the primary requisites of a good carbon are firstly physical stability and secondly high gold loading capacity and kinetic response. Both of these characteristics are however generally antipathetic due primarily to the very nature of the activation process i.e. activation at high temperatures is accompanied by a loss in physical strength. This situation immediately imposes some measure of compromise in the weighting of such properties.

GOLD ADSORPTION

Before discussing possible measuring procedures, it is necessary to first outline the factors affecting gold adsorption onto activated carbon. Due to the sometimes confounding effects of such variables it is not always possible to quantify each parameter in simple terms. At best, only general trends may be highlighted in a review of this nature. Furthermore, as many effects may be best described in terms of equilibrium data, such effects do not necessarily relate to the kinetics of gold adsorption in the same way.

Since various workers have used either the Freundlich or the Langmuir models in order to describe equilibrium adsorption data, it is most necessary to establish the reversibility of the adsorption reaction. In exhaustive tests lasting 3 months it was shown that carbon continued to adsorb gold from a solution containing no free cyanide and an irreversible mechanism was concluded[1]. Likewise, equilibrium experiments using different initial gold solution concentrations containing no free cyanide showed a considerable difference in adsorption capacities[2]. Similar tests however, when carried out in the presence of free cyanide yielded almost identical adsorption isotherms indicative of a reversible adsorption mechanism. The reversibility of the adsorption reaction would thus appear to be dependent on the presence of free cyanide[2].

Ionic strength of the adsorption medium.

The enhanced adsorption of gold and silver in the presence of added salts, and particularly Ca^{2+}, was first demonstrated by Davidson[3] and subsequently confirmed by other workers[1,4,5]. Figure 1 shows the effect of various ionic media of equal ionic strength on the adsorption of gold where it may be considered that the very significant effects demonstrated are due to an ionic strength effect. However, concentrations of KCl three times as high as $CaCl_2$ do not produce a loading enhancement of anything like as much. Furthermore, the enhancement of loading has a dramatic effect with only a very small quantity of added electrolyte which is not in keeping with a simple ionic-strength effect.

Further investigations[5] related to the adsorption of gold in the presence of various mono and divalent cations indicated the adsorption of these counter cations which were very pH dependent. At low gold loadings (< 2 kg/t) there was a negligible adsorption of counter cations between equilibrium pH 2-10,5 indicating that the gold on the carbon was present as a neutral species or ion-paired with H^+. However, at higher gold loadings (ca 10 kg/t) there was a significant adsorption of cations above pH 6 and a corresponding increase in gold loading when compared with pure solutions (Figure 2). Once again, there was very little gold loading enhancement in the presence of Li^+ and K^+ when compared with Ca^{2+} and Mg^{2+}. The enhancement in gold loading appeared to correlate with the charge of these cations and not on their size. Although any estimation regarding the molar adsorption ratio of gold and Ca^{2+} or Mg^{2+} was complicated by the possible precipitation of $CaCO_3$ or $MgCO_3$, one is forced to conclude that a divalent cation such as calcium is specifically involved in the adsorption of gold. Also, this sensitivity of gold adsorption to low concentrations of added electrolyte may

well explain in part at least the many confusing observations and conclusions reported by earlier investigators.

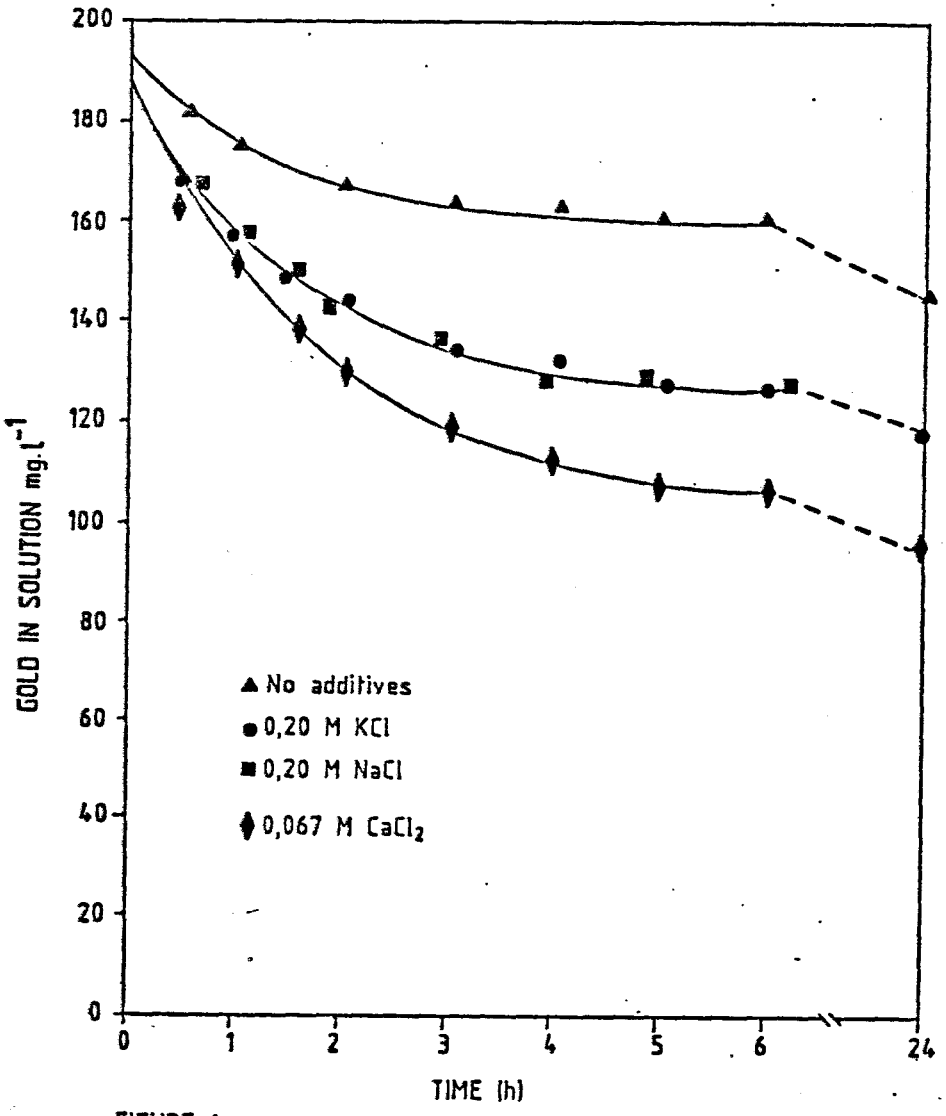

FIGURE 1

Rate Curves for the Adsorption of Aurocyanide on Carbon in the presence of Electrolytes at pH10 (FROM McDOUGALL etal [1])

Calcium carbonate deposition.

While the presence of a divalent cation like calcium would appear to play a significant role in enhancing gold adsorption, the simultaneous deposition of calcium carbonate which occurs on the carbon in a typical plant situation has a passivating effect on the rate of gold adsorption. Although this detrimental effect has not been satisfactorily documented and quantified, it would appear that calcium loadings of 10-20 kg/t have only a marginal effect on gold adsorption rates while loadings in excess of 50 kg/t result in a serious decline in adsorption rates. The presence of calcium carbonate does

not affect the equilibrium capacity of the carbon to adsorb gold which suggests that the passivating effect is purely physical.

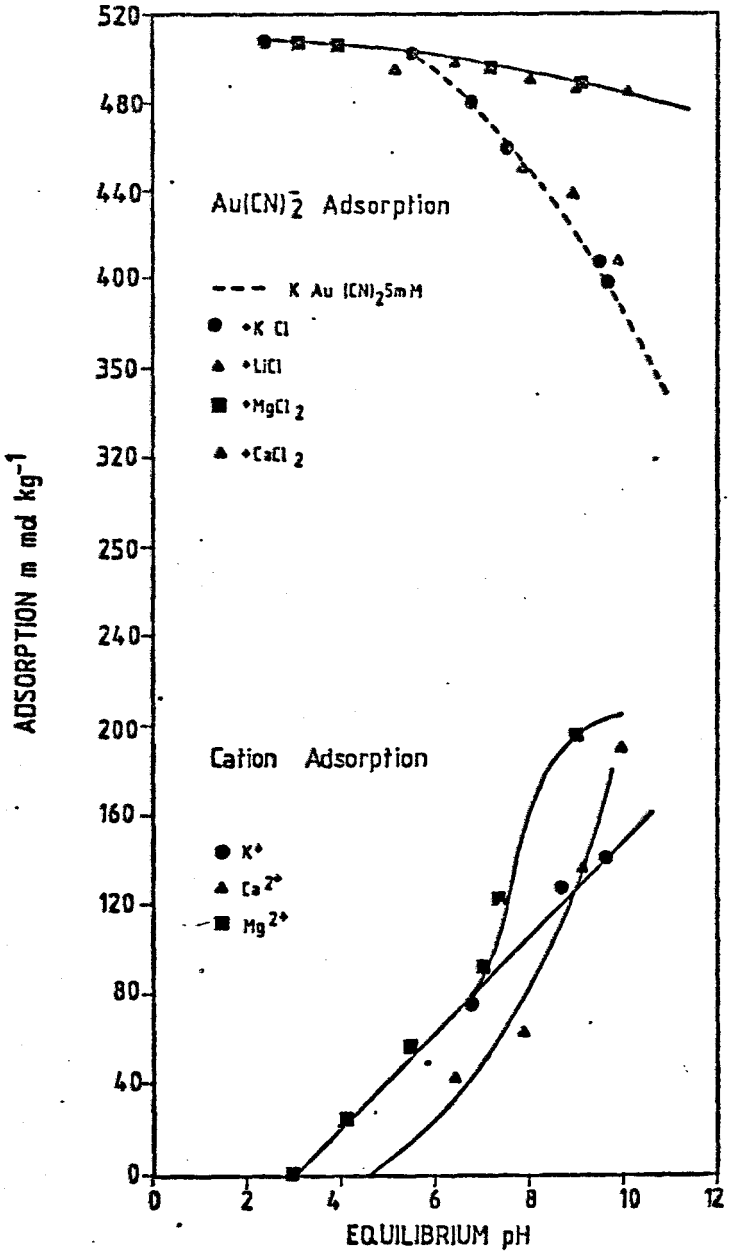

FIGURE 2.

Adsorption of Gold and Cation on Carbon at 25°
(FROM TSUCHIDA et al [5])

It is well known that activated carbon readily adsorbs oxygen and acts as an oxidation catalyst for the oxidation of free cyanide[6].

$$CN^- + \tfrac{1}{2}O_2 \rightleftharpoons CNO^-$$

$$CNO^- + 2H_2O \rightleftharpoons HCO_3^- + NH_3$$

It may thus be presumed that the free cyanide and oxygen present in typical plant circuits are the main source of such $CaCO_3$ fouling. Pilot CIP testwork[2] in which initial cyanide concentrations were controlled and cyanide degradation monitored certainly indicated poor metallurgical performance at high cyanide levels (ca 300 g/t NaCN) together with high loadings of calcium when compared with the exceptionally good metallurgical performance and low calcium loadings observed at lower cyanide levels (ca 100 g/t NaCN). This testwork further indicated that the marginal lowering of pH of the feed pulp had a most beneficial effect on both cyanide destruction as well as on gold adsorption while dramatically reducing calcium carbonate fouling (see below).

Based on solubility considerations the deposition of calcium carbonate is most sensitive to pH. As $[CO_3^{=}] \approx [HCO_3^{-}]$ at pH 8,3 it is evident that the loadings of calcium onto activated carbon becomes very pH dependent in the region pH 9-11 which is commonly found in gold plant dissolution circuits. This sensitivity to pH is well illustrated in Figure 2.

The acid washing of loaded carbon with dilute hydrochloric acid (3-5 vol. %) has been shown to effectively remove calcium carbonate from the carbon without eluting any gold or silver values. Acid washing of carbon prior to elution and thermal regeneration is also advisable as calcium is a particularly effective catalytic oxidant for the water gas reaction.

Cyanide concentration and temperature.

It has been amply demonstrated from a thermodynamic point of view that gold adsorption onto carbon is adversely affected by the presence of free cyanide in the adsorption medium and this effect becomes more pronounced with any increase in temperature. This effect is graphically illustrated in Figure 3 where the almost linear effect of temperature may be seen to increase from 0,50 kg Au/tC/°C in the absence of free cyanide to 0,78 kg/tC/°C in the presence of 200 g/t NaCN. Further equilibrium data showing the effect of free cyanide on gold adsorption at 25°C are presented in Figure 4 where it is evident that small additions of cyanide have a pronounced effect on gold loadings but that further increases in cyanide concentration above ca 100 g/t NaCN have a much smaller effect.

From kinetic considerations it may be expected that an increase in temperature would increase the initial rate of gold adsorption but that at gold loadings approaching equilibrium a serious decrease in adsorption rates would result. These effects are well illustrated in Figure 5 where the use of the Langmuir equation depicts the rate of gold adsorption onto carbon fairly well.

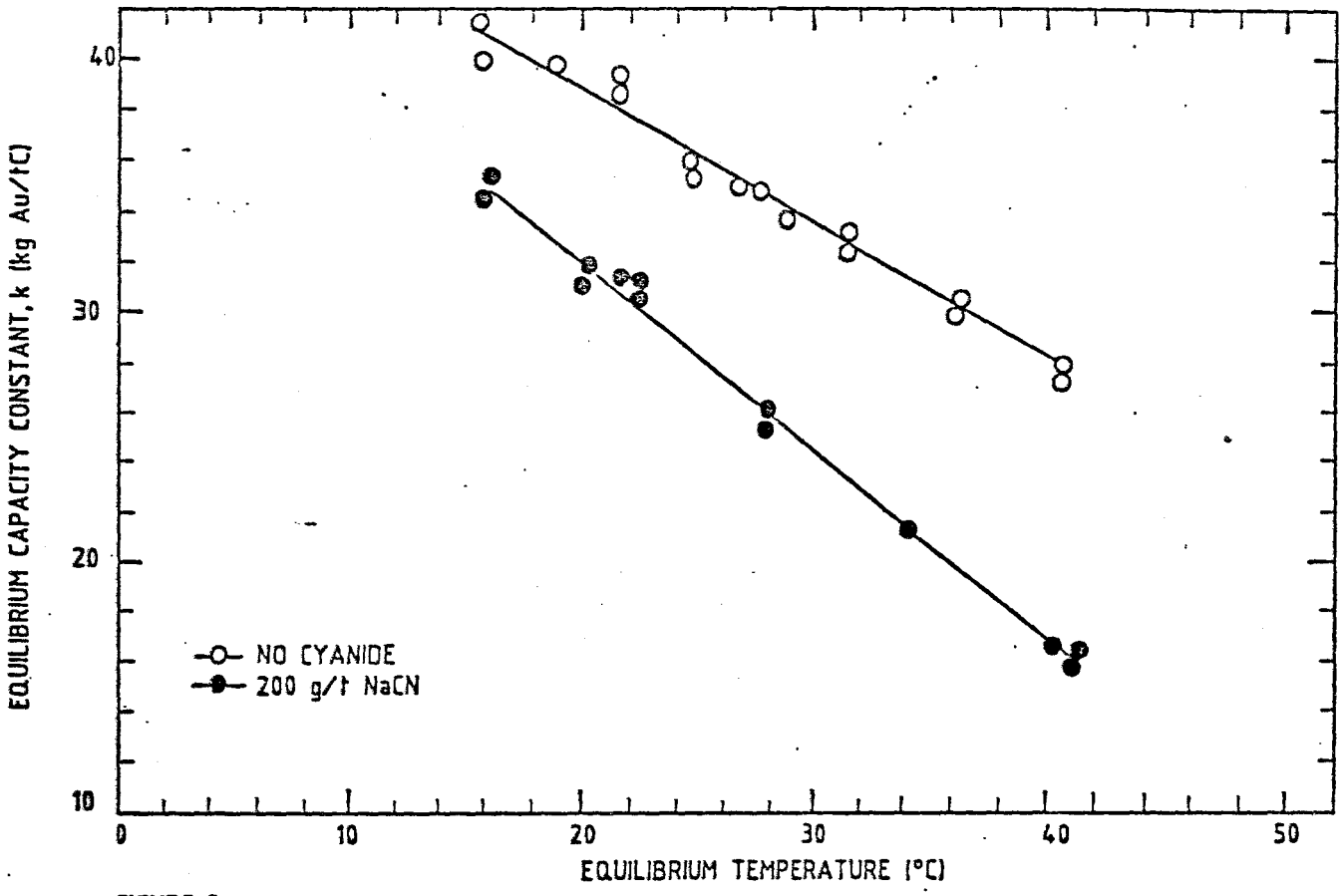

FIGURE 3
Effect of Temperature on Equilibrium Gold Loadings in Synthetic Plant Solution at pH 10 (FROM DAVIDSON etal [2])

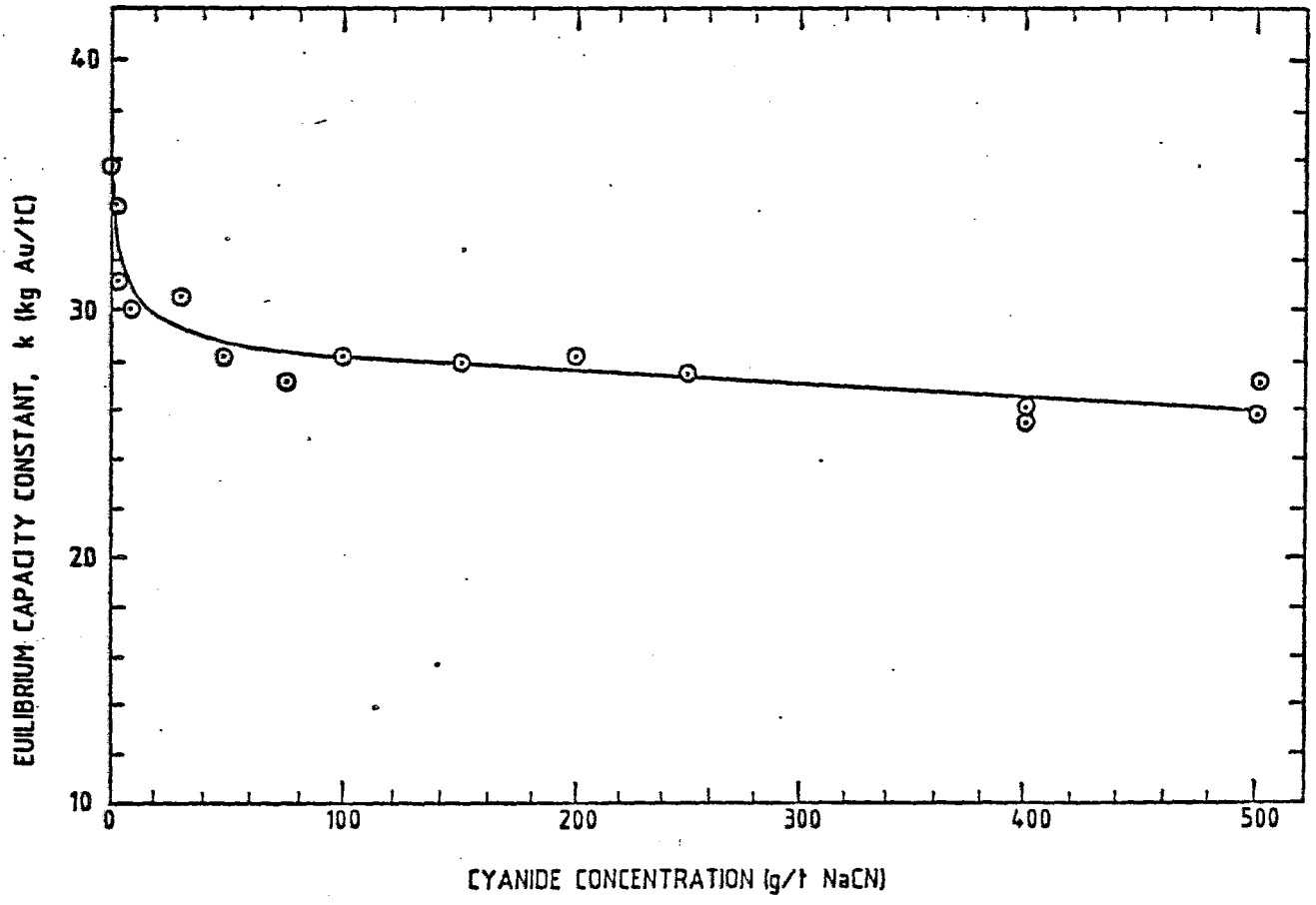

FIGURE. 4.
Effect of Cyanide Concentration on Equilibrium Gold Loadings at 25° C in Synthetic Plant Solution at pH 10 (FROM DAVIDSON etal [2])

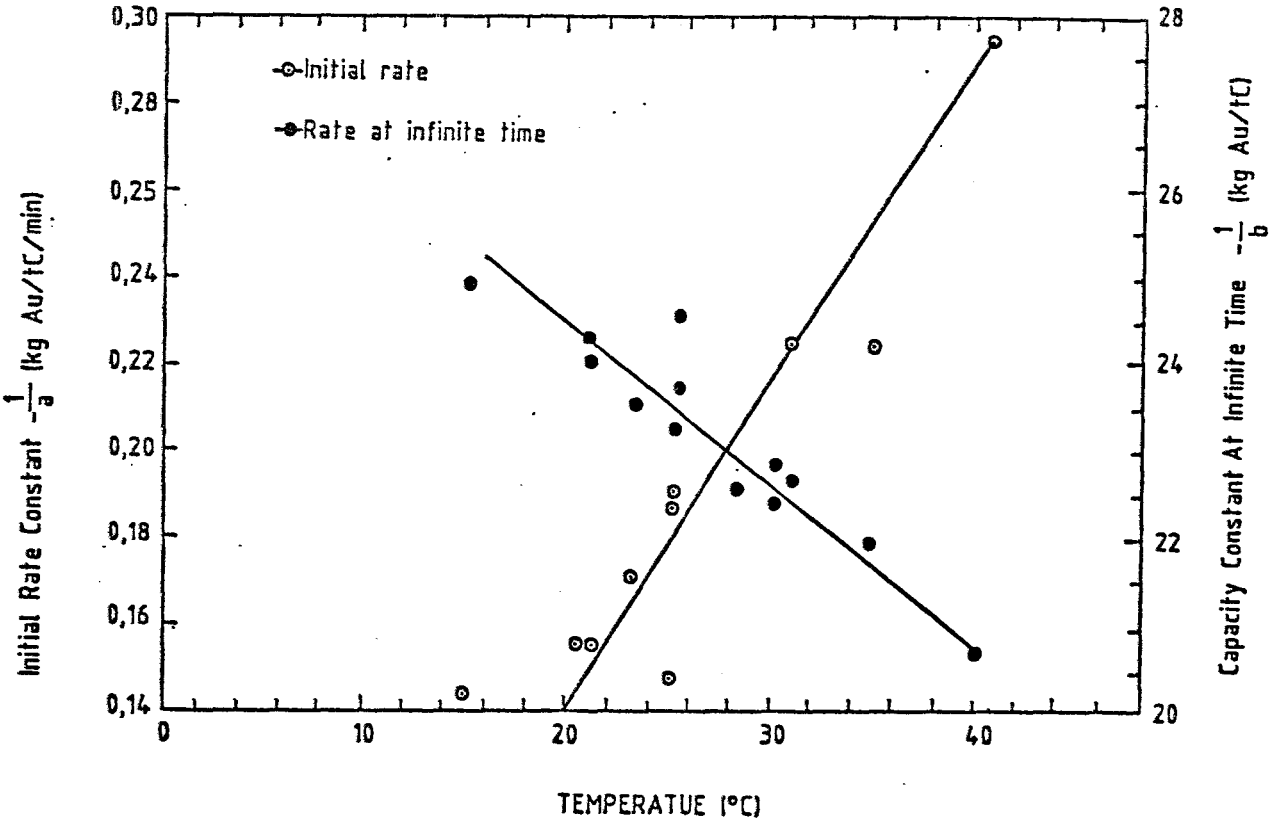

FIGURE 5

Effect of Temperature on Gold Adsorption Rates in Synthetic Plant Solution at pH10 (FROM DAVIDSON etal [2])

The free cyanide concentration also has a significant effect on the selectivity of carbon for gold when dealing with typical plant solutions or pulps but this aspect will be discussed later (see Other plant constituents).

Acidity of the adsorption medium.

It is well established that gold adsorption onto carbon is strongly enhanced by an increase in the acidity of the adsorption medium. Equilibrium data obtained in the absence of free cyanide (Figure 6) indicate a marginal decrease in gold capacity in the region pH 6-12, whereas the gold capacity increases strongly at lower pH levels. An equilibrium gold loading of 60 kg/t of carbon at pH 12 was reported which may be compared with a gold loading of 200 kg/t obtained at pH 1.

An many of the investigations related to pH have been carried out in synthetic solutions it is perhaps not wise to assume a similar response when dealing with complex solutions or pulps (see Other plant constituents). Both the presence of free cyanide and lime together with a heterogenous mixture

of metal cyanide complexes have a profound effect on gold adsorption. In the case of free cyanide, CN^- is protonated to the neutral HCN species at relatively high pH levels (pKa HCN = 9,5) thus lowering cyanide concentrations significantly when marginally lowering pH below pH 10,3 while at the same time increasing the catalytic oxidation of free cyanide to $CO_3^=$ and NH_3. The marginal lowering of pH below pH 10 will also have a significant effect on calcium carbonate fouling of the carbon (see Calcium carbonate deposition).

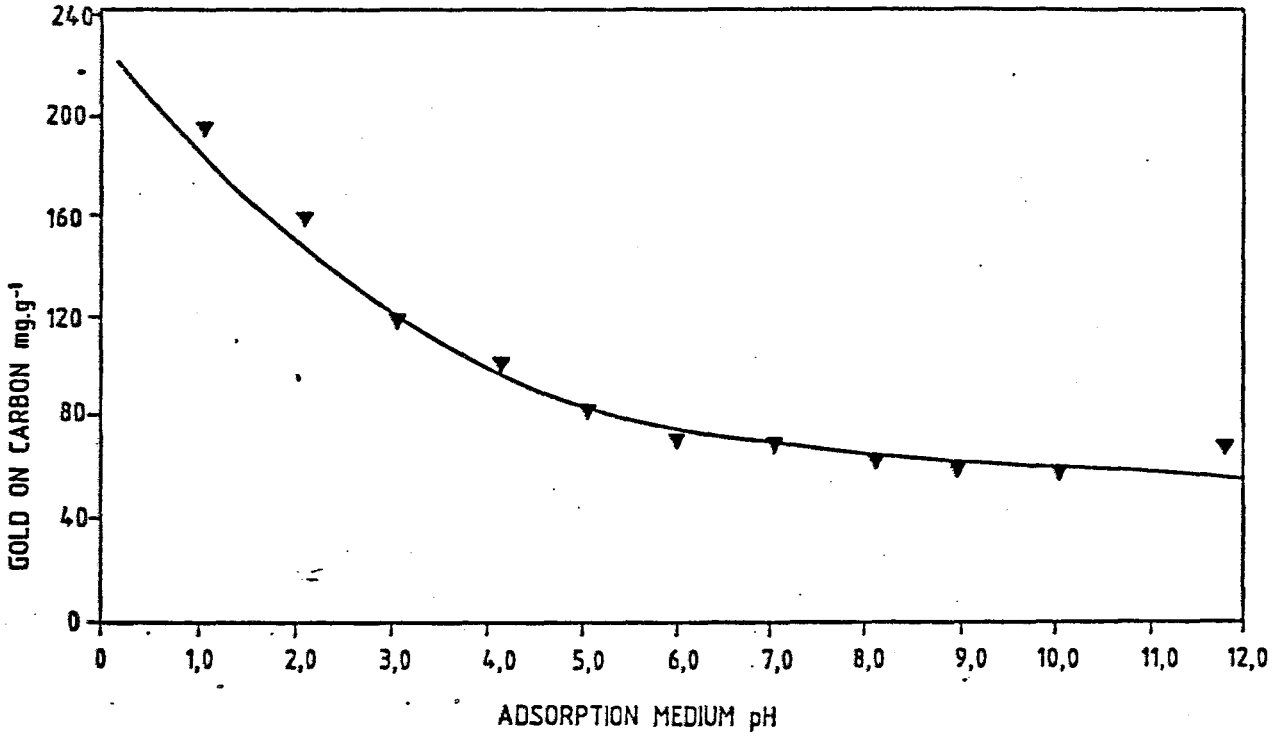

FIGURE 6
Effect of the pH Value of the Adsorption Medium on the Gold Capacity of the Carbon (FROM McDOUGALL et al [1])

Other plant constituents.

A number of base metals are usually leached with gold in cyanide solution, notably copper, nickel, cobalt, iron and zinc. Under typical plant conditions activated carbon is most selective in adsorbing gold from such complex mixtures. Selective metal adsorption appears to take the form of a complex displacement sequence with gold followed by silver being the most strongly adsorbed metal species. No well defined sequence of metal displacement may be formulated as the adsorption of such metal values appears to be most sensitive to such processing parameters as pH, free cyanide concentration, temperature and of course the relative concentration of these metal

values in a particular plant solution or pulp. In general, the selectivity of carbon for gold adsorption will increase with an increase in pH or free cyanide concentration.

In view of carbon's limited capacity to adsorb metal cyanide complexes, the adsorption of such values has a detrimental affect on gold adsorption by virtue of the selective replacement observed. While this effect is difficult to express in simple terms, any loss in selectivity for gold adsorption which may result from the lowering of pH or free cyanide concentration must be offset by the very much higher capacity of carbon effected by such changes.

The selective adsorption of metal cyanide complexes onto activated carbon would appear to relate to the solubility and charge of the adsorbed complex concerned. While the less soluble monovalent dicyano complexes of gold, silver and copper have all been shown to be strongly adsorbed by carbon, the more soluble divalent and trivalent tri and tetracyano complexes (e.g. Cu, Ni and Zn) are only weakly adsorbed, while the very soluble hexa and octacyano complexes (e.g. Co and Fe) are not adsorbed to any marked degree[7,8,9].

The adsorption of copper on carbon would appear to play a rather unique role in possibly aiding gold adsorption in some instances via its ability to catalyse the oxidation of free cyanide[6]. Under typical plant conditions copper is present as the $Cu(CN)_3^{2-}$ and $Cu(CN)_4^{3-}$ complexes and as such is only weakly adsorbed on carbon. On acidification or lowering of cyanide levels, however, $Cu(CN)_2^-$ is formed which is strongly adsorbed by carbon resulting in an increase in the rate of free cyanide destruction and subsequent improvement in gold adsorption efficiency. Obviously, a compromise between the beneficial effect of lower free cyanide concentrations and/or lower pH levels and the competitive adsorption of copper rather than gold will pertain[9]. The subsequent cold acid washing of such gold/copper loaded carbon in effluent treatment circuits low in free cyanide (Figure 7) is also of considerable interest when related to the chemistry of gold adsorption[8]. Under such conditions, the irreversible formation of an insoluble AuCNCuCN precipitate on the carbon followed by further gold adsorption in excess of any equilibrium loading capacity and akin to the formation of insoluble AuCN using hot acid as demonstrated by McDougall[10] is strongly suggested.

Effect of oxygen.

The role played by oxygen in gold adsorption is rather complex but nevertheless most significant. By virtue of carbon's ability to adsorb

oxygen from the adsorption medium, oxygen may be expected to play a significant role in cyanide oxidation and calcium carbonate deposition. Adsorbed oxygen may also influence the character of the surface active groups involved in gold adsorption if some sort of oxygen regeneration mechanism is operative. As may be expected, the roles played by oxygen dissolved in the adsorption medium as well as the oxygen adsorbed on the carbon surface are particularly difficult to define and quantify.

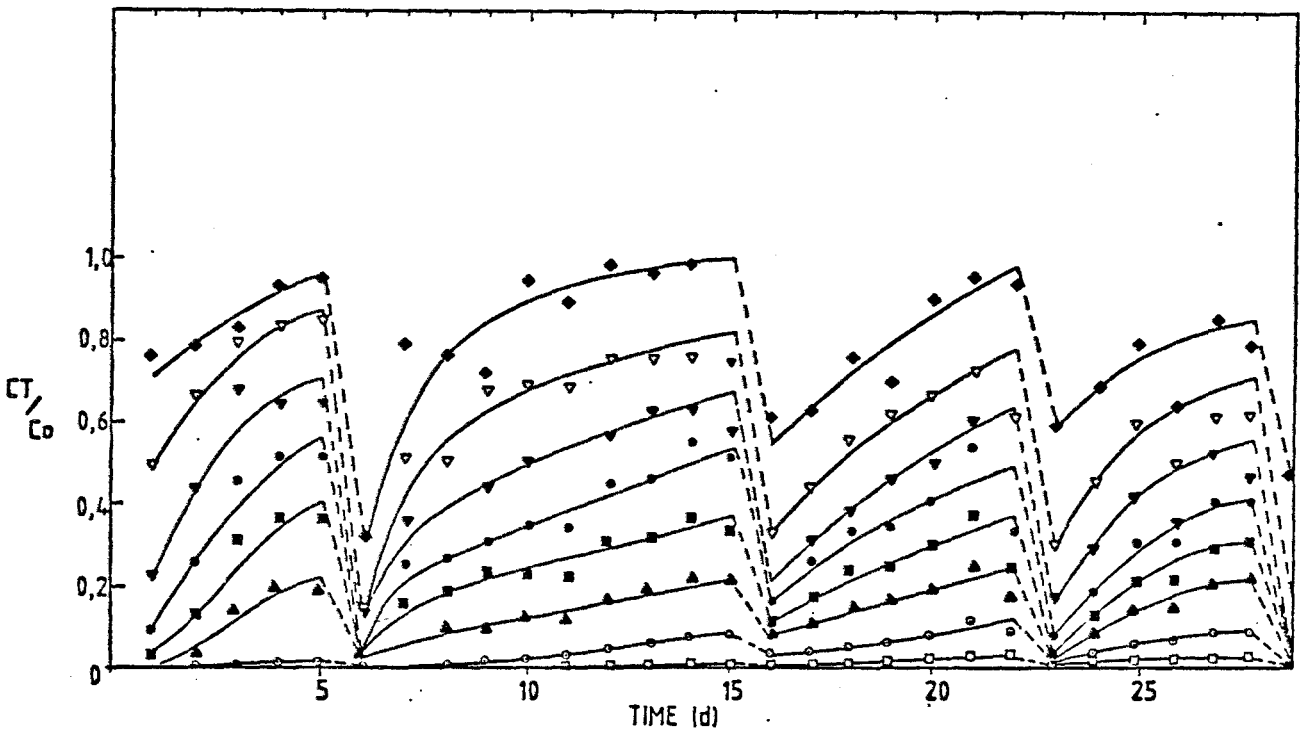

FIGURE 7

The effect of Periodic Acid Washing on the Gold Adsorption in a Fixed-bed Pilot Plant CT=gold concentration in effluent at specified depth of carbon bed: Co=gold concentration in column influent : breakthrough curves correspond to each metre of bed depth of carbon in the circuit as follows:
(FROM DAVIDSON AND STRONG [18])

Because of the complexities involved in an investigation of this nature, only limited information concerning the role played by oxygen is currently available. It has been demonstrated that gold adsorption is adversely affected on treating solutions devoid of oxygen e.g. Merrill barren solutions[11], while Tsuchida et al[5] showed that "deoxygenated" carbon displayed significantly lower gold and silver adsorption capacities when compared with normal carbon (Figure 8). Furthermore, when the "deoxygenated" carbon was exposed to oxygen bubbling through the solution, the loading of gold increased to that found on normal carbon.

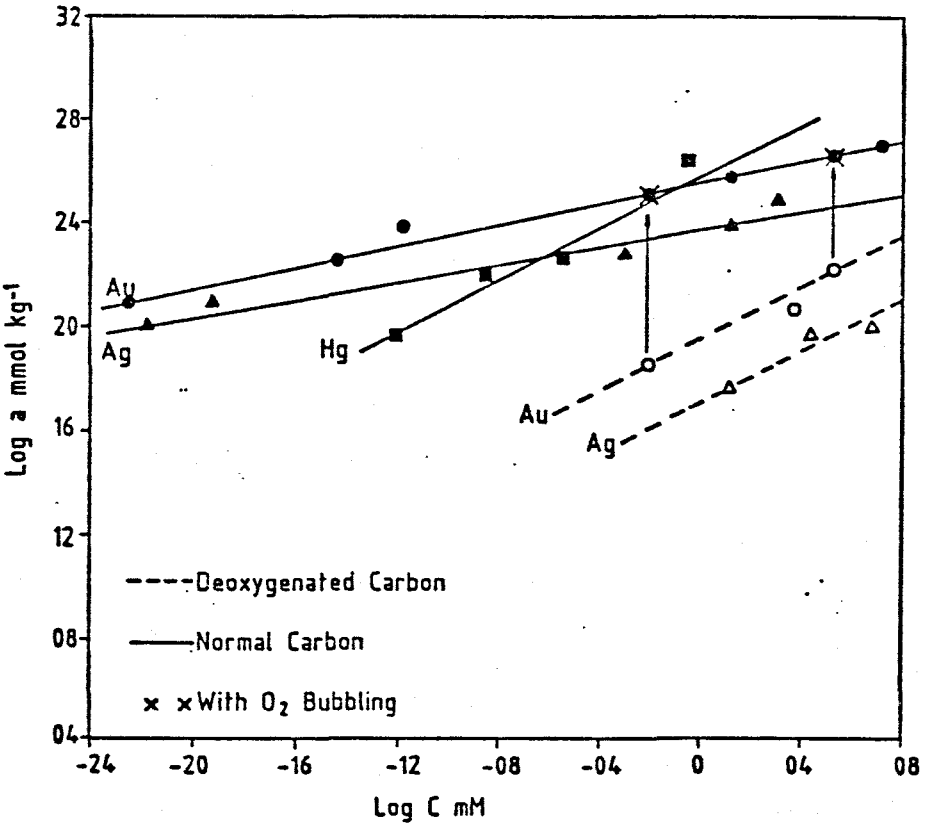

FIGURE 8

Adsorption Isotherm for loading of Gold, Silver and Mercury Cyanide Complexes onto Normal and Deoxygenated Carbon at 25°.
(FROM TSUCHIDA et al [15])

Organic fouling.

Activated carbon's ability to scavenge organic constituents from industrial effluents is well known and as such, the presence of such contaminants in a gold adsorption circuit will have some detrimental effect on gold adsorption. Recent studies[9,12,13] have identified a number of such poisons such as flotation reagents, lubricating oils and kerosene which are commonly found in gold plant circuits. Invariably, the rate of gold adsorption rather than the loading capacity of the carbon is seriously impaired by such poisons[9]. One possible mechanism of carbon "poisoning" that affects the rate of loading but not the loading capacity could be the adsorption of a film of the solvent onto the superficial surface of the carbon granule (which thereby creates a barrier to mass transfer) without significant penetration of the solvent into the pores of the granule. It is also significant that the influence of the organic solvents on the rate of loading is substantially

reduced in typical gold plant pulps[9], indicating that a portion of the organic solvents is adsorbed by the inert solids, thereby reducing the effective concentration in solution or on the carbon. The presence of the poisons in a CIP plant can be identified by a low extraction efficiency in the first few stages and, as the poison becomes adsorbed on the carbon, a gradually increasing extraction efficiency towards the latter stages of the plant. Since these poisons affect the rate of loading far more than the loading capacity, their influence can be counteracted to a certain extent simply by an increase in the concentration of carbon in those stages where the extraction efficiency is poor.

Physical effects.

In most CIP operations where the loading of gold is not allowed to approach equilibrium, film diffusion predominates in limiting the rate of gold extraction. Under such conditions, such physical aspects as mixing, pulp density, carbon particle size and carbon concentration play a most significant role and are likely to overshadow any chemical considerations. Features such as contactor design, choice of agitation and pulp density are particularly significant[9] while adsorption rates increase linearly with the reciprocal of the carbon particle size.

Conversely, in an adsorption circuit where the loading of gold does approach equilibrium, those factors influencing loading capacity become operative. As the loading in any stage approaches equilibrium, the rate of extraction decreases and therefore any factor that influences the loading capacity has a bearing on plant performance under these conditions. Such variables as ionic strength, pH value and temperature[9] are likely to play an important role.

TEST PROCEDURES

Determination of the Equilibrium Gold Adsorption Capacity Constant

Because gold adsorption onto activated carbon is most sensitive to such parameters as temperature, pH, cyanide concentration and to ionic strength, any method used for measuring the equilibrium adsorption capacity of carbon must involve the control of these parameters. The method as used at the Anglo American Research Laboratories (AARL) for the adjudication of commercial products is as follows:

1. Riffle out approximately 10 g of the carbon sample and crush the sample using a mortar and pestle before screening within the range 0,15-0,43 mm. The screened carbon is then boiled in a similar volume of 3 per cent hydrochloric acid (vol./vol.) for 10 minutes. Decant and

wash the carbon twice with equal volumes of cold distilled water before oven drying at 110°C overnight.

2. Prepare a borate buffer solution as follows:

Dissolve 15,46 g H_3BO_3 and 18,64 g KCl in approximately 4 ℓ of distilled water (solution A). Dissolve 10,00 g NaOH in 0,5 ℓ distilled water and add to solution A while stirring until pH 10,0 is reached (check with commercial buffer standard). Add 0,740 g $KAu(CN)_2$ and make up to 5 ℓ with distilled water i.e. 100 g/t Au. To every 1 ℓ of this bulk solution add 200 mg NaCN just prior to use.

3. Weigh out five portions of the prepared carbon as in (1) following cooling in a desicator, such that the masses are ca 0,10, 0,20, 0,30, 0,40 and 0,50 g respectively. The masses of each portion must be recorded to four decimal places.

4. Place each sample of carbon in a 250 mℓ conical flask add 100 mℓ of the buffered gold solution as prepared above (2). Stopper the flasks. Agitate the flasks in a thermostated shaking machine at 25°C for 20 hours at a medium speed.

5. Filter off the carbon quickly using Whatman No. 541 filter paper. After discarding the initial 10-20 mℓ filtrate, determine the concentration of gold in the filtrate in each case.

6. Calculate the mass of gold adsorbed by the carbon by means of the difference in the gold concentration between the head solution and the filtrate obtained from each sample. Determine the k value of the carbon by plotting the empirical Freundlich isotherm, $X/M = kc^{1/n}$, on log-log graph paper.

i.e. $\log X/M = \log k + 1/n \log c.$

where X = mass of gold loaded on carbon (mg)

M = mass of carbon sample (g)

c = equilibrium concentration of gold in filtrate (g/t)

k = constant defined as the carbon loading (mg Au/g C) in equilibrium with a residual solution gold concentration, c, of 1 g/t. This constant may be better expressed in kg Au/tC as the expression is more meaningful when related to a large scale CIP operation.

1/n = slope of plot as described above.

The relative standard deviation of the above method was shown to be 3,5 per cent.

Determination of Gold Adsorption Rate

As the measurement of adsorption rates are entirely empirical, many different methods may be used provided good reproducibility is attained. In this regard strict attention needs to be given to the particle size of the carbon sample and to the preparation of the carbon sample, while the control of temperature (Figure 5) is also necessary. The method developed at the Klipfontein Organic Products laboratories and adopted by AARL is as follows:

1. Screen out ca 2 g of wet carbon in the size range 1,40-1,70 mm. Remove carbon fines from the screened sample using ultrasonic cleaning until no further fines are evident (several decant washings may be necessary). After oven drying at 110°C weight out 1,000 g carbon sample for kinetic test. Moisten sample with 1 mℓ standard pH 10 borate buffer solution at least 1 h prior to use.
2. Prepare 1 ℓ of a fresh solution containing the following:

   10 mg/ℓ Au added as $K(Au(CN)_2)$
   100 mg/ℓ Ca added as $CaCl_2$
   100 mg/ℓ CN added as KCN

   Adjust the temperature of the test solution to 25°C before use.
3. Place 1 ℓ test solution in a stoppered oblong plastic container (150 mm x 150 mm x 140 mm) and add the prepared carbon sample while noting the time of addition. Immediately commence shaking on a linear shaker (Labotec Scientific Engineering (Pty) Limited) adjusted to 130 oscillations/min. The linear shaker is housed in a thermostated glove-box at 25°C. Withdraw 5 mℓ aliquots after 0, 15, 30, 45, 60, 75 and 90 min respectively for gold analysis.

The relative standard deviation of the above method was shown to be 3,4 per cent.

Determination of Attrition Resistance

Carbon hardness or attrition resistance is a most important characteristic in any gold recovery circuit. Empirical methods for the measurement of the attrition resistance such as the A.S.T.M. test involve the crushing action of a ball bearing on predried carbon. As the A.S.T.M. test bears little resemblance to actual conditions in a CIP circuit and is perhaps too drastic, more realistic conditions have been found to be more applicable. Two test procedures in common use are presented below.

The abrasion resistance method as developed by the Council for Mineral Technology involves the use of a laboratory float machine in order to simulate accelerated plant conditions. The method is as follows:

1) Sieve carbon to between 1,7-2,0 mm, wash and dry in an oven at 110°C for 16 hours.
2) Place 30 g of this washed and sized carbon in 3 ℓ deionised water contained in a 4 ℓ capacity float cell. A Denver laboratory float machine, size D-12, is fitted with a 9,5 cm stainless steel impeller and a 12,5 cm stainless steel hood.
3) With the air inlet in the open position and at a constant speed of 1200 rpm agitate the carbon and water for 60 minutes.
4) Wash carbon onto a 0,850 mm screen. Dry the +0,850 mm fraction in an oven at 110°C for 16 hours. Establish mass of this fraction and express as a percentage of the total carbon mass.

The measurement of carbon attrition as developed at the AARL makes use of a tumbling bottle as follows:

1. 50 g of oven dried carbon (overnight at 110°C) which was initially wet screened to 1,00-2,36 mm is agitated with 100 mℓ water in a 1 litre "Consol" preserve jar or bottle for a period of 24 h. Agitation is carried out by rotating the bottle, end over end, at a rate of 40 rpm (this is easily effected by wedging the bottle into a larger open ended plastic beaker or bottle and rolling on a set of rolls).
2. Following the agitation for a period of 24 h, the carbon sample is wet screened through a 1,00 mm screen and the undersize carbon recorded as a percentage mass distribution of the original carbon sample. The test is carried out in triplicate and the results averaged for specification purposes.

The relative standard deviation of the above method was 4,7 per cent.

Miscellaneous Tests

Other carbon specifications which may be required in certain instances include the following:

1. Particle size distribution. While the size distribution may be dictated to by the efficiency of the interstage screens on a particular circuit, the use of a 1,18-2,36 mm carbon is in common use in South Africa at present.
2. Bulk density. Bulk density data may be required in establishing plant balances.
3. Solubles and moisture content. For payment purposes the weight of carbon will be calculated on a water washed, dry solids basis.

4. Volatile content. The measurement of the carbon volatile content (B.S. 1016 : part 3 : 1965 : 3.4.2) provides a most useful method of maintaining quality control with regard to carbon regeneration.

CARBON SELECTION

With regard to the testing of carbon for gold adsorption, a subtle but different approach is required depending on whether virgin carbon products are being selected for a particular operation or whether used carbon is being tested in maintaining quality control. In the first instance, testing is very much an empirical procedure in which the physical and chemical properties and prices of commercial products are compared. As such, the tests concerned are relatively simple, although the weighting of the various properties is likely to be most complex. Due to this complexity, no finite formulations may be defined at the present time. Apart from the commercial price of the carbon, the chemical activity together with the physical hardness are perhaps the most important considerations in any carbon operation.

PLANT CARBON QUALITY CONTROL

Carbon quality while being perhaps the single most important aspect in any carbon operation, is very often the most neglected aspect. This situation relates particularly to the first generation CIP plants in South Africa in which in most instances the regeneration kilns and the elution columns were initially under-designed resulting in a bottleneck developing in both circuits. The consequence resulting from such bottlenecks has been the reuse of carbon that has not been efficiently eluted and furthermore not efficiently regenerated. Both aspects must be considered if meaningful quality control is to be carried out.

In many operations inefficient gold elution produces eluted carbons containing 100-300 g/t gold which if returned to the last stage of the adsorption circuit will not adsorb gold due to equilibrium rather than kinetic restraints. The gold content of the eluted and regenerated carbon is thus a most important consideration if carbon quality control is to be effective.

When on-site carbon testing is considered emphasis must be placed on the reproducibility of test procedures. The sensitivity of the carbon activity, both capacity and kinetic response, to temperature is most important if reliable control is to be attained. In this regard the testing of carbon is best carried out by the assay laboratory rather than by metallurgical staff.

It is not surprising to find that many of the chemical characteristics of eluted carbons taken from operating circuits are inter-related. Thus the

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carbon capacity and kinetic response relationship may be illustrated as in Figure 9. Likewise, the iodine number and kinetic response (Figure 10) and the volatile content and kinetic response (Figure 11) may also exhibit some relationship. The volatile content of carbons in a particular circuit would also appear to be a very simple means of assessing organic contamination and the efficiency of thermal regeneration.

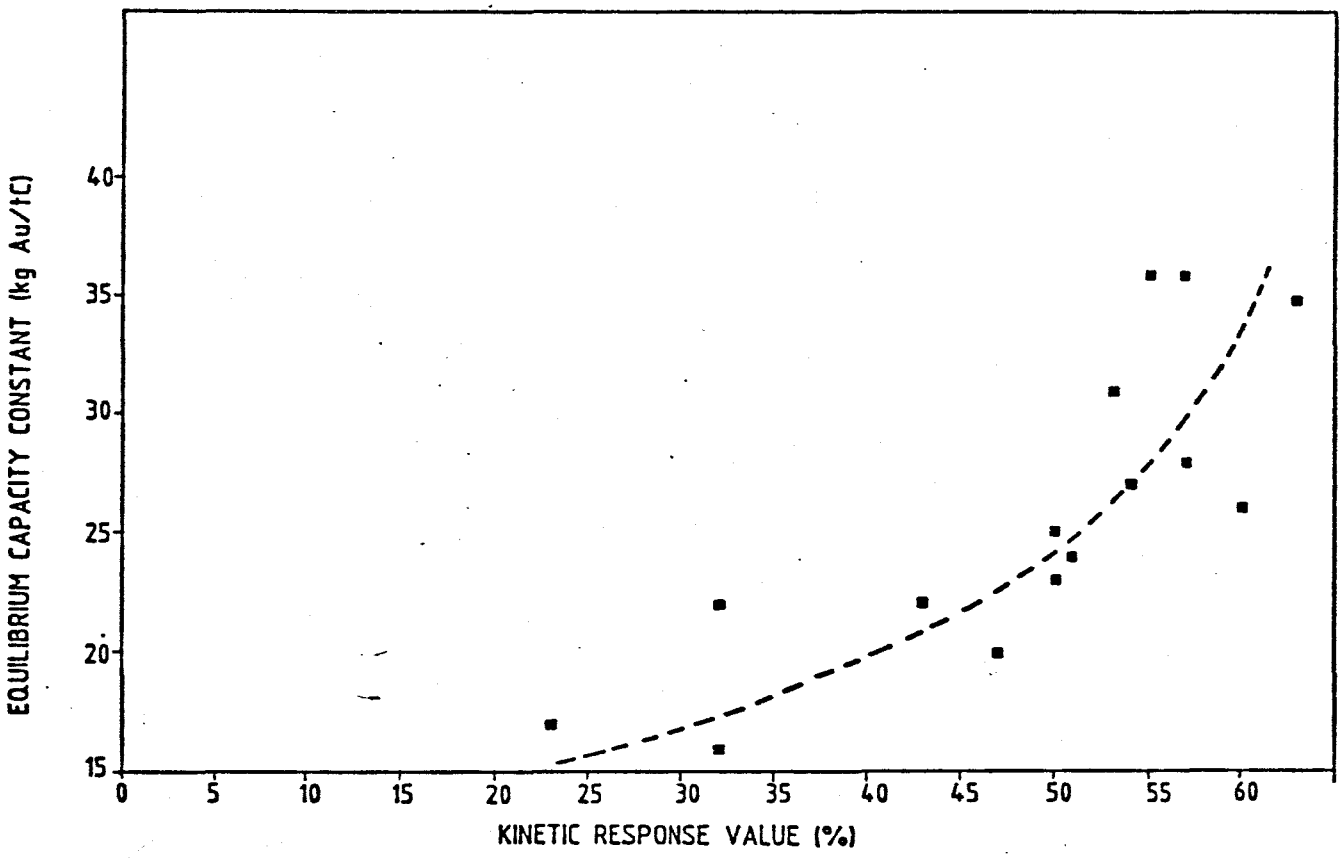

FIGURE 9

Correlation between the Capacity Constant and the Kinetic Response Value of Eluted Carbon Plant Samples.

With regard to the quality control of carbon on existing CIP plants, no general test procedures or specifications are currently in use. Each operating plant would appear to make use of some carbon testing although in many instances no quality control is exercised. In view of this situation, it is not easy to comment on quality control per se, suffice to say that in general it is sadly lacking in most operations, and if carried out at all is not used to great effect. The under-design and inefficiency in the elution and regeneration circuits is responsible in part at least for such short comings.

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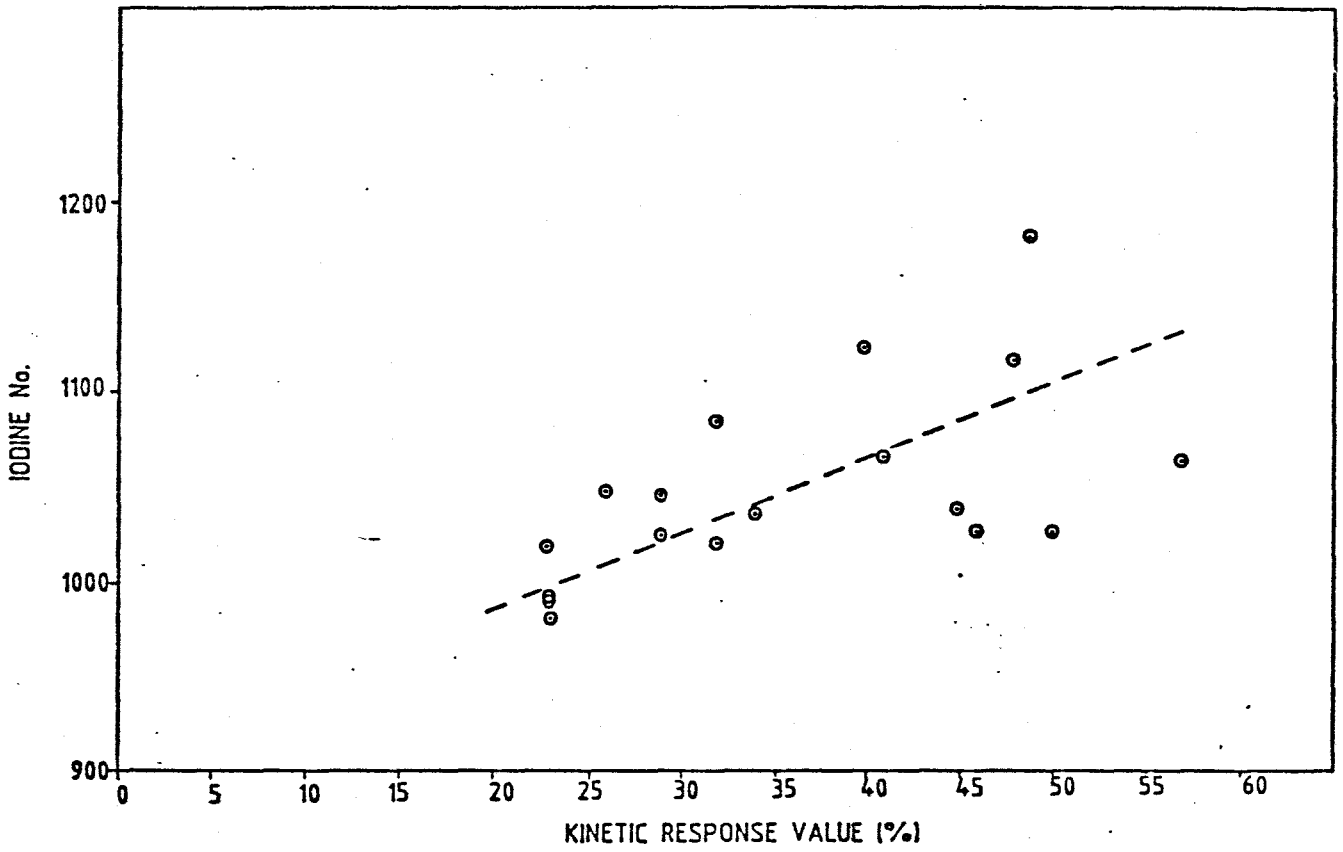

FIGURE 10
Correlation between the Iodine No. and the Kinetic Response Value of Eluted Carbon Plant Samples.

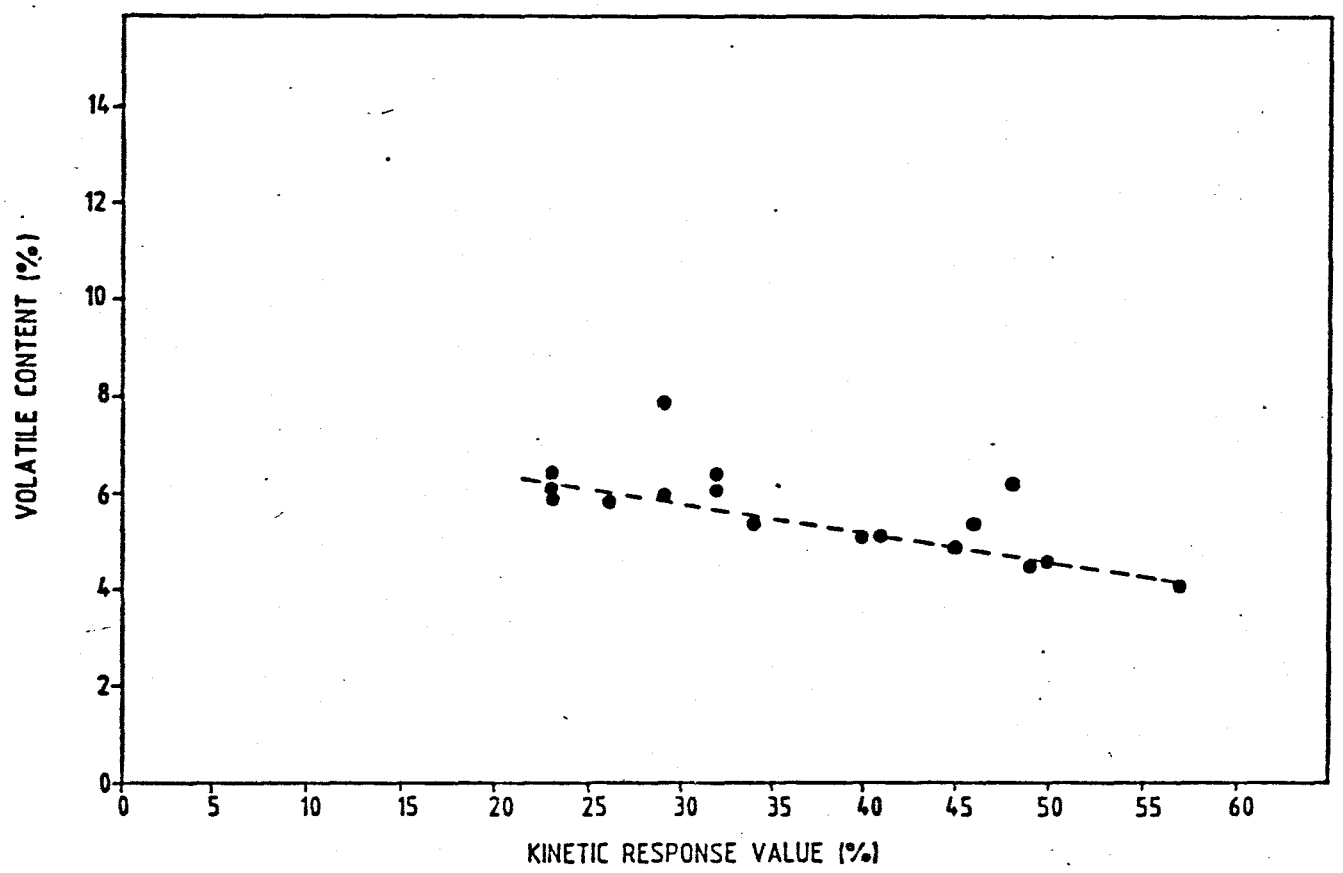

FIGURE 11
Correlation between the Volatile Content and the Kinetic Response Value of Eluted Carbon Plant Samples.

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CARBON-IN-PULP TECHNOLOGY FOR THE EXTRACTION OF GOLD

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THE SELECTION OF GRANULAR ACTIVATED CARBON FOR USE IN A CARBON-IN-PULP OPERATION

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ABSTRACT

With a view to characterising activated carbons with respect to their ability to adsorb gold from cyanide solutions, an attempt was made to establish correlations between the equilibrium capacity constant of the activated carbon for gold (k value) and five other parameters. These parameters were percentage ash content, iodine number, methylene blue value, surface area, and pH respectively.

Twenty four different proprietary brand activated carbon products were examined, the correlations being established by means of linear regression analysis. It was shown that a significant correlation only existed between k and pH value. This was interpreted in terms of the presence of oxide complexes on the activated carbon surface.

Further test specifications as developed at the Anglo American Research Laboratories are also outlined with a view to the selection of activated carbons suitable for use in a carbon-in-pulp operation. Methods for assessing both the capacity and the kinetic response of carbons for gold adsorption are presented, while the measurement of attrition resistance, particle size distribution and moisture and solubles content is also outlined. The measurement of the capacity constant for gold adsorption together with attrition resistance are considered to be necessary in the initial selection of a suitable carbon.

THE SELECTION OF GRANULAR ACTIVATED CARBON FOR USE IN A CARBON-IN-PULP OPERATION

INTRODUCTION

Due largely to the increase in the gold price over the past decade, the South African gold miming industry is currently undergoing a major swing towards the carbon-in-pulp process for the extraction of gold[1]. Thus it is, with the advent of several plants being currently commissioned that a large number of activated carbon samples have been tested in the Anglo American Research Laboratories over the past few years. As a very wide range of gold adsorption capacity data were found, it was thought worthwhile to attempt to relate these values to other carbon characteristics commonly used in the treatment of industrial effluents. This approach may be profitable in two ways. First, it may identify those parameters that exert a significant influence on the capacity of a carbon to adsorb gold. At the present time little is known of such parameters[2, 3]. Second, some insight may be gained into the nature of the active sites responsible for gold adsorption. The long term objective of this work is to understand the nature and mechanism of the formation of activity for gold adsorption sufficiently well to define the requirements for the manufacture of carbons of high capacity.

Activated carbons are made by heating organic materials in a mildly oxidizing atmosphere. In the course of activation the carbon develops a cellular structure with an extended surface area. In many ways activated carbon is a unique material with properties not possessed by any other class of organic substance. Since activated carbon usually contains considerable quantities of adsorbed oxygen and hydrogen the term carbon is perhaps a misnomer. It is essential to regard activated carbon as a complex organic polymer and not merely as an amorphous form of the element carbon. Activated carbon differs from other organic polymers in having an exceptionally large surface area and in being a good conductor of electricity. Many of its intriguing characteristics stem from the latter fact, as well as being related to the various functional groups

present on the carbon surface[4-8]. Its structure has often been compared with that of graphite and considered to be microcrystalline. A structure consisting of stacks of flat, aromatic, and for the most part heterocyclic highly cross-linked planes, has also been suggested[9].

When related to the adsorption of gold and the carbon-in-pulp process, the selection of suitable granular carbons presents a complex and sometimes confusing picture. Not only is the mechanism of gold adsorption and the functional groups taking part in the reaction on the carbon surface not well understood at the present time, but little selection other than by trial and error is documented in the literature. Furthermore, the technology itself is presently in a state of development. It is in this regard that the test specifications as formulated at the Anglo American Research Laboratories are presented in this publication. As may be expected the test procedures are provisional and largely empirical and designed primarily as a means of adjudicating commercial tenders within the Anglo American Corporation. As such the individual test procedures do not reproduce actual plant conditions and thus afford only a preliminary means of selecting a suitable carbon. Further on-site piloting would be recommended in most instances prior to final carbon selection.

THE RELATIONSHIP BETWEEN GOLD ADSORPTION ACTIVITY AND SOME OTHER PROPERTIES OF ACTIVATED CARBONS

PROGRAMME

Materials. 24 samples of activated carbon from a number of industrial suppliers were examined with regard to pH, gold adsorption, iodine number, methylene blue value, ash content and surface area. The relationship between the parameters was then examined by means of linear regression analysis.

Methods. Methods for the determination of the equlibrium capacity constant, k, for gold adsorption in borate buffer, iodine number, methylene blue value and surface area have been previously described[10-11]. The determination of ash content was performed according to the recommendations of Norit nv-Holland[11].

The pH of a water extract of each carbon was determined as described below. The method is based on procedures suggested by Norit (Holland) and Water Research Association (England).

2,0 g Carbon were mixed with 100 mL distilled water (ph = 5,5) in a beaker and agitated mechanically for 10 minutes at room temperature. The carbon was then immediately filtered-off and the pH of the filtrate measured to the nearest 0,1 ph. This technique affords a simple and reproducible method for the comparison of the relative "buffering capacity" of carbons.

The variation of pH with duration of wash was determined in the case of two carbons:

Time Min.	pH ± 0,05	
	A	B
1	9,8	2,7
5	9,9	2,6
10	9,9	2,6
20	9,9	2,6

Experimental evidence suggesting that pH may be related to acid/ base adsorption capacity was also derived. This was done by carrying out neutralization (by titration) of supernatant liquid obtained by boiling carbon in water for 5 mins. Neutralization (pH = 7) was obtained by addition of 0,1N NaOH or HCl according to the method described by Norit[11]. The results are displayed below:

Sample#	k (mg Au/gC)	pH	mg-equivalent acid*/base+
E	43	10,3	0,05*
X	38	9,6	0,03*
N	23	7,8	0,008*
G	2	2,3	0,16+

refer to Table I.

It is seen that a carbon of high pH has a correspondingly high capacity for acid adsorption, while a low pH carbon has a high capacity for base adsorption.

Data analysis. The data were examined by means of linear regression analysis whereby the correlation between the activity coefficient, treated as the dependent variable, and every other parameter taken singly was determined. Following this, the correlations between all pairs of parameters was also determined. The measure of the correlation was reflected by the correlation coefficient; the statistical significance of any correlation being established by determining the minimum value of the coefficient that is required for the correlation to attain a 95 per cent level of confidence. This minimum value was obtained from Student's t-test. It is pertinent to note here that statistical significance is not synonymous with a genuine physical-chemical relationship between the parameters examined. If reasonable grounds exist for the occurrence of such a relationship, then the correlation may indeed be physically significant and the square of the correlation coefficient may be interpreted as a measure of what fraction of the variance of the dependent variable is explained by the correlation.

RESULTS

Table I lists the values of the parameters measured for all 24 carbons. It is of interest to display the variation of k with surface area (Figure 1). The plot is essentially a scatter diagram and no obvious correlation is evident.

TABLE I - PARAMETERS USED FOR THE CHARACTERIZATION OF COMMERCIAL ACTIVE CARBONS

Sample	k (mg Au/g-C)	Ash (%)	Iodine No.	Methylene blue No.	Surface Area (m^2/g)	pH Value
A	21	16,4	591	136	631	5,9
B	9	0,3	533	14	484	5,8
C	9	2,7	891	29	1495	5,4
D	16	29	525	12	555	9,1
E	43	5,7	580	12	978	10,0
F	20	12,3	538	11	478	9,0
G	2	4,7	466	27	1216	2,7
H	24	3,6	247	3,0	684	9,4
I	41	7,2	172	22	1311	8,6
J	29	6,7	147	24	1861	8,2
K	45	5,6	294	20	890	9,9
L	26	2,9	405	24	1632	9,2
M	5	3,5	975	28	1459	5,8
N	23	8,5	313	29	1567	7,4
O	0	7,8	464	14	531	5,8
P	16	10,0	225	9,5	684	5,4
Q	7	1,9	138	5,3	1014	9,2
R	36	2,6	283	13	1317	10,1
S	29	6,1	980	28	1233	7,3
T	37	5,6	472	21	1008	6,7
U	35	5,1	275	22	902	6,5
V	38	4,7	194	18	1090	7,9
W	38	4,6	225	13	860	10,1
X	38	5,5	127	25	1108	7,6
Y*	27	8,6	309	2,5	513	10,9

* Prepared in Laboratories from wattle charcoal by steam activation at 800°C.

Table II lists the correlation coefficients obtained, first, between k and the other parameters, and, second, between pairs of parameters. The statistical significance (95 per cent level) of the correlation coefficient for 24 degrees of freedom is 0,40.

TABLE II - LIST OF CORRELATION COEFFICIENTS (r) BETWEEN PAIRS OF PARAMETERS OF ACTIVE CARBON

Dependent Variable	Independent Variable	r
k	Ash	-0,074
k	Iodine No.	-0,393
k	Methylene Blue	-0,133
k	Surface Area	0,137
k	pH	0,598
Ash	Iodine No.	0,077
Ash	Methylene Blue	0,807
Ash	Surface Area	-0,384
Ash	pH	0,072
Iodine No.	Methylene Blue	0,258
Iodine No.	Surface Area	0,051
Iodine No.	pH	-0,374
Methylene Blue	Surface Area	-0,153
Methylene Blue	pH	-0,145
Surface Area	pH	-0,083

Value of r is unchanged when dependent and independent variables are interchanged.

The only variable which emerged as being significantly correlated with k is the pH value. Approximately 30 per cent of the range of values of k can be ascribed to the variation of pH. As expected from Fig. 1, no significant correlation was found between k and surface area (r = 0,137 and r^2 = 0,019). There was one other significant correlation, viz

between the variables ash content and methylene blue value. The plot of k vs pH is shown in Figure 2.

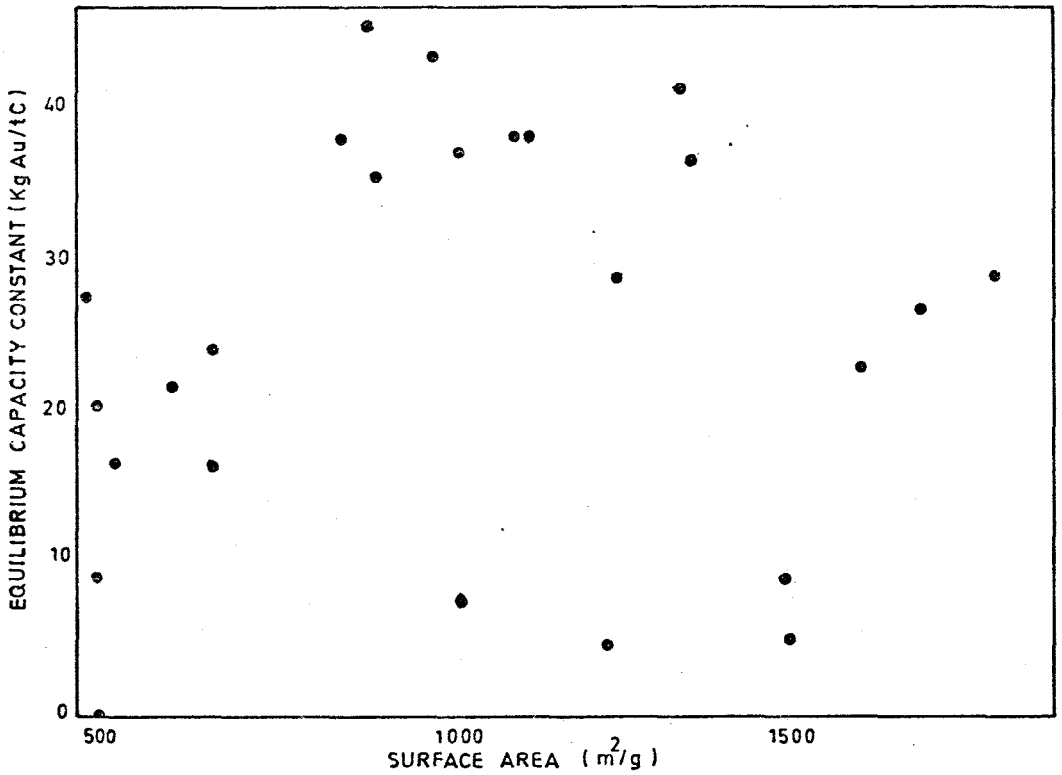

FIGURE 1 The Effect of Carbon Surface Area on the Adsorption of Gold

As the relationships between parameters may be perturbed to an unknown degree by the various types of carbonaceous substrates out of which the active carbons were prepared, it was decided to re-examine the correlations in the case of one group of carbons made from peat. The results of this analysis are presented in Table III. It is seen that whereas the correlation coefficients between k and surface area and pH are the same as for all the carbons taken as groups (Table II), the composite parameter pH-surface area correlates very well with k (Figure 3). It appears to account for 85 per cent of the variation of k. However, when applied to the whole group of carbons, this parameter did not yield an improved value of the correlation coefficient. Thus the true significance of pH-surface area is ambiguous. The correlation displayed in Figure 3 may be fortuitous; or it may be interpreted to show the composite parameter is of significance only when applied to carbons that are similar i.e. manufactured from a common starting material.

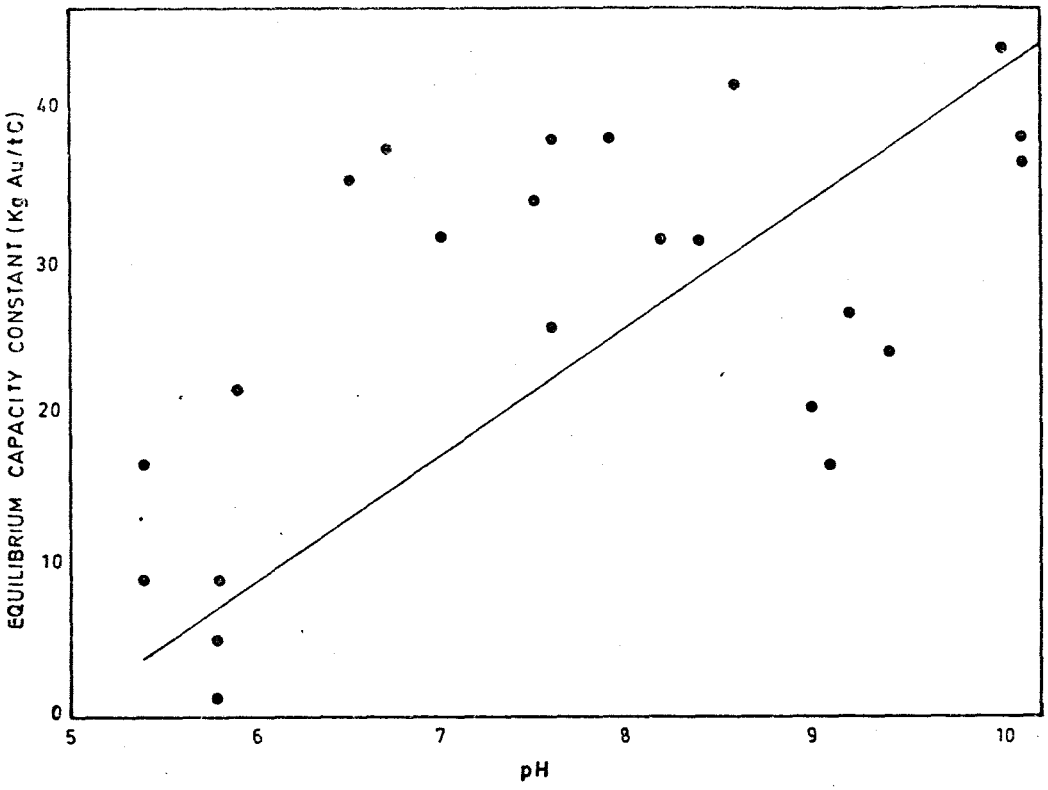

FIGURE 2 The Effect of Carbon "pH" on the Adsorption of Gold

TABLE III - CORRELATIONS BETWEEN SOME PARAMETERS OF CARBONS MADE FROM A COMMON STARTING MATERIAL

Parameters	r	r^2
Surface Area - k	0,129	0,017
pH - k	0,759	0,58
k - Surface Area . pH	0,923	0,85

DISCUSSION OF RESULTS

A correlation study such as this is based on two implicit assumptions. First, the dependent variable (k, in this case) can be uniquely described in terms of the parameters considered. Second, the parameters are the only significant ones. The first assumption is justifiable since, even in the absence of a unique relationship, the parameters that exert a significant influence on k may be identified. The second assumption is less secure, and it is quite probable that an accurate description of k

requires prior knowledge of at least the temperature and atmosphere of activation. In so far as the two assumptions are valid, some conclusions can be made.

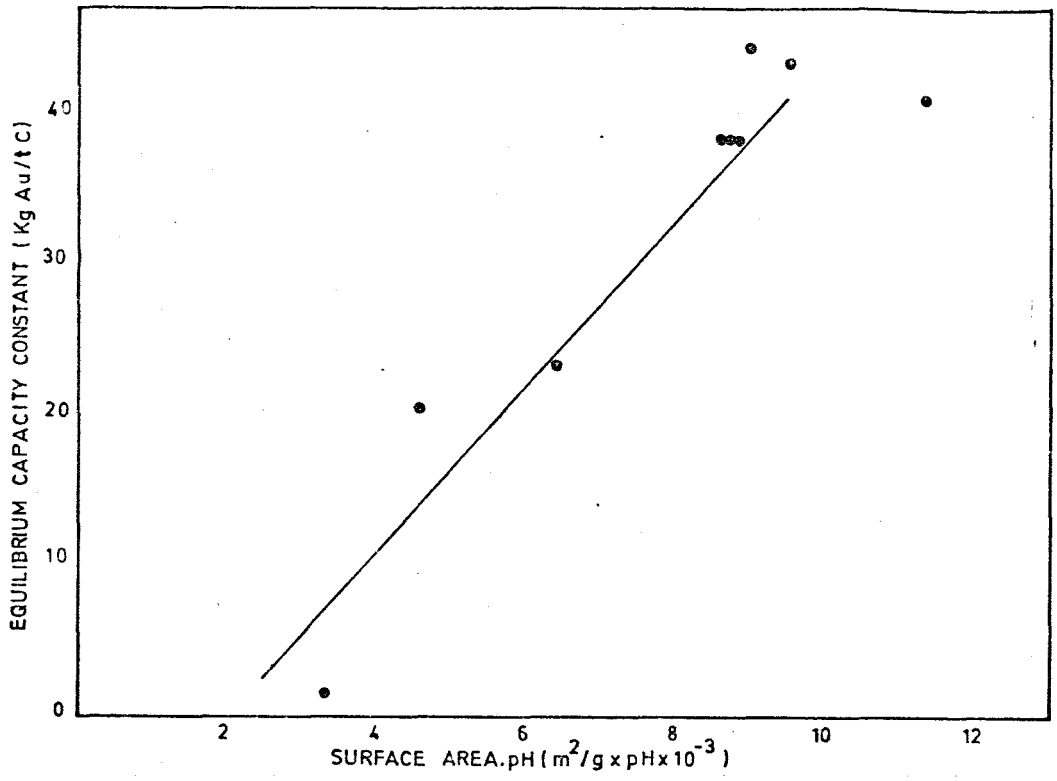

FIGURE 3 The Correlation Between Gold Adsorption and the Function of Surface Area and pH

1. The lack of any correlation between surface area and k is perhaps surprising. Activation is primarily associated with the creation of carbons of high surface area and porosity. It is seen that these parameters do not imply good adsorptive characteristics for gold in solution.

2. The correlation of k with pH is significant not only statistically but physically as well, since pH can be shown to influence the adsorptive properties of carbon. This effect arises through the presence of surface oxide groups.

   Oxygen may constitute up to 25 per cent by weight of an activated carbon, at least half of it being present in the form of surface groups, on the extended hexagonal microcrystalline lattice of

activated carbon[12]. That these groups can posses "basic" and "acidic" characters is well established[13]. Furthermore, it is known that the varying acid-base character of carbon surfaces is determined in the main by the temperature of activation.

Carbon prepared at high activation-oxidation temperatures ($> 500^{o}C$), known as H-carbons, adsorb primarily H ions from solutions, while carbons prepared at $< 500^{o}C$, known as L-carbons, adsorb hydroxyl ions.

The importance of the surface oxide groups lies in the fact that they impart polarity to the carbon surface. Thus sorption of non-polar species is impaired, but polar species may be sorbed to a greater extent, depending on the specific oxide-sorbate interaction[14].

It is concluded that, everything else being equal, the formation of H-type surface oxide groups is beneficial to the subsequent adsorption of the aurocyanide anion from solution. It is further concluded that the manufacture of carbons of high k-value by thermal activation requires $T > 750^{o}C$ since the adsorption of H-ions has been determined to reach a maximum in the vicinity of $800^{o}C$[15].

3. The observed correlation between ash content and methylene blue values may also have physical significance. The latter parameter is a measure of the adsorptive properties of carbon towards large molecules[11]. If this is assumed to depend on the presence of pores of large size, it follows that impurities in the carbon promote the formation of these pores. Pore structure results from the gasification of carbon by the activation atmosphere; the carbon surface is heterogeneous and all points of it are not equally amenable to gasification. The effect of impurities is to enhance the rate of gasification in their vicinity, and thereby promote the preferential removal of carbon at some point, but not at others. It follows that some measure of control over the pore structure can be effected by using impure carbons (i.e. not sugars or coconut shell) or by seeding the surface with metals that catalyze the gas-carbon reaction.

The above findings have both fundamental as well as practical significance. From the more fundamental point of view, a better understanding as to the nature of gold adsorption onto activated carbon has been established, while some of the requirements for the production of an H-type carbon with a high capacity for gold adsorption are indicated.

PROVISIONAL SPECIFICATIONS FOR ACTIVATED CARBON SUITABLE FOR USE IN A CIP RECOVERY CIRCUIT

INTRODUCTION

Because of the complex biophysical and biochemical character of activated carbon, the preliminary selection of a suitable carbon in quantifiable terms presents many difficulties. Perhaps the primary requisites of a good carbon are firstly physical stability and secondly high gold loading capacity. Both of these characteristics are however generally antipathetic due primarily to the very nature of the activation process i.e. activation at high temperatures is accompanied by a loss in physical strength. This situation immediately imposes some measure of compromise in the weighting of such properties.

Apart from the commercial price of the carbon concerned, other properties such as rate of adsorption, the particle size distribution and the moisture and soluble content (also considered in the purchase price) must be considered in selecting a suitable product. Perhaps even more important is the quality control peculiar to a product - this aspect is again most difficult to assess within a short time limit. Again, one particular carbon may respond better than another depending on the type of material being treated and only actual on-site testing would afford a means of quantifying this possibility.

Bearing in mind the complexities described above, the provisional specifications as developed at the Anglo American Research Laboratories may be outlined. While the specifications are used primarily for the adjudication of tenders within the Anglo American Corporation, modified

procedures may also be used for monitoring the quality of thermally regenerated carbon on existing CIP plants. Obviously such methods will involve simplified versions of such tests as the adsorption capacity and perhaps the attrition resistance and size distribution tests outlined below.

DETERMINATION OF THE EQUILIBRIUM GOLD ADSORPTION CAPACITY CONSTANT (k value)

1. Riffle out approximately 100 g of the carbon sample and then boil such a representative sample in a similar volumne of 3 per cent hydrochloric acid (vol./vol.) for 10 minutes. Decant and wash the carbon twice with equal volumnes of cold distilled water. The carbon is then crushed using a mortar and pestle and screened to within the range 0,15-0,43 mm. This screened carbon is then oven dried at 110 ^{o}C for 8 hours.

2. Prepare a borate buffer solution as follows:

   Dissolve 15,46 g H_3BO_3 and 18,64 g KCl in approximately 4 L of distilled water (solution A). Dissolve 10,00 g NaOH in 0,5 L distilled water and add to solution A while stirring until pH 10,0 is reached (check with commerical buffer standard). Add 0,740 g K $Au(CN)_2$ and make up to 5 L with distilled water i.e. 100 g/t Au. To every 1 L of this bulk solution add 0,200 g NaCN just prior to use.

3. Weigh out five portions of the prepared carbon as in (1) following cooling in a desicator, such that the masses are ca. 0,10, 0,20, 0,30, 0,40 and 0,50 g respectively. The masses of each portion must be recorded to four decimal places.

4. Place each sample of carbon in a 250 mL conical flask and add 100 mL of the buffered gold solution as prepared above (2). Stopper the flasks. Agitate the flasks in a thermostated shaking machine at 25^{o}C for 20 hours at a medium speed.

5. Filter off the carbon quickly under vacuum using Whatman No. 541 filter paper and determine the concentration of gold in the filtrate in each case.

6. Calculate the mass of gold adsorbed by the carbon by means of the difference in the gold concentration between the head solution and the filtrate obtained from each sample. Determine the k value of the carbon by plotting the empirical Freundlich isotherm, $X/M = kc^{1/n}$, on log-log graph paper.

   i.e. $\log X/M = \log k + 1/n \log c$.

   where X = mass of gold loaded on carbon (mg)
   M = mass of carbon sample (g)
   c = equilibrium concentration of gold in filtrate (g/t)
   k = constant defined as the carbon loading (mg Au/g C) in equilibrium with a residual solution gold concentration, c, of 1 g/t. This constant may be better expressed in kg Au/tC as the expression is more meaningful when related to a large scale CIP operation.
   1/n = slope of plot as described above.

DETERMINATION OF GOLD ADSORPTION RATES

1. The uncrushed carbon is wet screened as received to its designed specification (1,00-2,36 mm) and then oven dried at 110 °C for 8 hours.

2. Prepare 5 L borate buffer solution as described above but add only 0,074 g K $Au(CN)_2$ prior to making up to 5 L i.e. 10 g/t Au. To every 2 L of this bulk solution add 0,400 g NaCN just prior to use.

3. 1700 cm^3 of the gold/buffer solution is transferred to a 2 L beaker provided with a Perspex baffle (Figure 4) and the beakers placed in a water bath controlled at 25 °C and allowed to equilibrate at this temperature. The gold solution is agitated by means of an anchor-shaped stirrer rotating at 500 rpm (accurately measured with a tachometer) which is positioned 5 mm above the bottom of the 2 L beaker (Figure 5). Solution turbulance is effected by means of the Perspex baffle.

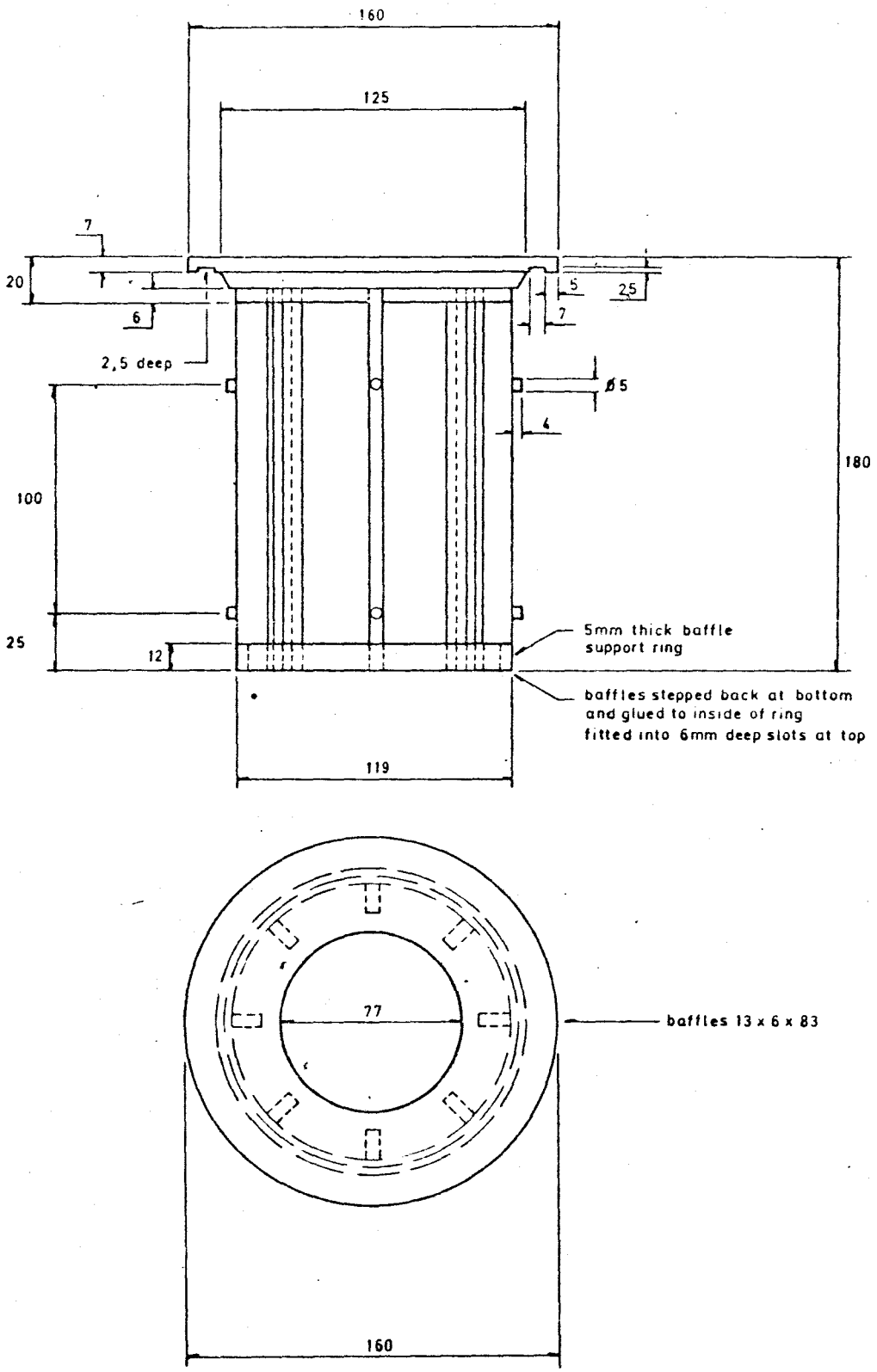

FIGURE 4 The Baffle Configuration used in the Determination of Gold Adsorption Rates

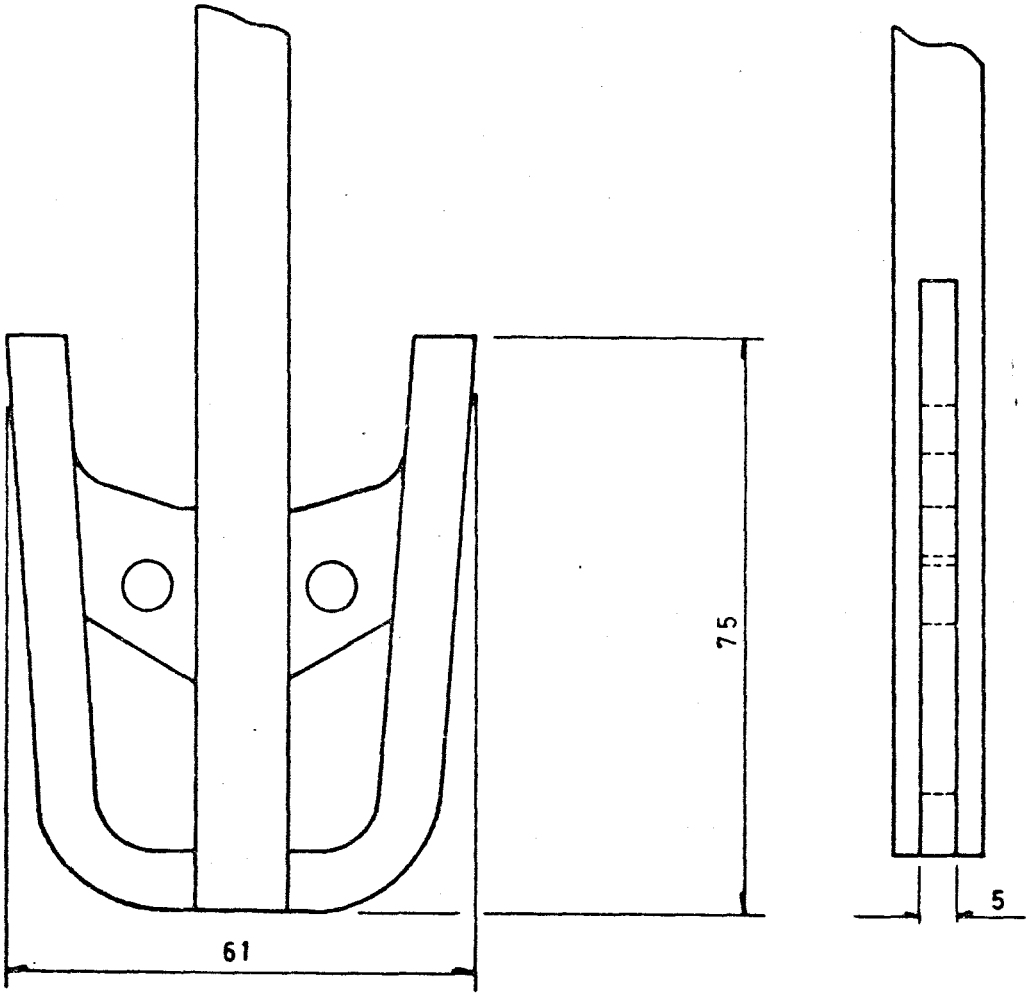

FIGURE 5 The Anchor Shaped Stirrer Used in the Determination of Gold Adsorption Rates (actual size)

4. Weigh out 5,000 g of the oven dried carbon (described above) after cooling in a dessicator and add it to the gold/buffer solution in the stirred beaker while noting the time.

5. Pipette out solution samples at specified time intervals following the addition of the carbon sample:

0,5 ; 1 and 2 hours (2 mL sample)
3 ; 4 and 5 hours (5 mL sample)
22 ; 26 and 30 hours (20 mL sample)

6. Analyze solution samples for gold and calculate the gold loading of the carbon at each time interval as follows:

$$\text{carbon loading} = \frac{Co - Ct}{m} \times V \quad \text{mg Au/gC}$$

where Co = initial solution gold concentration (g/t)
Ct = solution gold concentration after time, t (g/t)
V = solution volume (L)
m = mass of carbon used (g)

7. Obtain a straight line graph by plotting t/carbon loading vs t. Using a linear regression calculate the intercept on the y axis - the reciprocal of the value gives the rate constant at zero time "r" while the reciprocal of the slope of the linear plot gives "m" the loading of the carbon at infinite time.

DETERMINATION OF SIZE DISTRIBUTION

1. A 100g representative sample of dry activated carbon is screened using the following set of screens:-

From top to bottom:-

Equivalent Tyler Mesh Number	Opening in mm
8	2,36
10	1,70
12	1,40
14	1,18
16	1,00
20	0,85

2. The sample is screened on a Ro-Tap screening machine for 30 minutes. Any carbon particles which have stuck in the screen apertures are removed and added to the carbon that was retained by that particular screen.

3. The mass of carbon retained on each screen is expressed as a percentage of the total.

DETERMINATION OF ATTRITION RESISTANCE

1. 50 g of oven dried carbon (overnight at 110 °C) which was initially wet screened to 1,00-2,36 mm is agitated with 100 mL water in a 1 litre "Consol" preserve jar or bottle for a period of 24h. Agitation is carried out by rotating the bottle, end over end, at a rate of 40 r.p.m. (this is easily effected by wedging the bottle into a larger open ended plastic beaker or bottle and rolling on a set of rolls).

2. Following the agitation for a period of 24h, the carbon sample is wet screened through a 1,00 mm screen and the undersized carbon recorded as a percentage mass distribution of the original carbon sample. The test is carried out in triplicate and the results averaged for specification purposes (see below).

DETERMINATION OF SOLUBLES AND MOISTURE CONTENT

1. A representative sample of the carbon is accurately weighed in the "as received" form. The carbon is then transferred to a sintered glass crucible covered with distilled water and allowed to stand for 10 minutes.

2. The water is then filtered off under suction and the carbon covered with a fresh quantity of distilled water.

3. Step 2 is repeated until the carbon has received 5 washes.

4. The washed carbon is then transferred to a tared Petri dish of such size that the depth of carbon is not greater than 10 mm. The Petri dish is then placed in a ventilated oven and kept at 110 °C for 24 hours.

5. After cooling in a dessicator the dish, with carbon, is weighed. The dry weight of carbon is then expressed as a percentage of the "as received" weight.

SPECIFICATIONS

With reference to the prescribed testwork, the following provisional specifications for granular coconut shell activated carbon suitable for use in a CIP recovery circuit have been formulated:

Gold adsorption capacity. The Freundlich constant k in equilibrium with 1 g/t gold in solution shall be greater than 30 kg Au/tC.

Gold adsorption rates. The Langmuir reaction rate constant at zero time, r, shall be greater than 15 kg Au/tC/min, while the rate constant at infinite time, m, shall be greater than 12 kg Au/tC.

Size Distribution. The carbon is to be in the size range 1,00-2,36 mm. The maximum allowable fraction outside this size range shall be 3% by mass of the carbon.

Attrition resistance. The generation of fines following the tumbling bottle test shall average less than 2% by mass of the carbon.

Moisture and solubles content. For payment purposes the weight of the carbon will be calculated on a water washed, dry solids basis.

ACKNOWLEDGEMENTS

This paper is published with the permission of the management of the Anglo American Corporation. The author is indebted to Messrs. R.R. Liebenberg, M. Nkosi and T. Szendrei for various experimental data. Thanks are also due to the analytical laboratory of Anglo American Research Laboratories for the chemical analysis. Funding of the project by the Gold Division of the Anglo American Corporation is also gratefully acknowledged.

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ANGLO AMERICAN RESEARCH LABORATORIES

STANDARD ANALYTICAL PROCEDURES FOR EVALUATION OF ACTIVATED CARBON

AARL REFERENCE CA 68

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JOHANNESBURG
31st July, 1987
BG/mvdb/1

THE DETERMINATION OF ATTRITION RESISTANCE OF ACTIVATED CARBON, MESH SIZE 8 x 16 AND 6 x 12

1.0 PRINCIPLE

The activated carbon is agitated in the presence of water on a set of rollers. The fines generated during the tumbling bottle test, after 24 hours, are measured.

2.0 APPARATUS AND REAGENTS

2.1 A drying oven.

2.2 A set of rollers with timer.

2.3 A top loading balance, capable of measuring to 1 mg.

2.4 A 1 ℓ "Consol" preserve jar with lid.

2.5 A 5 ℓ plastic beaker (squat form).

2.6 A glass-fibre filter membrane : Gelman type GF/C was found to be suitable.

2.7 Suitable sets of standard sieves.

2.8 A desiccator.

2.9 Deionized water.

3.0 PROCEDURE

1. Screen through 1,19-2,38 mm sieves, 150 g of dry carbon (Note 5.1).

2. Thoroughly wash the carbon collected on the 1,19 mm sieve, to remove dust and small particles.

3. Dry at 105°C to constant mass. Cool in desiccator (Note 5.2).

4. Measure 50,00 g of the dried sample (Note 5.3). Place in the "Consol" jar and add 200 mℓ of water (Note 5.4). Seal and wedge the jar in the plastic beaker in a top/bottom, side/side manner to ensure an end-over-end rotation.

5. Place the sample on the set of rollers. Agitate at rate of 40 r.p.m. for a period of 24 hours.

6. Place a sieve of 1,19 mm size over a 2 litre beaker (Note 5.5) and transfer the wet carbon, from the jar, onto the sieve. Wash the carbon thoroughly with water (1,5 ℓ of water has been found to be sufficient).

7. Filter the washings, collected in the 2 ℓ beaker, through a GF/C filter membrane.

8. Transfer the GF/C membrane, with the fines of carbon, to the oven and dry, to constant mass, at 105°C.

9. Cool in a desiccator and measure the mass.

10. Perform the test in duplicate, and report the arithmetical mean of the results (Note 5.6).

4.0 CALCULATION

Calculate attrition resistance, as carbon loss (fines):

$$\text{Carbon fines (\%)} = \frac{A-B}{C} \times 100$$

Where:

A is the mass of filter membrane with carbon fines (g)
B is the mass of filter membrane (g)
and C is the mass of the dry carbon sample (50 g in the test, as described).

The statistical data for the test are : SD = 0,09% and RSD = 4,7%.

5.0 NOTES

5.1 When activated carbon of mesh size 6 x 12 is analysed, the sample is screened through 1,68-3,36 mm sieves.

5.2 Before starting the test, ensure that the desiccator and its drying material are in good condition.

5.3 Dry carbon is highly hygroscopic. An effort should be made to minimize the time taken for sample measurement, as the carbon mass increases rapidly on exposure to the atmosphere.

5.4 A large volume of water is used to prevent distortion of the result, due to high water adsorption.
Carbons from different manufacturers vary in their water adsorption capacity.

5.5 A sieve of 1,68 mm mesh size is used, when activated carbon of 6 x 12 is analysed.

5.6 If the difference between the duplicate results is greater than 0,2% the test should be repeated.

THE DETERMINATION OF GOLD ADSORPTION CAPACITY (K-VALUE) OF ACTIVATED CARBON

1.0 PRINCIPLE

The material is ground, and the carbon sieved through 0,43 mm mesh size is collected on 0,15 mm mesh (Note 7.1).

The activated carbon, after acid washing, is contacted with 100 mg/ℓ Au solution for 20 hours at 25°C.

A five point isotherm, in the range of 0,1 to 0,5 grams of carbon per 100 mℓ of a 100 mg/ℓ gold solution in the presence of cyanide, is plotted.

The gold equilibrium constant is calculated, using the Freundlich isotherm relationship: $X/M = kC^{1/n}$.

The X/M value defines the quantity of Au (mg) adsorbed on the carbon (g). The X/M value at 1 mg/ℓ equilibrium gold concentration, determined from the isotherm plot, is known as the "K-value".

2.0 APPARATUS

2.1 An analytical balance, capable of measuring to 0,1 mg.

2.2 A sieve, with aperture size 0,43 mm and 0,15 mm.

2.3 A drying oven.

2.4 A desiccator (Note 7.2).

2.5 A linear shaker-incubator with timer, adjusted to 130 oscillation/min. The Labotec scientific linear shaker was found to be suitable.

2.6 Erlenmeyer flasks, of 250 mℓ capacity.

2.7 A mechanical pestle and mortar.

2.8 A rotary sample splitter.

2.9 An atomic absorption spectrometer.

3.0 REAGENTS

All reagents are of analytical grade or equivalent. Where water is mentioned, deionized water is understood.

3.1 Sodium hydroxide solution : 1 mol NaOH/ℓ

Transfer 40,0 g of sodium hydroxide to a 500 mℓ beaker. Add 400 mℓ of water and stir to dissolve. Allow to cool. Transfer to a 1 ℓ volumetric flask and dilute to volume.

3.2 Borate buffer solution : pH 10

Transfer 6,184 g of boric acid (H_3BO_3) to a 2 ℓ beaker. Add 7,456 g of KCl. Dissolve in 1500 mℓ water. Adjust the pH to a value of 10,0 with the 1 mol/ℓ NaOH solution. Transfer to a 2 ℓ volumetric flask and dilute with water.

3.3 Gold standard stock solution : 100 mg Au/ℓ

To a 1 ℓ volumetric flask, add 0,1463 g of K $Au(CN)_2$, previously dried to constant mass. Dissolve, and dilute to volume with water.

1 mℓ = 0,100 mg Au

3.4 Potassium auro cyanide solution : 100 mg Au/ℓ

To a 1 ℓ volumetric flask, add 0,1463 g of K $Au(CN)_2$. Dissolve in the buffer solution. Add 200 mg of sodium cyanide. Mix to dissolve and dilute to volume (Note 7.3).

3.5 Sodium cyanide : NaCN crystals

3.6 Hydrochloric acid : 50% (v/v) HCl

Measure 500 mℓ of concentrated HCl into 1 ℓ volumetric flask. Dilute to volume with water.

3.7 Hydrochloric acid solution : 3% (v/v) HCl

Measure 30 ℓ of concentrated HCl into a 1 ℓ volumetric flask and dilute to volume with water.

3.8 Lanthanum nitrate solution : 20 g/ℓ La

Dissolve 62,3 g of La $(NO_3)_3.6H_2O$ in water. Dilute to 1 ℓ.

1 mℓ = 20 mg La

4.0 PROCEDURE

4.1 Using the rotary sample splitter, select a mass of about 100 g of the activated carbon.

4.2 Grind the carbon, using the mechanical mortar and pestle.

4.3 Screen the sample through the 0,43 mm sieve, and collect on a 0,15 mm sieve.

4.4 Transfer the collected carbon to a 400 mℓ beaker, washing the adhering material from the sieve with water. Add 100 mℓ of the 3% hydrochloric acid solution. Boil for 10 minutes. decant, and wash the carbon three times with water.

4.5 Transfer the carbon to the oven and dry at 105°C to constant mass. Cool in a desiccator.

4.6 Accurately measure 0,1000, 0,2000, 0,3000, 0,4000 and 0,5000 g of the sample and transfer each increment to a 250 mℓ Erlenmeyer flask (Note 7.4).

4.7 To each flask, add 100 mℓ of the potassium aurocyanide solution (3.4). Stopper the flasks.

4.8 Place the Erlenmeyer flasks on the linear shaker - incubator. Adjust the temperature to 25°C and the speed to 130 oscillation/min. Commence shaking (Note 7.5).

4.9 Continue the agitation of the flasks for 20 hours.

4.10 Filter the contents of each flask through a Whatman No. 542 filter membrane, into a 100 mℓ glass bottle (Note 7.6). Discard the first 10 mℓ of the filtrate.

4.11 Determine the equilibrium concentration of gold in each solution by direct atomic absorption measurement (Note 7.7), with reference to the calibration standard solutions (5.0). Application of background correction and La as interference suppressing agent, is essential.

4.12 Determine the Au concentration of the potassium aurocyanide solution (head solution), with each batch of carbon samples (Note 7.8).

5.0 PREPARATION OF THE CALIBRATION STANDARD SOLUTIONS

5.1 To a series of 100 mℓ volumetric flasks, transfer 0,5, 1,0, 2,0, 5,0 and 10,0 mℓ of the gold standard stock solution (3.3).

5.2 Add 10 mℓ of the La $(NO_3)_3$ solution.

5.3 Add 40 mℓ of the buffer solution.

5.4 Add 40 mℓ of 50% HCl.

5.5 Dilute to volume with water.

5.6 The prepared standard range corresponds to 0,5, 1,0, 2,0, 5,0 and 10,0 mg Au/ℓ.

5.7 Fresh calibration standard solutions should be prepared each week.

6.0 CALCULATION

To determine the gold adsorption capacity of the activated carbon, K-value, use the empirical Freundlich isotherm $\log X/M = \log kC^{1/n}$.

where: X is the mass of gold adsorbed on the carbon (mg)

M is the mass of the carbon taken (g)

C is the equilibrium Au concentration in the filtrate (mg/ℓ)

and k and n are constants.

The X/M value at 1 mg/ℓ equilibrium concentration, determined from the isotherm plot, is the K-value.

6.1 Record the accurate mass of carbon (g), taken for the determination, in column 1; symbol M in the Freundlich isotherm (Table I).

6.2 Enter the gold concentration (mg/ℓ) of the filtrate solution, in column 2; symbol C in the Freundlich isotherm.

6.3 Calculate the actual mass of gold (mg) in the solution, by multiplying the concentration (mg/ℓ) by 0,1, since only 100 mℓ of the potassium aurocyanide solution was used for the test. Enter the quantity thus derived (mg) in column 3.

6.4 Calculate the mass of gold adsorbed, by subtracting the equilibrium quantity (mg) from the quantity of gold in the head solution (mg). Record the value in column 4; symbol X in the Freundlich isotherm.

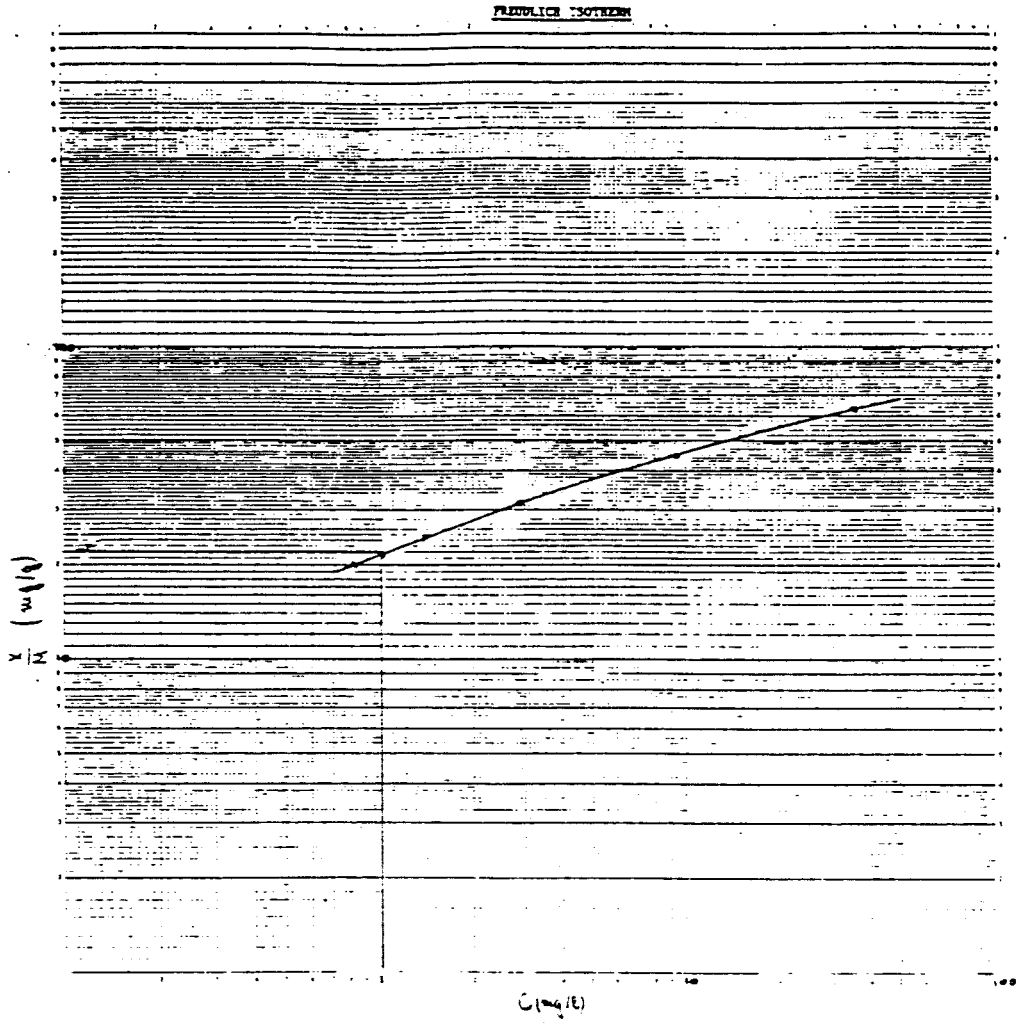
6.5 Calculate the relative quantity of gold adsorbed on the carbon (mg Au/gC), by dividing the figure in column 4 by the corresponding carbon mass in column 1. Record the value in column 5; symbol X/M in the Freundlich isotherm.

TABLE I

Carbon mass (g)	Au concentration in solution		Au adsorbed on carbon	
	Residual (mg/ℓ)	Actual (mg)	(mg)	(mg/g)
M	C		X	X/M
0	100,0	10,0	-	-
0,1028	35,4	3,54	6,46	62,84
0,2008	9,46	0,946	9,054	45,09
0,3016	2,88	0,288	9,712	32,20
0,4024	1,38	0,138	9,862	24,51
0,5018	0,825	0,083	9,917	19,75
Column 1	2	3	4	5

6.6 Plot on log/log graph paper the equilibrium Au concentration, C (mg/ℓ), along the x axis. Plot the relative quantity of adsorbed Au, X/M (mg/g), along the y axis.

6.7 Draw the best possible curve through all points.



6.8 To obtain the k-value, determine the mg Au/gC value, corresponding to 1,0 mg Au/ℓ equilibrium concentration: 22 kg Au/tC on the carbon used as example.

7.0 NOTES

7.1 Theoretically, equilibrium should be reached after 20 hours irrespective of particle size; but for better reproducibility, a uniformly reduced size fraction is taken for analysis.

7.2 Before starting the test, ensure that the desiccator and desiccant are in good condition.

7.3 Cyanide is highly toxic; it must be treated with caution. Always add cyanide salts to a solution of pH value greater than 10. Poisonous hydrogen cyanide gas can be released if the cyanide is added to water; 270 ppm in air is fatal. Before disposal, oxidize all solutions containing cyanide with $KMnO_4$. If cyanide solution is splashed on the skin, wash it off with water: 100 mg/ℓ CN^- can be absorbed dermally. If cyanide solution is swallowed, call a physician, and give warm salt water until the vomit fluid is clear. Never pipette cyanide solutions by mouth. Use an automatic sampler or pro-pipette for transfer of increments.

7.4 Dry carbon is very hygroscopic; allow minimum time for measurement of sample mass.

7.5 Ensure that during the test, the carbon does not collect in the neck of the Erlenmeyer flask; this can be a cause for negative bias.

7.6 Vacuum filtration is not essential, as the equilibrium is reached after 20 hours of contact, and the speed of filtering is reasonably fast.

7.7 When dilution is required, the buffer solution (3.2) should be used as diluent.

7.8 Head solution or blank is the 100 mℓ of potassium aurocyanide solution (3.4), without addition of activated carbon.

THE DETERMINATION OF GOLD ADSORPTION RATE (R-VALUE) OF ACTIVATED CARBON, MESH NUMBER 8 X 16 AND 6 X 12

1.0 PRINCIPLE

The activated carbon, after acid wash, is contacted with 10 mg/ℓ gold solution at 25°C, in the presence of potassium cyanide.

At periodic intervals, small aliquot portions of the solution are taken, and gold content is determined.

The rate of gold adsorption is obtained from a calibration graph, of per cent gold adsorbed on the carbon plotted against time.

The quantity of gold (%), adsorbed after 60 minutes, is known as "R-value".

2.0 APPARATUS

2.1 A set of sives.

2.1.1 For virgin activated carbon, of mesh number 6 x 12 : 2,0 and 2,38 mm.

2.1.2 For virgin activated carbon, of mesh number 8 x 16 : 1,41 and 1,68 mm.

2.2 An ultrasonic bath.

Brauson ultrasonic cleaner was found to be suitable.

2.3 An analytical balance, capable of measuring to 0,1 mg.

2.4 A drying oven.

2.5 A linear shaker-incubator with timer, adjusted to 130 oscillation/min.

The Labotec scientific linear shaker was found to be suitable.

2.6 An oblong plastic container: 95 x 130 x 210 mm or 150 x 150 x 140 mm.

The volume of the container must be 2-3 ℓ (Note 6.1).

2.7 A desiccator.

2.8 An AA spectrophotometer.

3.0 REAGENTS

All reagents are of analytical grade or equivalent. When water is mentioned, deionized water is understood.

3.1 Sodium hydroxide solution: 1 M NaOH

Dissolve 40,0 g of sodium hydroxide in 700 mℓ of water. Cool and dilute to 1 ℓ.

3.2 Borate buffer solution

Dissolve 3,09 g of H_3BO_3 and 3,73 of KCl in approximately 500 mℓ of water. Adjust the pH of the solution, using 1 M NaOH, to a pH value of 10. Dilute to 1 ℓ.

3.3 Gold solution: 10 mg/ℓ Au

Dissolve 0,074 g K Au $(CN)_2$ in 4 ℓ of water. Add 1,834 g of $CaCl_2.2H_2O$ or 1,384 g $CaCl_2$ (Note 6.2). To the solution add 1,251 g of KCN. Mix and dilute to 5 ℓ.

The concentration of the solution is: 10 mg/ℓ Au, 100 mg/ℓ Ca and 100 mg/ℓ CN^-.

3.4 Hydrochloric acid solution: 3% (v/v) HCl

Measure 30 mℓ of concentrated HCl into a 1 ℓ volumetric flask. Dilute to volume with water.

3.5 Lanthanum nitrate solution: 20 g/ℓ La

Dissolve 62,3 g of $La(NO_3)_3.6H_2O$ in water. Dilute to 1 ℓ.
1 mℓ = 20 mg La.

4.0 PROCEDURE

4.1 Using the rotary sample splitter, select mass of about 50 g of the activated carbon.

4.2 Screen the sample through the 1,68 mm (Note 6.3) and collect on 1,41 mm sieve (Note 6.4).

4.3 Transfer the carbon, collected on top of the 1,41 mm sieve, to a 100 mℓ beaker. Add 50 mℓ of the acid solution. Boil for 10 minutes. Decant and wash the carbon three times with water.

4.4 Cover the carbon with 50 mℓ of water. Position the beaker in the ultrasonic bath. After 2 minutes, decant the water and wash the carbon twice. Repeat the procedure four times (Note 6.5).

4.5 Transfer the carbon to the oven and dry at 105°C to constant mass. Cool in a desiccator (Note 6.6).

4.6 Measure, accurately, 1,000 g of the dried carbon sample (Note 6.7). Moisten the sample with 1 mℓ of the buffer solution, and allow to stand for ½ hour (Note 6.8).

4.7 Allow the flask, containing the gold solution, to stand for ½ hour in the incubator at 25°C, to adjust the temperature of the solution.

4.8 Measure 1 ℓ of the gold solution, using a measuring cylinder, into the oblong plastic container, and add the prepared carbon sample.

4.9 Position the container on the linear shaker-incubator, the temperature (25°C) and speed (130 oscillation/min) of which should have been adjusted previously. Commence shaking immediately.

4.10 Withdraw 5 mℓ aliquots after 0, 15, 30, 45, 60, 75 and 90 minutes (Note 6.9).

4.11 Proceed with the determination of gold, using drect atomic absorption measurement. Application of background correction and addition of $La(NO_3)_3$ solution, as interference suppresing agent, is essential (Note 6.10).

5.0 CALCULATION

5.1 The results from the atomic absorption measurement will give the residual gold in the solution (mg/ℓ Au) at the time of aliquot withdrawal (0, 15, 30, 45, 60, 70 and 90 min): Value B.

5.2 Subtract the residual gold concentration from the original gold content (10 mg/ℓ Au). Calculate, in percent, the quantity of gold adsorbed on the carbon at any given time:

$$\text{R Value (\%)} = \frac{(A - B)}{A} \times 100$$

where: A is the Au concentration (mg/ℓ) of the solution (10 mg/ℓ).

and B is the residual Au concentration (mg/ℓ) at the time of aliquot withdrawal

5.3 Plot a calibration curve of percentage gold adsorbed, against time, to obtain the mean percentage of gold adsorbed onto the carbon after 60 min. The statistical data for the test are: SD = 1,7% and RSD = 3,4%.

6.0 NOTES

6.1 The 2-3 ℓ volume of the plastic container was found to ensure sufficient sample agitation during the test.

6.2 The addition of $CaCl_2$ is to simulate the plant conditions; it also accelerates the gold adsorption rate.

6.3 A sieve of mesh size 2,38 mm should be used when 6 x 12 mesh activated carbon is tested.

6.4 A sieve of mesh size 2,00 mm should be used when 6 x 12 mesh carbon is tested.

6.5 Four washings were found to be sufficient for complete removal of dust and small particles.

6.6 Before starting the test, ensure that the desiccator and its drying material are in good condition.

6.7 Dry carbon is very hydroscopic; allow the minimum time for sample measurement.

6.8 The sample is wetted ½ hour prior to the test, to displace the air, entrapped in the carbon.

Buffer solution is used, so that, simultaneously with the air displacement, appropriate neutralisation of the acid washed carbon is ensured.

6.9 The original gold concentration is maintained, when small aliquot portions are withdrawn from the solution.

6.10 The final concentration of La in the solution should be 2 g/ℓ.

THE DETERMINATION OF SIZE DISTRIBUTION AND d_{50} NUMBER OF ACTIVATED CARBON MESH NUMBER 8 X 16 AND 6 X 12

1.0 PRINCIPLE

A representative sample of carbon is screened for 30 minutes through an appropriate set of sieves. The mass of each fraction is measured and expressed as a percentage of the original sample mass.

2.0 APPARATUS

2.1 A balance, accurate to 10 mg.

2.2 A Ro-Tap sieve shaker, with timer.

2.3 A series of sieves of aperture size as shown in Table I.

TABLE I

Carbon 8 x 16 Size of aperture (mm)	Carbon 6 x 12 Size of aperture (mm)
2,38	3,36
2,00	2,83
1,68	2,38
1,41	2,00
1,19	1,68
1,00	

2.4 A rotary sample splitter.

3.0 PROCEDURE

3.1 Assemble the sieves, in descending order of aperture size, with a receiving pan under the bottom sieve.

3.2 Using the rotary sample splitter, select a sample of about 100 g of the dry carbon. Measure a mass of 100 g to the nearest 10 mg.

3.3 Transfer the measured sample to the top sieve, and cover with a lid.

3.4 Position the assembly on the Ro-Tap shaker. Switch on, and allow the assembly to shake for 30 minutes.

3.5 Remove the sieve assembly and quantitatively collect each carbon fraction, using a brush.

Detach any carbon particles stuck in the screen apertures, and add to the fraction on the retaining screen.

3.6 Measure the mass of each carbon fraction.

4.0 CALCULATIONS

4.1 Calculate the particle size distribution as a percentage of the total mass:

$$\text{Particle size (\%)} = \frac{X}{M} \times 100$$

where: X is the mass (g) of each individual fraction
and M is the mass (g) of the total carbon sample.

4.2 Record the percent of fraction 2,38 mm as value A(%).

4.3 Add together the percent of the fractions 2,38 mm and 2,00 mm (Note 5.1). Record the value obtained as B(%).

4.4 To value B, add the percent of 1,68 mm fraction, to obtain value C(%).

4.5 Continue with the addition, until all fraction values are added together. Six values are recorded for activated carbon 8 x 16 mesh (Note 5.2).

TABLE II

Value Obtained (%)	A	B	C	D	E	F
Fractions added (mm)	2,38	2,38 2,00	2,38 2,00 1,68	2,38 2,00 1,68 1,41	2,38 2,00 1,68 1,41 1,19	2,38 2,00 1,68 1,41 1,19 1,00

4.6 Plot on graph paper the A, B, C, D, E and F values (%), along the y axis. Along the x axis, plot the corresponding aperture size (mm).

4.7 Draw the best possible curve through the points.

4.8 To obtain the d_{50} number, determine the screen size corresponding to 50% mass of activated carbon.

5.0 NOTES

5.1 For activated carbon of 6 x 12 mesh, the appropriate fractions are used (Table I).

5.2 Only five values; A, B, C, D and E are obtained for carbon of 6 x 12 mesh, as five screens are used for the determination of size distribution.

THE DETERMINATION OF SOLUBLES AND MOISTURE CONTENT IN ACTIVATED CARBON

1.0 PRINCIPLE

A representative sample of carbon is measured in the "as received" form, washed several times with water and dried. After cooling, the mass of the dried sample is measured. The solubles and moisture content is the loss in mass of the "as received" sample (Note 5.1).

2.0 APPARATUS

2.1 A balance, accurate to 0,1 mg.

2.2 A drying oven.

2.3 Sintered glass crucibles (Note 5.2) of porosity G4 (Note 5.3).

2.4 A vacuum filtration assembly.

2.5 A desiccator (Note 5.4).

3.0 PROCEDURE

Distilled or deionised water is used throughout the procedure.

3.1 Tare a clean, dry sintered glass crucible. Record the mass.

3.2 Measure 10 g of the carbon sample into the crucible.

3.3 Place the crucible in a 100 mℓ beaker. Cover with water and allow to stand for 10 minutes.

3.4 Filter off the water under vacuum.

3.5 Repeat steps 3.3 and 3.4 until the sample has been washed 6 times.

3.6 Transfer the crucible, with sample, to a ventilated oven at 105°C, and dry to constant mass.

3.7 Cool in a desiccator. Measure and record the mass.

3.8 Express the loss in mass as a percentage of the "as received" mass of the sample.

4.0 CALCULATION

$$\text{Solubles and moisture (\%)} = \frac{(A-B) - (C-B)}{A-B} \times 100$$

where A is the mass (g) of crucible and sample "as received"
B is the mass (g) of empty crucible
C is the mass (g) of crucible and sample after drying.

5.0 NOTES

5.1 When the sample is received in an open package, it is important to determine the solubles and moisture content as soon as possible, to minimise the absorption of atmospheric moisture.

5.2 A sintered glass crucible is superior to a petri dish: it entraps the carbon fines in the pores of the sinter. Thus it eliminates the loss of fine material and ensures more accurate results.

5.3 A crucible of porosity No. 4 was found to be adequate for retention of the fine dust.

5.4 Before starting the test, ensure that the desiccator and desiccant are in good condition.

PREPARED FOR CARBON MONOGRAPH (IN PRESS)

CHAPTER XX

CARBON MANAGEMENT

R.J. Davidson and
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20.1 Introduction

20.2 Daggafontein Process Description

20.3 Quality Control of Bulk Deliveries

20.4 Carbon Preparation and Make-up

20.5 Carbon Transport
Transport Water Circuit

20.6 Adsorption Circuit

20.7 Elution Circuit

20.8 Regeneration Circuit

20.9 Carbon Consumption

20.10 Carbon Selection

20.11 Conclusions

20.1 INTRODUCTION 2

With the rapid development of carbon technology in South Africa over the past decade culminating in some 30 larger and 30 smaller carbon based operations, a system of carbon management akin to trial and error has taken place in the industry. Many of the first generation circuits were and still are beset with major engineering problems making effective carbon management difficult. Due to the integral relationship existing in the many unit operations in such circuits, effective carbon management relies on the optimum performance of all of these unit operations. Carbon management is not only concerned with the physical aspects of carbon handling and housekeeping but also the metallurgical performance of each unit operation.

While it is not the purpose in the current review to accentuate poor plant practise, the intention is to rather generate a better understanding of the attributes related to effective carbon management. Because of the very different plant designs and scale of operation, as well as differences in types of material currently being treated in South Africa, any general overview of current carbon management is difficult to summarise. The intention is therefore to illustrate by way of example the management of carbon in a high tonnage dedicated well run operation of recent design. The Daggafontein CIP plant managed by the East Rand Gold and Uranium Company (ERGO) was selected for this purpose.

20.2 DAGGAFONTEIN PROCESS DESCRIPTION

20.2.1 Background

Two important technological advances made over the last 25 years were key factors in the conception of the Daggafontein project.

Firstly, the concept of large scale dump retreatment and the technology needed to transport over large distances, treat and dispose of enormous tonnages of slimes was pioneered and established by the ERGO operation. This has enabled large, low grade, slimes dam retreatment operations to be a viable proposition. The second key factor has been the development of the CIP/CIL gold recovery process which was vital to the Daggafontein project since the very low gold grades and low sulphide sulphur content of the feedstock makes treatment of the dams uneconomical by the conventional ERGO route.

These new technologies and the increase in gold price experienced in the late 1970's/early 1980's led to the formation of a joint venture between ERGO and East Daggafontein Mines to provide reserves, fund, design and construct a 1 000 000 t/m retreatment plant. Some 286 million tons will be treated by the Daggafontein plant which was constructed at a cost of approximately R150 million. The management of the project was conducted by Anglo American Corporation utilising ERGO metallurgical and engineering staff to provide specialist technical know-how and experience. The plant operates as a division of ERGO, falling under the direct control of ERGO management.

20.2.2 Process Review

Each of the two parallel streams referred to as North and South has a nominal capacity of 500 000 t/m (Figure XX.1). Each slimes treatment stream consists of a pre-leach thickener, a desanding and CIP/CIL stream, a fine carbon recovery section and a residue disposal facility, which is common to both streams.

The carbon treatment consists of a hot acid wash followed by an AARL type elution. Three identical parallel streams have been provided, giving a full standby facility. Thermal regeneration of the eluted carbon is accomplished by means of a rotary kiln.

Gold recovery from eluate is by zinc precipitation. The zinc-gold slime is acid washed, filtered and then despatched to ERGO Division for calcining and smelting.

Slimes Reclamation

During the projected life of the Daggafontein Division Plant, slimes dams from distances of up to 15 km away will be hydraulically reclaimed and pumped into the central plant for treatment.

The monitoring technique to be employed is similar to that developed at ERGO. High pressure water (2 MPa) is pumped from the central plant to the two reclamation sites where remotely operated monitor guns are utilised to slurry the dam material. The reclaimed pulp gravitates through a system of earth launders to a pumpstation located at the

lowest point on the perimeter of the dam. The elevation of the inlet to the pumpstation is calculated such that all material will gravitate to a single such station.

At the pumpstation the slurry passes over two Vibramech 6,4m x 2,4m vibrating screens for the removal of oversize tramp material (+2mm) for pump and pipeline protection. Envirotech rubber lined 'D' frame pumps have been selected for the transfer of material to the plant.

Pre-Leach Thickening

The material reclaimed from the dam sites tends to have fluctuating density with pulp ranging from 1,35 to 1,55 under normal conditions. In order that the CIL section of the plant is presented with material of a steady density and flow rate, two 137m peripheral drive thickeners are installed.

Slurry from the dams is received, via a distribution box, into a mixing tank where dilution water is added to produce a thickener feed pulp R.D. of 1,28. Lime is added to the mixing tank for primary pH modification.

Thickener overflow is recycled as dilution water for the feed slurry with excess supplementing the monitor water supply. 'D' frame pumps located in a ventilated chamber beneath the thickeners are used to transfer thickened material to the desanding section. A.C. variable speed underflow drives can be controlled on either volumetric flow measurement, or density measurement. The facility to bypass the thickeners and feed reclaimed slurry directly into the desanding area is incorporated.

Feed Preparation (desanding)

AAC/Delkor linear screens were selected to remove tramp material from the feed to CIP and four 20m^2 units are installed (two per stream).

Oversize material from the linear screens is removed by water wash and gravitates via launders to a sump from where it is pumped to dewatering screens. Screen underflow returns to supplement the launder flow with excess being used as pre-wash on the linear screens. The screen oversize is collected in a storage bin, and trucked to a stockpile.

Screened pulp is transferred to two "desanding" tanks (i.e. the first two units of each stream in the main tank farm). The tanks, 12m dia x 12m high are agitated using Prochem agitators each absorbing 62 kW. The facility for the measured and controlled addition of air, lime and calcium cyanide is provided in both tanks.

The second stage tank incorporates a screened outlet and the facility to purge any grit accumulations back to the linear screens. NKM mechanically swept screens with a cut size of 0,50mm are installed.

Incorporated into the design of this section is a calibration facility for all main slurry mass flow installations.

Adsorption

Each adsorption train comprises six 12m dia x 12m high tanks each fitted with Prochem agitators absorbing 32kW. Each adsorption tank is fitted with two 1,6m dia x 2,0m high NKM screens. Sala recessed impeller vertical spindle pumps are used for carbon transfer; the piping arrangement being such that low pumping heads are maintained keeping pump impeller speeds to a minimum.

The carbon concentration in the tanks is 18g/litre. Plant design, however, has been considered for carbon concentrations of 15g/litre up to 30g/litre.

Fine Carbon Recovery

The fine carbon screening and carbon scavenging from residue pulp is accomplished using 20m^2 linear screens in a similar manner to the desanding process. The washed oversize material is pumped to a three product screen for the separation of water, fine carbon and re-usable carbon. The re-usable carbon is fed into the seventh adsorption tank of the north stream while carbon fines are collected for further treatment.

A screening installation for adsorption area spillage is incorporated in this section of the plant, recovering carbon before returning spillage to the circuit.

Residue Disposal

The residue from the Daggafontein CIL plant is pumped to a tailings dam situated some 7 km away using up to eight Envirotech 'D' frame 300 x 300 pumps. The 500ha dam is constructed using cyclones in fashion similar to that employed at ERGO.

Acid Washing/Elution

The loaded carbon is washed and elutriated for the removal of slimes and woodchips before passing to a measuring hopper. A measured volume of carbon (12,5 t) is transferred to the acid washing column where it is treated with a 3% HCL solution at 60°C. The acid wash columns are constructed of carbon steel lined with Volco Durit D3. Candle screens and fittings within the column are fabricated from Hastelloy C276.

After neutralisation the carbon is transferred to the elution column where gold is eluted using the AARL type elution. Elution using soft water at temperatures of up to 140°C is carried out. The elution columns and fittings are fabricated from 316L stainless steel.

Heat to the process is derived from the combustion of Sasol gas. Three Wanson Thermopac 2000B units are used to heat a thermic oil ring main which feeds the shell and tube heat exchangers. The eluted carbon is transported under water, by pressurising the elution columns, to storage bins, prior to regeneration.

Three identical parallel streams have been provided giving a full standby facility.

Regeneration

Regeneration of the carbon is accomplished using equipment which has been standardised with the ERGO CIL plant and supplied by Wellman Engineering Africa.

Rotary vane feeders and dewatering screws transfer carbon from the storage bins into rotary predriers. The carbon is dried to 5 - 10 per cent moisture using burner off-gases from the gas fired regeneration kilns. Supplementary heat is provided by hot gas generators. The regeneration kilns provide a carbon temperature of ca 700°C for 20 minutes before cooling and quenching. The retorts of the regeneration kilns are fabricated from 321 stainless steel and are reinforced on the areas of maximum bending.

Incinerators have been provided in order to destroy noxious fumes produced during regeneration. (The Daggafontein plant is located adjacent to a residential area).

Gold Recovery

Eluate is received in an agitated storage tank prior to sampling and transfer to a precipitation feed tank. A zinc-dust and filter aid mixture is added as a slurry to three precipitation reactors. The precipitated slurry is dewatered using filter presses before acid treatment and refiltering.

Monitoring and Control

A network of eight PLC's control the process and transmit data to the CYGNUS supervisory system. The system not only allows for operator interface with the process, but provides an extensive data logging and manipulation facility.

20.3 BULK CARBON DELIVERIES

Perhaps the most important aspect of effective carbon management relates to the selection and quality of carbon entering the plant. While the most important aspect, this area of carbon management is the most neglected and least understood in the entire industry. In many operations, little or no effective quality control on bulk deliveries are carried out on a routine basis, primarily because of the costs involved and the inability at the present time of relating such control data to plant operation. This

unhappy circumstance does however place too much reliance in the hands of the carbon supplier rather than the users, and is no long term solution to upgrading the carbon quality used in the industry.

While the above situation may persist in many carbon based operations, the quality of carbons currently being marketed in South Africa continues to improve. Undoubtedly a better understanding of the qualities required by the industry is being developed. Specification limits with regard to the size, shape and hardness of suitable coconut based products are being formulated. Unfortunately the selection of the carbon activity required remains a debatable issue and requires further study. A recent pilot plant trial conducted by Mintek in which carbons of different activity were compared did however illustrate that carbon activity is a most important consideration. This is not surprising.

20.3.1 Plant Practise

Carbon as delivered to Daggafontein is stored in separate demarcated areas to facilitate separation of delivery batches. One bag of carbon is removed from each delivered batch and introduced to the fresh carbon make-up system. A representative sample of the entire bag is obtained via a sample cutter which is an integral part of the make-up system. The sample thus obtained is further riffled into four equal representative samples.

At the Daggafontein laboratories the sample is subjected to a size analysis in accordance with the Anglo American Research Laboratory specifications. This is then used to check the d_{50} and the percentage fines less than 1,19 mm. The sample is also subjected to activity testing. Any carbon not within specification, is then identified according to the source delivery batch. Suppliers are then asked to check results obtained using one of the remaining 3 samples. An independent laboratory is approached to resolve any discrepancies.

The quality control programme will later be expanded to include an accurate delivery mass and moisture content. Shape factor and hardness are only checked on stored batches of carbon if excessive carbon consumption is noted on the plant.

The above plant practise at Daggafontein is certainly to be recommended, although not enough emphasis is placed on the shape and hardness of bulk deliveries. While the activity of circuit carbon may be controlled in the regeneration circuit, the physical shape and hardness of virgin carbons can only deteriorate with use, and as such requires more attention with regard to both selection (20.10) and the sampling of bulk deliveries - great variations of both hardness and shape have been experienced in the industry due primarily to poor quality control on bulk deliveries.

20.4 CARBON PREPARATION AND MAKE-UP

Extended attrition studies carried out at AARL on numerous tender carbon samples (McArthur, 1987a) indicated an initial rate of carbon attrition some 5-8 times greater than the rate measured following extended attrition times (5-10d). This initial rate of carbon attrition was invariably accompanied by a reduction in gold adsorption activity of the residual carbon due to the loss of a small, low density soft but highly active component originally present in the commercial product.

The soft highly active carbon referred to would soon be lost in any contacting circuit incurring not only an initially high attritional loss, but also a considerable gold loss, particularly in high grade circuits as much of this material would have a rapid co-current movement in the adsorption circuit and be lost to the tailings residue. These findings confirm the view that any make-up carbon entering the adsorption circuit should first be mechanically conditioned. Besides improving metallurgical efficiency, the removal of soft flaky material would also improve the performance and maintenance of interstage screens.

The presence of this soft highly active carbon fraction in bulk deliveries should also be considered in any selection campaign (20.10). Following light attrition scrubbing incurring a mass loss of only 1-2%, McArthur showed that some tender samples suffered a 33% decrease in original kinetic activity. Loading capacities were likewise shown to decrease by as much as 20%.

20.4.1 Plant Practise

Carbon stockpiles at Daggafontein are operated on a first in, first out basis. The fresh carbon storage vessel is maintained at a constant level to hold approximately 40t of virgin carbon. (This is equivalent to two adsorption tanks inventory).

Fresh carbon make-up is controlled by the monthly carbon usage on the plant. The carbon requirements are calculated back to a daily average, and this predetermined amount is blowcased to the adsorption section every day. Currently, it works out to be approximately one bag of carbon per stream per day.

Virgin carbon is loaded into a pretreatment vessel where it is subjected to mechanical agitation in water at an approximate 5:1 volumetric ratio (water:carbon) for 5 minutes. This process removes friable particles and wets the carbon.

The carbon is then pumped, with the addition of dilution water, over a vibrating screen where the oversize product passes to the fresh carbon storage tank. Here, fresh carbon is stored under water before addition to the adsorption circuit. The screen undersize (-1mm) gravitates to the waste water sump before being pumped to the dirty water section for further clarification (20.5.2).

Attritional losses in the above carbon preparation circuit at Daggafontein are estimated at 3-5% of the total carbon loss (20.9). This estimate is significantly greater than the attritional losses measured on carbon tender samples at AARL where mass losses of 2-3% following a 5 day test procedure are commonly reported. Such discrepancies further emphasise the necessity of sampling bulk deliveries (20.3).

20.5 CARBON TRANSPORT

This aspect of carbon management relates particularly to the selection of carbon handling equipment commonly used in CIP circuits in order to minimise both carbon losses and to prevent equipment corrosion. Granular activated carbon is an extremely abrasive material as well as being highly corrosive towards mild steel due to its' anodic properties.

The handling characteristics of transporting carbon slurries in pipelines must also take into account such characteristics as linear flow velocities, water/carbon ratios and the choice of wide pipe radius bends. The emphasis in this review will be confined to actual transport equipment.

Blowcase

The blowcase uses compressed air or pressurised water as motive media applied in a specially designed pressure vessel to move the carbon slurry (Figure XX.2a). Carbon and water are slurried in the feed tank located above the pressure vessel. The slurry falls into the pressure vessel and the feeding valve is closed. Air or water pressure is applied and moves the slurry out through the discharge pipe. At the end of the transfer, the valve in the air/water line is closed and the vessel is vented/drained before starting a new cycle. The cycle is completely automated and controlled by timers.

At the beginning of the transfer, the concentration of the slurry is very high and could be close to 50% moisture (wet basis); thereafter, the excess water should dilute the slurry and, in a properly designed system, flush the pipelines at the end of the transfer. The required motive pressure for the air/water in most cases is minimum 1 bar above the static head from the blow-case to the discharge point.

This method of transport has found considerable success in carbon plants because of:

- no moving parts
- low maintenance (only on instruments)
- not limited by the discharge head
- less subject to erosion or corrosion problems

Eductor

Water jet educators for granular materials are similar to the ones installed in steam vacuum units (Figure XX.2b). A suction effect is created by pushing water at high velocity in a Venturi installed in the body of the ejector. The discharge of carbon from the feed hopper is controlled by the feed valve and the flow of water through the Venturi.

Water eductors have been used successfully in situations where motive water can be used in large quantity and is ready to be recycled; they are easy to operate, require minimum maintenance, easy to replace and of a low initial cost.

Their disadvantages can be listed as follows:

- subject to corrosion and erosion- the life of an eductor in continuous service has varied between one month to one year depending on the duty

- require large volumes of motive water at elevated pressure (as a function of the discharge head)

- efficiency drops off rapidly if the discharge head is located at more than 10m of water column

- use very diluted carbon slurry (below 0,10) and therefore, have a low delivery rate

- easily plugged by gravel, trash or any extraneous foreign material

Centrifugal Pump

Many types of pumps have been tried for the transport of carbon in slurry. The most successful pump has been the recessed impeller pump because it maintains a maximum clearance for the passage of carbon granules (Figure XX.2c). The speed of the pump should not exceed 1000 rpm to avoid excessive wear of the pump and to minimise degradation of the granules. A rubber lined impeller is recommended to improve the pump resistance to abrasion.

The major disadvantages of this type of equipment are:

- very low efficiency - discharge head limited to 15m of water column

- corrosion and erosion problems (impeller and casing)

- dilution water to the pump must be controlled very carefully to avoid any feed overload

- density of the carbon slurry is very low due to the excess of water dilution in the pump needed for correct operation. Therefore, the carbon delivery rate is as low as with a water ejector.

The operating characteristics of the recessed impeller pump vary considerably from one manufacturer to another and, therefore, it is wise to ask for confirmation of design criteria before making any type of decision. In continuous service operation, e.g. under the quench tank, this equipment has given full satisfaction in many carbon reactivation plants; the life of the pump has been reported to be one - two years.

Continuous Dewatering Screw Feeder

Continuous dewatering screws used for feeding carbon to regeneration kilns have proven to be most satisfactory for reducing the water content to at least 50% on a wet basis (Figure XX.2d). The inclined screw conveyor which is fitted with a box-like back end, is fairly insensitive to fluctuation in the feed storage bin, up to the limits of its design capacity.

This equipment is relatively inexpensive, very reliable to provide continuous dewatering operation at low maintenance costs. The feed rate to the reactivation furnace is not controlled by the screw itself but by a timer operated valve; therefore, it provides a very accurate control. In addition to providing the feed - dewatering function, this equipment is also designed to serve as a water sealing device and it eliminates the need for a rotary air lock valve.

20.5.1 Equipment Selection

Based on pilot plant testwork at ERGO and Chemviron Engineering assistance, the following carbon transportation systems were adopted for Daggafontein:

a) Interstage Pumping - Sala STGVA vertical spindle pumps fitted with all metal impellers. For interstage pumping with up to a 15m static head and 10 - 60 g/ℓ carbon in slurry, this pump gave the best performance. (Based on lowest carbon abrasion).

b) Carbon in water pumping - the Morris 3HS10 recessed impeller pump gave the best performance. The unit must be operated with a 7 - 10:1 water to carbon ratio to minimise carbon attrition. (Quench tanks, fresh carbon make-up).

c) Blowcases - $1m^3$ blowcases ($1m^3$ carbon capacity) were utilised for heads above 15m and to minimise dilution of slurry circuits. (Fresh and regenerated carbon).

d) Eductors were tried at Ergo CIL plant for discharging quench tank carbon. They were found to block easily, causing frequent downtime for removal and cleaning operations. These downtime periods affected kiln throughput, and because of this, the eductors were removed from service. Because of this, eductors were not considered for Daggafontein.

20.5.2 Transport Water Circuit

Due to the often involved water circuits in many major gold metallurgical complexes, the containment and reuse of carbon transport water is most essential. Beside significant savings in water, such transport waters usually contain both granular and carbon fines which may seriously affect other gold recovery circuits e.g. conventional filter plants and milling circuits. Such carbon may also contain significant gold values arising from plant spillages or from circuits transporting loaded carbon.

During the initial design phase for Daggafontein, the one area that was doubtful was the so called dirty water purification plant. After visiting several plants, it became apparent that this was a "grey" area concerning design, and no one could boast about a very good system. The dirty water plant design at Daggafontein was based on the best available information, but did shortly after commissioning, undergo several changes. These included re-routing of piping and the replacement of a sieve bend with a Velmet rotary drum screen.

Acid wash and elution overflow tanks, regeneration spillage, eluted carbon storage overflow and fresh and regenerated carbon make-up spillage is all routed to the fresh carbon pretreatment tank. This allows all carbon bearing water sources to be screened over the fresh carbon pretreatment screen (1mm slotted aperture), before being routed to the waste water sump. The +1mm usable carbon is then blowcased back to adsorption. Regenerated carbon screen underflows are also routed to the waste water sump. All waste water sump products are pumped over the rotary drum screen (0,3mm aperture) where intermediate carbon is separated from the water. This carbon product is stockpiled, and the water in the dirty water storage tank is used for blowcasing, column transport, and for maintaining levels in carbon storage vessels. (All carbon is stored under water). A side stream continuous operation removes some of the carbon fines by means of a plate and frame pressure filter. The quality of the dirty water is difficult to ascertain, but no detrimental effects have been found on the plant.

20.6 ADSORPTION CIRCUIT

Both the carbon concentration and the rate and mode of carbon transfer in the adsorption circuit play significant roles in metallurgical efficiency and bullion lock-up. Simulation studies by Nicol et al., 1984 and Stange et al., 1990 indicated the use of an even distribution of carbon in circuits where no leaching takes place, while Stange et al., 1990 suggested the use of higher carbon concentrations in the latter contacting stages in circuits where leaching is still in progress i.e. CIL type circuits. In the case of carbon transfer, this latter study also indicated using a sequential mode of carbon transfer where carbon is first transferred to elution and then proceeds in logical sequence through the contacting circuit before finally adding regenerated carbon to the last contacting stage (the study also indicated the improved performance obtained in a carousel type mode of operation i.e. 100% carbon transfer).

Carbon management in this circuit is essentially related to the control and movement of the carbon inventory in the contacting circuit in order to effect the most cost effective metallurgical efficiency. As the overall profitability of a CIP plant is affected by a vast number of variables many of which are constrained by plant design and throughput, the emphasis in this review will relate more particularly to the physical movement and control of the carbon inventory.

Monitoring of the carbon inventory distribution in the adsorption circuit is currently carried out by the taking of dip samples through the contacting stages. This is generally performed manually once or twice per shift, but as infrequently as once daily in some circuits. Sampling usually involves the taking of a 1-2ℓ grab sample 1-2m below the pulp surface before volumetrically measuring the washed carbon.

Whilst this method of monitoring carbon distribution may be adequate in some circuits where carbon distribution is "reasonably" uniform due to efficient mixing and mass transfer, unreliable data is often recorded in circuits where poor mixing is evident - the taking of a 1ℓ grab sample from a 2×10^6ℓ contacting stage at ERGO is a good example. In such circuits the taking of larger and/or more frequent samples may be required. The development of a reliable "plant-proof" on-line monitor would appear to be justified in many circuits (Mintek are expected to market an on-line ultrasonic carbon concentration meter in early 1990).

20.6.1 Feed Preparation

One of the most crucial factors in CIL/CIP adsorption systems, is to ensure that no near size material becomes entrained in the carbon circuit. Coarse material (consisting typically of wood fibre, calcium sulphate, coal/ash/clinker and quartz grits), unless removed prior to adsorption, will accumulate in the circuit and have a number of detrimental effects. The following are a few examples-

1. Interstage screen throughput reduces and the wear on screen surfaces increases.

2. Elution costs per yield increases as the carbon treated becomes steadily diluted with near size tramp material.

3. Efficiencies are adversely affected in both elution and adsorption circuits.

4. Regeneration kiln descaling operations have to be performed more frequently.

At Daggafontein, feed to the desanding section arises from the preleach thickeners underflow, or in the event of bypassing of the thickener, directly from the reclamation pumpstations.

The material is discharged into a distribution box/launder system before gravitating to four 20m² AAC/Delkor linear screens. The oversize (+0,5mm) is removed via a water wash. Underflow passes to a main sump before being pumped to the first of two, desanding tanks.

The desanding tanks have two main functions. The first tank is utilised to pH condition the pulp in preparation for leaching (pH 10-10,5), and to provide pre-leach oxidation by sparging air into the pulp. The second desanding tank also utilises air sparging, but the main purpose is to act as a screening vessel with the ability to purge back over the linear screens. Thus any material not removed by linear screen, and trapped in the second desanding vessel by NKM interstage screens, may be purged back over the linear screens. The desanding tanks' NKM screen aperture is 0,5mm (adsorption is kept between 0,65 - 0,85mm).

Spillage from the desanding tank overflow is kept separate from adsorption spillage, and is returned to the front of the desanding circuit (i.e. desanding linear screens).

20.6.2 Interstage Screening

The interstage screen operation is the most critical of all unit processes in a CIP/CIL plant. Inefficient screening results in the co-current movement of carbon and even small carbon leakages can cause large increases in soluble loss and gold lock-up.

The NKM interstage screens designed and used by the Daggafontein operation are undoubtedly a major breakthrough in high tonnage screening. They satisfy the four main criteria in screen selection namely:

- Reliable and effective performance
- High unit capacities
- Robust construction
- Cost effectiveness

A diagrammatic representation of the screen is shown in Figure XX.3. The screen consists of three main components.

- The mounting frame, pulp chamber and outlet volute.
- The central drive unit and agitator arrangement.
- The replaceable wedge wire screen section.

Regular checking of interstage screens for blinding and/or excessive aperture wear is an important control measure if steep carbon profiles are to be maintained.

Interstage screens at Daggafontein are fitted with type 430 stainless steel wedge wire screens. The desanding tank screens have a nominal aperture of 0,5mm, which wears to 0,65mm within approximately 2½ months. The adsorption screens have an aperture of 0,65mm which is allowed to wear to 0,85mm (ca 8 months) before being removed from service. (Desanding screens are moved to the adsorption circuit on reaching an aperture of 0,65mm).

Monitoring exercises established that each interstage screen must be removed for cleaning approximately every 10 days. Unit cleaning time has reduced from 4 hours to 15 minutes, simply by replacing manual wire brushing with high pressure water cleaning. After a screen has been cleaned, the aperture is checked using a magnified optical measuring device. Sealing gaskets are replaced on the screen unit each time that the unit is cleaned. This reduces the possibility of carbon escaping through the interstage screen seal on the transfer launder to the downstream adsorption tank.

Apart from regular screen maintenance, production personnel check for interstage screen leakage twice per shift. This is done using a test sieve of 0,8mm aperture.

Interstage screen leakage detectors developed by Mintek were used originally, but the unreliability and high instrumentation maintenance costs shifted the system back to manual checking.

20.6.3 Carbon Inventory Control and Sampling

All interstage carbon transfer pumps at Daggafontein have the facility to be started independently from the control room. Each pump can be allocated a variable run time, which through experience allows for good control of carbon movement and maintenance of carbon profiles.

The last carbon stage vessel (before transfer to acid wash) is normally kept at a slightly higher concentration than the rest of the adsorption tanks (22 g/ℓ as opposed to 18 g/ℓ). Carbon is first moved into the acid wash/elution circuit from this vessel before the rest of the tankfarm profiles are corrected. This circumvents dilution of loaded carbon into the elution circuit, from the second adsorption vessel. Regenerated carbon is blowcased continually into the last adsorption tank as and when this carbon arises. Fresh carbon make-up is performed on dayshift only.

Accurate carbon sampling is dependant on the homogeneous dispersion of carbon in the slurry, which is totally dependant on the mixing system used. Daggafontein adsorption vessels are fitted with Prochem single impeller agitators (20.2). The last adsorption tank is fitted with sampling points down the side of the tank. Samples were initially taken from these points, giving an indication of the carbon concentration throughout the height of the tank. Long sampling devices were also made up and the top of the tank sampled at different places, ranging in depth from 1-2m. All of the samples taken were very similar in terms of carbon concentration. It was however ascertained that the most representative samples must be taken 1m below the slurry surface. To further improve the sampling of the carbon concentration, stainless steel sampling devices were made up for each adsorption vessel, and these are permanently chained in place. Samples are taken every 4 hours on a routine basis, and these samples kept and composited over a 24 hour period for analysis.

Average gold loading profiles for the period April-September, 1990 from the North and South tankfarms at Daggafontein are presented in Table XX.1 where the relatively steep and consistent profiles in both circuits may be noted. The low gold loadings in the final contacting stages (41 g/t) producing soluble losses of ca 0,005 g/t gold are also highly commendable. These data are an excellent example of effective carbon inventory control and efficient carbon elution (20.7) in a low grade dump treatment operation.

20.7 ELUTION CIRCUIT

Due to equilibrium rather than kinetic restraints, soluble losses in the adsorption circuit are largely dependent on efficient gold elution. Since carbon loadings of less than 100 g/t gold are required to contain a gold soluble loss of 0,01 g/t in the final stage (Stange et. al, 1990), eluted carbon containing less than 50 g/t gold would be required as considerable back-mixing takes place in the contacting circuit. While this statement may not apply to all CIP circuits, it does emphasise the important role played by the elution circuit. A survey of 19 elution circuits (Chapter X111) indicated that in many of the circuits reviewed, eluted carbon containing in excess of 50 g/t gold was being returned to respective adsorption circuits. While it is perhaps difficult to relate the soluble losses incurred in these circuits to any single process variable, inefficient gold elution is likely to be a major cause. Furthermore, as carbon movement rates are largely defined by the soluble gold losses incurred in the adsorption circuit, inefficient gold elution will increase carbon movement rates and so lower gold loadings thus increasing elution and regeneration costs as well as increasing carbon losses (20,9).

20.7.1 Plant Practise

Because of the batch nature of the elution process and the importance of attaining efficient gold elution (see above), the representative sampling of both loaded and eluted carbons is a prerequisite in the effective metallurgical control of the elution circuit.

The transfer of loaded carbon from the tankfarm to the acid wash building at Daggafontein takes approximately 4-5 hours. Manual sampling is performed by taking an hourly cut sample from the entire width of the loaded carbon screen. Acid washed and eluted carbon samples are taken manually during carbon transfer from acid wash column to elution column and elution column to eluted carbon storage respectivley. These samples are also composite samples, taken every 10 minutes during a transfer (a transfer takes approximately 60 minutes).

Initially, all carbon sampling was automated, with the loaded carbon sample being taken via a sample cutter, and the acid washed and eluted carbon samples via high pressure poppet samplers, installed in the respective transfer lines. All the sampling equipment had a common failure, which required high maintenance to prevent sample leakage. Maintenance costs, operational problems, and continual clean up of carbon spillage resulted in a change to manual sampling.

Manual carbon sampling during a transfer operation only commences 5 minutes after the start of the respective transfer. Testwork has established that the small quantity of carbon, lying below the internal column candle distributors, has a high gold value. Throughout the rest of the column, acid washed and eluted grades are fairly consistent. The 5 minute delay before commencing carbon sampling ensures that this material does not adversely bias the respective assay values.

Each batch of carbon is sampled and analysed for gold before and after acid washing, elution and regeneration. The regenerated carbon is also tested for activity using the ACIX method developed by National Chemical Products. Composite samples are collected over a week from all the points and tested for activity (20.8).

Monthly composited carbon samples of loaded, acid washed, eluted and regenerated are tested for Au, Ca, Cu, Ni, Fe, Si and Zn. Typical loadings are shown in Table XX.2 where it may be seen that acid washing removes 60-70% of the calcium and that only minor amounts of base metals are removed by acid washing; but that 80-90% of the copper and nickel are removed during elution.

Eluted carbon grades at Daggafontein are usually below 20 g/t (year to date average 16 g/t), with a corresponding elution efficiency of above 95%.

20.8 REGENERATION CIRCUIT

As the quantity of fresh make-up carbon entering the adsorption circuit seldom exceeds 5% of the total carbon feed, it is easily seen that both the activity and hardness of the circuit carbon is largely defined by the conditions used for carbon regeneration. These conditions in turn will influence the metallurgical efficiency effected in the adsorption circuit. The basic function of carbon management in this regard is to define such conditions and to maintain the quality of this circuit carbon.

Following a survey of 19 large CIP operations (Table XX.3), it may be seen that carbon regeneration temperatures in the range 600-750°C are in common use. When reviewing the methods used for carbon quality control in these circuits, various methods of kinetic activity testing were invariably carried out either by metallurgical staff, mine assay laboratories or by carbon suppliers. Only in a few circuits were such parameters as volatile content, loading capacities, iodine numbers or carbon hardness monitored as a means of maintaining carbon quality.

Carbon quality is perhaps the single most important aspect in any carbon based operation. As stated above, while kinetic activity testing is generally carried out in most circuits, little attention is focussed on monitoring carbon hardness primarily because of associated costs and the difficulty in relating such data to plant performance. While this common practise of monitoring only carbon activity may be a practical means of effecting quality control using low regeneration temperatures (600-700°C), the use of higher regenerating temperatures and/or longer residence times and the effect on carbon hardness should be considered.

The effect of acid washing as a means of maintaining carbon hardness should also be emphasised. As calcium, magnesium and iron are well known catalytic oxidants for the water gas reaction, acid washing prior to thermal regeneration will allow the use of higher regeneration temperatures and so effect more efficient carbon regeneration while not sacrificing physical hardness. Thermogravimetric evidence supporting this conclusion is well illustrated in Figure XX.4.

The beneficial effects of acid washing before rather than after regeneration on maintaining the quality of carbon taken from Vaal Reefs No. 8 CIP was further demonstrated by McArthur, 1987b (Table XX.4). The relatively high mass and attritional losses incurred when acid washing after regenerating at 750-850°C may be noted. Likewise, the decrease in loading capacity following extended carbon attrition would suggest localised surface activity which is readily lost by attrition.

Acid washing of carbon is also a recognised form of chemical regeneration which may well diminish the frequency of thermal regeneration in many instances. Carefully monitored carbon in pulp pilot operations at AARL indicated that non-regenerated carbon may be re-used at least 5-6 times with little fall-off in carbon activity being observed when using a hot acid wash (90°C) followed by the AARL elution. Acid washing of carbon following gold elution is not advisable as severe chloride corrosion in the furnace may be encountered.

20.8.1 Plant Practise

Indirect gas fired rotary kilns are used at Daggafontein for regeneration (20.2). Two kilns are utilised, (one per carbon stream) at a carbon throughput of approximately 520 kg/h/kiln. Many methods of checking and controlling throughput and carbon temperature in the kiln were tried, and the following found to be the most practical:-

The regeneration kiln discharges carbon into a quench tank from where the carbon is pumped, using recessed impeller morris pumps, over 1mm vibrating screens before being stored in a pachuca. The quench tank pumps remove carbon faster than the kiln discharge, which means that the quench tanks are usually empty of carbon. An impeller device (bindicator) has been installed into the quench tank at a predetermined height. If the quench tank pumps are stopped, carbon builds up in the quench tank and on reaching the bindicator, a torque alarm is sounded. Timing from pump stop to alarm allows the operator to calculate the carbon throughput of the respective kiln.

This method allows on-line throughput measurement, without handling carbon, and requires only a stopwatch and a pump stop and start. Any deviations are immediately corrected by adjusting the feed to the kiln. Measurements are carried out every 4 hours.

The kiln zone temperatures are automatically controlled, and the controllers, linked with accurate kiln feed rates allow for the desired carbon regeneration temperature to be maintained. To further supplement the system, operators have been trained to determine the correct carbon colour for normal operating conditions.

Any surges in the carbon flow prior to regeneration will require bypassing the kilns, as all the carbon fed to the kilns is regenerated at the required conditions; thus maintaining the philosophy "rather regenerate some of the carbon well, than all of the carbon badly".

20.8.2 Kinetic Activity Measurement

Due to the variable activity of virgin carbons (McArthur 1987a) the common practise of relating the activity of plant carbon to that of virgin carbon entails the use of an unnecessary variable and is thus most difficult to justify. As the measurement of kinetic activity is a purely empirical function, carbon activity in the regeneration circuit may be more effectively controlled using this function only. Furthermore as virgin carbon activity is reduced considerably by mechanical attrition, any attempt to regenerate circuit carbon to 100% of virgin activity will result in an unnecessary loss in carbon hardness due to over-activation.

The kinetic activity of circuit carbon at Daggafontein is maintained above 50% using the ACIX testing procedure. This level of carbon activity has been found to adequately maintain an adsorption efficiency of 98% in the contacting circuit.

20.9 CARBON CONSUMPTION

The consumption of carbon which may be seen to vary considerably in various operations (Table XX.5) constitutes a very significant operating cost. At a consumption of 60 g/t and a carbon price of R7/kg a direct reagent cost of R0,42/t is incurred. This of course does not reflect the contained gold lost with this degraded carbon - if this carbon had a gold content of 100 g/t, this would incur a further loss in revenue of R0,18/t treated. These data emphasise the importance of effective carbon selection and management.

Due to the difficulties associated with monitoring carbon losses, quantitative data relating carbon losses in associated unit operations of CIP plants are sadly lacking. In a serious attempt at rationalising these carbon losses (Sorenson, 1988) a review of 18 large CIP operations indicated carbon losses ranging from 23-263 g/t with an average loss of 69 ± 59 g/t. A further simulated study (Sorenson, 1989) of a plant treating 100 000 t/month and having a carbon consumption of 55 g/t indicated the following breakdown in carbon consumption : mixing (48,2%), pumping (0,7%), Kambalda mechanically swept screens (5,1%), regeneration (41,7%), quenching (2,7%), elution (0,9%) and acid washing (0,7%). This assessment would indicate that ca 50% of carbon is lost in the adsorption circuit via mechanical attrition while ca 40% is lost via oxidation in the regeneration circuit.

Apart from demonstrating a linear relationship between carbon inventory and ore treatment rate, the above studies were not able to identify any single process variable which had a significant influence on carbon consumption. The sample techniques currently in use would appear unable to adequately quantify the carbon inventory in the adsorption circuit. Moreover, this circumstance could result in an unaccounted gold loss/gain on an annual basis of as much as 180 kg in some circuits.

Quantification of areas of carbon losses has been very difficult to ascertain at Daggafontein. Poor reproducibility of sampling campaigns have highlighted the difficulties in quantifying these exact areas. The trend appears to be approximately 3 - 5% during the makeup sequence. (Fresh carbon is agitated under water and screened at 1mm before being fed to the adsorption section). The remaining carbon losses then appear to be equally split between adsorption and regeneration.

20.10 CARBON SELECTION

As stated earlier in this review, perhaps the most important aspect of effective carbon management relates to the selection of suitable carbon for a particular operation. Although the empirical test procedures generally used for carbon evaluation are difficult to relate to plant operation, the test data does provide considerable information concerning the physical and chemical characteristics of such commercial products thus affording some means of adjudication. This data besides being relatively costly (ca R1500/carbon sample at AARL) is however used only to rank tender samples in terms of each parameter tested viz. hardness, shape, activity, particle size etc. as no simple means of carbon evaluation has generally been accepted. Because of this difficulty in carbon evaluation, much of the data is not effectively used while in some instances inferior products may be purchased.

In the present review, an attempt is made to "weight" and to normalise the various chemical and physical parameters used in the adjudication of tender samples together with the "weighting" of the commercial price of the commodity so as to rationalise carbon selection using only technical and commercial considerations. By way of example, data using proposed new specifications (McArthur, 1987a) for the 18 tender samples submitted to AARL in July, 1988 are used to illustrate this weighting exercise.

The four principle parameters related to carbon evaluation viz. attrition hardness, shape, loading capacity and kinetic response may be normalised and "weighted" using a point count index totalling 100 points where the sum of the four weighted factors also totals 100 points. Each parameter may thus be defined (depending on the selected weighting factors) within specification limits as follows:

a) $$\text{Shape Index} = \frac{\text{Specification limit - shape factor}}{\text{Specification limit - optimum shape factor}} \times \text{weighting factor}$$

the specification limit used was 10%, while an optimum shape factor of 0% was selected.

b) $$\text{Attrition Index} = \frac{\text{Specification limit - attrition resistance}}{\text{Specification limit - optimum attrition resistance}} \times \text{weighting}$$

the specification limit used was 3,0%, while the optimum attrition resistance was adjudged to be 2,0%.

c) Capacity Index = $\frac{\text{K value - specification limit}}{\text{Optimum K value - specification limit}}$ x weighting factor

the specification limit used was 15 kg Au/tC, while an optimum K value of 19 kg Au/tC was selected.

d) Kinetic Index = $\frac{\text{R value - specification limit}}{\text{optimum R value - specification}}$ x weighting factor

The specification limit of 50% was used while an optimum R value of 62% was selected.

NOTE: as may be seen from the above 100 point count index, penalty points are incurred where carbon samples fail to meet certain specification limits.

As the moisture and solubles content and the fines content (less than 1,2 mm) of tender samples must also be considered in any carbon evaluation, these values expressed in percentage terms may be conveniently subtracted from the evaluation index which may then be expressed as follows:

Evaluation Index = Shape Index + Attrition Index
+ Capacity Index + Kinetic Index
- % moisture and solubles content
- % fines content

Using a selected price weighting factor each carbon may be "weighted" in terms of price using a price index as follows:

Price Index = $\frac{\text{PH - PS}}{\text{PH - PL}}$ x weighting factor

where PH = highest carbon price
PL = lowest carbon price
PS = sample carbon price

Having thus normalised and "weighted" the physical and chemical parameters of the carbon samples using an Evaluation Index, the Price Index may be added to give a Final Index which may be used to evaluate carbon tender samples i.e.

Final Index = Evaluation Index + Price Index.

When evaluating the 18 tender carbon samples using the proposed new specifications, the following weighting factors were assumed:

	"Weighting" Factor
Shape factor	25
Attrition hardness	25
Loading capacity	25
Kinetic response	25
Total	100

From the normalised test data presented in Table XX.4 using the procedure outlined, the following comments may be made:

a) a reasonably good correlation exists between the kinetic response and the CTC activity of samples from each major supplier; this correlation is not as marked in terms of carbon capacity;

b) samples supplied by A have generally a high fines content and poor shape characteristics;

c) samples suppled by B have excellent capacity characteristics and good shape factors;

d) All C samples have excellent attrition resistance;

e) All E samples have low fines content and good shape characteristics;

f) the excellent Evaluation Index of the C3 carbon and the consistent quality of the B carbons may be noted;

g) a good correlation exists between the Evaluation Index and the CTC activity and price (see below) in the case of the A, C, D & E samples.

In the present investigation, a price "weighting" factor of 50 was assumed which may be equated to a maximum of 33 1/3% of the Final Index. From the normalised data incorporating the Price Index presented in Table XX.5, several further comments may be made:

a) but for the B and C samples, no significant changes in carbon ranking are evident when comparing the Evaluation Index (Table XX.4) with the Final Index;

b) a good correlation exists between the Price Index and the CTC activity of samples from each major supplier.

While the weighting of the various parameters have been perhaps arbitrily chosen in the above selection exercise, the procedure does illustrate a means of selecting carbon which is based on only technical and commercial considerations. The development of a computer based software package whereby these weighting factors could readily be adjusted would be a logical step in further aiding effective carbon selection.

As may be seen, the above selection exercise was carried out using samples of carbon submitted for tender purposes and as such may not represent the quality of bulk deliveries. In order to obviate such an eventuality it is recommended that samples rather be taken from bulk deliveries in any serious selection campaign. Furthermore, this sampling of bulk deliveries as an on-going exercise will be the most effective means of improving the quality of carbon as supplied to the industry.

Because carbon consumption is such a significant reagent cost, the weighting of the price of the commercial product needs careful consideration. Thus the purchase of a good quality product with good hardness and shape characteristics at R8/kg incurring a consumption of 30 g/t (i.e. R0,24/t treated) is more cost effective than the use of a softer product priced at R4/kg but which results in a consumption of 100 g/t (i.e. R0,40/t treated). Needless to say this does not take into account the additional gold lost using the inferior product.

NOTE: Due to the presence of soft highly active material present in most commercial products (McArthur, 1987a), some form of attrition scrubbing prior to the monitoring of carbon activity is strongly recommended in any selection campaign. As this soft material is readily lost in the contacting circuit, this procedure more closely simulates the activity of the carbon when in actual use.

CONCLUSIONS

Being a dedicated high tonnage operation handling large quantities of carbon, Daggafontein has demonstrated that with effective carbon management, carbon consumption can be maintained at a cost effective level in a CIP type operation. The latest consumption data at Daggafontein indicates a carbon consumption of 26 g/t treated. Not withstanding the excellent carbon management displayed at Daggafontein, several improvements with regard to further limiting carbon losses warrant further emphasis when related to the industry in general:

Carbon Size

While the use of 8 x 16 mesh carbons (d_{50} 1,6-1,8 mm) are currently in general use in South AFrica, the advent of wedge-wire mechanically swept screens perhaps requires a slightly coarser carbon with d_{50} 1,8-2,0mm.

Carbon Shape

The use of wedge-wire interstage screens immediately places more emphasis on carbon shape. Purchasing specifications related to carbon shape require further consideration as some bulk deliveries have been shown to contain high flake contents.

Carbon Selection

Further attention to carbon selection based on the sampling of bulk deliveries is urgently needed in order to upgrade the quality of carbon used in the industry.

A carbon consumption of 10 g/t treated is attainable?

ACKNOWLEDGEMENTS

The authors wish to thank the Anglo American Corporation of South Africa Limited, Gold Division and the East Rand Gold and Uranium Company Limited for permission to publish this paper.

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TABLE XX.1

GOLD LOADING PROFILES AT DAGGAFONTEIN
FOR THE PERIOD APRIL-SEPTEMBER, 1989

Monthly Average	Loading Profile (g/t)					
	North Tankfarm					
	N3	N4	N5	N6	N7	N8
April	456	243	191	105	46	26
May	459	371	233	131	72	40
June	465	313	173	113	67	35
July	482	276	138	92	55	39
August	439	301	171	101	73	52
September	412	266	176	117	79	51
6 Month Average	452	295	180	110	65	41
	South Tankfarm					
	S3	S4	S5	S6	S7	S8
April	462	323	174	94	67	32
May	456	291	152	73	46	55
June	432	272	165	106	50	27
July	340	192	125	88	80	35
August	439	250	148	80	53	42
September	396	268	161	111	79	56
6 Month Average	421	266	154	92	63	41

TABLE XX.2

COMPOSITION OF CIRCUIT CARBONS AT DAGGAFONTEIN FOR THE PERIOD APRIL-AUGUST, 1988

Element	Carbon Composition (g/t)					
	Loaded Carbon		Acid Washed Carbon		Eluted Carbon	
	North	South	North	South	North	South
Ca	18 117	20 456	6 541	13 530	4 944	6 261
Ni	15 545	15 637	12 792	13 499	1 070	643
Cu	548	988	497	953	93	62
Fe	1 096	1 390	1 020	1 374	858	843
Si	4 314	6 258	3 457	4 462	2 346	3 266

TABLE XX.3

CARBON REGENERATION AT VARIOUS CARBON BASED OPERATIONS

Circuit Description	Pre-Acid Wash (Yes/No)	Pre-Drying (Yes/No)	Kiln Type	Regeneration Conditions	
				Carbon Temp. (°C)	Approx. Res. Time at Temp. (min)
Ergo	Yes	Yes	Rotary	730	40
Simmergo	No	No	Rotary	650	15
Daggafontein	Yes	Yes	Rotary	700	20
New Brand	Yes	No	Rotary	600-650	NA
Brand Calcine	Yes	No	Rotary	660-730	NA
City Deep	Yes	Yes	Rintoul	700	30
Crown Mines	Yes	Yes	Rintoul	700-750	90
Crown Mines	Yes	No	Rotary	750	
Harmony		Yes	Rintoul	700	60
Western Areas	No	No	Rotary	600-650	
Doornkop	No	No	Rotary	720	10
Grootvlei	Yes	No	Rotary	650	
Beatrix	No	Yes	Rotary	600	
St. Helena	No	No	Rotary	710	
Kinross	No		Rotary	590-610	
Af Lease	Yes	No	Rotary	600	3-5
VR No. 8	No	No	Rotary	600	10
VR No. 9	No	No	Rotary	400	10
WDL No. 1	Yes	Yes	Rintoul		
WDL No. 3	Yes	Yes	Rintoul	650	30

TABLE XX.4

THE EFFECT OF ACID WASHING ON CARBON QUALITY

Test Parameters Measured	Regeneration Temperature (30 min. contact time + steam)					
	750°C		800°C		850°C	
Mass loss (%)						
acid wash before regen.	0,36		3,15		3,25	
acid wash after regen.	0,73		3,14		6,23	
$ Attrition resistance (%)						
acid wash before regen.	0,9		1,0		1,7	
acid wash after regen.	1,8		1,4		2,6	
* Kinetic activity (%)						
acid wash before regen.	52	(43)	53	(48)	56	(56)
acid wash after regen.	51	(50)	51	(53)	56	(56)
* Carbon capacity (kg Au/tc)						
acid wash before regen.	20	(18)	20	(18)	19	(19)
acid wash after regen.	24	(17)	24	(17)	24	(17)

$ attrition measured after 5 d using standard AARL procedure.

* kinetic activity and carbon capacity measured both before and (after) extended attrition using AARL procedures.

TABLE XX.5

CARBON CONSUMPTIONS IN VARIOUS CIP OPERATIONS*

Circuit Description	Consumption (g/t - type)	Circuit Description	Consumption (g/t - type)
Ergo	45 - YAO + 205C	New Brand Mod. 1	31 - G210
Simmergo	76 - YAO	New Brand Mod. 2	24 - G209
Daggafontein	32 - YAO + ANK10	New Brand Mod. 3	45 - YAO
City Deep	47 - ANK11	Brand Calcine	46 - G210
Crown	80 - TN5	Grootvlei	45-60 - ANK11
Harmony	54 - GRC22 + ANK11	Beatrix	80 - TN5
Western Areas	26 - ANK11	St. Helena	60 - TN5
Doornkop	28 - ANK11	Kinross	90-150 - TN5
Vaal Reefs No. 8	39 - ANK11	WDL No. 1	48-G210+GRC22
Vaal Reefs No. 9	43 - ANK11 + NORIT	Af. Lease	85-G210

* 6 month survey ending December 1988.

TABLE XX.6: CARBON EVALUATION USING NORMALISED LABORATORY TEST DATA

Sample Identification*	Shape Index	Attrition Index	Capacity Index	Kinetic Index	Moisture & Solubles	Fines Content	Evaluation Index	Ranking No.
A1	-11,3	17,5	6,3	- 2,1	1,5	3,9	5,0	15
A2	-20,3	- 5,0	6,3	0	2,8	6,7	-28,5	18
A3	- 6,3	10,0	12,5	22,9	9,5	4,6	25,0	12
A4	- 4,3	0	12,5	18,8	1,9	4,3	20,8	13
A5	- 8,8	10,0	25,0	22,9	3,3	3,2	42,6	5
B1	14,8	15,0	25,0	0	2,3	0,6	51,9	2
B2	18,5	0	18,8	6,3	1,8	0,4	41,4	6
B3	14,5	7,5	18,8	12,5	2,9	1,1	49,3	3
C1	4,0	20,0	6,3	- 8,3	2,4	3,3	16,3	14
C2	7,5	17,5	6,3	6,3	1,2	3,3	33,1	11
C3	12,5	22,5	12,5	25,0	1,2	1,0	70,3	1
D1	- 0,8	-10,0	12,5	0	5,2	1,7	- 5,2	17
D2	12,8	0	12,5	12,5	3,2	0,2	34,4	8
D3	7,5	- 5,0	12,5	22,9	3,6	0,4	33,9	10
E1	14,3	5,0	- 6,3	- 6,3	3,5	0,1	3,1	16
E2	14,8	12,5	0	12,5	5,5	0	34,3	9
E3	19,0	7,5	6,3	16,7	4,9	0	44,6	4
F1	- 0,8	17,5	25,0	0	3,1	1,3	37,3	7

* Samples listed in order of increasing CTC activity for each major supplier

TABLE XX.7

FINAL CARBON SELECTION USING CURRENT WEIGHTED COMMODITY PRICES

Sample Identification	Evaluation Index (total 50)	Price Index (total 150)	Final Index Ranking	Final
A1	5,0	46,5	51,5	14
A2	-28,5	43,2	14,7	18
A3	25,0	36,8	61,8	9
A4	20,8	31,7	52,5	13
A5	42,6	26,4	69,0	5
B1	51,9	19,2	71,1	3
B2	41,4	12,1	53,5	11
B3	49,3	0	49,3	15
C1	16,3	50,0	66,3	7
C2	33,1	38,5	71,6	2
C3	70,3	23,3	93,6	1
D1	5,2	28,8	23,6	17
D2	34,4	23,3	57,7	10
D3	33,9	18,8	52,7	12
E1	3,1	37,6	40,7	16
E2	34,3	31,3	65,6	8
E3	44,6	22,6	67,2	6
F1	37,3	33,4	70,7	4

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FIGURE XX.1. SIMPLIFIED DAGGAFONTEIN FLOWSHEET.

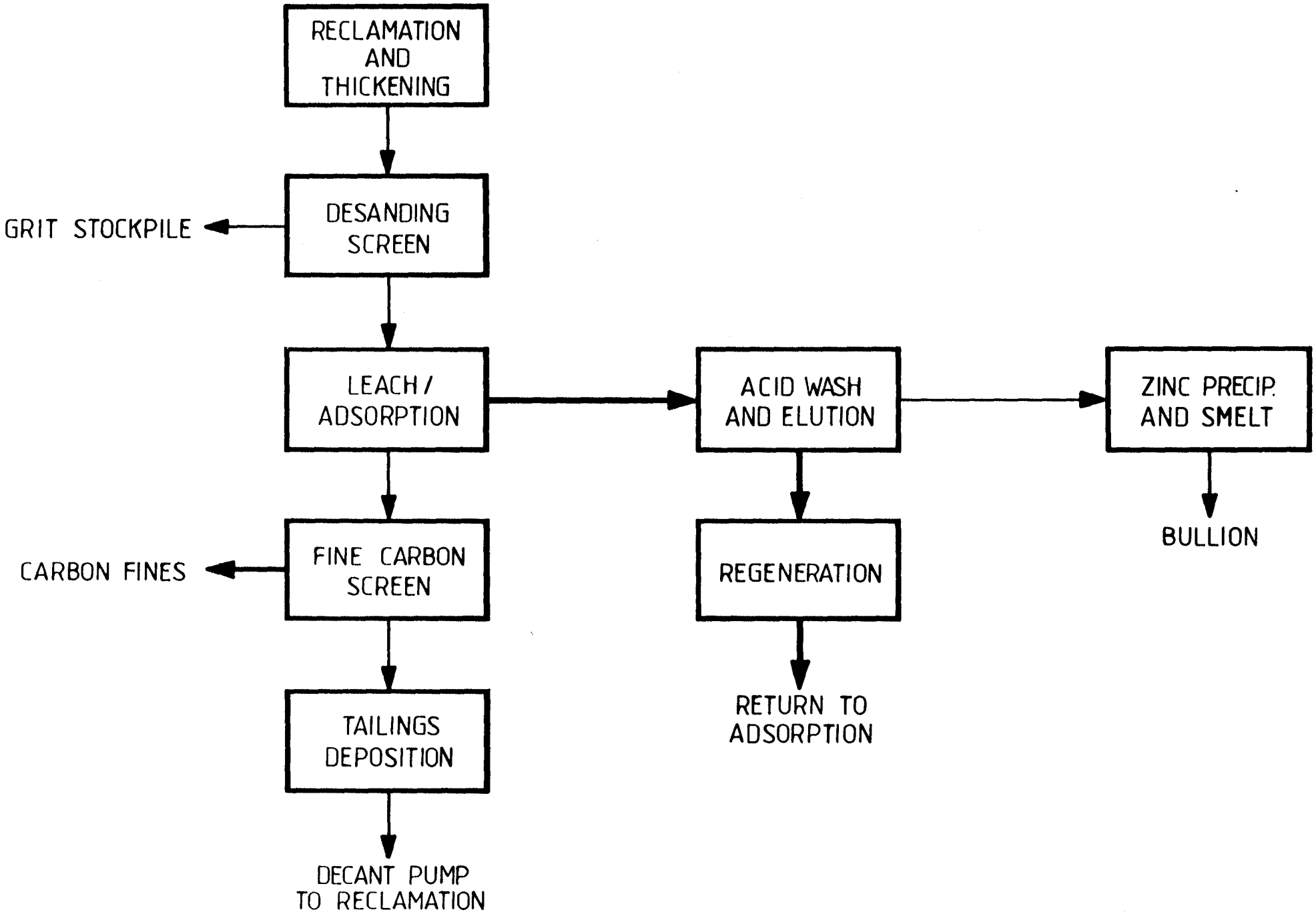
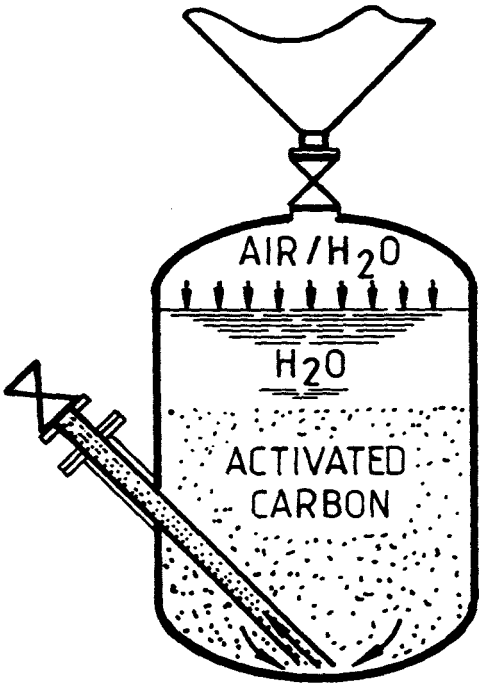
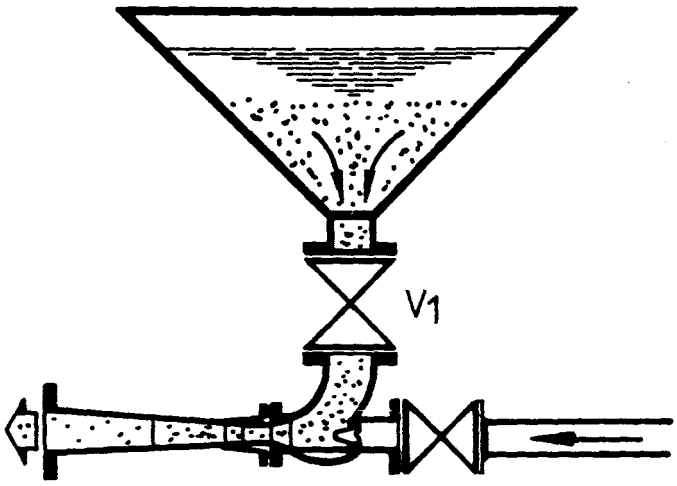


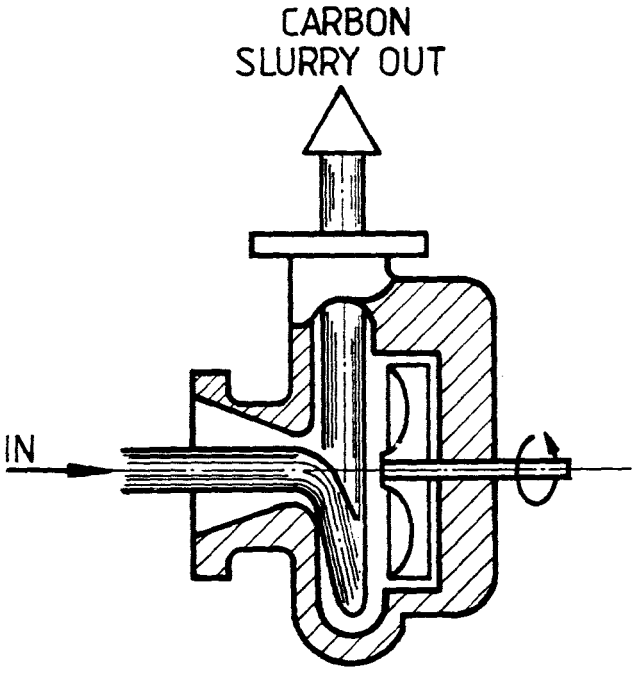
FIGURE XX. 2. EQUIPMENT USED FOR CARBON TRANSPORT



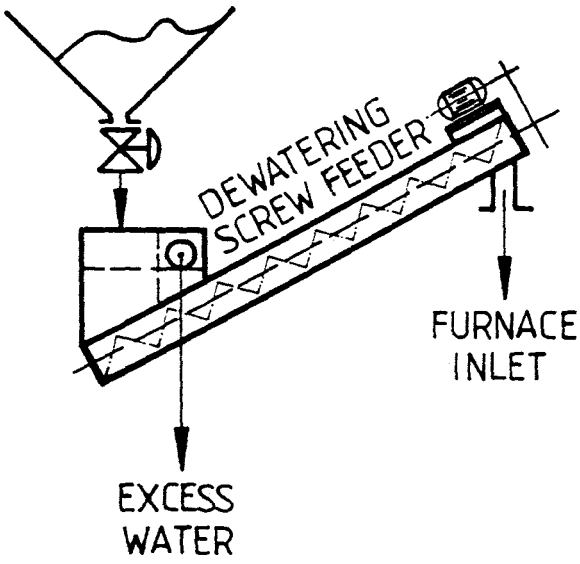
(a) BLOWCASE.



(b) EDUCTOR.



(c) RECESSED IMPELLER PUMP.



(d) INCLINED SCREW FEEDER.

FIGURE XX.3. MODIFIED ANGLO N.K.M. INTERSTAGE SCREEN.

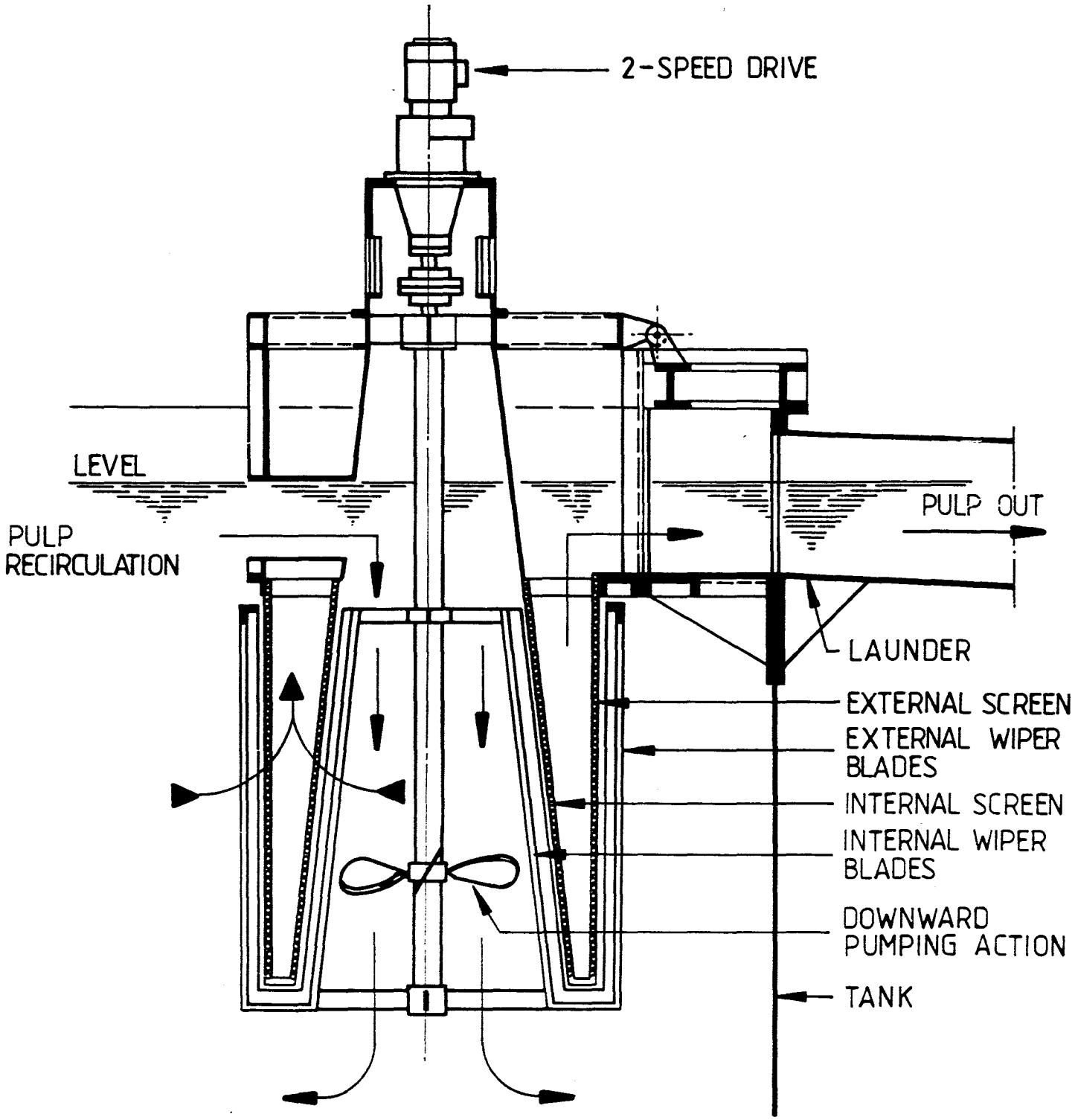
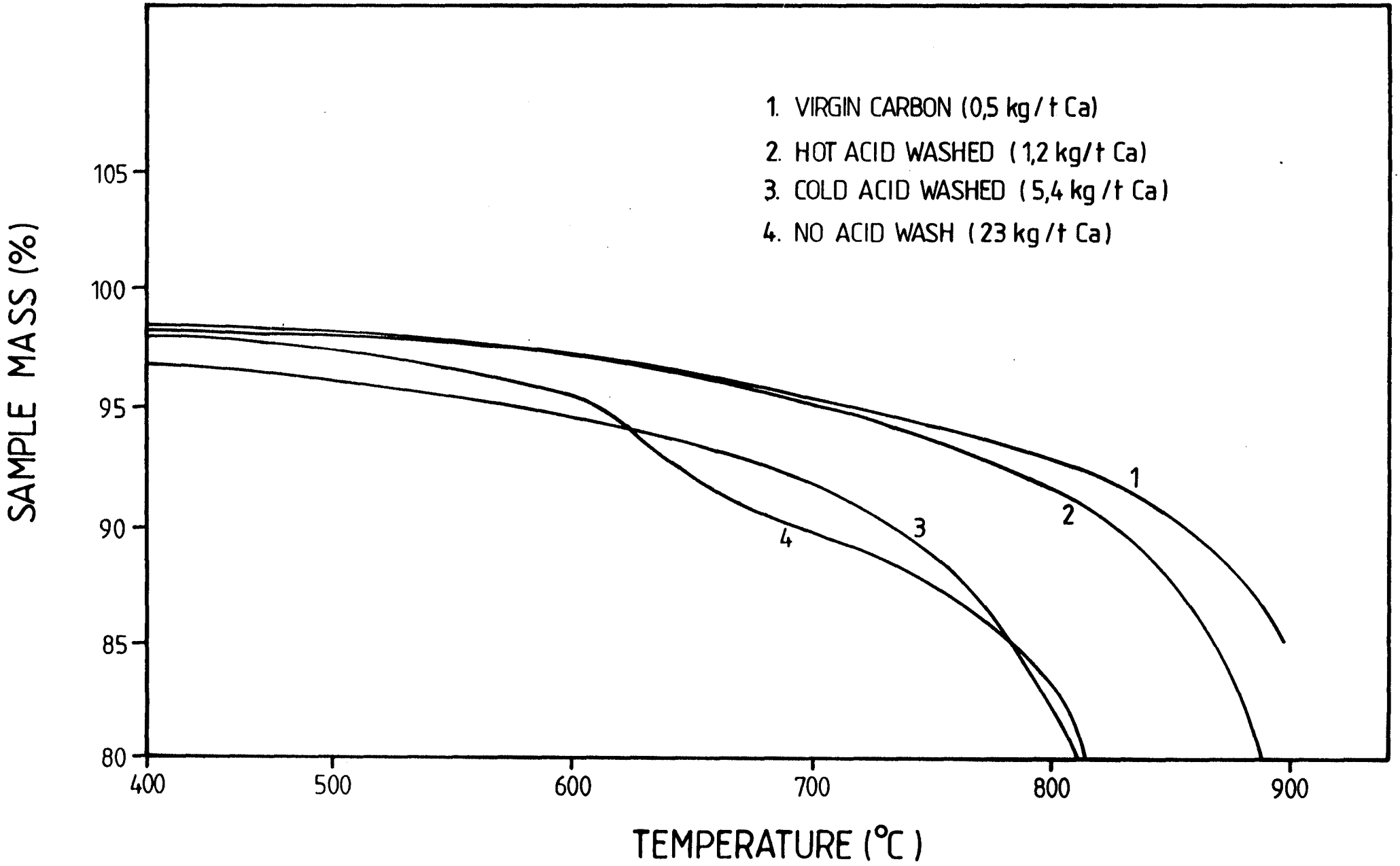


FIGURE XX.4. THERMOGRAVIMETRIC ANALYSIS OF ELUTED CARBON SAMPLES ORGINATING FROM A CARBON-IN-PULP PLANT.



7.0 GENESIS OF GOLD IN THE WITWATERSRAND

The appended investigation is an attempt by an extraction metallurgist to relate the chemistry of gold extraction by activated carbon (algal biomass) to that of the mineralogy of the Witwatersrand.

J. S. Afr. Inst. Min. Metall., vol. 90, no. 3.
Mar. 1990. pp. 53-57.

A pre-burial adsorption model for the genesis of gold in the Witwatersrand

by R.J. DAVIDSON*

SYNOPSIS

The chemistry related to the adsorption of gold, uranium, and transition-metal ions onto algal biomass (activated carbon) is related to the genesis of the Witwatersrand.

Detrital gold being washed into the early basin, together with cyanide generated in the atmosphere, would have solubilized as the stable aurocyanide complex, the necessary oxygen being produced from the photosynthesis of carbon dioxide by primitive algae. Pebble-type conglomerates, while providing for the free movement of solutions containing the vital nutrients for prolific algal growth, would have screened off the sunlight, which would otherwise have degraded any free cyanide in solution. With the subsequent decomposition of the algal deposits, it is surmised that carbon-rich layers having adsorptive properties similar to those of activated carbon formed in the conglomerates. Under these conditions, gold (silver) in solution would have been adsorbed selectively as the cyanide complex, together with uranium as the carbonate complex. Any transition-metal ions present would also have been adsorbed onto such a substrate.

The subsequent burial and compression of the gold-rich conglomerate with temperatures rising to about 400°C would then have reduced the adsorbed gold to the metal in a single segregated gold-silver metal phase (95 to 99 per cent of the contained gold), while leaving behind a sulphide phase rich in iron, base metals, and arsenic and containing some gold and silver. The recrystallization of this latter phase may explain the presence of the rather consistent contents of secondary pyrite in the Witwatersrand conglomerates. Similarly, uranium minerals, some rich in titanium and carbon, may also have formed.

An adsorption model would explain the very consistent trends in the gold-to-silver ratios of individual reefs in the Witwatersrand, which suggest an extensive hydrothermal system approaching isothermal equilibrium. Also, as gold grades increase, so silver grades generally decrease, indicating the sequential displacement of silver by gold as classically obtained with activated carbon.

SAMEVATTING

Die chemie verbonde aan die adsorpsie van goud, uraan en oorgangsmetaalione op 'n algale biomassa (geaktiveerde koolstof) hou verband met die oorsprong van die Witwatersrand.

Detritale goud wat in die vroeë kom ingespoel is, het saam met sianied wat in die atmosfeer gegenereer is, oplosbaar geraak as die stabiele ourosianiedkompleks. Die nodige suurstof is verskaf deur die fotosintese van koolstofdioksied deur primitiewe alge. Spoelklipagtige konglomerate wat vir die vrye beweging van oplossings met die lewensbelangrike voedingstowwe vir die weelderige groei van alge gesorg het, het die sonlig afgekeer wat andersins enige vry opgeloste sianied gedegradeer het. Met die daaropvolgende ontbinding van die algafsettings het daar vermoedelik koolstofryke lae met adsorpsie-eienskappe soortgelyk aan dié van geaktiveerde koolstof in die konglomerate gevorm. Onder hierdie omstandighede sou opgeloste goud (silwer) selektief as die sianiedkompleks, saam met uraan as die karbonaatkompleks geadsorbeer word. Enige aanwesige oorgangsmetaalione sou ook op so 'n substratum geadsorbeer word.

Die daaropvolgende begrawing en saampersing van die goudryke konglomeraat met temperature wat tot ongeveer 400 °C gestyg het, sou dan die geadsorbeerde goud tot die metaal gereduseer het in 'n enkele gesegregeerde goud-silwermetaalfase (95 tot 99 persent van die ingeslote goud), terwyl dit 'n sulfiedfase ryk aan yster, onedelmetale en arseen agtergelaat het wat ook 'n bietjie goud en silwer bevat het. Die herkristallisering van laasgenoemde fase kan moontlik die aanwesigheid van 'n taamlik konstante sekondêre pirietinhoud in die Witwatersrandse konglomerate verklaar. Uraanminerale, waarvan sommige ryk aan titaan en koolstof is, kan ook op dieselfde wyse gevorm gewees het.

'n Adsorpsiemodel sal die uiters konstante tendense in die goud-silwer verhouding van individuele riwwe van die Witwatersrand verklaar wat dui op 'n uitgebreide hidrotermiese stelsel wat na aan isotermiese ewewig kom. Die silwergrade wat oor die algemeen afneem namate die goudgrade toeneem, dui ook op die sekwensiële verplasing van silwer deur goud wat algemeen met geaktiveerde koolstof verkry word.

Introduction

Recent studies[1] on the genesis of the gold in the Witwatersrand tend to indicate that hydrothermal fluids have affected a large part of the basin and, as such, raise serious questions concerning the current placer model[2]. The predictive capacity of the placer model is questioned here, and an alternative chemical viewpoint is presented that emphasizes the possible role played by activated carbon in the genesis of the Witwatersrand.

A rather incongruous relationship between the noble metal gold and the common element carbon is often apparent in many gold-bearing orebodies. Not only are gold and carbon very often found together in the same stratigraphical horizon in nature—perhaps indicating some sort of genetic relationship—but activated carbon is well known as a very selective adsorbant for gold and is widely used in many extraction circuits. Questions must therefore be asked as to whether the chemistry involved in the extraction of gold by activated carbon bears any relationship to the genesis of the gold in the Witwatersrand. The present paper is a preliminary attempt to answer some of these questions.

* 120 13th Street, Victory Park, Johannesburg 2095.
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Activated Carbon

In many ways activated carbon is a unique material with properties not possessed by any other class of organic substance. Since activated carbon usually contains considerable quantities of adsorbed oxygen and hydrogen, the term *carbon* is perhaps a misnomer. It is essential to regard activated carbon as a complex organic polymer, and not merely as an amorphous form of the element carbon. Activated carbon, however, differs from other organic polymers in having an exceptionally large surface area and in being a good conductor of electricity. Many of its intriguing characteristics stem from the latter fact, as well as being related to the various functional groups present on the carbon surface. A better understanding of both the complex physical and chemical characteristics of activated carbon is thus a prerequisite to an understanding of the factors influencing gold adsorption and to the formulation of any adsorption mechanism.

Activated carbon can be made by the heating of virtually any carbonaceous material in a mildly oxidizing atmosphere. The physical and chemical properties of the activated carbon are influenced to a large degree by both the nature of the source material and the conditions of activation. During the course of activation, the carbon develops a porous graphitic structure of molecular dimensions with an extraordinarily large internal surface area on which adsorption can take place. The quantity, size, and shape of these pores distinguish both the various types of activated carbons and their subsequent application. In general, raw materials with a dense, cellular structure produce a hard, brittle product, and it is for this reason that carbons made from coconut shell are used almost exclusively in the gold-mining industry.

Activated carbons are, however, also produced in nature[3]; many peats, brown coals, and even decomposed vegetation display similar adsorptive properties, particularly with regard to the selective adsorption of gold. In particular, algal biomass[4] has been shown to be a binding medium for transition metals, as well as for gold. Moreover, gold is efficiently and reversibly bound to the algal biomass from gold solutions of as low as 10^{-9} M. The ligating atom in the biomass appears to be sulphur.

Adsorption of Gold

The selective adsorption of metal–cyanide complexes onto activated carbon appears to be related to the solubility and charge of the ion-paired adsorbed species. While the less soluble monovalent dicyano complexes of gold, silver, and copper have all been shown to be strongly adsorbed by carbon, the more soluble divalent and trivalent tri- and tetracyano complexes (e.g. copper, nickel, and zinc) are only weakly adsorbed, while the very soluble hexa- and octacyano complexes (e.g. cobalt and iron) are not adsorbed to any marked degree.

Under plant conditions, the selective adsorption of gold onto activated carbon is thought to be related to the solubility of the ion-paired species Ca^{2+} $[Au(CN)_2^-]_2$. Conditions favouring gold adsorption include low temperature, low cyanide concentrations, low alkalinity, and the presence of an adsorption medium of high ionic strength containing Ca^{2+} ions. Also, the concentration of free cyanide has a significant effect on the selectivity of carbon for gold from typical plant solutions or pulps. Under plant conditions, activated carbon is most selective in adsorbing gold from such complex mixtures. Selective metal adsorption appears to take the form of a complex displacement sequence, gold followed by silver being the most strongly adsorbed metal species. No well-defined sequence of metal displacement can be formulated since the adsorption of these metal values is sensitive to such processing parameters as pH, free cyanide concentration, temperature and, of course, the relative concentration of the metal values in a particular plant solution or pulp. In general, the selectivity of carbon for gold increases with an increase in pH or in free cyanide concentration.

Adsorption of Uranium

Attempts at the Anglo American Research Laboratories (AARL) to adsorb uranium onto activated carbon under oxidizing acid conditions have not been successful. However, activated carbon pretreated with titanium chloride or with zirconium chloride has been shown to be very efficient at scavenging uranium.

As tne reduced U^{+4} carbonate complex of uranium is less soluble than the U^{-6} complex, the adsorption of uranium onto activated carbon under alkaline conditions as the carbonate complex can be expected to be more efficient. (A large commercial plant in Japan is using titanium-impregnated carbon to recover uranium from sea water.)

Proposed Model

It seems likely that solar ultraviolet light and lightning were the main sources of energy to affect the chemistry of the early atmosphere[5]. In Miller's classic experiments (1953), in which mixtures of CH_4, NH_3, H_2, and H_2O were subjected to electrical discharges, substantial quantities of organic polymers and amino acids, together with large amounts of cyanide and aldehydes, were formed. The organic chemistry of the early oceans was thus probably most complex. Organic compounds including cyanide were almost certainly generated in the atmosphere and carried into the oceans.

In the case of the Witwatersrand inland sea, it is thus probable that gold in the form of the very stable $Au(CN)_2^-$ complex ($k = 10^{-38}$) tended to concentrate in solution. Any detrital gold being washed into this environment may thus have been solubilized, the necessary oxidant being produced from the photosynthesis of CO_2 by the living algae:

$$xCO_2 + yH_2O + hv \rightarrow Cx(H_2O)_y + O_2.$$

Thus, while the early atmosphere may have been mildly reducing and deficient in oxygen, the Witwatersrand basin may well have been saturated with oxygen via this reaction.

It should be noted that $Au(CN)_2^-$ is stable towards high doses of ultraviolet light and ozone, while $Ag(CN)_2^-$ ($k = 10^{-22}$) becomes partially decomposed.

Hydrothermal springs from faults along the edge of the basin[6] could have been an additional source of gold.

In regard to the free movement of solutions through an original sedimentary geological formation, only rudites such as a pebble type of conglomerate would afford such free flow. The presence of solutions containing vital

nutrients for the proliferation of primitive algal growth (e.g. potassium and ammonia) may explain the very wide extent of such algal growth in the Witwatersrand conglomerates some 3000 million years ago. Little evidence for such algal growth is present in the quartzites and sandstones. Furthermore, a pebble type of formation would have screened off sunlight, which would otherwise have degraded any cyanide in solution.

With the subsequent decomposition of such algal deposits under what were thought to be mildly oxidizing conditions, a deposit of active biomass displaying adsorptive properties may have formed. Under these conditions, gold (silver) in solution as the cyanide complex would have been selectively adsorbed onto this active substrate.

Furthermore, the active substrate, in which the ligating atom appears to be sulphur, would also have scavenged the iron and base metals present in very dilute solution. Uranium present as the carbonate complex may also have been adsorbed onto the carbon-rich biomass.

The subsequent burial and compression of the gold-rich conglomerate with temperatures rising to about 400 °C would then have reduced the adsorbed gold to the metal as a single gold–silver metal phase (95 to 99 per cent of the contained gold), leaving behind a sulphur-rich phase containing iron, base metals, and arsenic[7] with some gold and silver. The recrystallization of this latter phase may explain the presence of the rather consistent secondary pyrite found in the Witwatersrand conglomerates, and may also explain the sulphurization of the heavy ferruginous minerals as proposed by Ramdohr[8]. Similarly, uranium minerals, some rich in titanium and carbon, may also have formed.

Gold-to-Silver Ratios

The most consistent trends in the gold-to-silver ratios of individual reefs in the Witwatersrand basin are explained satisfactorily only in terms of an extensive hydrothermal system approaching isothermal equilibrium. In a major analytical survey[9] of the economic reefs being mined in the Anglo American Group, a near-perfect linear relationship was obtained in all instances for gold grades plotted against total precious-metal values (gold + silver) (Table I). As an example, the values obtained from the Leader Reef in the Welkom area are presented in Fig. 1. This graph also shows that, in all the reefs considered, as the gold grades increased, so also did the gold-to-silver ratio, as illustrated in Fig. 2. The sequential displacement of silver by gold as classically obtained using activated carbon is strongly suggested.

While it would appear that 90 to 95 per cent of the contained gold in the Witwatersrand reefs is present as 20 to 50 μm electrum particles with gold-to-silver ratios approaching 10 in the case of the higher-grade reefs, a small proportion (1 to 2 per cent) of the gold is associated with the pyritic phases, together with relatively high base-metal and arsenic values. The gold-to-silver ratios in these phases are very different, falling in the region between 0,6 and 0,9. The low solubility of silver sulphide would explain the concentration of silver in these pyritic phases.

Particle Size

Both the small size and the very often irregular shapes of the particles of detrital gold found in the Witwatersrand are difficult to explain according to a placer model. On the other hand, gold adsorbed onto activated carbon as the aurocyanide complex and subsequently heated to 400 °C, which would reduce the gold to the metal, may well resemble these particles of detrital gold. At ambient temperatures, activated carbon is not able to reduce gold to the metal when adsorbed as the aurocyanide since the reduction potential of $Au(CN)_2^{-1}$ is −0,6 V (n.H.e) compared with 0,0 V for activated carbon[10]. However, on being heated to about 400 °C, the gold that had migrated to the surface of the carbon would have been reduced to the metal in small irregular-shaped particles. A gold–silver mixture adsorbed onto activated carbon as the respective cyanide complexes in the ratio 10 Au : 1 Ag has been observed to respond in a similar way. It was also observed that the particle size of the reduced gold was very dependent on the particle size of the carbon used. (Synthetic ore aimed at simulating a milled reef is prepared at AARL using loaded activated carbon that is calcined subsequently to remove the carbon.)

TABLE I
THE LINEAR RELATIONSHIP BETWEEN GOLD AND SILVER IN VARIOUS WITWATERSRAND REEFS

Reef	Plot of Y(Au g/t)* = *a* X(Au g/t + Ag g/t)† + *b*		
	a	*b*	Correlation coefficient
Western Holdings			
Leader Reef	0,906	−0,228	0,994
Vaal Reefs			
Elsburg Reef	0,909	−0,497	0,999
Ventersdorp Contact	0,899	−0,368	0,997
Vaal Reefs	0,912	−0,479	0,999
Western Deep Levels			
Ventersdorp Contact	0,929	−1,297	0,999
Carbon Leader	0,913	−0,658	0,999
Welkom			
Basal Reef	0,923	−0,540	0,999
Leader Reef	0,910	−0,360	0,999

* Fire assay followed by parting
† Fire assay (gravimetric)

Chemical Associations

Gold occurs in two important end-member chemical associations in the Witwatersrand[2]: the first, corresponding to between 95 and 99 per cent of the contained gold, is present as discrete particles of electrum and is characterized by Au–C in carbon-rich seams or layers, and the second is Au–Fe–S–As, being characterized by pyrite-bearing conglomerates. Similarly, uranium has two important associations: U–Ti in the conglomerates, and U–C in the carbon-rich seams. These chemical associations are found in reefs throughout the basin, but they are not mutually exclusive and are rarely discrete.

Dilute Solutions

Activated carbon is currently used to scavenge gold efficiently from dam return circuits with a gold concentration of only 20 ng/t (10^{-9}), where loading on the

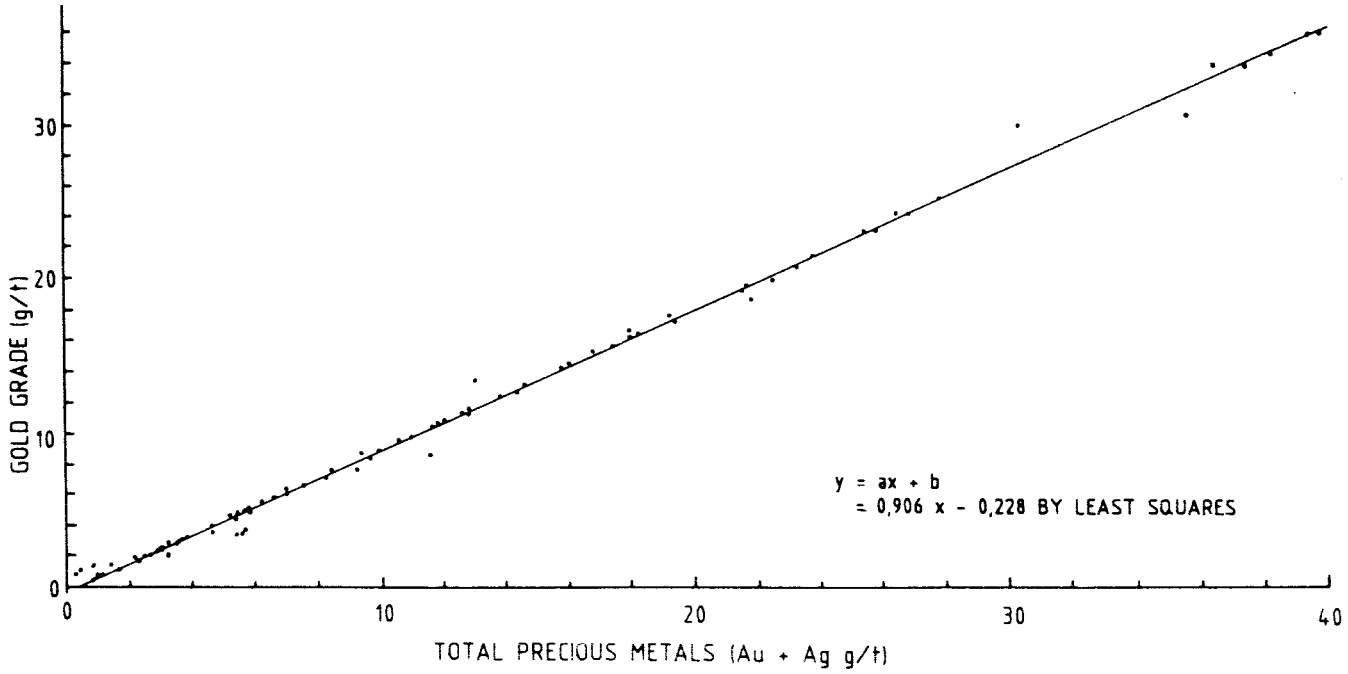

Fig. 1—The linear relationship between the gold and silver values found in the Welkom Leader Reef

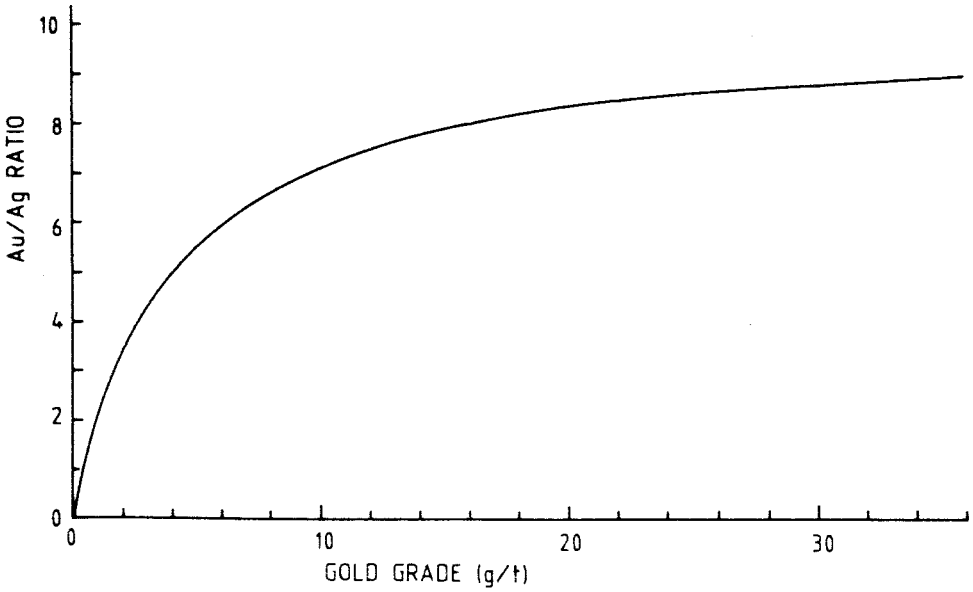

Fig. 2—The increase in gold-to-silver ratio with gold grade in the Welkom Leader Reef, illustrating the sequential displacement of silver by gold

carbon has approached x50 000, i.e. giving a gold loading on the carbon of 1 kg/t. The concentration of gold in sea water is estimated to be 5 pg/t (10^{-12}). Also, algal biomass[4] has been shown to efficiently and reversibly adsorb gold present as $Au(CN)_2^-$ at concentrations as low as 10^{-9} M.

Carbon Deportment and Decomposition

Any decaying algal deposit would have become sedimented to the bottom of the conglomerate prior to burial, with the pebbles acting as a trap for any such material. Subsequent decomposition (and activation) of the biomass with carbon being lost as carbon dioxide to the atmosphere was most likely. (Activated carbon is a most effective scavenger for oxygen.) With this decomposition, high residual potassium and aluminium values would have occurred, which may explain the presence of the phyllosilicates found as cementing material in the conglomerates. Following adsorption, high gold values, particularly towards the footwall of the reef, may also have occurred.

Other Lixiviants

While the proposed adsorption model would suggest that gold is adsorbed as an aurocyanide complex, other possible lixiviants should also be considered.

(1) For gold dissolved as the $AuCl_4^-$ complex, an acidic chloride medium is required. This complex is strongly adsorbed onto activated carbon. Having a reduction

potential E° = 1,0 V, it is reduced directly to the metal by activated carbon (E° = 0,0 V). However, the silver complex $AgCl_2^-$ (E° = 0,2 V) is soluble only in the presence of excessive salt concentrations (200 to 300 g of NaCl per litre) and, even if silver had been adsorbed onto carbon, its sequential displacement by gold would not have been possible because of the direct reduction of both the gold and the silver to their respective metals. No adsorption of sulphide ions would occur in an acidic chloride medium because of the instability of sulphide ions in such a medium.

(2) Gold forms a very soluble sulphide, AuS^-, which is stable under alkaline conditions but which is not adsorbed onto activated carbon. However, silver sulphide is very insoluble and, if adsorbed onto carbon, is difficult to elute. The selective adsorption of gold and silver onto activated carbon as sulphides would thus appear to have been impossible.

(3) While gold and silver are both soluble in ammoniacal solution as their thiosulphate complexes, little information on the adsorption of these complexes onto activated carbon is currently available. Alkaline sodium thiosulphate at ambient temperature is an excellent eluant for silver adsorbed onto carbon as the cyanide complex[11], suggesting that the silver–thiosulphate complex is not adsorbed by activated carbon.

Discussion

The present preliminary investigation is an attempt by an extraction metallurgist to relate the chemistry of gold extraction by activated carbon to that of the mineralogy of the Witwatersrand—a subject with which the author is not familiar. Nevertheless, while certain hydrothermal models for the Witwatersrand appear to be gaining some credibility, they are limited to post-burial effects. The present adsorption model suggests that pre-burial effects should also be recognized. Both the extent and the uniformity of the gold-bearing conglomerates in the Witwatersrand, together with the very consistent gold–silver values found, tend to support such a model.

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Jet cutting

BHRA, The Fluid Engineering Centre, has released the final details of its Tenth International Symposium on Jet Cutting Technology.

The organizers have selected the Sonesta Hotel in Amsterdam, Holland, as a suitable venue for this meeting. The Symposium will be held in the Sonesta from 30th October to 2nd November, 1990, a fortnight earlier than originally scheduled.

Jetting techniques are now extensively employed for a wide range of tasks—from cutting difficult materials in hazardous environments to dismantling reactor vessels, as well as solving many cleaning and descaling problems. Today, the jetting industry is working to improve the profitability and efficiency of its systems in order to extend the variety of applications of this valuable cold-cutting technique. This meeting will enable researchers, systems manufacturers, contractors, and end users to discuss current practice as well as explore opportunities for the future.

Further details can be obtained from

The Tenth Jet Cutting Symposium Organizer
BHRA, The Fluid Engineering Centre
Cranfield
Bedford MK43 0AJ
England.

Telephone (0234) 750422
Telex 825059
Fax (0234) 750074.

CHAPTER II

IMPROVEMENTS IN GOLD EXTRACTION

1.0 INTENSIVE CYANIDATION OF GRAVITY CONCENTRATES

Due to the health hazard associated with the amalgamation of gold plant gravity concentrates, an alternative process route using cyanide was investigated at the AARL in 1976-78. The advantages sought from such a route included low-risk working conditions, together with the elimination of the large recycling load of concentrates in the gravity circuit and perhaps a less labour intensive operation. A gold recovery equal to or better than amalgamation was also a prime requirement of such an alternative process.

Following an in-depth study (see appended paper), a suitable reactor was developed (SA Patent No. 77/1777) resulting in the commissioning of full scale plants at Welkom Gold Mine, Vaal Reefs Mining and Exploration, Western Holdings Gold Mine and Dome Mines (Canada). Average gold recoveries of greater than 99% indicated in the South African operations with leach residues of 10-50 g/t gold being reported. The AARL electrowinning cell copyright which has since been used in several CIP operations was also developed during this investigation.

The intensive cyanidation of gold-plant gravity concentrates

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SYNOPSIS

The development of a cyanidation route for the treatment of gold-plant gravity concentrates is described. This work was carried out as a natural consequence of earlier improvements to the recovery of gold in the gravity sections of existing gold plants. Concern about the environmental aspects of the conventional amalgamation process was further motivation in the development of an alternative procedure.

It was found that the partial pressure of oxygen in solution was the most critical reaction-rate determining factor. The rate of gold dissolution was shown to be controlled by the reaction rate below approximately 25 °C and by diffusion above that temperature. Thus, the efficiency of pulp agitation and oxygen partial pressure permitted a high degree of control over the rate of gold dissolution. The use of oxygen rather than air significantly increased the dissolution rate of both gold and silver. The solubilities of sodium, calcium, and potassium aurocyanide complexes under simulated reaction conditions were studied, and it was established that the values for the sodium and calcium complexes were limited, requiring the use of lower pulp densities and higher reaction temperatures for the treatment of very high-grade concentrates. The presence of excessive amounts of tramp iron in such concentrates of up to 25 per cent by mass was found to result in poor gold dissolution owing to the cementation of gold under certain reaction conditions. The maintenance of high oxygen partial pressures, together with high cyanide concentrations, high pH levels, and the presence of calcium in the system, all aided in passivating the tramp iron. Copper, gold, and iron were shown to be the main cyanide-consuming elements.

Optimum gold dissolution was obtained in a closed bench-scale reactor incorporating a flotation-machine type of agitator for effective pulp agitation and oxygenation, the oxygen being recirculated. The reactor was designed to operate under a very small positive pressure. After reaction times of 2 to 6 hours at temperatures in the region of 30 °C, both the gold and the silver dissolution averaged 99,5 per cent. In the case of very high-grade concentrates, it was shown that gold could be floated in the same reactor by drawing off the flotation concentrate through a port in the reactor. A flotation recovery of 87,4 per cent of the gold in 3,9 per cent by mass was obtained.

The recovery of osmiridium from cyanide-leach residues and the electrowinning of gold from the pregnant liquors were also considered as processing aspects in the treatment of gold-plant belt concentrates. Final table tailings with a gold concentration of 10 to 12 g/t were obtained.

The results of the investigation strongly suggested that an alternative cyanidation route for the treatment of gold-plant concentrates was technically feasible, provided the system included effective agitation and oxygenation. In the case of lower-grade concentrates, cyanidation followed by the electrowinning of gold appears to be the simplest route, while, for high-grade concentrates, flotation and smelting of the concentrate followed by cyanidation of the flotation tailings are suggested. The use of oxygen rather than air for sparging the pulp would depend largely on the grade of material to be treated. A preliminary comparison of the costs related to the amalgamation and cyanidation processes for the recovery of gold from such concentrates indicated very similar capital and operating costs. It was concluded that other considerations, e.g. toxicity, safety, and lower labour requirements, were more important than such small cost differences. Furthermore, the 'opening' of the gold-plant gravity circuit, which would be made possible by a cyanidation route, would further reduce the operating costs because of the subsequent decrease in the tonnage to be treated.

SAMEVATTING

Die ontwikkeling van 'n sianiedproses, vir die behandeling van goudaanlegswaarkonsentraat, word beskryf. Die werk was 'n natuurlike voortvloeiing van vroeëre verbeteringe van die herwinning van goud in gravitasie-afdelings van bestaande goudaanleggings. Daarbenewens was besorgdheid oor omgewingsbesoedeling deur die konvensionele amalgameringsproses verdere motivering vir die ontwikkeling van 'n alternatiewe prosedure.

Dit was bevind dat die parsiële druk van suurstof in oplossing die mees kritiese reaksietempobepalende faktor was. Onder ongeveer 25 °C was die tempo waarteen goud opgelos het reaksietempobeheerd, terwyl dit diffusiebeheerd was bokant hierdie temperatuur. Die doeltreffendheid van pulpberoering en suurstof parsiële druk het dus 'n hoë graad van beheer oor die oplossingstempo van goud moontlik gemaak. Die gebruik van suurstof, eerder as lug, het die oplossingstempo van beide goud en silwer aansienlik verhoog. Die oplosbaarheid van natrium- kalsium- en kaliumaurosianied-komplekse was onder gesimuleerde reaksieomstandighede bestudeer en dit was vasgestel dat die oplosbaarheidswaardes vir natrium- en kalsiumkomplekse beperk was en dat die gebruik van laer pulp digthede en 'n hoër reaksie temperatuur nodig was vir die behandeling van baie hoëgraad-konsentraate. Dit was bevind dat die teenwoordigheid van oormatige hoeveelhede dwaalyster (tot 25 persent volgens massa) 'n swak oplossing van goud tot gevolg het, omdat dit onder sekere omstandighede die sementasie van goud veroorsaak het. Die handhawing van hoë suurstof parsiële druk, tesame met hoë sianiedkonsentrasies, hoë pH-waardes en die teenwoordigheid van kalsium in die sisteem het alles bygedra tot die passivering van dwaalsyter. Koper, goud en yster was aangedui as die hoof sianied-verbruikende elemente.

Optimum oplossing van goud was verkry in 'n geslote bankreaktor met 'n flottasie masjien-tipe roerder vir effektiewe pulpberoering en verbinding met suurstof. Suurstof was deur die reaktor hersirkuleer. Die reaktor was ontwerp om onder 'n baie lae positiewe druk te werk. Na reaksietye van 2-6 uur teen temperature in die omgewing van 30 °C, het gemiddeld 99,5 persent van beide die goud en silwer opgelos. In die geval van baie hoëgraad-konsentrate was dit aangetoon dat flottasie van goud in dieselfde reaktor gedoen kan word deur die geflotteerde konsentraat deur 'n opening in die reaktor af te trek. Die konsentraat wat 3,9 persent van die massa was, het 87,4 persent van die goud bevat.

Die herwinning van osmiridium uit sianiedloogoorblyfsels en die eletrolitiese herwinning van goud uit dragtige loogoplossing was ook na gekyk as prosesaspekte in die behandling van goudaanlegbandkonsentrate. Finale tafeluitskotte van 10 tot 12 g/t goud was verkry.

Die resultate van die ondersoek dui sterk daarop dat indien effektiewe beroering en oksigenering in die sisteem ingesluit kan word, 'n alternatiewe sianiedproses vir die behandeling van goudaanlegkonsentrate tegnies moontlik is. In die geval van laegraad-konsentrate blyk sianidering gevolg deur elektrolitiese herwinning van goud die eenvoudigste proses te wees terwyl in die geval van hoëgraad-konsentrate, flottasie en smelting van die konsentraat, gevolg deur sianidering van die flottasie-uitskotte aanbeveel word. Die gebruik om suurstof, eerder as lug, deur die pulp to blaas sal grootliks afhang van die graad van die materiaal wat behandel moet word. 'n Voorlopige vergelykende studie van die kostes verbonde aan die amalgamering en sianideringsprosesse vir die herwinning van goud uit sulke konsentrate het groot ooreenkoms in die kapitaal en lopende kostes aangedui. Die gevolgtrekking was gemaak dat faktore soos giftigheid, veiligheid en arbeid benodig, belangriker is as die klein kosteverskille. Verder sal die onderbreking van 'n goudaanleggravitasiebaan, moontlik gemaak deur die daarstelling van 'n sianied-proses, die lopende koste verder verlaag deur die vermindering van die massa wat behandel moet word.

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Introduction

When mining operations first commenced on the Witwatersrand, the method of extracting gold from the ore was by stamp-milling and amalgamation[1]. With the introduction of cyanidation in 1890, the amalgamation of stamp-mill pulp was still retained so that as much gold as possible could be recovered at an early stage. With the later development of the closed gravity circuit, classified concentrates were again treated by amalgamation, and this practice is in use at the present time. The process involves the batch treatment of high-grade riffle-belt concentrates with mercury in amalgamation barrels. The barrel is rotated for a period ranging between 14 and 18 hours in order to liberate and polish the gold particles. Suitable quantities of lime, caustic soda, cyanide, and detergent are usually added to assist in this cleaning and polishing of the free gold. An adequate amount of mercury is then added to the barrel, and the rotation is continued for several hours during which time the mercury and gold form a soft amalgam. After amalgamation, the barrel is emptied over a single amalgam plate and the soft amalgam is then scraped by hand from the plate surface. Alternatively, the barrel contents are discharged through a hydrocyclone, and the amalgam is recovered from the spigot discharge. After being cleaned, the amalgam is filtered and retorted. The retorting of the pressed amalgam is carried out in an electric furnace at approximately 600 °C, the condensed mercury being collected in buckets and recycled, while the sponge gold is smelted and cast into bars. Mercury losses during retorting should not normally exceed 0.02 g per ton of ore milled. The recovery of gold from gravity concentrates using amalgamation is about 96 to 97 per cent, although the overall recovery may be even higher because the amalgam tailings are returned to the head of the gravity circuit. The process may involve a considerable health hazard related to mercury ingestion, and is at the same time very labour-intensive.

As a result of earlier improvements to the recovery of gold in the gravity sections of existing gold plants[2] and the consequent need to treat larger tonnages of concentrates, an alternative processing route for gold-plant gravity concentrates was indicated. Concern about the pollution and health aspects of the amalgamation process was further motivation in the development of an alternative procedure. The advantages sought from such a route included low-risk working conditions, together with the elimination of the large recycling load of concentrates in the gold-plant gravity circuit and perhaps a less labour-intensive operation. A gold recovery equal to or better than the amalgamation route was also a prime requirement of such an alternative process. Earlier work carried out in these laboratories on the recovery of gold from both belt-concentrator tailings and amalgamation tailings had indicated that the cyanidation of high-grade gravity concentrates merited further investigation. In view of the varied and relatively long history of the gold-mining industry in South Africa, it may have been expected that the cyanidation of high-grade concentrates was well documented in the literature. However, a literature survey carried out in early 1974 failed to indicated any relevant information in this respect. The only useful information was derived from numerous publications dealing with investigations on the dissolution of pure gold, but the extent to which such data could be applied to gold concentrates had yet to be established. Concurrently with the literature survey, work started on the on-site sampling of gold-plant belt concentrates and preliminary cyanidation tests.

The preliminary cyanidation testwork, which was carried out in stirred beakers, was most significant in that a very wide range of gold extractions, indicating very poor reproducibility, was obtained. This poor reproducibility suggested that some hitherto unforeseen experimental variable (or variables) was playing a critical role in the gold dissolution. Further investigations directed towards a better understanding of such reaction parameters were thus most necessary.

Composition of Belt Concentrates

Weekly composite samples of belt concentrates from Welkom Gold Mine, designated W1 to W6, were washed in acetone and dried at 110 °C prior to blending and splitting. The detailed analysis of four of these samples is presented in Table I, from which it can be seen that the gold content varied from 6 to 22 kg/t, reflecting the extreme difficulty in obtaining representative samples of such high-grade materials.

The average U_3O_8 content can be seen to be approximately 23 kg/t, while the average platinum content of 44 g/t indicates a value for the total platinum-group metals (PGM) according to Cousins[3] of approximately 277 g/t. The presence of such high concentrations of uranium and PGM was indicative of the recycling of a high percentage of material in the gold-plant gravity circuit. The relatively large amounts of metallic tramp iron present (19 per cent) should be noted, because such material, if not effectively passivated, will precipitate the gold from the cyanide solution. The presence of copper (0,46 per cent), nickel (0,13 per cent), cobalt (0,10 per cent), and arsenic (0.29 per cent) should also be noted as possible major cyanide-consuming elements. The samples studied had a very similar composition with regard to their pyrite (35 per cent) and silica (33 per cent) contents.

Table II presents the screen analysis of samples W1 to W4, showing that, on average, 7,2 per cent by mass of the material is coarser than 295 μm while containing only 4.4 per cent of the total gold. On average, 6,1 per cent of the material is below 43 μm, but it contains as much as 26,1 per cent of the total gold content. Furthermore, on average 85 per cent of the particulate gold present is smaller than 147 μm.

Because of the presence of considerable concentrations of PGM in such concentrates, the analyses for residual gold, silver, and platinum were carried out by acid digestion of the samples followed by atomic-absorption analysis.

Cyanidation of Gravity Concentrates

Consumption of Cyanide

Because of the excessive cyanide consumptions observed in the treatment of gravity concentrates, the

TABLE I

THE CHEMICAL COMPOSITION OF FOUR COMPOSITE SAMPLES OF GOLD-PLANT BELT CONCENTRATES

Sample	Semi-quantitative spectrographic analysis (g/t)																						
	Zr	Mn	Ba	Pb	Zn	Sb	Sn	Li	Be	Mo	Nb	Cu	Ni	La	Ag	Co	As	Bi	V	Y	Sr	Na	Ca	Ti
W1	4000	4000	150	5000	80	< 50	80	< 100	< 0,1	5	< 100	1500	600	< 100	100	400	400	3	50	100	< 50	150	200	200
W2	4000	3000	100	8000	100	< 50	70	< 100	< 0,1	8	< 100	3000	1000	< 100	400	800	1000	6	100	200	< 50	150	200	200
W3	6000	4000	100	3000	60	< 50	50	< 100	< 0,1	8	< 100	1500	700	< 100	300	400	400	3	70	100	< 50	150	200	200
W4	1500	3500	200	6000	250	60	100	< 100	< 0,1	5	< 100	900	3000	100	250	400	5000	8	15	400	< 10	100	2000	200

	Semi-quantitative spectrographic analysis (g/t)								X-ray fluoros. (kg/t)			Chemical analysis (%)								Fire assay (g/t)			
Sample	Al	Ge	Fe	Si	Cr	Mg	B	W	U_3O_8	ThO_2	Rb	SiO_2	Fe	Fe°	Co	Cu	Ni	S	As	Au	Ag	Pt	Pd
W1	> 3000	< 10	> 3000	> 3000	2000	2500	< 10	—	18,3	—	—	46,9	41,8	21,9	0,08	0,48	0,10	13,0	0,24	6300	671	40	2
W2	> 3000	< 10	> 3000	> 3000	3000	2000	< 10	—	20,0	1,4	0,59	31,3	39,3	15,4	0,15	0,65	0,17	22,5	0,43	22000	2500	39	4
W3	> 3000	< 10	> 3000	> 3000	2500	1000	< 10	—	32,0	2,3	1,21	34,4	33,0	13,4	0,08	0,28	0,10	18,4	0,19	12200	1200	44	2
W4	> 3000	10	> 3000	> 3000	2000	500	< 10	1000	16,8	1,2	0,82	22,3	46,4	25,0	0,08	0,43	0,13	21,8	0,31	19400	2200	54	—

TABLE II

PARTICLE-SIZE AND GOLD-DISTRIBUTION ANALYSES OF FOUR COMPOSITE SAMPLES OF GOLD-PLANT BELT CONCENTRATES

Size fraction (μm)	Sample W1			Sample W2			Sample W3			Sample W4		
	Mass distr. (%)	TPM* (kg/t)	% of total TPM	Mass distr. (%)	TPM* (kg/t)	% of total TPM	Mass distr. (%)	TPM* (kg/t)	% of total TPM	Mass distr. (%)	TPM* (kg/t)	% of total TPM
+295	5,8	2,4	1,9	5,9	11,8	3,0	5,9	7,3	3,2	11,5	15,6	9,3
+208—295	10,5	2,3	3,2	7,1	10,7	3,2	7,1	7,3	3,9	9,1	11,8	5,5
+147—208	22,3	2,7	8,0	14,0	8,4	5,1	15,3	5,7	6,6	16,2	11,0	9,2
+104—147	22,2	4,5	13,2	16,9	5,4	4,0	20,2	5,7	8,6	24,4	12,0	15,1
+74—104	21,6	7,5	21,6	23,7	15,7	16,1	25,6	11,2	21,5	17,8	18,0	16,5
+53—74	8,5	12,9	14,6	15,6	28,7	19,4	13,6	17,5	17,9	11,8	28,6	17,4
+43—53	4,9	23,3	14,9	8,2	39,9	14,2	5,7	23,3	10,0	4,2	38,7	8,4
—43	4,2	40,4	22,6	8,6	93,9	35,0	6,6	57,1	28,3	5,0	72,2	18,6

*Total precious metals

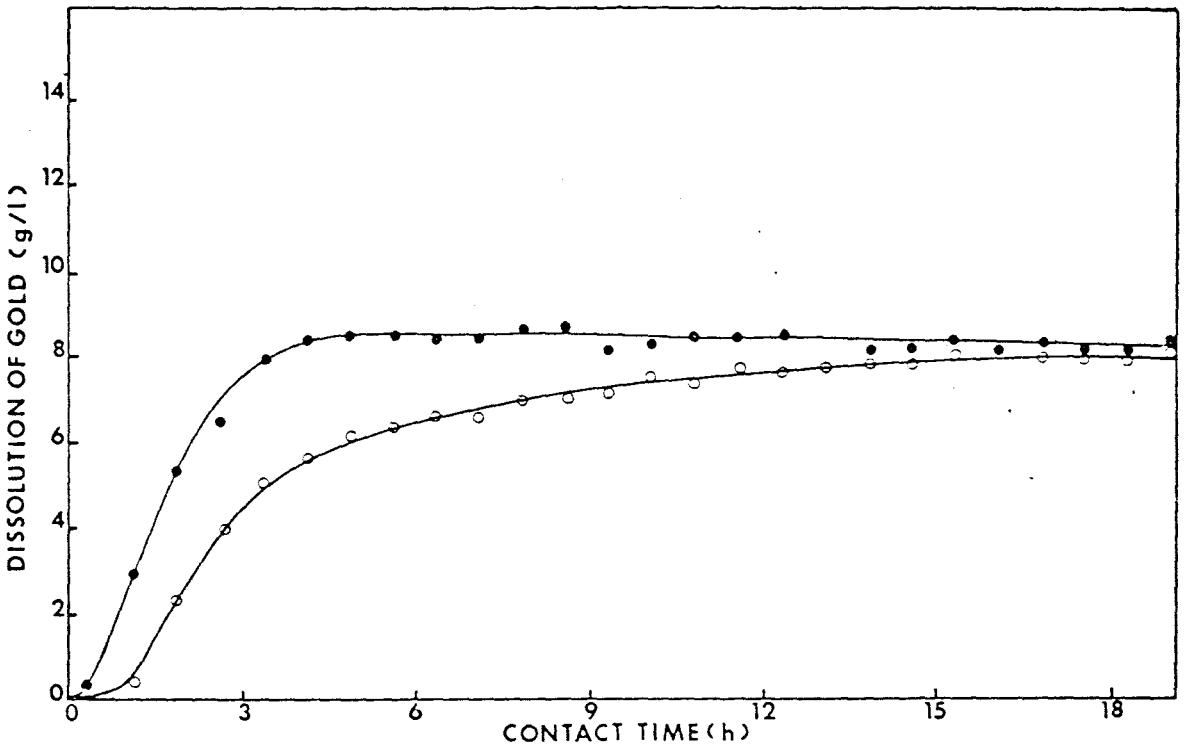

Fig. 1—Dissolution of gold from sample W1
● Oxygen flow-rate 180 cm³/min ○ Air flow-rate 180 cm³/min NaCN addition 25 kg/t pH value 12,4 to 12,7 Liquid: solid 1:1

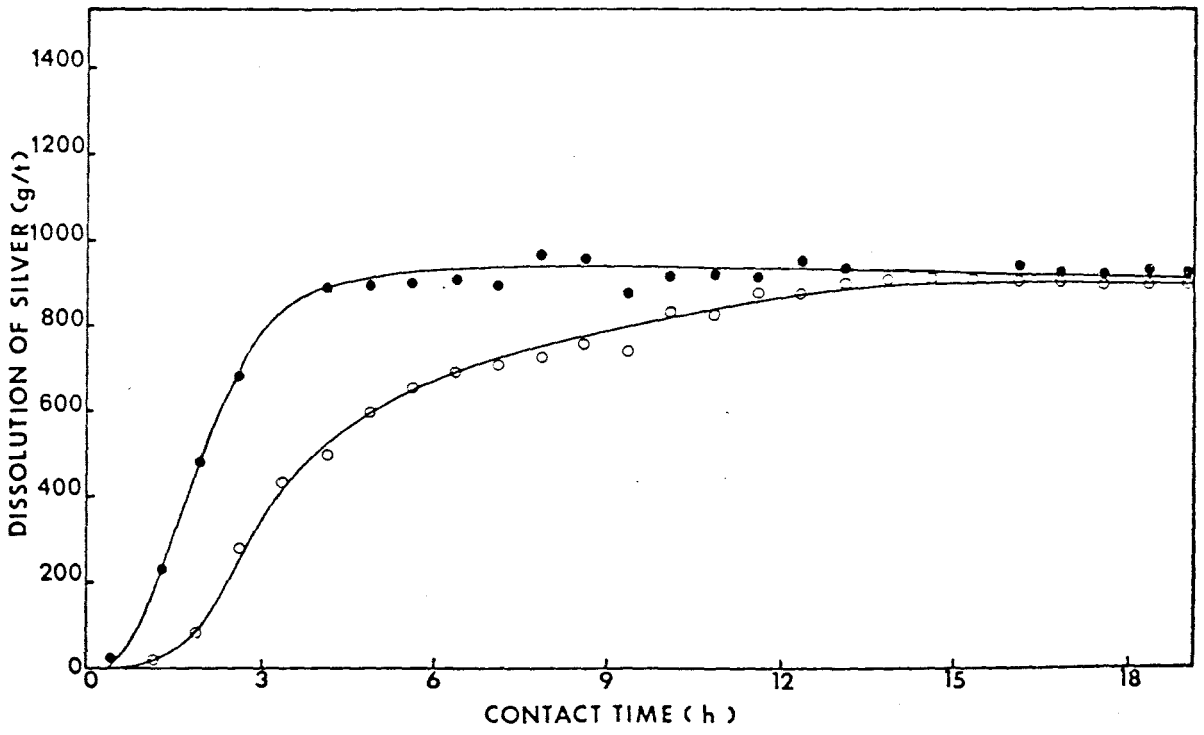

Fig. 2—Dissolution of silver from sample W1
● Oxygen flow-rate 180 cm³/min ○ Air flow-rate 180 cm³/min NaCN addition 25 kg/t pH value 12,4 to 12,7 Liquid: solid 1:1

TABLE III

CYANIDATION OF SAMPLE W1 GOLD-PLANT BELT CONCENTRATE AT ≈25 °C

Test no.	Residue		Extraction		Sodium cyanide		Gas flow-rate		Conditions	Final $\frac{Fe}{Cu}$ in soln (g/t)	Initial pH	Final pH
	Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	Added (kg/t)	Consumed (kg/t)	Oxygen (cm^3/min)	Air (cm^3/min)				
1 M	22,6	4,6								650		
1 NM	17,7	4,8			25,0	20,0	180	—	50 g/t Pb* added	———	12,4	12,7
1 R	29,0	5,0	99,6	99,4						5000		
2 M	227	14,1										
2 NM	589	23,2			25,0	19,5	—	180	2,5 kg/t CaO added	—	—	—
2 R	320	10,0	95,0	99,3								
3 M	105	12,0								360		
3 NM	285	21,2			25,0	19,5	—	180	2,5 kg/t CaO added	———	12,5	12,7
3 R	140	20,3	97,5	98,1						4900		
4 M	33,5	4,2								1560		
4 NM	21,1	6,2			25,0	23,1	180	—	50 g/t Pb* added	———	12,4	12,6
4 R	38,2	6,2	99,7	99,0						4900		
5 M	492	703								1450		
5 NM	32,5	28,0			25,0	24,9	180	—	H_2SO_4 added	———	11,4	10,3
5 R	339	750	93,1	20						4100		
6 M	370	84,1								2300		
6 NM	1123	167			25,0	25,1	20	—	Reduced flow-rate (50 g/t Pb*)	———	12,3	12,6
6 R	729	148	87	80						3600		
7 M	32,3	4,6							Increased cyanide addition	2500		
7 NM	19,1	7,1			35,0	31,2	20	—	(50 g/t Pb*)	———	12,3	12,9
7 R	24,4	5,4	99,7	98,2						4900		
8 M	392	43,1								570		
8 NM	654	82,1			35,0	25,0	20	—	Air pretreat for 12 h (50 g/t Pb*)	———	12,1	12,4
8 R	554	64,2	92,1	82						4100		
9 M	33,0	2,6								1700		
9 NM	39,2	2,3			25,0	24,3	180	—	No Pb added	———	12,0	12,5
9 R	33,1	2,1	99,5	99,5						4700		

M Magnetic fraction residue
NM Non-magnetic fraction residue
R Composite residue
* Lead added as $Pb(NO_3)_2$

562.

dissolution of gold and silver together with iron, copper, cobalt, nickel, and arsenic was investigated as a function of time so that the major cyanicides could be defined. The use of oxygen rather than air to aid the gold dissolution was also briefly considered.

Concentrate samples of 2 kg each were leached with 2 litres of cyanide solution in a covered 5-litre beaker. The pulp was well stirred by a large paddle-type stirrer, while the pulp was sparged at a constant flow-rate of gas through a sintered-glass gas-distribution thimble. The reaction was carried out at ambient temperature, and, unless otherwise specified, the leaching times were confined to 20 hours. A micropump was employed to bleed the leach system continuously of filtered solution at a rate of 12,5 ml/h, and the solution was collected in a fraction collector for atomic-absorption analysis. After cyanidation the pulp was filtered, the filtrate being titrated for residual cyanide and the washed residue assayed for gold and silver. In certain instances, the leach residues were further separated into magnetic and non-magnetic fractions prior to analysis. Deionized water was used throughout the investigation.

Figs. 1 and 2 illustrate the dissolution of gold and silver respectively from sample W1 (gold 6,3 kg/t), either oxygen or air being used to oxygenate the pulp. It can be clearly seen that oxygen has a markedly beneficial effect on the kinetics of both gold and silver leaching. After 20 hours, an extraction of 99,6 per cent of the gold and 99,4 per cent of the silver using oxygen can be compared with an extraction of 97,5 per cent of the gold and 98 per cent of the silver using air (Table III).

In a further test run in which the pH value was maintained at approximately 10,3 by the use of sulphuric acid (Test No. 5, Table III), a gold extraction of 93,2 per cent and a silver extraction of only 20 per cent were obtained. Such poor dissolution, together with the very high gold and silver contents of the magnetic fraction of the residue and the high soluble iron but low copper content of the leach liquor, strongly suggested the cementation of gold, silver, and even copper onto the metallic tramp iron present. This conclusion is supported by earlier findings in which iron powder was added to the leach pulp prior to cyanidation at pH 10,5. The deleterious effect of the iron powder was most marked when air rather than oxygen was used to oxygenate the pulp, indicating that the oxidizing potential of the system plays a vital role in such a mechanism. The reduction in the oxygen flow-rate from 180 cm^3/min to 20 cm^3/min (Test No. 6, Table III) further substantiates the above suggested mechanism, where again a poor gold extraction (87 per cent) was obtained together with abnormally high amounts of soluble iron in solution. Such poor recoveries can, however, be improved by the use of higher cyanide additions (Test No. 7, Table III) but result in an excessive consumption of cyanide. Alternatively, prolonged pre-aeration of the pulp (Test No. 8, Table III), producing an excessive iron hydroxide precipitate, resulted in a marked decrease in soluble iron. The resulting poor gold recovery (92 per cent) is not easily explained, but excessive adsorption of gold and silver by such a precipitate (very high positive zeta potential) is certainly suggested. The addition of lead in the form of lead nitrate prior to cyanidation appeared to have little or no effect on the gold dissolution under these specified conditions (compare Test Nos. 1 and 9, Table III). Likewise, small additions of lime had no noticeable effect on the gold dissolution (see later).

An effort was made to approximate the contributions to the total cyanide consumption made by gold, silver, copper, and iron, neglecting the small contribution made by cobalt, nickel, and arsenic as well as that lost by oxidation. The following co-ordination numbers were assumed for the corresponding cyanide complexes: 6 (iron), 3,6 (copper), 2 (gold), and 2 (silver). The solution concentration of the respective elements at the end of each 20 hour leach (Table III) were then used in the calculation of the cyanide consumption. In most instances, the calculated consumption agreed fairly well with the observed consumption. Thus, for the best gold extraction obtained (Test No. 7, Table III), it was shown that 63 per cent of the cyanide was consumed by copper, 19 per cent by gold, 15 per cent by iron, and 3 per cent by silver. These contributions can be compared with a very poor gold extraction (Test No. 6, Table III), where only 41 per cent of the cyanide was consumed by copper, 12 per cent by gold, 2 per cent by silver, and as much as 45 per cent by iron. This excessive cyanide consumption by iron is in many instances associated with poor gold and silver extractions.

The leaching of a second sample of belt concentrate designated W2 (gold 22 kg/t) using variable oxygen flow-rates and cyanide additions (Test Nos. 1-3, Table IV) yielded very poor gold and silver extractions. Such poor extractions, together with slower gold dissolution rates and the observation that there was sometimes a white crystalline precipitate in the pregnant liquor, suggested the saturation of the liquor with some gold complex. The subsequent analysis of such a dried precipitate indicated the formula $NaAu(CN)_2$. It is interesting to note that, in those tests in which excessive additions of sodium cyanide were not made (Test No. 1, Table IV), the saturation of the pregnant liquor occurred at a gold concentration of approximately 13 g/l. However, when the addition of cyanide was excessive (Test No. 2, Table IV), only about 2 g/l gold was found in solution. This suggests that the solubility of the sodium aurocyanide complex decreases considerably with increasing cyanide concentration owing to a common ion effect. Such an effect may well be significant at the start of a reaction, when cyanide concentrations are relatively very high.

In a demonstration that the above solubility effect slows down and even halts the dissolution of gold and silver, leaches were conducted with equimolar additions of potassium cyanide under similar conditions. (From the available information it appears that potassium aurocyanide is approximately 6 to 7 times more soluble than sodium aurocyanide.) It was immediately evident (Test No. 5, Table IV) that, under the prescribed experimental conditions, i.e., a liquid:solid ratio of 1:1 and 25°C, the solubility of the gold cyanide complex plays a most significant role in the leaching of such rich concentrates. A gold recovery of 99,9 per cent, together with an iron

TABLE IV

CYANIDATION OF SAMPLE W2 GOLD-PLANT BELT CONCENTRATE AT ≈25°C

Test no.	Residue		Extraction		Cyanide		Oxygen flow-rate	Conditions	Fe Final — in Cu soln (g/t)	Initial pH	Final pH
	Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	Added (kg/t)	Consumed (kg/t)	(cm^3/min)				
1 M	3620	585							2000		
1 NM	10800	1022			30,0	30,0	20	Pretreat with O_2 for 12 h	——	12,1	12,7
1 R	8400	928	63	64	(NaCN)				3000		
2 M	4500	966							2300		
2 NM	12900	303			70,0	42,2	180	NaCN added to excess	——	12,0	12,6
2 R	10844	2375	53	60	(NaCN)				4400		
3 M*	3225	320					500	Initially large O_2 flow reduced to	1000		
3 NM	8132	866			35,0		Reduced to	180 cm^3/min after 0,75 h	——	12,0	12,6
3 R	5700	600	75	75	(NaCN)	30,1	180		3900		
4 M*	2425	203							940		
4 NM	7200	697			35,0	26,3	180	25 l/t of H_2O_2 added	——	12,0	12,6
4 R	6340	629	73	75	(NaCN)				5900		
5 M	20,3	5,2						Equimolar quantity of KCN to compare	1080		
5 NM	16,4	3,1			47,0	42,5	180	with 35 kg/t NaCN	——	12,6	13,2
5 R	15,1	2,0	99,9	99,9	(KCN)				4400		
6 M	1000	110							240		
6 NM	1300	150			35,0	34,9	180	25 l/t of H_2O_2 added	——	12,2	12,9
6 R	1600	180	92,7	92,5	(KCN)				4500		
7 M	350	35						KCN added in batches;	100		
7 NM	1500	200			35,0	34,0	180	25 l/t of H_2O_2 added	——	12,5	13,1
7 R	650	40	97,0	98,3	(KCN)				3300		

* Reaction time of 6 h
M Magnetic fraction residue
NM Non-magnetic fraction residue
R Composite residue

564.

concentration of only 1080 g/t in the final leach liquor, should be noted. The dissolution of gold, silver, copper, and iron from sample W2 using potassium cyanide while sparging with oxygen is illustrated in Fig. 3. The addition of small quantities of hydrogen peroxide (Test Nos. 6-7, Table IV) can be seen to have little effect on the gold dissolution, but significantly decreases the iron dissolution.

Solubility of Sodium and Potassium Aurocyanide

During the course of the present investigation on the dissolution of gold from high-grade belt concentrates, precipitates of aurocyanides frequently occurred. Such precipitates invariably coincided with poor gold extraction, which necessitated an investigation into the factors affecting the solubility of aurocyanide. Literature data on the subject revealed very little information, and only the solubility of potassium aurocyanide at 25 °C could be derived (from the works of Basset and Corbet[1]).

Both potassium and sodium aurocyanide were prepared by evaporation of pregnant liquors, these being obtained by the leaching of gold concentrates with sodium or potassium cyanide respectively. The condensed solutions were chilled, and the precipitated aurocyanide salts were filtered off, washed with ice water, and dried at 110 °C. Analysis of the salts showed that small quantities of alkalies and silver, and traces of iron, were still present. It was assumed that such small quantities of impurities would have no significant bearing on the solubility data.

A thermostatically controlled water-bath connected to a refrigeration unit was used for temperature equilibration. Small Erlenmeyer flasks containing water or leach liquor were placed in the water-bath and saturated somewhat above the required temperature with the respective aurocyanide. The required temperature was allowed to stabilize for at least 6 hours while the flasks were shaken frequently. Then 10 ml quantities of saturated solution were pipetted into 100 ml weighed volumetric flasks, the pipette always being heated to above the test temperature to prevent crystallization of the aurocyanide during pipetting. After the sample had been weighed, the flasks were filled with distilled water and analysed for gold. The sodium and potassium concentrations were determined by atomic-absorption spectrophotometry. The error introduced by pipetting at temperatures other than the standard temperature of the pipette was found to be less than 1 per cent.

The effect of temperature on the solubilities of potassium and sodium aurocyanide in solutions of different cation concentrations is shown in Fig. 4, while Fig. 5 shows the dependence of the sodium aurocyanide solubility on the total concentration of sodium ions.

The two diagrams show that both the temperature change (Fig. 4) and the total cation concentration (Fig. 5) have a considerable effect on the solubility of the aurocyanide. Although the solubilities of both potassium and sodium aurocyanide increase with increasing temperature, the rate of increase is significantly greater for the potassium salt. The detrimental effect of the increase in cation concentration is probably related to both the salt effect and the common ion effect; there may even be a significant influence of anions such as cyanide or hydroxyl ion. Moreover, it has been observed that the aurocyanide concentration in saturated solutions is also affected by the total salt concentration. However, the influence of these parameters was not

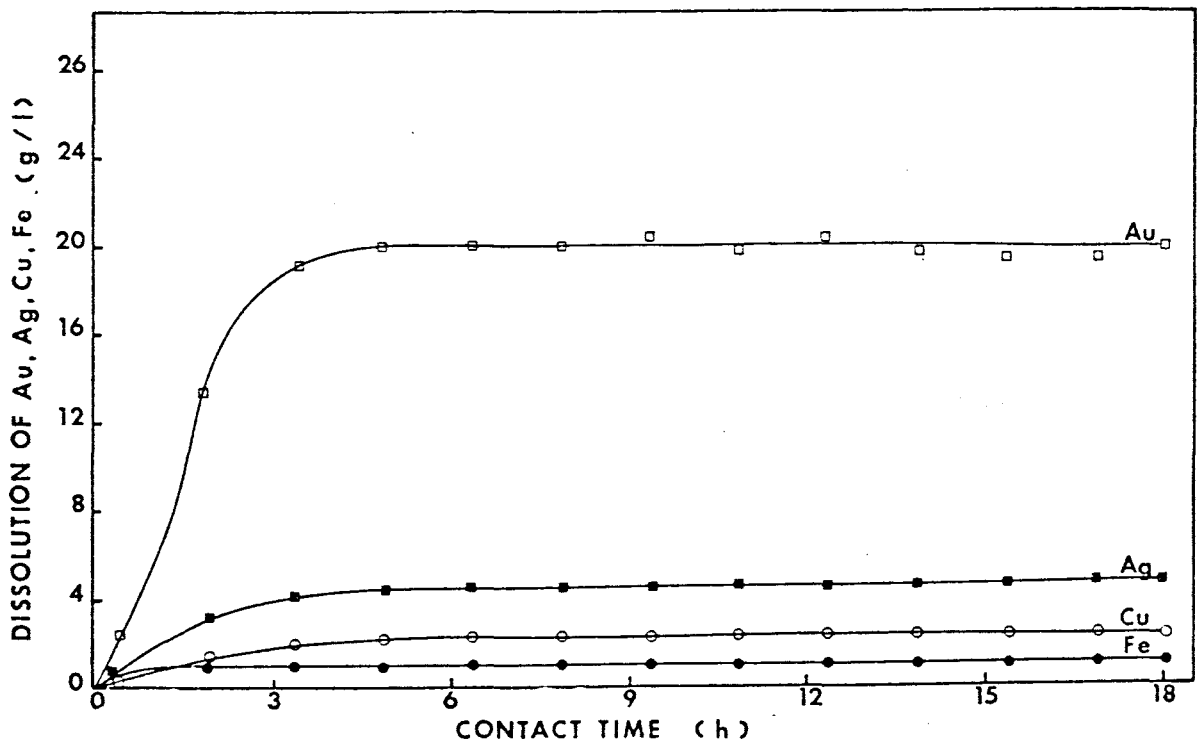

Fig. 3—Dissolution of gold, silver, copper, and iron from sample W2
KCN addition 47 kg/t Oxygen flow-rate 180 cm³/min Liquid: solid 1:1 pH value 12,6 to 13,2

investigated because of the extent and complexity of the work involved.

The increase in pH accompanying the increase in cation concentration (Fig. 5) is probably not due to a general dependence of aurocyanide solubility on pH, but rather to the manner of gold dissolution in cyanide solutions according to the equation

$$Au+2CN^-+\tfrac{1}{2}H_2O+\tfrac{1}{4}O_2=Au(CN)_2^-+OH^-, \quad . . \quad (1)$$

One mole of hydroxyl ion is produced for each 2 moles of cyanide ions consumed. Similarly, as mentioned before, hydroxyl ions are produced during the dissolution of iron and copper by cyanide.

It is perhaps surprising that saturation of the leach liquor should limit the gold dissolution during the leaching of concentrates. From available thermodynamic data, reaction (1) has a highly negative free energy ($\Delta G°_{298}=-24$ kcal/mol), even for saturated aurocyanide solutions. Thus, precipitation of aurocyanide during gold dissolution should act as a gold sink preventing attainment of equilibrium, rather than a dissolution inhibitor. Probably, the formation of a diffusion barrier

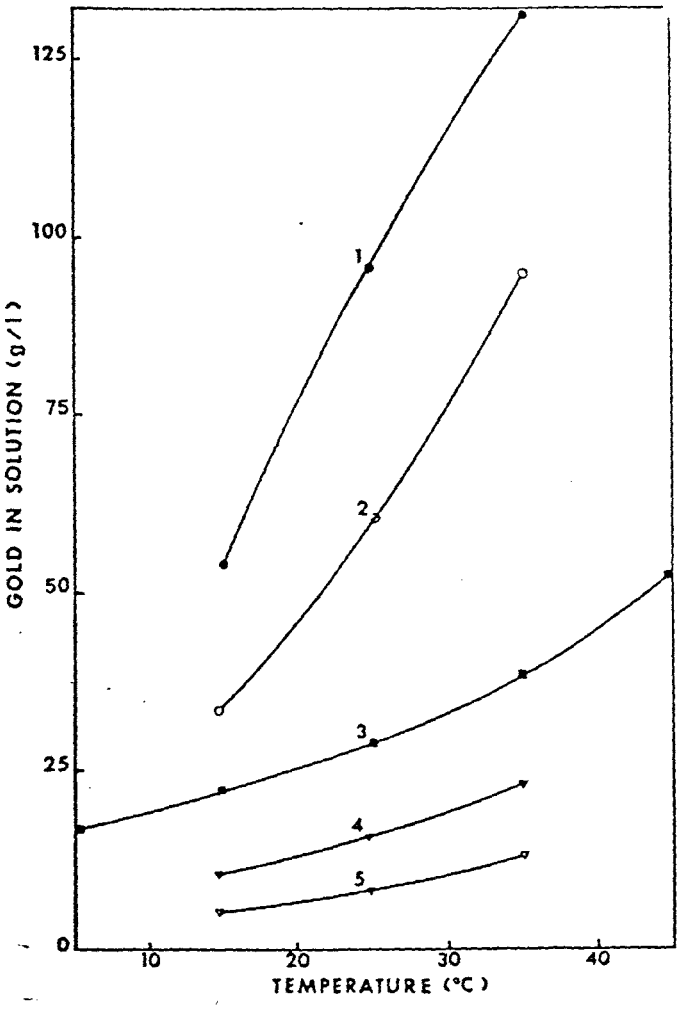

Fig. 4—Effect of temperature on the solubility of sodium and potassium aurocyanide

- ● $K[Au(CN)_2]$—K^+ increasing from 49 to 76 g/l
- ○ $K[Au(CN)_2]$—K^+ increasing from 64 to 86 g/l
- ■ $Na[Au(CN)_2]$—Na^+ increasing from 2,3 to 6,7 g/l
- ▼ $Na[Au(CN)_2]$—Na^+ increasing from 6,5 to 8,0 g/l
- ▽ $Na[Au(CN)_2]$—Na^+ increasing from 12,4 to 14,4 g/l

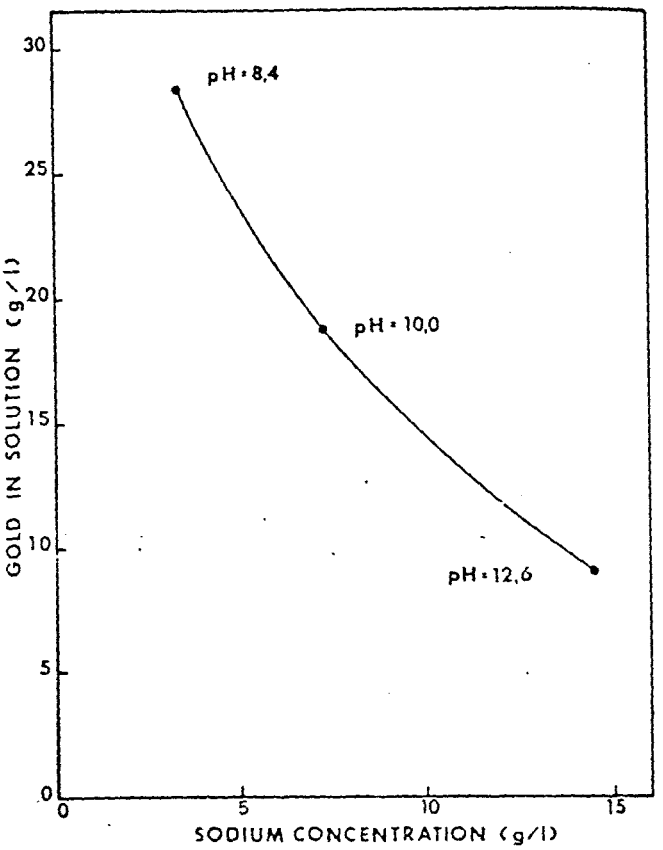

Fig. 5—Effect of sodium concentration on the solubility of aurocyanide at 25 °C

by precipitated aurocyanide on the reaction surface (passivation) is responsible.

Because of the cessation of gold dissolution as soon as the precipitation of aurocyanide occurs, the solubility of the gold complex is of major importance in the cyanidation of concentrates. Consequently, with very rich concentrates (20 kg/t gold) the liquid:solid ratio becomes very critical. The liquid:solid ratio in turn has a pronounced effect on the size of leaching equipment required. For example, if a concentrate containing 20 kg/t is being leached with 25 kg/t sodium cyanide, the sodium concentration will be 11,7 g/l at a liquid:solid ratio of 1:1 and 5,9 g/l at a ratio of 2:1. If all the gold dissolved, a gold concentration of 20 g/l for the lower liquid:solid ratio and 10 g/l for the higher would be obtained. From curve 5 in Fig. 4 it is obvious that just over half the gold can dissolve at the low liquid:solid ratio even at 35 °C, whereas a temperature in excess of 30 °C is necessary to achieve complete gold dissolution at the higher ratio.

A liquid:solid ratio of approximately 3:1 would be required if leaching were carried out at room temperature. On the other hand, if potassium cyanide were used at room temperature, the liquid:solid ratio could be reduced to as little as 1:2. Although solubility data are generally represented as grams of solids in 100 grams of solution, the present representation of grams per litre was adopted for reasons of simplicity. Values can be converted to grams of solids per 100 grams of solution by the assumption of an increase in specific gravity from 1,02 to 1,03 g/cm³ for sodium aurocyanide solutions and

1,11 to 1,19 g/cm³ for potassium aurocyanide solutions in the temperature range 15 to 35°C.

It is clear from the results presented that high-grade gold concentrates pose certain practical limitations on the choice of leaching conditions when sodium cyanide is used as a leaching agent. Both higher reaction temperatures (30 to 40°C) and higher liquid:solid ratios (>2:1) are indicated if precipitation of the gold is to be avoided. While the use of potassium cyanide as an alternative leaching agent may well allow a far greater latitude in the choice of such reaction parameters, the additional cost and availability of potassium cyanide may militate against its use. The present programme was confined to the solubility of potassium and sodium aurocyanides, illustrating the important role played by solubility in the intensive cyanidation of gold concentrates. Such findings immediately suggest that the presence of calcium or magnesium may likewise play a very important role in the cyanidation of gold concentrates, and further investigation in this direction appears to be necessary (see later).

The Effect of Oxygenation and Agitation

The earlier investigations had indicated that pulp aeration and agitation may play a very vital role in effecting efficient gold and silver dissolution from such high-grade material. It was thus necessary to investigate such variables using several means of mechanical agitation. The use of paddle agitation, a Super-agitator (Denver Equipment nomenclature), and a flotation machine was considered in the present investigation, which was confined to the treatment of sample W3 (gold 12,2 kg/t and silver 1200 g/t) at a liquid:solid ratio of 2:1 unless otherwise specified.

For paddle agitation, a series of experiments was designed so that variable flow-rates of oxygen and air, together with the efficiency of agitation at various reaction times, could be investigated. Cyanidation was carried out in a loosely covered 2-litre glass beaker maintained at 25°C in a water-bath in which the pulp was agitated by means of a paddle. Oxygen or air was added in controlled quantities through a sintered-glass gas distributor located near the bottom of the vessel. Two different impeller speeds and two different paddle sizes were investigated.

From Table V it can be seen that the kinetics of gold and silver dissolution are again faster with oxygen than with air, but that, at longer reaction times, the residual values obtained with air are similar to those obtained with oxygen (compare Test No. 7 with Test No. 15). Furthermore, the lower stirring speed (400 r/min) gave consistently poor gold extractions as compared with the higher speed (800 r/min). The effect of paddle size, although significant, is less obvious. On the other hand, the amount of oxygen added, either in the form of air or as pure oxygen, had no significant effect on gold extractions down to a critical level, which in the prescribed experimental conditions was approximately 10 cm³/min (see Test Nos. 11 and 12). Iron dissolution can be seen to be higher when the final cyanide concentration was low (Table V, Test Nos. 14, 14B, 16, 16B, 17, 19, 21, and 23), the lower sodium cyanide limit appearing to be around 3 g/l. This appeared to indicate that cementation may occur at low concentrations of residual cyanide. Table V also shows that the cyanide consumption is not significantly affected by either air or oxygen. Longer reaction times increase the cyanide consumption marginally.

As the above results again emphasize the importance of efficient agitation and aeration, a similar exercise was carried out with a small Super-agitator type of system. The Super-agitator vessel was loosely covered, and air or oxygen was added at a controlled rate by means of a sintered-glass distributor located near the bottom of the vessel. The impeller speed was controlled at 1800 r/min, and the temperature was maintained at 25°C in all cases.

As Table VI shows, the kinetics of gold and silver dissolution are faster with oxygen than with air, but, given sufficient time, air produced residual values similar to that of oxygen. An increase in the leaching time from 6 to 16 hours lowered the residual values marginally when oxygen was used but dramatically when air was used. An increase in the leaching time from 16 to 24 hours using air decreased the residual values only marginally. Residual values tended towards a minimum, possibly determined by the amount of occluded gold present in the concentrate sample. With regard to the dissolution of iron and copper, it can be seen that the iron dissolution increases with time and is lower with oxygen than with air for shorter reaction times (6 hours). As before, when the final cyanide concentration was reduced, the iron dissolution was found to be higher, which again indicated that cementation occurs at lower cyanide concentrations, confirming previous observations. The copper dissolution, on the other hand, appeared to be quite unaffected by such changes in reaction variables.

As the above results again emphasized the importance of efficient agitation and oxygenation, a third mechanism, viz a conventional flotation machine, was also investigated. The leaching vessel was completely sealed, apart from a small vent hole. Air or oxygen was fed conventionally at a controlled rate through the agitator shaft of the flotation machine. The impeller speed was 1800 r/min and the reaction time 6 hours in all cases. The temperature was controlled by a water jacket. As shown in Table VII, the kinetics of gold and silver dissolution are faster with oxygen than with air, the effects being more marked with sodium cyanide than with potassium cyanide. An increase in the temperature from 25 to 35°C using oxygen also reduced the residual gold value from 32,6 to 18,6 g/t, and an increase in the cyanide addition from 30 to 70 kg/t reduced the residual gold value from 32,6 to 20,6 g/t — an effect that is probably purely kinetic. Potassium cyanide at a liquid:solid ratio of 1:2 gave lower residual values than sodium cyanide at a liquid:solid ratio of 2:1 for a given temperature and cyanide addition, both with air and with oxygen. The iron dissolution was again lower with oxygen than with air (possibly a cementation effect), and higher with potassium cyanide than with sodium cyanide (possibly owing to higher pH). An increase in the temperature from 25 to 35°C had no effect, but an

TABLE V

CYANIDATION OF SAMPLE W3 CONCENTRATE USING A PADDLE IMPELLER

Test no.	Stirrer (r/min)	Impeller paddle size	Residue		Extraction		Sodium cyanide		Gas flow-rate		Reaction time (h)	Final Cu in solution (g/t)	Final Fe in solution (g/t)
			Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	Added (kg/t)	Consumed (kg/t)	O_2 (cm³/min)	Air flow-rate (cm³/min)			
1	400	small	6993 TPM		—	—	25,0	15,6	20	—	6	1305	249
2	400	large	1497 TPM		—	—	25,0	17,1	20	—	6	1300	257
3	400	small	6024 TPM		—	—	25,0	14,7	10	—	6	1080	415
4	400	large	1997 TPM		—	—	25,0	16,5	10	—	6	1240	270
5	400	small	5197 TPM		—	—	25,0	15,5	100	—	6	1350	230
6	400	large	1848 TPM		—	—	25,0	16,1	100	—	6	1350	210
7	800	small	59,5	2,5	99,5	99,8	25,0	16,6	100	—	6	1360	190
8	800	large	79,0	3,0	99,4	99,7	25,0	17,0	100	—	6	1380	230
9	800	small	76,0	6,0	99,4	99,5	25,0	16,2	20	—	6	1380	190
10	800	large	81,0	3,5	99,4	99,7	25,0	16,6	20	—	6	1380	170
11	800	small	1025 TPM		—	—	25,0	18,0	10	—	6	1290	240
12	800	large	120	5,0	99,0	99,6	25,0	13,1	10	—	6	1410	190
13	800	large	82,5	7,0	99,4	99,5	30,0	18,0	—	500	24	1500	180
14	800	small	178	17,5	98,6	98,7	25,0	24,8	—	500	16	1420	790
15	800	large	60,0	9,0	99,5	99,3	30,0	18,6	—	100	24	1460	170
16	800	small	176	14,3	98,6	99,0	25,0	24,4	—	100	16	1390	820
17	400	large	5625 TPM		—	—	30,0	24,8	—	500	16	1215	1140
18	400	small	3404 TPM		—	—	30,0	15,7	—	500	16	1410	280
19	400	small	4132 TPM		—	—	30,0	27,1	—	100	16	1260	1435
20	400	large	84,0	10,0	99,4	99,3	30,0	18,4	—	100	16	1590	330
11B	800	small	114	23,7	99,0	98,2	30,0	24,9	—	500	16	1200	920
16B	800	small	149	26,2	98,9	98,0	30,0	25,5	—	100	16	1300	910
21	800	small	3279 TPM		—	—	25,0	25,0	—	500	6	9650	1230
22	800	large	284	85,3	97,5	92,7	30,0	18,0	—	500	6	1260	248
23	800	small	4685 TPM		—	—	25,0	25,0	—	100	6	8400	1330
24	800	large	585	120	95,1	90,3	30,0	16,8	—	100	6	1276	228

TABLE VI

CYANIDATION OF SAMPLE W3 CONCENTRATE USING A SUPER-AGITATOR*

Test no.	Residue		Extraction		Sodium cyanide		Gas flow-rate		Reaction time (h)	Final Cu in solution (g/t)	Final Fe in solution (g/t)
	Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	Added (kg/t)	Consumed (kg/t)	Oxygen (cm³/min)	Air (cm³/min)			
1†	144	18,0	98,6	98,4	25,0	22,2	200	—	6	1060	920
2	25,4	7,5	99,8	99,4	30,0	17,7	200	—	6	1360	100
3	19,1	6,8	99,9	99,5	35,0	22,0	200	—	16	1390	580
4	203	20,0	98,5	98,5	35,0	17,8	—	1000	6	1230	241
5	27,5	6,3	99,8	99,4	35,0	18,7	—	1000	16	1426	312
6	24,5	5,4	99,8	99,3	35,0	19,3	—	1000	24	1370	345

*Nomenclature of Denver Equipment
†Slow and inefficient filtration, and lower than normal residual cyanide

increase in the sodium cyanide addition from 30 to 70 kg/t again increased the iron dissolution considerably. The copper dissolution again appeared to be unaffected by changes in reaction conditions.

While the evidence presented for each type of agitator system is perhaps self evident, no definite conclusion can be drawn at this stage concerning the relative effectiveness of the flotation machine and the Super-agitator, the only conclusive results being that the flotation machine gave lower residual values when using air than did the Super-agitator for short reaction times, indicating better aeration characteristics. It can reasonably be expected that, when optimum conditions are used, the flotation-machine system will yield residual gold values of between 10 and 20 g/t after relatively short reaction times. While the present exercise indicated very clearly that the type of agitation is an important operational parameter in the cyanidation of gold-plant concentrates, further testwork on a larger scale would be required before the merits of the flotation-machine type of agitator can be evaluated.

Effects of Temperature and Calcium

It had been earlier established that both the liquid:solid ratio and the temperature of the pulp have a very direct bearing on the solubility of the aurocyanide complex during gold dissolution. Furthermore, the effects of calcium on such a leaching system and its effect on gold solubility had yet to be established, and the significance of such reaction parameters had only been deduced by indirect means. Further investigation under actual leaching conditions was warranted.

The investigation was thus primarily concerned with an elucidation of the effects of temperature and calcium on the kinetics of gold dissolution. The investigation was confined to the treatment of sample W4 (gold 19,4 kg/t) using the Super-agitator and the flotation-machine type of agitator as previously described.

For cyanidation carried out in the flotation-machine type of agitator, the experimental conditions and results are summarized in Table VIII, which shows the effects of temperature on gold dissolution, together with the use of deionized water and of air and oxygen aeration. Likewise, Table IX shows the effect of temperature on gold dissolution in the presence of 0,28 g/l calcium added in the form of gold-plant mill return water. The above effects are also illustrated graphically in Fig. 6.

It can readily be seen that, in both instances (Tables VIII and IX), temperature has a very important kinetic effect on the dissolution of gold at temperatures below 20°C. Gold extractions can be seen to decline considerably below 20°C, but were little affected in the range 25 to 50°C when oxygen aeration was used. However, the differences in the rate of reaction between air and oxygen were markedly lower at the higher reaction temperatures, sparging with air yielding residual gold values of 114 g/t at 25°C and 30 g/t at 45°C after a contact time of 6 hours, which can be compared with 15 g/t at 25°C and 15 g/t at 45°C using oxygen. The use of mill return water (Table IX) appeared to have no detrimental effect on the gold dissolution.

When the experimental programme was repeated with

TABLE VII

THE CYANIDATION OF SAMPLE W3 CONCENTRATE USING A FLOTATION MACHINE TYPE OF AGITATOR

Residue		Extraction		Cyanide		Gas flow-rate		Liquid solid ratio	Temp. (°C)	Final pH	Final Cu in solution (g/t)	Final Fe in solution (g/t)
Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	Added (kg/t)*	Consumed (kg/t)*	Oxygen (cm^3/min)	Air (cm^3/min)					
50,6	22,0	99,5	98,3	25,0 (NaCN)	16,6	50	—	2/1	25	12,1	1340	160
32,0	6,6	99,7	99,5	30,0 "	17,0	100	—	2/1	25	12,2	1230	178
121	—	†98,9	—	25,0 "	17,3	—	250	2/1	25	12,2	1270	300
44,0	12,6	99,6	98,9	30,0 "	18,4	—	500	2/1	25	12,2	1300	220
20,6	4,3	99,8	99,7	70,0 "	37,0	100	—	2/1	25	—	1290	1468
18,8	4,6	99,8	99,5	30,0 "	18,3	100	—	2/1	35	12,1	1235	165
19,5	10,6	99,8	99,0	25,0 (KCN)	20,8	250	—	0,5/1	25	13,1	4820	1420
14,6	4,4	99,9	99,6	30,0 (KCN)	23,2	250	—	0,5/1	25	13,1	4890	1875
56,8	17,4	99,5	97,7	30,0 (KCN)	22,6	—	1250	0,5/1	25	13,1	4680	2050

*Expressed as equivalent NaCN
†Slow and inefficient filtration

a Super-agitator, it became evident that no conclusions could be drawn from the results owing to poor reproducibility. The difficulty in obtaining reproducible results was ascribed to the efficiency of agitation. To be consistently efficient, and especially to avoid 'dead' spaces where unleached material may collect, the Super-agitator system appears to require careful design, which could not be obtained with the device used in this testwork.

When the present results concerning the effect of temperature on gold dissolution are compared with the findings of Julian and Smart[5], who studied the dissolution of pure gold, similar trends are most evident. Their results, which are presented in Fig. 7, show that below 20°C the rate of gold dissolution increases much more rapidly with temperature than it does above approximately 30°C. In the temperature range 20 to 30°C, a transition region between the fast and the slow reaction is apparent. The present investigations on the temperature dependence of leaching rates of gold from concentrates and that of pure gold show that below about 20°C the leaching rates are most sensitive to temperature

TABLE VIII

EFFECT OF TEMPERATURE ON GOLD DISSOLUTION FROM BELT CONCENTRATES USING A FLOTATION-MACHINE TYPE OF AGITATOR

Temperature (°C)	Residue		Extraction		Residual NaCN (g/l)	Gas flow-rate		Reaction time (h)
	Au (g/t)	Ag (g/t)	Au (%)	Ag (%)		Oxygen (cm³/min)	Air (cm³/min)	
25	15,0	2,6	99,9	99,9	6,9	200		6
35	19,2	2,8	99,9	99,9	7,2	200		6
45	15,1	2,0	99,9	99,9	5,8	200		6
25	114	29,8	99,4	98,6	7,0		1000	6
35	70,0	26,0	99,6	98,8	6,2		1000	6
45	30,2	13,3	99,8	99,4	6,8		1000	6
15	446	54,4	97,6	98,9	8,5	200		2
20	170	31,6	99,1	99,4	7,8	200		2
25	128	21,1	99,3	99,6	8,1	200		2
30	26,1	10,1	99,9	99,8	8,0	200		2
35	81,4	23,0	99,6	99,5	7,7	200		2
40	58,3	21,1	99,7	99,6	7,6	200		2
45	79,2	21,2	99,6	99,6	8,0	200		2
50	110	24,5	99,4	99,5	7,8	200		2

Material: Sample W4 (19,4 kg/t Au)
Conditions: Liquid: solid ratio 2:1
NaCN addition 40 kg/t

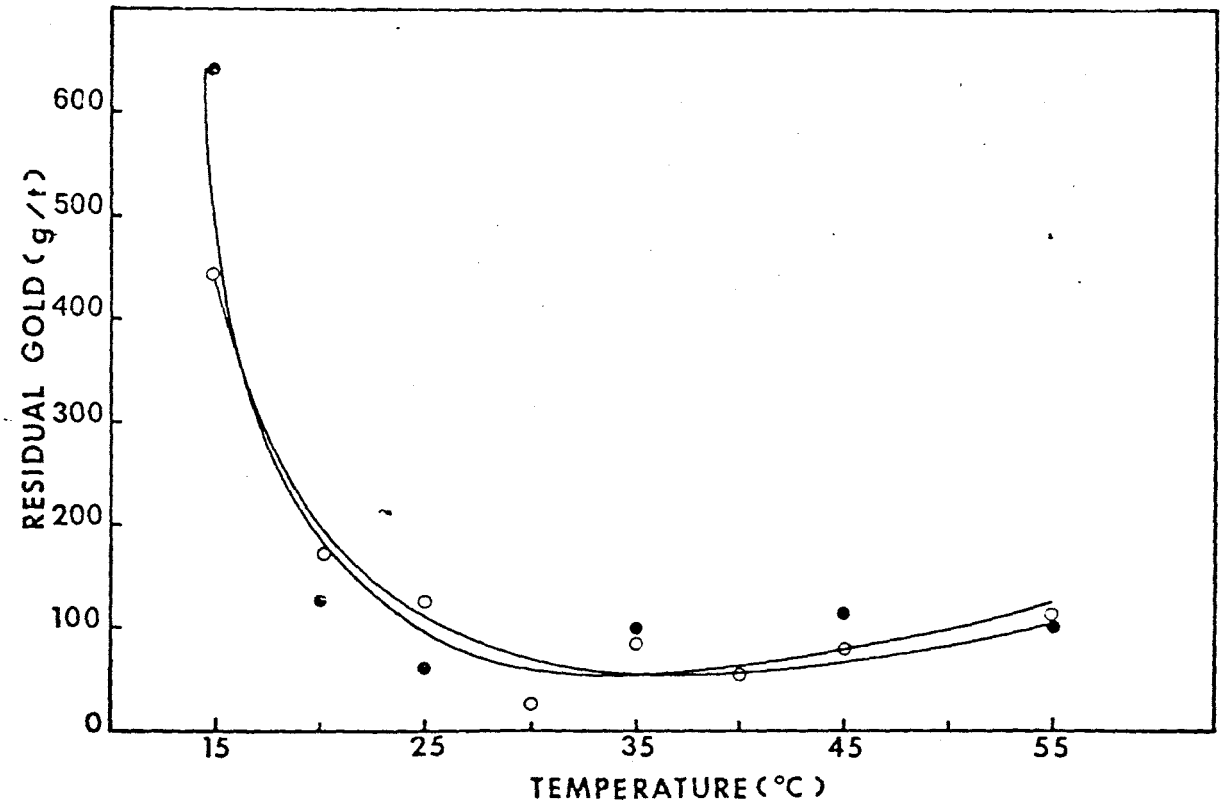

Fig. 6—Effect of temperature on the dissolution of gold from sample W5
O Using deionized water ● Using mill return water NaCN addition 40 kg/t Oxygen flow-rate 200 cm³/min
Liquid: solid 2:1 Contact time 2h

changes. On the other hand, above about 25°C there is no significant increase in the rate of gold dissolution with temperature.

The chemical behaviour of the gold in concentrates during leaching is similar to that of pure gold, and all the reactions that apply to pure gold appear to be valid for the gold present in concentrates. Numerous investigators have revealed that the dissolution of pure gold is usually controlled by diffusion. At low temperatures the rate control of gold dissolution can change from diffusion to reaction rate control. This change-over can be aptly demonstrated from the data of Julian and Smart. If the data illustrated in Fig. 7 are plotted in the form of log (rate) against the reciprocal of absolute temperature, a curve similar in shape to that presented in Fig. 7 is obtained. However, the portions below about 15°C and from 25 to 70°C yield straight lines with different slopes. From these respective slopes, an activation energy of about 13 kcal/mol can be derived from the line through the low temperature points, and approximately 2 kcal/mol for the temperature range 25 to 70°C. For the type of heterogeneous reaction in the liquid phase that

TABLE IX

EFFECT OF MILL RETURN WATER ON GOLD DISSOLUTION FROM BELT CONCENTRATES AT VARIOUS TEMPERATURES USING A FLOTATION-MACHINE TYPE OF AGITATOR

Type of water	Temperature (°C)	Residue		Extraction		Residual (NaCN) (g/l)
		Au (g/t)	Ag (g/t)	Au (%)	Ag (%)	
Deionized	15	446	54,4	97,6	98,9	8,5
Deionized	20	170	31,6	99,1	99,4	7,8
Deionized	25	128	21,0	99,3	99,6	8,1
Deionized	35	81,4	23,0	99,6	99,5	7,7
Deionized	45	79,2	21,2	99,6	99,6	8,0
Deionized	55	110	24,5	99,4	99,5	7,8
Mill Return	15	640	65,0	96,6	98,7	8,2
Mill Return	20	128	23,8	99,3	99,5	7,9
Mill Return	25	82,9	14,7	99,7	99,7	7,9
Mill Return	35	104	28,3	99,4	99,4	8,4
Mill Return	45	114	14,0	99,4	99,7	8,1
Mill Return	55	99.0	30,0	99,5	99,4	8.2

Material: Sample W4 (19,4 kg/t Au)
Conditions: Liquid: solid ratio 2:1
NaCN addition 40 kg/t
Oxygen flow-rate 200 cm³/min
Reaction time 2 h

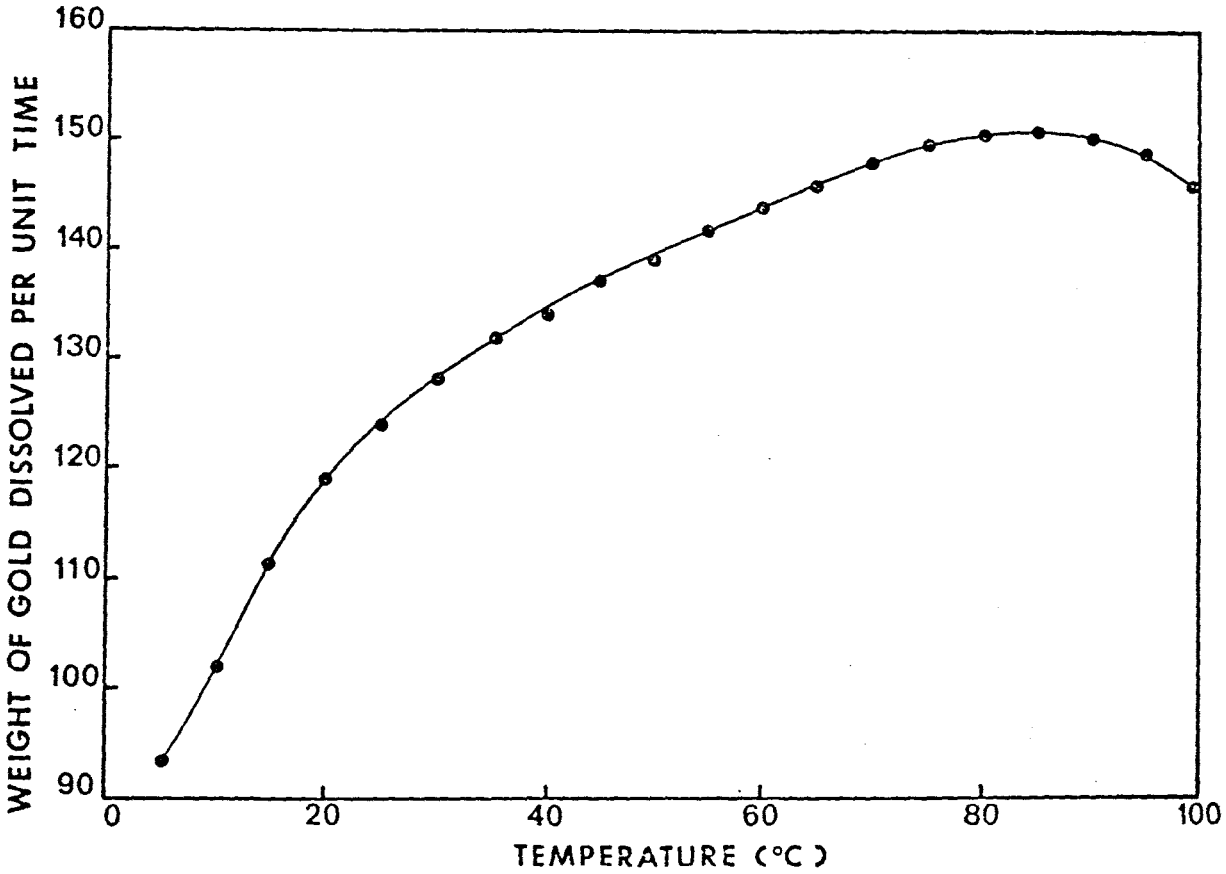

Fig. 7—Effect of temperature on the dissolution of pure gold (after Julian and Smart[5])

is responsible for gold dissolution, the higher activation energy is characteristic for a reaction rate controlled process, while the lower one is typical for diffusion controlled systems. Evidence for a change from reaction rate control at low temperature to diffusion control at the higher temperature was also found in the present investigation, i.e., below 20°C the gold dissolution rate had a strong dependence on temperature, whilst above about 20°C no change of gold dissolution rates with increase in temperature could be determined. Further evidence regarding diffusion control above 20°C was that, under normal operating conditions, increased agitation and aeration and an increase in oxygen partial pressure resulted in increases in the gold dissolution rates.

Other Reaction Variables

It became most apparent during the course of the present investigation that the reaction variables including pH, cyanide concentration, and time had very complex compounding effects upon one another, and as such were difficult to define. As the definition of these compounding effects necessitated a large number of tests, further study was confined to simple tests involving the rolling of bottles. Some 500 to 600 individual tests were conducted (the equipment being housed in a thermostatically controlled cabinet). For the sake of brevity, only the more pertinent features of this study are discussed here.

The effect of reaction time on the extraction of gold from various concentrates in the presence of air is shown in Fig. 8. It can be seen that, after an initial acceleration

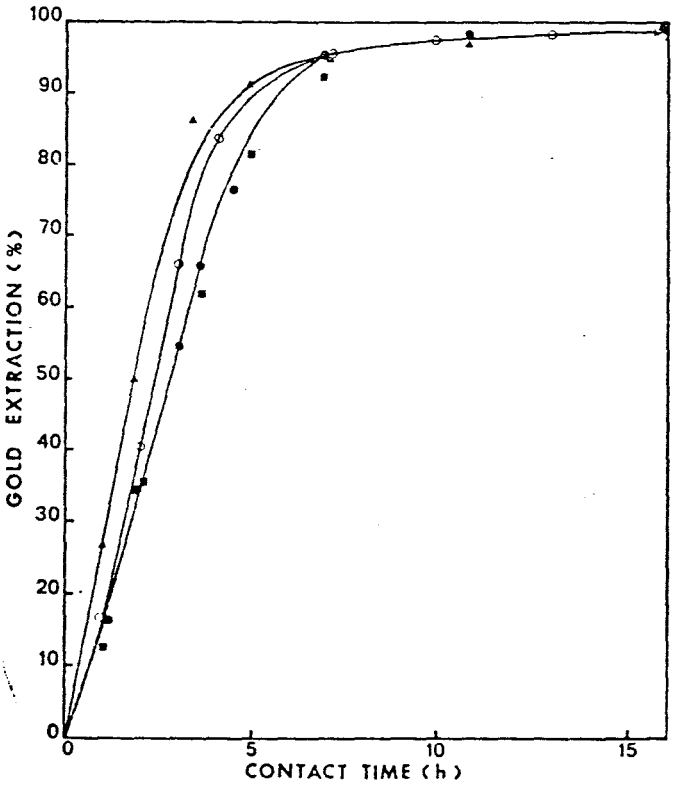

Fig. 8—Effect of contact time on the dissolution of gold from belt concentrate at 20 C

O Sample W1 (Au 6,3 kg/t) ▲ Sample W3 (Au 12,2 kg/t)
● Sample W2 (Au 22 kg/t) ■ Sample W4 (Au 19,4 kg/t)

period, the rate of gold dissolution is almost constant for a considerable time (1 to 4 hours), and that this rate is roughly proportional to the original gold concentration, i.e., the higher the head grade, the faster the dissolution. In other words, the percentage extraction with time is roughly the same even for concentrates with very different gold contents. This cannot be strictly true for reasons that will be given later.

As expected, these reaction rates, which would yield a gold dissolution of 100 per cent in 4 to 5 hours irrespective of the head value, do not continue. After about 70 to 80 per cent of the gold has been dissolved, the rate of dissolution decreases significantly and becomes very slow after about 7 hours, when about 95 per cent of the gold has been dissolved. Approximately the same time is required for the dissolution of most of the remaining 5 per cent as for the dissolution of the first 95 per cent. Under similar experimental conditions but for extended leaching times (48 to 96 hours), the final residual gold values were approximately 10 to 20 g/t for all the concentrates so far investigated (W1 to W4). These residual gold values appear to be independent, within reasonable limits, of the magnitude of the reaction variables; that is, under all reasonable operating conditions, a final residual gold value, which is characteristic of the particular concentrate, can be achieved if leaching is continued for a long enough time. It also means that, at extended leaching times, all the confounding effects disappear. Thus, the effectiveness of any variation in the operating conditions can be assessed by the leaching time required to reach residual gold values of 10 to 20 g/t in the present instance or by how closely this residual value can be approached in a given time.

Generally, the dissolution of silver followed the same trend as that of gold, but the dissolution of silver appeared to be somewhat slower than the dissolution of gold and, during the progress of the reaction, the gold: silver residue ratio decreased from its original value of about 9:1 to roughly 5:1.

The leaching of sized fractions was undertaken for two reasons. First, it was suspected that the coarser gold might be responsible for the slower leaching rates observed after about 70 to 80 per cent of the gold had been dissolved. Second, it was hoped that this exercise might throw more light on the nature of the refractory or residual gold. In the past, speculations about the nature of this gold had ranged from its being occluded gold to its being gold in solid solution or even adsorbed gold. The observation that milling in cyanide or milling followed by cyanidation lowered the residual value considerably hinted strongly at occluded gold. Occluded gold particles were also observed during microprobe analyses of the leach residues.

A size analysis and gold distribution of the fractions used in the present investigation are presented in Table II, and the results of leaching tests are shown in Fig. 9. As expected, the $+147\ \mu m$ size fraction, which contained the lowest concentration of gold, had the slowest leaching rate, whilst the $-74\ \mu m$ fraction leached the fastest. In spite of the large difference in the head grade of gold between the coarsest and the finest fractions, the actual residual gold value after about 3 hours leaching

was already lower for the finest fraction. In addition to being the fraction with the slowest leaching rate, the +147 μm material was also responsible for the greatest proportion of residual gold.

The effect of cyanide concentration can best be demonstrated by a graph of residual gold versus residual cyanide concentration. Such a graph is given in Fig. 10, which shows that, under the specified conditions, increasing concentrations of residual cyanide have a very significant effect on the rate of gold dissolution when the leaching is done in the presence of air, but only a very small effect on the gold dissolution rate when the leaching is carried out in the presence of oxygen. In both instances, the use of high cyanide additions invariably resulted in a higher consumption of cyanide, while, if the cyanide concentration dropped below approximately 0.5 g/l, the gold dissolution almost ceased. It was also found during this investigation that very little difference in gold dissolution rates were observed when 'flake' cyanide (50 per cent calcium cyanide) rather than sodium cyanide was used. In fact, a very beneficial effect was noted in that the solubility of the iron was markedly depressed by the use of calcium cyanide. Another advantage in the use of calcium cyanide rather than sodium cyanide is the increased solubility of calcium aurocyanide. Because analytical grades of both calcium cyanide and calcium aurocyanide could not be obtained at the time and were difficult to prepare, the solubility of calcium aurocyanide was estimated by an indirect method. A high-grade gold concentrate was leached with Aero Brand calcium cyanide (S.A. Cyanamid (Pty) Limited) at a low liquid:solid ratio, and the maximum amount of gold that could be dissolved was considered to indicate the solubility of calcium aurocyanide under such leaching conditions. Thus, at a temperature of 20 °C, a solution with a calcium concentration of about 25 g/l contained 18 g/l gold. This is more than double the quantity of gold that could be contained by 1 litre of sodium aurocyanide solution with a sodium concentration of about 15 g/l. However, the solubility of calcium aurocyanide is much lower than that of potassium aurocyanide. A solution containing approximately 75 g/l calcium as calcium aurocyanide could dissolve only about 7 g/l gold at approximately 20 °C. Other cyanide reagents besides those discussed above have no practical significance and were therefore not investigated.

Leaching data on the effect of pulp pH on gold dissolution are presented in Table X, from which it is most evident that increasing pH has a significant detrimental effect on the rate of gold extraction under the specified conditions. The addition of a protective alkali, following the standard practice in plants with low-grade circuits, would therefore be detrimental only if applied to the leaching of gold concentrates, since the resultant pH during the leach is quite sufficient to prevent the evolution of cyanide. The opposite should rather be carried out; that is, acid should be added, but the danger of lowering the pH into a region where the evolution of hydrogen cyanide becomes significant would militate against such a practice except under strictly controlled

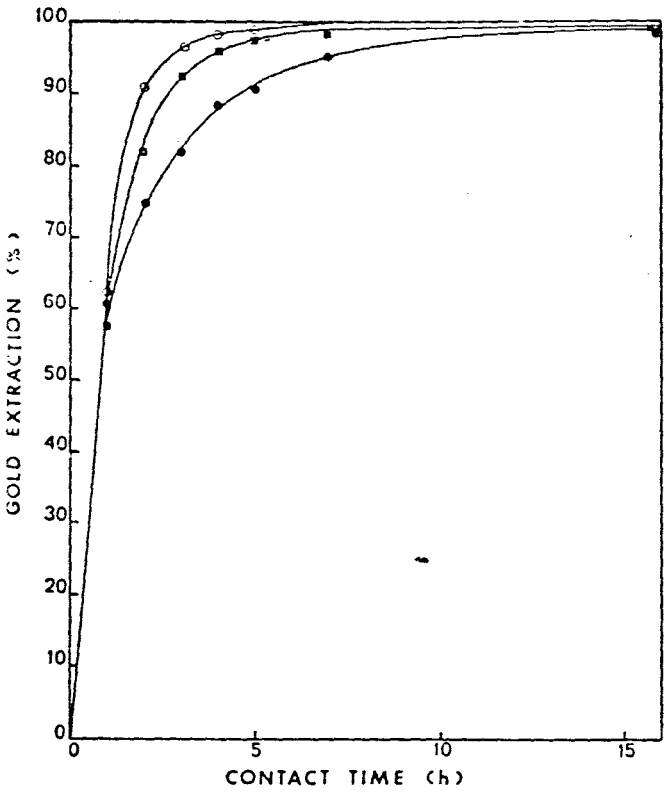

Fig. 9—Effect of particle size on the dissolution of gold from sample W4 at 30 C
○ —74 μm (Au 117 kg/t) ● +147 μm (Au 45 kg/t)
▣ +74—147 μm (Au 13,8 kg/t)
NaCN addition 40 kg/t Air aeration in rolling bottle
Liquid: solid 3:1

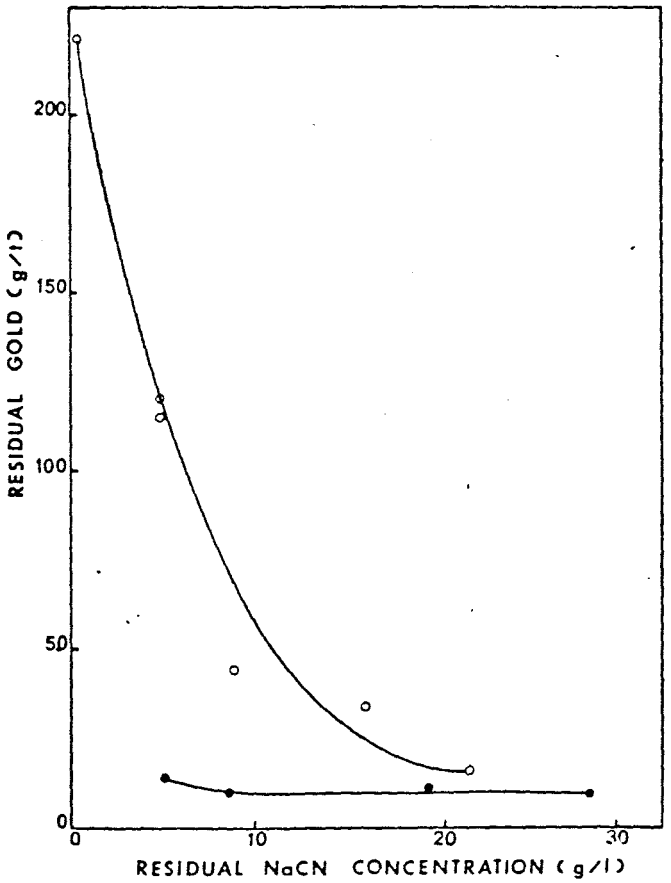

Fig. 10—Effect of residual cyanide concentration on the dissolution of gold from sample W3 at 20 C
○ Aeration ● Oxygenation
Liquid: solid 2:1 Contact time 1h

conditions. Although the reasons for the high pH value observed during the leaching of concentrates are twofold, they are attributable to the same source. At the beginning of the reaction when cyanide is added to the pulp, some hydrolysis of the cyanide takes place according to the equation

$$CN^- + H_2O = HCN + OH^- \quad \ldots\ldots\ldots\ldots \quad (2)$$

The extent of the hydrolysis is demonstrated in the case of sodium cyanide and potassium cyanide in Table XI. It can be seen that, at the high cyanide addition required for the leaching of gold concentrates, a safe pH value is obtained even before the start of the reaction, and the addition of a protective alkali is no longer necessary. The rapidly ensuing gold dissolution reaction (1) produces more hydroxyl ions, which accounts for a rise in pH during the course of leaching. Thus, the higher the gold tenor of the concentrate, the higher the resultant pH will be and the slower the resultant leach rate.

Since increasing cyanide concentration also causes increasing cyanide consumption, especially when air-sparging is carried out, improvement of the gold dissolution rate by this method is not economically attractive owing to the current costs of reagents. The effects of pH become really detrimental only at a pH level of more than 12,5. Such high pH values can be obtained only from very rich concentrates and fairly low liquid:solid ratios. Even temperature, provided it is maintained above 25°C, does not appear to provide a ready means of controlling the reaction. This leaves oxygen partial pressure as the one variable of those currently investigated for effective rate control. Another factor that in practice is closely allied with oxygen partial pressure and that will permit a considerable amount of rate control is the type and rate of pulp agitation. Theoretical considerations indicate that the rate and type of stirring may even permit the same control of the rate of gold dissolution as can be effected by oxygen partial pressures. To what extent use will be made of either or both of these factors in improving the rate of gold dissolution depends entirely on economic considerations.

TABLE X

EFFECT OF pH ON GOLD DISSOLUTION FROM BELT CONCENTRATES

Contact time (h)	6	16	6	16
pH	Residual Au (g/t)		Au Recovery (%)	
11,5	180	55	99,1	99,7
12,3	340	95	98,2	99,5
12,7	3 100	240	84	98,3
12,9	6 400	440	67	97,7
13,0	9 000	750	54	96,1

Conditions: Cyanide addition 60 kg/t; Temperature 40°C; Material Sample W4; Liquid: solid ratio 2:1

TABLE XI

CHANGE OF pH WITH INCREASING CYANIDE CONCENTRATION AT 20°C

Cyanide concentration (g/l NaCN/KCN)	pH with NaCN	pH with KCN
1	10,7	10,7
2	10,9	10,9
3	11,0	11,0
5	11,1	11,1
7	11,1	11,1
10	11,2	11,2
15	11,3	11,3
25	11,3	11,4
35	11,3	11,5
50	11,3	11,5

Cyanidation and Flotation in a Bench-scale Pilot Reactor

The efficiency of a flotation-machine type of agitator for the cyanidation of gold-plant belt concentrates having already been established, a bench-scale pilot reactor was constructed. It consisted basically of a stainless-steel leaching vessel and a Denver D12 flotation-machine agitator. The aim of the present investigation was to obtain reaction data on a larger scale for certain parameters that would affect the design of an envisaged full-scale plant. Because of gold solubility problems associated with the treatment of very rich high-grade concentrates, which would consequently seriously affect the capacity of such a plant, the feasibility of floating gold before cyanidation was also investigated.

Equipment and Procedure

The leaching apparatus is shown in Fig. 11. The vessel was fitted initially with a heating element coupled to an automatic temperature controller. However, it was found that temperatures increased during the course of the reaction owing to the violent agitation, and the heating element was subsequently replaced by a cooling coil.

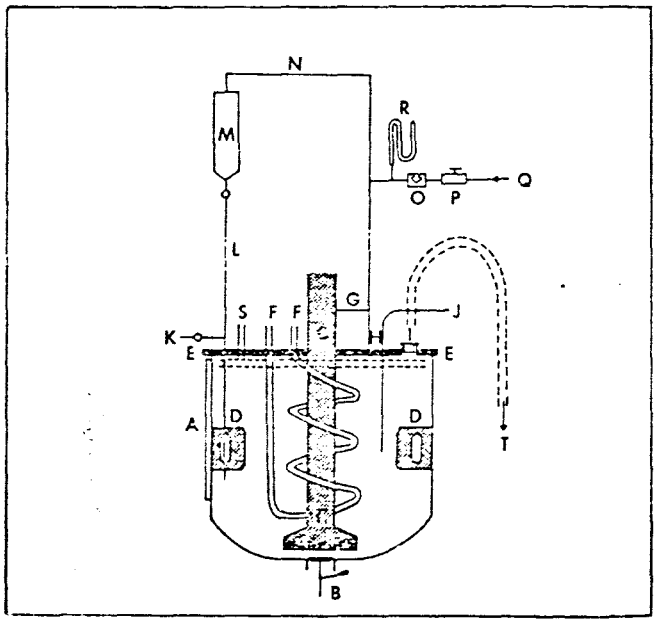

Fig. 11—The intensive cyanidation bench-scale reactor (10 kg capacity)

A Perspex window, 5 cm by 20 cm
B Discharge mechanism
C Denver D12 flotation impeller
D Baffles
E Flanged cover plus gasket
F Cooling coil
G Air/oxygen feed
H Oxygen recycle
I Solids-feeding port/flotation port
J Temperature control
K Sampling/decantation tube
L Cyanide-solution feed
M Cyanide-solution storage
N Pressure balancing line
O Flowmeter
P Pressure regulator
Q Air/oxygen supply
R Manometer
S Washing ring
T Flotation concentrate

When air was used, the system was open to the atmosphere, but, when oxygen was used, the vessel was sealed and oxygen was fed automatically through a pressure regulator, the pressure in the vessel being kept constant at approximately 7 to 14 kPa and oxygen being added only as it was consumed.

At the end of each leaching experiment, the solids were washed by decantation as follows: the solids were allowed to settle for 15 minutes for the first two decantations and for 10 minutes for subsequent decantations; the cycle of washing and decantation was continued until a neutral wash water was obtained, after which the residue in the vessel was discharged, filtered, and dried.

The on-site flotation of fourteen 1 kg samples of high-grade belt concentrates (gold 11 to 40 kg/t) by a well-established gold-flotation procedure yielded average gold recoveries of 88 per cent at rougher-concentrate grades averaging 48 per cent gold. The cyanidation of the resulting flotation tailings (gold 0,4 to 1,9 kg/t) was subsequently carried out in the bench reactor. The feasibility of floating gold direct from the bench reactor was also studied (see later).

Cyanidation

When a high-grade gravity concentrate (sample W4 — gold 19,4 kg/t) was leached using oxygen sparging at a liquid : solid ratio of 2 : 1 and controlling the temperature in the region 30 to 35 °C, gold extractions of 99.9 per cent were obtained after a contact time of 4 hours (Table XII). Similarly, for concentrate flotation tailings containing 432 g/t gold (Table XIII), the overall gold recoveries averaged 99,8 per cent when the flotation was taken into account. Once again, the use of oxygen would appear to be advantageous where high-grade concentrates are treated directly by cyanidation, but, with lower-grade flotation tailings, air or a short period of sparging with oxygen followed by air would appear to be adequate. The lowest value for residual gold obtained in the present exercise (6,9 g/t) was obtained after sparging with oxygen for 1 hour followed by aeration for 16 hours. From a technical point of view, the present investigation indicated that a very simple operation yielding exceptionally high gold recoveries is technically feasible.

Direct Flotation from Reactor

Sample W6 of gold-plant belt concentrate (gold 2,5 kg/t) was submitted to flotation in the cyanidation reactor (Fig. 11), whereby the excess froth was piped off through a 25 mm plastic tube connected to the ore-feeding port in the lid of the vessel. After a flotation procedure involving 25 minutes, a concentrate grade of 7,3 per cent gold was obtained, representing 3,9 per cent of the original mass. The gold recovery was 87,4 per cent with tailings assaying 421 g/t gold. As such a flotation recovery compared very favourably with the more conventional procedure, it would appear to be recommended, especially for high-grade concentrates. Although the cyanide consumption was reduced only marginally, the cyanidation times can be expected to be considerably shorter. The possiblity of floating gold direct from the cyanidation vessel thus appeared very attractive because of its simplicity and relative effectiveness.

Recovery of Osmiridium from Leach Residues

Three composite samples of belt concentrates designated W1 to W3 were initially leached by intensive cyanidation, and the residues were tabled for the recovery of osmiridium. Approximately 10 kg of each

TABLE XII

CYANIDATION OF BELT CONCENTRATES IN THE BENCH-SCALE REACTOR

Contact time (h)	Temperature rise (°C)	Residual Au (g/t)	Au extraction (%)	Sodium cyanide		Fines decanted (% of total)
				Added (kg/t)	Consumed (kg/t)	
3	16-53 No control	124	99,4	40,0	28,0	3,8
6	13-59 No control	270	98,6	40,0	34,4	6,7
3	18-56 No control	67,9	99,6	40,0	31,4	6,7
6	30-35 Controlled	939	95,3	40,0	40,0	5,5
4	30-35 Controlled	26,9	99,9	55,7	40,5	5,2

Material: Sample W4 (19,4 kg/t Au)
Conditions: Liquid : solid ratio 2:1
Oxygen aeration

TABLE XIII

CYANIDATION OF BELT-CONCENTRATE FLOTATION TAILINGS IN THE BENCH-SCALE REACTOR

Contact time (h)	Temperature (°C)	Oxidizing agent	Impeller speed (r/min)	Residual Au (g/t)	Au extraction (%)	Sodium cyanide		Fines decanted (% of total)
						Added (kg/t)	Consumed (kg/t)	
4	20-25	Air	1450	9,1	97,8	30,0	21,9	7,5
4	30-35	Air	1450	38,9	90,8	37,5	25,7	5,8
4	40-45	Air	1450	6,5	98,5	30,0	18,1	7,6
3	30-35	Oxygen	1450	9,7	97,7	30,0	18,5	7,2
1	30-35	Oxygen	1450	9,6	97,7	35,0	25,2	7,4
1	30-35	Oxygen	1100	33,8	92,0	35,0	23,5	5,1
1	30-35	Oxygen	800	15,9	96,2	35,0	24,5	5,8
1+16*	30-35	Oxygen + Overnight air	1450	6,9	98,4	25,0+10*	24,3+8,1*	6,0

*1 hour of oxygen followed by 16 hours of air.
Material: Flotation tailings (432 g/t Au)
Conditions: Liquid : solid ratio 1:1

leach residue composite was tabled on a small Wilfley shaking table, and concentrate and tailing fractions were taken for subsequent analysis. The results are presented in Table XIV, where it can be seen that an average platinum recovery (osmiridium) of 76 per cent was obtained in a mass distribution of 3,7 per cent. The concentrate fraction can also be seen to contain on average 24 per cent of the total gold and 49 per cent of the total uranium values. The results show remarkably good reproducibility with regard to the recovery of platinum and uranium. Such findings indicate that a very large percentage of the heavy-mineral phases are currently recirculating in the existing closed gold-plant gravity circuit. A mineralogical examination of various residue fractions indicated that the gold was almost exclusively occluded in the uraninite and pyrite phases. As the tabling of leach residues yielded tailings containing 10 to 12 g/t gold, roughly half of which was estimated to be occluded in the large recycling uraninite phase, the eventual opening of the gravity circuit should roughly halve the subsequent residual gold values.

The evidence presented, together with the findings on the cyanidation of gravity concentrates given earlier, shows that, in the case of composite belt concentrates, intensive cyanidation followed by a tabling operation for the recovery of osmiridium may eventually yield residues containing as little as 5 g/t gold. It is difficult to assess at present whether such residues merit further milling to reduce the gold values even further, or whether the simple calcining of such residues in a conventional acid plant would not be a more practical operation.

Electrowinning of Gold from Pregnant Liquor

In view of the high-tenor gold solutions resulting from the cyanidation of gold-plant concentrates, as well as the relatively high residual cyanide concentrations of such solutions, the electrowinning of gold was considered to be a better alternative than zinc precipitation. As a result of investigation at these laboratories on the electrowinning of gold from relatively concentrated cyanide solutions, a cell consisting of an inner cathode compartment separated from an outer annular anode compartment by a semi-permeable cation-exchange membrane was developed (Fig. 12). The use of a semi-permeable membrane allows for the regeneration of cyanide in the catholyte. Gold and silver were plated onto mild-steel wool, and the iron was subsequently removed from the product either by fusion and fluxing or by leaching with acid. It was shown that, with con-

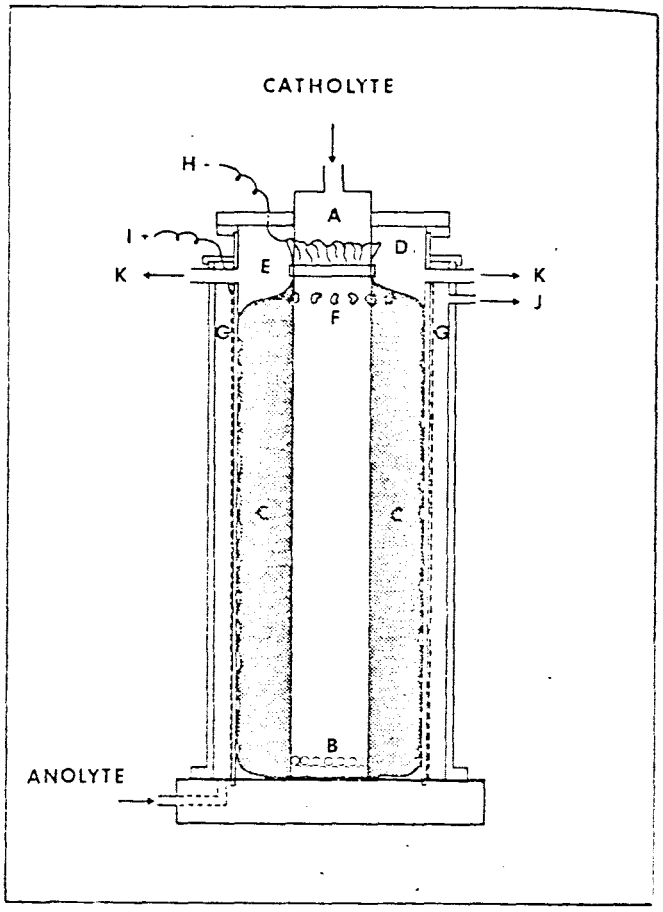

Fig. 12—The steel-wool gold-electrowinning cell

A PVC inlet distribution tube B Catholyte distribution ports
C Mild-steel wool D Sock made from filter cloth
E Band for securing filter sock F Catholyte outlet holes
G Annular perforated stainless-steel anode enclosing the cation-exchange membrane
H Cathode connection I Anode connection
J Recirculating NaOH/Na_2SO_4 anolyte
K Recirculating catholyte

TABLE XIV

RECOVERY OF OSMIRIDIUM FROM BELT-CONCENTRATE LEACH RESIDUES

Material	Mass distribution (%)	Concentrate grades			Distribution			
		Au (g/t)	Pt (kg/t)	U_3O_8 (%)	Au (%)	Pt (%)	U_3O_8 (%)	Pyrite (%)
Sample W1								
Concentrate	2,5	405	1868	345	28,0	77,0	47,0	1,0
Middling 1	16,6	26,1	26,4	—	12,2	7,4	—	24,5
Middling 2	74,5	24,8	10,9	—	52,1	13,6	—	67,0
Tails	6,4	41,8	11,4	—	7,7	1,0	—	7,5
Sample W2								
Concentrate	5,1	83	655	295	25,0	80,0	53,0	2,1
Middling	19,8	14,5	17,8	35	18,5	8,5	24,4	23,4
Tails	75,1	12,4	6,4	8,6	56,5	11,5	22,6	74,5
Sample W3								
Concentrate	3,6	76	625	221	20,0	70,0	47,0	1,8
Middling	22,3	14,3	16,5	21	22,0	10,2	27,0	28,2
Tails	74,1	11,1	93	6	58,0	20,8	26,0	70,0

centrated solutions exceeding 1 g/l gold, 100 per cent current efficiency was obtained, but this efficiency decreased at lower concentrations and was about 40 per cent at 150 g/t gold. The mild-steel wool was the type used for cleaning, and could be loaded with at least ten times its own mass of gold plus silver. Minor amounts (less than 0,1 per cent) of copper and zinc, and traces of nickel were deposited, but a gross amount of mercury originating from amalgamation (2,8 per cent) was co-plated with the gold. For leach liquors approximating 2 g/l gold, the cost of electrolysis plus smelting to yield bullion was estimated to be approximately R1,22 per kilogram of gold recovered for a large scale of operation. The cost of power for electrolysis was approximately 0,3 cents per kilogram of gold.

Discussion

The results presented in the investigation strongly suggest that an alternative cyanidation route for the treatment of gold-plant concentrates is technically feasible, provided effective agitation and oxygenation of the system are afforded. In the case of lower-grade concentrates, cyanidation followed by the electrowinning of gold would appear to be the simplest route, while, for high-grade concentrates, flotation and smelting of a concentrate followed by the cyanidation of flotation tailings are suggested. Again, the use of oxygen rather than of air for sparging the pulp will depend largely on the grade of the material to be treated.

A preliminary study of the costs of the amalgamation and cyanidation processes for the recovery of gold from gravity concentrates was also undertaken at these laboratories. For the sake of comparison, a plant treating approximately 2 t of concentrates per day was considered. The total capital costs (1977) were estimated to be as follows:

Amalgamation	R247 000
Cyanidation	R293 000
Cyanidation plus flotation from reactor	R294 000
Cyanidation plus separate flotation circuit	R328 000.

The following operating costs per ton of concentrate (1977) were estimated:

Amalgamation	R26,4
Cyanidation (air)	R24,0
Cyanidation (oxygen)	R26,6
Flotation plus cyanidation (air)	R24,4
Flotation plus cyanidation (oxygen)	R26,4.

Thus, while capital costs favour the amalgamation route, operating costs favour cyanidation, although differences in operating costs are relatively small. However, small additional costs would be incurred in the electrowinning of gold from solution in the cyanide routes, which would reduce still further the differences in operating costs for the alternative processes. It was concluded that, in the selection of the best route for the treatment of concentrates, other considerations, e.g. toxicity control, safety, and lower labour requirements, are more important than the small cost differences. Furthermore, the cyanidation route and the flotation plus cyanidation route are expected to achieve a gold extraction of more than 99,5 per cent. This is somewhat higher than for amalgamation, which is claimed to extract 96 to 97 per cent of the gold, the balance being partially extracted elsewhere in the circuit. Amalgamation has the further disadvantage compared with the other routes of involving a fairly large and uncertain circulating load in the gold-plant gravity circuit; 'opening' this circuit would considerably reduce the capital and operating costs for the cyanidation route because of the decreased tonnage to be treated.

While many of the aspects related to the current investigation concerned the possible cementation of gold onto the large amount of tramp iron present in the material studied, the subsequent 'opening' of the gold-plant gravity circuit may well change the composition of such concentrates considerably. This aspect alone could modify the conclusions drawn from the present investigation. Only a full-scale plant operation with careful monitoring of the process variables and the composition of the material treated would provide all or some of the answers necessary for such an appraisal.

Sequel

As a result of the investigation described here, a full-scale plant for the intensive cyanidation of gold-plant gravity concentrates was commissioned at the Welkom Gold Mining Company in March 1977. At the time of writing, average gold recoveries of more than 99 per cent were indicated, with gold in the leach residues running at 10 to 50 g/t.

Acknowledgement

The authors are indebted to the Analytical Laboratory of Anglo American Research Laboratories for the chemical analyses and to the management of Anglo American Research Laboratories for permission to publish these findings.

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PATENT SPECIFICATION

Country: SOUTH AFRICA

Application Number: 77/1777

Date of Filing: 24 MARCH, 1977

Name of Applicant: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

Name of Inventor:
1. RAYMOND JOHN DAVIDSON
2. CHRISTIAN GHISLAIN SCHMIDT

Title of Invention: "APPARATUS FOR CARRYING OUT FLOTATION AND/OR LEACHING PROCESSES"

~~Provisional~~/*Complete Specification as Filed*

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JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

579.

FILING DATE:

REPUBLIC OF SOUTH AFRICA

THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 77/1777

TITLE OF INVENTION: "APPARATUS FOR CARRYING OUT FLOTATION AND/OR LEACHING PROCESSES"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 MAIN STREET,
JOHANNESBURG.

KWE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

Brill Bros. F 8812

BACKGROUND OF THE INVENTION

THIS invention relates to flotation and leaching processes, and to apparatus for carrying out such processes.

SUMMARY OF THE INVENTION

The invention provides apparatus for carrying out a flotation or leaching process which comprises a closed vessel for receiving the pulp for the process, means to produce a froth from the pulp in the vessel, and a first outlet near the upper end of the vessel through which the froth is withdrawn.

The means for producing the froth is preferably a flotation type impeller. Such impellers are well known in the art.

The apparatus has particular use in the flotation of gold concentrates, particularly high grade gold concentrates. Such gold concentrates may be subjected to either a leaching process or a flotation and leaching stages. The froth produced during flotation is drawn off through the first outlet for further treatment.

The first outlet is preferably directly connected to a pipe which may be removed so that the pulp can be introduced into the vessel through the outlet aperture. With the pipe connected to the outlet, since the vessel is closed, it is apparent that the apparatus has security advantages

in the flotation of valuable metals such as gold and safety advantages when the process requires the use of noxious gases. The apparatus also has the facility to circulate the reactive gases such as oxygen which may be required in any leaching process.

The invention also provides a flotation process which includes the steps of introducing a pulp into a confined space, producing a froth from the pulp in the space, and withdrawing froth from the upper end of the space.

The process may include the steps of regulating the pressure and the temperature inside the space. In the case of the flotation of gold concentrates, the gold in the tailings from the flotation may be recovered using the cyanidation process.

The invention further extends to a method of leaching high grade gold concentrates which includes the steps of introducing the concentrate into a closed vessel, adding cyanide to the concentrate, and agitating the resulting pulp.

The aeration of the pulp using oxygen rather than air may be beneficially used in such a case.

DESCRIPTION OF THE INVENTION

The invention is further described by way of examples with reference to the accompanying drawing which schematically illustrates in cross section apparatus according to the invention.

Referring to the drawing there is shown a closed stainless steel vessel 10 for receiving the pulp for a flotation or leaching process, a pipe 12 leading from the vessel and a Denver D12 flotation type impeller 14. Froth formed during flotation is removed from the vessel through the pipe 12. The pipe may be removed from the vessel to enable the pulp to be introduced into the vessel through the resulting port 16 in the vessel wall. A cooling coil 18 is provided to minimise temperature rises during flotation. The vessel is further provided with baffles 20 on its side walls to regulate the flow of the pulp inside the vessel and a discharge outlet 22 for the removal of tailings from the flotation.

Gas for the flotation or leaching process is introduced into the vessel through the impeller in the usual way. The gas may be recycled, if so desired, by maintaining a slight positive pressure inside the sealed vessel. Use is made of a pressure regulator to control the pressure inside the vessel, and the gas being introduced into the vessel through the regulator.

This apparatus was used for the flotation of a gold concentrate (40,4 kg/ton gold). A 20 percent solids pulp was made of the concentrate and the pH adjusted to 10 with calcium oxide. To the pulp was added the following ingredients:

1. 100 g/t sodium cyanide as a 7 percent solution.
2. 500 g/t Aerofloat 208 as a 10 percent solution.
3. 50 g/t Aerofloat 25, undiluted.
4. 24 cc/t Teefrcth D, undiluted.

The pulp was introduced into the vessel through the port 16 and the pipe was then fitted to this port. The pulp was conditioned for ten minutes and then floated for 25 minutes using air. The froth formed during flotation was withdrawn from the vessel through the pipe 12 from whence it was delivered to further processing steps. The flotation yielded a gold recovery of over 85 percent. The gold in the tailings in the vessel was recovered by adding sodium cyanide to the pulp and agitating for a further 16 hours. An overall gold recovery exceeding 99.5 percent by this method was obtained.

The apparatus was also used for the leaching of a high grade gold concentrate (12 200 g/t gold and 1 200 g/t silver). A 20 percent solids pulp was made of the concentrate and the pulp was initially purged with oxygen for 10 min. Sodium cyanide was then added at a level of 30 kg/t, and the reaction

vessel closed. A small positive pressure of oxygen of approximately 1 psi was maintained in the vessel in order to balance the amount of oxygen consumed during the reaction. The temperature of the reaction was maintained at 35°C. After 6 hours of agitation and aeration, it was found that 99,8 percent of the contained gold and 99,5 percent of the contained silver in the original material was rendered soluble.

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

1.

Apparatus for carrying out a flotation of leaching process which comprises a closed vessel for receiving the pulp for the process, means to produce a froth from the pulp in the vessel, and a first outlet near the upper end of the vessel through which the froth is withdrawn.

2.

Apparatus according to claim 1 in which the froth producing means comprises a flotation type impeller.

3.

Apparatus according to claim 1 or 2 which includes means to regulate the temperature inside the vessel during the process.

4.

Apparatus according to claim 3 in which the temperature regulating means comprises a cooling coil.

5.

Apparatus according to any one of claims 1 to 4 which includes baffle means inside the vessel to regulate the flow of the pulp inside the vessel.

6.

Apparatus according to any one of claims 1 to 5 which includes means to introduce gas for the process into the vessel.

7.

Apparatus according to claim 6 which includes a pressure regulator, the gas being introduced into the vessel through the regulator so that the pressure inside the vessel can be controlled.

8.

Apparatus according to any one of claims 1 to 7 which includes a second outlet at the lower end of the vessel for the extraction of tailings from the vessel.

9.

Apparatus for carrying out a flotation or leaching process substantially as herein described with reference to the accompanying drawing.

10.

A flotation process which includes the steps of introducing a pulp into a confined space, producing a froth from the pulp in the space, and withdrawing froth from the upper end of the space.

11.

A process according to claim 10 which includes the step of regulating the temperature inside the space.

12.

A process according to claim 10 or 11 which includes the step of regulating the pressure inside the space.

13.

A process according to any one of claims 10 to 12 for the flotation of gold concentrates which includes the step after the withdrawal of the froth of recovering the gold in the tailings with a cyanidation process.

14.

A flotation process substantially as herein described with reference to the accompanying drawing.

15.

A method of leaching high grade gold concentrates which includes the steps of introducing the concentrate into a closed vessel, adding cyanide to the concentrate, and agitating the resulting pulp.

16.

A method according to claim 15 which includes the step of aerating the pulp with oxygen during the agitation.

17.

A method of leaching high grade gold concentrates substantially as herein described with reference to the accompanying drawings.

DATED THIS 22nd DAY OF MARCH , 1978.

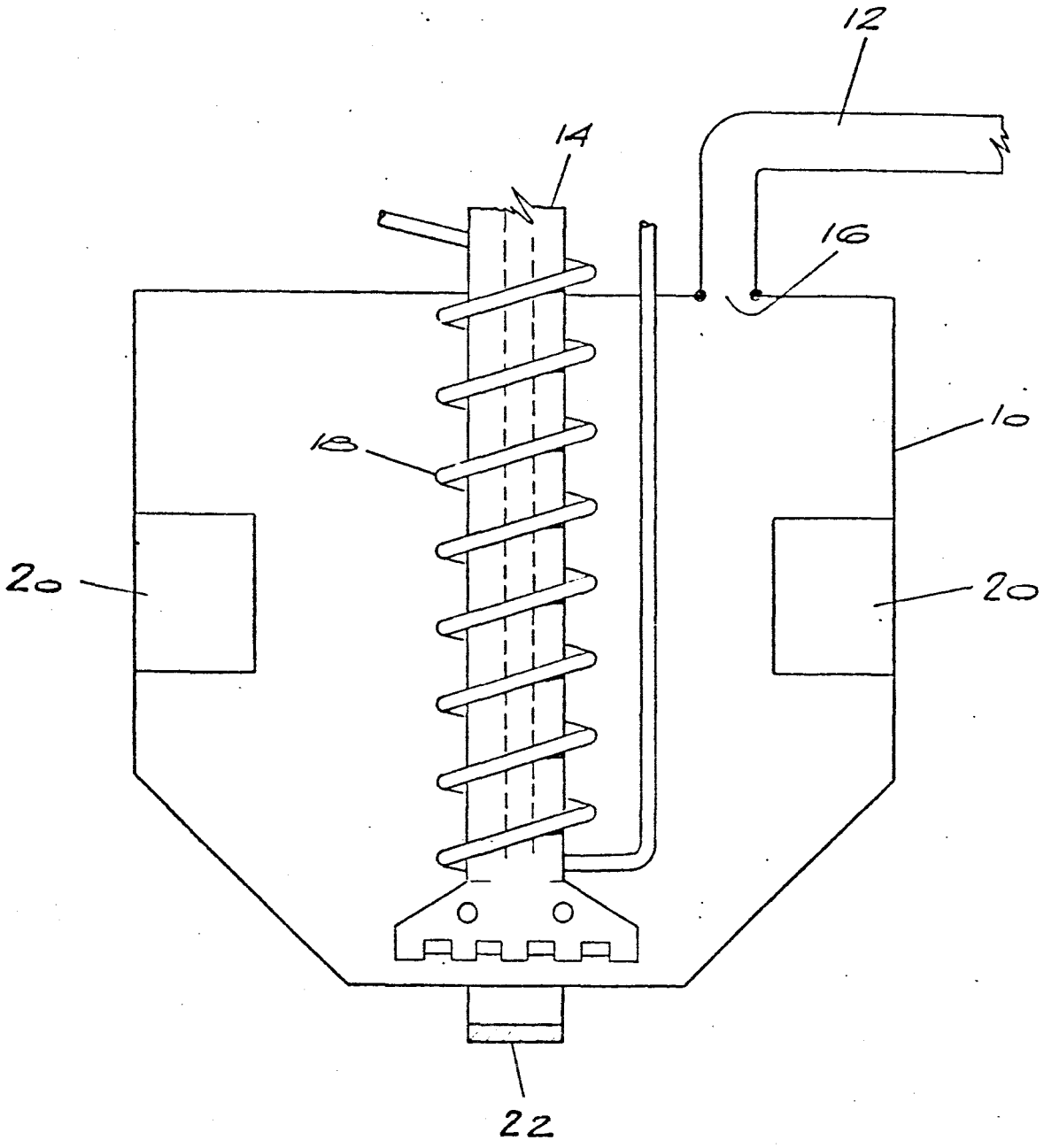
SPOOR AND FISHER

APPLICANT'S PATENT ATTORNEYS

COMPLETE SPECIFICATION NO. 77/1777

ONE SHEET

589.



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APPLICANT'S PATENT ATTORNEY

New gold production method means you can end health and pollution dangers

590.

And it costs no more with intensive cyanidation

Most of the developmental work in intensive cyanidation was carried out in South Africa. The Dome operation in Timmins will comprise three basic steps – intensive cyanidation in covered vessels, liquid/solid separation on a belt filter, and recovery of the gold by electrowinning.

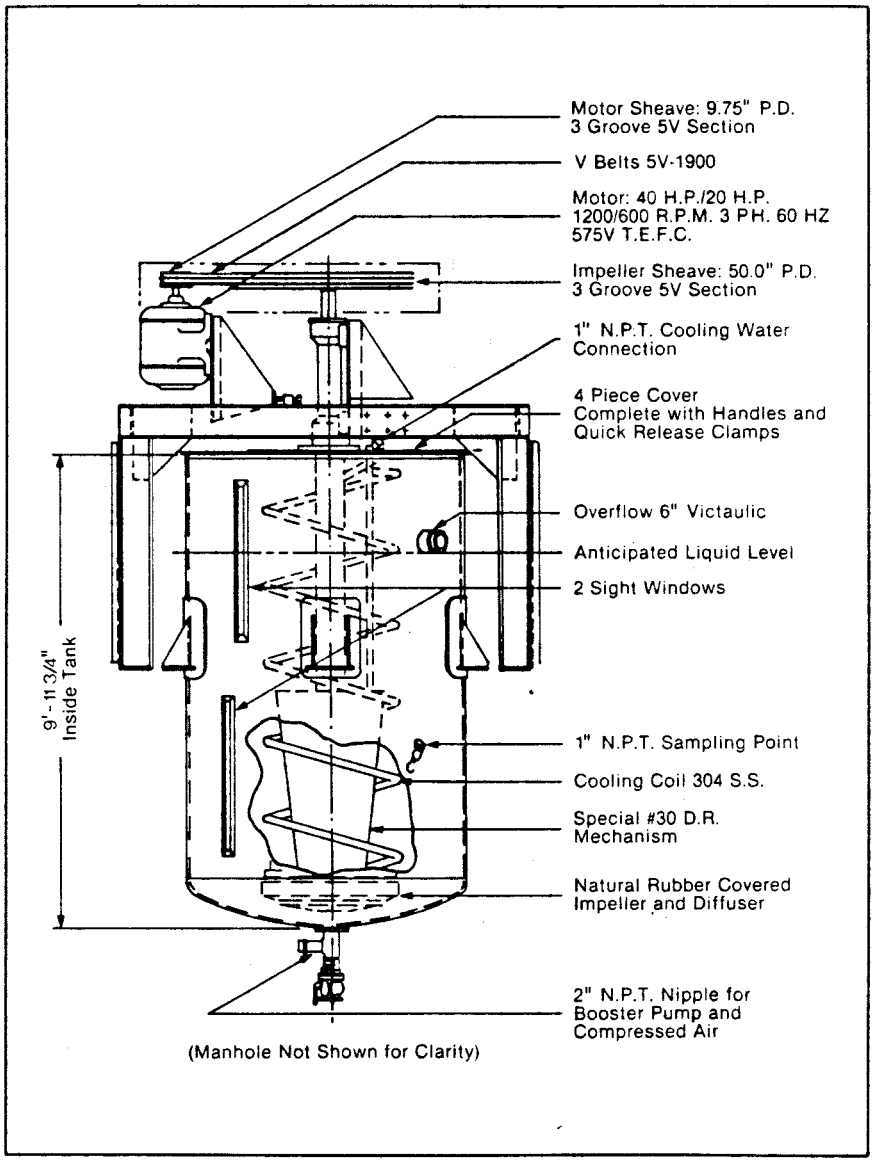
New facilities being constructed by Dome Mines in Timmins will permit the recovery of gold by a new safer method. The technique removes the health and pollution problems associated with mercury amalgamation.

And production will cost no more than traditional methods, says Ken Thomas of Toronto consultants Kilborn Ltd.

Intensive cyanidation, as the process is called, is the batch treatment of gravity concentrate in covered, mechanically agitated vessels. The solution is processed for 18 hours at pH 11-12 and ambient temperature in a sodium cyanide solution concentration of 0.5 per cent to 1.0 per cent. Thereafter the enriched gold solution is separated from denuded concentrate by either filtration or washing in a cone with barren solution.

Reactor vessels

The intensive cyanidation full scale plant will comprise three major processing steps. First, there will be intensive cyanidation in covered vessels followed by liquid/solid separation on a belt filter, and thereafter recovery of gold onto steel wool cathodes by electrowinning. Concentrate is pumped to one of two 1.8 m diameter by 3 m high covered leach reactor vessels for intensive cyanidation. The second of these vessels serves as a storage vessel for filtered solids and as a standby reactor. It is anticipated that 3.5 to 5.5 t of concentrate wil be treated per batch. The tank is constructed of 0.64 cm welded steel plate with a dished bottom lined with 0.64 cm thick cold bonded natural rubber. An additional 0.64 cm thick rubber liner is added in areas of high abrasion around the tank bottom.



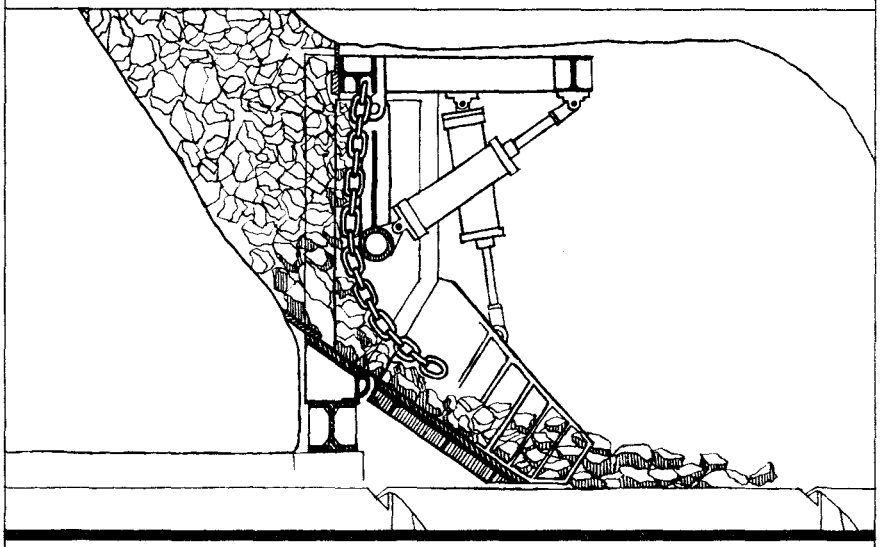
Central to the Dome Mines intensive cyanidation process is the leach reactor shown here. The technical feasibility of the process was established by the Ontario Research Foundation and Dawson Metallurgical Labs of Murray, Utah.

Baffle plates

Tank accessories include a sampling port, over-flow line and a manhole for feeding bulk lime and cyanide. Three 15 cm wide removable baffle plates are also incorporated and extend the height of the tank. The tank cover, split into four pieces for easy removal, is clamped to the reactor. Two sight windows are located on the tank wall for operator convenience. The tank cover has connections for pulp feed, pulp recirculation, a pH probe and a temperature probe.

A 10 cm vent is also provided for connecting to the centrifugal venting fan. The flotation type agitator mechanism is a suspended Denver No. 30 D-R, with a combination pulp recirculation well and air pipe, 69 cm diameter natural rubber covered cast iron skeleton 9 bladed receded disc impeller and diffuser. Sheaves are sized to turn the impeller at 115 rpm for startup and 230 rpm for leaching.

Versatile Chutes as proven in Sweden's Kiruna Mine.



Air-powered Dual Action Chutes are now available in four sizes and provide flexibility for Side Loading and In-line Loading of mine cars or trucks. They may be direct air operated from the chute loading gallery or drift, with local remote pendant servo control or distant remote control utilizing closed circuit television.

The unique "variable throat" cylinder actuated "chain curtain" provides maximum control over fines which tend to flush, yet allows big muck to pass safely.

The active "lip" extends down into the car body allowing "soft loading". This eliminates spillage and reduces car abuse and maintenance problems.

Simultaneous operation of the lip and throat cylinders provides excellent coordinated control of all "run of mine" muck.

Access is available through the chain curtain for chute blasting and provides for maximum personnel safety.

The standardized design using, structural steel members, provides ease of installation with a minimum of preparatory work. Component sub assemblies of reasonable size and weight may be transported and mounted with ease.

All components may be removed and used at new locations where rapid mining cycles dictate.

Wear lining using standardized shapes assures long life and good resistance to chute blasting. All mechanisms are top mounted allowing serviceability without interference with production cycles.

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Dissipate heat

Integral with the mechanism is a 2.5 cm diameter, 304 stainless steel cooling coil, which runs the length of the mechanism, to dissipate heat generated by the mechanism. A discharge port and valve are located on the tank bottom, and a flush point is incorporated for flushing with air or water prior to transferring to the belt filter.

After testing a 2 m^2 belt filter was selected. Early filtration testwork indicated extremely poor filtration rates, but this problem was overcome by reducing the tip speed of the impeller in the flotation machine to 290 mpmin., found to be just sufficient for solids suspension.

Steel wool

The solution is transferred from the gold solution tank to a 1.4 m^3 electrowinning cell manufactured by Custom Equipment Corporation of Salt Lake City. The cell contains 0.7 m^3 of steel wool, arranged as seven cathode baskets between eight anodes. The cathodes draw a total of 250 amps.

The results of electrowinning testwork, conducted at Dawson Metallurgical of Dawson, Utah, showed that more than 98 per cent of the gold and silver can be deposited on a steel wool cathode with little or no base metal co-deposition. Retention time required in the electrowinning cell was between 50 to 60 minutes to achieve the target barren of not more than 5 ppm Au.

About 60 amps is effectively used for deposition, gold being deposited at 7 g per ampere hour. To otbain 3.4 kg production of gold per day will require the cell being on line 9 hours. Baffles will be incorporated in the cell in an attempt to give a loading profile on the cathodes. It is estimated between 6800 to 9000 l of gold solution will be treated through the cell, equivalent to a solution flowrate of 13.5 l per minute. Testwork also indicated a cathode residence time of 63 minutes, equivalent to a solution flowrate of 13.5 l per minute.

Labor requirements for the two processes, amalgamation and intensive cyanidation, are considered comparable, one man per day in full attendance during day shift. Capital costs for the two processes are also similar at $400,000, excluding common items such as buildings and access ways.

The intensive cyanidation operation will require one full-shift per day, comparable to the mercury amalgamation employment. At $400,000, the capital cost requirements, exclusive of common items such as buildings and access ways, are also comparable.

2.0 SULPHATION ROASTING

Pilot plant investigations into the sulphation roasting of pyrite concentrates in order to recover base metal values indicated increased recoveries of gold (SA Patent No. 78/0688) and uranium (SA Patent No. 77/0995) when compared with conventional roasting procedures. These findings resulted in full scale trials at Vaal Reefs which were not satisfactorily concluded.

PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	78/0688
Date of Filing:	January 29, 1979
Name of Applicant:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
Name of Inventors:	1. RAYMOND JOHN DAVIDSON 2. NORBERT WALTER HANF 3. CHRISTIAN GHISLAIN SCHMIDT
Title of Invention:	"GOLD EXTRACTION"

~~Provisional~~/*Complete Specification as Filed*

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Brill Bros H 1490

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PATENT ATTORNEYS
JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

594.

FILING DATE:

REPUBLIC OF SOUTH AFRICA

THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 78/0688

TITLE OF INVENTION: "GOLD EXTRACTION"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street, JOHANNESBURG.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

Brill Bros. F 8812

THIS invention relates to the extraction of gold values from pyrite (FeS_2) concentrates.

Pyrite concentrates are obtained, for example, by flotation of pyrite ores. Such concentrates contain at least 20% by weight sulphur and often gold, uranium and base metals in addition to iron. This invention is concerned with the improved extraction of gold values from gold-bearing pyrite concentrates.

According to the present invention, a process for extracting gold values from a gold-bearing pyrite concentrate includes the steps of roasting the concentrate at a temperature in the temperature range 600°C to 750°C and leaching the gold values, for example by means of a cyanide leach, from the roasted concentrate. The roasting converts sulphur values in the concentrate to sulphur dioxide.

It has surprisingly been found that roasting of the concentrate in the temperature range specified above improves significantly the extraction of gold values from the concentrate. The preferred temperature range is 650 to 750°C, typically 650 to 700°C.

The roast may be carried out in the presence of an alkali metal salt or hydroxide which further improves the extraction of gold values from the concentrate.

The roast is preferably carried out in a standard fluidised bed roaster such as those illustrated in the article of R.B. Thompson and R. Roesner "Fluid Bed Roasting - Principles and Practice" in "Extractive Metallurgy of Copper, Nickel and Cobalt" by Paul Queneau, Interscience 1961.

The preferred alkali metal is sodium and the preferred salt is the sulphate. Only small quantities of the salt or hydroxide are required to improve the gold extraction. The salt or hydroxide is typically used in an amount of 1 to 10%, preferably 1 to 5%, by weight of the concentrate.

The residence time in the roaster will vary according to the nature of the concentrate. In general, the longer the residence time, the better the final extraction.

The pyrite concentrate is generally fed to the roasting step either as a dry mass or as a slurry. When the concentrate is in the form of a slurry, the water content will generally be less than 50% by weight of the slurry.

The pyrite concentrate will typically have a sulphur content of 20 to 50%, generally 30 to 40%, by weight.

If the roasted concentrate contains significant amounts of uranium in addition to the gold values, it is desirable first to leach the uranium values from the roasted concentrate prior to the gold leaching step. Leaching of these uranium values can readily be achieved by means of an acidic leach, typically a sulphuric acid leach at a pH not exceeding 3. The residue from the acid leach may be subjected to the conventional cyanide leach for gold dissolution. The cyanide leach, as is known in the art, involves leaching with a cyanide, typically a sodium cyanide, leach solution of pH 10 to 12. An excess, e.g. 0,2g/l excess, of cyanide is usually used. The gold values go into solution and may be recovered therefrom by known techniques such as precipitation with zinc or absorption onto activated carbon.

The invention will be illustrated by means of the following example. A pyrite concentrate, obtained from the flotation of a pyrite ore, was found on analysis to contain 600 grams of uranium per ton, calculated as U_3O_8, 15 grams of gold per ton, and a sulphur content of 35% by weight.

The pyrite concentrate was roasted in a standard fluidised bed roaster at various temperatures for an average period of 3 hours (a) in the

absence of sodium sulphate and (b) in the presence of sodium sulphate. The roasted concentrate was subjected to a sulphuric acid leach at pH 3 for 60 minutes to leach uranium values from the concentrate. The residue from the uranium leach was subjected to a standard sodium cyanide leach at pH 10 - 12 to recover the gold values. The results obtained were as follows:

Roasting Temperature	$NaSO_4$%	Gold Recovery %
820	0	83
665	0	87
680	2	89
690	4	89
660	6	88

The effect of roasting temperature together with that of the addition of sodium sulphate is further exemplified in the Table below. Again a standard fluidised bed roaster was used in the experiments and the gold leach was a standard sodium cyanide leach at pH 10 - 12. The leach time was 16 hours.

Test	Roasting temp. (°C)	Mean Roasting Time (h)	Na_2SO_4 Addition (%)	Au Extraction (%)
1	690	2	0	87,1
1	690	2	1	88,3
1	690	2	2	89,6
2	850	4	0	75,6
2	650	4	0	86,2
2	650	4	2	89,3
3	820	2	0	83,2
3	690	2	0	87,4
3	690	2	4	88,8
3	650	2	2	89,7
4	690	2	0	90,2
4	690	2	2	92,1
4	690	4	2	94,0

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

598.

1.
A process for extracting gold values from a gold-bearing pyrite concentrate including the steps of roasting the concentrate at a temperature in the range 600 to 750°C and leaching the gold values from the roasted concentrate.

2.
A process according to claim 1 wherein the roast is carried out at a temperature in the range 650 to 750°C.

3.
A process according to claim 1 wherein the roast is carried out at a temperature in the range 650 to 700°C.

4.
A process according to claim 1 wherein the roast is carried out at a temperature of 690°C.

5.
A process according to any one of the preceding claims wherein the roast is carried out in the presence of an alkali metal salt or hydroxide.

6.
A process according to any one of claims 1 to 4 wherein the roast is carried out in the presence of an alkali metal sulphate.

7.
A process according to any one of claims 1 to 4 wherein the roast is carried out in the presence of sodium sulphate.

8.
A process according to any one of the preceding claims wherein the pyrite concentrate contains 20 to 50% by weight sulphur.

9.
A process according to any one of claims 1 to 7 wherein the concentrate contains 30 to 40% by weight sulphur.

10.
A process according to any one of the preceding claims wherein the concentrate also contains uranium values and these uranium values are leached from the concentrate prior to the gold leach.

11.
A process according to any one of the preceding claims wherein the gold values are leached by means of a cyanide leach.

12.
A process according to claim 1 and substantially as herein described.

13.
Gold values when recovered by a process according to any one of the preceding claims.

DATED THIS 29TH DAY OF JANUARY, 1979.

SPOOR AND FISHER
APPLICANT'S PATENT ATTORNEYS

PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	77/0995
Date of Filing:	18th February, 1977
Name of Applicant:	Anglo American Corporation of South Africa Limited
Name of Inventor:	1. Raymond John Davidson 2. Norbert Walter Hanf 3. Christian Ghislain Schmidt
Title of Invention:	"URANIUM EXTRACTION"

xx~~Provisional~~/*Complete Specification as Filed*

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JOHANNESBURG OFFICE
HYDE PARK CORNER, SANDTON

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 77/0995

TITLE OF INVENTION: "URANIUM EXTRACTION"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street, Johannesburg

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

THIS invention relates to the extraction of uranium values from pyrite (FeS_2) concentrates.

Pyrite concentrates are obtained, for example, by flotation of pyrite ores. Such concentrates contain at least 20% by weight sulphur and often gold, uranium and base metals in addition to iron. Some pyrite concentrates contain a relatively high uranium content, i.e. a uranium content of at least 50 grams per ton, calculated as U_3O_8. This invention is concerned with the extraction of uranium values from pyrite concentrates which contain at least 50 grams per ton of uranium, calculated as U_3O_8, and more particularly with such concentrates which have a uranium content in the range 50 grams to 2000 grams per ton, again calculated as U_3O_8.

According to the present invention, a process for extracting uranium values from a pyrite concentrate as defined above includes the steps of roasting the concentrate to convert the uranium to uranium sulphate, and solubilising uranium sulphate in the roasted concentrate by means of an acid leach of pH not exceeding 3.

The first step in the process is the roasting of the pyrite concentrate. This roast converts the uranium values in the

concentrate to uranium sulphate. The roast is preferably carried out at a temperature in the range of 650° to 750°C and more preferably in the range 650 to 700°C. Roasting temperatures in excess of 750°C are undesirable because refractories tend to form at these temperatures and this is disadvantageous.

The roast is preferably carried out in a standard fluidised bed roaster such as those illustrated in the article of R.B. Thompson and R. Roesner "Fluid Bed Roasting - Principles and Practice" in "Extractive Metallurgy of Copper, Nickel and Cobalt" by Paul Queneau, Interscience 1961.

The extraction of uranium values from the pyrite concentrate is improved by carrying out the roast in the presence of a suitable alkali metal salt such as sulphate or carbonate or alkali metal hydroxide. The preferred alkali metal is sodium and the preferred salt is the sulphate. Only small quantities of the salt or hydroxide are required to improve the extraction. The salt or hydroxide is typically used in an amount of 1 to 10%, preferably 1 to 5%, by weight of the concentrate.

The residence time in the roaster will vary according to the nature of the concentrate and the degree of conversion of uranium to uranium sulphate required. In general, the longer the residence time, the better the final extraction.

The uranium sulphate is extracted from the roasted concentrate by means of an acid leach of pH not exceeding 3. The pH of the acid leach is typically in the range 1 to 3. The preferred acid leach is sulphuric acid. The leach time can vary, but is typically 20 - 120 minutes.

The pyrite concentrate is generally fed to the roasting step either as a dry mass or as a slurry. When the concentrate is in the form of a slurry, the water content will generally be less than 50% by weight of the slurry.

The pyrite concentrate will typically have a sulphur content of 20 to 50%, generally 30 to 40%, by weight.

Any base metals such as copper, cobolt and nickel present in the pyrite concentrate are simultaneously extracted with the uranium values. The uranium values from the uranium rich leach solution may be recovered by methods known in the art such as solvent extraction or iron exchange.

The process of the invention provides a very effective

extraction of uranium values from pyrite concentrates and a far better extraction than has been achieved in the past using a strong acidic leach on its own.

The invention is illustrated by means of the following examples.

EXAMPLE 1

A pyrite concentrate, obtained from the flotation of a pryite ore was found on analysis to contain 600 grams per ton of uranium, calculated as U_3O_8, and a sulphur content of 35% by weight.

The pyrite concentrate was roasted in a standard fluidised bed roaster at 650°C for an average period of 3 hours in the presence of sodium sulphate. The sodium sulphate was present in an amount of 2% by weight of the concentrate. The roasted concentrate was then subjected to a sulphuric acid leach of pH3 for 60 minutes.

Analysis of the leach solution showed a uranium extraction in excess of 90%.

EXAMPLE 2

A pyrite concentrate was found on analysis to contain 41% by weight sulphur and 600 grams per ton of uranium, calculated as U_3O_8.

This concentrate was subjected to the same procedure as Example 1, save that in a first experiment a roast temperature of 680°C in the absence of sodium sulphate was used and in a second experiment a temperature of 690°C in the presence of 4% by weight sodium sulphate was used. The leach time in each case was 20 minutes. The results obtained were as follows:

Roast Temperature	Na_2SO_4 %	% U Extraction
680	0	25
690	4	91

EXAMPLE 3

A pyrite concentrate was found on analysis to contain 24% by weight sulphur and 300 grams per ton of uranium, calculated as U_3O_8.

This concentrate was subjected to the same procedure as in Example 1, save that the temperatures and sodium sulphate contents as set out below and a sulphuric acid leach solution of pH 1,1 were used:

Roast Temperature	Na_2SO_4 %	% U Extraction
665	0	41
680	2	48
680	4	73
660	6	84

HAVING HOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS :

1.

A process for extracting uranium values from a pyrite concentrate containing at least 20% by weight sulphur and at least 50 grams per ton of uranium, calculated as U_3O_8, including the steps of roasting the concentrate to convert the uranium to uranium sulphate, and solubilising the uranium sulphate in the roasted concentrate by means of an acid leach solution of pH not exceeding 3.

2.

A process according to claim 1 wherein the roast is carried out at a temperature in the range 650 to 750°C.

3.

A process according to claim 1 wherein the roast is carried out at a temperature in the range 650 to 700°C.

4.

A process according to any one of the preceding claims wherein the pyrite concentrate contains 20 to 50% by weight sulphur.

5.

A process according to any one of claims 1 to 3 wherein the pyrite concentrate contains 30 to 40% by weight sulphur.

6.

A process according to any one of the preceding claims wherein the pyrite concentrate contains uranium calculated at U_3O_8, in the range of 50 grams to 2000 grams per ton.

7.

A process according to any one of the preceding claims wherein the pH of the leach solution is in the range 1 to 3.

8.

A process according to any one of the preceding claims wherein the leach solution is a sulphuric acid leach.

9.

A process according to claim 1 and substantially as herein described with reference to any one of the Examples.

10.

Uranium values extracted from a pyrite concentrate by a process of any one of the preceding claims.

DATED THE 8th day of FEBRUARY, 1978

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APPLICANT's PATENT ATTORNEYS

3.0 DUAL GOLD AND URANIUM LEACH

According to the appended invention (SA Patent No. 77/1778), a process for extracting gold and uranium from an ore contaning them both comprises the steps of pulping the finely comminuted ore with a suitable cyanide solution at an alkaline pH, acidifying the pulp for uranium dissolution, adding activated carbon for gold recovery to the pulp, and recovering gold from the activated carbon and uranium from solution.

PATENT SPECIFICATION

Country:	SOUTH AFRICA
Application Number:	77/1778
Date of Filing:	20th March, 1978.
Name of Applicant:	ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED
Name of Inventor:	(1) GEORGE STEPHEN JAMES (2) RAYMOND JOHN DAVIDSON
Title of Invention:	"GOLD AND URANIUM EXTRACTION"

~~Provisional~~/Complete Specification as Filed

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HYDE PARK CORNER, SANDTON

611.

FILING DATE:

REPUBLIC OF SOUTH AFRICA
THE PATENTS ACT 1952

COMPLETE SPECIFICATION

APPLICATION No.: 77/1778

TITLE OF INVENTION: "GOLD AND URANIUM EXTRACTION"

NAME OF APPLICANT: ANGLO AMERICAN CORPORATION OF SOUTH AFRICA LIMITED

ADDRESS: 44 Main Street, Johannesburg.

I/WE DO HEREBY DECLARE THIS INVENTION, THE MANNER IN WHICH AND THE METHOD BY WHICH IT IS TO BE PERFORMED, TO BE PARTICULARLY DESCRIBED AND ASCERTAINED IN AND BY THE FOLLOWING STATEMENT:

Field of the Invention.

THIS invention relates to the extraction of gold and uranium values from ores containing both of them.

The term "ore" is intended in the context of this specification to embrace, besides conventional ores, other materials such as imcompletely extracted tailings, concentrates and other materials which contain both gold and uranium.

Background of the Invention.

Many gold mines in South Africa produce an ore which contains both gold and uranium. Where it is considered economical to recover both the gold and the uranium, the finely ground ore is pulped with a leach solution. In many cases the ore is first leached with a cyanide solution, the gold-bearing solution is filtered from the remaining solids, the solution is clarified and subjected to the conventional Crowe-Merrill for the recovery of gold, and the remaining solids are subsequently treated for uranium recovery. In this subsequent process the solids are leached in an acid solution, the uranium-bearing solution is filtered from the remaining solids and the clear solution treated with either an ion exchange process or a solvent extraction process for the recovery of uranium. The

solid residue, being acidic, is treated with lime and goes to waste.

In some cases a reverse leach order is practised in that the uranium leach precedes the gold leach.

In general solids-liquid separation, including filtration, takes place between the two leach stages and at the end of the final leach, and it is usual to suffer some loss of the desired values in the solution retained in the solid fraction.

Clearly it would be an advantage to eliminate at least one of the separation steps.

In order to eliminate the separation step in uranium leaching the so-called "resin-in-pulp" process has been devised. The variants of this process have one thing in common: resin for adsorbing uranium values is contacted with a pulp and not with a clarified solution.

With the advent of activated carbon designed to adsorb gold from cyanide leach solutions "carbon-in-pulp" processes have been proposed and used. Again the carbon for adsorbing the gold is contacted with a pulp and not with a clarified solution.

In known carbon-in-pulp processes pulp at an alkaline pH, which has been chosen for optimum gold solubilization, is contacted with suitably activated carbon.

It has also been proposed to combine a conventional alkaline cyanidation process for the extraction of gold with an alkaline carbonate leach of the extraction of uranium. This process has met with problems amongst them that sulphides have to be floated off and these may carry gold values with them. Also the optimum temperature conditions for gold and uranium leaching are not the same and clarification is still an essential step in the process.

If one wished to eliminate costly and frequently wasteful clarification in gold and uranium leaching a possible step would be to combine the existing resin-in-pulp and carbon-in-pulp processes into a single continuous process. Bearing in mind the requirement that the carbon-in-pulp process requires alkaline conditions, the obvious link-up would be an acid leach with resin-in-pulp recovery of uranium combined with an alkaline cyanide leach coupled with a carbon-in-pulp process.

The reverse process of an alkaline gold leach followed by an acid leach would not have been followed because of a widely-held belief that gold would precipitate out of solution

upon acidification of a cyanide pulp.

The present invention is predicated by the discovery that contrary to acceptec belief gold is not readily precipitated on acidification of dilute cyanide solutions. In other words it is feasible to dissolve gold and uranium in a continuous leaching operation which starts off highly alkaline and ends up at an acid pH for optimum uranium dissolution.

Summary of the Invention.

According to the invention a process for extracting gold and uranium from an ore containing them both comprises the steps of pulping the finely comminuted ore with a suitable cyanide solution at an alkaline pH, acidifying the pulp for uranium dissolution, adding carbon activated for gold recovery to the pulp at a suitable stage, separating the loaded actived carbon from the pulp, and recovering gold from the activated carbon and uranium from solution.

The activated carbon may be added before or at the start of acidification and it may be in a granular form which may be separated from the pulp by screening.

The invention also provides that the uranium may be recovered by a resin-in-pulp process. In this event the resin may be added to

the pulp during or after acidification and the gold bearing activated carbon and uranium bearing resin recovered separately or together from the pulp. One method of recovering them is by screening and then separating them in a suitable manner.

Description of the Drawing.

The drawing illustrates a proposed flow-sheet according to the invention.

Description of Preferred Embodiments.

EXAMPLE 1.

A sample of high grade ore (10,8 g/t Au; 0,325 kg/t U_3O_8) was pulped with a solution containing 0,30 kg/t NaCN to a liquids:solids ratio of 1:1. CaO was added in an amount of 1 kg/t.

During cyanidation the sample was agitated and aerated. The agitation was continued for 17 hours at ambient temperature after which it was found that 94,2% of the original gold had dissolved (equivalent to 10,17 g/t).

At this point 100 g/t of granular activated carbon was added to the pulp. The pulp temperature was raised to 60°C and

the pH was lowered to 1,4 using 1,5 kg/t H_2SO_4. When the pH was measured as being 1,6 and after 30 min. 5 kg/t of MnO_2 was added and the pH maintained at 1,8 for a further 6 hours.

The activated carbon was then separated from the pulp by a simple screening process. Analysis of the residue indicated that 99,6% of the gold originally dissolved during cyanidation was now absorbed on to the activated carbon. This is equal to a final gold recovery of 93,8% from this particular ore and means that the process results in negligible gold loss due to acidification.

An ion exchange resin in the form of granules of Amberlite IRA 400 was then added to the pulp in an amount of 100 g/t and after 1 hour the granules were screened off. It was found that 97% of the uranium in solution could be recovered in this way. This is equivalent to a total uranium recovery of 75,4%.

A paralled test without the addition of the granules of activated carbon showed that at the end of the acid leaching stage only 83% of the total gold was in solution.

Note that no attempt was made to optimize uranium leaching and that such optimization could lead to improved uranium recoveries in the above described example.

EXAMPLE 2.

A sample of tailings from a slimes dam was treated.

The sample analysed:

Au 1,16 g/t

Ag 0,09 g/t

U_3O_8 85 g/t

and a moisture content of 15,6%.

A sample of 400 g was leached for 3 hours with 200 g/t NaCN at pH 8 in the presence of 3 g activated carbon (grade G210) using a liquids/solids ratio of 1:1. After the three hours in a rolling bottle the pH was dropped to 2,0 by the addition of 8,9 ml of 10% (V/V) H_2SO_4. Also 6 g of Amberlite IRA 400 ion exchange resin was added to the bottle which was rolled for 1 hour..

The carbon and the resin were then separated from the pulp by screening and from each other by hand sorting, dried at 120°C and the carbon analysed for gold. The solution was filtered off and analysed for gold and uranium. The washed filter cake was dried at 120°C and analysed.

The filter cake analysed 18 g/t of U_3O_8 and 0,74 g/t of Au.

The carbon had adsorbed on to it Au equivalent to 0,44 g/t of the gold in the original sample.

The filtrate solution had a final pH of 2,6 and analysed as follows:

U_3O_8 1,5 ppm
Fe 210 ppm
Ni 160 ppm
Co 55 ppm
Cu 35 ppm
Au less than 0,01 ppm.

In other words practically all the gold dissolved in this process was adsorbed by the activated carbon. It is estimated that 97,7% of the dissolved uranium was recovered.

FLOWSHEET.

The attached drawing shows one proposed simplified flow sheet for a gold and uranium recovery plant. No cognisance is taken of the recovery of pyrite or any other desirable material.

In the drawing milled ore (or slimes from a slimes dam) is first pulped and passed into a suitable vessel, such as a pachuka with the addition of the requisite amounts of NaCN and lime.

After an adequate cyanidation period, the pulp is transferred to a rubber-lined pachuca for acid leaching. Activated carbon is added to the pulp at this stage. At an appropriate pH resin beads are also added to the pulp.

After a suitable leaching time the loaded carbon and resin are recovered from the pulp. These two may then be separated.

The gold and uranium are then recovered by means of any suitable process.

It is possible that there may be an advantage in some cases in adding the activated carbon to the first pachuca at some stage of cyanidation. Also the resin may be added at the start of acidification.

during leaching

HAVING NOW IN DETAIL DESCRIBED AND ASCERTAINED OUR SAID INVENTION AND THE MANNER IN WHICH THE SAME IS TO BE PERFORMED, WE DECLARE THAT WHAT WE CLAIM IS:

1.

A process for extracting gold and uranium from an ore containing them both comprising the steps of pulping the finely comminuted ore with a suitable cyanide solution at an alkaline pH, acidifying the pulp for uranium dissolution, adding carbon activated for gold recovery to the pulp at a suitable stage, separating the loaded activated carbon from the pulp, and recovering gold from the activated carbon and uranium from solution.

2.

The process claimed in claim 1 in which the activated carbon is added at the start of acidification.

3.

The process claimed in claim 1 in which the activated carbon is added during cyanidation.

4.

The process claimed in any of the above claims in which the uranium is recovered in a resin-in-pulp process.

5.

The process claimed in claim 4 in which the resin is added to the pulp during acidification and the activated carbon and resin are r covered separately or together from the pulp.

6.

The process claimed in claim 5 in which activated carbon and resin are recovered from the pulp together and subsequently separated.

7.

A process for extracting gold and uranium from an ore containing them both substantially as herein described with reference to either one of the examples.

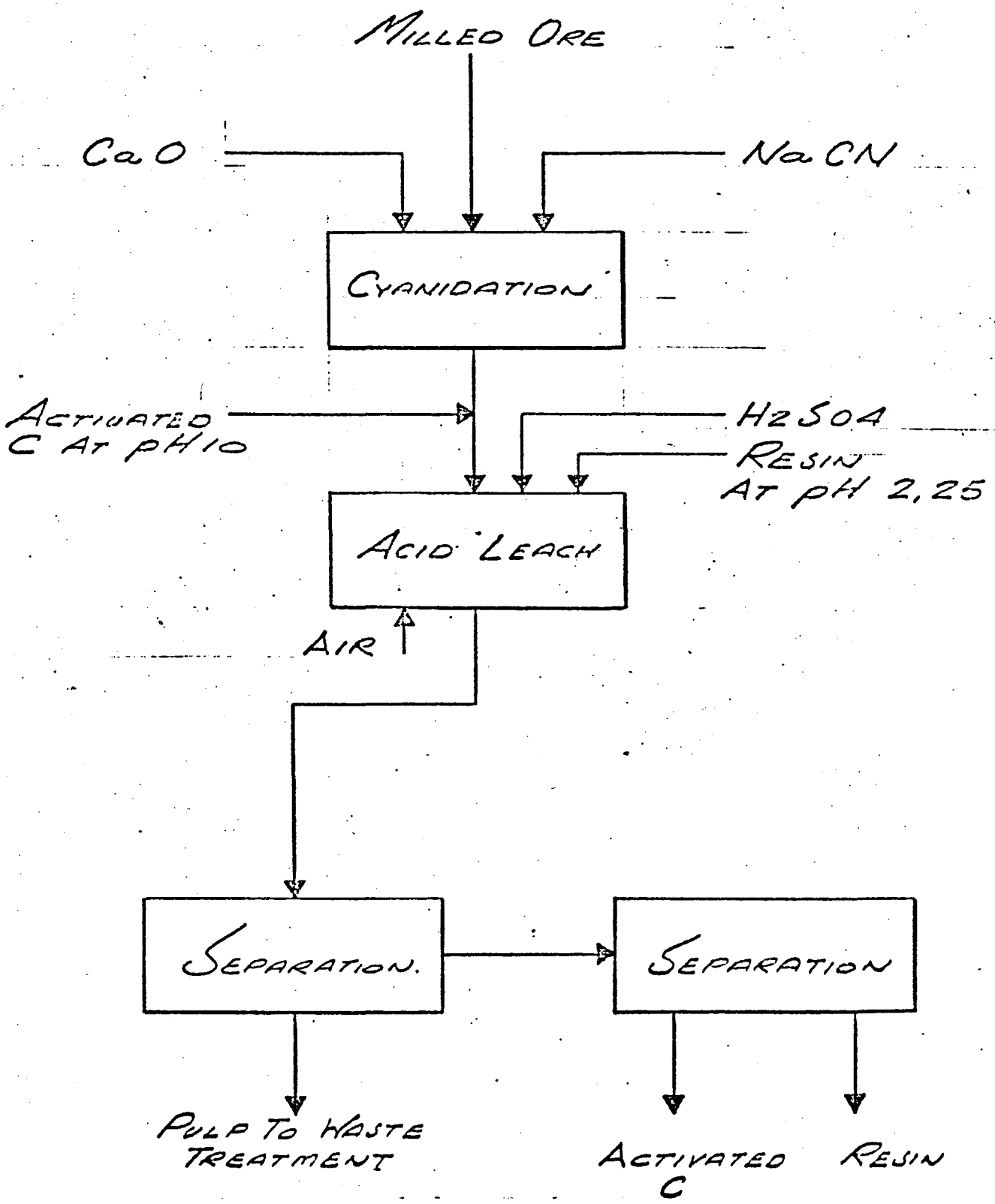
9.

A process for extracting gold and uranium from an ore containing them both substantially as herein described with reference to the accompanying flow sheet..

DATED THIS 20th DAY OF March, 1978.

SPOOR AND FISHER

APPLICANT'S PATENT ATTORNEYS


MILLED ORE
CaO
NaCN
CYANIDATION
ACTIVATED C AT pH 10
H2SO4
RESIN AT pH 2,25
ACID LEACH
AIR
SEPARATION.
SEPARATION
PULP TO WASTE TREATMENT
ACTIVATED C
RESIN