

**COMPARISON OF PROCESS ALTERNATIVES FOR
GOLD RECOVERY FROM CYANIDE LEACH SOLUTIONS**

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ABSTRACT

Both conventional and nonconventional process alternatives for gold recovery from cyanide heap leach solutions are discussed in terms of possible flowsheets. Conventional process alternatives include carbon adsorption and zinc dust precipitation, while nonconventional process alternatives include resin ion exchange, solvent extraction, and direct electrowinning. These strategies for gold recovery are compared on the basis of technological features and on the basis of process economics. Review of the technology status and process economics suggests that improvements for conventional process alternatives are possible and that further development of nonconventional process alternatives is warranted.

INTRODUCTION

Cyanidation of gold and silver ores by conventional mining and milling techniques has been practiced for over a century. During the past decade, heap leaching has emerged as a viable process for treating certain gold and silver ores (Chamberlain and Pojar, 1981). The rapid development of heap leaching for gold and silver ores is largely a result of lower capital and operating costs and faster start-up time than that offered by conventional mining and milling. In general, these features make heap leaching an ideal approach for exploiting either small or large low-grade disseminated deposits of gold and silver considered to be uneconomic by conventional methods. The development of heap leaching technology has evolved from the operations initiated in the western United States.

The basic components of a typical heap leaching operation are depicted in Figure 1. Run-of-mine or crushed ore (usually 1 to 5 ppm gold) is delivered to specifically prepared impervious pads. Alkaline cyanide solution is applied to the surface of the heap and allowed to percolate through the heap (typically 10-30 ft high). Pregnant leach solution which drains from the heap is collected and stored in tanks of sufficient size or in collection reservoirs. Gold and silver are recovered from the pregnant solution and the barren solution is recirculated back to the heaps after reagent make-up.

Carbon adsorption and zinc precipitation are the principal methods used to recover gold and

silver from heap leach solutions; however, there are several proven alternatives already available and others emerging that show promise. Some of the alternative methods for recovering gold and silver include: resin ion exchange, direct electrowinning, solvent extraction, aluminum precipitation, and sulfide precipitation.

With every recovery option, except carbon and resin adsorption, suspended solids must be eliminated from the process stream. Typical heap leach solutions may contain anywhere from 10 to 300 ppm total suspended solids but can fluctuate widely from these values depending on such factors as weather, ore type, age of the heaps being leached, and ore agglomeration. Therefore, the flowsheet shown in Figure 1 includes a step for clarifying the pregnant liquor. The clarifier should be capable of reducing the TSS concentration in the range of 2-5 ppm.

Even though there has been considerable interest in developing alternatives to cyanidation (i.e., thiosulfate, thiourea, and brines), dilute alkaline cyanide solution remains the preferred lixiviant for leaching gold and silver ores. Table 1 reports the concentrations of selected components in some actual solutions from heap leaching operations using alkaline cyanide.

New developments in heap leaching have dealt with improving the leaching part of the operation and have not examined metal recovery from solution. For example, there have been major attempts to improve the design and construction of heaps, to stabilize slimes (agglomeration) for improved

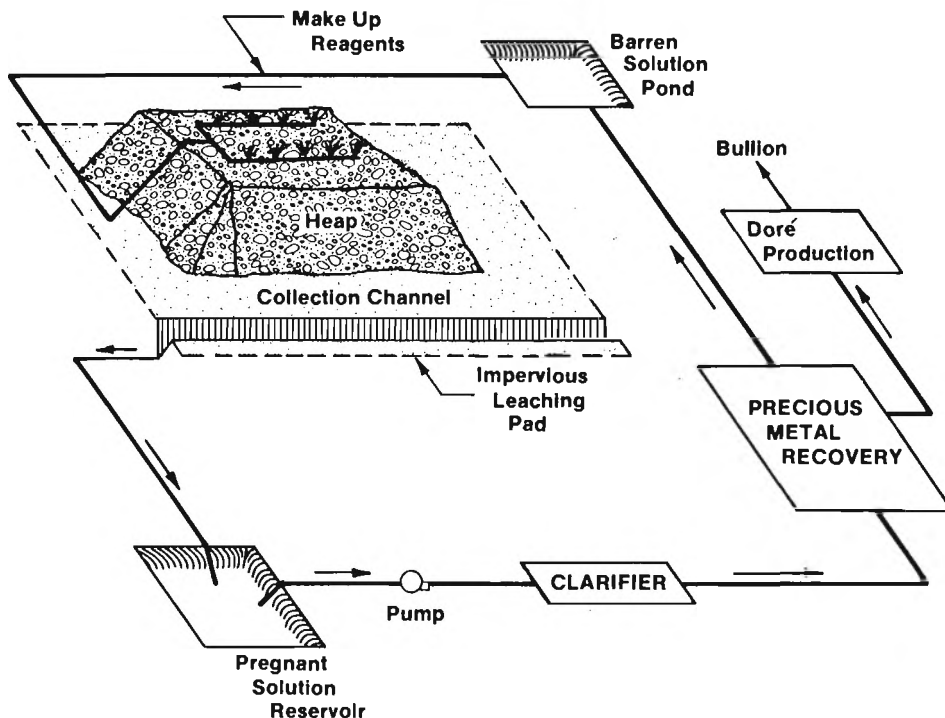


Figure 1. General layout of a heap leaching operation.

HEAP AND DUMP LEACHING PRACTICE

Table 1. Typical Flows and Leach Solution Composition for Heap Leaching Operations

Plant	Flowrate gpm	Concentration							
		Au ppm	Ag ppm	Cu ppm	Hg ppm	Fe ppm	Hardness ppm	CN g/l	pH
A	500	1.0	-	<5	1-8	<1	350	-	10.5
B	1350	1-2	2.1	24.1	<1	<1	313	0.24	9.8
C	500	1.0	1.4	20.0	6.0	-	900	0.75	10.0
D*	35	1.5	0.7	4.1	75.0	0.39	164	0.14	9.3

*Test heap

solution percolation and enhanced gold and silver recovery, and to modify solution chemistry to increase metal extraction. Recently there have been both laboratory and pilot-scale studies directed at improving metal recovery from cyanide solution (i.e., resin ion exchange (Fleming, 1982), solvent extraction (Mooiman and Miller, 1983), and direct electrowinning (HSA Reactors Limited, 1982)).

This paper reviews the technological status of selected recovery options and presents an economic analysis in terms of operating and capital costs of these processes. Promising process alternatives are examined and areas for future research and technological development are identified.

TECHNOLOGY STATUS

Gold recovery from alkaline cyanide solution is typically accomplished by carbon adsorption or zinc cementation (Merrill Crowe). These and other emerging technologies, particularly resin ion exchange, solvent extraction, and direct electrowinning, are discussed, emphasizing the unique features of each process alternative.

Carbon in Column (CIC)

The essence of the carbon-in-column process alternative is the adsorption of gold from alkaline cyanide solution by activated carbon. A concep-

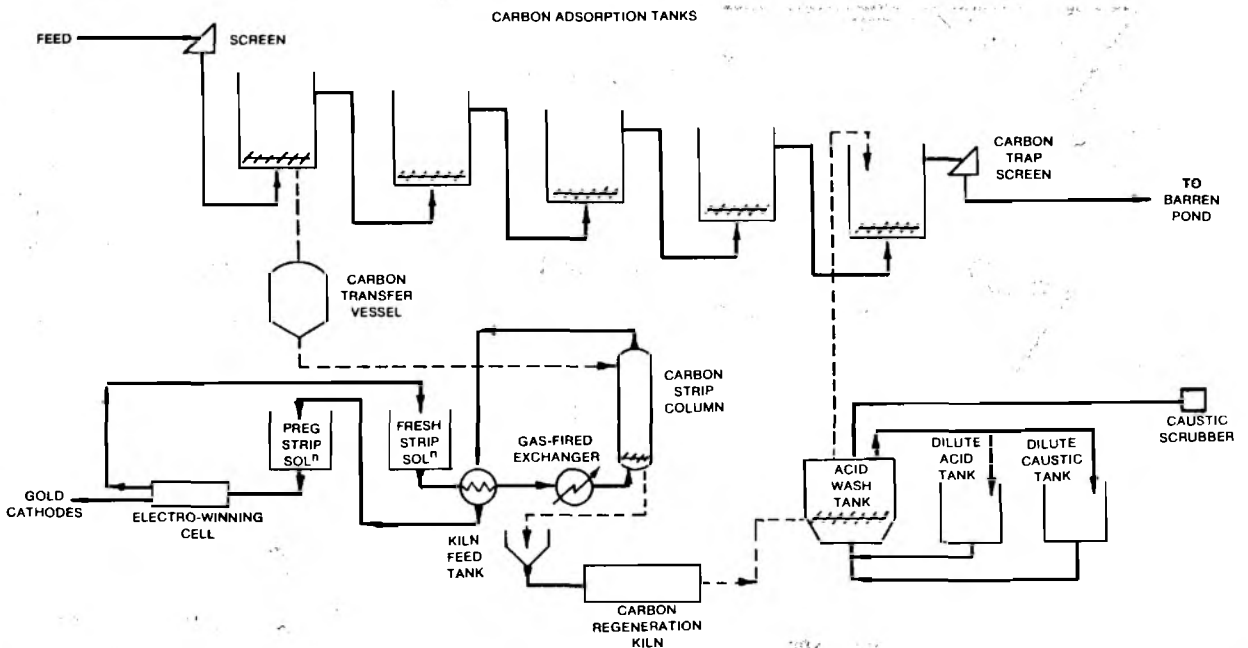


Figure 2. Flowsheet for gold recovery by the carbon-in-column process.

tual flowsheet for this process is presented in Figure 2. For typical carbon-in-column operation, the cyanide leach solution is pumped upward through columns at a flowrate of 15 to 25 gpm/ft², fluidizing the packed bed of activated carbon (16x30 mesh). A carbon-loading level, frequently of 200 oz/ton, is achieved while the carbon is advanced countercurrent to the solution through a staged circuit. The carbon is advanced at a rate necessary to achieve the desired level of loading. Excessive loading generally results in gold losses to the barren solution. The staged circuit could be replaced by a continuous ion exchange contactor, resulting in a substantially smaller plant and easier operation.

The loaded carbon is educted to the elution column and the gold may be stripped by different techniques as suggested in Table 2. A common procedure is to strip with 0.1% NaOH and 0.1% NaCN at 130°C and 2 bed volumes of strip solution per hour. The concentration of the pregnant strip solution is generally suitable for electrowinning of the precious metals.

Activated carbon prepared from coconut shells is preferred for gold recovery from alkaline cyanide solutions. These carbons are distinguished by high surface areas (> 1000 m²/g), a narrow range of pore sizes (10-20 Å) and good mechanical

strength. Carbon regeneration is required to remove organics and restore the intrinsic chemical activity of the carbon. After removal of carbonate deposits by an acid wash (if necessary), the carbon is activated by a thermal treatment in a kiln at 600-800°C.

Complete understanding of the gold/cyanide adsorption and desorption reactions is lacking. High surface areas and pore diffusion are important aspects of the adsorption process. A number of adsorption mechanisms have been suggested to describe the nature of the adsorbed species:

- * Anion Exchange, Au(CN)₂⁻
- * Precipitation of Insoluble AuCN
- * Pore Reduction to Au⁰
- * Cluster Formation

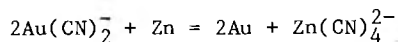
Generally it has been found that the CN/Au ratio of the adsorbed species lies between 2/1 and 1/1 and investigators have been inclined to subscribe to the idea that adsorption involves cluster formation. This proposed mechanism together with different site functionality and pore diffusion transport provides great flexibility in accounting for the behavior of the system -- slow loading rates, some chemical specificity, and severe conditions for stripping.

Table 2. Gold Stripping from Loaded Carbon

Process	Eluant	Temperature	Time	Comments
USBM ⁽¹⁹⁾ Homestake	1% NaOH 0.2% NaCN	90°C	50-60 hr	Slow desorption,
AARL ⁽³⁾ South Africa	1% NaOH 5% NaCN presoak plus water	110°C	6 hr	Autoclave elution under pressure
Alligator ⁽¹⁴⁾ Ridge	1% NaOH 0.1% NaCN	120°C (25 psi)	9 hr	0.5 hr presoak
Smoky ⁽¹⁰⁾ Valley	1% NaOH 0.1% NaCN	88°C	52 hr	1 day presoak
Ortiz ⁽⁷⁾	1% NaOH 0.5% Ethanol	85°C	24 hr	Safety
Battle ⁽⁹⁾ Mountain	20% Ethanol 1% NaOH 0.1% NaCN	77°C	24 hr	Safety
Anglo ⁽¹¹⁾ American	90% acetone ethanol methanol	70-90°C	6 hr	Safety, Reflux solvent, Vapor losses
Murdoch ⁽¹⁷⁾	20-40% organic	25°C	8 hr	Safety Organic losses New technology

Merrill-Crowe Precipitation (MC)

The Merrill-Crowe zinc-dust precipitation process is a mature and well-known technology consisting of solution clarification, deaeration, precious metal precipitation, and precipitate filtration. A conceptual flowsheet for the Merrill-Crowe process is presented in Figure 3. Cementation of gold onto a metallic zinc surface is represented by the following reaction:



A similar reaction can be written for silver precipitation. Historically, gold cementation with zinc shavings was practiced in South Africa as early as 1888 (Clennell, 1915). Zinc-dust precipitation known as the Merrill system was introduced in 1897 in the United States and is the basis of modern practice. T. B. Crowe in 1918 patented the process for removing dissolved oxygen from gold-bearing solution prior to addition of the precipitating agent. This approach improved the efficiency of gold precipitation and decreased zinc consumption. Early cyanide practice involved contacting zinc shavings with a solution of lead acetate to establish a powerful galvanic couple between the finely divided lead deposit and the zinc surface. This procedure provides a substrate capable of recovering gold from solutions low in free cyanide concentration. Lead nitrate or acetate is added along with Merrillite (zinc dust) in many modern recovery plants.

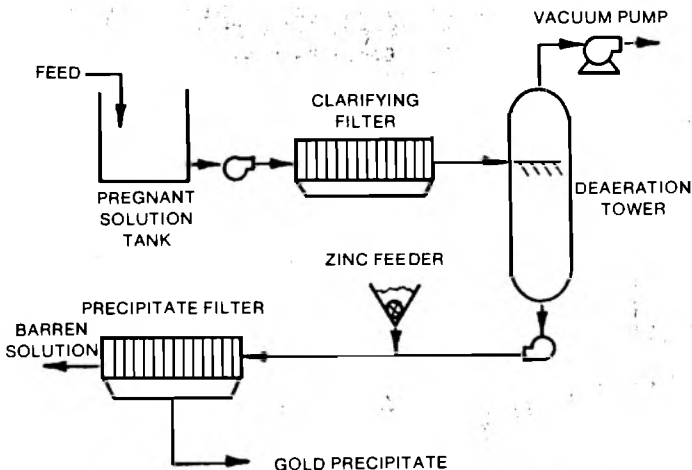


Figure 3. Flowsheet for gold recovery by the Merrill-Crowe process.

Clarified pregnant solution is drawn by vacuum into the Crowe tower to lower the dissolved oxygen concentration to less than 0.5 mg/l. Zinc dust is added at an unspecified rate, far in excess of stoichiometry, to the pregnant solution in such a way as to limit the ingress of atmospheric oxygen. The slurry of precious metal precipitate and

unreacted zinc is pumped directly to a precoat pressure filter system. Zinc dust can be added to the precoat material (i.e., diatomaceous earth) as a safety measure. Typical precipitate analyses vary between 10% and 40% Au. It is clear from the high zinc consumption that further research effort is warranted which could result in a substantial reduction of operating costs.

The grade of the precipitate can be increased by either consuming excess zinc in the filter cake by washing with pregnant solution while the zinc feeder is off or dissolving excess zinc with sulfuric acid after removing the precipitate from the filter press. The precious metal filter cake is then smelted into bars.

Silver is often recovered by zinc precipitation instead of carbon adsorption because of the high silver concentration and the poor silver-loading characteristics of carbon.

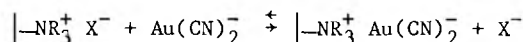
Resin In Column (RIC)

Ion-exchange resins for the recovery of gold from cyanide solution is an emerging technology which is receiving considerable attention. The flowsheet for the resin-in-column plant is similar to that for a carbon-in-column (CIC) plant. See Figure 4. The loading section is virtually identical; however, the elution (stripping) section is different as elevated temperatures and pressures are not required as in carbon stripping. Pressurized elution vessels and heat exchangers are therefore unnecessary. Again the continuous ion exchange contactors can be considered for resin-in-column as they were as in the CIC case.

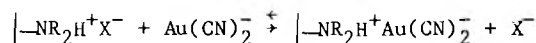
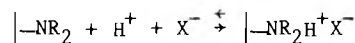
Even though the exchange kinetics are faster than the kinetics of adsorption by carbon, it does not necessarily mean that less resin and smaller loading columns can be used. To the contrary, as resin beads are smaller and generally less dense than carbon granules, slower flowrates (typically 6 gpm/ft² compared to 16 gpm/ft² for carbon) are required to prevent the resin bed from being inadvertently transported to the next stage. In other words, flowrate limitations rather than exchange kinetics determine the reactor size. This problem could be solved if dense and/or large resin beads were available. Research to this end is in progress, and the results to date are very promising (Hendriksz, 1983).

Strong or weak base resins can be used, and the loading reactions for these resins can be described as follows, where $[-]$ represents the polymeric resin matrix.

Strong Base



Weak Base



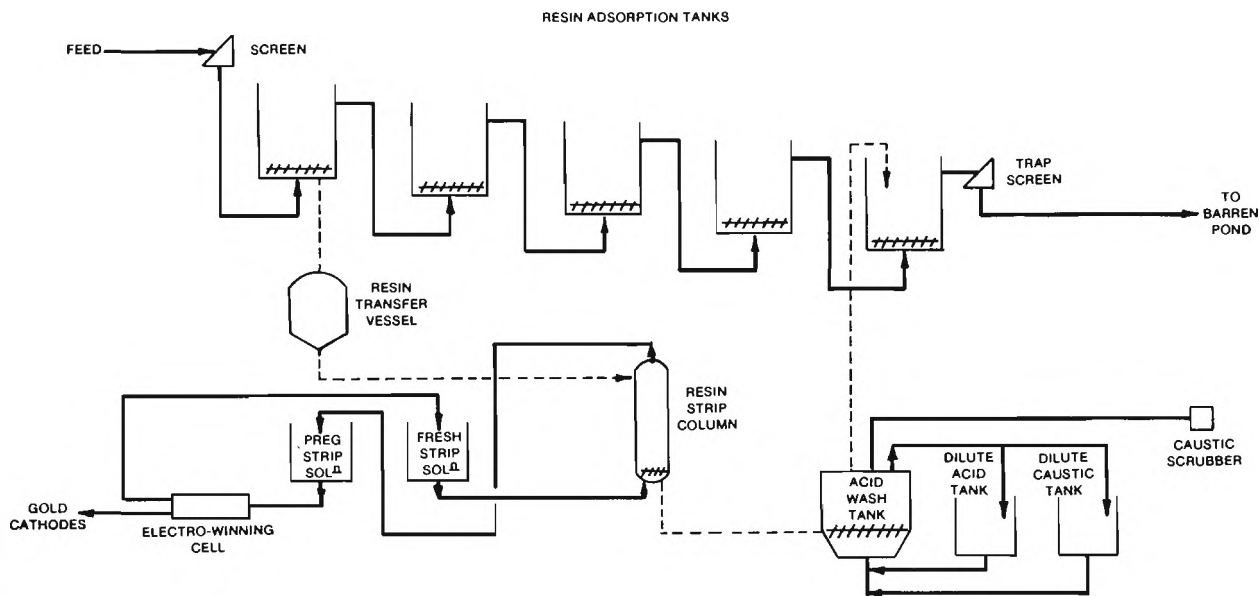


Figure 4. Conceptual flowsheet for gold recovery by the resin-in-column process.

Similar reactions occur for the extraction of silver from cyanide solutions. The weak base resins require protonation in order to extract anions. This occurs for pH values of less than 10, and, therefore, depending on the resin, some pH adjustment of the pregnant leach liquor might be required in order to obtain acceptable loadings on weak base resins.

Generally, the pH performance of these amine-type resins can be described by the plot shown in Figure 5.

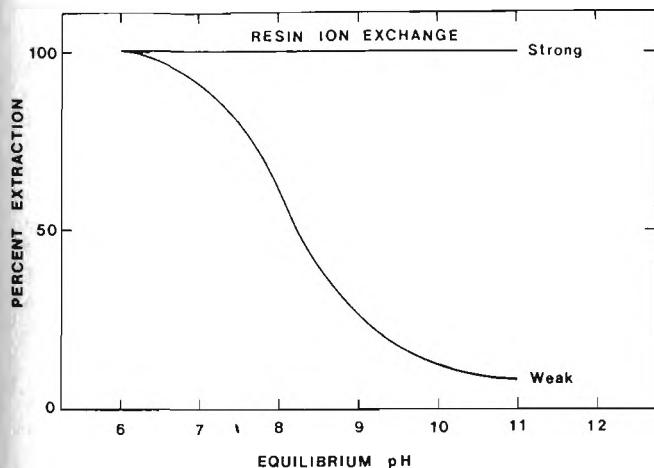
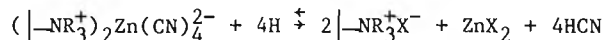
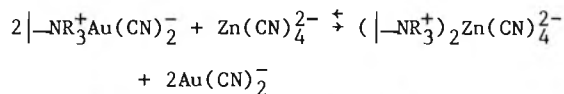


Figure 5. The effect of pH on the extraction of $\text{Au}(\text{CN})_2^-$ by strong and weak base resins.

It is clear that the elution procedure for the RIC process will depend on the type of resins, i.e., weak or strong base. Weak base resin can be easily eluted by using a dilute caustic solution at ambient temperature. On the other hand, it is more difficult to elute the strong base resins. Researchers at MINTEK, South Africa, have been able to elute $\text{Au}(\text{CN})_2^-$ from strong base resins using concentrated $\text{Zn}(\text{CN})_4^{2-}$ solutions (Fleming and Nicol, 1981).



This procedure has proved effective but has the disadvantage that the resin needs to be regenerated using acid which destroys the zinc cyanide complex producing HCN gas which must then be reabsorbed into NaOH to be returned to the process.

Solvent Extraction (SX)

This technology is still in its infancy, a suitable extractant having only recently been identified (Mooiman and Miller, 1983). Therefore many of the engineering aspects have not been studied and in this paper we shall present a rather simple flowsheet, incorporating the most common contactors, namely, mixer-settlers. This approach permits at least a first-order evaluation of the capital and operating costs for a gold solvent extraction plant. The conceptual flowsheet, shown in Figure 6, is described on the next page.

The pregnant leach solution from the heap will have to be filtered if the suspended solids content is too high, in order to avoid excessive solvent losses (by adsorption onto the solid particles). In addition, suspended solids promote phase disengagement problems and the formation of interfacial crud. The clarified solution is contacted with the organic phase in a series of two mixer-settlers with the organic phase flowing countercurrent to the aqueous phase. The mixer-settlers will have to be designed to allow organic recycle to achieve an organic:aqueous phase ratio of at least 1:1. The raffinate will have to be passed through a coalescence unit in order to limit the amount of entrained organic. A scavenging unit such as an activated-carbon bed might be used. The carbon would be periodically steam-stripped, thereby returning the organic to the circuit. In the solvent extraction circuit, the loaded organic from extraction is stripped with a 1-10% caustic solution at an organic-to-aqueous flow ratio of at least 10:1 in order to achieve the desired concentration of the gold values. The loaded strip solution is then passed to a conventional gold electrowinning cell.

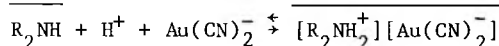
It should be appreciated that the mixer-settler might not, in fact, be the best contactor for a gold solvent extraction circuit which requires high aqueous-to-organic flow ratios and very low organic losses. Among the contactors that could be considered are raining drop, centrifugal, or pulsed columns.

If contacting of the aqueous and organic phases proves to be difficult, another process option is

to impregnate a porous polystyrene bead with the extractant, producing an impregnated resin, and to follow a resin-in-column (RIC) procedure. However, several advantages such as capacity, kinetics, selectivity, etc. over conventional RIC technology will have to be proved if this approach is to be adopted.

Many of the technological features of the gold SX process are described in another paper presented at the 1983 SME Fall Meeting (Mooiman and Miller, 1983). A brief summary of these features will be presented below.

The organic phase consists of a weak base amine whose basicity, with respect to gold extraction, has been increased by the addition of an organic phosphorous oxide modifier, such as TBP, all in an appropriate diluent. The gold-loading step can be performed at about pH 9.5 depending on the amine type and concentration, the modifier type and concentration, and possibly the diluent type. Small acid additions to the leach solution might be necessary in some cases to decrease the pH to a value in the range of pH 9-10 to maximize loading. The loading reaction for a secondary amine can be described as follows:



where R = C₁₀-C₁₂ alkyl group. Good extraction is achieved at pH 9.5 even for feed solutions containing 10 ppm as shown in Table 3.

Stripping can be accomplished at pH values where the amine deprotonates, i.e., pH 12-13 or

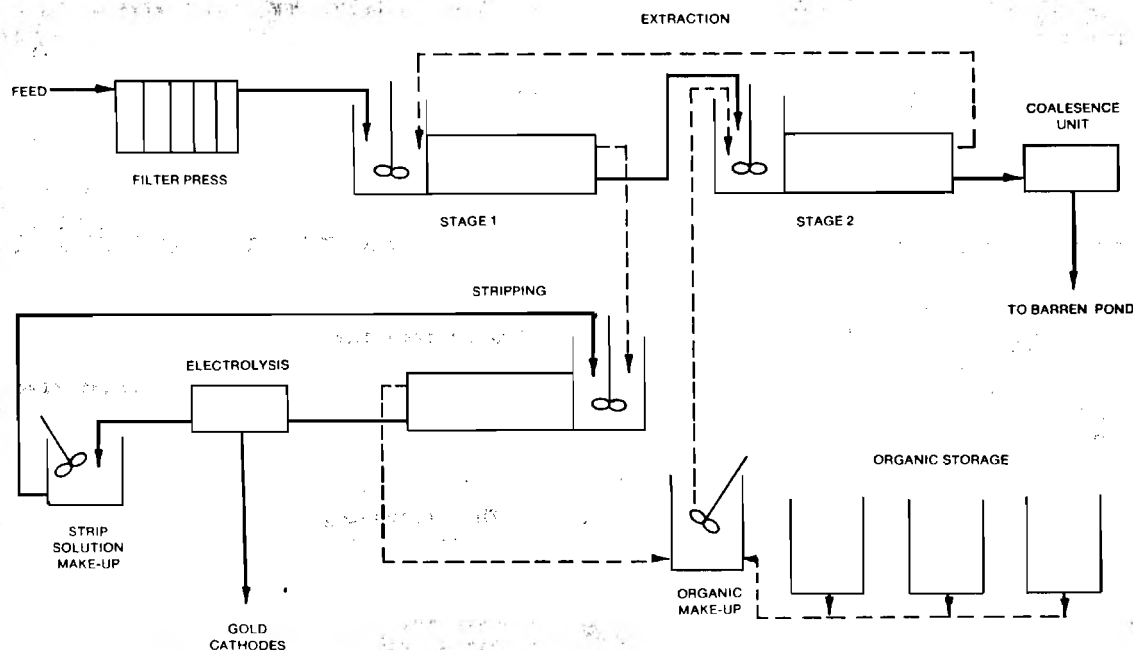


Figure 6. Conceptual flowsheet for gold recovery using solvent extraction.

Table 3. Gold Extraction from Cyanide Solution with a Modified Secondary Amine

Equilibrium pH	Percent Extraction from 10 ppm Gold Solution
9.5	100.0
10.0	87.3
10.5	73.2
11.0	27.1
11.5	11.8

with a caustic solution of about 0.1-0.5%. The loading and stripping reaction rates are relatively fast, and therefore long contact times are not necessary. Excellent selectivity for gold over other metal cyanoanions present in alkaline cyanide leach solutions, e.g., $\text{Fe}(\text{CN})_6^{3-}$, $\text{Cu}(\text{CN})_2^-$ etc. has been found which should allow for the production of a relatively pure gold product.

This technology has some interesting possibilities, and at the moment research and development activities are in progress.

Direct Electrowinning (DEW)

Electrowinning of gold according to the Zadra technique is practiced using rich eluates generated by the stripping of carbon from carbon adsorption circuits. In the case of heap leaching and as discussed in this paper, concentrated aurocyanide solutions can be produced by the following technologies:

- * carbon-in-column
- * resin-in-column
- * solvent extraction

Paul et al. (1983) defined a "concentrated" aurocyanide solution as having gold concentrations ranging from 50 to 2000 ppm. At these concentrations gold can be electrodeposited onto encased stainless-steelwool cathodes. The direct electrowinning (DEW) of gold from dilute aurocyanide solution represents an emerging technology which has been suggested for heap leaching operations. Dilute aurocyanide solutions from heaps are defined as those having gold concentrations in the range of 1 to 10 ppm Au.

A direct electrowinning system, from which gold is recovered in the metallic state must have extended surface area electrodes and can be basically categorized as a flowthrough supported fiber electrode or as a particulate (packed-bed) electrode. In both cases, the electrodes are characterized as conductive material having extended surface areas (i.e. fiber or particulate carbon). The porous nature and the large active electrode

area serve to increase the mass transfer limiting currents and thus the rate of electrodeposition by

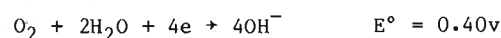
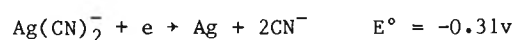
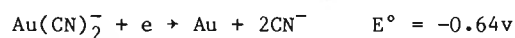
- * reduction of boundary layer thickness as a result of the increased flow velocity imposed by the flow of solution through macro- and micro-channels of the electrode.
- * increased specific surface area.

These factors may make it practical to electrowin metals at low concentrations from aqueous solution.

The DEW process for recovering precious metals from heap leaching circuits is illustrated in Figure 7. Leach solution from the heaps is clarified to remove suspended solids and is pumped to the electrolysis cells. Gold and silver are electrochemically deposited onto the extended surface area (ESA) cathodes and the barren solution is returned to the heaps after reagent make-up. The precious metal product can be recovered in several ways depending on the electrode system:

- * For the particulate packed-bed electrode, the metal can be recovered by melting the carbon/precious metal particles in a furnace.
- * For the supported, fibrous flowthrough electrodes, the metal can be removed by electrolytic stripping (i.e., the loaded cathode can be connected anodically and the metal deposited on a solid starting sheet cathode).
- * Alternately, a chemical stripping approach can be used on either electrode system.

In the alkaline cyanide leach solution, the cathodic reactions of interest are:



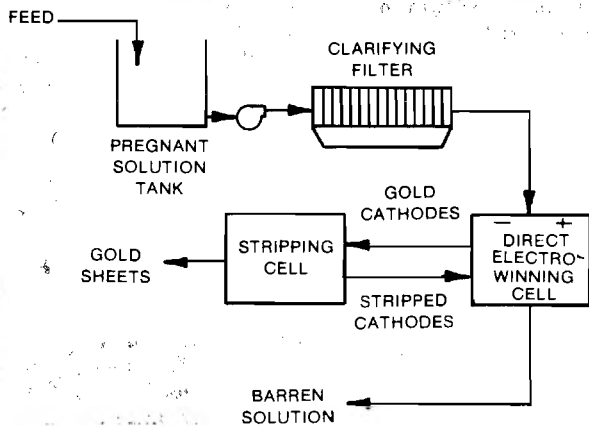
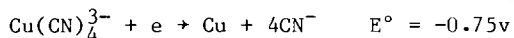


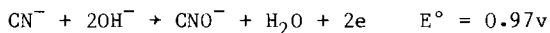
Figure 7. Conceptual flowsheet for gold recovery by direct electrowinning.

As reported in Table 1 heap leach solutions may contain significant concentrations of cuprocyanide species which can undergo cathodic reduction as well,

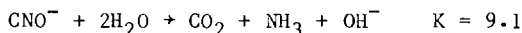


The reduction of the $\text{Cu}(\text{CN})_2^-$ has a standard electrode potential of $E^\circ = -0.43\text{v}$. In either case, Au, Ag, and Cu will probably co-deposit at the operating potentials.

Oxygen is generated at the anode and cyanide ion is oxidized according to the following anodic reaction:



The cyanate ion reacts in the bulk solution to yield CO_2 and NH_3 as follows:



Cyanide destruction during DEW can be avoided by using a diaphragm electrolysis cell which isolates the anodic compartment from the gold-bearing cyanide solution.

HSA Reactors Limited (1982) of Toronto, Canada, manufacture and market a high-surface-area (HSA) carbon-filter electrolysis system. This system has been piloted at a heap-leaching operation in Nevada. It is reported that existing systems can accommodate solution flow rates of up to 400 gpm and can achieve acceptable metal loadings from precious metal solutions containing as little as 0.01 oz/ton.

Maslii and coworkers (1970) have examined the application of extended surface area electrodes in gold hydrometallurgy, operating cells since the early 1970s. The carbon-graphite fiber version contains 10-micron diameter threads giving a specific surface of $0.2 \text{ m}^2/\text{g}$. The capacity of these cells is projected to reach 180 gpm.

Summary of Technological Features

1) Features of Conventional Technology

Merrill Crowe Process

- Fast Reaction Kinetics
- Passivation-TSS, O_2 , Poisons
- Low Selectivity
- Ease of Operation
- Good Accounting, Low Inventory

Carbon in Column

- Slow Adsorption/Desorption Kinetics
- Excellent Recovery
- Low Sensitivity to TSS
- Fair Selectivity
- Carbon Transport-Regeneration
- Poor Accounting, High Inventory

2) Features of New Technology

Resin in Column

- Intermediate Reaction Rates
- Insensitive to TSS
- Fair Selectivity
- Size/Density

Solvent Extraction

- Fast Reaction Kinetics
- Sensitive to TSS
- Good Selectivity
- Organic Losses

Direct Electrowinning

- Fast Reaction Kinetics
- Sensitive to TSS
- Fair Selectivity
- Low Capacity

ECONOMIC ANALYSIS

As detailed in the first part of this paper, some of the new technologies, for example resin-in-column (RIC), appear to offer considerable technical advantage over the well-established procedures such as Merrill-Crowe (MC). However, it is interesting to examine how these new technologies compare economically, and this is the objective of this section of the paper.

In order to present a fair comparison between the various technologies, the capital and operating costs of only the gold recovery section were calculated. Ore preparation, leaching, tailings disposal, solution ponds, etc., were considered the same for all processes. Also, the cost estimation procedure was based on a theoretical feed solution containing only gold. Silver recovery was not considered. All calculations were based on volumetric flowrates of pregnant leach liquor

and concentration of gold in the leach liquor. Solution flowrates in the 100-1000 gpm range were chosen, and two typical gold concentrations, 1 and 5 ppm, were used. These values should cover the wide range of operating conditions typically experienced in gold heap-leaching practice.

It is not intended that these capital and operating costs be definitive for any particular plant, but rather they serve as a basis of comparison of a particular processing strategy with respect to others. Any definitive cost study comparing process alternatives for a particular applications should of course involve the costing of the overall process, and naturally the extent of gold recovery would feature significantly in the calculation. However, recovery figures for new process options are often hard to come by and are not featured in our calculations. Unfortunately the costs for the DEW process were not available for inclusion in this report.

Capital Costs

The simplified flowsheets illustrated in Figures 2 to 5 were costed on the following basis: The capital cost estimates for the 100 and 1000 gpm extreme cases were obtained by costing the major items of equipment required for each process alternative. Engineering and construction costs were estimated. Piping, instrumentation, electrical, and ancillaries were factored from the cost of the major items of equipment. The cost of buildings and civils were excluded. The capital costs for the intermediate 500-gpm case were factored from the 1000-gpm costs using a cost exponent of 0.67, i.e., $(500/1000)^{0.67}$.

It should be appreciated that all these plants were sized and costed from the flowsheets. However, it should be noted that package processing units are available which are ideal for low flowrates and far less expensive than building a plant for a specific duty. For example, an MC unit that treats 50 gpm of leach solution can be obtained for approximately \$25,000. The cost of these package units were not considered in this evaluation. Capital cost results are shown in Table 4.

Some interesting observations can be made from Figure 8 which shows capital cost versus solution flowrate for a 1 ppm Au feed. For all flowrates, the RIC option appears to be the most attractive

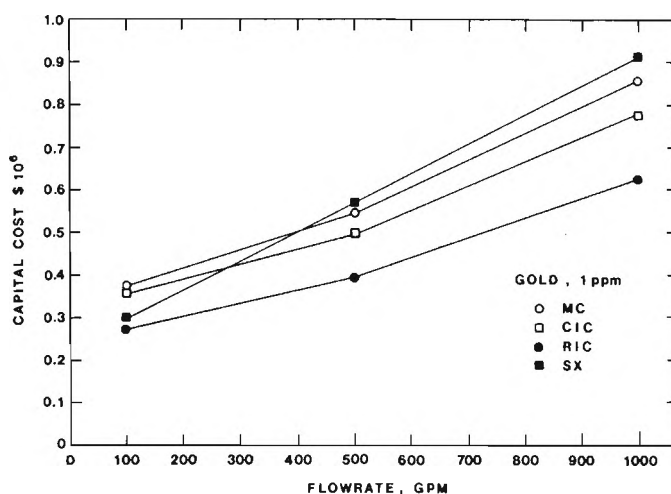


Figure 8. Capital cost as a function of solution flowrate at 1 ppm gold for different process alternatives.

as it has the lowest capital cost. It is less expensive than the CIC process because the RIC elution circuit does not require heat exchangers to heat the eluant and it does not require a regeneration kiln -- a particularly expensive item. At the lower flowrates (100 gpm), SX appears more economical than CIC or MC, but, at higher flowrates, the capital cost rapidly rises, making it economically unattractive.

In the CIC circuit some saving could be obtained if a vertical kiln were used instead of a horizontal kiln. The vertical kiln is approximately 30% less expensive and has a far lower operating cost. However, its efficiency and effect on carbon attrition is not known and will have to be determined. Considerable savings for both the RIC and CIC could be realized if a well-developed continuous ion exchange contactor were used instead of the discrete column arrangement as conventionally used in CIC circuits. Initial calculations indicate that overall capital cost savings of about 25% for both RIC and CIC could be obtained if a continuous contactor were installed. Some of the advantages that a continuous contactor would offer are:

Table 4. Capital Cost in 1983 U.S. Dollars for Different Gold Recovery Options

[Au]	Flowrate	CIC	MC	RIC	SX
1 ppm	100 gpm	360,000	364,000	283,000	309,000
1 ppm	1000 gpm	783,000	858,000	623,000	910,000
5 ppm	100 gpm	494,000	364,000	360,000	309,000
5 ppm	1000 gpm	1,046,000	858,000	820,000	910,000

- * lower operating cost
- * reduced plant area
- * high extraction efficiencies
- * simpler operation
- * lower carbon/resin degradation
- * high security for loaded carbon/resin
- * lower carbon/resin inventory and reduced gold lock-up
- * no dilution of process stream (due to the use of eductors)

Capital cost versus concentration of gold in the feed is depicted in Figure 9. It is interesting to note that the capital cost for both the RIC and CIC processes increase as the gold in the feed increases. As the gold concentration in the feed increases, the rate of gold loading on resin and carbon increases at a slower rate, and the carbon and resin will have to be moved faster through the circuit. Therefore, a larger stripping and/or regeneration section will be required, resulting in an increased capital cost. In the MC process, increasing gold concentration is simply handled by increasing the zinc addition rate. In solvent extraction the phase ratio could be decreased or the gold loading in the organic could be increased, except that the latter could increase gold losses via the entrained organic. The results in Figure 9 indicate that, at high gold concentrations (> 5 ppm), MC processing is the preferred procedure -- a conclusion that has been borne out in industrial practice.

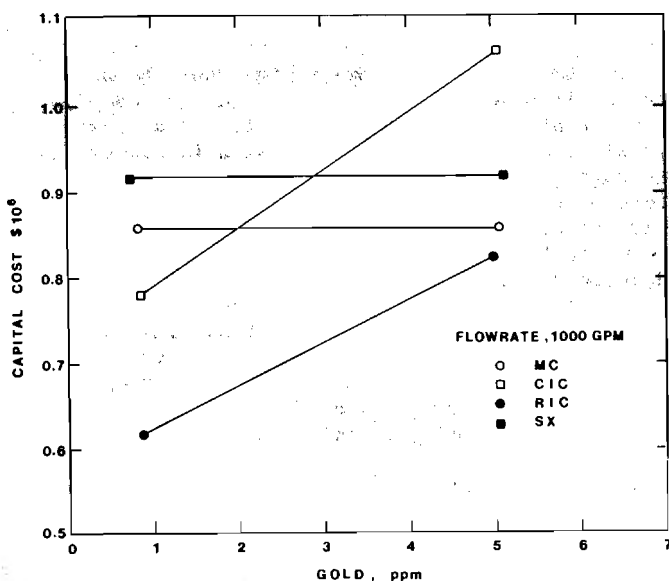


Figure 9. Capital cost as a function of solution concentration for different process alternatives at 1000 gpm.

Operating Costs

The basis for the calculation of operating costs is available from the authors on request.

The results from these detailed calculations of operating cost for each process are presented in Tables 5 through 8. The operating costs in \$/1000 gal and \$/annum are considered for a 1000 gpm plant and gold concentrations of 1 and 5 ppm. As can be seen in comparing carbon-in-column with Merrill Crowe, there is little difference in total operating costs for conventional technology and 1 ppm. However, a significant increase in carbon-in-column cost occurs for higher gold tenors. The reason for this cost increment is associated with the necessity to advance the carbon at a faster rate for higher concentrations and attendant increase in stripping and regeneration cost. However, all the carbon need not be regenerated on every cycle. The operating costs for a Merrill-Crowe process are particularly sensitive to zinc consumption which increases as the gold tenor rises.

Table 5. MERRILL-CROWE OPERATING COSTS

	1000 gpm -- 1 ppm		1000 gpm -- 5 ppm	
	\$/1000 gal	\$/year*	\$/1000 gal	\$/year*
Labor	0.40	201,600	0.40	201,600
Chemicals				
Zinc	0.10-0.20	75,600	0.20-0.40	151,200
Filter Aid	0.15-0.30	113,400	0.15-0.30	113,400
Power	0.10	50,400	0.10	50,400
Total	0.75-1.00	441,000	0.85-1.20	516,300

*Average cost per year

Table 6. CARBON-IN-COLUMN OPERATING COSTS

	1000 gpm -- 1 ppm		1000 gpm -- 5 ppm	
	\$/1000 gal	\$/year	\$/1000 gal	\$/year
Labor	.64	322,500	0.64	322,500
Chemicals				
Carbon	0.006	3,000	0.03	15,100
Eluate Make-Up	0.006	3,000	0.03	16,400
Steel Wool	0.001	500	0.005	2,500
Power				
Heating for Elution	0.03	15,100	0.14	70,500
E.W. Cell		300	0.004	2,000
Carbon Regeneration	0.20	100,800	0.40	201,600
Motors and Pumps	0.05	25,200	0.07	35,300
Total	0.926	470,400	1.314	665,900

Of the newer technologies, RIC appears particularly attractive. Solvent extraction, on the other hand, has a larger operating cost, a large component of which is contributed by the assumed organic loss. These organic losses and attendant operating costs could be improved if a coalescence unit were introduced into the circuit.

Table 7. RESIN-IN-COLUMN OPERATING COSTS

	1000 gpm -- 1 ppm		1000 gpm -- 5 ppm	
	\$/1000 gal	\$/year	\$/1000 gal	\$/year
Labor	0.64	322,500	0.64	322,500
Chemicals				
Resin	0.028	1,400	0.14	70,500
Eluate Make-Up	0.006	3,000	0.03	15,100
Steel Wool	0.001	500	0.005	2,500
pH Adjustment	0.02	10,100	0.02	10,100
Power				
E.W. Cell		300	0.004	2,000
Motors and Pumps	0.05	25,200	0.07	35,300
Total	0.746	363,000	0.908	458,000

Table 8. SOLVENT EXTRACTION OPERATING COSTS

	1000 gpm -- 1 ppm		1000 gpm -- 5 ppm	
	\$/1000 gal	\$/year*	\$/1000 gal	\$/year*
Labor	.064	322,500	0.64	322,500
Chemicals				
Filter Aid	0.15-0.30	113,400	0.15-0.30	113,400
Organic	0.83	418,000	0.83	418,300
Acid (pH adj.)	0.02	10,100	0.02	10,100
Strip Sol'n Make-Up	0.007	3,500	0.035	17,600
Steel Wool	0.001	500	0.001	500
Power				
E.W. Cell		300	0.004	2,100
Motors and Pumps	0.05	25,200	0.05	25,200
Total	1.69-1.848	893,500	1.73-1.89	909,700

*Average cost per year

Overall Cost Comparison

A convenient way to make an economic assessment of the process technologies is by considering the cumulative costs (capital plus yearly operating costs) that accrue during a period of operation. Such a comparison is presented in Figures 10 and 11. On these figures, the capital cost is represented by the intercept on the ordinate. The yearly operating costs are all in 1983 U.S. dollars.

The comparison indicates that at low gold concentration (1 ppm) there is little difference between MC and CIC in terms of overall cost. However, at increased gold concentrations (5 ppm), the cost for CIC is significantly greater than for MC. Therefore, of the conventional technologies, MC is the preferred method of gold recovery, especially at higher gold tenors.

In addition, this comparison points out that additional work will be required to make solvent extraction a competitive process alternative for heap leach solutions. It should be noted, however, that for gold concentrations exceeding 5

ppm, solvent extraction could become economically competitive. On the other hand, significant savings may be realized by the utilization of resin-in-column technology. The RIC process alternative offers an exciting process alternative for the recovery of gold from cyanide leach solutions.

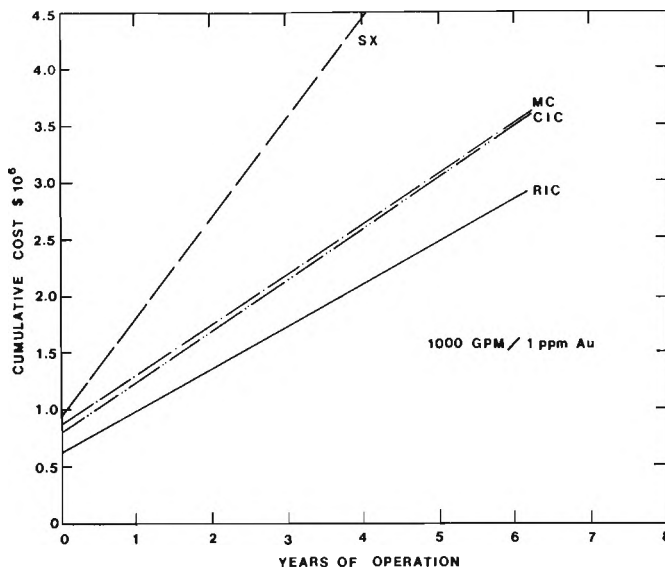


Figure 10. Cumulative cost for different process alternatives at 1000 gpm/1 ppm gold.

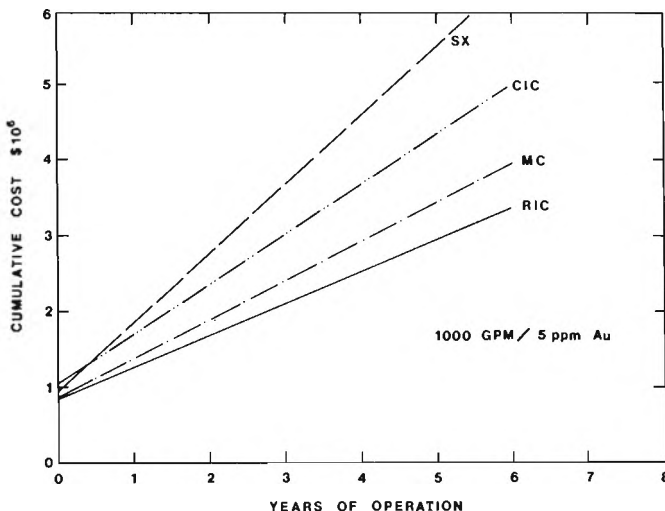


Figure 11. Cumulative cost for different process alternatives at 1000 gpm/5 ppm gold.

SUMMARY AND CONCLUSIONS

The potential for technology change in the recovery of gold from cyanide heap leach solution is high. This change could be via developments in conventional technology. In the case of the carbon-in-column process, improved reactor design (continuous contactors) and improved rate/selectivity of the adsorption/desorption reactions may provide for the advance in process technology. Probably the most significant feature of carbon recovery technology is the essentially complete removal of gold from 1 ppm leach solutions. Electrolysis from the strip solution usually results in a satisfactory product quality for bullion production.

In the case of the Merrill-Crowe process, care must be taken to avoid the passivation and/or poisoning of zinc dust in order to achieve successful recovery from dilute heap leach solutions. The cementation reaction is fast, but the precipitation product quality can be low, especially for recovery from dilute solutions containing impurities. Electrolysis is not required in the Merrill-Crowe process. Although the processes differ significantly from a technology viewpoint, capital and operating costs are similar. It seems that the operating costs at 5 ppm gold would be somewhat higher for a carbon plant than for a Merrill-Crowe plant.

Development of new technology is in progress both in the U.S. and abroad. Some new process alternatives include resin-in-column, solvent extraction, and direct electrowinning. The most advanced of these alternatives is the resin-in-column process. Physical properties of the resin beads, selectivity and ease of stripping are areas of concern for the future development of the resin-in-column technology. Cost analyses suggest that such a process may have an advantage over conventional technology for gold recovery from heap leach solutions. New modified amines and other extractants have been discovered for selective solvent extraction of gold from alkaline cyanide solutions. With such extractants it should be possible to produce high-purity concentrated gold solutions. Engineering aspects of the solvent extraction system must be considered further. Solvent extraction seems to be competitive from a cost standpoint at concentrations exceeding 5 ppm. Direct electrowinning is possible with fiber electrodes and some pilot-scale programs are in progress to allow for better estimates of the potential of this technology.

Although many advances have been made in the past decade, exciting possibilities still exist for further improvement and innovation in gold recovery technology.

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