# The process design of gold leaching and carbon-in-pulp circuits

by W. Stange*

## Introduction

Assuming that a gold ore has been effectively ground to ensure maximum economic liberation of gold, the circuits that will have the most effect on the successful operation of a gold plant will be that of the leaching and carbon-in-pulp circuit (CIP). The reagent and utilities operating costs associated with leaching, adsorption, elution and regeneration would typically make up some 15% of total operating cost, whilst the capital costs associated with these areas is about 16% of the total. Although not largest (capital and operating costs associated with comminution are generally dominant) these items represent a significant proportion of capital and operating cost.

Much more importantly, these plant areas represent the primary gold recovery process and their technical and operational efficiencies will have a significant impact on overall plant efficiency. The objective during process design of these sections is thus to develop a design which provides maximum technical and economic efficiency and which is robust to potential changes in ore throughput, mineralogical characteristics and head-grade. Experience has shown that, particularly for longer life and higher-grade projects typical of the South African underground gold mining industry, small changes in recovery and efficiency have a significant value over the life of the project.

This paper describes the process issues that should be considered during the design of leach and CIP/CIL circuits in order to ensure that these objectives are met. In addition, simulation and modelling techniques that can assist with this design process are also described. Figure 1 provides an illustration of the generic process used to design a chemical or metallurgical plant. The shaded activities represent those areas and applications where computer aids have been developed and are readily available and can provide significant benefit to the design process. This general approach has been termed Computer-Aided Process Engineering (CAPE) by the American Institute of Chemical Engineers.

Computer aids can be applied in a similar manner in the metallurgical industry to facilitate the design process. However, it must be kept in mind that high quality experimental data will be required in order to make effective use of simulation, particularly in the case of the leaching and CIP/CIL processes. These data allow the simulator to be calibrated for the particular scenario that is to be examined and could be obtained from the laboratory, pilot plant and/or full-scale plant. In the case of leaching and CIP/CIL, such data are relatively easy to obtain for existing plants, assuming that the problem is one of the design of an upgrade for an existing plant. In the case of a new plant, such data are much more difficult to come by, particularly if the ore is from a deep underground mine. In this case, the most appropriate approach is to use experimental data obtained from a similar pulp on a currently operating plant.

## Process overview and description

### *The CIP process*

A block-flow diagram of a typical CIP plant for a non-refractory gold ore is shown in Figure 2. Table I and Table II illustrate the capital and operating cost breakdowns for a typical South African gold plant. These figures are not a standard but reflect the nature of the ore and the design basis for a particular situation.

The ore is first reduced in size (typically 80% passing 75μm) to ensure that all non-refractory gold is readily accessible for cyanide leaching. There are several variations of comminution circuits used in the gold industry, such as:

- Multi-stage crushing and pebble and/or ball milling circuits, typical of older installations.

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 *SA ISSN 0038–223X/3.00 + 0.00. Paper received Jul. 1998; revised paper received Aug. 1998.*

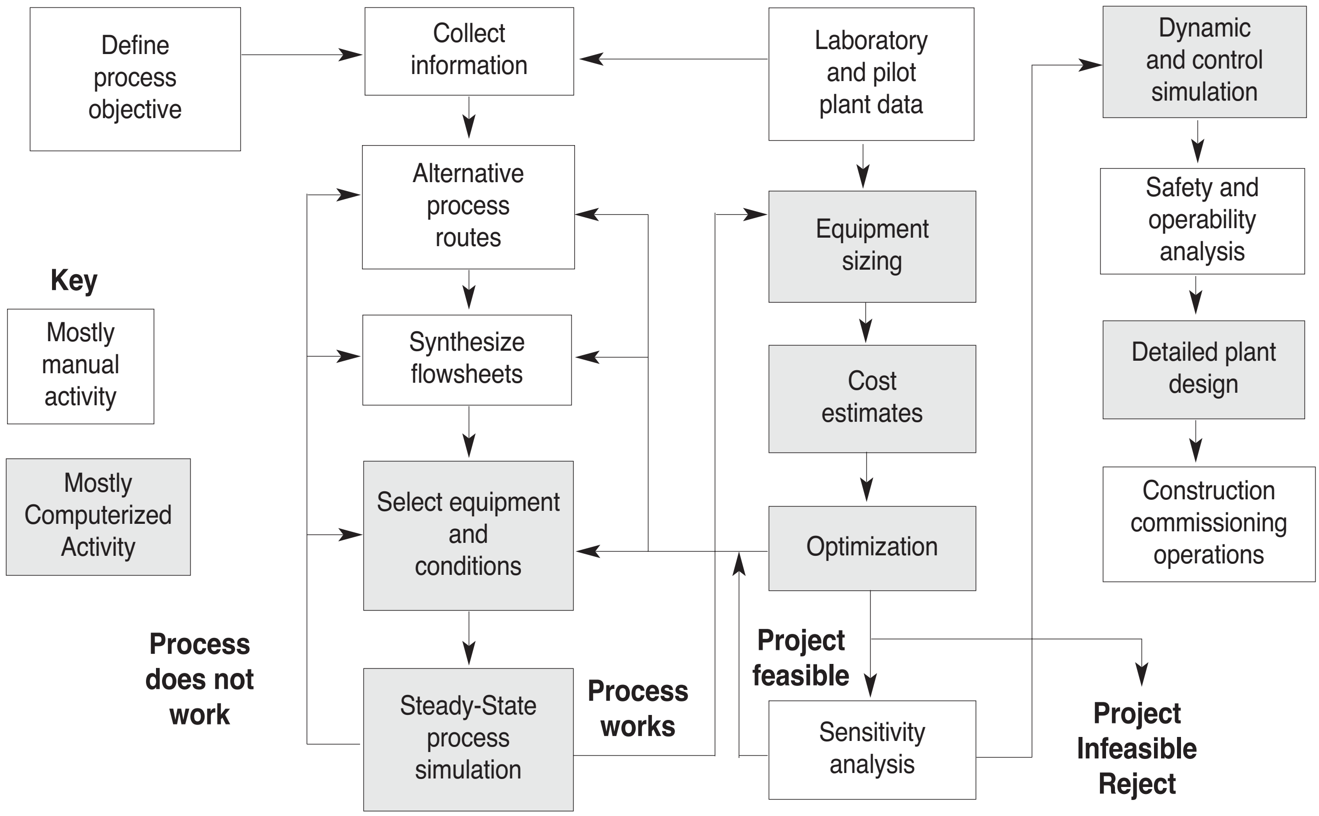


**Figure 1—Computer-Aided Process Engineering (Cape)**

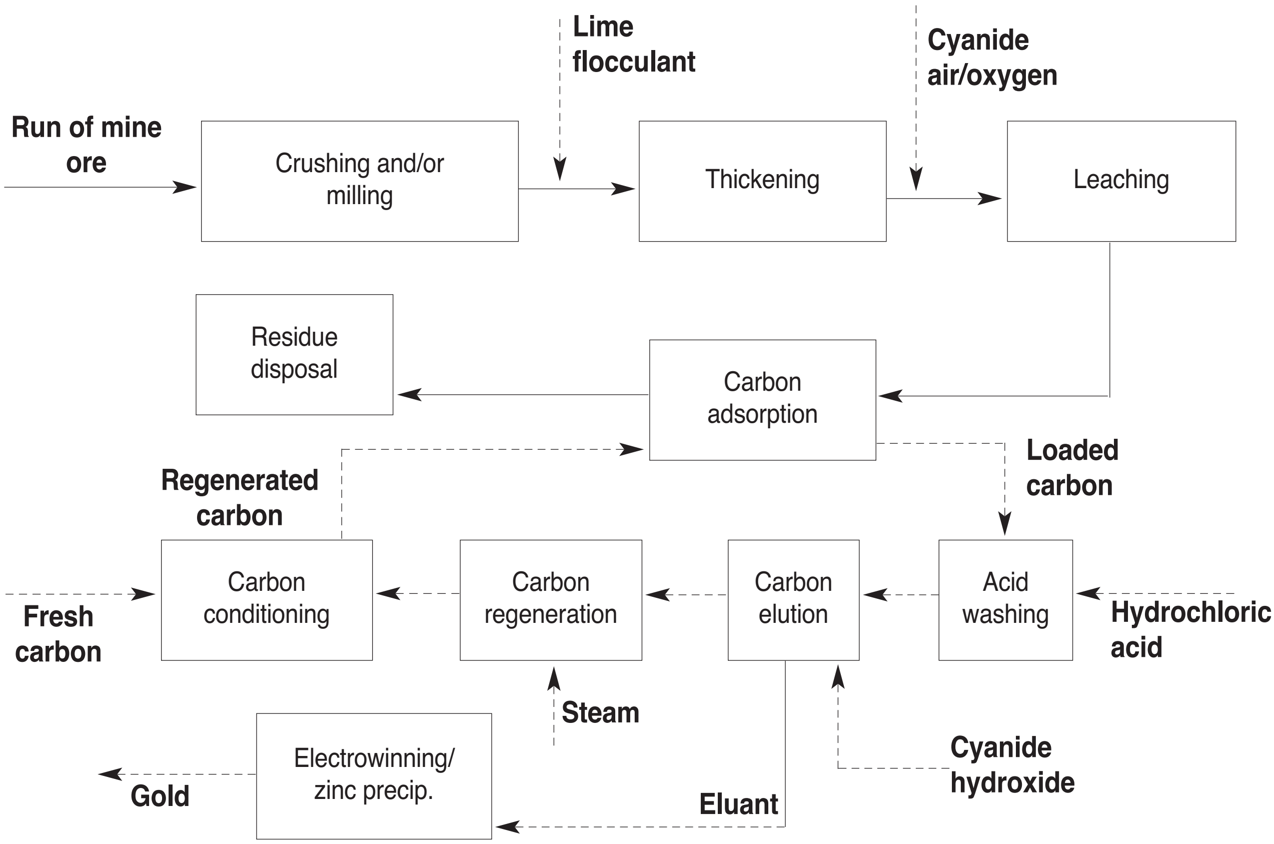


**Figure 2—The Carbon-In-Pulp (CIP) process**

- Single-stage semi-autogenous (SAG) or autogenous (AG) circuits and primary single stage SAG circuits followed by secondary ball milling.

After comminution the pulp is normally dilute and thickening (6-12% solids by mass) is performed to increase the pulp density to about 50% solids by mass. This reduces the size of the leaching plant that would be required as well as reduces the amount of leaching reagents required. Many gold plants use conventional thickening with flocculant addition. High rate thickeners, as well as de-watering systems that use cyclones only or cyclones and high-rate thickeners can also be used in principle.

Leaching reagents in the form of cyanide and an oxidant such as air or oxygen are added after thickening. Leaching takes place in a series of agitated leach reactors or pachucas. Before leaching the pH of the pulp is normally adjusted to a value of around 9.5–11 to ensure minimum loss of cyanide as hydrogen cyanide.

The leaching of gold can be conveniently represented by the Elsener equation:

$$4Au + 8NaCN + O_2 + 2H_2O \rightarrow 4NaAu(CN)_2 + 4NaOH \quad [1]$$

*Table I*

**Typical capital cost breakdown**

| Item | % of capital |
|---|---|
| General site facilities | 7.4 |
| Services facilities | 7.5 |
| General piping utilities | 6.5 |
| General electrical | 4.8 |
| Process control | 1.1 |
| Waste reclamation and delivery | 0.9 |
| Ore-delivery | 2.7 |
| Primary crushing | 2.7 |
| Ores Storage | 2.7 |
| Milling | 21.1 |
| Thickening | 5.1 |
| Leaching | 7.5 |
| Adsorption | 5.2 |
| Elution | 3.5 |
| Gold refining | 4.2 |
| Reagent make-up and dosing | 1.4 |
| Residue disposal | 1.0 |
| Tailings dam | 5.5 |
| Indirects | 9.2 |

*Table II*

**Typical operating cost breakdown**

| Item | % of operating |
|---|---|
| Reagents and chemicals | 12.2 |
| Consumables | 16.3 |
| Utilities | 26.7 |
| Maintenance | 3.38 |
| Labour | 37.9 |
| Tailings dam | 0.8 |
| Assays | 2.8 |

Although air agitated leach tanks were commonly used in the industry, mechanically agitated reactors are preferred due to lower operating costs. Typical residence times in the leach section may range from 20–40 hours depending on the head-grade and nature of the ore and the number of leach reactors in series may range from 6 to 12.

After leaching, the pulp is passed over a feed screen to ensure removal of tramp material such as wood-chips, plastics and grit larger than about 0.6 mm. This is done to minimize screen blocking in the adsorption section. After feed pre-screening the pulp flows through a cascade of well-mixed adsorption tanks, typically 6 to 8 in number. Conventional practice is to have a mean pulp residence time of about an hour in each tank. The tanks are normally mechanically agitated and each will contain a batch of carbon of concentration 10–25 grams of carbon per litre of pulp (0.5 to 1.2% by volume carbon). The carbon is retained in each reactor by means of screens having an aperture of 0.6–0.8 mm which allows pulp to flow through and out of the reactor whilst retaining the carbon in the reactor.

The gold aurocyanide complex in the aqueous phase is readily adsorbed onto the activated carbon. By the time the pulp leaves the last tank in the adsorption cascade the concentration of gold in the aqueous phase is typically between 0.001 and 0.02 ppm with a value of 0.01–0.005 ppm regarded as a practically achievable value for most well-designed and operated plants. This represents a gold recovery of 90–99% depending on the tenor of the feed solution. These high recoveries are achieved because the carbon is transferred countercurrent to the main pulp flow. Transfer is normally accomplished by pumping pulp and the associated carbon in a countercurrent direction. Due to the abrasive nature of the pulp a certain amount of degradation of carbon takes place within the adsorption section. Carbon fines that are produced by abrasion pass through the inter-stage screens and move co-current with the pulp down the adsorption cascade. The barren pulp is screened at 0.6–0.8 mm to recover undersize carbon. Tailings are pumped to a slimes dam for disposal.

The carbon in the reactor into which pregnant pulp is fed (referred to as the first reactor) becomes highly loaded with gold. Loaded carbon values on operating plants range from 300 to 20 000 grams of gold per ton of carbon, a concentrating factor of about 1 000–1 500. A portion of the loaded carbon is periodically removed from the first adsorption reactor. This loaded carbon may then be subjected to an acid wash by treating the carbon with a hot or cold solution of dilute hydrochloric acid. This removes $CaCO_3$ that has precipitated on the carbon, as well as cleaning fines out of the carbon pores.

The acid washed carbon is then eluted. This is done by contacting the carbon with a solution of sodium cyanide (0.1 to 2% by mass) and sodium hydroxide (0.1 to 2% by mass) at high temperatures (90–120°C). This results in the reversal of the adsorption process with most of the gold desorbing from the carbon back into solution. This produces a small volume of solution with a high gold concentration. The gold is recovered from this solution by electrowinning, zinc precipitation or refining technology such as the Mintek Minataur process.

The eluted carbon may still contain various organic contaminants. These are removed by thermal regeneration of the eluted carbon, typically in a rotary kiln at temperatures of 650–750°C. Regeneration is carried out in a steam atmosphere to minimize carbon degradation due to oxidation. The eluted and regenerated carbon is screened at about 0.8 mm to remove any undersize carbon. New carbon is added to the regenerated carbon to balance carbon losses from the circuit. This mixture is then added to the last tank in the adsorption cascade.

It has been noted that a small amount of leaching takes place in CIP plants, due to the additional residence time of pulp in an environment appropriate for leaching in the CIP reactors. This additional leaching provides additional economic benefit.

### *The CIL process*

The carbon-in-leach (CIL) process, illustrated in Figure 3 is a variation of the CIP process. In this process carbon is added directly to the leach circuit so that the leaching and adsorption processes proceed simultaneously. Capital cost is reduced, as only one set of agitators for both leaching and adsorption is required, as opposed to a set of agitators for leaching and a set for adsorption.

There are, however, disadvantages inherent to the CIL process. For example, the agitated tanks need to be large to provide sufficient residence time for leaching. This results in low carbon concentrations in the vessels making counter-current transfer of the carbon and good operational control of the carbon inventory more difficult. Other serious operational problems, such as the flattening of the adsorption profile due to concurrent leaching and adsorption leading to a reduction in the driving force for adsorption, have been encountered with the CIL process and have been described by Bailey (1987). Experimental work has shown that the CIL process tends to be less efficient, in terms of gold recovery, compared to the conventional leach-CIP route (Davidson 1988).

The applications for which CIL has definite advantages are:

- The treatment of low-grade reclaimed dump material where long leaching times are not normally required to leach all the gold which is recoverable by cyanidation.
- The treatment of complex ores which contain constituents (such as carbonaceous material) which tend to adsorb or precipitate the gold aurocyanide complex. If leaching is performed in the presence of activated carbon the aurocyanide complex is preferentially adsorbed onto the activated carbon, resulting in higher gold recoveries.

## Establishing a process design basis

### *Leach parameters and test-work*

Young (1987) provides a comprehensive discussion of South African cyanidation practice. The most important parameters affecting leach performance are the following:

- *Lime addition*
  In order to create an aqueous environment with a pH of 9.5 to 11 so that minimum cyanide is lost by conversion to HCN gas, lime is normally added to modify the pH to this level. A CaO content of 150–200 ppm is normally required to achieve this. The actual

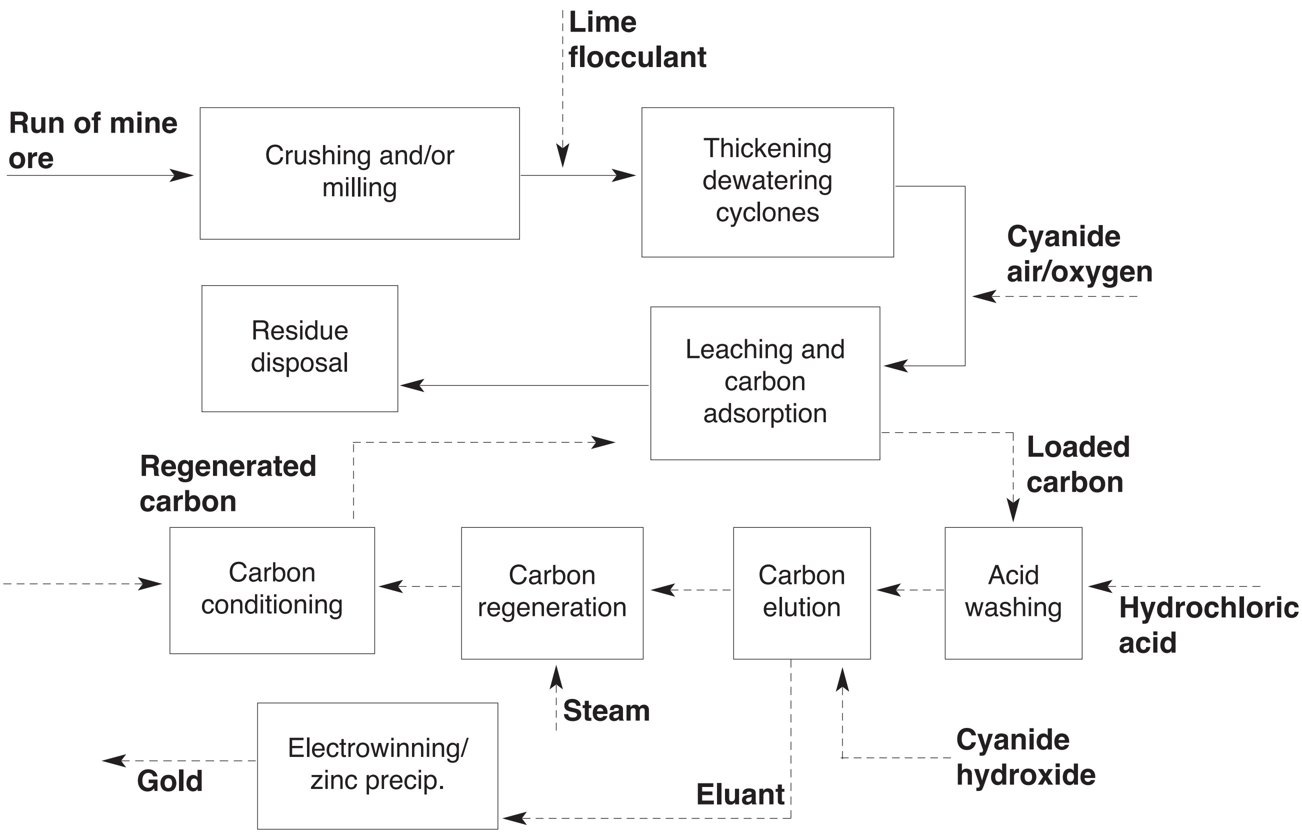


**Figure 3—The Carbon-In-Leach (CIL) Process**

lime addition rate required to achieve this will depend on the specific properties of the pulp. Typical lime addition rates vary from 700–1500 g/ton of 100% CaO equivalent.

- *Grind size*
  The influence of grind on the residue value that can be attained by direct cyanide leaching is illustrated in principle in Figure 4. The position and shape of this curve is dependent on the particular ore.
- *Cyanide consumption*
  There are many components in a typical ore that will consume cyanide by side reactions. In addition, cyanide is also lost by hydrolysis. In order to achieve effective leaching, cyanide concentration (expressed as 100% NaCN) must be maintained at a minimum level. This minimum value may typically be in the order of 120 ppm in the last leach vessel. A cyanide concentration of 200–250 ppm may be required in the initial leach vessel to achieve this. The actual rate of cyanide consumption for a specific ore will vary depending on its characteristics. Typical cyanide addition rates vary from 150 to 500 g/ton of 100% NaCN equivalent.
- *Oxygen demand*
  Oxygen is a crucial reagent for leaching. Pulps may contain organic and inorganic components that consume oxygen, reducing the overall level of dissolved oxygen in the pulp and thus reducing leach kinetics and/or leach efficiency. When the use of air agitated pachucas was common oxygen was almost always present in sufficient quantities due to the vast quantities of air pumped into the pulp for agitation purposes. However, with the increasingly common use of mechanical agitation for leaching vessels, oxygen is often supplied as a reagent.
- *Agitation*
  Due to the mechanisms of gold leaching, a certain minimum rate of agitation is required. There have been cases reported where agitation at low rates (to save power) has resulted in poor leach recoveries.
- *Residence time*
  Depending on whether the ore is a low-grade dump material or a high-grade ROM material, varying residence times of leaching will be required for acceptable leach recoveries. Typical residence times may vary from 12 to 48 hours.
- *Pulp density*
  Pulp density affects viscosity considerably, which has an impact on gold leaching. It has been found that pulp densities which are too high, as well as those which are too low can affect gold leaching performance in a negative manner. Overly dense pulps hinder mass transfer whilst dilute pulps result in a loss of ore leaching residence time as well as high reagent addition rates. A compromise between these two effects must be achieved and a w/s ratio of around 1.0–1.1 has been found to be effective.

In order to determine the above parameters for a particular pulp, it is crucial that experimental work be performed. As with all metallurgical test-work it is most important that the work be performed on a representative sample. In addition, there is an inherent level of scatter and noise in this type of work. Multiple repeats of identical tests will allow some confidence in the reproducibility of the work.

Such test-work can normally be executed quite rapidly and cost-effectively with a minimum of complex equipment. Some commercial laboratories, such as Mintek, offer a structured test-work programme that will identify the optimum parameters for leaching a particular ore in an effective manner.

It has been found that rolling bottle leach tests (ideally performed in a constant temperature environment) offer a simple manner of performing leach tests that produce results comparable with those that would be obtained on a full-scale plant treating the same ore. Tests performed with vigorously stirred reactors on a laboratory scale may result in optimistic predictions of leaching rates due to the much higher mixing intensities produced in stirred laboratory reactors.

### Modelling of leach kinetics

A number of useful kinetic models have been developed which can help to interpret experimental leaching data and which can also be used to predict the performance of a continuous leaching plant from batch experimental data. Woollacott, Stange and King (1990) have discussed these models. A summary of the model equations is provided here for convenience.

Where:

- $S$ is the gold residue value in the ore (g/ton)
- $S_t$ is the gold residue value in the ore in the batch reactor at time $t$ (g/ton)

| Model | Batch Expression |
|---|---|
| Mintek: $R_L = k_p\left(S - S_m\right)^2$ | $S(t) = S_m + \dfrac{\left(S_o - S_m\right)}{1 + \left(S_o - S_m\right)k_p t}$ |
| Brittan: $R_L = \left(S - S_m\right)e^{b_1\left(S - S_m\right) - b_2}$ | Must be integrated numerically |
| Loveday: Distribution of zero order rates described by the Schumann distribution | $S(t) = \left(S_o - S_m\right)\left(1 - \dfrac{n}{n+1}k_{\max}t\right) + S_m\left(t < \dfrac{1}{k_{\max}}\right)$ <br> $S(t) = \dfrac{\left(S_o - S_m\right)}{n+1}\left(\dfrac{1}{k_{\max}t}\right)^n + S_m\left(t > \dfrac{1}{k_{\max}}\right)$ |

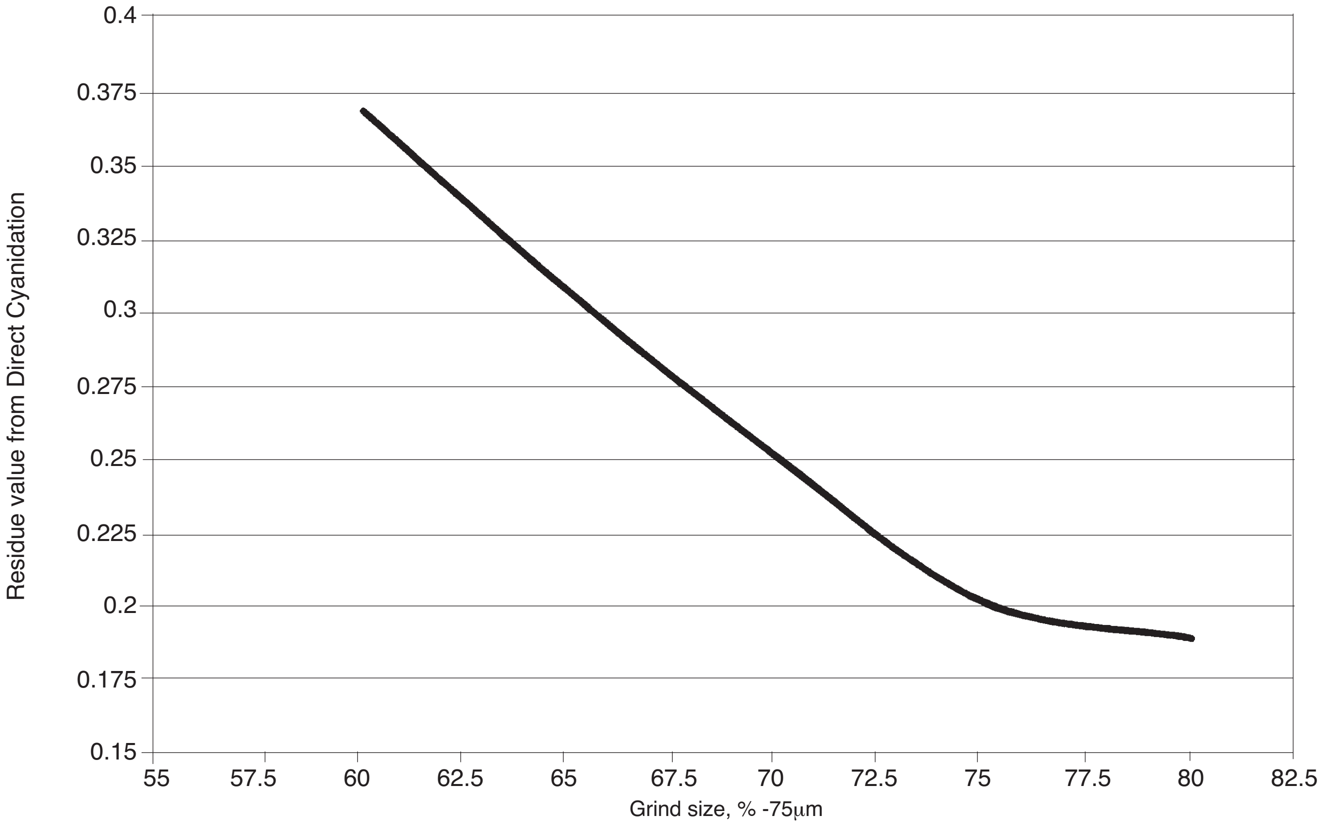


**Figure 4—Influence of grind size on leach recovery**

- $S_o$ is the head-grade of gold used in the batch leach test (g/ton)
- $S_m$ is the ultimate residue value i.e. the gold concentration in the residue after an 'infinite' period of leaching. In this case, 80 hours of batch leaching is a good approximation to infinity. This value will vary as a function of ore particle size distribution as well as possibly other leach conditions
- $k_p$ is the leaching rate constant for the Mintek model
- $b_1$ and $b_2$ are leaching rate parameters for the Brittan model
- $n$ and $k_{max}$ are the parameters describing the distribution of zero order leaching rates for the Loveday model.

These parameters can then be easily estimated from batch leaching data using the Solver capability in a spreadsheet like Excel. Ideally, a single set of parameters should be estimated from multiple experimental data sets. Care should be taken when fitting the models that good predictions of the data towards the end of leaching, as the leach curve is flattening off, are obtained. This is important as often the intention of fitting the models is to find the minimum residence time required to provide the economically optimum leach recovery and the residence time required to obtain this. This difficulty is compounded by the fact that batch leach data are often noisy and inconsistent toward the end of the leach.

Once the model parameters have been estimated they can be used to predict the performance of a continuous leach plant under different conditions. The real value of this approach is that it allows the sensitivity of process performance to changes in design parameters (and the associated costs) to be quickly and easily ascertained. The appropriate forms of the leaching models must be used to predict CSTR performance; the models in the Table are for batch leaching. These models are documented in detail in Stange (1988).

It must be emphasized that the kinetics of gold leaching for a real pulp and leach solution is complex and, to the author's knowledge, no fundamental predictive model has been developed. The models discussed here are phenomenological rather than fundamental. The implications for this are that the parameters in the models may depend on actual experimental conditions, such as the ore size distribution, rate of agitation and other chemical and physical effects. It is thus critically important that the laboratory tests, from which these parameters are estimated, are performed under conditions that are as close as possible to those that will apply on the full-scale plant.

In the case where no material is available for test-work the only way in which design parameters can be developed is to base them on industry standards, or by performing test-work on an ore known to be similar. This is inherently more risky and appropriate safety factors must be used in the design.

If the CIL case is being examined, similar test-work needs to be performed to establish leach performance under CIL conditions. The principles are similar to those described above. However, variations in the test-work procedure, analytical procedures and data analysis may vary because of the presence of carbon during leaching.

### The CIP/CIL circuit

It is important to understand some of the fundamental physical and chemical parameters affecting adsorption, acid-washing, elution and regeneration. These are discussed in detail by Bailey (1987) so will be summarized here.

#### Parameters affecting adsorption

The equilibrium capacity of activated carbon for the adsorption of gold is influenced by a number of factors. These include, but are not limited to:

- Temperature
- The nature of the raw material used to manufacture the carbon
- The activation conditions used during carbon manufacture
- pH
- The concentration of free cyanide ions and spectator ions such as $Ca^{2+}$, $Na^{+}$ and $K^{+}$
- The presence of organic solvents such as acetone, ethanol and acetonitrile
- Presence of organic and inorganic foulants such as xanthates, calcium carbonates, and so on.

Woollacott and Guzman (1993) have quantified some of the more important effects on the equilibrium behaviour of the carbon-aurocyanide system under typical industrial adsorption conditions while Van der Merwe (1991) quantified equilibrium influences under typical elution conditions.

Adams (1990) has shown that gold adsorbs onto activated carbon as the ion pair:

$$M^{n+}\left[Au(CN)_2^-\right]_n \qquad [2]$$

Where $M^{n+}$ is the appropriate metal ion such as $Ca^{2+}$, $Na^{+}$ or $K^{+}$. The aurocyanide is only transformed into AuCN or metallic gold by high temperatures or in a strong acid solution or a combination of both. At ambient temperature and in an alkaline solution, conditions typical of a CIP/CIL adsorption circuit, the aurocyanide complex is adsorbed in a fully reversible manner.

Parameters that affect the kinetics of adsorption include:

- Carbon particle size. In general, the smaller the carbon particles, the higher the rate of adsorption
- Temperature
- The surface roughness of the carbon which affects the film mass transfer coefficient for adsorption
- The internal surface and pore diffusion properties, affected by the physico-chemical nature of the carbon which will be influenced by the raw material properties and manufacturing methods used
- The degree of agitation of the pulp in which adsorption is taking place. Under industrial conditions it is suspected that adsorption rates are limited by film-transfer. It is thus expected that increases in agitation intensity will result in improved adsorption rates
- Oxygen concentration
- Solution pH. The lower the pH, the more strongly adsorption takes place
- Competing ions that are adsorbed with gold—like silver, copper and nickel. If these elements are present in sufficiently high concentrations, gold adsorption may be hindered
- Physical blocking of the carbon pores by fine solids or precipitated material.

#### Parameters affecting elution

Elution, from a fundamental perspective, is simply the reverse of adsorption. Thus, parameters which influence adsorption tend to influence elution as well. The various elution processes (Batch AARL, Zadra, Continuous AARL and Micron) are merely engineering systems that allow the various parameters that influence the adsorption of gold onto carbon to be manipulated in a manner which results in desorption or elution. Elution is typically promoted by:

- High temperatures
- High free cyanide and/or hydroxide conditions
- Low ionic strength.

#### Parameters affecting acid washing

During adsorption other inorganic species, notably calcium, adsorbs on the carbon. Other organic species that tend to foul the carbon are silica, alumina and iron oxide. As with organic species, if these are left to build up, the carbon will become poisoned and gold adsorption efficiency will be significantly reduced. Acid washing with HCl is able to remove most of these precipitates and to thus improve adsorption performance. The primary parameters affecting acid washing are:

- *Acid strength*
  This may range from 3% to 10% HCl, depending on the degree of inorganic fouling, as well as the temperature of washing
- *Temperature*
  As expected, it is generally found that higher temperatures accelerate the rate of removal of inorganic species. In practice, acid washing is performed at temperatures ranging from ambient to 90°C. One of the consequences of hot acid washing is that the adsorbed gold in the form of the aurocyanide ion may be decomposed to a species which is much more thoroughly adsorbed (Adams 1990). This makes elution with no or low cyanide levels unachievable.

#### Parameters affecting regeneration

Regeneration is a key unit process if efficiency of gold adsorption is to be maintained over time. The key purpose of regeneration is to remove organic adsorbates which have adsorbed on the carbon and which if left on the carbon would build up to a stage where the carbon would be so poisoned that efficient adsorption of gold would no longer be possible. Regeneration is performed at high temperatures (600–800°C) in a steam atmosphere to minimize the destruction of the activated carbon by oxidation at these temperatures. The primary parameters affecting regeneration are:

- *Carbon feed moisture*
  The carbon must have sufficient moisture content to generate sufficient steam within the kiln, but also not overly wet which will hinder the attainment of elution temperatures or increase the cost of attaining these temperatures

- ➤ *Temperature*
  An appropriate temperature profile must be maintained in the kiln, avoiding cold patches and thus inconsistent regeneration
- ➤ *Residence time*
  A residence time must be established which is consistent with the proposed temperature and degree of regeneration required. Temperature and residence time interact in that the same degree of regeneration is achieved by longer residence times at lower temperatures or higher temperatures at shorter residence times. The optimum set of parameters is that which balances operating and capital costs in the most effective way, as well as minimizing breakage to the carbon by over-oxidation of the carbon.

### *CIP/CIL process design issues*

It is relatively easy to arrive at the optimum parameters for the design of a leach circuit, and these can often be derived from relatively simple test-work, as just described. The development of optimum design parameters is significantly more complex due to the dynamic and interactive nature of the CIP process. This complexity arises due to the interrelationship between the various parameters influencing the performance of a CIP plant, such as:

- ➤ The gold content of the eluted and regenerated carbon
- ➤ The gold content of the solution feeding CIP adsorption
- ➤ The extent of leaching within the adsorption system
- ➤ Carbon activity and properties
- ➤ The solution tails barren value that needs to be obtained
- ➤ The carbon inventory in each stage
- ➤ The rate and mode of carbon movement through the adsorption stage
- ➤ Gold lock-up in loaded carbon in the adsorption, elution and regeneration plant
- ➤ The number and size of the adsorption reactors
- ➤ The elution and regeneration technology to be used, which determines the flexibility and cycle of operation
- ➤ The capacity of the elution and regeneration plants. If undersized, these circuits may limit the rate of transfer that is used on the CIP/CIL plant.

The only manner in which these interactions can effectively be studied in a quantitative manner is by the use of an appropriate mathematical model. Bailey (1987) provides a comprehensive evaluation of the interactions between many of these factors, using a rather simplistic model. Stange, Woollacott and King (1990) explore these issues with a model that was developed and validated against industrial data.

The major interactions in a CIP/CIL plant that have been identified follow.

- ➤ *Number of adsorption stages*
  As the number of adsorption stages increase, the total amount of carbon in these stages required to adsorb the same amount of gold is decreased. This implies that the total gold locked up in adsorption, as well as the total carbon inventory used decreases. This also influences gold loss on abraded carbon, as the rate of carbon abrasion in the adsorption circuit is influenced by the total amount of carbon in the adsorption circuit. This effect diminishes with the increasing number of stages so that a point is reached (around 8 to 10 stages) where insignificant benefits are obtained from increasing the number of stages as the capital and operating savings obtained are outweighed by the incremental capital required to install more adsorption tanks.
- ➤ *Carbon residence time*
  Carbon adsorbs many other species such as calcium and organics from the pulp in the adsorption plant. These species are poisons in that they tend to reduce the carbon's capacity for gold adsorption. The longer the carbon remains in contact with pulp, the more poisons are adsorbed. Thus, longer contact between the pulp and the carbon will invariably result in more poisoning and poorer adsorption of gold. The total carbon contact time (in fact a distribution of contact times) will be influenced by the number of stages, the amount of carbon in each stage, the mode of carbon transfer as well as the total rate at which carbon is moved.
- ➤ *Loaded carbon value*
  The degree to which the carbon in contact with the pregnant liquor will load is strongly influenced by the gold tenor in this solution. Industrial practice has shown that loaded carbon values range from about 800–1 200 times higher than the gold tenor in solution; thus for a pregnant solution value of 1 g/m$^3$, a loaded carbon value in the range of 800 g/ton to 1 200 g/ton will result.

  This is a rule of thumb only and is more applicable to early adsorption plants that were designed using the 'standards' of 1 hour pulp residence time per stage, 20 g/l carbon concentration per stage and 6–8 adsorption stages. Many operating parameters will influence this loading value; for example the 'Pump-Cell' concept (Whyte, Dempsey and Stange 1990) in which smaller high-carbon concentration reactors are utilized allows a significant improvement in loaded carbon values to be attained compared to industry practice. The degree of poisoning of the carbon will also influence the actual loaded value to a significant extent. The performance of equilibrium isotherm tests in the laboratory will allow the expected loaded carbon values to be ascertained with a greater degree of insight.

  The primary benefit associated with higher loaded carbon values is that the carbon transfer rate can be reduced significantly for the same amount of gold recovered. Disadvantages are that gold lock-up in the adsorption circuit may rise and longer times or higher temperatures may be required to elute the higher loaded carbon to the same target values.
- ➤ *Barren value*
  Adsorption plant design and operation has developed to the stage where it is not unreasonable to expect gold tenors in the solution in adsorption tails to be less than 0.005 g/m$^3$. This value will more easily be achieved in the cases where:

- There is sufficient carbon in the adsorption plant
- Good eluted carbon assays (less than 40 g/ton) are routinely achieved
- There is a minimum of leaching in the last few adsorption stages
- The activity of the carbon is consistently high through good acid washing and regeneration practice
- There is minimum co-current flow of carbon through the circuit caused by leakage of carbon through inter-stage screens. Leakage results from holed screens due to poorly maintained inter-stage screens or due to off-specification carbon (or equivalently poor carbon preparation) that is of a shape or size such that it passes through the apertures in an inter-stage screen. This is of particular importance for slotted inter-stage screens such as those based on wedge-wire.

Due to the many interactive parameters that need to be determined for a CIP/CIL plant, it is essential to use a model or simulator that allows the interactions between design parameters to be established. More importantly, the financial ramifications of the technical interactions must be quantified so that a plant design that is both technically appropriate, as well as financially optimum can be developed.

### *Simulation and modelling of CIP/CIL circuits*

Much effort has been put into the development of appropriate models, laboratory and plant test-work procedures and parameter estimation techniques over the last 10 to 15 years. This implies that a suite of validated tools is available which can be used for both effective CIP/CIL plant design, as well as to optimize the operation of such plants.

Interested readers are referred to the references and bibliography that provide details on the relevant publications.

## A techno-economic case study

The focus of this case study is to highlight the effect that various parameters have on project economics. The plant simulated was typical of large-scale dump re-treatment plants. All scenarios were evaluated using the techno-economic simulator described by Stange (1998). The simulator was configured using the appropriate parameters, a simulation was initiated which resulted in the calculation of technical and economic parameters for the particular configuration. The key performance indicators were noted, the configuration was changed to reflect the next set of plant conditions to be studied, and the simulation calculation was re-initiated.

It was found by laboratory test-work on plant pulp and carbon that the equilibrium isotherm was most effectively described by the Langmuir isotherm with the following parameters:

$$y_e = \frac{19\ 900\ C_e}{0.6 + C_e} \qquad [3]$$

Kinetic data measured from the plant was analysed and it was found that a simple first-order rate model fitted the experimental data well and that there was little justification, from the perspective of the fits obtained, to use a more complex model. This was predominantly due to the low solution tenors, and therefore resulting low carbon loadings, on the plant.

The adsorption kinetics could therefore be described by:

$R = k_2\ C$

Where:

- ➤ $R$ is the rate of adsorption (g Au/ton carbon/hour)
- ➤ $C$ is the solution tenor (g Au/ton solution)
- ➤ and $k_2$ is the rate constant. A value of 157 (in units of tons and hours) was estimated from the experimental data

The base case plant configuration was as follows:

- ➤ Ore flow rate of 700 TPH at a head-grade of 1.0 g/ton and a w/s ratio of 1:1
- ➤ 2 leaching vessels of size 960 $m^3$
- ➤ 6 adsorption reactors of size 450 $m^3$ at a carbon concentration of 20 g/l
- ➤ A carousel mode of carbon movement was assumed with a 24-hour adsorption cycle. This results in a carbon transfer rate to elution of 9 000 kg per day
- ➤ An eluted carbon loading of 50 g/ton was assumed.

These conditions resulted in the solution tenor in the adsorption plant tails of 0.009 g/$m^3$ and a solids residue of 0.344 g/ton in the solids in the adsorption plant tails slurry.

Where required, the NPV was calculated over a period of 20 years using a discount factor of 15%. NPVs were also calculated relevant to the base case. Capital cost outflows were assumed to be spent over the first two years. No revenue or operating costs were included in the base case NPV calculation; only capital cost outflows were used in the calculation making the NPV for the base case negative. Calculations of the NPV for subsequent situations included the effect of operating costs and revenues relative to the base case. That is, for each year, the net cash inflow for a given scenario was calculated as follows:

Cash inflow = Revenue for scenario
–Total operating cost for scenario
–(Revenue for base case–Operating costs for base case).

This was done as the author's experience has shown that the large revenue values generated for profits from precious metal plants tend to swamp the capital and operating costs when evaluated using DCF calculations in the standard manner. In addition, the main objective is to compare various scenarios, rather than to calculate absolute financial measures.

### *The effect of the number of adsorption stages*

It is well known that as the number of adsorption stages are increased, the carbon inventory in each adsorption reactor can be decreased for a constant adsorption performance i.e. the same solution tails tenor (Bailey 1987). The decreasing carbon inventory results in higher loaded carbon values transferred to elution, as the mass of carbon to be eluted decreases. This results in the requirement for a smaller elution and regeneration plant, as well as lower elution and regeneration operating costs. In a CIL plant, increasing the number of adsorption stages will also increase the total pulp residence time and therefore the amount of leaching which takes place.

Case studies were conducted by varying the number of

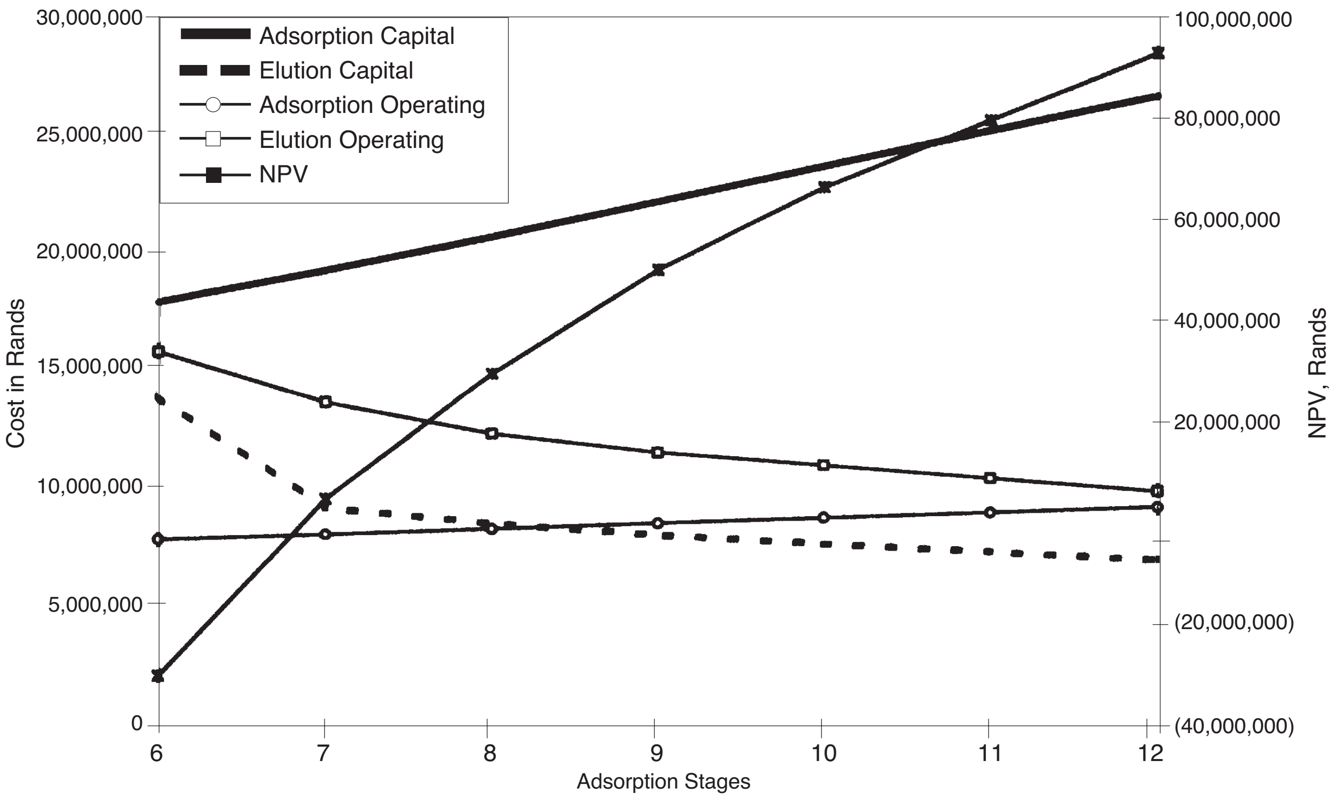


**Figure 5—Effect of number of adsorption stages on CIL plant economics**

adsorption stages, and then adjusting the amount of carbon in each adsorption reactor to keep the tails solution tenor at 0.009 g/m$^3$, i.e., comparable to the base case.

Figure 5 illustrates the effect of increasing the number of adsorption reactors on various financial measures. The capital costs for the various sections are shown as total costs in rands, whilst operating costs are shown as annual costs in rands. As expected, both the adsorption capital and operating costs increase as the number of adsorption reactors are increased. This is compensated for by a fall in elution capital and operating costs.

The overall result is a significant increase in the project NPV as the number of adsorption reactors are increased. This is due to both the decrease in elution capital and operating costs, as well as the increased leaching due to the increased residence time.

Some caution must be exercised when interpreting these results. It is well known that carbon fouls progressively with respect to its gold adsorption ability due to adsorption of both inorganic and organic foulants. Thus, as the number of adsorption stages increases and the adsorption cycle is kept constant, the average carbon residence time would increase. This would tend to exacerbate fouling and thus reduce average adsorption kinetics as the number of adsorption stages is increased. This effect was ignored although the simulator does have the ability to deal with stage-wise varying kinetics.

### *Effect of the number of leaching stages*

When capital is restricted or the grade of the deposit is low, the industry has tended to move toward a CIL situation rather than a leaching plant followed by an adsorption plant. However, this has various operational disadvantages (Bailey 1987) as well as potential revenue disadvantages. Although often difficult to measure in a laboratory environment, practical experience has shown that most gold ores will continue to leach whilst exposed to an alkaline environment containing cyanide. Thus, longer residence times lead to lower residue values. The disadvantage of the CIL approach, apart from the flat adsorption profile, is that it results (Stange, Woollacott and Stange 1990) in a lower overall leach residence time.

This case study examines the economic advantages, even with a low-grade ore, of the traditional leach-CIP approach as opposed to the CIL approach.

Figure 6 illustrates the effects of adding leaching stages to the plant. As expected, the capital and operating costs of leaching rise linearly with increasing stages. The adsorption capital cost is seen to fall slightly as the number of leaching stages increases. This is due to the fact that as the number of leach stages increases, less leaching occurs in the adsorption train, resulting in a significantly steeper carbon loading profile. This results in less gold lock-up in adsorption and therefore less adsorption working capital*.

The NPV rises sharply as the number of leaching stages are increased due to two primary effects:

1) The increased leaching extraction due to an increase in residence time
2) The increase in adsorption efficiency as less leaching takes place in the adsorption plant.

**In this financial model, gold locked up in adsorption and elution is deemed to represent working capital equal to the value of the gold locked up, rather than as an operating cost.*

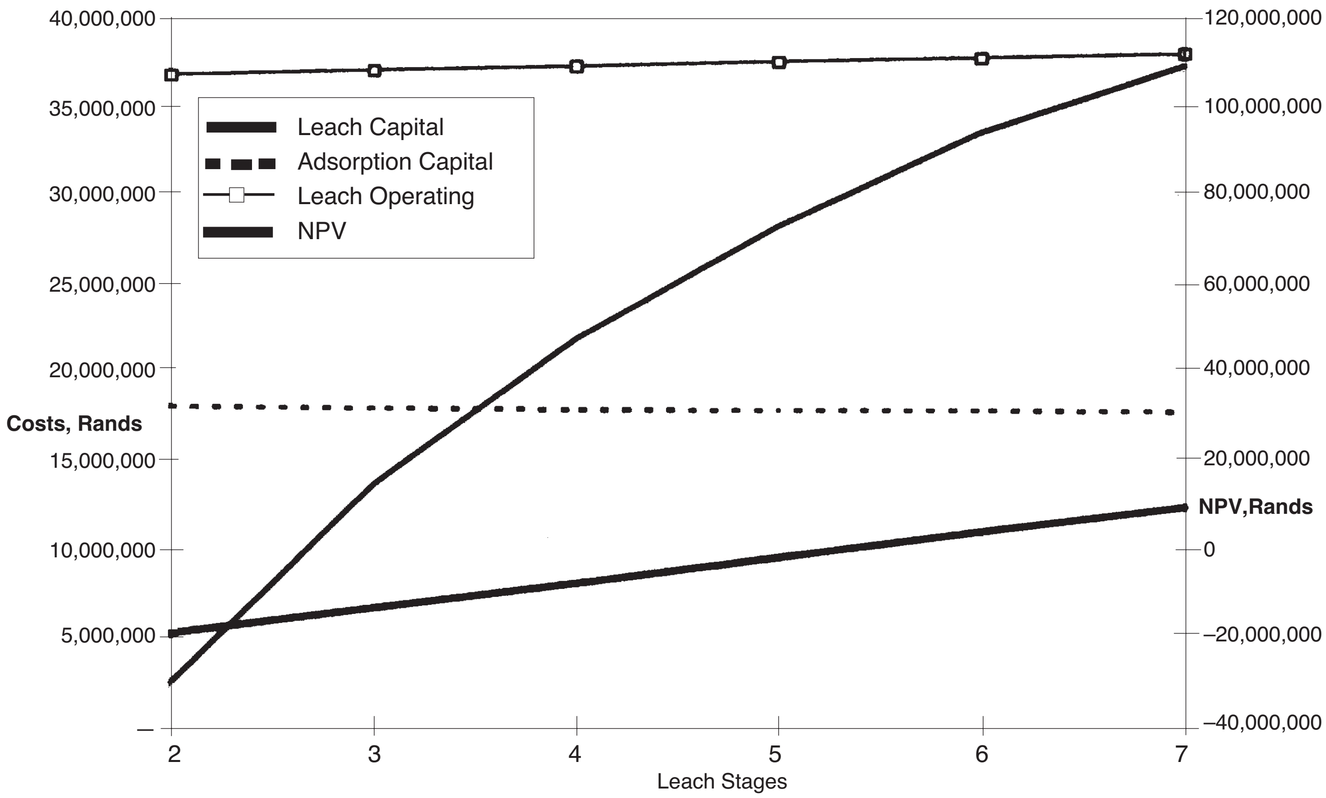


**Figure 6—Effect of leaching stages on CIL/CIP economic**

### *Pump cell versus conventional adsorption plant*

In a high-intensity carbon adsorption plant, such as the pump-cell, the design is such that the plant can be operated at carbon concentrations significantly higher than those employed on conventional plants (Whyte, Dempsey and Stange 1990). This results in a significantly smaller adsorption and elution plant for the same duty. This has advantages; experimental work has shown that the kinetics of adsorption are improved by higher specific mixing power which increases as the size of the adsorption reactor decreases.

The disadvantage with the pump-cell approach for a CIL project is that the overall leaching residence time will be decreased, as the size of the adsorption reactors will be significantly smaller. Thus, to be comparable to the base case more leaching reactors will be required for the pump-cell plant.

Simulation showed that increasing the number of leaching stages from 2 to 6 x 960m$^3$ reactors with a pump-cell adsorption plant consisting of 6 x 100m$^3$ adsorption reactors operating at a carbon concentration of 37 g/l resulted in similar leaching and adsorption extractions. The following Table shows the differences between the base case and pump-cell financial parameters.

It can be seen that leaching capital and operating costs increase due to the increase in the number of adsorption stages. However, decreases in adsorption and elution capital and operating costs due to the smaller physical size and associated performance benefits of the pump-cell plant result in significant decreases in both total operating and capital costs.

This NPV for the pump-cell plant is significantly higher due to both lower capital and operating costs as well as:

1) A longer overall leach residence time as six leach stages are used instead of two. There is, therefore, a resulting increase in the overall leach extraction.
2) The improved kinetics in the pump-cell plant, as well as the increase in the number of leaching stages result in less leaching in the adsorption plant, and consequent improvements in adsorption performance.

### Process engineering methodology

'The successful operation of a carbon-in-pulp plant is centred on the adsorption section which is expected to extract more than 99.6% of the gold present in solution' (Bailey 1987).

This quote reflects the approach that the author has developed towards the design of leaching and CIP/CIL plants. In summary:

*Table III*

| | Base case | Pump cell |
|---|---|---|
| Leach Capex | 5,212,000 | 10,890,000 |
| Adsorption Capex | 17,963,000 | 12,027,000 |
| Elution Capex | 13,935,000 | 6,817,000 |
| **Total Capex** | **37,110,000** | **29,734,000** |
| Leach Opex | 36,755,000 | 37,753,000 |
| Adsorption Opex | 7,911,000 | 7,069,000 |
| Elution Opex | 15,859,000 | 9,620,000 |
| **Total Opex** | **60,525,000** | **54,442,000** |
| **NPV** | **(30,165,028)** | **70,547,878** |

1) Through laboratory, pilot or plant test-work gather sufficient information which will allow a simulator of the proposed plant to be calibrated with confidence, as well as to provide confidence in the design basis.
2) Use the simulator and appropriate financial models to generate an optimum technical and financial design basis.
3) In terms of CIP/CIL plant design, the focus is on optimizing the adsorption circuit technical performance and then ensuring that equipment is sized and selected which ensures that the acid-washing, elution and regeneration unit processes will operate in the manner required.
4) The trade-offs between CIP and CIL are ideally explored using a financial and technical simulator to ensure that the financially optimum approach is used for the particular set of project circumstances.
5) In a similar manner, trade-offs between adopting a more conservative approach to adsorption design* versus the more aggressive pump-cell approach can be examined using financial and technical models. This allows the reward associated with increasing technical risk to be evaluated in a more quantifiable manner. Through this approach, the author has found that many of the financial benefits associated with the pump-cell approach can be attained by more aggressive design parameters using conventional equipment and a carousel system, without the need of a full pump-cell implementation.
6) The most valuable aspect of using a simulator is that it allows calculation of different scenarios allowing the sensitivity of different design decisions to be easily evaluated. As all real plants will see a range of conditions during their lifetime, this allows plants to be designed which are robust to the degree required. The cost of this robustness can be evaluated with the use of the financial model.

Once an appropriate design basis has been established, this is used for the normal process design and engineering activities which take place. The nature and detail of these activities depend on the objectives of the study, i.e. whether it is a feasibility study or a final design. The process engineering aspects of such studies normally consist of the following activities.

➤ *Development of process design criteria*
These criteria define the duty which the plant must perform, the efficiency required, as well as the philosophy which will be adopted in terms of operability, staffing, maintenance, level of automation and so on. These criteria should be developed in close consultation with the client and with full awareness of what state-of-the-art practices and technologies allow.

➤ *Development of process flow diagrams (PFDs)*
PFDs define the basis of the process under consideration and any decisions made during their development will have far-reaching consequences. During this phase it is appropriate that process engineers who have significant experience of gold plants interact extensively with the operator in order to ensure that the proposed process flowsheets meet their requirements. Visits to operating installations, and with equipment vendors, to review recent and innovative developments in the industry can also be useful to ensure that the plant design reflects current best operating practice. Equipment lists are developed from these flowsheets.

➤ *Process descriptions*
These are narratives describing exactly how the process works and are considered to be documentation supporting the PFDs.

➤ *Mass and energy balances*
Development of mass and energy for the developed flowsheets, using the developed process design basis and criteria. This task is made easy by the use of mass and energy balance calculation software.

➤ *Equipment selection and sizing*
Equipment can be sized based on the calculated mass and energy balances for the plant. Equipment selection, where there are competing options, may be based on:

- The preferences of the designers or the operators
- A lifecycle financial evaluation of the alternatives
- Awareness of current industry practices
- Vendor credibility and technical reputation.

What is most important is that the decisions made reflect the outcome of a debate between all stakeholders.

➤ *Equipment specifications*
Once equipment has been sized and selected, equipment specification packages which can be used as the basis for competitive tenders are developed and submitted to selected vendors who are asked to provide quotes or tenders based on this information. Once returned, the appropriate vendor is selected by a committee which evaluates each quote or tender on pre-defined technical, commercial and strategic criteria.

➤ *Layouts and general arrangements*
Once details of selected equipment are available, plant layouts and general arrangements can be completed. Again, strong interaction between various engineering disciplines, process engineers and the future operating company or staff are needed to ensure that the resulting plant layout meets the requirements of all stakeholders. Visiting similar installations to see what works and what doesn't, in practice, as well as benefiting from experiences of other parties may help to improve layout significantly.
It has become commonplace to use 3D CAD tools to provide effective visualization of layout decisions. This technology has the potential, if used appropriately, to provide significantly improved designs as well as to improve the operability of the resulting installation.

➤ *Process and instrumentation diagram* (P&ID) *development*
P&IDs are typically developed for more detailed design

**Conventional design is the use of the 'formula' used to design many CIP plants i.e., 6-8 stages, 1 hour pulp residence time per stage, 20-25 g/l of carbon per stage with a loaded carbon value of 1 000 times the solution feed tenor.*

studies. Each PFD is typically decomposed into several P&IDs where each P&ID illustrates:

- All mechanical equipment in that area, including standby equipment with major parameters associated with each item of equipment. Each item of equipment is uniquely identified by whatever tagging convention has been selected
- Instrumentation and major control loops
- Valves and piping, including materials of construction, type in the case of valves and size. An appropriate tagging convention is used to ensure that this information can be portrayed in the most concise and effective manner.

A number of detailed lists can be derived from the P&IDs (equipment, instruments, process signals). After appropriate specification these lists are used as the basis on which items are cost estimated and procured.

A process and control philosophy is developed for each P&ID which describes the manner in which that plant area is intended to be operated and controlled in detail. Process control functional specifications are developed from this which serve as the basis for the implementation of the process control system.

Experience has shown that P&ID development can be accelerated significantly, and the quality of the output improved significantly, by utilization of tools like dynamic simulation as well as the application of CAD tools which support the development of 'intelligent' P&IDs.

➤ *Engineering design and cost estimation*
The information derived from the process engineering activities are used as a basis for engineering designs in the areas of civil, structural, mechanical, electrical and process control. Depending on the type of study (pre-feasibility, feasibility etc.), different cost estimation techniques can be used. These range from factoring costs based on major equipment costs, to detailed costs where a full commercial enquiry is issued for each item identified by taking off items from the defined process.

## Conclusions

This paper has described in some detail the issues related to the design of gold leach and CIP/CIL plants. It has been shown that due to the complex and interactive nature of these processes, a good basic understanding of the factors which affect the process phenomena is essential. In addition, the development of a good design basis by an appropriate laboratory, pilot or full-scale test-work is crucial.

The application of modelling and simulation techniques (technical and financial) allows these data to be used to evaluate the technical and financial interactions within the design space under consideration, allowing a more effective design to be synthesized. Standard process engineering methodologies can then be used to convert this information into an appropriately designed, engineered and costed process plant.

## References


ADAMS, M. The chemistry of the CIP process. Ph.D. thesis, University of the Witwatersrand, 1990.

BAILEY, P. 'Application of activated carbon to gold recovery'. *The extractive metallurgy of gold in South Africa*. Stanley, G.G. (ed.), Johannesburg, *South African Institute of Mining and Metallurgy*, 1987. Chap. 9.

DAVIDSON, R. Anglo American Research Laboratories. Personal communication. (1988).

JOHNS, M.W. The simulation of gold adsorption by carbon using a film diffusion model. M.Sc. thesis, University of the Witwatersrand, Johannesburg 1987.

KNOTTENBELT, P.R. Population balance models for particulate processes in hydrometallurgy. M.Sc. thesis. University of the Witwatersrand, 1995.

LE ROUX, J.D. The modelling of the effect of organic fouling on the adsorption of gold cyanide by activated carbon in a batch reactor. M.Sc. thesis, University of the Witwatersrand, 1989.

MENNE, D. Predicting and assessing carbon-in-pulp circuit performance. *XIV International Mineral Processing Congress*, Toronto, Canada, 1982. pp. 17–23.

NICOL, M.J., FLEMING, C.A. and CROMBERGE, G. The adsorption of gold cyanide onto activated carbon. I. The kinetics of adsorption from pulps, *J. S. Afr. Inst. Min. Metall.*, vol. 84, no. 2, 1984, pp. 50–54.

NICOL, M.J., FLEMING, C.A., and CROMBERGE, G. The adsorption of gold cyanide onto activated carbon. II. Application of the kinetic model to multistage adsorption circuits. *J. S. Afr. Inst. Min. Metall.*, vol. 84, no. 3, 1984. pp. 70–78.

PEEL, R.G. and BENEDEK A. A simplified driving force model for activated carbon adsorption. *Canadian J. Chem. Eng*. vol. 59, 1981. pp. 688–692.

STANGE, W. 'Towards more effective simulation of mineral processing systems'. M.Sc. dissertation, University of the Witwatersrand, 1988.

STANGE, W. 'A Techno-Economic Framework for the design and optimization of the Carbon-In-Pulp Process'. Ph.D. thesis, University of the Witwatersrand, 1998.

STANGE, W. and KING, R.P. A population balance approach to the modelling of the CIP process. Proceedings of the 20th International Symposium on the Application of Computers and Mathematics in the Mineral Industry (APCOM). *SAIMM*, Johannesburg, October 1987. pp. 209–221.

STANGE, W.W. and KING, R.P. The Optimization of the CIP process. *XXII International APCOM Symposium*, Technical University of Berlin. (1990).

STANGE, W.W., KING, R.P. and WOOLLACOTT, L.C. Towards more effective simulation of CIP and CIL processes. 2. A population-balance-based simulation approach. *J. S. Afr. Inst. Min. Metall.*, vol. 90, no. 11. Nov 1990. pp. 307–314.

STANGE, W.W., WOOLLACOTT, L.C. and KING, R.P. Towards more effective simulation of CIP and CIL processes. 3. Validation and use of a new simulator. *J. S. Afr. Inst. Min. Metall.*, vol. 90, no. 12. Dec. 1990. pp. 323–331.

VAN DER MERWE, P.F. Fundamentals of the elution of gold cyanide from activated carbon. Ph.D. thesis, University of Stellenbosch, 1991.

VAN DEVENTER, J.S.J. Kinetic models for the adsorption of metal cyanides on activated carbon. University of Stellenbosch, Ph.D. thesis, 1984. 337 pp.

VAN LIER, W.C. On the kinetics of adsorption on activated carbon from the aqueous phase. In: *Activated Carbon—A fascinating material*. (Eds.) A. Capelle and F. De Vooys, Norit, Amersfoort, 1983.

WHYTE, R.M., DEMPSEY, P. and STANGE, W. The development and testing of the AAC 'Pump-Cell' at Vaal Reefs Exploration and Mining Company. International Deep Level Mining Conference, *SAIMM*, Johannesburg, September 1990.

WILLIAMS, D. and GLASSER, D. The modelling and simulation of processes for the adsorption of gold by activated carbon. *J. S. Afr. Inst. Min. Metall.*, vol. 85. no. 8, 1985. pp. 237–243.

WILLIS, J.A., MOLONEY, M.J., MORRISON, R.D., CORRANS, I.J. and NAPIER-MUNN, T.J. Development of a dynamic simulator for CIP/CIL plant optimization and design. *Australian Institute of Mining and Metallurgy, Extractive Metallurgy Conference*, Perth, October 1991

WOOLLACOTT, L.C., and AFEWU, K.I. Scale-up procedures for gold-adsorption systems. Part 1 : Adsorption kinetics, *J. S. Afr. Inst. Min. Metall.*, vol. 95, no 4, 1995. pp. 167–177.

WOOLLACOTT, L.C. and ERASMUS, C.S. The distribution of gold on loaded carbon. *J. S. Afr. Inst. Min. Metall.*, vol. 92, no. 7, 1992. pp. 177–182.

WOOLLACOTT, L.C. and GUZMAN, G.N. 'The isotherm shift in carbon-in-pulp adsorption circuits'. *J. S. Afr. Inst. Min. Metall.*, vol. 93, no 8, 1993. pp. 185–193.

WOOLLACOTT, L.C., STANGE, W. and KING, R.P. 'Towards more effective simulation of CIP-CIL processes, Part I: The modelling of adsorption and leaching'. *Journal of the South African Institute of Mining and Metallurgy*, October 1990.

YOUNG, G.J.C. 'Cyanidation'. The extraction metallurgy of gold in South Africa. Stanley, G.G. (Ed.), Johannesburg, *South African Institute of Mining and Metallurgy*, 1987. Chap. 6. ◆

## Solution to mine water problem on the cards*

According to Mintek's 1998 Annual Review, Savmin water-treatment technology developed by Mintek, the Wren Group and Savannah Mining (Australia), which has been tested successfully at the mine, could provide a solution to the mine-water problem at Grootvlei Gold Mine on the East Rand; the dissolution of which is fast reaching a showdown between the mining company and government. Results of the tests have conferred sufficient confidence for the consortium to plan further tests on other arisings of polluted mine water (mainly from gold and coal mining operations).

Grootvlei pumps up to 120 million litres per day of underground water containing high concentrations of iron, calcium, and sulphate ions. The water is currently treated to reduce the suspended solids and iron content before it is released into the Blesbokspruit wetland. However, the levels of dissolved calcium, magnesium, and sulphate in the discharged water make it unfit for human consumption or industrial use, and inhospitable to aquatic life.

Grootvlei is a marginal mine, and the de-watering costs (running into millions of Rands) are currently subsidized by the government, but pumping subsidies have been falling, affecting the profitability of the mine.

In order to sustain the subsidy, the mine has to demonstrate a viable long-term solution to the water problem. A decision to stop pumping would not only result in the loss of at least 6000 jobs in mining and support industries in the economically depressed area, but the water table would rise and, within a few years, polluted water would spill over into the Blesbokspruit or the Nigel region to the south of the mine.

In the Savmin process, once all the heavy minerals and magnesium have been removed, calcium and sulphate are

*Evaluation of the SAVMIN process for treating acid mine drainage at Grootvlei Gold Mine*

precipitated in the form of the mineral ettringite. After solid-liquid separation, the water only requires treatment with carbon dioxide to make it totally acceptable for disposal into sensitive natural water courses. Further simple processing would make the water suitable for domestic consumption. The operating costs of the Savmin process are comparable to the Rand Water Board's price for fresh water, and the capital cost is expected to be significantly lower than that of other available technologies.

Estimates of the total quantity of the mine drainage water on the Witwatersrand vary considerably, but it would be enough to supply water to several million people, if purified.

Contact Dr Mike Dry for further information ◆

* *Issued by: Communications Division Mintek, Tel: (011) 709-4111 Fax: (011) 709-4326*

## Women on the move at Chemical Engineering, UCT

Lucy Wood, final year Chemical Engineering Student at the University of Cape Town's Chemical Engineering Department was selected as the winner of the SASTECH Award for the best undergraduate Practical Training Report.

This is the third year in succession that UCT has won the SASTECH award and it is the fifth year in a row that the award has been won by a female student.

Currently, a final-year student at UCT on a bursary from Anglogold, Lucy Wood was educated in Johannesburg at St Mary's School for Girls.

'My parents had a profound effect on my career choice. My father has a Ph.D. in Chemistry and has worked in the Chemical Engineering Industry for a number of years, whilst my mother is an English-History major and has worked as a teacher and journalist for many years. They have always encouraged me to work hard and develop all areas of my life.'

Lucy did her compulsory third-year vacation job at ERGO (East Rand Gold and Uranium Co. Ltd) in Brakpan. Her project was CIL Elution Optimization and she worked under the supervision of Clint Armstrong and Harry Bezemer.

'The Anglo Student Training Programme has been excellent and has stretched me both mentally and physically. The standard of work that they expect is very high, and we were required to do presentations in front of directors and our peers, which obviously encouraged excellence. UCT also instilled in me a self-motivating and well-rounded attitude to my studies. The Chemical Engineering Department taught me skills like time management, and application of theory to practical situations and the first principle engineering concepts.'

Lucy received her award at the South African Institution of Chemical Engineers' Annual Dinner held last year on the Ist October in Johannesburg. ◆