CHAPTER 10.12

Carbon-in-Leach and Carbon-in-Pulp Circuit Design

Stuart Smith

Several factors influence the design of a carbon adsorption circuit outside the realm of the pure process engineer's task of providing a functional design. These include capital expenditure (CAPEX), operating expenditure (OPEX), and risk. The design has to provide the desired financial outcome, not the optimum process solution.

The design of a facility is often compromised by the need to minimize entry capital. The discipline of process engineering is similarly compromised. A successful design needs to address the engineered solution as a holistic blend of process engineering and fiscal engineering.

To effect a design, a set of process design criteria (PDC) is required. Much of the PDC comes from metallurgical test work. Other components, such as design throughput, come from analysis undertaken during the various phases of study and project definition.

PROCESS DESCRIPTION—CIL AND CIP

An example of a generic carbon-in-pulp (CIP) process flowsheet diagram (PFD) is presented in Figure 1. The only difference between the CIP and carbon-in-leach (CIL) flow sheet is that CIL would not have the three leach tanks prior to the carbon tanks.

Ground or scrubbed ores in slurry form are sent to the trash screen. Typically, the pH of the slurry has previously been adjusted to provide protective alkalinity prior to the addition of cyanide. The pulp density of the slurry usually lies in the range of 48% to 50% solids w/w if subjected to pre-leach thickening. If the slurry results directly from a cyclone overflow, it can be expected to be in the range of 40% to 42% solids w/w.

The trash screen scalps out oversize material and directs it to an oversize bunker. Trash screen undersize slurry is directed to the first of the CIL/CIP tanks via a distributor, which allows the slurry to be directed to the second tank should the first tank be off-line. The distributor often receives cyanide solution, pH modifier, and other streams such as sump pump spillage return or spent eluate return.

Tanks are typically agitated with axial flow mechanical agitators submerged in the slurry. The tanks are generally configured with a height-to-diameter aspect ratio of around unity. Agitators are centrally and vertically mounted in the open tank, with three or more baffles provided around the tank periphery to stop vortexing.

Air and/or oxygen gas is injected into the tanks to provide oxygen for leaching. Various arrangements exist, from open pipes simply submerged in the tank to down-agitator shaft installations and more elaborate diffusers. Some ores respond favorably to elevated oxygen concentrations in that leach kinetics and/or extraction is improved. Cyanide consumption may be reduced if reactive sulfide minerals are present. Some ores require such high levels of oxygen that the volumes of air that would be required to satisfy the demand would impede the agitator operation, in which case, use of oxygen gas or some other oxidant is mandatory.

The slurry passes from one tank to another usually by means of a series of launders. The launders are configured so that tanks can be bypassed. A staggered tank arrangement simplifies this bypass arrangement and is the most common layout applied. An alternative is a carousel arrangement, which is described briefly below.

In each of the carbon contactors, some form of inter-tank screen resides to retain the activated carbon. The inter-tank screen aperture is coarser than the trash screen so that all of the ore particles should be able to pass the inter-tank screen. The flow of slurry through the inter-tank screen holds the carbon against the screen surfaces; consequently, some form of "sweeping" is required to keep the screen area open.

As the precious metals leach from the ore, the slurry passes from one tank to the next. Transfer is usually by gravity flow. In some instances, flow is assisted or totally dependent on the use of pumping inter-tank screens.

In the pure CIL case, the slurry is contacted with carbon at all times as the precious metals leach. This results in lower solution tenors of precious metals, as the carbon is present at all stages of leaching. The slurry continues to flow down the tank farm, through the inter-tank screens, to eventually discharge over the carbon safety screen. The carbon safety screen's function is to capture activated carbon that may have

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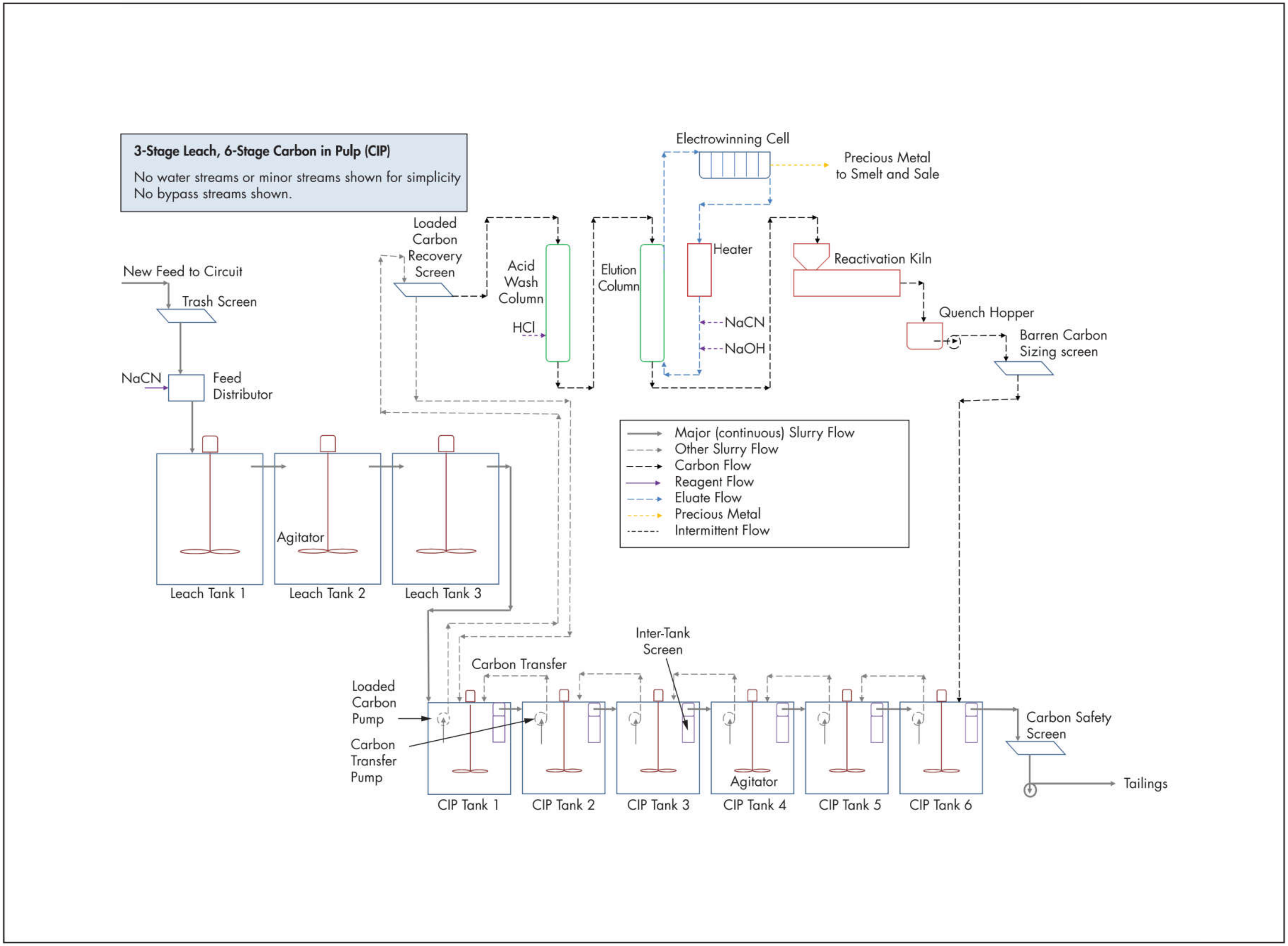

Figure 1 Generic CIP process flow-sheet diagram

escaped the tank farm because of a hole or a badly seated inter-tank screen.

In the pure CIP case, the slurry flows through the leach tanks before entering the carbon contactors. The precious metal leaching is effectively complete when the slurry contacts the carbon. The resulting solution tenors are much higher than a comparative CIL case. This results in higher carbon loadings of precious metals.

Barren carbon (ex-elution/reactivation or new) is added to the last contactor. This carbon, having a low precious metal loading, will continue to adsorb the low-solution concentrations in the last tank. As this carbon loads up, it will start to approach the equilibrium loading associated with the solution tenor of the tank in which it resides, becoming ineffective. Carbon is transferred to the next upstream tank, where the solution tenor is higher and the carbon loading is now lower than the equilibrium loading. Hence, the carbon will adsorb the precious metals until it again starts to approach equilibrium. The countercurrent transfer of carbon continues until the carbon reaches the first carbon contactor and becomes fully loaded with precious metals.

Carbon is transferred by pumping slurry and the carbon contained in it from the body of the downstream contactor up and into the upstream contactor. This carbon transfer is usually achieved by use of an airlift or a recessed impeller pump. The transfer results in slurry being transferred upstream as well. This results in dilution of the upstream tank solution tenor and this is reflected in the inefficiency of the contactor circuit. This back-mixed slurry will pass back through the inter-tank screen and return to the tank from which it originated.

Loaded carbon in the first contactor is recovered by use of an airlift or a recessed impeller pump. The slurry is pumped to the loaded carbon recovery screen. Here, the loaded carbon is separated and washed while the slurry component flows back to the tank from where it came. The washed loaded carbon gravitates to the acid-wash column/hopper. Some operations perform the acid wash after elution, others beforehand. The PFDs presented here assume that acid washing is pre-elution.

Carbon is acid washed and then transferred to the elution column as a water–carbon slurry. The elution circuit is generally an Anglo American Research Laboratories– (AARL) or Zadra–based process. These processes are described in more detail below. Both utilize elevated temperatures and high-sodium-concentration liquors to desorb the precious metals off the carbon and back into a high-grade solution.

The high-grade solution is passed through electrowinning cells where the precious metals are precipitated onto the cell cathodes. The metal is periodically recovered from the cathodes and smelted to produce doré bars of varying purity as final mine product. Refining is typically conducted by specialist refining companies off-site.

Eluted carbon is transferred as a water–carbon slurry to the reactivation kiln. The carbon is dewatered over a screen and reports to the kiln feed hopper. Heat and a reducing steam environment is applied by the kiln to increase the carbon activity. The carbon is typically quenched in good-quality water and then transferred to the last contactor as a water–carbon slurry.

The carbon is screened prior to reentering the last carbon contactor to remove fines. Some operations screen the kiln discharge dry. This is a more efficient sizing operation than the wet process.

FACILITY DESIGN

The following sections discuss the various aspects that need to be considered when selecting a design philosophy for a CIL or CIP facility. The advantages and disadvantages are noted for consideration.

The design process needs to consider the variability of the ores that will be fed to the carbon circuit. Considerations include

- Variable metal flow due to differential ore throughput for, say, oxides compared to harder primary ores;
- Preg-robbing tendencies that would indictate CIL over CIP;
- Ores that are sensitive to the actual solution tenor present (Le Chatelier's principle), suggesting that CIL may be advantageous over CIP;
- A CIP facility that requires less carbon to be eluted, as there is more precious metal adsorbed per unit mass of carbon;
- A CIP having less carbon and precious metal inventory;
- The financing strategy to be undertaken influencing the process design;
- Downstream processes such as cyanide detoxification, tailings deposition, and water balance considerations.

Pulp Density Considerations

The pulp density of the slurry to be leached and processed will influence the design of the carbon circuit. In some low-CAPEX designs, the comminution circuit hydrocylone (cyclone) classification is compromised and runs at high overflow pulp densities. The benefit is the deletion of a pre-leach thickener.

The advantages of low-pulp-density circuits are

- Reduced entry capital with regard to thickening,
- Enhanced precious metal and reagent recycle if a tailings thickener is employed,
- Reduced viscosity, and
- Reduced preg-robbing (reduced solution tenor).

Elevated pulp-density circuits have advantages of

- Higher solution tenors increasing carbon loadings,
- Reduced tank volumes,
- Reduced reagent consumption,
- Reduced solution losses, and
- Improved tailings deposition.

Trash Screening

The task of the trash screen is to remove the near-size particles that may block the inter-tank screens as well as coarse oversize that can cause tank sanding issues and elution circuit problems.

Trash screens are typically linear vibrating screens (high frequency) fitted with long-life polymer decks. Wire decks are sometimes used but are generally higher wear and harder to maintain. Linear belt screens are also used, particularly when large quantities of vegetation is present. Trash should not be directed to the carbon circuit bund, as cleanup of spills will result in the trash entering the carbon circuit.

The unit process of trash screening is one that is often compromised, both in design and in operation. Design compromises include

Figure 2 Contaminated activated carbon

- Using minimal size equipment to reduce CAPEX but also to address engineering considerations such as vibrating load; and
- Use of slotted polymer screens to reduce the total screen dimensions, as the slots provide increased open area over square apertures.

Slotted decks should be avoided when processing ores with vegetation present, significant nonelectric explosive waste (nonel), or flaky minerals such as micas. Slotted trash screen decks should not be used when the inter-tank screen apertures are square.

Sieve bends should never be used as a trash screen. The action of a sieve bend is such that to cut at, say, 600 μm, a 1,000-μm slot is required. A sieve bend will allow a minor percentage of material up to the slot size to pass, which is only exacerbated as the screen wears.

Given the recirculating nature of the carbon circuit, any minor bypassing of trash will quickly result in a buildup in the circuit, which is difficult to reduce and almost impossible to totally remove. The issue with coarse buildup in the circuit can be a major operational issue and one not to be underestimated.

Figure 2 presents the deck of a loaded carbon recovery screen. It is apparent that the activated carbon has a lot of coarse sands mixed in and the screen apertures are blinded with sands as well as fine carbon. The sand in the carbon causes issues in the acid wash, elution, and reactivation circuits as do ore slimes that are carried over because of poor screen efficiency. Silicate fouling of carbon can result, and this has an irrecoverable impact on the carbon circuit. The heart of the carbon circuit is the carbon itself, and the cleanliness of this carbon is key.

Tankage and Agitators

CIL/CIP tanks are most commonly agitated with axial flow mechanical agitators. Various impeller configurations are employed. Other less common agitation alternatives are

- Draft tubes, which involve elevated CAPEX but can have OPEX benefits over open tank agitation;
- Air sparge, which is a high-OPEX option and is impractical for large tanks; and
- A Pachuca tank (air), which encompasses high OPEX and is not suited to large tanks.

As tank height-to-diameter ratios increase above unity, the effective cost of the tankage increases per unit volume. This includes costs assocated with structural steel, electrical systems, and in supporting vibrating loads. Open agitated tanks having a height-to-diameter ratio nearing unity have a capital advantage over draft-tube or pachuca/air-agitated tanks. The latter agitation methods require tanks of greater height-to-diameter ratio.

Tank materials of construction are almost exclusively mild steel. The high pH of the process provides corrosion protection for unlined or uncoated mild steel. Under some circumstances, anaerobic biological oxidation can occur. The biomass establishes a favorable environment under an isolating film on the inside of the tank wall, which results in pitting corrosion. Elevated pH values of +10.5 are often necessary to reduce the impacts of bacterial attack (Mardhiha et al. 2014).

Coarse abrasive slurries can wear the lower strakes of mild steel tanks. Poor trash screening practices exacerbate this wear issue. Such coarse material can also settle out in the corners of the tank. This can result in entrapment of activated carbon and exacerbate corrosion activity.

Carbon tanks can be configured in a carousel arrangement such that the tanks are fed by a distributor to the first contactor. Slurry then flows to the next tank in the carousel, and so on. Instead of transferring carbon, the slurry flow path is changed; consequently, what was the second tank now becomes the first tank in the series. What was the first tank is drained or carbon transferred for elution. The tank is then reestablished with the addition of barren carbon and then the tank is inserted to the end of the carousel as the last tank. The cycle repeats and the tanks are repeatedly moved up in the sequence to maintain the carbon contact with higher and higher solution tenors.

Oxygen Source and Introduction

The leaching of gold and silver requires oxidation of the metals prior to complexation with cyanide. Oxygen is the typical oxidant and can be supplied by the ambient atmosphere, aeration into the tanks, oxygen gas introduction to the tanks, or disassociation of hydrogen peroxide introduced to the tanks.

Part of the metallurgical test-work phase of a project should include measurement of the oxygen demand of the ore (Fraser 1984). Equilibrium dissolved-oxygen concentration decreases with rising temperature and increasing altitude, and so the site conditions also need to be considered. If oxygen gas is used over air, there may be advantages such as

- Faster leach kinetics (elevated carbon loadings and reduced tankage volume),
- Improved circuit efficiency (reduced back-mixing),
- Improved final extractions,
- Reduced soluble losses (steeper solution tenor profile), and
- Reduced cyanide consumption.

Disadvantages can include

- Increased capital cost,
- The need to observe specific design considerations,
- Elevated operating cost potential, and
- Gold passivation and elevated cyanide consumption if the conditions are too aggressive.

Oxygen adsorption into the slurry from the ambient atmosphere will only provide adequate oxygen supply to the most benign of oxygen consumers. As modern plant tankage has increased in size, the lower surface area of the tank per unit volume of slurry contained has made this option impractical. Air/oxygen sparging options include

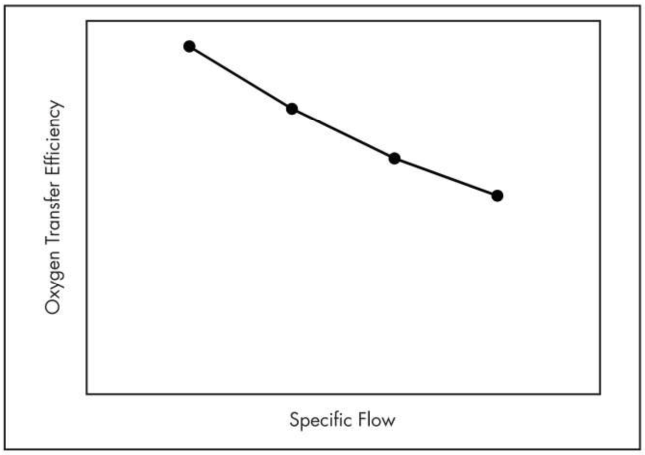

Figure 3 Aeration efficiency

- Proprietary lance-based sparging methods,
- Open-lance-based options,
- Gas introduced down the center of the tank agitator shaft,
- Sparge rings under the tank agitator, and
- Shear reactors.

Figure 3 shows test-work data for a given open-sparge system as the airflow increases. A nominal 100% increase in the airflow results in a decrease in transfer efficiency of around 25%. Doubling the air addition, therefore, only provides a 50%–60% increase in oxygen mass transfer.

Air-based systems may achieve 5%–15% utilization for oxygen transfer, although vendors of sparging equipment may claim higher values. Oxygen-gas-based systems often achieve utilizations of greater than 80%, even for moderately basic lance systems. Oxygen transfer is more efficient because the volumes introduced are lower than the air equivalent, and the equilibrium concentration of oxygen systems is some five times greater than air.

The design of the oxygen supply needs to consider the following:

- The oxygen demand, temperature, and viscosity of the slurry
- Cost of generation of air or oxygen
- Oxygen supply purity
- Altitude
- Utilization/transfer efficiency

Inter-Tank Screens and Launders

The task of the inter-tank screen is to retain the coarser activated carbon in the carbon contactor while allowing the slurry to pass. Various configurations of inter-tank screens have been developed over the years and have included options such as air-swept woven or wedge-wire screens. Cylindrical, flat panel, and V-shaped options exist. Mechanically swept screens are the most common. These are generally vertical cylinder screens of woven or wedge wire and have a set of external noncontacting, rotating wiper arms.

Some mechanically wiped screens include what is effectively a low-head/high-volume pump impeller to provide the small amount of head necessary to overcome the head loss of the inter-tank screens and launders. Capital cost and complexity of the screen are off-set by savings in the tankage and tank-top steel costs.

The inter-tank screen working area has to be estimated with consideration of the volume of slurry being processed, but also the viscosity and the back-mixed volume resulting from carbon transfer. The inter-tank screen apertures are normally square for woven wire or slotted for wedge wire. The woven-wire aperture is normally 1 × 1 mm, whereas wedge-wire slots may be finer in the short dimension, around 800–900 µm.

Inter-Tank Carbon Transfer

The two dominant methods for transfering carbon between the carbon contactors are airlifts or recessed impeller pumps. The advantages of the airlift are low CAPEX, simplicity, and relatively low maintenance. Recessed impeller pumps are higher CAPEX but are less abrasive on the carbon and are capable of higher flow rates.

Carbon transfers are often conducted in short periods of time because it is believed that this minimizes the upset conditions due to back-mixing. It is also common to paritally empty a contactor of carbon prior to transferring carbon to it from downstream to minimize short-circuiting of carbon. Both of these practices actually reduce the efficiency of the carbon circuit and are to be avoided as a basis of design or operation. Carbon transfers should be designed to be continuous but with adequate capacity to catch up in the event of circuit disturbances. This diminishes CAPEX of the transfer equipment, and inter-tank screen capacity is reduced.

Carbon Safety Screen

It is common practice to use a vibrating screen for the carbon safety screen duty. It should also be the same model and size as the trash screen to allow commonality of spares. The carbon safety screen aperture is often the same as the inter-tank screens or, preferably, slightly coarser. Slotted polyurethane panels are most commonly used.

Screen Aperture Selection

Several factors must be considered regarding screen aperture shape, materials of construction, and wear. The plant designer must allow for practical considerations and not just design for a "new" plant where reliable aperture dimensions are present.

Screen aperture shapes for the trash, inter-tank, and carbon safety screens, and even the loaded carbon recovery screen, need to be considered holistically:

- The trash screen aperture must be the finest used of all the various applications.
- Square apertures may be necessary to deal with certain trash types.
- If a square aperture inter-tank screen is to be used, then the trash screen aperture must be square to save passing of flaky material that will not pass the inter-tank screen.
- The inter-tank screen aperture should be at least 100 µm coarser than the trash screen. The trash screen will wear and pass coarse particles with time that a new inter-tank screen cannot.
- The carbon safety screen should be 100–200 µm or even coarser than the inter-tank screen. As inter-tank screens wear, they will pass trash and coarse minerals that have entered the circuit, resulting in excessive fines being caught by the carbon safety screen.
- The loaded carbon recovery screen should be the same aperture as the inter-tank screen at most, but preferably 100 µm or so finer to minimize the recirculation of

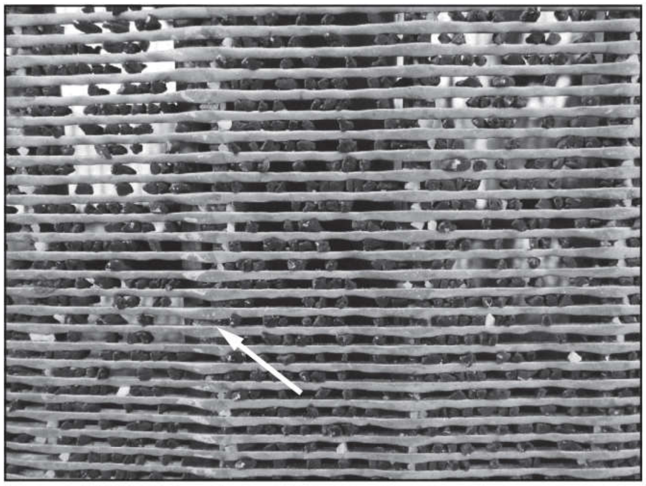

Figure 4 Blinded and worn inter-tank screen

near-size carbon too coarse to pass the inter-tank screen but too fine to be retained by the partially worn loaded carbon screen.

- The barren carbon sizing screen should be the aperture of the inter-tank screen plus 200 µm. This deals with unavoidable inter-tank screen wear.

Figure 4 presents an example of a partially blinded and badly worn wedge-wire inter-tank screen. The effective short dimension of these apertures is approximately 2 mm, double the original new-screen aperture. The white arrow points to areas retaining the original sub-1-mm aperture adjacent to one of the internal support wires.

Even if the allowances for wear in the various screening applications suggested above were applied, such high wear would still result in losses of fine carbon from the contactors and excessive amounts of carbon and trash being recovered on the carbon safety screen. The recovered carbon is too fine to use back in the circuit unless the inter-tank screens are replaced. This is an example of why practical considerations need to be made when selecting screen apertures.

LOADED CARBON RECOVERY, ACID WASH, AND ELUTION

Loaded Carbon Recovery

The first stage of the acid wash and elution process is the loaded carbon recovery step. Loaded carbon recovery is usually undertaken in 4–6 hours. The first contactor should have adequate carbon present so that partial removal takes the population of carbon down to nominal design levels to minimize the impact to contactor performance.

Recessed impeller pumps are the most common means of loaded carbon recovery. The loaded carbon recovery screen location and height usually preclude the use of airlifts. The loaded carbon recovery screen must produce a very clean carbon product. The efficiency of the acid wash, elution, and even reactivation process are all impacted by the quality of the carbon recovered. Carbon containing a large percentage of fines can

- Consume additional acid,
- Generate soluble silicates that permanently foul the carbon and cause electrowinning issues,
- Result in corrosion issues,
- Aggravate heat exchanger and solution heater blockages,
- Reduce heat transfer in the elution and reactivation kiln due to accrection, and
- Make cell cleanout and final smelting onerous.

The slurry containing the loaded carbon should be sent back to the tank it originated from to avoid short-circuiting (downstream return) or total residence time compromises (upstream return).

Acid Wash

As most CIL and CIP operations use quicklime or hydrated lime for pH control, the activated carbon can foul with calcium carbonate. The acid wash is for removing the calcium carbonate and other acid-soluble foulants, typically prior to elution to keep these ions out of the elution circuit.

The acid wash vessel is normally configured to receive the high-value loaded carbon directly from the loaded carbon screen discharge chute to avoid abrasion from transfers. Acid washing can be conducted in open-top vessels, low-aspect closed tanks, or columns that often have an aspect ratio the same or similar to the elution column. Many operations have used a rubber-lined column to undertake both acid washing and then elution. Although cost-effective with regard to equipment, there remains a risk of acid and cyanide mixing should the acid wash/elution sequencing be upset.

Acid washing is usually conducted using a nominal 3% w/w hydrochloric acid solution. Materials of construction therefore need to be resistant to low pH and also chloride attack, which negates the use of typical stainless-steel wetted parts. Fiberglass, high-density polyethylene and rubber-lined steel are common materials.

Dilute acid is recirculated, generally upflow, through the bed of carbon, typically for 2 hours at 2 BV/h (bed volumes per hour). Raw water is then used to rinse the carbon and reduce the pH. The water rinse continues until the pH of the acid wash discharge is greater than 5 or preferably higher.

It has become more common to pass a final dilute sodium hydroxide rinse through the carbon to ensure that the pH is alkaline at the end of the wash-rinse cycle. Once the acid wash is complete, the carbon can either be transferred to the elution column, or if the plant is single column, the carbon remains in situ, ready for elution.

Elution Options

Several methods are used to elute precious metals from activated carbon. The three most common methods discussed in this section are atmospheric Zadra, pressure Zadra, and AARL elution. Other methods employed include the Micron Research method (alcohol-based elution), sodium sulfide–based elution (Adams 1994), variants that use alcohols and other accelerants, and variants that use pressurised electrowinning (Fast 1989).

Atmospheric Zadra

The Zadra process was developed by J.B. Zadra and colleagues while working for the U.S. Bureau of Mines in the 1950s (Zadra et al. 1952). Atmospheric pressure Zadra stripping is considered to be the first commercially successful process developed for stripping gold from carbon. The Zadra process consists of continuous (upflow) circulation of a nominal 1% sodium hydroxide and 0.1%–0.5% sodium cyanide solution at as high a temperature as possible before the solution starts to boil or flash, which is generally around 98°C, depending on the altitude of the installation. Flow rates are in the 2–2.5 BV/h rate, and the circuit needs to operate for 2 to 3 days to achieve low barren-carbon values.

The reagents are mixed in a sump and circulated as the solution is heated. The eluate is pumped upflow through the elution column. The aspect ratio of the column should be at a ratio of 6:1 height to diameter to facilitate plug flow. As the solution circulates, the precious metals are desorbed and enter the eluate primarily due to the sodium saturation coupled with the high temperature as well as the influence of the elevated pH.

The eluate discharges the top of the elution column and is directed to electrowinning. The electrowinning cell (or cells) results in the cathodic precipitation of the precious metals from solution. The elevated pH provides protection to the anodes from acid attack due to the evolution of hydrogen. A supplementary sodium hydroxide dose is common during the elution process to replace consumed reagent due to acid consumption.

The barren eluate is pumped through a heater/heat exchanger to replace any heat lost and then recirculated back up through the column. This cycling of the eluate is continued until the pregnant solution has a gold grade on the order of 5–10 mg/L and/or the electrowinning discharge is less than 5 mg/L. Heating is stopped and the solution continues to circulate to lower the temperature of the eluate and ensure that the heating circuit has no hot spots, which can result in damage to the heater. Once cooled, the carbon can be pumped or pressure transferred to the reactivation kiln. Pressure transfer is the preferred method, as it results is less carbon attrition.

Figure 5 shows the more complex pressure Zadra flow sheet, which is described in the following section. However, this flow sheet is effectively the same as the atmospheric Zadra, except for the heat exchangers shown in gray. The advantage of the atmospheric Zadra is low-entry capital and simplicity of operation, resulting in high reliability and utilization. The disadvantages include low elution rate, sensitivity to elevation (elution temperature), low electrowinning efficiency, elevated barren carbon grades, and low carbon activity post-elution.

Pressure Zadra

The pressure Zadra process is basically the same process as the atmospheric Zadra except that the heating and column components are pressurized to allow temperatures in excess of the solution atmospheric boiling point to be utilized, typically 110°–135°C. This greatly accelerates the elution rate such that the actual elution stage can be conducted in 12–16 hours. Allowing for transfers of carbon to and from the column and heat-up time, a pressure Zadra elution can generally be completed in 18–20 hours.

The use of pressurized equipment has increased mechanical design demands, which also adds to the CAPEX and complexity. In some jurisdictions, the temperature and pressure may influence what operator and maintenance expertise is necessary to operate a pressure Zadra (or AARL) plant. It is necessary for the designer to be familiar with such regulations because they may influence design and process selection.

Although there is additional cost to a pressure Zadra, a much smaller-capacity facility can be utilized compared to an atmospheric unit, given that three times the elution frequency is achieved. Pressure Zadra has the following advantages over atmospheric Zadra:

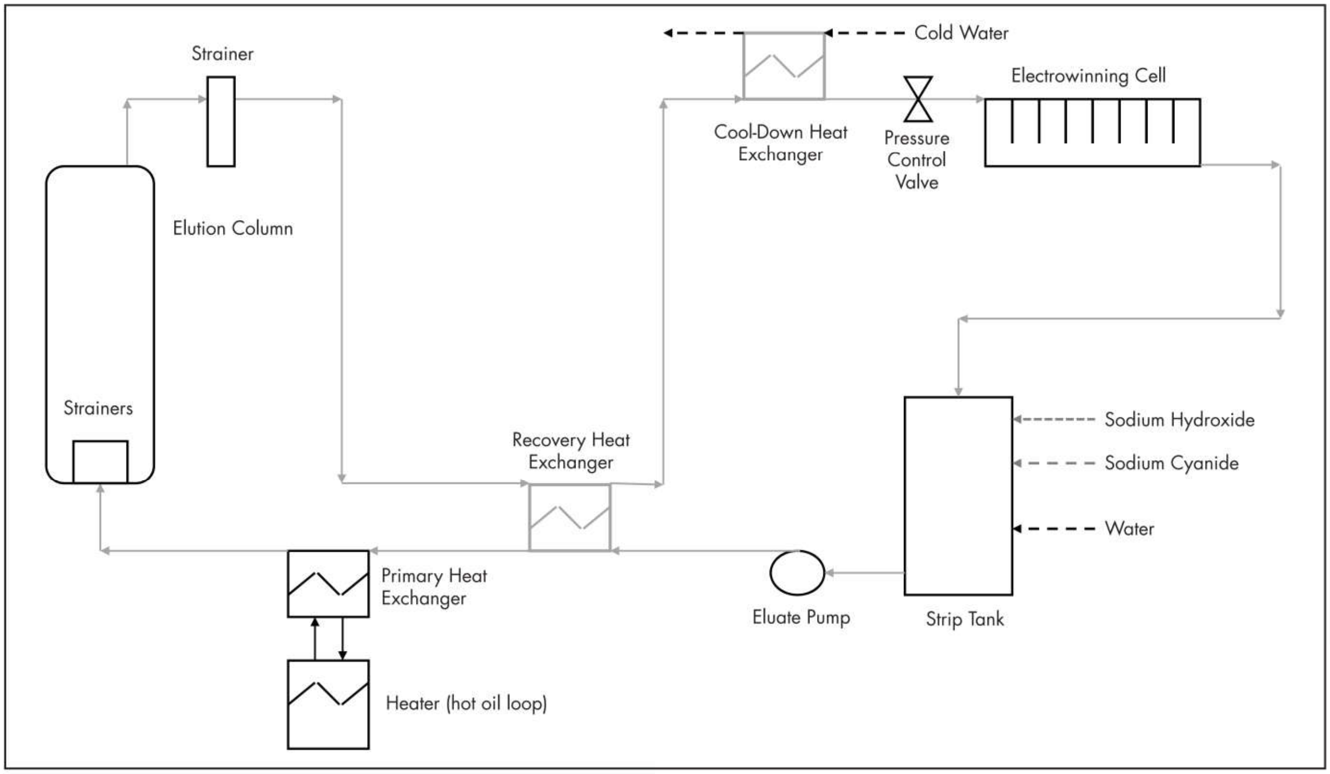

Figure 5 Pressure Zadra flow sheet

- Capital per unit mass of carbon processed per unit time is typically lower.
- Smaller batches of carbon reduce carbon inventory.
- CIL/CIP inventory management is simplified.

Consequently, most modern Zadra facilities are of the pressure variety. Figure 5 presents a typical configuration.

Given that electrowinning is conducted at atmospheric pressure, there is a need to reduce the temperature of the pressure Zadra eluate as it leaves the top of the elution column prior to electrowinning. This requires the use of additional heat exchangers and/or a flash vessel to cool the eluate.

The barren eluate from the electrowinning cell is circulated via the eluate tank to a heat exchanger and is preheated by the pregnant eluate discharging the column on its way to electrowinning. The barren eluate then passes through the heating equipment (electric, gas, or diesel-fired heater/boiler) where it is heated to the final elution temperature. Eluate is then directed to the base of the elution column and flows upward through the carbon. The pregnant eluate discharges through the column top and the cycle is repeated.

To achieve a rapid cooldown of a pressure Zadra, it is common to add a cooldown heat exchanger employing cold raw water. Once the circuit is cooled below the flash point, the carbon can be transferred to the reactivation kiln.

Anglo American Research Laboratories Elution

The AARL elution process was developed in the 1970s and gained acceptance in the 1980s. The process is somewhat different from the Zadra-based processes in that the elution and electrowinning steps are isolated from each other. The AARL process uses low-ionic-strength (deionized) water to undertake elution post heat-up and pretreatment with reagents.

An example of an AARL flow sheet is presented in Figure 6. The AARL process has three basic steps. The first is the presoak step, which involves circulating a sodium hydroxide and cyanide solution in a closed loop through the column and the solution heating circuit to achieve temperature. Typically 1–2% w/v of sodium hydroxide and similar concentrations of sodium cyanide are used in the presoak. It is common for operations to reduce the sodium cyanide concentrations to 0.5% or even lower post-commissioning. This presoak stage is very similar to the initial heat-up stage of a Zadra-based process. The flow rate of presoak solution and eluate is on the order of 2–2.5 BV/h.

Once the elution column discharge is ~95°C, the flow path of the solution is diverted and the second stage of actual elution commences. Low-ionic-strength water (eluate) is pumped through a recovery heat exchanger, which scavenges heat from the presoak solution exiting the column. The cooled presoak solution is directed to the eluate tank via a pressure control valve.

The preheated water (eluate) now passes through the primary heat exchanger (heater) where it is heated to the elution temperature, typically in the 110° to 140°C range. The circuit is pressurized to stop the solution from flashing off. The water (eluate) is pumped upflow through the column, desorbing the precious metals from the carbon. Pregnant eluate exits the top of the column and passes through the recovery heat exchanger, giving up heat to the cold, new water entering the circuit, before being directed to the pregnant eluate tank. The efficiency of the circuit is such that typically only 8 BV of

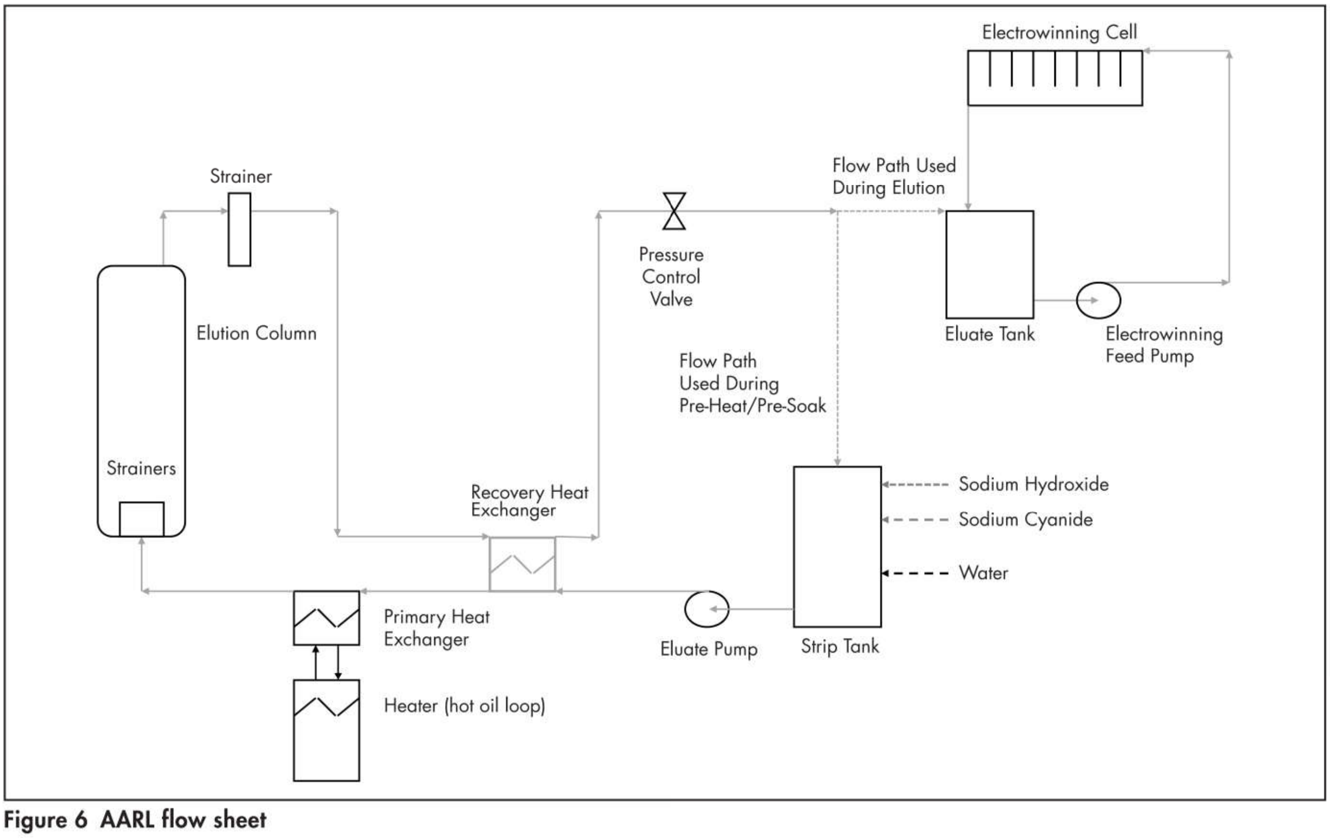

Figure 6 AARL flow sheet

eluate are required to complete an elution, taking only 4 or so hours post-presoak/preheat.

The last stage of the elution is the cooldown stage. Once cooled, the carbon can be transferred to the reactivation kiln. An AARL circuit can process up to three batches of carbon per day at 100% utilization. However, circuits are generally designed to process between 1 and 1.5 batches per day and sized accordingly. In practice, a maximum of 2.5–2.8 elution cycles per day are achievable.

The eluate is stored in the eluate tank from which it is pumped to the electrowinning circuit. The electrowinning tail recirculates back to the eluate tank, and the solution is recirculated until the required barren eluate grade is achieved. As the electrowinning cycle is independent of elution, it is possible to utilize multiple electrowinning tanks to allow multiple elution cycles to be run. One eluate tank can be filled or emptied while another tank is being used for the actual electrowin cycle itself.

The efficiency of the process is very dependent on the quality of the elution water used. If water having elevated levels of sodium, calcium, or magnesium are applied, the elution rate is significantly impacted, as is the final barren carbon grade. A reverse osmosis or nanofiltration plant is sometimes required to provide the necessary volume of quality water.

This is where Zadra-based processes have an advantage over AARL: The Zadra process recirculates around 1–2 BV of eluate, whereas the AARL process requires 8–10 BV of good-quality water for elution. This water demand can add to CAPEX and OPEX associated with the AARL option.

Split AARL

A variation of the AARL process is the split AARL process, which

- Conserves water (saves 3–4 BV),
- Generates higher-grade eluate,
- Makes for more efficient electrowinning, and
- Reduces recycle of values in barren eluate return to the adsorption circuit.

The circuit effectively operates the same as a conventional AARL except that the last few BVs of low-grade eluate are directed to a "starter" or "lean" eluate tank instead of the eluate tank.

The presoak step and preheat of the next elution cycle are conducted with new water. Once the elution steps from presoak/preheat to elution proper are completed, the low-grade "starter" eluate is then used. Once the starter eluate is expended, new water is consumed to complete the elution. The increased complexity and cost of the split AARL is offset by savings in water treatment and a minor offset in electrowinning capital.

Advantages and Disadvantages

When considering the two most common elution options, pressure Zadra or AARL variants, the following advantages and disadvantages should be considered:

- **Water quantity and quality.** The AARL process requires a larger volume of quality water.
- **Upgrade/expansion.** An AARL facility is typically easier to expand.
- **Flexibility.** Projects that have wide variation in feed grades or throughput are better serviced with the more-flexible AARL option.
- **Water balance.** The AARL option can influence the process water balance.
- **Capital cost.** An AARL circuit has a higher capital cost *for the same batch capacity* than a Zadra equivalent. If the AARL is sized for multiple elutions per day, the effective CAPEX can be less than a Zadra circuit.
- **Performance.** An AARL circuit will typically provide lower barren carbon grades than a Zadra circuit.
- **Reactivation.** The AARL barren carbon contains less-soluble salts. This generally results in elevated carbon activity prior to and after reactivation.
- **Operating cost.** The operating costs are so similar that when all facets are considered, there is often no clear advantage between AARL and Zadra. Often the AARL is suggested to have a higher OPEX, but AARL advantages can be shown to reduce operating costs in other areas.

ELECTROWINNING

It is assumed in this section that eluates are derived from the most common Zadra or AARL elution variants and not alcohol or other elution processes. Electrowinning is usually conducted in square or slightly rectangular cross-sectional cells. The anodes and cathodes are placed at right angles to the eluate flow. The eluate is introduced at one end of the cell and passes through the anodes and cathodes, exiting the cell partially or near fully depleted of precious metals. A direct current potential of around 3–5 V is applied between the anodes and cathodes. Current demands vary depending on the size of the cells and may range from a few hundred amperes up to 6,000 amperes or beyond.

Circular (radial flow), tubular (axial flow), and other cell variants have been tested over the years with varying success. The rectangular cell remains the most common, mainly due to the ease of cleaning, the layout affording close packing of cathodes and anodes, and simple control of the eluate flow path.

Anodes are generally stainless-steel mesh or punched plate. Cathodes are either mild steel wool or knitted stainless-steel wire. The mild-steel-wool options provide a high surface area. This affords a low current density and a competent precious metal deposit with good adherence to the wool. Stainless wire cathodes present a higher current density due to the reduced area. This results in a poorly adhering precious metal deposit that can be relatively easily removed from the cathode by brushing or hosing.

Mild steel cathodes are subjected to oxidation or acid treatment prior to smelting. Direct smelting is also practiced. As the mild steel wool is consumed, it must be replaced after each cycle. Stainless-steel cathodes are washed to dislodge the precious metal. The precious metal sludge is collected, filtered, and dried, and then the sludge is smelted to produce doré. The washed cathodes are returned to the electrowinning cell for reuse.

Design Considerations

Elution Process

The type of elution being undertaken needs to be considered during the electrowinning cell design process. A Zadra variant requires the barren electrowinning eluate to be as low a grade as reasonably practical to increase the efficiency of the elution itself. Therefore, the electrowinning cell should have a pass efficiency of 90% or higher.

AARL processes are not as dependent on the pass efficiency, as the eluate is just recycled back to the eluate tank. As long as the electrowinning circuit is designed to recirculate an adequate flow and electrowin the precious metal in the allotted time, the number of cathodes is not generally a constraint.

Cell Flow Rate

The cell flow rate should be in the range of 250 to 500 L/m^2 of cross-sectional area per minute (Filmer 1982). Too high a flow rate can result in abrasion of the precious metal precipitate such that it dislodges and forms a slime in the bottom of the cell, settles in the eluate tank itself, or is recirculated to the elution.

Current Demands

The electrowinning process results in oxygen evolution at the anode according to the following equation:

$$4OH^- \rightarrow O_2 + 2H_2O + 4e^-$$

At the cathode, evolution of hydrogen occurs according to this equation:

$$2e^- + 2H_2O \rightarrow H_2 + 2OH^-$$

Gold (or other metal) deposition occurs at the cathode according to the following equation:

$$e^- + Au(CN)^-_2 \rightarrow Au + 2CN^-$$

There is an imbalance in the amount of hydroxide consumed and generated, such that supplementary sodium hydroxide is dosed part way thought the electrowinning process or elution to maintain alkalinity.

Other metal precipitation at the cathode, including silver, copper, and even mercury, results in allowances being made in the current supply to the cell. Many designers assume a fixed current efficiency for gold and silver when calculating the rectifier capacity. Values of 10% for gold and 15% for silver are common values applied independent of eluate grade. However, the current efficiency relevant to the metals is a function of the solution grade and is not a constant. Figure 7 presents field data plotted on a linear axis for a rectangular electrowinning cell fitted with stainless-steel knitted cathodes. The constituents of the eluate, relative concentrations, and influence they may have on current efficiency should be considered prior to deciding how current efficiency is to be applied to sizing a rectifier.

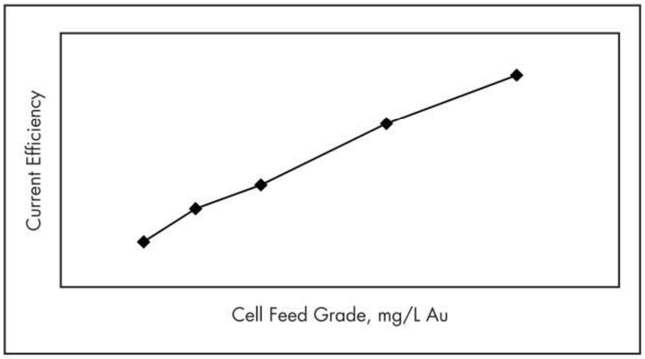

Figure 7 Current efficiency versus electrowinning cell feed grade

Other Design Features

Electrowinning cells should have anodes and cathodes spaced as tightly as possible without risk of short-circuiting of fluid flow or electrical current. Cell bodies need to be able to be easily cleaned to remove metal sludge and mineral fines, and cells should be located at a height to allow free draining. In addition, mercury will elute from the carbon and present to the electrowinning cell. Capture and control of emissions of mercury is required if present (retort, sulfur-impregnated carbon adsorption, wet scrubbing).

Eluate tanks necessitate a means of easy cleaning to remove precious metal and mineral sludge with consideration of a potential toxic atmosphere. Eluate tanks also require consideration of mercury presence.

REACTIVATION AND REGENERATION OF CARBON

Reactivation and regeneration are two different processes that are often referred to erroneously as one and the same (Claflin et al. 2015). *Regeneration* is the displacement of adsorbed species present on the carbon that does provide an increase in carbon activity. Regeneration may not provide the necessary return to activity of the fouled carbon, particularly if pyrolysis products remain. *Reactivation* is considered the process where the conditions are such that the water–gas reaction is initiated, typically at temperatures greater than 700°C (up to 750°C) in the presence of H_2O (steam):

$$C + H_2O \rightarrow H_2 + CO$$

The conditions need to be retained for a period of time until the pyrolysis products are removed. Additional time is potentially disadvantageous should the carbon hardness be impacted or the micropores be allowed to grow to mesopores (Claflin et al. 2015). It is therefore important that the carbon post-elution is subjected to the correct conditions to achieve reactivation and not just regeneration. The designer needs to specify the reactivation kiln such that the *temperature of the carbon is achieved and maintained for the appropriate time at this temperature.* Too often, it is the nominal kiln residence time and temperature of the kiln shell that are accepted as proxies for the required carbon conditions. That is, the conditions needed to be experienced by the carbon in the kiln are not the same as the kiln operating conditions.

Reactivation Circuit Design

The type of feed to the CIL/CIP, water quality, and other factors such as potential oil contamination from the mine will determine the need and frequency of reactivation. Circuit feed emanating from flotation processes will contain elevated levels of reagents that can easily foul the carbon. Water high in calcium or other species can similarly foul the carbon chemically or physically (carbonate deposition). Whereas acid washing may remove such carbonate buildup, other contaminants such as humic acids and tannins will remain.

Unless the ore is particularly clean, the water quality is excellent or the demand for reduced capital is critical, designing for less than 100% reactivation of the elution capacity is to be avoided. On exiting the reactivation kiln, the carbon needs to be cooled in an inert atmosphere or, more often, it is quenched in water. It is important that the quench water is of good quality. Poor-quality water can itself result in reduced activity.

Carbon Sizing

The reactivated carbon has to be screened to remove any fines prior to introduction back into the CIL/CIP. Fines introduced to the CIL/CIP can partially load up with precious metals and then discharge to tails, resulting in losses.

Quenched carbon is usually pumped as a slurry from the quench tank to the carbon sizing screen located on top of the CIL/CIP. The screen oversize drops into the last CIL/CIP tank. Wet carbon tends to ball up once dewatered on a vibrating screen. Clusters of wet carbon migrate along the screen and the resulting sizing efficiency is low. The use of water sprays and, in some instances, wash troughs provide some benefit.

The most effective sizing method is to screen the dry carbon post-reactivation. Minor fine ash and carbon will adhere to the oversize, but after the carbon is slurried and pumped to the CIL/CIP, a simple dewatering screen will remove most of these misreporting fines.

DESIGN OF A CARBON CIRCUIT—CASE STUDY

The following sections describe *one method* for selecting the basis of a basic carbon circuit design. Where appropriate, comment is made as to alternative methods that can be applied. The design assumes free milling ore with low silver grade, and simple metallurgy with no complicating issues such as poor water quality, high viscosity, or high cyanide-soluble copper levels.

Data Required

The design of a carbon circuit requires a selection of process design criteria, much of which is the result of metallurgical test work undertaken on representative ore samples. Process design criteria required include:

- A proposed production schedule reporting head grade and ore types to be processed as a function of time.
- Viscosity data at various pulp densities and shear rates and at the appropriate pH using the appropriate pH modifier for the various ore types.
- Leach data at the appropriate conditions of pulp density, pH, oxygenation/aeration, and reagent dose.
- Carbon characterization data. The methodology used to design the circuit will dictate the data required. Typically, carbon kinetic data are required as a minimum (Fleming k and n values). Kinetic data are required for all metals of interest.
- Carbon equilibrium data should also be determined.

Carbon characterizations must be determined at the actual size distribution of the carbon to be applied to the design and after the carbon has been attritioned and equilibrated with a sodium cyanide solution. Use of certain carbon adsorption procedures that specify pulverizing the carbon are not applicable to CIL/CIP (or heap leach) design. It is preferable to use a slurry and not a solution to determine the carbon characteristics, as a slurry will typically provide lower values with regard to both kinetics and equilibrium loadings.

Example Design Case Basis

The design cases evaluated are based on the following:

- Seven cases evaluating a throughput of 4 Mt/yr (million metric tons per year) over 8,000 operating hours per year on an ore having a head grade of 3 g/t Au, insignificant silver, and a moderate gold leach rate.
- Seven additional cases evaluating a throughput of 4 Mt/yr over 8,000 operating hours per year on an ore having a head grade of 4 g/t Au, insignificant silver, and a slower gold leach rate.
- Ore specific gravity of 2.7.
- CIL/CIP feed at 42% solids w/w.
- Split AARL elution.
- Return eluate having a grade of 3 mg/L Au.
- Process water having a grade of 0.01 mg/L Au.
- 24-hour residence time.
- No leaching has occurred prior to the tank farm.
- Kinetic leach data available and leach model fitted.
- Equilibrium Freundlich isotherm data available.
- Decay parameters applied to carbon kinetic parameters based on benchmarked data.
- Tails solution grade of 0.010 mg/L targeted.

Model Structure

Several methods are applied to model CIL and CIP performance. The most basic models consider only the leaching and carbon adsorption processes. Constant carbon kinetics are considered and a final loaded carbon grade is assumed based on benchmarking or the use of "factors" or generic equations. Other models, including the model applied herein, consider carbon equilibria, carbon kinetic decay, and other influences such as back-mixing from carbon-transfer operations.

A leach equation is fitted to the test-work data to provide an estimate of the gold leaching from the solids phase and into solution as a function of time.

Carbon performance decays as the carbon progresses up the tank train. The loadings in the plant will not be as high as determined in the laboratory using fresh carbon. The age of the carbon and the other species that may be present (silver and copper as well as water quality) influence carbon performance. The Fleming k and n values are discounted from the test work using factors based on operating experience and decayed as a function of carbon residence time.

While the model used herein considers a constant rate of decay in carbon characteristics, it is often found that the first carbon contactor is the one in which the carbon suffers the most due to the exposure of contaminants in the new circuit feed.

The carbon loading in each contactor is compared to the loadings predicted by the application of a Freundlich isotherm for the solution concentration calculated for each contactor. In the example herein, it has been assumed that the carbon loading should not exceed 70% of the equilibrium loading determined by test work. The use of such a discount is based on plant observations and experience.

Such adjustments provide some allowance for real-world plant operational issues compared to the laboratory environment, such as uneven carbon populations from one tank to another.

The CIL/CIP model applied herein considers the influence of back-mixing due to carbon transfer. In those cases where carbon populations are low, the amount of slurry that has to be back-mixed (transferred) to the preceding tank will be higher than those cases where populations are higher. This back-mixing influences the residence time distribution of the ore and will flatten the concentration profile of the gold in solution and therefore the corresponding carbon loadings.

Lower final loaded carbon grades require higher elution rates to recover a given mass of precious metal and therefore require elevated carbon transfer rates. Therefore, higher final carbon loadings will provide a benefit with regard to elution, electrowinning, and the reactivation plant by reducing the required capacity.

The model also considers the influence of precious metal in the process water entering the facility due to dilution (water sprays) as well as in the actual feed stream from the comminution circuit itself. Recycled eluate is also considered. The model assumes a split AARL circuit with a return barren eluate grade of 3 g/m^3. The amount of recycle is a function of the elution type and the elution rate required, and consequently a function of the loaded carbon grade achieved for each case.

Design Exercise for 3-g/t Head Grade

Cases Evaluated

The model was applied to assess seven plant arrangements, or cases. All cases have a total residence time of 24 hours. The cases are summarized as follows:

- **Case 1.** Pure CIL comprising six tanks. All tanks have carbon present.
- **Case 2.** Hybrid CIL comprising six tanks. The first tank does not contain carbon but all subsequent tanks do. Tank sizes and configuration are the same as for case 1.
- **Case 3.** Pure CIL comprising seven tanks. All tanks have carbon present. The tanks are smaller than those modeled for cases 1 and 2.
- **Case 4.** Hybrid CIL comprising seven tanks. The first tank does not contain carbon but all subsequent tanks do.
- **Case 5.** Pure CIL comprising eight tanks. All tanks have carbon present. The tanks are smaller than those modeled for cases 3 and 4.
- **Case 6.** Hybrid CIL comprising eight tanks. The first tank does not contain carbon but all subsequent tanks do.
- **Case 7.** CIP comprising four leach tanks (residence time of 18 hours) and six adsorption tanks (residence time of 1 hour each for a total of 6 hours), having a total residence time of 24 hours. This circuit is not a "pure" CIP, because if it were, the leach residence would be 24 hours alone. This configuration modeled can be considered a "practical" CIP design, where the minor leaching in the latter stages of the processing are accepted to reduce plant capital cost.

All cases assume the same carbon population in each tank as relevant to the case evaluated and constant carbon transfer. Design is based on 100% utilization, as is necessary to correctly size a facility.

Performance

Three plots of the solids grade, solution grade, and the carbon grade by tank are presented as Figure 8, Figure 9, and Figure 10 for select cases.

Figure 8 shows typical profiles for the various phases of solids, solution, and carbon. A carbon loading of just under 2,400 g/t has been achieved. When compared to Figure 9, one leach and five contactors, it can be seen that case 2 has achieved slightly higher carbon loading of just under 2,500 g/t. The contact solution grade (the solution grade equilibrating with the first carbon contact) for case 1 is just under 0.48 g/m^3, whereas for case 2 it is 0.54 g/m^3, allowing for a higher carbon loading.

As a result, case 2 has a slightly lower elution demand, which is advantageous in both elution capacity required and the amount of back-mixing that will be experienced. Consequently, case 2 may be considered preferable over case 1. However, the practical consideration is that the carbon profile is very steep for case 2 compared to case 1. If carbon populations are not properly managed, or if a tank were to be taken off-line for maintenance as will happen in normal operations, the impact this would have on the case-2 arrangement will be much more significant than it would be to case 1. Therefore, the consideration as to the most appropriate arrangement needs to consider the practical aspects and not just the theoretical and steady-state model output.

Figure 10 presents the CIP case. As there is no carbon present in the first four tanks, the solution grade can be seen to climb in excess of 1.9 g/m^3. The resulting solution grade in contact with the carbon in the first contactor is around 0.8 g/m^3 compared to the values of around 0.5 g/m^3 seen in case 1 and case 2. This results in higher carbon loadings, reduced elution demands, and reduced back-mixing associated with carbon transfers.

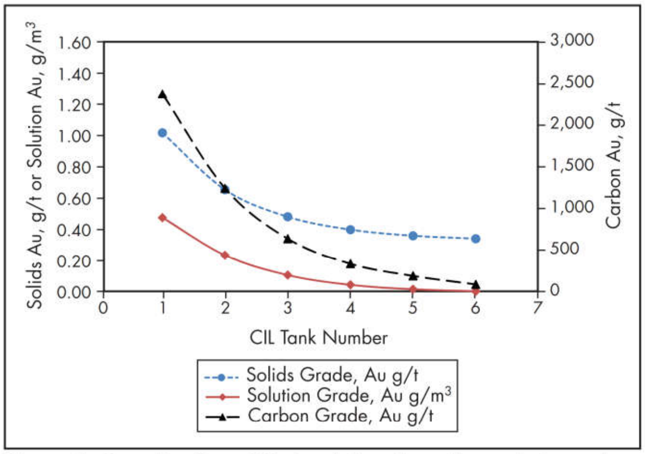

Figure 8 Case 1—Pure CIL, 3-g/t head grade, and six tanks

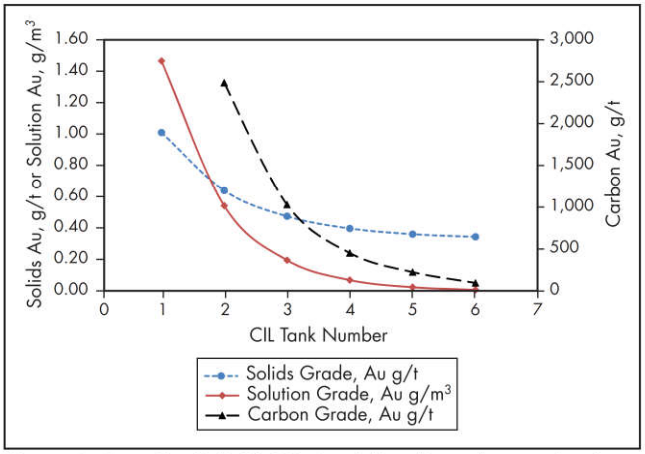

Figure 9 Case 2—Hybrid CIL, 3-g/t head grade, one leach tank, and five carbon contactors

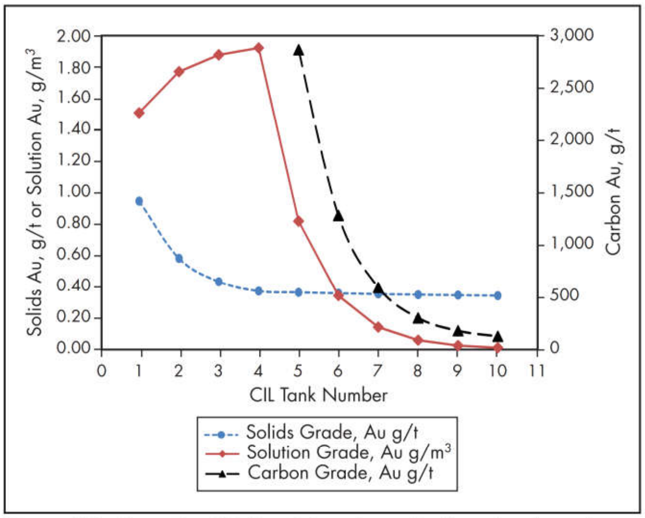

Figure 10 Case 7—CIP, 3-g/t head grade, four leach tanks, and six carbon contactors

The carbon profile is steep, and the loss of a carbon contactor due to maintenance will impact the profile as a consequence. However, in a CIP operation, it is easier to recover from such upsets by adding the small amount of displaced carbon to the other tanks or, in fact, adding new carbon, because the carbon inventory is so much lower than that of an equivalent CIL that the inventory is more easily managed.

Table 1 presents several inputs and outputs that allow the performance of the various cases to be evaluated. The tonnage rate, head grade, residence time, and pulp density have all been held constant in each of the cases modeled, as has the barren carbon grade, which is equivalent to the introduced carbon grade. Notably, the lower the introduced carbon grade, the lower the elution rate required for a fixed flow of precious metal, given that the loaded carbon grade is "fixed" by the contact solution grade. A more effective elution that provides lower barren carbons also means a smaller elution demand, and this is an important consideration when evaluating the type of elution circuit to be applied.

The model has been run to achieve the same soluble loss of 0.010 g/m^3 for each case. Given that the tailings pulp density is the same for each case, then the soluble loss for each case is also the same. In effect, the carbon circuit options are comparable based on the solution losses.

Referring to Table 1:

- The contact solution grade increases as the number of tanks is increased. Having a leach tank present escalates the contact solution grade and provides for an elevated carbon loading.
- The elution rate diminishes as the number of tanks increases/contact solution increases/loaded carbon grade increases.
- The carbon population decreases for the CIL cases as the number of tanks increases, as does the carbon residence time. However, the back-mixing due to carbon transfers increases as the number of CIL tanks increases.
- The back mixing results in a loss of extraction, which increases with the number of CIL tanks.
- A single leach tank in the CIL cases reduces the back-mixing and improves the extraction.
- Both the gold and carbon inventory decrease as the number of CIL tanks increases. The use of a single leach tank in the CIL cases escalates the gold inventory for comparable cases 1 and 2, but decreases the gold inventory for the comparable case 3 and 4 as well as comparable cases 5 and 6.

Table 1 Case performance summary for 3-g/t head grade

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7
	CIL	Hybrid	CIL	Hybrid	CIL	Hybrid	CIP
Parameter	6 Adsorption	1 Leach 5 Adsorption	7 Adsorption	1 Leach 6 Adsorption	8 Adsorption	1 Leach 7 Adsorption	4 Leach 6 Adsorption
Tonnage rate, t/h	500	500	500	500	500	500	500
Head grade, Au g/t	3.00	3.00	3.00	3.00	3.00	3.00	3.00
Residence time, h	24	24	24	24	24	24	24
Tails pulp density, % solids w/w	41.8	41.8	41.8	41.8	41.8	41.8	41.8
Contact solution grade, mg/L	0.48	0.54	0.51	0.60	0.54	0.64	0.82
Solution loss (liquor grade) mg/L	0.0100	0.0100	0.0100	0.0100	0.0100	0.0100	0.0100
Carbon final loading, Au g/t	2,382	2,484	2,434	2,570	2,474	2,635	2,863
Barren carbon loading, Au g/t	100	100	100	100	100	100	100
Elution rate, kg/h	590	559	577	538	568	523	480
Elution rate, t/d	14.2	13.4	13.8	12.9	13.6	12.6	11.5
Carbon population, g/L slurry	9.0	10.7	7.5	8.4	6.5	7.0	16.5
Carbon residence time, days	13	14	11	12	10	10	8
Back-mixing (carbon transfer), m^3/h	65.8	52.1	77.2	64.2	86.8	74.6	29.1
Extraction from solids, no back-mix, %	88.58	88.58	88.58	88.58	88.58	88.58	88.58
Extraction with back-mixing, %	88.44	88.49	88.41	88.47	88.38	88.44	88.57
Loss due to back-mixing, %	0.14	0.09	0.17	0.11	0.20	0.14	0.01
Gold inventory, oz	5,040	5,260	4,093	4,086	3,509	3,403	2,495
Carbon inventory, t	189	189	158	152	138	130	87

Table 2 Financial considerations for 3-g/t head grade

Parameter	Case 1 CIL 6 Adsorption	Case 2 Hybrid 1 Leach 5 Adsorption	Case 3 CIL 7 Adsorption	Case 4 Hybrid 1 Leach 6 Adsorption	Case 5 CIL 8 Adsorption	Case 6 Hybrid 1 Leach 7 Adsorption	Case 7 CIP 4 Leach 6 Adsorption
Extraction loss (8,000 h/yr), oz/yr	552	339	665	444	766	541	31
Loss in revenue, US thousand $/yr	662	407	798	532	920	649	37
Gold inventory, oz	5,040	5,260	4,093	4,086	3,509	3,403	2,495
Value of gold inventory, US million $	6.05	6.31	4.91	4.90	4.21	4.08	2.99
Incremental value over CIP case 7, US million $	3.05	3.32	1.92	1.91	1.22	1.09	0
Carbon inventory (including elution and reactivation), t	232	229	199	190	179	167	122
Value of carbon inventory, US thousand $	139	137	120	114	107	100	73

- The CIP case presents with reduced elution rates, lowest gold and carbon inventory, and, importantly, the lowest back-mixing and highest extraction.

The capital cost of the tank farm will grow as the number of tanks increases for a given residence time. The hybrid CIL options (cases 2, 4, and 6) would be a marginally lower cost than the corresponding pure CIL options (cases 1, 3, and 5) because there is minor savings in inter-tank screen costs and transfer pumping.

The CIP case would have the higher capital associated with the tank farm. There will be savings in some areas compared to the CIL options (screen access and transfer pumping), but the increased tankage costs will generally negate these benefits.

The capital cost of the elution, electrowinning, and reactivation kiln facilities will cost less as the number of tanks increases. This provides some capital cost offset when considering CIL/CIP tank configurations for a new project.

Although the pure capital cost is one of the most important considerations when evaluating plant design options, other factors will significantly influence the cash flow of the project, particularly in the all-important start-up and finance payback period. These factors can be considered as working capital and can be expected to influence the terms of any finance agreements thay may be required to fund a project; the establishment of the inventory will delay payments to the financiers but will require operating costs to be sunk.

Table 2 summarizes some of these factors and quantifies them based on a gold price of US$1,200/oz and a carbon cost of US$600/t. The "extraction loss," based on an 8,000-hour operating year, is the difference in the theoretical leach extraction without back-mixing compared to what is actually achieved when back-mixing is considered. The CIP case has a minor loss in revenue because of this inefficiency, whereas the CIL options, particularly the higher number of tanks options, suggest elevated losses.

The value of the gold inventory is a significant cost. As the tank numbers increase, the inventory is reduced. This is a "one-off" cost to the project, because once the inventory is established, it remains until the end of the mine life. However, this can be considered as cash flow that is sunk in the first month or two of the project life. The CIP presents the best case with regard to inventory lockup, as it presents the lowest value.

The lower-capital-cost CIL cases (cases 1 and 2) in reality have an additional inventory-based cash-flow burden compared to the more elaborate CIL cases or the CIP case. So although cases 1 and 2 may have the lowest capital and the least loss due to back-mixing of the CIL cases, cases 1 and 2 also have the highest inventory to contend with. It is likely that the cost of inventory for cases 1 and 2 would exceed the incremental capital associated with cases 3 and 4.

The hybrid CIL described in case 2 may still appear appealing as the lowest entry capital cost option compared to the most expensive CIP option, case 7. Case 2 will have a loss of nominally US$400,000/yr in extraction and require an additional US$3.3 million to be sunk as gold inventory compared to CIP case 7. This is nominally US$4.5 million over a three-year non-discounted payback period, which is a significant disadvantage to the CIL case. In addition, the CIP will be more robust to operate and will require lower cost elution, electrowinning, and reactivation facilities with associated reduced operating costs.

It is therefore necessary to consider not just the basic metallurgical demands and the entry capital cost when designing a CIL or CIP facility, but also the operability, inventory, and losses due to the operating mechanisms of the design cases to reach the most appropriate financial case.

A further consideration in circuit selection is the variability of the ore blends that the plant will experience. Particularly in the payback period of the operation, it is common for operations to experience soft oxides at start-up, potentially of lower grade than deeper sulfide ores. Then as the open pit develops, the sulfides present and require different operating conditions or a different CIL/CIP design. They may require more comminution but may also have different leaching, adsorption, and carbon management demands. It is therefore paramount for the designer to fully understand the production schedule implications with regard to what the design must contend with as a function of time.

The influence of carbon inventory itself (i.e., cost of carbon) are insignificant compared to the capital cost and gold inventory considerations for cases 1 to 7.

Design Exercise for 4-g/t Head Grade

To present the influence that variable parameters can have on the CIL/CIP design, a second series of cases was run. The same plant configurations were used as had been designed for cases 1–7 presented earlier. However, the head grade of the ore was increased to 4 g/t and the leach rate moderated, as could be expected if the various plant configurations were to experience a change in ore type.

The cases here are designated with an "H" to infer "high-grade." The 3-g/t head grade case 1 is directly comparable with the 4-g/t case 1H.

In summary:

- **Case 1H.** Pure CIL comprising six tanks.
- **Case 2H.** Hybrid CIL comprising six tanks.
- **Case 3H.** Pure CIL comprising seven tanks.
- **Case 4H.** Hybrid CIL comprising seven tanks.
- **Case 5H.** Pure CIL comprising eight tanks.
- **Case 6H.** Hybrid CIL comprising eight tanks.
- **Case 7H.** CIP comprising four leach tanks (residence time of 18 hours) and six adsorption tanks (residence time of 1 hour each for a total of 6 hours), having a total residence time of 24 hours.

Table 3 presents the inputs and outputs so the cases can be compared. The same general trends as were found for the 3-g/t modeling are presented here. Some notable differences resulted from slower leaching and higher quantity of precious metal to be processed, the most significant of which are listed here:

- The carbon residence times for the CIL options have all increased from nominally 10–13 days (cases 1 to 6) up to 20–28 days for cases 1H to 6H. These carbon residence times are considered too high and, as such, these CIL options would need to have lower carbon loadings applied and much higher elution rates enforced in reality.
- The slower leach rate and higher grade is pushing more gold to the tail end of the circuit. This is advantageous to the CIP case. The contact solution grades for the CIL cases are lower here than what they were for the 3-g/t cases even though the head grade is higher, resulting in lower carbon loadings but still having to process around 30% more precious metal.
- The CIL carbon populations have increased significantly by around a factor of 3, whereas the CIP population has increased by a factor of 30%, in line with the incremental precious metal flow.
- Gold inventory for the CIL options has increased dramatically.

Cases 1H to 6H show that if a circuit had been selected based on the case 1–6 models, then the circuit would have been compromised by the higher-grade but slower-leaching ore. The likely practical outcome would have been to maximize the elution rate, maximize carbon population to the point of maximum carbon residence time, and suffer the soluble losses that would occur if the ore change had not been allowed in the design. This is where an AARL elution circuit would prove advantageous over a Zadra facility, as increased elution rates would be more easily achieved.

Table 4 summarizes the same values as had been presented in Table 2. The loss in revenue per year due to back-mixing has jumped for all seven cases, but is most significant for the CIL cases. The gold inventory has increased slightly for CIP case 7H, but for the CIL cases, the inventory has increased by a factor of 3. There is now a very high proportion of working capital to be considered if the lower-capital-cost CIL options are to be considered.

The CIP (case 7H) in this example has handled the change in characteristics of the ore with little impact. The higher-grade and slower-leaching ore is easily managed. However, the CIL options, although probably competitive for the 3-g/t scenarios, would in reality not be applicable to the 4-g/t cases without significant modification.

If this 4-g/t ore were to be harder than the 3-g/t ore, and it was to be processed at a lower throughput, then the CIL options would be expected to perform more effectively.

Table 3 Case performance summary for 4-g/t head grade

	Case 1H	Case 2H	Case 3H	Case 4H	Case 5H	Case 6H	Case 7H
	CIL	Hybrid	CIL	Hybrid	CIL	Hybrid	CIP
Parameter	6 Adsorption	1 Leach 5 Adsorption	7 Adsorption	1 Leach 6 Adsorption	8 Adsorption	1 Leach 7 Adsorption	4 Leach 6 Adsorption
Tonnage rate, t/h	500	500	500	500	500	500	500
Head grade, Au g/t	4.00	4.00	4.00	4.00	4.00	4.00	4.00
Residence time, h	24	24	24	24	24	24	24
Tails pulp density, % solids w/w	41.7	41.7	41.7	41.7	41.7	41.7	41.8
Contact solution grade, mg/L	0.35	0.47	0.35	0.49	0.35	0.50	0.94
Solution loss (liquor grade), mg/L	0.0100	0.0100	0.0100	0.0100	0.0100	0.0100	0.0100
Carbon final loading, Au g/t	2,142	2,370	2,140	2,401	2,139	2,419	3,000
Barren carbon loading, Au g/t	100	100	100	100	100	100	100
Elution rate, kg/h	899	801	903	791	906	786	619
Elution rate, t/d	21.6	19.2	21.7	19.0	21.7	18.9	14.9
Carbon population, g/L slurry	28.1	30.5	23.6	25.2	20.6	21.9	20.7
Carbon residence time, days	28	28	23	24	20	21	7
Back-mixing (carbon transfer), m^3/h	32.0	26.3	38.2	31.4	44.1	36.0	29.9
Extraction from solids, no back-mix, %	89.70	89.70	89.70	89.70	89.70	89.70	89.70
Extraction with back-mixing, %	89.45	89.54	89.40	89.49	89.35	89.46	89.67
Loss due to back-mixing, %	0.25	0.16	0.30	0.21	0.35	0.24	0.03
Gold inventory, oz	15,630	14,885	12,590	12,018	10,602	10,271	3,110
Carbon inventory, t	593	537	499	456	435	404	110

Table 4 Financial considertations for 4-g/t head grade

	Case 1H	Case 2H	Case 3H	Case 4H	Case 5H	Case 6H	Case 7H
	CIL	Hybrid	CIL	Hybrid	CIL	Hybrid	CIP
		1 Leach		1 Leach		1 Leach	4 Leach
Parameter	6 Adsorption	5 Adsorption	7 Adsorption	6 Adsorption	8 Adsorption	7 Adsorption	6 Adsorption
Extraction loss (8,000 h/yr), oz/yr	1,272	838	1,549	1,058	1,791	1,249	168
Loss in revenue, US million $/yr	1.53	1.01	1.86	1.27	2.15	1.50	202
Gold inventory, oz	15,630	14,885	12,590	12,018	10,602	10,271	3,110
Value of gold inventory, US million $	18.76	17.86	15.11	14.42	12.72	12.33	3.73
Incremental value over CIP case 7H, US million $	15.02	14.13	11.38	10.69	8.99	8.59	—
Carbon inventory (incuding elution and reactivation), t	658	594	564	513	500	460	154
Value of carbon inventory, US thousand $	395	357	339	308	300	276	92

If leach kinetics enhancement could be undertaken, for instance, by adding oxygen to the leach, then again the CIL performance would improve. More of the gold would leach earlier and this would result in much lower carbon inventory and higher loaded carbon grades. This point also highlights why increased leach kinetics may be beneficial even if the final extraction is the same—the adsorption circuit is more efficient, and both elution demands and back-mixing are reduced.

If the period of time that the 4-g/t ore were to be experienced was short or the 4-g/t material could be blended, then the CIL options again may be positive. These points all highlight the need for the designer to understand the ores to be processed and how the ore variability may influence the design.

Although the scenarios run for the 4-g/t head grade support CIP, it is often the case that for faster leaching and perhaps lower-grade ores there will be a bias back to CIL preference over CIP. If ores show any preg-robbing tendencies, then full CIL will be the preferred flow sheet and the designer will need to establish how a CIL facility can be designed/configured to deal with issues such as elevated grades and slow-leaching components. Use of different carbon populations in the different tanks and higher elution rates at lower carbon loadings are options that could be explored.

In summary, the designer should evaluate the plant configuration options with due consideration of the ore characteristics, how these characteristics vary, and the production schedule. Assuming a configuration and simply applying it to a life-of-mine composite or some other "general" ore can be expected to result in a suboptimal outcome. This is particularly true if the pure entry capital of the tank farm is used in isolation as a decision-making tool.

The designer also has to be aware of the financial environment. Cost of finance, finance type and terms, influence of working capital (inventory), and capital cost all playing a part in the flow-sheet decision process. Consideration as to operating cost influences (reduced stripping rate due to higher carbon loadings) also play a part in the financial process considerations. The process is not being designed to be the most metallurgically efficient or to have the lowest capital or operating cost, but to make the best or most robust return under the prevailing financial constraints.

Silver, Copper, and Other Species

The cases presented previously assumed no complications from other species in solution. In some instances, such as if there are high values of silver and/or copper in solution, there will be increased complexity in the modeling and design process, as these metals may adsorb and retard gold adsorption, or even desorb and free up sites for the gold under certain conditions. This then infers additional steps that would need to be considered in the modeling of the gold adsorption itself.

If low levels of silver are present and they have been included in the carbon characterization test work for gold, then the design process remains relatively simple. The circuit can be designed for the gold component, and the silver balance (loaded carbon grade of silver) can be back-calculated based on the carbon flow rate required to manage the gold. Only if the equilibrium loadings of silver are exceeded is there a need to increase the carbon flow rate over that required for the gold component alone such that the silver is managed. In such cases, the silver loadings will drive the elution circuit capacity and not the gold.

Higher silver-grade ores can become problematic. The carbon flow rates may become so great as to make the circuit impractical. Examples exist in industry where silver desorption is witnessed as the carbon is progressed or transferred upstream. Gold adsorption still occurs by utilizing the sites vacated by the silver, but the interim desorption of the silver results in the gold having an apparent reduction in kinetics. The silver is displaced downstream where it loads onto carbon again, and thus an internal recycle of silver presents as it is again moved forward with the carbon. There will be a high soluble loss of silver under such scenarios, as this behavior occurs when the carbon circuit cannot deal with the excessive quantities of silver present.

In such cases, consideration should be given to countercurrent-decantation/Merrill–Crowe-type flow sheets or hybrid flow sheets instead of carbon-only-based flow sheets. Alternatively, additional carbon characterization test work is required and a comprehensive understanding of the silver variability in the plant feed is necessary to retain carbon-based process routes.

Copper presents high adsorption kinetics but can also be controlled to a degree with free cyanide concentration. Elevated cyanide concentration results in the $Cu(CN)_4^{3-}$ complex being dominant over the $Cu(CN)_3^{2-}$ complex. The $Cu(CN)_4^{3-}$ complex has a low affinity for carbon and, if adsorbed, will give up sites for gold and silver adsorption. There is a cost associated with the cyanide demand, and if copper levels are too high, the copper will continue to dominate the carbon and adversely impact the gold and silver loadings. High cyanide-soluble copper ores require detailed evaluation before effective circuit modeling and confidence in design can be achieved. Cyanide consumption for both copper dissolution and carbon management often make such high copper ores uneconomic.

The site water itself can influence the design of a carbon-based circuit. The cations Ca^{2+}, Na^{+}, and Mg^{2+} are all known to influence carbon adsorption characteristics as will the CN^{-} radical itself. pH also impacts carbon adsorption characteristics. For these reasons, it is key that the designer has considered the influences with adequate evaluation at the laboratory stage of development to ensure that carbon performance is understood.

Model Alternatives

The CIL/CIP model described in this chapter is just one of many methods used to estimate circuit performance or develop a design. Simpler models are used and are applied, as are more complex models. The simplest models have their place but are generally limited to study-level evaluations or are applied to ores for which the confidence in predicted performance is high. Examples include a project expansion on the same ore type, a project restart where operating history from the past is available, or processing an ore from a region of reliable characterization, such as many of the oxide ores from the Western Australian Goldfields.

Models often do not consider the equilibrium loadings. The design needs to consider the kinetics to understand how much carbon is required and how quickly it needs to be progressed through the plant. The design also needs to ensure that the loadings do not exceed what is chemically possible with regard to loading—and this implies that equilibrium data and assessment are key in optimizing a design.

Simple models include the following:

- Carbon loading factored based on the head grade of the ore. This type of model does not allow for pulp density influences, tank configuration, or competing species. As presented according to the previous modeling, the grade of the solution in contact with the carbon in the first contactor influences the final loading; consequently, models that do not estimate this solution value result in escalated risk when estimating the achievable final carbon loading.
- Carbon loading factored on solution grade. The precious metal leached is estimated to provide a solution grade that is fed to the first carbon contactor. A factor is then applied to give the anticipated carbon loadings. The issue with this method is that it is the solution tenor in the first contactor in contact with the carbon that dictates the carbon loading, not the grade of the introduced solution. Carbon activity, population, and how much additional leaching occurs in the first contactor will influence the actual solution tenor in contact with the carbon.
- Carbon loading based on a "generic" algorithm that is based on a head-grade or solution-grade assumption. Some models use an equation based on a large number of historical observations to estimate carbon loadings, for both gold and silver. Such models assume the ore is therefore generic. If the ore being assessed is an outlier, then deviation from the generic algorithm can be expected. This may be positive or negative in outcome with regard to loading achieved, but even a positive/conservative outcome means an over-designed plant and is therefore suboptimal.
- Stage efficiency. Most circuits are observed to operate with a solution-stage efficiency of around 50%. That is, tank $n + 1$ will have a solution tenor of 50% of tank n. The various cases presented (cases 1 to 7H) had a range of stage efficiency of 40%–65%, suggesting that a fixed-stage efficiency is invalid. The leach rate, number of tanks, and configuration all give different efficiencies to achieve the same soluble loss of 0.01 g/m^3.

Such methods have their place, but the designer needs to appreciate that the method applied to design a CIL or CIP circuit may have great bearing on the confidence and reliability of the outcome. Reliable design requires reliable inputs. Representative test work remains the key to reliably estimating plant performance, particularly for the more complex and atypical ore types.

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