CHAPTER 10.8

Agitated Bioleach Reactors

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BRIEF HISTORY OF BIOPROCESSING USING AGITATED REACTORS

This chapter describes the design and operational aspects of agitated reactors using bacteria for minerals processing and focuses on the processing of refractory gold concentrates, as there have been few commercial applications using bioreactors for treatment of base metal concentrates. Several terms are commonly used to describe bio-hydrometallurgical processes, with *biooxidation* or *bacterial oxidation* and *bioleaching* or *bacterial leaching* being favored. Biooxidation or bacterial oxidation generally refers to the use of bacteria for solubilization of gangue elements, such as arsenopyrite or pyrite, to expose inert metals, such as gold, for subsequent recovery. In biooxidation processing, the bacteria are only responsible for removing mineralization that occludes the inert metals. The term *bioleaching* or *bacterial leaching* is generally used for bioprocesses in which the metals of interest are made soluble in the bioleach liquor (e.g., copper, nickel, or zinc from chalcopyrite, pentlandite, and sphalerite, and so on). For the discussions in this chapter concerning agitated bioreactors, these definitions are used as interchangeable terms unless specified otherwise.

The role of bacteria in the oxidation of minerals and metal extraction was identified many years prior to their commercial application in the agitated bioreactors currently used in minerals processing operations (Colmer and Hinkel 1947). The use of the Bacfox process at the Buffelsfontein uranium plant in South Africa during the 1970s was perhaps the first case of engineering an environment in which a bacterial process of a few days' residence time formed part of an overall commercial process (James 1976). The commercial development of an agitated reactor process using bacteria for the treatment of high-value arsenical gold concentrates was driven by the need to introduce an economic and environmental alternative to roasting. Biooxidation using reactors could satisfy both of these criteria as a suitable method of oxidation. Not only was the process economic, but also, from an environmental perspective, the arsenic made soluble by biooxidation could be neutralized with low-cost limestone and stabilized downstream as solid ferric arsenate. In 1986, the first biooxidation plant for refractory gold concentrate treatment was commissioned at Fairview in South Africa (Van Aswegen et al. 1989). It was followed by other plants throughout the world. Currently, more than 20 refractory gold projects worldwide use biooxidation with agitated, aerated reactors. These plants are listed in Table 1 with their locations and various operational parameters. As a result of more than 30 years of commercial application, biooxidation is now well accepted as a pretreatment method for removal of refractory components from gold concentrates prior to cyanidation.

The adoption of the technology for base metal recovery and other elements such as radio-actinides has not yet found widespread commercial use. An excellent review by Watling cites examples of treatment applications for polymetallic concentrates, tailings, black shales, and mine wastes, indicating the limits of commercial bio-hydrometallurgical processes, other than gold, for which agitated reactors have been used (Watling 2015). Base metal processes have been developed to pilot or demonstration scale for the bioleaching of copper (Van Staden et al. 2000), nickel (Dew and Miller 1997), and zinc sulfide concentrates (Sandstrom and Petersson 1997) using mesophiles, moderate thermophiles, and thermophiles, and a process for the extraction of cobalt from pyrite concentrate was commercialized in 1998 (Briggs and Millard 1997). Bioleaching using agitated reactors for the removal of arsenic as a contaminant of nickel concentrates has been proposed (Fewings et al. 2015), and more recently, a project has proceeded to construction status (Neale et al. 2015), but the operational status and related information for these projects has yet to be reported.

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Table 1 Chronology of recognized commercial bacterial oxidation plants for refractory gold processing

Plant and Location	Operating Years	Design Capacity, stpd concentrate	Total Reactor Volume, m^3	Approximate Percentage S^{2-} in Feed	Technology Supplier	Reason for Closure	References
Fairview, South Africa	1986–present	10 (1986) 35 (1991) 40 (1994) 55 (1999)	1,260	18–24	Biox	N/A	Van Aswegen 1993; Van Aswegen et al. 2007
Tonkin Springs, Nevada, USA	1989–1990	1,500 (whole ore)	8,800	1.3	Own installation	Funding constraints and technical issues	Foo et al. 1990
Sao Bento, Brazil	1990–present	150 (1990) 300 (1994) 380 (1997)	1,300	19	Biox	Bacterial oxidation section shut down for energy saving	Suttill 1990; Van Aswegen et al. 2007
Harbour Lights, Western Australia	1992–1994	40	980	18	Biox	Ore depleted	Brierley 1995
Wiluna, Western Australia	1993–present	115 (1993) 158 (1995)	4,230	24	Biox	N/A	Brown et al. 1994; Stephenson and Kelson 1997; Odd et al. 1993
Sansu, Ashanti, Ghana	1994–present	720 (1994) 960 (1995)	21,600	11	Biox	N/A	Nicholson et al. 1994
Youanmi, Western Australia	1994–1998	120	3,000	28 (partial oxidation)	BacTech	Ore depleted (high mining costs)	Budden and Bunyard 1994; Miller 1997
Olympiada, Siberia	1997–present	200 (approx.)	N/A	N/A	BioNord	N/A	Sovmen et al. 2008
Proano, Tamboraque, Peru	1999–2003	60	1,572	30	Biox	Closed but reopened	Loayza and Ly 1999
Beaconsfield, Tasmania	2000–present	60–70	2,241	27–34	BacTech Mintek	N/A	Pinches et al. 2000
Laizhou, Shandong Province, China	2001–2010	100 (2001) 200 (2008)	4,050	21–25	BacTech Mintek	Change of owner (on care and maintenance)	Miller et al. 2004
Suzdal, Kazakhstan	2005–present	196 (2005) 520 (2009)	7,800	12	Biox	N/A	Van Aswegen et al. 2007
Fosterville, Australia	2005–present	211	5,400	21	Biox	N/A	Whincup and Binks 2004; Amiconi et al. 2004
Bogoso, Ghana	2006–present	825	21,000	20	Biox	N/A	Van Niekerk 2012
Jinfeng, China	2007–present	790	16,000	9.4	Biox	N/A	Van Niekerk 2012
Kokpatas, Uzbekistan	2009–present	2,138	43,200	20	Biox	N/A	Van Niekerk 2012
Agnes, South Africa	2010–present	20	396	30	Biox	N/A	Van Niekerk 2012
Runruno, Philippines	2014–present	404	10,800	17	Biox	N/A	Olivier and Jardine 2014

Adapted from Miller and Brown 2005
N/A = not applicable.

GENERIC FLOW SHEET SHOWING UNITS AND BASIC PROCESSING

Bioleaching and biooxidation processing using agitated reactors has been confined to concentrate treatment, with a single exception using whole ore as feed at Tonkin Springs (Nevada, United States) in 1989 (Foo et al. 1990). A block flow sheet of typical unit operations involved in biooxidation for treatment of refractory gold concentrate is shown in Figure 1, whereas a flow-sheet example of bioleaching for recovery of base metals (copper) and precious metals from a polymetallic concentrate is given in Figure 2.

The upstream unit operations for concentrate treatment to provide feed for biooxidation or bioleach processing are common to all projects in terms of crushing, grinding, gravity gold recovery if appropriate (Holder 2007), and concentrate production by flotation. Concentrate feed slurry is drawn continuously from the stock tanks and diluted to the working pulp density for bioprocessing. A typical biooxidation plant consists of one or more trains of six agitated, aerated reactors with three reactors operating in parallel as primary reactors and three operating in series as secondary reactors. The reaction is exothermic and reactors are cooled with water to maintain a set temperature, generally between 42°C and 50°C, depending on the bacterial culture in use (Brown et al. 1994; Budden and Bunyard 1994).

After bioprocessing, a countercurrent decantation (CCD) circuit can be used to wash the solids and separate the liquor for further treatment. Filtration and washing can be used as an alternative to a CCD circuit (Miller et al. 2004). For biooxidation, the liquor containing acidic ferric and arsenic elements is neutralized in stages by addition of lime or limestone to produce a stable ferric arsenate cake or slurry for disposal and a clean water for reuse within the plant (Nyombolo et al. 2000). The biooxidized solids are neutralized with lime and subjected to cyanidation treatment for recovery of gold and silver values. In the case of bioleaching, the liquor containing the metals of value is subjected to routines of selective precipitation and/or solvent extraction electrowinning (SX-EW) to produce salable metal products (Briggs and Millard 1997).

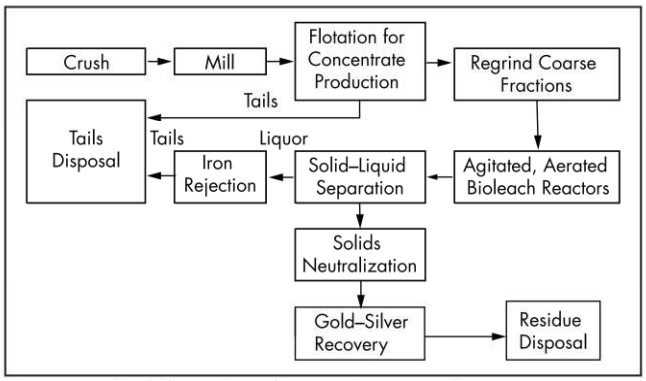

Figure 1 Block flow sheet for precious metal concentrate treatment using biooxidation in reactors

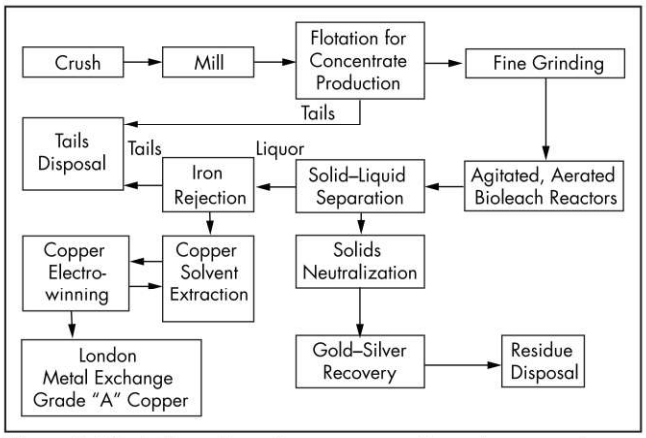

Figure 2 Block flow sheet for copper and precious metal concentrate treatment using bioleaching in reactors

If the bioleached solids contain precious metals, the solids can be subjected to cyanidation or otherwise disposed.

DESIGN CRITERIA FROM TEST WORK

The emphasis on test work is to deliver design criteria of required accuracy in a timely and cost-effective manner. Various phases of test-work investigations are normally undertaken through several sequential steps. Whether it is a requirement that every step be undertaken is project specific and typically depends on the novelty of the project and potential differences from existing commercial bioprocessing operations. An example of the sequence of test steps together with typical schedules and material requirements is given in Table 2.

Some biooxidation projects have been successfully scaled up using criteria derived from conducting only small-scale laboratory batch tests (Pinches et al. 2000). Often this is undertaken because insufficient material is available for more-detailed test studies or because sufficient confidence and knowledge is available from treating similar ore types of similar composition from other projects that use biooxidation (Miller et al. 2004). In such cases, the design criteria are based on the results from available batch testing combined with assumptions that allow the design envelope to be sufficiently flexible to cater to less well-characterized feedstocks.

Table 2 Phases of test work showing typical sample requirements and schedules for project development stages

Test Phase	Deliverable	Scope of Work	Sample	Schedule, months
1	Process amenability	Screening and testing different mixed cultures with ore or concentrate samples	2–5 kg	3–4
2	Scoping study data	Batch testing using bioreactors of a few liters' volume and first metal recovery tests	5–20 kg	4–6
3	Scoping–prefeasibility study data	Batch testing using larger bioreactors, giving greater amounts of liquor and residues for metal recovery and environmental testing	20–100 kg	6–8
4	Prefeasibility–feasibility study data	Semicontinuous–continuous testing using a train of laboratory-scale bioreactors with larger-scale batch downstream testing	100–500 kg	6–12
5	Feasibility	Fully integrated, continuous demonstration-scale pilot plant in purpose-built facility	20 t+	9–18

Residence time calculations for a commercial plant can be modeled from continuous pilot-plant campaigns using commercial feedstock by producing a series of operating curves (Van Aswegen et al. 2007). The theoretical modeling of biooxidation and bioleaching processes for design purposes can be complex, partly because of the partition of the bacterial population between solids and liquid; both phases also supply substrate for growth and division, complicating modeling exercises (Gonzalez et al. 2004; Dinkla et al. 2013).

FEEDWATER QUALITY, DILUTION, AND RESIDENCE TIME

An important criterion for bioprocessing is the need for a reasonable quality of makeup water. Although this can be true for certain metallurgical processing operations, largely because of issues of corrosion and scaling, the possibility of certain ions being detrimental because of toxicity is a further consideration in bioprocessing. Water enters the process from several sources:

- Water associated with the concentrate from upstream processing
- Dilution water added to achieve the working pulp density in the reactors
- Water used for reagent makeup, such as nutrients and limestone slurry

An early area for concern was the potential toxicity that residual flotation reagents associated with the concentrates may have on bacterial activity. Conducting toxicity testing using the intended flotation reagents is standard protocol during project development phases. Although there has been little evidence that carryover of these reagents can cause

toxicity issues, many operations have facilities for a concentrate regrind or a re-upping step, and this would serve to dilute reagents from flotation. Cyanide, which is highly toxic in all forms to the bacteria, can be used as a pyrite depressant in flotation and should be avoided when producing concentrate feeds for bioprocessing. Van Aswegen et al. (2007) list the following reagents that may show toxic effects toward the culture:

- Cyanide and thiocyanate
- Grease and oil
- Detergents, solvents, and degreasing components
- Biocide descaling reagents and other water treatment reagents
- Certain flocculants, flotation reagents, and nutrients

Chloride is the most common detrimental ion that can be present at elevated levels in some waters. Historically, some biooxidation processes have operated well in arid regions where water makeup is limited and only relatively saline water can be used (Bell and Quan 1997). Van Aswegen et al. (2007) suggest that 5,000 ppm chloride is a maximum acceptable limit for chloride in the biooxidation operation using Biox technology with a total dissolved solid of 2,100 ppm and that arsenic (V) concentrations of 15–20 g/L are tolerated.

Accurate metering of dilution water to achieve the required pulp density of feed to biooxidation is critical, as the pulp density of the operation is a key parameter in dictating the rate of oxidation and operational efficiency. The parameters of feed pulp density, residence time, and pulp feed rate are the major process parameters for control. If the operating pulp density and/or pulp flow rate is too low, the efficiency of the plant suffers as oxidation is achieved well within the residence time and the mass transfer capabilities of the reactors are underutilized. Conversely, if the pulp density or/and throughput are increased, only a partial oxidation is achieved within the residence time and final metal recovery may decrease. It is unusual for the pulp density of biooxidation feed to exceed 20%, as above this value and irrespective of concentrate type, the rate of oxidation appears to decrease (Bailey and Hansford 1993).

TANK CONFIGURATIONS AND MATERIALS OF CONSTRUCTION

The process objective is to maximize the effectiveness of reactor volume, with limiting factors in primary reactors typically being related to oxygen transfer and in secondary reactors being particle size or inhibition because of increasing ionic composition of the liquor. Reactors are normally of similar dimensions but with variations in agitator, cooling, and aeration duty. A typical tank configuration consists of half of the entire reactor volume being primary reactors operating in parallel with feed being divided equally. The remaining tanks consist of reactors operating in series in which the partially oxidized pulp exiting the primary reactors is combined and fed to the remaining reactors operating in series. A configuration of this type ensures that sufficient residence time is available in the primary reactors for growth and division of the bacterial population. For many plants, a three-day residence time in the primary reactors is common. Some plants will amend the tank configurations to operate four reactors as primary reactors and only two reactors as secondary reactors, or to install seven reactors per train with four as primary reactors and three as secondary reactors (Van Aswegen et al. 2007; Miller and Brown 2005). Such variations can allow a greater tonnage throughput, resulting in an overall increase in metal production but may sacrifice recovery if residence time is too low. Operating a plant with a low residence time requires a greater fineness of control to ensure a stable operation.

The acidic nature of the pulp and often the consideration of water quality require the use of rubber-lined steel or specialized steels as materials of construction. This includes SAF 2205 and other high-grade stainless steels (Ritchie and Barter 1997). The largest tank sizes believed to be in use are 1,500 m^3 in size at the Kokpatas project in Uzbekistan commissioned in 2009. Typically, the water quality in terms of chloride concentration, in combination with the acidity, dictate the grade of stainless steel required. Specialty steels with higher tensile strength are becoming more favored. Although they are more expensive on a weight basis, they can be more cost-effective because a thinner plate thickness is required.

Acid-resistant construction materials are also required in the interface operations of solid–liquid separation, residue washing, and liquor neutralization. Thickeners used in countercurrent washing require acid proofing, and primary neutralization tanks are normally constructed from stainless steel. The bioreactors are generally contained in a single concrete bund area sized to contain the entire volume of a single reactor for emergency draining purposes. This concrete must also be acid-proofed, as must all sump areas likely to be in contact with acidic pulp. A high-quality epoxy coating can be used for concrete protection. Galvanized steels are not usually permitted in the biooxidation area because of the possibility of arsenic solutions reacting with the zinc to produce toxic arsenious gas. Walkways and areas of the superstructure located above the reactors also require corrosion protection with relevant coatings. The use of plastics in certain areas of some newer plants has become more common in recent years as a lower-cost alternative to steels. Smaller tankage, such as that in nutrient makeup or acid, is constructed from epoxy or equivalent plastics.

FEED PARTICLE SIZE AND FINE GRINDING OF CONCENTRATE FEEDS

Because coarse fractions are present in most concentrate feeds, it is normal practice to regrind coarse components as part of the feed preparation before concentrate is stored in stock tanks. Components coarser than 120 μm are generally reground to avoid the following:

- Solids settling in the bioreactors forming large fillets over time between the baffles at the base of the reactor. This gives a considerable metal "lock-up" within the biooxidation process overtime.
- Coarse components may contain higher gold values with a higher specific gravity that settle more readily in reactors.
- When a substantial amount of the reactor volume is occupied by settled solids, these solids becomes severely compacted and are difficult to resuspend. The reactor must be taken off-line and drained, and the solids must be removed manually.
- An extensive buildup of solids reduces the reactor volume, making it impossible to achieve the required throughput and meet design specification.

Regrinding even larger proportions of the concentrate to improve oxidation kinetics is a favored method for either increasing plant throughput above design specifications or

ensuring the complete oxidation of more refractory minerals. Although theoretically the throughput design basis of primary reactors is fixed, the conservative warranties offered for oxygen transfer are usually exceeded in practice, making particle size the rate-limiting factor for oxidation. Without regrinding the concentrate, most of the fractions would readily meet the criteria for suspension by the specified agitation duty, but particle size can be the rate-limiting step in oxidation kinetics (Holder 2007).

AGITATION AND AERATION

Agitators are used to both suspend solids and disperse air. Air is typically provided using low-pressure blowers and is delivered to a ring main beneath the agitator for dispersion. The air exiting the blowers is at an elevated temperature and is usually precooled before delivery to the reactors. Because the oxidation process results in a decreased solids density through each stage of the process, accompanied by lower air demands for reaction, each reactor stage will have different agitation and aeration needs. Air is therefore metered to reactors according to the oxidation requirements for a reactor.

Agitation with solids suspension and high aeration in large reactors have created a unique duty for the type of agitators used in biooxidation. The hydrofoil designs developed in the 1980s have been traditionally favored for this three-phase mixing duty (Lally 1987; Fraser 1993; Kubera and Oldshue 1992). Gas-transfer coefficients are similar or superior to turbines but with lower power requirements, whereas solids suspension is homogeneous. However, large airflows are a recognized characteristic of bacterial oxidation reactors, owing to typical utilization factors of only between 30% and 40%, combined with the high demand for oxygen in the reaction. Air-to-sulfide ratios of between 25,000 and 30,000 nm^3/t (normal cubic meters per metric ton) of sulfur are common values in design (Miller and Brown 2005). The agitator must disperse the air while ensuring that flooding is avoided and solids suspension is maintained. The size of the agitator is thus a function of the air-expansion thermodynamics. A specific agitator power of 30–35 nm^3/h of air per installed kilowatt is often used for first basis agitator sizing, and it has been implied that the conventional power requirement for agitation on traditional plants is between 0.2–0.3 kW/m^3 of reactor volume (Miller and Brown 2005).

Airflows and agitator power requirements in design are sized accordingly to the requirements for each reactor stage in the process. It is normal practice for the first secondary reactor to have an agitator of comparable size to the primary reactors to allow for maintenance, such that the secondary reactor can also have the duty of a primary reactor. Typically, two-thirds of the oxidation is completed in the primary reactor stage. Traditionally, the agitators have used fixed speed, with adjustments made only by changing the drive belt and the gear wheels. The use of variable-speed drives is now gaining favor in new plants because it allows the optimization of the mixing duty of each reactor. Although it may add to the plant capital cost, the operating cost savings from optimizing power draw for each individual reactor can be considerable. The failures of agitators and associated equipment such as gearboxes in biooxidation plants has also been higher than expected because of high bending moment, torque, and thrust. More recently, the use of dual impellers has been favored and reported as being more cost-effective while achieving greater operational robustness. Such improvements to agitation/aeration systems are claimed to reduce agitator power requirements by 20% while achieving the same oxygen mass transfer requirements (Olivier and Jardine 2014). Experts suggest that this dual-impeller system also offers a reduction in bending moment, torque, and thrust that will improve the mechanical life of the agitator system.

Several considerations are necessary in the design specification and operation of the low-pressure air blowers used to deliver air to the reactors. First, design calculations for the total air blower requirement should include the site elevation of the project to account for the lower oxygen content of air at heights above sea level. Second, one must determine the number of blowers to be used to deliver the total air required. Optimization of capital costs suggests the use of a fewer number of large blowers. By contrast, optimization of operating costs suggests the use of a larger number of smaller blowers, such that the number of blowers operating can be managed according to variations in air demand because of feed variations. The turndown ratio of blowers is generally limited, representing a fixed operating cost when in use. The siting of blowers also needs consideration. Usually the blowers are separately housed for noise abatement and positioned away from areas prone to high dust levels. Cable routings may also need consideration to reduce the potential for electrical interference.

COOLING SYSTEMS AND WATER REQUIREMENT

The heat generated from the exothermic reactions is often a few megawatts of energy. Heat is removed in situ using cooling water to retain a set-point temperature conducive to bacterial activity. The set-point temperature chosen depends on the microbial population used in the process as specified by the technology vendor. The optimum set-point temperature for many plants using mesophilic cultures is generally between 40° and 42°C, but for plants using moderate thermophiles, a temperature of about 50°C is typically the optimum set point.

Bundles of cooling water tubes inside the reactor are favored for heat removal and often positioned at the walls of the tank to act as baffles. A less sophisticated method applied at one project has been to use a film of cascading water on the outside of the reactors (Budden and Bunyard 1994).

Cooling water bundles are prefabricated off-site from a similar grade of stainless steel to be used in the reactors. The cooling water bundles contribute a significant cost to the overall capital cost of the reactor. This is because of the substantial number of coils required, combined with the labor content for custom fabrication.

Some heat is removed from the reactor by evaporative loss, by the effects of aeration, as well as by heat being lost through the tank walls. The amount of heat removed in this way depends on the local climate but is generally low. Concerns are often raised that operating reactors in extremely cold climates is untenable because of high heat losses from tank walls, preventing the temperature from being maintained. Heat balance calculations, as well as data from commercial plants that operate in extremely cold climates with high wind chill factors, has shown this not to be the case and heat removal is always required. An exception may be the final reactor where oxidation is nearing completion and heat generation is comparatively low. These reactors may require external jacketing to retain more heat or, depending on plant design, hot cooling water exiting the primary reactors' cooling water circuit can be redirected to these reactors before being returning to the cooling tower.

The heat generated is low grade and is not generally useful to transfer to other areas of the plant. The heat is removed from the cooling water using conventional forced-draft towers in a closed circuit. Failure of the cooling water supply can have a severe effect, with an excursion of temperature above the set point significantly impairing the process. In general terms and by empirical observation, a temperature rise between approximately 3° and 5°C is the maximum tolerance before significant process impairment occurs. Because the cooling towers are open to atmosphere, biocide and other chemicals such as antiscalants must generally be added to mitigate fouling and maintain the heat transfer duty. The particulate load of the water also increases because of the scrubbing action of the water as it passes through the tower. In-line ultraviolet irradiation devices can reduce the need for biocides (Holder 2007), and sand filters can reduce the particulate matter collected in the water over time. For some projects operating in regions of elevated temperature and humidity, the use of cooling towers is much less effective because of the cooling requirement operating close to the ambient wet-bulb temperature, and refrigeration may be a preferable concept for consideration. Closed-circuit refrigeration can also be favored for reducing needs for water treatment associated with cooling water towers and occupies a smaller footprint.

FEED SUPPLY, REAGENT ADDITIONS, SPILLAGE HANDLING, AND PULP TRANSFER

Primary reactors receive diluted concentrate feed slurry from a header feedbox. Acid or limestone additions are made, if required, to each reactor to adjust the operating pH value to a set point. The header box located above the nest of primary reactors is fitted with a splitter arrangement such that feed can be distributed equally into the primary reactors. A timed actuator is often fitted to allow feed rates to be adjusted between reactors when required. Nutrient solution is also dosed to the header box to mix with the concentrate before entering the primary reactors. Nutrients are predissolved in a mixing tank before reporting to a nutrient holding tank ready for use. The nutrients resemble an agriculture fertilizer mix consisting of ammonium, phosphate, magnesium, and potassium salts and can be delivered to the site either as agricultural-grade chemicals in bags or as a specified premixed pellet for dissolution. The bacteria utilize carbon from carbon dioxide to form new cells, and the carbon dioxide formed by the reaction of acid with the carbonates present in the concentrate may be an important supplement to the natural supply of carbon dioxide that is present in the air supplied to the reactors for oxidation. A very low percentage of carbonates is ideally required in the feed, which is neutralized by the natural acidity generated from oxidation in the primary reactors to supply this need (Van Aswegen et al. 2007). If the concentrate feed contains excessive carbonates, then acid addition will be made to the header feedbox to neutralize excess alkalinity. Concentrated sulfuric acid is typically metered and pumped directly from an acid storage vessel to the header feedbox.

The ability to add limestone slurry to all reactors is common, with limestone being supplied through a ring main. Limestone additions are made to reactors if the final acidity level produced by oxidation becomes too high and the pH value is below the optimal value for bacterial activity as specified by the technology vendor. A typical range of pH values for primary reactor operations may be 1.5–2.5 depending on feedstock and culture type. A typical range of pH values for oxidized pulp emerging from the last secondary reactor may be 0.8–1.8, again depending on technology vendor specification for optimum microbial activity and the type of concentrate feedstock being treated.

To reduce reagent additions, and when feasible, many operating plants aim to minimize the need for pH adjustments and achieve a good operating acid balance in which the natural acidity generated from oxidation is used to neutralize any excess carbonate in the concentrate feed. Excessive acid addition to the concentrate feed will require excessive use of limestone in downstream neutralization. Excessive limestone addition to reactors also gives rise to excessive gypsum and ferric hydroxide precipitates, creating an extra mass of solid to be carried forward into downstream processing. There has been little evidence that such precipitates hinder subsequent gold extraction by cyanidation.

Formation of a stable foam on the surface of reactors tends to occur when changes are made to the process. Such changes may include a sudden variation in feed rate or type of feed, or introduction of a toxin, all of which may provoke a microbial response, resulting in a highly stable foam. Antifoams can be used for foam suppression, but the choice of antifoam is limited to those considered nontoxic. Such antifoams are likely to be more expensive, and the amount that can be used may be limited. Many plants have a high freeboard allowance on reactors to allow foam to be contained in the reactors as part of normal operational practice. Some plants have deep launders that transfer pulp between reactors, which helps to remove foam from the primary reactors where it is often more prevalent. Water sprays positioned on top of reactors or launders have also proved successful in combatting excessive foaming. If foaming is not controlled and there is extensive overflow from the top of reactors, the foam collects in the bund area and is pumped back to the reactors. The reason for the occurrence of foam must be understood and addressed, as it should only be a temporary feature of an operation while undergoing change from a steady-state condition.

Various methods have been adopted for transferring pulp between reactor stages, with airlifts or gravity flow by launders being favored. It is important to maintain a consistent level of pulp in reactors while achieving good pulp transfer, and airlifts can be problematic for this reason (Holder 2007). Launders are a simpler, effective alternative, subject to being correctly designed to prevent solids settlement during longer sections. Slurry transfer using centrifugal pumps has also been used in two plants (Spencer et al. 1997). During plant operation, occasionally significant quantities of pulp must be transferred from one reactor to another. For example, this is required when starting up a reactor after scheduled emptying and maintenance, or for cross-inoculation of reactor contents to aid recovery from an unscheduled shutdown. For this reason, it is useful to be able to connect reactors with temporary pipework and valving arrangements at the reactors' base to rapidly transfer large amounts of pulp between the reactors.

SOLID–LIQUID SEPARATION, LIQUOR NEUTRALIZATION, AND WATER REUSE

After biooxidation, the liquor is separated from the oxidized solids, which are also washed in preparation for lime guarding and cyanidation. CCD or thickening followed by wash filtration can be used. CCD circuits were favored in early plants, but filtration with washing is also considered as an established alternative. Some plants have experienced gold losses from

entrained solids reporting to the thickener or CCD liquor overflow ahead of liquor neutralization, resulting in the use of a polishing filter to mitigate such losses (Holder 2007). Current generally accepted practice is to include a polishing filter or separate settlement catchment tank to capture any entrained solids that would represent a gold loss. This polishing filter or catchment tank is also important to prevent solid losses occurring in the overflow for periods when the CCD circuit is unstable.

Neutralization of bioliquor is typically conducted in four sequential reactor stages with a total nominal residence time of about six hours. Reactors are agitated and a small amount of aeration supplied to convert any residual ferrous iron present to the ferric form (Nyombolo et al. 2000). Limestone is generally the reagent of choice, but the superior neutralizing qualities of lime can be more economical if reagent transportation costs are a significant factor for the project. Typical limestone use ranges from 400 to 800 kg/t of concentrate feed, depending on the sulfide mineralogy and sulfide concentrations present in the feed. Despite the high limestone use, local limestone is usually available to reduce supply costs. Alternative sources of alkalinity have been used on some projects, including the use of flotation tails high in calcite providing a source of carbonate that reduces the total quantity of commercial limestone or lime required (Van Aswegen et al. 2007). When using commercial limestone, it is either delivered finely ground and pneumatically discharged into an on-site silo or delivered as crushed limestone and stockpiled. In the latter case, on-site grinding is conducted prior to preparation of a slurry, which is then delivered to the ring main for use in liquor neutralization or pH adjustment of bioreactors.

In general, when arsenic is present in the liquor, a minimum molar ratio of iron to arsenic of 3:1 is necessary during neutralization to produce a ferric arsenate compound with long-term stability for disposal (Ringwood 1995; Nyombolo et al. 2000). Monitoring of such ferric arsenic wastes over time has verified the stability of these wastes (Kraus and Ettel 1989; Broadhurst 1993; Van Aswegen et al. 2007). The solids produced from neutralization of biooxidation liquor contains gypsum as the major compound, with solid ferric hydroxide species in lesser amounts and ferric arsenate present as a minor compound in the total waste.

The neutralized liquor containing the precipitated solids reports to a thickener for settling followed either by filtration or discharging the thickened underflow directly to impoundment. The clarified solution from the thickener together with filtrates or return water after impoundment is transferred to process water storage for reuse. Creating a high underflow density of the precipitate together with poor filtration characteristics can cause problems in plant operations. Plate-and-frame filters are generally favored for filtration duty, and slow filtration times result in relatively large filtration areas. Although the final filter cake is quite stable, large amounts of water are contained within the chemical matrix; this usually represents the major source of permanent water loss from the circuit. The neutralization process generally results in clean water that is of good quality but that is high in calcium and has an increased tendency for scaling.

Whether water is reclaimed from the ferric precipitate impoundment area or retrieved from filtration at plant level is project specific and related to arrangements and siting for tailings disposal. The expense of construction and the permitting of tailings impoundments can lead to severe restrictions in volumes available for tailings disposal, which encourages filtration and stacking of tailings to reduce the final volume for impoundment.

In many plants, the water circuit for biooxidation is separate from the water circuit for cyanidation. This is necessary to prevent contamination of toxic cyanide species to the biooxidation circuit (Van Aswegen et al. 2007). Technologies are available for tertiary detoxification for degradation of thiocyanates. Use of such technologies is likely to increase because they involve a single water circuit of less complexity and reduce costs (Van Buuren 2014).

RESIDUE TREATMENT PRIOR TO CYANIDATION AND BASE METAL RECOVERY FROM BIOLIQUOR

As biooxidation residue often contains some transient sulfur species, lime guarding prior to cyanidation is a critical step to reduce downstream cyanide consumption (Hackl 1989; Van Aswegen et al. 2007; Miller and Brown 2005). Methods to protect residues from consuming excessive cyanide include more intensive lime guarding, or extension in biooxidation residence time (converting transient sulfur species to sulfates); and fine grinding of the feed, which may also improve the kinetics of oxidation as well as give a more benign oxidized residue for cyanidation.

For base metal bioleaching, these elements become soluble in the leach for recovery with conventional SX-EW or selective precipitation routines. An example of this was shown by the Kasese Cobalt operation with sequential removal of iron, zinc, and copper from the bioleach liquor, followed by cobalt SX-EW (Briggs and Millard 1997). These downstream metal rejection and recovery routines contrast with the biooxidation for precious metal processing, in which gold, silver, and platinum group elements are exposed by removal of sulfides but remain inert in the oxidized residue for later downstream recovery.

TAILINGS DISPOSAL AND OTHER DISCHARGE CONSIDERATIONS

Two distinct types of tailings are produced from bioprocessing. One is the final leached residue after metal extraction, which may have also included cyanidation and detoxification treatment. The second tailing is the ferric precipitate from liquor neutralization, which is likely to contain stabilized ferric arsenate if derived from refractory gold concentrate processing. Disposal methods of each of the tailings will vary according to their characteristics and the specific project requirements to meet permitting and monitoring requirements. Separate disposal facilities are generally favored. Leach residues are particulate, which is relatively easy to filter and of smaller volume. By contrast, the ferric or ferric arsenate precipitate is voluminous, contains significant amounts of water, and is more difficult to thicken to a high density and to filter. Depending on local laws, stabilized ferric arsenate meeting U.S. Environmental Protection Agency or equivalent disposal requirements may be suitable for landfill, whereas cyanidation residues are more likely to require lined dams with greater monitoring and reporting requirements.

A further issue that favors separate disposal of the two tailings is water reclamation. Even after a typical detoxification process, the cyanide residue will still contain thiocyanate species that are toxic to bioprocessing, and this water cannot be reused in bioprocessing unless tertiary treatment is used to degrade thiocyanate species. For traditional plants,

reclamation of any water from a tails dam containing cyanide residues is limited for reuse in downstream processing. Conversely, water derived from the bioleach liquor after neutralization can be reclaimed and used in all process areas.

GENERIC CAPITAL AND OPERATING COSTS

Capital Costs

Published information on costs for biooxidation or bioleaching is limited. Recently, Van Niekerk reported the percentage of total capital cost that various areas of equipment contribute to the total cost for bacterial oxidation plants using Biox technology (Van Niekerk 2015). This information is portrayed in Table 3, indicating as expected that the tanks, blowers, and agitators are the major contributors to equipment cost.

Biooxidation reactors typically contribute about half of the equipment cost (Van Aswegen et al. 2007). The author's experience suggests that the capital cost of a reactor can be divided into three components with roughly equal contributions: (1) the materials of reactor construction and fabrication, (2) the cost of the internal fixtures and fabrication for cooling and air provision, and (3) the combined cost of the agitator and gearbox.

In 1996, published information on the capital costs of various biooxidation plants using different technology providers showed a range of costs according to concentrate throughput and sulfur content (Poulin and Lawrence 1996). These historic costs varied from US$2.4 million for a plant treating 40 stpd (short tons per day) concentrate rated at 7 stpd sulfur oxidation to US$25 million for a plant treating 760 stpd rated at 85 stpd sulfur oxidation. These were cited as installed costs for the total biooxidation plant. Table 4 indicates the cost values for four biooxidation facilities given in this publication and escalated to current 2017 values by factoring. Cost factoring is an inferior method for cost escalation because of the time span involved, but it still suggests a broad first estimate of possible costs for biooxidation plants. Most biooxidation plants are considered on a common cost basis in terms of the metric tons of sulfide treated per annum. Using this basis and factored data from published information, a capital cost range of US$1,100 million to US$1,700 million per annual metric ton of sulfide oxidized is shown in Table 4.

In support of these estimates, the author can offer an indicative range of costs based on values obtained by examining a portfolio of biooxidation projects or cost studies for potential projects conducted at feasibility level over a 20-year period. Projects showing extreme variations for capital costs were discounted because of distortion factors unique to certain projects. The following table shows the result from this exercise factored to 2017 values, with battery limits from receipt of thickener underflow to discharge of oxidized solids from washing in preparation for cyanidation:

Capital Cost per Annual Metric Ton of Sulfide Oxidized, US$	Capital Cost per Annual Metric Ton of Concentrate Treated at 25% S, US$
1,400–2,200	350–550

The capital cost of neutralization of bioleach liquor is also included. Assuming the treatment of a concentrate containing 25% sulfur, the cost per annual metric ton of concentrate is also indicated. Although the preceding table is only intended to provide a first-estimate indicative cost guide, it is somewhat supportive of the published costs for biooxidation plants when factored to a 2017 cost basis.

Operating Costs

Various authors have highlighted the contribution made by different areas to biooxidation operating costs. Three such studies are compared in Table 5, indicating reagents are the largest contributor to oxidation costs (Ritchie and Barter 1997; Batty and Post 1999; Van Aswegen et al. 2007). The use of reagents is project specific; limestone for neutralization typically is the largest contributing factor because of the large amounts required.

When examining the operating cost contributions, one must recognize that many of the operating costs are fixed costs; regardless of tonnage throughput, the annual cost does not vary significantly. For example, most equipment has minimal turndown capability, so that power usage is not directly related to throughput, and labor requirement and maintenance costs are often insensitive to throughput rate.

Relatively little published data are available concerning operating costs per metric ton of concentrate or metric ton of sulfide oxidized. The 1996 study referred to earlier (Poulin and Lawrence 1996) highlights two such costs from different biooxidation plants, which when converted to current values, suggest cost equivalents of US$67 and US$110 per metric ton

Table 3 Breakdown of capital cost as a function of equipment for a typical Biox project

Equipment Area	Contribution to Equipment Cost, %
Tanks	28
Blowers	23
Agitators	19
CCD thickeners	11
Stainless-steel piping	6
Waste residue tailings thickener	5
Pumps	4
Cooling towers	2
General	2

Table 4 Results from escalation of published capital costs for biooxidation plants to 2017 values

Project Number	Capacity, t/d concentrate	Sulfide, % concentrate	Oxidation, % concentrate	Sulfide Oxidation, t/a sulfide	Capital Costs		
					Original, million US$	Escalated for 2017, million US$	Escalated for 2017, t/a sulfide, million US$
1	40	20	88	2,441	2.4	4.1	1,700
2	115	24	90	8,613	5.6	9.3	1,100
3	760	11	98	29,442	25.0	40.5	1,400
4	120	28	32	3,728	3.9	6.3	1,700

Table 5 Comparison of published information showing contribution of cost components to biooxidation operating costs

Operating Cost Area	Study by		
	Batty and Post 1999, %	Van Aswegen et al. 2007, %	Ritchie and Barter 1997, %
Blowers	10	—	15
Agitators	5	—	9
Other power	6	—	6
Total power	21	34	30
Reagents	50	33	58
Labor	14	13	7
Maintenance	4	15	5
Analyses	6	—	—
Other	5	5	—
Total	**100**	**100**	**100**

of concentrate for the two different facilities. If the sulfide content and extent of oxidation are considered, these values in terms of cost per metric ton of sulfide oxidized can be recalculated to give values of US$310 and US$463 per metric ton of sulfide oxidized, respectively.

Based on experience but not verified by others, the author can offer an indicative range of operating costs obtained by examining a portfolio of biooxidation projects or cost studies for potential projects conducted at feasibility level over a 20-year period. The output from this exercise is given in the following table, suggesting a range of operating costs of between US$75 and US$175 per metric ton of concentrate, assuming a 25% sulfide content. This is equivalent to US$300–US$700 per metric ton of sulfide oxidized. Much of this variation is due to different reagent requirements and the unit power costs for the project.

Operating Cost per Ton of Sulfide Oxidized, US$	Operating Cost per Ton of Concentrate Treated at 25% S, US$
300–700	75–175

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