Improving the process performance of gold cyanide leaching reactors

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Cyanide leaching of gold is carried out in very large mechanically stirred tank reactors. The reactor size can be up to 5000 m³. A typical agitator configuration is two downward-pumping hydrofoil impellers, which requires a low mixing power input and results in low shear rates inside the reactor. This results in an economically efficient process with low levels of carbon attrition, to thus minimizing carbon and gold losses in CIL applications.

However, this kind of reactor design has some disadvantages. Efficient gas dispersion and suspension of coarse particles are hard to obtain with a low mixing power input. Thus many industrial cyanide leaching reactors currently operate at low oxygen utilization efficiencies and suffer problems with the sanding of solids at the bottom of the reactor.

In this investigation, different ways of improving solids suspension and the gas-to-liquid mass transfer properties of a typical cyanide leaching reactor configuration were studied by experimental testing and CFD modelling. The results showed that with a novel corrugated tank bottom structure, the mixing power input required to provide uniform solids suspension was reduced by 64% compared to a flat-bottom tank. Due to the lower mixing power, the intensity of attrition of activated carbon was also significantly decreased. Gas dispersion inside the reactor was enhanced with a specially designed gas dispersion chamber. An approximately 22% improvement in gas-to-liquid mass transfer was obtained with the same mixing power input by utilizing the patented Outotec OKTOP® Dispersion Chamber for gas feed.

The process performance of gold cyanide leaching reactors can be significantly improved by implementing these developments. Considerable economic savings can be achieved due to the reduction in operating costs together with decreased losses of carbon and gold.

Introduction

In the gold mining industry today, cyanide leaching is the prevailing method for solubilizing and recovering gold. Most of the new gold that is mined each year is processed via adsorption of gold cyanide onto activated carbon (Fleming et al. 2011). Typically, the carbon in pulp (CIP) or carbon in leach (CIL) process is used for gold recovery. These processes have proven to be metallurgically efficient and mechanically robust.

In order to avoid gold and carbon losses due to attrition of activated carbon, low mixing power inputs and shear rates are preferred in CIP and CIL reactors. Consequently, since efficient gas dispersion and suspension of coarse particles are hard to obtain with low mixing power input, many industrial leaching reactors operate at low oxygen utilization efficiencies and suffer problems with sanding of solids at the bottom of the reactor. Low oxygen utilization efficiency increases operating costs, since the air or oxygen feed needs to be high. In addition, a high gas feed decreases the solids suspension induced by the agitator, which can decrease the effective reactor volume and thus the residence time of coarse solid particles.

Conventional gold cyanide leaching reactor

Due to the inherently low concentration of gold in ore and the relatively long retention time required for gold to dissolve, gold cyanide leaching reactors are usually very large, with a size of up to 5000 m³. For the same reasons, the solids content in tank leaching is kept at a high level, around 40 wt.%. Currently, a typical agitator configuration in a cyanide leaching reactor is two downward-pumping hydrofoil impellers, as illustrated in Figure 1. Hydrofoil impellers generate a highly axial flow pattern for a low power draw. This makes them a good choice for most solids suspension mixing applications (Seal and Kehn, 2013). Installation of two impellers on one shaft is preferred, especially if the slurry surface height is greater than the tank diameter.
Figure 1 – Outotec OKTOP® cyanide leaching reactor with dual OKTOP®3200 hydrofoil impellers

An agitator with dual hydrofoil impellers has some clear advantages in cyanide leaching. Firstly, the requirements for mixing power and impeller torque are significantly lower than for other impeller types. The typical installed mixer motor power in a cyanide leaching reactor is less than 0.1 kW/m³; in very large-scale reactors it can even be below 0.05 kW/m³. In principle, the lowest initial capital costs and annual operating expenses for a reactor are achieved with minimized power and torque requirements. In addition, hydrofoil impellers generate a low shear rate inside the reactor. This, together with the low mixing power intensity, results in a low level of carbon attrition, thus minimizing losses of carbon and gold in carbon in pulp (CIP) and carbon in leach (CIL) applications.

However, this reactor design also has some drawbacks. The strong correlation between solids particle size and the required mixing power for solids suspension makes it difficult to achieve a homogeneous suspension. An occasional small increase in the feed solids particle size can thus lead to sanding of solids at the bottom of the reactor. The gas-to-liquid mass transfer in an agitated reactor is determined by the mixing power intensity and gas feed rate. Thus it is not possible to achieve a high oxygen utilization efficiency with a low mixing power. In addition, the presence of gas affects the ability of the impellers to suspend solids, which may also cause sanding, lower the effective reactor volume, and restrict the amount of gas that can be fed into the reactor.

Gas-to-liquid mass transfer

Since the cyanide leaching of gold requires oxygen, aeration and gas dispersion play an important role. Several studies have shown that the rate of dissolution of gold in cyanide solution is directly proportional to the amount of oxygen present in solution (Kondos et al. 1996; Ling et al. 1996; Lima and Hodouin 2005). The volumetric mass transfer coefficient $k_La$ can be used as a quantitative measure of the mass transfer capability of a reactor. According to Filippou et al. (2000), the most important factors that affects the rate of oxygen mass transfer in hydrometallurgical applications are the reactor configuration and geometry. The general equation for the estimation of $k_La$ that is frequently found in the
literature (Equation [1]) shows that $k_{La}$ is a function of mixing power intensity $(P/V)$ and superficial gas velocity $v_s$. The coefficients $A$, $B$, and $C$ are reactor- and application-specific.

$$k_{La} = A \left( \frac{P}{V} \right)^B v_s^C \quad [1]$$

Increasing the gas-to-liquid mass transfer rate in gold cyanide leaching reactors by increasing the mixing power intensity is not a good option, since it increases the attrition of activated carbon and consumption of electricity. On the other hand, increasing oxygen mass transfer by increasing the gas feed rate decreases the ability of the impellers to suspend solids and is thus restricted. Excessive gas feed with low utilization efficiency also increases capital costs of equipment and operating expenses.

Reactor gas dispersion efficiency can be improved using different sparger arrangements that inject the gas as finer bubbles than a plain gas feed pipe. However, the use of spargers, especially the types that use a high gas feed velocity for creating small bubbles, increases the capital and operating expenses. This is due to the higher pressure requirement for the gas feed because of the pressure loss in the sparger nozzles. In addition, the nozzles tend to block up easily and require regular maintenance.

**Gas dispersion chamber**

One option to improve the gas dispersion properties of a conventional cyanide leaching reactor is to implement a special gas dispersion chamber developed by Latva-Kokko et al. (2014). This patented mechanical gas sparger is installed below the lower main impeller, as shown in Figure 2. Mixing power intensity inside the chamber is approximately ten times higher than elsewhere in the reactor. Gas fed through the chamber is dispersed into fine bubbles before it enters the main circulation inside the reactor. The chamber is located so as to minimize its effect on the solids suspension induced by the main impeller. Owing to the rising gas bubbles and macro flow pattern inside the reactor, lightweight activated carbon particles do not travel extensively through the dispersion chamber. Thus the utilization of the dispersion chamber does not significantly increase the attrition of activated carbon.

![Figure 2 – Outotec OKTOP® Dispersion Chamber located below the lower main impeller](image-url)
The performance of the gas dispersion chamber was evaluated by oxygen utilization efficiency measurements. The experiments were conducted in a cylindrical transparent reactor with a diameter of 780 mm and effective volume of 373 L. The reactor was equipped with four baffles and agitated with a dual Outotec OKTOP®3200 impeller. For determining utilization efficiency, the steady-state sodium sulphite oxidation method was used. The air feed was 2.4 Nm$^3$/h, which gives the same superficial gas velocity, 0.14 cm/s, as an air feed of 400 Nm$^3$/h to a full-scale cyanide leaching tank with a 10 m diameter. The temperature was kept at 25°C and the sodium sulphate concentration above 10 g/L. The effect of the gas dispersion chamber on oxygen utilization efficiency is shown in Figure 3.

As shown in Figure 3, oxygen mass transfer can be greatly increased by using the gas dispersion chamber together with a hydrofoil agitator. For example, with a 126 W mixing power input, a 22% improvement in oxygen utilization efficiency was obtained. In order to achieve the same increase without the dispersion chamber, mixing power intensity would need to be increased by 38%, which would increase the attrition of activated carbon significantly.

In addition, the effect of the dispersion chamber on the attrition of activated carbon was tested with a similar reactor configuration. The modified wet attrition test method developed by van Dam (1993) was used. Quartz sand with a particle size range between 50 and 200 μm was used as an abrasive medium. The solid content of the slurry was 41 wt% sand, and the activated carbon content 10 g/L. The air feed and agitator configuration were kept the same as in the oxygen utilization efficiency measurement described above. Norit GCN-612G carbon was wetted and screened with an 840 μm woven wire sieve before the tests to remove fines. After 24 hours of mixing, the sample was screened through the same screen again and the carbon content of the solids that passed through the sieve was analysed. Carbon attrition with the dual hydrofoil impeller (rotation speed 260 r/min) without the dispersion chamber was 2.3 wt%. A similar mixing power input with the dual hydrofoil impeller including the dispersion chamber (rotation speed 250 r/min) induced a carbon attrition of 2.4%. When the agitator without the dispersion chamber was rotated at 291 r/min, which gives the same oxygen utilization efficiency as 250 r/min with the dispersion chamber, the carbon attrition was 4.9 wt%.

**Solids suspension**

The agitator configuration and impeller type for the gold cyanide leaching reactor have been optimized for solids suspension with minimum mixing power requirements. It may seem, therefore, that there is not much room for further
improvement. However, the solid suspension performance of a mixing tank can also be influenced by the tank geometry, especially by the shape of the tank bottom. This topic has been previously studied by Chudacek (1985), who proposed a cone-and-fillet tank bottom structure to be used with a square-pitch propeller to significantly improve the solids suspension. A similar approach, with a totally new tank bottom structure, was utilized in this study in order to further improve the solids suspension induced by the hydrofoil impeller.

**Corrugated tank bottom**

The novel patent-pending tank bottom structure developed by Outotec is shown in Figure 4. The tank bottom has a corrugated formation comprising alternate consecutive ridges and valleys that extend radially in relation to the centre of the bottom. The bottom structure straightens, concentrates, and channels the flow of the slurry and thus increases the velocity of the flow. The structure simultaneously turns the downward flow generated by the hydrofoil impeller smoothly upwards and channels it. This increases the liquid flow velocities near the tank bottom as well as in the upward direction in comparison to a flat- or dished-bottom tank.

**CFD modelling**

The performance of the corrugated tank bottom structure was studied by single-phase CFD modelling. Calculations were made for a cylindrical reactor that had a diameter and solution surface height of 8.5 m. The reactor was fully baffled and agitated with an OKTOP®3200 hydrofoil impeller. The impeller diameter was 3.5 m and the rotation speed 32 r/min. A comparison was made between flat, conventional dished (Klopperform), and corrugated tank bottom structures. The results are shown in Figures 5 and 6.
As shown in Figure 6, with the corrugated tank bottom structure the average fluid velocity near the tank bottom is approximately 1.3 m/s. This is significantly higher than the average velocities of less than 0.8 m/s obtained with flat- or dished-bottom tanks. With the corrugated tank bottom, the velocity of the rising flow also stays at a higher level in the upper half of the tank, as shown in Figure 5. Due to these higher velocities, solid particles are more efficiently suspended inside the reactor and sanding is less likely to occur.

**Experimental testing**

Testing was conducted in order to verify the performance of the corrugated tank bottom structure. The tests were carried out with laboratory-scale equipment using an OKTOP®3200 axially downward pumping hydrofoil impeller with a diameter of 145 mm. Tank 1 had a flat bottom, tank 2 a dished bottom, and tank 3 a corrugated bottom structure. All tanks had a diameter of 362 mm and were loaded with 37.3 L of quartz sand in water slurry. Thus the surface height was slightly different in each tank. Other tank dimensions, i.e., the baffle plates, impeller, and its placement, were kept constant. The solid content of the test slurry was 400 g/L. The particle diameter range of the sand was 125–185 μm.

The impeller rotation speeds required for complete off-bottom suspension (N_{js}) and uniform suspension were determined visually. Transparent tanks were attached to a torque table that was used to measure the absorbed mixing power in each case. These results are shown in Table I. The photographs of the tests in Figure 7 show the suspension cloud height with each bottom structure at a rotation speed equal to the N_{js} state with the flat-bottom tank.
Table I. Impeller rotation speed and absorbed mixing power required for complete off-bottom suspension and uniform suspension of sand slurry (400 g/L, 125–185 μm)

<table>
<thead>
<tr>
<th></th>
<th>Flat bottom</th>
<th>Dished bottom</th>
<th>Corrugated bottom</th>
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<tr>
<td><strong>Off-bottom suspension</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>r/min</td>
<td>330</td>
<td>390</td>
<td>285</td>
</tr>
<tr>
<td>W</td>
<td>5.37</td>
<td>8.87</td>
<td>3.46</td>
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<tr>
<td><strong>Uniform suspension</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>r/min</td>
<td>490</td>
<td>500</td>
<td>350</td>
</tr>
<tr>
<td>W</td>
<td>17.6</td>
<td>18.7</td>
<td>6.41</td>
</tr>
</tbody>
</table>

As shown in Table I, the mixing power input required to provide a uniform solids suspension was 64% smaller with the corrugated tank bottom structure compared to the flat-bottom tank. Similarly, a complete off-bottom suspension was achieved with 36% lower mixing power. The performance of the dished bottom was the worst in this comparison, due to a stagnant region directly below the impeller and the highest slurry surface height with the same volume.

Continuous suspension level measurement and control of mixing

The particle size distribution and solids concentration have a strong effect on the degree of suspension. In continuous industrial operation, both of these properties typically fluctuate. Thus the required mixing power also alters over time. Outotec has developed a measurement technology for the continuous surveillance and control of the degree of solids suspension inside a slurry reactor. The measurement is made by electrical impedance tomography using probes that are integrated into the reactor, as shown in Figure 8. The cloud height of the slurry is measured using OKTOP® CloudSense, and the thickness of the settled layer of solids at the bottom of the tank with an OKTOP® SandSense probe. The mixing power can be controlled based on these measurements through the variable-speed drive of the mixer motor.
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With continuous suspension level measurement, the mixing power intensity inside the reactor can be optimized. Thus, harmful attrition of activated carbon is minimized without the risk of sanding. Latva-Kokko et al. (2014) have reported one industrial application where a 54% saving in mixing energy consumption was achieved with this technology compared to operation with the nominal rotation speed of the impeller.

Attrition of activated carbon

Attrition of activated carbon causes gold and carbon losses in CIP and CIL processes. Typically, attrition values in CIP and CIL processes are 40–60 g of carbon per ton of ore (Marsden and House, 2006). According to van Dam (1994), the gold loading takes place predominantly in the outer shell of the carbon particles and thus abrasion fines have a relatively high gold content. In financial terms, gold losses are more serious than carbon losses.

Coetzee and Cloete (1989) studied the attrition of carbon in an agitated tank with different types of impellers. They concluded that the proportion of carbon fines ($X$) is correlated with the total amount of energy transferred to the suspension by the mixer ($E$) after the initial conditioning phase, according to Equation [2]. The same equation was valid for all the impellers studied.

$$X = 0.056E^{0.6}$$  \[2\]

Based on the correlation presented in Equation [2], the 64% reduction in mixing power input that could be obtained with the corrugated tank bottom structure would generate 46% lower attrition of activated carbon compared to a conventional reactor design. In fact, this is a rather conservative estimate, since the attrition tests carried out by Coetzee and Cloete did not include the effect of the ore on the attrition rate. Based on the dispersion chamber tests described earlier, the effect of increasing the mixing energy on carbon attrition was significantly higher when sand particles were present in the slurry.
Gold and carbon losses due to attrition of activated carbon

In order to evaluate the financial effect of carbon attrition, a dynamic carbon management model developed in-house was used to calculate gold and carbon losses for two attrition rates in a CIP process. The compared attrition rates were 60 g and 32.5 g per ton of ore.

The attrition effect was calculated for a six-reactor CIP train. The effective volume of each reactor was 1500 m$^3$. The slurry flow was 350 m$^3$/h, containing 45 wt% solids. The specific gravity of the solution was 1 kg/L, and that of the solids 3.5 kg/L, leading to a solution flow of 284 m$^3$/h. The adsorbed gold amount was 2.2 mg/L (solution). An annual running time of 8000 hours was applied.

The activated carbon concentration in each reactor was 5 g/L (7.5 t per reactor), i.e. a total of 45 t of carbon in the train at the beginning of the process. Approximately 95 wt% of the carbon was transferred in 6 hours each day. The carbon was pumped at the same time from each reactor. The loaded carbon contained 2714 g Au per ton of carbon, and the regenerated carbon 20 g/t.

The attrition rate was the same in each reactor. In addition to the gold loading on retained carbon, the retention time of the pulverized carbon in the system was taken into account in the gold loading.

The results are shown in Table II. The annual production of gold without any carbon attrition would be 4984 kg. The loss of gold with the pulverized carbon at a 60 g/t attrition rate would be 64 kg/a (2060 oz), decreasing to 35 kg/a (1122 oz) with a 32.5 g/t attrition rate. The carbon loss would be 111 t/a and 60 t/a at attrition rates of 60 g/t and 32.5 g/t, respectively. The annual saving due to the reduced attrition and energy consumption would be US$1.4 million, using a gold price of US$1200 per ounce, a carbon price of US$1000 per ton, and an energy price of US$50 per MWh.

<table>
<thead>
<tr>
<th>Carbon attrition</th>
<th>Value</th>
<th>$ million</th>
<th>Value</th>
<th>$ million</th>
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<tbody>
<tr>
<td></td>
<td></td>
<td>32.5 g/t</td>
<td>60 g/t</td>
<td></td>
</tr>
<tr>
<td>Gold loss</td>
<td>35 kg</td>
<td>1.35</td>
<td>64 kg</td>
<td>2.47</td>
</tr>
<tr>
<td>Activated carbon loss</td>
<td>60 t</td>
<td>0.06</td>
<td>111 t</td>
<td>0.11</td>
</tr>
<tr>
<td>Mixing energy consumption</td>
<td>2592 MWh</td>
<td>0.13</td>
<td>7200 MWh</td>
<td>0.36</td>
</tr>
<tr>
<td>Total value</td>
<td></td>
<td>1.54</td>
<td></td>
<td>2.94</td>
</tr>
</tbody>
</table>

Conclusions

The process performance of gold cyanide leaching reactors can be significantly improved by implementing the developments described in this article. Considerable economic savings can be achieved due to the reduction in operating costs together with decreased carbon and gold losses. With gas dispersion chamber reactors, the oxygen utilization efficiency can be improved without a major increase in the carbon attrition rate. This improvement might also make it possible to use air instead of oxygen in some leaching applications. The required mixing power intensity, which is directly linked to gold loss by carbon attrition, can be greatly reduced with the novel corrugated tank bottom structure. Finally, the mixing power intensity during operation can be optimized using continuous suspension level measurement and active mixing control. All of these developments can be combined and installed in a conventional cyanide leaching reactor as required, as shown in Figure 9.
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Figure 9 – Outotec OKTOP® cyanide leaching reactor equipped with OKTOP® DispersionChamber, corrugated tank bottom, and continuous suspension level measurement

References


The Author