Solvent extraction (SX) reagent selection for high temperature, acid, chloride and Cu PLS at Port Pirie and its impact on electrowinning (EW)

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Nyrstar’s Port Pirie Copper Plant produces copper cathode from the lead smelting process copper/lead matte by-product. More than 98% Cu is recovered from the matte by a unique mixed chloride-sulphate leach technology (formerly known as the BHAS process) followed by conventional copper solvent extraction and electrowinning processes.

The highly effective agitated chloride-sulphate leach process presents several challenges for the operation of the downstream Cu SX-EW plant, namely it:

- Produces a high temperature PLS with high concentrations of acid, chloride and copper
- Requires effective solids liquid separation and soluble silica control
- Requires high O/A ratios and high concentrations of extractant

The strategy that was undertaken to examine the historical performance of the Port Pirie Cu SX plant and the development of a planned approach on implementing effective measures to combat the high chloride transfer to the EW process is outlined in this paper. This involved determining the plant organic constituents and understanding the characteristics that presented the underlying causes to many of the Cu SX-EW plant problems being encountered. Operational data collected during the Cu SX plant reagent transition phase is presented. The data illustrate the changes in organic chemistry and organic health that ultimately produced a very different EW electrolyte to that previously experienced in this operation. A key result of this reagent change and attention to detail in Cu SX was the significant reduction of the chloride concentration in the Port Pirie EW plant electrolyte. From historical levels of >200 ppm, which required the use of titanium cathodes, electrolyte chloride levels were reduced to less than 30 ppm.
Introduction

The Port Pirie smelter is located on the eastern side of the Spencer Gulf in South Australia, a little over 200 kilometres north of Adelaide. It has a rich history, being the nearest seaport to Broken Hill, where the rich ore field containing lead and zinc was discovered in 1883. The smelter received its first load of ore in 1889 and in due course Port Pirie evolved to be the centre of a lead smelting and refining industry. Over 100 years later what started out as a minor smelter grew to become one of the world’s largest.

Originally built in 1889 with a capacity of 80 000 tonnes of lead per year the smelter was progressively expanded in the 1950s and 1960s. The lead refinery itself was largely rebuilt in 1998 and has a demonstrated production capacity of approximately 270 000 tonnes per year of lead metal and alloys, although the lead blast furnace and sinter plant capacity limits annual production to approximately 235 000 tonnes. The precious metals refinery in which gold and refined silver are recovered after extraction in the lead refinery was also largely rebuilt in 1998. The current zinc production facilities were commissioned in 1967 and have a capacity of 45 000 tonnes per year.

Copper production facilities were commissioned in 1984 and have a capacity of approximately 4 500 tonnes per year of copper cathode. The copper units enter the site through numerous concentrates and are largely ‘free’. This makes the manufacture of copper at the smelter very economical at this small scale. In addition to lead, zinc and copper metal, Port Pirie also produces approximately 80 000 tonnes of sulphuric acid, approximately 11.5 million troy ounces of refined silver and approximately 16 000 troy ounces of gold.

Port Pirie Cu plant process description

The feed material for the Cu plant is primarily a copper-lead matte by-product from the lead blast furnace typically containing about 35% copper, 38% lead and 12% sulphur with the balance being minor elements such as arsenic, zinc, tin, silicon, antimony, iron, cobalt, cadmium and nickel.

In the 1970s the BHAS Research Department developed a process for leaching the copper matte using an acidic chloride/sulphate solution and oxygen. The successful laboratory work led to the construction and operation of a pilot plant during 1980 and the BHAS Board approved a commercial plant in September 1982. This plant was designed to have a nominal capacity of 4 000 tonnes EW Cu cathode per annum and was commissioned in 1984. The copper plant consists of five sections: grinding, leaching/solids liquid separation, SX, EW and residue neutralization. A brief description of each section follows.

Grinding

The grinding mill treats over 12 000 tonnes of crushed matte per year in a conventional ball mill charged with 90 mm diameter forged steel balls, operating in closed circuit with a spiral classifier. Thickened matte slurry, at 100%–75 micron, is pumped to a 24-hour capacity surge tank, where it is continuously agitated to prevent settling.

Leaching chemistry

The BHAS leaching process centres on an oxidative chloride/sulphate leach of the matte to extract copper into solution, leaving a solid lead and silver rich residue. Leaching is conducted at atmospheric pressure at 85°C. The overall chemistry of the process is summarized by the Equation:
The copper dissolution reaction can be described in two stages:

Equation [1] implies the dependency of the reaction on the availability of acid and oxygen. However, what is not shown is the role of the chloride in the system. The conversion of chalcocite to covellite, Equation [2], is rate-controlled by the amount of oxygen that can be dissolved into the system and available to the sulphide surfaces. Distribution of oxygen is manipulated so as to provide maximum possible oxidant in the initial leaching reactors, and ensures that approximately 50% Cu dissolution and 100% lead sulphate conversion is achieved within the first few hours of leaching.

The rate limiting parameter for the oxidation of covellite to Cu²⁺(aq), Equation [3], is the concentration of chloride in the system. The role of chloride in sulphide copper leaching is important to minimize the re-precipitation of covellite from Cu⁺(aq) if contacted with elemental sulphur:

Equation [4] high levels of arsenic also promote Cu re-precipitation in the BHAS process to form a Cu/As compound (2). It is thought that Cu⁺(aq) is unstable in low chloride concentrations, therefore chloride concentrations at the Nernst boundary layer must be sufficient to complex the Cu⁺(aq) prior to transport into the bulk solution where the oxidation of Cu⁺(aq) to Cu²⁺(aq) occurs, as:

Thus, the level of chloride in the BHAS leach process is maintained at around 20 gpl in the leach liquor to enable sufficient Cu⁺(aq) to be complexed and minimize the contact potential with S⁰ and As during leaching.

**Primary leach**

Matte slurry is fed at approximately 35 tonnes per day to six first stage leach reactors in series, each with a nominal residence time of around 40 minutes, with the leached slurry gravitating through the leach circuit via covered launders. The slurry is diluted with reheated SX raffinate and concentrated sulphuric acid, to 8% solids. Compressed oxygen is dispersed through each of the first five reactors. Leaching is maintained at an optimum temperature of 85°C with the addition of steam.

The overflow from the final leach reactor containing 7% solids gravitates to the primary leach thickener where flocculant, Magnafloc E10, is added to assist with solids liquid (S/L) separation. The hot thickener overflow, pregnant leach solution (PLS), is pumped to a void cooling tower where it is cooled so that the temperature in the PLS Tank is around 40°C. The cooled PLS solution is clarified to reduce the final level of suspended solids to SX.

**Second stage leach—high acid repulp**

The primary leach thickener underflow at 35–40% solids is pumped to the acid repulp reactor, similar to the primary leach reactors, but with a larger residence time of around 4–5 hours. All
of the chloride, as salt, and additional acid is added at this point, to ensure a high acid, high density, high chloride repulp of the leach slurry. Oxygen is also dispersed through this reactor. The repульped slurry is then washed with all the returning SX raffinate by mixing in a wash mixing tank (repульp mixer) before gravitating to the repульp thickener where Magnafloc E10 is again added.

To aid in silica control, Polysil® is also added to the repульp mixer along with the returning SX raffinate and the repульp reactor overflow. The repульp thickener overflow is directed to the first leach reactor in the primary leach. The underflow is pumped to an agitated neutralization tank.

Copper SX

The solvent extraction circuit is configured as a conventional 2E x 1W x 2S circuit. The PLS containing around 40 g/l Cu, 20 g/l acid and 20 g/l chloride is pumped to two extract stages in series. The loaded organic from E1 is washed with acidified water in order to reduce impurity transfer (particularly chloride) then stripped in two stages with spent electrolyte returning from EW.

Due to the high copper content of the PLS, a higher than ‘typical’ extractant concentration of approximatley 30–35 volume % (in Shellsol 2046 diluent) is employed in conjunction with a high O/A ratio of 4:1. This provides a nominal Cu recovery of approximately 70%. While this may be considered low by industry standards, the high PLS acidity makes further copper recovery very challenging without making significant circuit modifications or without neutralization.

Copper electrowinning (EW)

The Cu from SX is recovered by EW process employing the ISA process using 30 cells each containing 40 stainless steel or (later) titanium cathode plates and 41 anodes of 6% antimonial lead. More recently a Pb/Ca/Sn composition cast anode has been trialled. FC-1100 is added to control acid mist.

Residue treatment

The repульp thickener underflow contains recoverable lead and silver. To render the thickener underflow solids suitable for recirculation through the smelter, it is necessary that all of the free acid and most of the Cu process solution associated with it be removed. It is also desirable that prior to discharging the solution associated with the thickener underflow most of the dissolved metals be precipitated out. Neutralization of repульp thickener underflow is conducted continuously by controlled addition of milk of lime and the neutralized slurry is filtered on an Andritz belt filter. The residue discharges from the filter at 60–70% solids and is recycled to the lead plant. The metal free filtrate collects in a filtrate tank and is returned to the Andritz filter and is returned to on-site waste collection.

Summary of process difficulties

The use of the highly effective chloride/sulphate leach process produced a warm pregnant leach solution (PLS) high in acid, chloride and copper content, which resulted in several challenges for the operation of the Cu SX-EW plant requiring, for example, effective solids liquid separation and soluble silica control; high concentrations of SX extractant and high organic/aqueous (O/A) ratios in extraction; and careful monitoring of organic health with regular, effective, organic treatment.
Following commissioning of the SX circuit in 1984, the process experienced a gradual deterioration in the quality of the organic phase, so causing an increase in the transfer of chloride to the EW circuit. As chloride levels in the EW electrolyte surpassed 100 ppm, the original 316L cathodes were replaced by titanium cathode mother plates. In 1990 the original Acorga P5100 extractant was replaced by Acorga M5640. Over the following years, the plant struggled with low nett copper transfer rates and high EW chloride levels in the range of 200–500 ppm, which resulted in high chlorine gas concentrations in the surrounding environment causing corrosive damage to the EW infrastructure.

In August 2003 a sample of Port Pirie plant organic was analysed by Cognis in Tucson to determine reasons for the poor performance by analysing the composition of the circuit organic. The main findings were as follows:

- The organic phase was dark orange in colour. It was extremely viscous and exhibited very poor organic continuous phase separation behaviour, >15 minutes at 25°C. Aqueous continuous phase separation time was reasonable.
- 16.2% of the total oxime in the circuit was present as the nitro-C9 aldoxime—a very high level—causing the organic phase to exhibit a very high viscosity (and a dark orange colour).
- The TXIB:C9 aldoxime ratio was significantly higher (0.38) in the plant organic compared to that in Acorga M5640 (0.28) as a result of the differential degradation rate of C9 aldoxime compared to the TXIB modifier. This had caused the viscosity of the plant organic to increase with time and assisted in promoting nitration.
- Despite the reoximation process, the concentration of C9 aldehyde in the circuit was quite high (2.3%) compared to other circuits elsewhere. The same conditions that favour nitration also promote hydrolysis of the aldoxime to the aldehyde (warm temperature and high acid concentrations in the aqueous phase). Given the operating conditions at Port Pirie higher hydrolysis levels than conventional circuits may be expected, but the measured levels seemed excessive, especially as Port Pirie practised continuous re-oximation of the plant organic.
- The organic also contained significant amounts of the 2-cyano-4-nonylphenol and 5-nonylsalicylamide, which is unusual. They are typically the result of thermal degradation of the C9 aldoxime at temperatures higher than those said to be experienced in the Port Pirie operation. They were estimated to be present at about 10–15% of the level of the C9 aldoxime, thus making a significant contribution to the viscosity of the organic phase.
- Copper strip kinetics were surprisingly slow, 62% @ 30 seconds. Typical circuit organics would be expected to be in the range of 95% @ 30 seconds.
- The overall net transfer capability of the organic was low. It did not strip well due to the presence of the nitro-C9 aldoxime, which is an extremely strong copper extractant. This resulted in high stripped organic values. The net Cu transfer was further exacerbated by the low PLS pH, which in conjunction with a highly modified circuit organic provided a very ‘shallow’ equilibrium extraction isotherm.
- The organic did not respond well to clay treatment probably due to the presence of unfilterable entrained aqueous solution which deactivates clay. The entrained aqueous may have existed as micro-emulsions formed due to the presence of the degradation products or contaminants.

**Reagent selection and Isocalc® TM modelling—proposal to optimize the plant operation**

Cognis has experience with the effects of nitration at other Cu SX plants and has conducted laboratory degradation studies using different extractant compositions to determine which extractants were more resistant to nitration. 

SOLVENT EXTRACTION (SX) REAGENT SELECTION
Clearly, the modified aldoximes, LIX®622N and Acorga®M5640, and the modified ketoxime, LIX®84-I plus TDA, were much more sensitive to degradation/nitration than the non-modified reagents, LIX®984N and LIX®84-I.

Thus, Cognis recommended that (a) Port Pirie cease addition of the modified reagent Acorga M5640 and employ the more stable nonmodified reagent LIX® 984, and (b) Port Pirie cease re-oximation. Cognis have much experience in re-oximation, and attempts to re-oximate heavily degraded Cu circuit organics typically result in an organic that demonstrates very poor physical performance.

LIX 984 was selected for several reasons:

• LIX 984 is a 50/50 blend of ketoxime and C12 aldoxime, both of which are more stable than the C9 aldoxime. LIX 984 does not contain a modifier
• Acorga M5640 contains both a C9 aldoxime and modifier, TXIB
• TXIB (and all modifiers in general) promote nitration where nitrates are present in the aqueous solution
• LIX 984 showed good performance in the screening tests in Tucson.

In order to support the reagent conversion, a technical service package was negotiated. Port Pirie accepted the Cognis offer and LIX® reagent addition began in May 2004.

Pre-conversion plant survey

In April 2004 Cognis visited the Port Pirie smelter to conduct a pre LIX reagent addition plant survey and to collect a second plant organic sample to provide a baseline for the conversion. Analysis of the plant organic revealed that the Acorga M5640 concentration was approximately 43 vol% (24.3 g/l Cu). However, the concentration of Cu in the E1 organic was only 9.05 g/l, which calculates to a copper loading of only 37%. The Cu net transfer was very poor, at approximately 0.14 g/l Cu per % oxime due in part to the presence of the nitro-C9 aldoxime and the weak reagent blend (due to excess TXIB modifier). All mixers were operated aqueous continuous due to the extremely slow phase separation under organic continuous mixing.

The phase disengagement times (PDTs) of the plant loaded organic with both the plant PLS and a synthetic PLS were measured in the laboratory under both aqueous and organic continuous conditions at room temperature (23°C). For comparative purposes, fresh samples of LIX984 and Acorga M5640 were also tested. The results are provided in Table I.

Figure 1. Effects of modifiers on the stability of oximes in the presence of nitrate
Under aqueous continuous mixing conditions, the PDT was measured at similar rates to that as experienced in the operating plant, whereas under organic continuous conditions the emulsion demonstrated very poor phase separation behaviour taking several minutes for the emulsion to indicate any signs that the phases would separate; and after 24 hours the aqueous phase contained a considerable amount of ‘rag’ and an indefinable interface. This indicated high levels of organic contaminants were interfering with the coalescence of aqueous droplets within the organic continuous emulsion. Clay treatment—even at high dosage (10 wt% clay, typically 2–5 wt% is sufficient for heavily contaminated organics)—failed to improve the PDT.

Viscosity and specific gravity (SG) measurements were conducted on the plant loaded organic and compared to fresh samples of LIX 984 and Acorga M5640 at 43 vol%, with all sample temperatures equilibrated in the laboratory (23°C). The resultant viscosities are listed in Table II.

The viscosity and SG of the plant organic were much higher than that for fresh LIX 984 and fresh Acorga M5640. Both the high viscosity and SG resulted from an accumulation of degradation products that no longer participated in the extraction and stripping chemistry but which had a profound impact on phase separation, aqueous entrainment and plant performance.

The physical appearance of the plant organic was unusual in that it had a distinct dark orange/red coloration when in a thin film, and it tended to coat all surfaces with which it came in contact. This orange/red colouration is consistent with the presence of nitration products. It had a very syrup-like consistency and a distinct smell. When filtering plant organic, the samples required extensive time to filter through Whatman 1PS filter paper and there was a reasonable amount of aqueous remaining in the filter paper. This would tend to indicate that the aqueous formed a very fine micro-emulsion in the organic phase that was not readily released. After the initial plant survey it was concluded that the plant organic was in extremely poor shape.

Reagent changeover period and resultant plant operation
Following the addition of LIX 984 to the circuit in April 2004, plant monitoring and organic sampling were conducted on a regular basis to track the chemical composition as well as the

<table>
<thead>
<tr>
<th>Table I</th>
<th>Laboratory PDT results pre-LIX reagent addition</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>PLS/continuity</td>
</tr>
<tr>
<td>Plant O/C</td>
<td>Did not separate</td>
</tr>
<tr>
<td>Plant A/C</td>
<td>126</td>
</tr>
<tr>
<td>Synthetic O/C</td>
<td>Did not separate</td>
</tr>
<tr>
<td>Synthetic A/C</td>
<td>71</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table II</th>
<th>Organic viscosity results pre-LIX reagent addition</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Sample</td>
</tr>
<tr>
<td>SG (@ 43 vol%)</td>
<td>0.93</td>
</tr>
<tr>
<td>Viscosity (cSt)</td>
<td>26.48</td>
</tr>
</tbody>
</table>
The cessation of re-oximation and the conversion to LIX reagent has gradually improved the ‘health’ of the plant organic. This is evident by the lower concentrations of Acorga M5640 (C9 aldoxime) and TXIB with time. The TXIB and M5640 concentrations had decreased from 13.12% and 34.9% in August 2003 to 0.32% and 0.2% in September 2008 respectively, indicating a significant change from modified to unmodified plant organic.

The level of C9 nitro aldoxime decreased at a lesser rate than the Acorga M5640 and TXIB. This is consistent with the level of residual copper on the stripped organic. Nitration appeared to occur on an infrequent basis, suggesting that Port Pirie was getting nitrate into the system via ‘specific events’ throughout this period.

The chemical performance of the Port Pirie Cu SX plant was monitored regularly and compared with the health of the plant solvent. Plant surveys and chemical analysis were used to monitor the important parameters such as Cu recovery, reagent concentration, E1 organic loading as % of max load and net Cu transfer. Survey data are listed in Table IV. The reagent net transfer has improved as the LIX reagent conversion has progressed due to the reduction in the concentration of nitro oxime species present in the plant organic and an increase in the transfer capacity. The E1 organic loading has also increased from ~60% to above 80%. (The low Oct 07 levels were due to a dip in PLS Cu.)

Although organic loading is highly dependent on the Cu and acid concentration in the PLS, by reducing the TXIB content in the organic the E1 organic loading was increased, both of which reduce the potential for crud formation.

The extraction and stripping kinetics of the Port Pirie plant organic have improved significantly since the initial addition of the LIX reagent, as shown in Table V. Surface active impurities crowd and compete with the oximes at the interface, adversely affecting extraction and stripping kinetics. The reduction in the concentration of these surface active impurities in the plant organic has improved the extraction and stripping kinetics.

### Table III

Reagent concentrations in the Port Pirie plant organic post-LIX reagent addition

<table>
<thead>
<tr>
<th>Vol% or Wt/Vol%*</th>
<th>LIX 84-I</th>
<th>M5640</th>
<th>LIX 860-I</th>
<th>Nitro Oxime*</th>
<th>TXIB*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Aug 03 (pre-LIX)</td>
<td>-</td>
<td>34.9</td>
<td>-</td>
<td>3.1</td>
<td>13.12</td>
</tr>
<tr>
<td>Jan 05</td>
<td>7.6</td>
<td>10.9</td>
<td>9.8</td>
<td>3.0</td>
<td>7.78</td>
</tr>
<tr>
<td>Jul 05</td>
<td>11.0</td>
<td>6.5</td>
<td>11.9</td>
<td>2.8</td>
<td>5.76</td>
</tr>
<tr>
<td>Oct 05</td>
<td>12.3</td>
<td>4.3</td>
<td>12.4</td>
<td>2.2</td>
<td>3.88</td>
</tr>
<tr>
<td>Mar 06</td>
<td>14.0</td>
<td>3.1</td>
<td>12.4</td>
<td>2.1</td>
<td>1.96</td>
</tr>
<tr>
<td>Jun 06</td>
<td>12.8</td>
<td>2.6</td>
<td>12.9</td>
<td>2.0</td>
<td>1.64</td>
</tr>
<tr>
<td>Dec 06</td>
<td>8.2</td>
<td>1.8</td>
<td>16.4</td>
<td>2.4</td>
<td>1.75</td>
</tr>
<tr>
<td>Feb 07</td>
<td>7.0</td>
<td>1.6</td>
<td>18.1</td>
<td>1.8</td>
<td>1.27</td>
</tr>
<tr>
<td>Sep 08</td>
<td>6.9</td>
<td>0.2</td>
<td>20.5</td>
<td>1.2</td>
<td>0.32</td>
</tr>
</tbody>
</table>

Note. ^ Concentration is calculated as an M5640 equivalent since the majority of the nitrated material is nitro C9 aldoxime.

The LIX reagent addition blend changed from LIX 984 to LIX 973 in April 2006.
Clay treatment on the Sept 2008 sample increased kinetics to 99.6% and 98.9% respectively. The plant physical performance was monitored by measuring operating parameters such as emulsion temperature, mixer continuity, mixer O/A ratio, emulsion PDT under both AC and OC mixing (before and after clay treatment), and organic viscosity. Laboratory tests were conducted at room temperature ($22 \pm 2^\circ C$).

Plant physical parameters such as temperature, mixer continuity and mixer O/A ratio remained relatively constant for all sampling dates. The temperature was $40 \pm 2^\circ C$, the mixer continuity was always aqueous continuous, and the O/A ratio in the mixers was in the range 1.0 to 1.3. The plant PDT trends were consistent with that for the laboratory PDT. (Table VI)
As the LIX reagent conversion progressed, the PDT of the Port Pirie plant organic decreased significantly. This reduction in PDT corresponded with the reduction in degradation products and impurities in the plant organic. The Port Pirie Cu SX plant is now at a stage where it can convert all mixers to OC, which is the preferred operation for Cu SX, especially operations that have high levels of silica and solids present in the PLS.

Initial clay treatment tests conducted in August 2003 proved unsuccessful, possibly due to the presence of micro-emulsions in the plant organic, as witnessed previously with nitrated oximes. Later clay treatment proved successful as the levels of nitrated oxime and other impurities declined, as shown in Table VII.

After 2006, the clay treatment process became a part of normal operation at Port Pirie, which also aided in improving the physical performance of the plant organic. As the LIX reagent conversion progressed, the concentration of degradation products and contaminants decreased, resulting in a subsequent reduction in plant organic viscosity, as shown in Table VIII.

The chloride concentration in the electrolyte is compared against the % LIX reagent in the organic below (Figure 2).

Table VII
Clay treatment results

<table>
<thead>
<tr>
<th>Date</th>
<th>Extraction (OC)</th>
<th>Strip (AC)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Pre-CT</td>
<td>Post-CT</td>
</tr>
<tr>
<td>April 04</td>
<td>No break</td>
<td>440</td>
</tr>
<tr>
<td>March 06</td>
<td>920 sec</td>
<td>22</td>
</tr>
<tr>
<td>June 06</td>
<td>627 sec</td>
<td>22</td>
</tr>
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Table VIII
Viscosity of Port Pirie plant organic in cSt at 25°C

<table>
<thead>
<tr>
<th>Date</th>
<th>Viscosity (cSt)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Apr 04</td>
<td>26.48</td>
</tr>
<tr>
<td>Sep 04</td>
<td>19.43</td>
</tr>
<tr>
<td>Jan 05</td>
<td>15.06</td>
</tr>
<tr>
<td>Jul 05</td>
<td>15.13</td>
</tr>
<tr>
<td>Oct 05</td>
<td>13.90</td>
</tr>
<tr>
<td>Mar 06</td>
<td>13.15</td>
</tr>
<tr>
<td>Jun 06</td>
<td>12.20</td>
</tr>
<tr>
<td>Dec 06</td>
<td>9.50</td>
</tr>
<tr>
<td>Feb 07</td>
<td>8.93</td>
</tr>
<tr>
<td>Sep 08</td>
<td>6.66</td>
</tr>
</tbody>
</table>

Figure 2. Chloride concentration in the electrolyte
The chloride concentration in the PLS remained consistent throughout the LIX conversion period in the range 15–25 gpl.

It is evident that as the LIX reagent conversion progressed, the chloride concentration in the electrolyte has decreased significantly from unacceptable levels of ~200 ppm, to approximately 100 ppm with less than 50% LIX reagent in the circuit, to less than 30 ppm when operating with more than 90% LIX reagent in the circuit. This reduction in electrolyte chloride corresponded exceptionally well with the reduction in nitration products, TXIB and other impurities in the plant organic.

Conclusions

The conversion of the Port Pirie plant organic from a modified C9 aldoxime to a mainly unmodified blend of C12 aldoxime and ketoxime, plus the decommissioning of the re-oximation stage, has directly resulted in significant improvements in the health of the plant organic. Since the modified Acorga reagent M5640 was replaced by the non-modified LIX reagent the health of the SX organic and the performance of the circuit have both significantly improved:

- The strip point has decreased as the level of nitrated oxime has declined
- Extraction and strip kinetics have improved
- Net transfer has increased to greater than 0.2 gpl Cu/vol%
- Reagent strength has decreased to less than 35 vol%
- Wash stage acid has been reduced from 40 gpl to 15 gpl, resulting in savings on neutralization and reduced soluble Cu losses
- Organic phase viscosity has declined to a quarter of the viscosity of the pre-LIX period
- Organic continuous phase separation time has very significantly improved
- Clay treatment is now used to maintain organic health
- And most importantly, aqueous entrainments through to EW have significantly declined, so reducing EW chloride from more than 200 ppm to less than 30 ppm, and improving the EW atmosphere and environment.

References

Phil Crane

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Phil graduated from Birmingham University with BSc Hons in Minerals Engineering in 1975; with final year studies on copper solvent extraction. He worked in metallurgical plants and in research and development in Zambia and Iran; then in the design and supply of solids-liquid separation and flotation process equipment in both England and later South Africa. Moved to Cognis, then Henkel in 1988; working in the PGM flotation field and supporting Henkel’s copper and uranium solvent extraction business activities in southern Africa. He then transferred to Australia in 1996 and from 2002 Phil has been responsible for Cognis Mining Chemical Technology’s business and technical support in the Asia-Pacific region, specifically in business development and customer’s plant troubleshooting.