DESIGN OF COPPER-COBALT HYDROMETALLURGICAL CIRCUITS

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1. ABSTRACT

Copper-Cobalt hydrometallurgy has seen a resurgence in recent years with the development of projects in Zambia and the DRC. Many of the flow sheets are unique in their use of hydrometallurgical techniques to improve performance compared to the older styles in current use.

Many of the cobalt purification by precipitation steps have been replaced with solvent extraction; while the remaining precipitation stages have been enhanced with use of oxidants to improve impurity removal. Application of recent developments in precipitation circuit designs has improved solid-liquid separation performance by better crystal growth. Operating experience from two of these circuits has shown the benefits of the new technologies and the improved overall recovery of both copper and cobalt into higher quality products.

The emphasis has moved from direct copper electrowinning / stripping with multiple stages of precipitation for removal of impurities, and concentration of cobalt; to innovative use of solvent extraction and ion exchange with higher cobalt recoveries to final product. The poor quality direct electro-won copper has been supplanted with solvent extraction-electrowinning; with consequent production of LME “A” quality metal.
2. INTRODUCTION

Copper-Cobalt hydrometallurgy has seen resurgence in recent years with the development of projects in Zambia and the Democratic Republic of Congo DRC (formerly Zaire). Many of the flow sheets are unique in their use of hydrometallurgical techniques to improve performance, compared to the older styles in current use. The emphasis has moved from direct copper electrowinning / stripping followed by multiple stages of precipitation for removal of impurities and concentration of cobalt; to innovative use of solvent extraction and ion exchange with higher cobalt recoveries to final product.

Cobalt is recovered as a by/co-product of copper production and is generally treated as a commercial bonus in the metallurgical copper circuits. As a result it has not been until recent newer projects have come into the development programme, that a significant focus on cobalt hydrometallurgy has become more important. For some projects in the DRC the potential cobalt income is now of the same order of magnitude as the copper income (Anon, 2008). Maximising the recovery and value addition to the product is a significant driver for hydromet process development.

It is impossible to divorce the selection of the cobalt recovery flowsheet from the specific markets to be targeted. However there are a number of common unit operations; that are used to remove specific groups of ions; that have all been advanced in the most recent round of process designs. Many projects have taken increased development times in order to pilot the cobalt recovery part of the process plant. The commonly used unit operations and their relative place in the process chain are discussed in the body of this paper.
3. COBALT PRODUCT SELECTION

The selection of the cobalt product to be produced on site is vitally important in the development of the required flow sheet. This motherhood statement is often overlooked by the project owners who assume that the cobalt market is similar to other metal commodities. It is however an extremely fragmented market, with many different product types, purity requirements and market demands. One of the more important drivers for recent projects of modest size has been the availability of excess cobalt metal refining capacity in Europe and China. Both of these markets will take an impure cobalt intermediate salt and refine to final metal. The available capacity is ultimately limited, and new large projects will need to address the market directly with finished (or semi finished) products for direct input to customer processes. The diversification of the BHP Billiton Yabulu refinery into cobalt chemicals is a good example of this change in product emphasis.

The potential hydrometallurgical cobalt products are extremely wide ranging. However it is the major current markets that most green fields project are targeting. The more significant cobalt products include:

- Cobalt Salts
  - Hydroxide from either lime, magnesia or NaOH precipitation
  - Carbonate usually using soda ash for precipitation

- Value Added products
  - CoO from calcination of either hydroxide or carbonate
  - Sulphate from purified leach liquors and evaporative crystallisation

- Cobalt metal has many grades of product depending on chemical purity and physical condition. Some potential metal producers are considering the full range of post-production metal enhancement with:
  - Hydrogen de-gassing furnaces
  - Size control
  - Surface burnishing
  - Specific packaging for market.

It is evident that there is no single target product similar to LME “A” grade copper. As a result there are many possible cobalt flow sheets and processes to generate a marketable product, that is optimum of the specific operation. There are however a number of elements that are specifically targeted to improve the realised value of the cobalt. The major ones include:

- Copper
- Iron
- Calcium
- Zinc
- Magnesium
- Manganese
Specific products require the removal of these ions to a greater or lesser extent. Particularly with intermediate salts (for further refining) the level of penalty metals can be in the order of 2% to 4% without too high a reduction in income. On the other hand production of super high quality electrowon metal needs the preparation of high purity solutions with controlled maximum amounts of all the ions mentioned.
4. COBALT LEACHING

Particularly in the DRC and to some extent in Zambia, the oxide ores with copper and cobalt are leached in atmospheric systems. The dominant mineralisation is malachite/azurite with accessory chrysocolla and minor other secondary copper minerals. The cobalt is present as heterogenite with cobalt in both the Co\(^{2+}\) and Co\(^{3+}\) oxidation state. The cobaltic minerals are about 50 per cent of the total cobalt; and are not direct acid leachable at normal temperatures and pressures. Alternate methods of enhancing the cobalt leaching are required.

The leach trains are designed to leach the copper oxides in mildly oxidizing conditions; to recover the minor cuprite, chalcocite and native copper. The leach time can be extended if there is significant chrysocolla present (Miller, 2005). Copper leaching is generally taken close to completion in four to eight hours. Those projects that do not have significant secondary copper minerals can achieve high leaching efficiencies in as little as two hours from the rapidly leaching malachite and azurite (Crease, 2006).

About 50 per cent of the cobalt is leached (along with the copper) mainly from the Co\(^{2+}\) minerals. The other cobaltic minerals need to be reduced to the 2+ state in order for them to leach. This is achieved with controlled reductive leaching. The main reductant used to date has been sodium meta bi-sulphite (SMBS: Na\(_2\)S\(_2\)O\(_5\)) (Mwema et al). In an acid solution the SMBS disassociates to form SO\(_{2\text{aq}}\) which lowers the Eh in solution and reduces the cobalt oxidation state. SMBS is costly and is only partially utilised with side reactions producing sulphuric acid – particularly in the presence of manganese ions in solution.

A more recent development has been the use of gas from a sulphur burning acid plant (SO\(_2\) and N\(_2\)) as a direct Eh management tool. The gas is injected into the later parts of the leach train to enhance the cobalt leaching. The SO\(_2\) is readily soluble and reduces the cobalt while undergoing conversion to sulphate with acid production. The large volumes of nitrogen (about 88% v/v) can be an issue with scrubbing of the SO\(_2\) from the solution, if the gas is not dispersed properly. One operation in Zambia has used a Pressure Gas Disperser (PGD) in order to create ultra fine gas dispersion using a high velocity pressurised jet of solution to entrain the gas (John, 2006). Another alternative is to disperse the gas into a recycle stream of the leach slurry via a venturi eductor/mixer. Other newer projects are considering the use of liquefied SO\(_2\) to remove the voluminous nitrogen (Grosse, 2007). This eliminates the nitrogen scrubbing and makes the gas dispersion much more controlled.
Venturi Eductor/Mixer for Gas Dispersion into Leach Slurry

Two projects in Zambia, at Chambishi and Nkana, calcine a copper/cobalt/pyrite concentrate in a sulphation roast, using fluid bed contactors (Sole et al., 2005). The copper and cobalt are in the oxide/sulphate form and are leached (along with a lot of the iron) with dilute sulphuric acid. Since the cobalt has been reduced in the sulphation roast to the Co$^{2+}$ form it is readily soluble in the acid solution. Both projects use recycled solution ‘barren’ in copper. This ‘preg builds’ the cobalt solution levels to +10 g/L for subsequent recovery.

Other projects also undertake ‘preg building’ to increase the cobalt concentration in solution. Most often this is accomplished by recycling a majority of the copper SX raffinate back to the leach. A smaller bleed stream is treated for further copper removal and recovery of the cobalt. The technique has some issues with the control of cobalt soluble loss in the solid/liquid separation steps. Most the oxide ores are very weathered and have poor filtration and thickening characteristics. As a result all newer projects have opted for CCD trains. The wash liquor needs to be fresh water or recycled cobalt plant effluent, low in cobalt, to enable the recovery of the dissolved cobalt along with the copper. These CCD trains occupy a significant proportion of the project footprint and contribute a large proportion of the project capital costs.
5. COPPER RECOVERY AND REMOVAL

Primary copper removal

The primary metal production from most projects is copper. The existing Zambian and DRC operations used direct electro-winning and electro-stripping to produce impure metal; that generally had to be re-refined to produce a saleable product. The electrowin and electro-strip process is also inefficient with current efficiencies as low as 65%. Both Zambian cobalt projects have committed to changing the copper recovery to solvent extraction and electrowinning (Miller and Nisbett 2005; Kordosky, 2008; Sole et al, 2005). The Nkana project has already been in operation for some time with excellent results (Mwakila, 2008). The copper is LME“A” quality and the plant copper production has increased seventy five per cent. The production increase is directly linked to the ability to add more copper into the circuit with the removal by SX. The direct EW tankhouse has been reconfigured from starter sheets to stainless steel cathodes; and the capacity increased from 14, 000 tpa to 30, 000 tpa in a reduced footprint (Kordosky, 2008).

![Rebuilt Nkana 30 000 tpa EW](image)

Secondary copper removal

The new generation of projects has gone directly for primary copper recovery from solution by solvent extraction and subsequent electrowinning on to stainless steel cathodes. In conventional SX plants this will recover up to 98% of the copper – still leaving 0.10 g/L to 0.20 g/L copper in the final raffinate; that progresses to the cobalt recovery. This level of copper is an issue with cobalt product quality and needs to be reduced to less than 1 ppm (Kongolo et al, 2005).

Preparation of copper SX plant ‘raffinate’ for downstream cobalt production needs the copper to be reduced to less than 1 ppm for metal electrowinning. In the past this was achieved by
co-precipitation of the iron and any residual copper at pH +4.5; and disposal of the gypsum/iron/copper solids. The copper was a direct loss from the process.

Traditional single stage copper SX will produce the raffinate described above. This represents a significant loss of copper that could be recovered into the primary product. Cost benefit analyses on three projects to date have shown that a unique SX Split-Circuit™ can be justified (Miller and Nisbett, 2006). This circuit (shown in Figure 1.0) takes the cobalt bleed steam and removes the copper to consistently less than 10 ppm. The copper is recovered as EW LME “A”. The key to the SX plant performance is to integrate it with the precipitation removal of the other ions particularly iron.

![Figure 1: Copper-Cobalt Split-Circuit™](image)

The iron is precipitated at pH 3.2 with only marginal loss of copper. The iron free high pH solution is contacted with standard copper SX extractant (in a second SX section – SX2) to remove the copper quantitatively. The copper in the SX2 raffinate is thought to come mainly from the entrainment of EW electrolyte in the stripped organic. The copper ‘free’ solution is sent on to further purification. Should the final route be to metal, the solution quality of < 1 ppm copper can be achieved with a much smaller IX plant or less reliance placed on purification via cobalt SX.
Kabwe Integrated Split-Circuit™ SX Plant – Large Units SX1, Small Units SX2.
Using MMS Side-Feed™ Mixer-Settlers.
6. SOLUTION PURIFICATION

Iron Removal

Iron removal has been undertaken for many years in many hydrometallurgical process plants. The classic method is air oxidation to ferric and precipitation with lime and or limestone. All the current and previous Zambian and DRC project use this basic method. However the older style plants all suffer from the usual problems of:

- downstream gypsum precipitation and
- fouling of process equipment and pipes.

Many of the new projects are using more recent process developments to minimise the calcium over-saturation and subsequent gypsum precipitation issues. The major focus has been on:

- Use of air/SO$_2$ as an enhanced oxidant for the iron Multiple stage addition of precipitant (Papangelakis, 2004; Demopoulos, 2004)
- Higher temperature operation to enhance kinetics of crystal growth and kinetics (ibid)
- Recycle of seed crystals (ibid)
- Use of High Density Sludge HDS™ style techniques to improve precipitant utilisation and further reduce calcium over-saturation (HGE, 2007).

The air/SO$_2$ system has been the subject of many papers in the recent past (Krause E, 2007; Ho and Ring 2007; Ferron CJ and Turner D, 1999). The system is reasonably robust and can reduce iron to < 5 ppm with <1 ppm a common result. The major engineering issues with the system are:

- The very high gas volumes, when using acid plant feed gas,
- The change from reaction rate limited to gas dispersion limited in vessels over ten cubic meters volume (Van Royeen, Archer and Fox, 2007)
- The competing kinetics of gas solubility reduction and crystal growth kinetics with increasing temperature.

Optimum temperature is quite solution chemistry dependent but generally falls in the range of 45$^\circ$C to 50$^\circ$C (Krause, 2007). The effect of ions in solution is quite marked with reduction in Fe oxidation rates of between 10% to 25% with H$^+$, Cu$^{2+}$, Co$^{2+}$, Cl$^-$ (Krause, 2007). As these ions are in most solutions the rate of Fe oxidation needs to be confirmed from laboratory test work on ‘real’ process solution.

The gas flow issues, in the precipitation tanks, are quite important as they are sufficiently high to cause both SO$_2$ stripping from the solution and to flood impellers – even with appropriately designed ones. The key control criterion is the SO$_2$/O$_2$ ratio that can seldom be greater than 1:5. These factors are drivers for eliminating significant gas volume by using liquified SO$_2$ from the acid plant gas. The volume reduction will to a large degree overcome the solubility issues and subsequent OH&S issues with SO$_2$ in the ambient air. At least two new projects are using this technique for enhanced iron removal.

Multiple stage reagent addition is just good chemical engineering to control the over concentration of calcium. Papangelakis (2004) has shown the benefits of the method in the control of calcium concentration in the final solution. He has also recommended that a final
stage of solution maturation without reagent addition be used to further minimise the calcium saturation.

Higher temperature operation has also been adopted in a number of projects where the benefits were required. The first of these was in Zambia where direct steam injection was used to raise the temperature. Both Zambian roast-leach projects have enhanced PLS temperature (as a result of leaching hot calcine) and their subsequent iron removal is generally good without the use of oxidants other than air.

Crystal seed recycle has been practiced for many years to allow growth of larger crystals and minimisation of downstream gypsum fouling. The operations at the three Australian Nickel plants showed how important this unit operation is to achieving high plant utilisation. None of the plants utilised all the techniques mentioned here, in their original design. Plant shut downs of one to two days every three to four weeks were common; until better control of the iron precipitation step was achieved. Seed recycle is an integral part of this control and has been included in all new projects.

The HDS™ technique (HGE, 2007) and other similar patented processes (RAMS, 2007) were developed primarily to provide larger crystals from dilute water treatment solutions. The key process is to have large seed recycles (up to 2000% from dilute solutions) combined with a modified reagent addition regime. The first reactor receives the recycled crystals and the precipitant. Here the precipitant is adsorbed on to the crystal seeds. In the second reactor(s) the process solution is added and reacts directly on the particle surface growing there; in preference to fresh nucleation in the solution. This technique claims (Gabb et al, 1995) to reduce downstream calcium precipitation and to improve solid-liquid separation rates. Many variations on this theme are available as technology packages. Most need to develop specific ‘recipes’ to address the solution chemistry to be used. However almost all the new projects have incorporated very flexible systems of tankage, reagent addition, seed recycle and oxidation intensity; to allow them to adjust their plant to suit changes in the actual chemistry presented.

**Manganese Removal**

Manganese removal is done mainly to achieve levels that are suitable for the product specification. It is accomplished at a higher pH than iron and has greater potential to co-precipitate cobalt. As a result it is often done as a separate step to iron and the resultant solids recycled back to the leach to recover some of the precipitated cobalt. This is similar to the technique used at Bulong for recovery of co-precipitated nickel (O’Callaghan, 2003). The process used to date in Zambia and DRC has been a combined iron and manganese precipitation. This has removed the copper and some zinc; but also co-precipitated significant cobalt which has been lost. Enhanced methods of manganese removal have been developed to minimise this loss.

The amount of manganese that can be precipitated depends on the level of Mn$^{4+}$ present. But due to the reductive leaching for cobalt this is generally a small portion of the total manganese. The air/SO$_2$ oxidation system has been investigated (Zhang W, Singh P and Muir D, 2001; Ferron CJ and Turner D, 1999; Wythe J and Vegter NM, Lunt et al, 2001; Schulze-Messing J, 2006) and included in a number of circuits to lower the manganese concentrations. The level of reduction needs to be determined for the specific product. In many cases removal to 1 g/L is sufficient if subsequent operation is via Co SX-EW or for
production of an impure intermediate product. Some manganese is acceptable in Co EW electrolyte provided that anode maintenance is regular and the deposited MnO\(_2\) is removed.

If manganese removal is required to lower levels, then it has been found that the rate of reduction is enhanced if a two stage process is used (Van Royen J, Archer S and Fox M, 2007). A primary stage to reach 1 g/L; followed by a solid-liquid separation step. The liquor is then treated in a second oxidative step to remove manganese to ppm levels. The kinetics are slow and long residence times are required. The same issues with gypsum management and gas volumes need to be addressed in the design of this circuit.

Bulong used an alternate method by precipitating the cobalt as a sulphide and leaving the manganese in solution. The sulphide was re-leached in a small autoclave. This process was complex and involved more stages of treatment. As a result it was discontinued in favour of producing an intermediate sulphide product for sale.

**Zinc Removal**

At this point it is possible to precipitate a reasonable quality cobalt salt without further purification. However if zinc levels are elevated in solution they will be elevated in the product as the zinc will precipitate with the cobalt. Zinc removal in the traditional circuits has been via lime precipitation at elevated pH and recycle of the solids to the leach for cobalt recovery. This has proved to be marginal at best and a circulating load of zinc has built up. Both Chambishi and Nkana use a D2EHPA SX to extract the zinc prior to cobalt metal production Sole et al., 2005). The Nkana plant is a simple 1E+1S that takes out a portion of the zinc to stabilise the circulating concentration to a low enough level to be acceptable. Chambishi have a more sophisticated SX plant with multiple stages of extraction, scrubbing, stripping and washing. They also include an HCl regeneration stage where extracted iron is removed from the D2EHPA.

Other newer projects are also considering the benefits of Zn SX, ahead of the cobalt purification plant. Cyanex 272 has been proposed for this duty (Tinkler et al, 2007) with specific pH and scrubbing conditions to target zinc rather than cobalt. Both D2EHPA and C272 suffer from issues of calcium saturation in the strip solution, that must be addressed by large volumes of Zn stripping solution to limit the calcium to below saturation conditions in this waste stream. The solution can not readily be reused for CCD wash as the zinc content will build up with the circulating load. As a result further processing may be required to precipitate the zinc so that the water can be reused in the process.

A small D2HEPA Zn SX was in operation at Bulong, that removed this ion before cobalt EW. It suffered severely from reagent poisoning with ferric whenever the iron and sulphide precipitation plants were not operated well. No HCl regeneration was carried out.
7. COBALT CONCENTRATION

Solvent Extraction

At this stage the older Co EW plants would direct electrowin from the cobalt solution. Some control of nickel would be required, generally on a side stream to reduce the cost of the IX plant. Cobalt precipitation and re-leaching was required to obtain the required water balance, neutralise the acid from EW and to increase the cobalt concentration in the advance electrolyte. These precipitation steps also lead to cobalt losses – especially from the less than perfect re-leach step.

More recent projects are considering the production of high quality metal by EW of an SX electrolyte. The preferred reagent is Cyanex 272 or its analogues. The cobalt is removed from the PLS at elevated pH; and stripped in a very dilute acid electrolyte. The EW of cobalt in un-divide cells is well understood, and the maximum operating concentration of acid is around 8 g/L in the spent electrolyte. This corresponds to a cell and SX delta cobalt of around 5 g/L.

The engineering issues are around the selection of the pH control chemical – ammonia gas, ammonia solution, NaOH solution or others. The lack of an ammonia gas infrastructure in Africa is leading many operations to consider ammonia solution (Tati Nickel) or dilute NaOH (Sole et al 2005). In all cases there is an issue of residue disposal unless steps are taken to eliminate the common ions: ammonium or sodium from the effluent.

Precipitation

Cobalt precipitation is also undertaken at this stage for production of a saleable salt or to reject water. Precipitation has been conducted traditionally with lime slurry when re-leaching; or sodium carbonate for a saleable product. However a number of projects have considered the use of MgO instead of sodium carbonate. This is largely based on the costs of the reagents and the higher utilisation of MgO than Na₂CO₃.

African Style NaOH Addition System
The precipitation is also likely to be conducted at elevated temperatures of up to 80°C. This is used to enhance crystal growth rates and morphology; and to achieve extremely high recoveries from the enhanced kinetics. Crystal recycle is also an integral part of process to achieve coarser product size distribution that has enhanced dewatering characteristics.

**Other Ionic Controls**

Control of other ionic species is not generally practiced specifically unless one or another is introduced as part of the process reagent(s). One operation in Zambia could not use local limestone as it was running 4% acid soluble oxide zinc in the material. Likewise introduction of ammonia or sodium carbonate / hydroxide have issues with the disposal of tails having high available nitrogen or high sulphate (as the sodium salt).

Aluminium and silica are generally precipitated with the higher pH reactions. At a pH of 4.5 most of the aluminium and silica precipitate with the iron and/or manganese removal. This is non specific removal as both co-precipitate.

Use of MgO for precipitation of cobalt hydroxide introduces Mg into the solution. This can be removed by lime precipitation of magnesium hydroxide but at the cost of the lime to force the reaction. Disposal of the calcium saturated water is now possible once it is re-acidified to pH 6.0 to 8.0.
Many of the ancillary process steps are designed to minimise the effects of one process on the succeeding process(s). The significant number of plant designs that are using sequential solvent extraction processes, will all require the minimisation of reagent carry over. Bulong suffered from carry over of Cyanex 272 from the cobalt SX into the Versatic 10 nickel SX (O’Callaghan, 2003). Insufficient consideration of this interaction has lead to other operations also having difficult operations (Kasese, 2007). Newer plant designs are utilising some or all of the following process steps to reduce the intermixing of SX reagents:

- Settler designs with lower entrainment losses such as the MMS Side-Feed™ settler
- After-settlers and coalescors for removal of bulk entrained organic
- Use of a diluent scrub stage for recovery of Cyanex 272. This was combined with a saponification process at Bulong to recover the C272 in a concentrated stream (O’Callaghan, 2003).
- Dual media filtration to remove organics to < 5 ppm entrained
- Carbon adsorption to reduce organics to < 1 ppm total entrained and dissolved

Other process steps are designed to provide higher plant utilisation between shut downs for gypsum removal:

- Clean-in-place acid circulation systems for dissolving gypsum from heat exchangers and key process pumps
- ‘spare’ mixer-settler unit for on-line clean out – especially for zinc D2EHPA stripping stages.

There is generally a need to address the issue of silica in the primary copper SX plant PLS. The silica will be removed in the iron precipitation, but only after the primary copper extraction has been done. Plant operation in organic continuity may be required; and the control of entrainment in the loaded organic is necessary. Most new copper SX plants are either including loaded organic coalescing or making design provision to allow easy retrofitting of the coalescing system.
9. PLANT EXPERIENCE

It is important to look at the experience of existing plants to try to build on the knowledge base that exists. Many errors of engineering application can be avoided if the lessons can be included in the new plant designs. Most of the unit operations have been run for long periods at one level of sophistication or another.

**Australian Nickel Operations.**

The Australian HPAL nickel operations have all suffered to some degree from significant issues of gypsum precipitation; which has resulted in lower than design utilisation and subsequent cash flow issues. This is a prime lesson in making sure that all the points regarding process optimisation are used for the iron / manganese precipitation processes.

None of the operations practice copper recovery specifically; and copper is removed by precipitation as part of the iron removal process. Bulong had a final copper clean up prior to Co EW using IX resins.

**Cawse**

The Cawse plant used an intermediate precipitation of mixed Ni-Co hydroxide with re-leach in ammonia. Ni and Co were separated in ammonia solution using a LIX reagent. The cobalt was subsequently precipitated as a salt for sale. Other than the gypsum issues there are no specific items that can be taken to new cobalt production facilities.

**Bulong**

Bulong went the full route to cobalt EW metal production in the initial design and operation. This proved to be problematic due to the high reagent costs and large number of operators required to keep the section running. The long train of processes to provide an electrolyte for EW metal production was such that it created one of the bottle necks in the plant. It was subsequently modified to produce an intermediate sulphide salt. The main issues to take from this operation are the simplification of the process with an intermediate product, and the difficulty in operating the long train of processes continually.

As part of their in house development to improve plant utilisation Bulong undertook a six month commercial trial of an anti scaling reagent (O’Callaghan, 2003). This was successful in reducing the build up gypsum in the process vessels and pipes; and increased their run time from three weeks to over two months. Although expensive, the reagent was cost justified on the basis of the reduced clean out costs and improved plant utilisation and total production.

**African experiences**

**Confidential Client DRC**

This project is planning to make a high grade cobalt metal product for market. They currently operate a facility making an intermediate cobalt basic sulphate. As part of this process the cobalt is reduced with acid plant SO\(_2\) gas stream; and iron is precipitated with air and lime, using indirect steam heating. The operating plant has developed a number of techniques for minimising the downstream precipitation of gypsum.

A semi commercial pilot plant has been running for more than eighteen months producing 500 kg/day of high grade cobalt metal. This process uses the sequential zinc SX extraction with D2EHPA and cobalt concentration with a Cyanex 272 analogue. Nickel, magnesium and other metal rejection is high from a solution that is essentially iron free. Adjustment of pH in
the Co SX is with NaOH in dilute solution. This pilot plant is being used to develop design criteria for a full scale operation to produce 10 000 tpa cobalt metal.

*Tati*

LionOre (now Norilsk) have constructed and operated a semi commercial pilot plant for their Activox™ process at Tati Nickel in Botswana. The process involves:

- Initial copper removal by single stage SX-EW,
- Iron / copper / manganese / silica precipitation in two separate stages at different pH,
- Cobalt recovery with Cyanex 272 SX and
- Subsequent Nickel recovery with Versatic Acid 10 (the same circuit as used at Bulong).

Cobalt is precipitated as the carbonate for sale as an intermediate product. Many of the innovations in iron precipitation and gypsum management are included in the plant design. This combined with other proprietary designs has meant that the pilot plant has run without significant gypsum precipitation in the cobalt or nickel SX plants.

*Sable Zinc Kabwe*

The SZK operation has the first Split-Circuit™ copper SX to produce very low copper concentrations in the cobalt plant feed (Kordosky, 2008). It has also been the first to start up with SO$_2$ reductive leaching and air/SO$_2$ iron and manganese oxidative precipitation. To date the leaching has been successful while the precipitation is still undergoing parameter optimisation and operational adjustments. The use of seed recycles and an elevated temperature has meant that there is only slight gypsum scaling in the secondary copper SX plant.

*Nkana*

Nkana operations have improved with the conversion of the copper electrowin – electro-strip operation to a Split-Circuit™ copper SX and stainless steel cathode EW plant (Mwakila et al,
2008). The copper recovery has improved 5% and the cobalt recovery has also improved by a further +5%. The copper product is LME “A”. The increased cobalt recovery is attributed to the reduction in pH required for the iron precipitation stage and the subsequent gain in cobalt no longer co-precipitated with the iron. Further cobalt production has been possible by adding a bleed of raffinate from the newly commissioned Nkana leach plant as the make up water to the calcine leach. This has further increased cobalt production by around ten per cent.
10. CONCLUSIONS

The selection of an appropriate cobalt recovery process is a tortuous path threading between product specification, leach solution chemistry, cost/benefit analysis and operational complexity. There is no one process selection that is appropriate to all operations – even those with similar chemistry.

Unit operations for use in cobalt circuits have undergone major development since the last significant cobalt plant was constructed. As a result the operations of the next generation of cobalt plants are likely to be more stable with higher recoveries in the long term; but fraught with the usual un-foreseen issues in the period immediately after commissioning.

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