Introduction

The increased demand for copper and cobalt together with the increase in political stability in the region, have resulted in renewed interest in developing the copper and cobalt deposits in the Katanga Province of the Democratic Republic of the Congo (DRC). A number of projects, in various stages of development, are currently underway in the DRC. The primary product from most of these projects is copper; however, cobalt is often a significant and sometimes major contributor to the project revenue.

Cobalt in the form of heterogenite is commonly associated with copper bearing oxide ores and hence it is commonly co-extracted with copper. In order to extract cobalt with copper, a reducing environment is created for reduction of Co$^{3+}$ to Co$^{2+}$ as divalent cobalt is soluble at the pH values at which copper dissolution is conducted. Cobalt is then generally recovered from a copper solvent extraction (SX) raffinate bleed stream.

The general trend in the development of these copper/cobalt projects is to produce an intermediate cobalt product at start-up and progress to cobalt metal production once the operation has proven to be viable for a period. The decision not to go all the way to cobalt metal is generally due to the perceived risk associated with operating a complex cobalt refinery in central Africa. Shifting from production of intermediate cobalt products to pure metal is largely driven by increased revenues.

The decision to produce an intermediate cobalt product vs. metal significantly affects the processing routes employed subsequent to copper solvent extraction (SX). If an intermediate product is to be produced, then removal of iron, manganese and copper will occur with little if no emphasis placed on reducing the levels of other impurities. Production of cathode necessitates stringent impurity control in the cobalt electrolyte, which is critical in ensuring...
that cathode metal specifications are met. Typical impurities, other than those mentioned
above, are nickel, zinc and aluminum. The methods for removal of these are precipitation, SX
and/or ion exchange (IX), and the choice of technology depends on impurity concentration,
relative capital costs for removal and associated operational preferences.

This paper discusses some of the processing options available for cobalt production and
provides an evaluation of the associated costs, risks and benefits applicable to the DRC.

**Democratic Republic of the Congo**

**Introduction**

The Democratic Republic of the Congo was initially established as the Belgian Congo in
1908. It subsequently gained independence in 1960. The early years of independence were
marked by social and political turmoil.

In a coup in 1965, army colonel Joseph Mobutu seized power, renaming himself Mobutu
Sese Seko and the country Zaire. Under his presidency, social and economic conditions in the
country deteriorated further. In 1997, Mobuto was toppled by a rebel leader, Laurent Kabila.
Kabila restored the name of Congo in the form of the Democratic Republic of the Congo
(DRC). He in turn faced a rebel movement backed by Rwanda and Uganda. Troops from a
number of African allies were flown into the country and the rebellion was put down and a
cease fire signed in 1999. Kabila himself was assassinated in early 2001, and his son, Joseph
Kabila, was installed as President of the Republic.

Joseph Kabila subsequently negotiated the removal of Rwandan forces, and a peace
agreement was signed in South Africa in 2002. A transitional government was formed with
the objective of eventually holding democratic elections in the country. These elections were
held in July 2006, and resulted in the two leading presidential contenders each receiving less
than 50% of the vote, although Joseph Kabila did gain a significant majority over his rivals. In
the terms of the election rules, a run-off between the two preferred candidates took place in
November 2006 with Kabila gaining an outright majority.

**Economy**

Since 2001, when Joseph Kabila became president, the economic policy of the country has
aimed at stabilization. Since about that date, there is little doubt that the mining and
construction sectors are improving, albeit from a low base. Gross national income is US$120
per capita, but since 2002, the economy has shown signs of expansion with a GDP growth in
2002 of 3.5%, 2003: 5.7%, 2004: 6.8%, 2005: 6.5%. The latter two figures are provisional.

The independent central bank has bought inflation under control in recent years, again
indicating a normalizing of the economic and social situation within the country.

The corporate tax and investment codes are being revised to continue to liberalize the
domestic business environment and attract foreign investment, and an increasing number of
foreign companies are establishing themselves in the country.

**Gecamines**

Prior to the departure of the Belgians from the Congo in 1960, the copper industry in the
Katanga area was owned and operated by Union Miniere de Haut Katanga (UMHK), a
company dating from 1906. After independence, nationalization of the copper industry was
inevitable, and UMHK was eventually dispossessed of its ownership of the industry, and a
state owned company, initially Generale Congolaise des Minerals and later La Generale des
Carriers et des Mines (Gecamines) took over as owners and operators of all copper mines in Katanga. Gecamines initially prospered throughout the 1970s and 1980s, but since the late 1980s, the company endured through a series of increasing crises, until finally, by the late 1990s, metal production was only 35 000 t of copper and 3 940 t of cobalt, compared to 437 000 t of copper in 1987, plus 11 900 t of cobalt in the same year.

As a result of this poor performance, the Government began to seek ways of improving the situation both within Gecamines and within the copper industry as a whole. As a result of these initiatives, foreign companies have been encouraged to enter into agreements with Gecamines in an attempt to restore the industry. After some stumbling, these initiatives are now beginning to bear fruit.

**Code Minier**

A Code Minier has been introduced by the Government and provides a transparent legal framework for investors in mining. The code has been gazetted in French and English, making it easier for foreign Anglophone companies to review the legislation.

**Cobalt processing options**

In the DRC there are typically three major saleable product types produced, namely:

- A cobalt concentrate via gravity separation
- An intermediate cobalt salt
- Cobalt metal.

The value of the product increases as the level of refining increases; however, this is also proportional to the capital and operating cost associated with the facility required to produce the end product. The complexity of the operation also increases significantly from concentrate production through intermediate to pure metal.

**Introduction**

Gravity separation is by far the cheapest and lowest cost method of producing a saleable product from ore. Not all ore types are amenable to gravity separation and generally low recoveries are achieved even for those materials that are amenable. Product grade and recovery are ore specific and generally quite poor. Dense media separation is the most common technology used and is often employed as a preliminary measure to generate cash flow before proceeding to the production of a purer product.

This method is mentioned for completeness and will not be discussed further.

There are a number of different cobalt intermediate products that can be produced, namely:

- Cobalt hydroxide
- Cobalt carbonate
- Cobalt sulphide
- Cobalt sulphate.

In addition, there are a number of refined cobalt metal products that can be produced from a copper raffinate bleed stream or any of the above intermediates:

- Cobalt cathode by electrowinning (EW)
- Cobalt powder/briquettes by hydrogen reduction.

Cobalt refining is not only complex, but product types vary considerably in terms of purity (and hence market premium and transport costs) and by-product/effluent generation. It is important, therefore, to establish factors with the greatest importance to and impact on the project to be able to identify the optimum solution. Four main categories should be considered:
• Operationally safe, robust and simple
• Technically robust and proven technology
• Environmentally compliant
• Commercially acceptable.

Unfortunately there is no single technology, flowsheet and product that rank highly in all of the above categories. It is therefore necessary to select a flowsheet that has an ‘acceptable ranking’ in all categories while providing the best return to the project. A decision diagram (Figure 1) illustrates the options, ranked in environmental and operability terms.

Options producing high levels of soluble salt form the left-hand leg of the decision chart, and those processes that minimize the production of soluble salt effluent are indicated in the right-hand leg. Those processes on the right of the decision chart reduce soluble salt generation by an order of magnitude. It should be noted that none of the options provides a zero discharge scenario.

Process options with lower risk profiles are located at the top of the chart, whereas higher risk options such as sulphide precipitation with hydrogen sulphide gas are located in the bottom. For the purpose of this ranking, process risk includes operational safety, operability and robustness.

Intermediate product options

There are four main cobalt intermediate product forms that can be produced from either selective precipitation or evaporation/crystallization techniques:
• Cobalt hydroxide (precipitation using lime, magnesia or caustic)
• Cobalt carbonate (precipitation using sodium carbonate)
• Cobalt sulphide (precipitation using sodium hydrosulphide)
• Cobalt sulphate (crystallization of a cobalt sulphate solution).

For all of the above products the option exists to produce a crude or ‘pure’ form. A crude form is defined as the product obtained after limited impurity removal (only iron); a pure product is defined as the product obtained after removal of the major typical impurities, namely, iron, manganese, copper and aluminium.

The process description for impurity removal is similar for all four intermediate products and is summarized below.

**Impurity removal prior to intermediate product production**

**Fe/Mn/(Al) removal**

Iron and manganese (and aluminium in the presence of phosphate) are removed from the cobalt bleed stream (copper SX raffinate) using a patented process that uses an air/SO$_2$ mixture to oxidize and precipitate Fe and Mn/Al in a single step. It must be noted that no Mn or Al removal is required when producing a CoS product as these two metals do not precipitate as sulphides.

This section is usually equipped with approximately four agitated, overflow tanks in series. The feed liquor is often heated to around 50°C via indirect heating and a residence time of up to 5 hours is required for substantial (>90%) manganese removal.

The tanks will be equipped with gas spargers to inject the air/SO$_2$ mixture. The addition of the air/SO$_2$ gas mixture is controlled by adding the air at a fixed volume and then controlling the SO$_2$ into the tank at an approximate SO$_2$ content of ~0.5% by volume. The redox controller ensures that the reaction remains in an oxidizing regime with a measured redox potential above 600 mV by controlling the flow of SO$_2$.

The pH is controlled by adding limestone to the solution to increase the pH to around 2.5 to 3.0 (a pH of ~3.5 is necessary for aluminum removal in the presence of phosphate). The pH control at this stage is important to ensure that as little Co as possible is precipitated with the Fe and Mn (Al).

The slurry coming from the Fe/Mn/(Al) removal tanks flows by gravity to a high rate thickener where the supernatant liquor is decanted from the precipitated solids. The solids are concentrated in the underflow to around 30%, following which they are pumped to the Fe/Mn/(Al) removal filters. A thickener underflow recycle ratio of approximately 200% is required for crystal growth.

The Fe/Mn/(Al) removal filters are used to separate the supernatant liquor left in the thickened slurry from the solids. Two filters will normally be used in this application: one operating and a standby. A counter current washing regime is used to minimize the wash solution required while still achieving high wash efficiencies.

The washed cake is disposed of with the neutralized leach residue.

**Secondary copper solvent extraction**

Copper in solution can be removed via precipitation with lime (similar to Fe and Al removal) or via SX. The SX route is generally the preferred route as the capital and operating costs associated with the precipitation and thickening route are usually higher, and it imposes a significant recirculating load of cobalt.
The aqueous liquor from the Fe/Mn/(Al) thickener is contacted with a bleed of the stripped organic from the primary copper SX circuit to remove minor quantities of copper. This increases copper recovery and minimizes contamination of the cobalt product.

Copper extraction is usually completed in one or two mixer-settlers. Cu-loaded organic is mixed with the loaded organic from the primary Cu SX in the primary loaded organic wash or strip mixer-settler.

**Aluminium removal**

This step is often not required as aluminium is often precipitated to a sufficient extent as the phosphate complex during iron removal. Al precipitation is not required when producing CoS as Al is not precipitated as the sulphide. Al precipitation is completed in agitated tanks with a combined residence time of around two hours. The precipitation is run at an elevated temperature of around 50°C usually via direct steam injection, and controlled to a pH of around 5.5 using slaked lime (Ca(OH)\(_2\)).

The underflow is concentrated to around 40% solids in the Al removal thickener. A large underflow recycle is required to aid crystal growth and the underflow is recycled to the first tank in the Fe and Mn removal stage.

The thickener O/F will still contain solids that need to be removed from the process before the solution is pumped to the cobalt precipitation area. A pinned bed type clarifier is used as it can efficiently take out these solids in a single stage.

**Intermediate product production**

**Cobalt hydroxide precipitation**

The Co(OH)\(_2\) intermediate product is formed by contacting Fe, Mn, Al and Cu free solution in a series of agitated tanks at elevated temperature with a slurry of Ca(OH)\(_2\).

\[
\text{CoSO}_4(aq) + \text{Ca(OH)}_2(s) + 2\text{H}_2\text{O} \rightarrow \text{Co(OH)}_2(s) + \text{CaSO}_4 \cdot 2\text{H}_2\text{O}(s)
\]

The reaction co-precipitates gypsum, which contaminates and significantly lowers the grade of the product. Magnesium and residual manganese are also partially precipitated as hydroxides, the extent of which is dependent on the terminal pH of the reaction sequence. The combined cobalt hydroxide and gypsum slurry is thickened, filtered and washed, and the filter cake dried before being loaded into bulk bags for dispatch.

A preferred method for producing the hydroxide product is by using milk of magnesia as the precipitant. This route avoids gypsum contamination of the product, thereby producing a purer cobalt hydroxide product than the milk of lime precipitation.

Sodium hydroxide may also be used as the precipitant with advantages similar to magnesia. However, this will result in a final effluent containing sodium sulphate, which will have to be disposed of or contained.

For all three precipitants, product purity may be increased further by utilizing a two-stage precipitation process. The intent here is to precipitate ~90% of the cobalt at pH ~7.5 in the first stage as a relatively pure cobalt hydroxide product, low in manganese and magnesium. The resultant slurry is thickened, filtered, bagged and dried for dispatch.

The first stage thickener overflow solution is then subjected to a second precipitation step to recover the remaining ~10% of cobalt. This is achieved with milk of lime addition at pH 8.0 to 8.5. The slurry is thickened and returned to a suitable place in the cobalt purification circuit.
**Cobalt carbonate precipitation**

A pure CoCO₃ intermediate product is formed by contacting Fe, Mn, Al and Cu free solution in a series of agitated tanks at elevated temperature with a slurry of Na₂CO₃. The resulting CoCO₃ slurry is thickened, filtered and dried before being loaded into bulk bags for dispatch. If the product is to be refined further on site the drying is generally not required.

**Cobalt sulphide precipitation**

Cobalt sulphide can be precipitated using sodium hydrosulphide (NaHS) or hydrogen sulphide (H₂S). While it is possible to perform a hydrogen sulphide cobalt precipitation at atmospheric pressure, this would recover only approximately 85 to 90% of the cobalt to the sulphide. The remainder would thus need to be recovered by an additional precipitation step using, for example, milk of lime, and perhaps recycled in this instance to the Cu scavenger neutralization circuit for redissolution and subsequent reprecipitation.

The precipitation of a sulphide intermediate using hydrogen sulphide is a series of relatively complex and, due to the toxic nature of hydrogen sulphide, high risk operations. The requirement for hydrogen for the generation of hydrogen sulphide necessitates the installation and operation of a hydrogen plant in addition to a hydrogen sulphide plant. The large volume of purge gas for maintenance operations may also dictate the requirement for a small nitrogen (air separation) plant. Each of these gas plants is complex and intricate, requiring a skilled labor complement for their reliable and safe operation. For these reasons the use of hydrogen sulphide will not be discussed further.

Neutralized secondary SX raffinate constitutes the feed to the Co sulphide precipitation circuit. (Note that no specific Fe, Mn or Al removal step is required as Fe(II), Mn and Al will not precipitate as a sulphide; however, some Fe(III) and Al removal may occur during neutralization.)

Pure CoS intermediate product is formed by contacting Fe(III)-free solution in a series of agitated tanks at elevated temperature with a solution of NaHS. The resulting CoS slurry is thickened and filtered, and the filter cake loaded into bulk bags for dispatch.

**Cobalt sulphate crystallization**

Iron, manganese, aluminium and copper free solution constitutes the feed to the CoSO₄ crystallizer plant.

The CoSO₄ intermediate product is formed by first concentrating the feed solution in a large falling film evaporator, and then crystallizing the concentrated liquor in a smaller evaporator/crystallizer. Washing, drying and bagging facilities are provided to ensure a pure crystal product with minimum contamination and moisture.

The operation of the crystallizer is complicated by the fact that a calcium saturated solution is formed, resulting in gypsum precipitation and scaling in the crystallizer. This process will also result in the co-crystallization of all impurities with the cobalt product, including magnesium and calcium resulting in a low grade product.

**Intermediate product comparison**

The four product options listed above are compared and discussed with reference to a number of key project drivers.

**Capital cost (excluding effluent treatment)**

Capital costs associated with the carbonate and hydroxide routes are similar. Both routes are only partially selective for manganese and magnesium and thus a similar degree of purification is required prior to final product precipitation. The sulphate route has the highest
capital cost due to the requirement for an evaporator and crystallizer in addition to the preceding impurity removal steps. The sulphide flowsheet is the cheapest option in terms of capital cost, despite the requirement for closed vessels and off-gas scrubbing equipment. Cobalt sulphide is selectively precipitated in the presence of manganese and aluminium and thus the removal of these elements is not required.

Operating cost
The sulphate route has the advantage that no reagents are required for final product production. However, a significant amount of energy is required for solution evaporation, and this cost makes it the most expensive to operate. Operating costs for hydroxide, carbonate and sulphide are similar; if lime is utilized for hydroxide precipitation that route is slightly cheaper.

Metal recovery
The recoveries of all four routes are similar. For the hydroxide route a two stage precipitation process is recommended to recover and recycle soluble cobalt. The disadvantage of the sulphide route is that any off specification material is difficult to redissolve without specialized equipment (autoclave).

This is often the most important factor to consider when choosing a preferred route. HSE issues are more prevalent for the sulphide intermediate due to the potential for H₂S generation. In addition to this, the use of NaHS will result in a final effluent containing sodium sulphate, which will have to be disposed of or contained. The problem associated with sodium sulphate disposal is common for the hydroxide-caustic and carbonate route as well and can often be a significant additional cost. Hydroxide (lime/magnesia) and sulphate intermediates usually have lower HSE issues.

Product purity
The sulphide product contains the highest percentage of cobalt followed by the hydroxide (caustic or magnesia), carbonate then the low-grade hydroxide (lime), while sulphate has the lowest contained cobalt. The major impurity associated with the sulphide intermediate is magnesium; with sulphate it is manganese, magnesium and calcium; with carbonate it is manganese and calcium; and with the hydroxide it is manganese and either calcium or magnesium, depending on the reagent used for precipitation, i.e. lime or magnesia.

Apart from the rejection of Al and Fe(II) by sulphide precipitation, there is negligible selectivity in this application for the remaining minor elements (Cu, Zn, Ni) by all options.

Operability and maintenance
The sulphide process requires off-gas scrubbing and tight pH control during precipitation. In addition to this, the handling and corrosive nature of NaHS must be considered. For the production of sulphate, the operation of a crystallizer and evaporator is relatively complex and maintenance intensive so is rated more difficult than the carbonate and hydroxide routes. Large recycles required for the hydroxide route make the circuit cumbersome from an operability perspective, and the relatively poor settling and filtration characteristics of the hydroxide product aggravate this. Descaling will be required in all cases.

Sales and marketing
This will depend on product purity, but in general terms marketing information indicates that the customer base for hydroxide, carbonate and sulphate is larger than for the sulphide. The value received from sales for carbonate and sulphate is higher than that for hydroxide and sulphide.
**Future cobalt metal production**

In a number of cases clients wish to adopt a strategy of cobalt intermediate production in the initial years of project life with the conversion to metal production as revenue from copper and intermediate sales stabilizes and operational experience is gained. The sulphide intermediate would potentially require ultra-fine grinding followed by pressure oxidative leaching. Intermediate hydroxide and sulphate products would require additional manganese impurity removal steps prior to cobalt EW. The carbonate would be the ‘easiest’ to refine. Those products containing significant gypsum will obviously produce calcium saturated liquors upon dissolution, requiring additional consideration and control within the cobalt refinery.

The above comparison is summarized in Table I below; the intermediate options are rated 1–4, with 4 being the preferred option.

**Recommended intermediate product**

The cobalt sulphate route is eliminated on the basis of cost and the production of a low purity product. Cobalt sulphide production is by far the most complex process, and HSE as well as operability issues eliminate this option.

The two best intermediate product routes are the carbonate and the hydroxide. The choice between the two has to be made considering environmental issues associated with the disposal of sodium sulphate effluent. If the plant is close to a large water body, the disposal of the effluent via dilution may be possible. However, with increasingly stringent environmental legislation, this is becoming less likely unless the effluent can be pumped to the sea. The production of the hydroxide via magnesia (MgO) precipitation will result in a reasonably high quality product, and the liming of the effluent to a pH of around 10.5 will result in a magnesium free solution, which can be disposed of a conventional tailing dam. The precipitate from the neutralization can be stored in a relatively small lined facility.

The production of cobalt hydroxide using a two-stage process is in most cases the recommended route.

**Pure metal product options**

Cobalt metal can be produced via hydrogen reduction (powder or briquettes) or EW (full plate cathode or rounds). EW is believed to be the preferred method in almost all cases in terms of lower capital and operating cost, higher metal recoveries and less stringent operability and

### Table I
Cobalt intermediate product ratings

<table>
<thead>
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<th>Parameter</th>
<th>Co(OH)₂</th>
<th>CoCO₃</th>
<th>CoS</th>
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<td>Capex</td>
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</tr>
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<td>Future metal production</td>
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<td>3</td>
</tr>
<tr>
<td>Total</td>
<td>20</td>
<td>27</td>
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</tr>
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</table>
maintenance requirements, and for the purpose of this comparison only EW will be considered. The production of cobalt rounds is a patented process, so this paper will concentrate on the production of full plate cathode.

Two different processing routes will be considered; these are described below:

- Cobalt hydroxide precipitation followed by redissolution, impurity removal and EW
- Cobalt SX, impurity removal and EW.

**Precipitation and redissolution**

**Cobalt precipitation**

The primary cobalt hydroxide precipitate is the feed material to the cobalt metal plant. The use of magnesia is unnecessary and a cheaper precipitant, i.e., milk of lime, is more suitable, although calcium saturation issues and larger volumes of redissolution residue must be considered. Primary precipitate, in the form of a filter cake, is advanced to the cobalt redissolution circuit; secondary precipitate (high manganese) is recycled to the iron/manganese/(aluminum) precipitation circuit feed for recovery of the contained cobalt.

**Cobalt redissolution**

Primary cobalt hydroxide filter cake is dissolved using cobalt spent electrolyte and sulphuric acid in a cascade of agitated tanks. Oxidation and partial neutralization of the leached slurry may be used to scavenge iron and manganese using milk of lime and air/SO₂ gas mixture.

Aluminum will redissolve and will remain largely ‘inert’ in the refinery, unable to plate with the cobalt. It will either exit as entrained liquor in the redissolution residue filter cake, or as part of the spent electrolyte bleed.

Leached slurry is thickened and filtered and the filter cake is recycled upstream to recover any undissolved cobalt. Filtrate is clarified in a polishing filter for final solids removal, and is then diluted with clean water to ensure liquor advancing to the IX circuits is comfortably below calcium saturation.

**Zinc ion exchange (IX)**

Zinc and residual copper are removed from the cobalt rich liquor using a macroporous aminophosphonic acid chelating resin within fixed-bed columns operated in a typical lead-lag configuration. A split elution technique using sulphuric acid is used to maximize zinc-cobalt selectivity. Weak eluate is recycled to cobalt precipitation, and strong eluate is disposed via tails neutralization. Regeneration of the resin is achieved using a caustic solution.

**Nickel ion exchange (IX)**

Nickel is partially removed from the zinc and copper-free cobalt rich liquor using a macroporous bis-picolyamine chelating resin. As electrowinning is selective for cobalt over nickel, removal of nickel to ~0.2 g/l is generally sufficient to meet cathode specification.

Similar configuration and elution/regeneration techniques to the Zn IX circuit are used; however, the inherently low nickel-cobalt selectivity and high unit cost of the resin, encourage the use of smaller bed volumes and a consequent increased number of columns.

**Cobalt electrowinning (EW)**

Electrowinning is conducted at ~70°C in standard undivided cells. Stainless steel permanent blanks are used as cathodes, with Pb-Sn-Ca anodes at 120 mm cathode-cathode spacing. A
rich electrolyte tenor of at least 50 g/l Co is achieved by adjusting spent electrolyte recycle to the re-solution circuit. This permits EW operation with a cell drop of 5 g/l, producing a spent electrolyte of 45 g/l Co and ~8 g/l H2SO4. Under these conditions, a current density of 250 A/m² will typically give a current efficiency of ~80%.

The use of divided cell technology, particularly anode bags, would potentially allow higher cell drops to be achieved (>20 g/l), thereby significantly reducing the flow through the purification circuit. However, given the highly stressed nature of full-deposit cobalt cathode and the resultant potential for frequent damage to the surrounding bags, significant development is required to reduce process risk before this technology can be considered for application in the DRC.

Cathodes are harvested live on a six/seven-day cycle. They are washed with hot water to remove surface electrolyte and are then manually stripped. Stripped blanks can be treated with dilute nitric acid for removal of residual metal deposits. Prior to return to the cells, stripped blanks are dipped in gelatine to create a favorable deposit-blank bond strength.

Stripped cobalt cathode pieces are crushed to -70 mm, vacuum degassed to remove contained hydrogen and nitrogen, and burnished in rotating drums to achieve the desired product finish. Final product is packaged in sealed drums for dispatch.

Manganese that is oxidized at the anode forms insoluble MnO₂. This is periodically washed from the anodes and hosed from the cell bottoms.

Spent electrolyte is returned to the redissolution circuit as required. Surplus liquor is bled to the cobalt precipitation circuit, providing an important outlet for such species as magnesium, aluminium, sodium, sulphate and chloride.

**Cobalt SX EW**

For this refining configuration, liquor exiting the secondary copper SX stage forms the feed to the cobalt SX circuit. Sufficient removal of organic from the secondary copper SX raffinate must be performed to limit the transfer of copper extractant (oxime) to the Cobalt SX organic. A diluent wash is considered to have good potential in this application due to the availability of relatively large volumes of make-up diluent required in the primary copper SX circuit.

**Cobalt SX**

Cobalt is recovered from the partially purified bleed liquor in a multiple-stage SX circuit utilizing a phosphinic acid organic extractant dissolved in a suitable hydrocarbon diluent. Zinc, iron, aluminum, copper and manganese are all co-extracted, while calcium, nickel and some magnesium are rejected.

The extraction circuit usually consists of four countercurrent stages. Each stage comprises a single mixer providing a total of 1 minute mixing time and a settler for phase disengagement. To maximize cobalt-nickel selectivity, pH is controlled within each stage by direct addition of caustic solution to the mixers.

Two-stage organic removal is used for the raffinate, generally comprising a diluent wash stage and Jameson flotation cell. Clean raffinate can be recycled within the process (e.g. CCD wash water), with surplus advanced to tails neutralization. The use of sodium hydroxide results in a raffinate containing sodium sulphate which, as per several of the intermediate options, will have to be disposed of or contained.

Care must be used to avoid recycling cobalt SX extractant directly into the copper SX circuits. Circuits containing solids generally act as good organic adsorbents, thereby significantly reducing this risk.
Cobalt loaded organic is scrubbed in a single stage to remove co-extracted and entrained nickel (and magnesium). Demineralized water and sulfuric acid are contacted with the loaded organic at a suitably low pH to achieve the required Co:Ni ratio on the scrubbed organic. A bleed of cobalt loaded strip liquor can be added to the scrub stage to improve scrubbing efficiency if required. Scrub raffinate is returned to the first cobalt extraction stage.

Scrubbed organic is stripped in two countercurrent stages to remove all loaded metals. Cobalt spent electrolyte is used as strip liquor, the flow rate of which is adjusted to achieve the required cobalt tenor for downstream electrowinning. Dilute sulphuric acid solution is used to control pH in the first strip stage to ensure required stripping efficiency is achieved. Stripped organic is returned to extraction via a stripped organic tank.

Zinc SX

Zinc (and ferric iron) is removed from the cobalt SX loaded strip liquor in a multiple-stage SX circuit. While Di-Ethyl Hexyl Phosphoric Acid (DEHPA) is an option in this application, the use of phosphinic acid is preferred due to the ability to reject calcium, relative ease of iron stripping, and improved compatibility with cobalt SX (no extractant cross-contamination issues, ability to use same crud treatment equipment, etc.).

The phosphinic acid extracts all zinc and ferric, whilst aluminium, copper, manganese and cobalt are rejected. To ensure the target iron level is maintained in the zinc SX raffinate, all ferrous in the cobalt loaded strip liquor is oxidized to the ferric form just prior to zinc extraction using hydrogen peroxide in an in-line mixer.

Zinc is extracted in three countercurrent pH-controlled stages. Thorough removal of organic is required for the downstream circuits, so a three-stage system is used generally consisting of a Jameson cell, liquid-liquid coalescer and carbon columns.

Zinc loaded organic is scrubbed with water and sulphuric acid in a single stage to remove co-extracted and entrained cobalt. Scrub raffinate is returned to the first zinc extraction stage. Scrubbed organic is stripped with acidified water in two countercurrent stages to remove all loaded metals. Stripped organic is returned to extraction via a stripped organic tank.

Entrained organic in the zinc loaded strip liquor is recovered using a liquid-liquid coalescer, and the cleaned liquor is disposed via tails neutralization.

Copper IX

Residual copper is removed from the zinc SX raffinate using a macroporous amino-phosphonic acid chelating resin within fixed-bed columns operated in a typical lead-lag configuration. A single stage elution is used due to the ability to recover all eluted copper (and cobalt) upstream. Regeneration of the resin is achieved using a caustic solution.

Cobalt EW

Cobalt electrowinning is carried out in identical fashion to that described in the precipitation/re-dissolution process.

Spent electrolyte is returned to the cobalt SX strip circuit as required. Surplus liquor is bled to the cobalt SX extraction circuit (blended with PLS) providing the only significant outlet for such species as magnesium, aluminum, sodium, sulphate and chloride.

Metal production route comparison

The two general production routes described above are compared and discussed with reference to a number of key project drivers.
Capital cost (excluding effluent disposal)

Capital costs associated with both the precipitation and solvent extraction routes are similar. For the second route, the additional cost of the cobalt SX circuit (organic inventory) is effectively offset by the elimination of the nickel IX circuit, comprising a very expensive resin inventory of its own.

Being solely dependant upon fixed bed IX, the precipitation/redissolution route is considered to be more capital sensitive to the florate and composition of the cobalt rich liquor feeding the purification steps. For this reason, the adoption of divided cell technology to increase cell drop in the cobalt tank house would bring significant capital savings to the precipitation/redissolution route (with smaller savings to the SX route).

Operating cost

The SX route consumes much larger amounts of caustic than the precipitation/redissolution based route due to its use for primary cobalt recovery in the cobalt SX circuit. This is offset by lower consumption of lime (effectively replaced by caustic) sulphuric acid and limestone (reduced recirculating loads of cobalt and impurities). The operating cost differential between the two options is therefore highly sensitive to the prevailing price of caustic; below US$800/t (including delivery) the SX route is generally favoured.

Metal recovery

Due to the higher selectivity of the impurity removal steps, cobalt recovery for the SX route is somewhat higher (~10%) than that for the precipitation/redissolution route. Cobalt loss with both nickel and zinc is reduced considerably by the use of SX. Cobalt losses are also reduced due to a significant reduction in the recirculating load of cobalt (elimination of secondary cobalt precipitate, cobalt re-solution residue, weak eluates and a smaller electrolyte bleed).

Product purity

Both production routes have been compared on the basis of equivalent product purity. The SX route, however, provides additional flexibility with respect to impurity removal, and could tolerate higher than design levels of both zinc and nickel with greater ease than the precipitation/re-solution based route.

Operability and maintenance

Both processing routes may be considered complex refineries, requiring steady operation and tight control to achieve desired performance. The elimination of virtually all solid-liquid circuits and consequent batch filtration steps is expected to provide an operability advantage to the SX route.

The SX route also produces cobalt rich liquor (electrolyte) very low in calcium, as opposed to the high calcium liquor produced via the precipitation/redissolution route. The need for periodic de-scaling of the cobalt precipitation and redissolution circuits, and the potential for downstream gypsum precipitation within the IX and EW circuits places the precipitation/resolution route at a further disadvantage in terms of both operability and maintenance.

Health, safety and environmental issues

The presence of larger volumes of organic solutions in the SX route requires increased measures to control personnel exposure and fire risks.
Environmentally, the SX route produces an effluent containing large amounts of sodium sulphate that must be adequately disposed of or contained, whereas in its place the precipitation/redissolution route produces gypsum, the disposal of which is quite straightforward.

The above comparison is summarized in Table II below; the metal production options are rated 1 or 2, with 2 being the preferred option.

**Recommended metal production process**

Technically, the SX based route is equal or superior on most parameters, and excluding the cost of effluent treatment, would be expected to give a superior return to the precipitation/redissolution based route.

As with the choice of intermediate, however, the selection of a refining route is driven by the costs and environmental issues associated with the disposal of a sodium sulphate containing effluent.

The use of ammonia in place of caustic for the SX route would allow ammonia recovery using a lime-boil process, producing a gypsum residue for disposal. The low concentration and high volume of effluent to be treated, however, do not make this option particularly attractive, and a cheap source of low pressure steam would be essential.

On this basis, the SX route is effectively eliminated, and the recommended route is precipitation of a cobalt hydroxide using lime, redissolution in spent electrolyte, purification and metal recovery via electrowinning.

**Comparative nett present value (NPV) between hydroxide intermediate and metal**

A high level NPV model has been applied to the following scenarios:
- Cobalt hydroxide intermediate production
- Cobalt hydroxide intermediate production followed by redissolution and cathode production (i.e. conversion to a metal production circuit after operating an intermediate production circuit)
- Cobalt cathode production via intermediate production (i.e. direct metal production with no saleable intermediate).

The following assumptions have been applied
- 15% discount rate
- 9 000 tpa cobalt metal production
- US$16/lb cobalt metal
- Cobalt hydroxide intermediate realizes 70% of the contained metal value.

Table III summarizes the input data for the NPV calculation.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Precipitation and redissolution</th>
<th>SX</th>
</tr>
</thead>
<tbody>
<tr>
<td>Capex</td>
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<td>1</td>
</tr>
<tr>
<td>Opex</td>
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<td>2</td>
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<td>2</td>
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<tr>
<td>Purity</td>
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<td>2</td>
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<td>2</td>
</tr>
<tr>
<td>HSE</td>
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<td>1</td>
</tr>
<tr>
<td>Total</td>
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<td>10</td>
</tr>
</tbody>
</table>
Capital cost

The costs listed above are high level estimates and meant for comparative purposes. The estimate consists of costs for earthworks, civils, structural steel, plate work, mechanical equipment, piping, electrical and instrumentation and buildings. It excludes indirect costs but includes a 15% contingency. The battery limits covered are from the copper SX raffinate pond to product dispatch, all internal recycles are costed, and effluent reporting to the copper circuit is costed at a convenient point between the two plants. All reagent storage and utilities are included.

The cost for the ‘cathode from intermediate’ option is US$12.4 million higher than the ‘cathode via redissolution’ option, and this additional amount covers a second stage cobalt precipitation, drying and bagging facility.

Operating cost

The operating cost is defined as the incremental cost (total cost being for the copper and cobalt facility) required to produce the cobalt product. This includes labour, power, operating and maintenance spares, vehicles, reagents (including SO₂ for leaching), consumables (including laboratory) and other.

The operating cost estimate is to the same level of accuracy as the capital cost and is meant for comparative purposes.

Production and revenue

A cobalt production of 9,000 tpa has been assumed at a cobalt price of US$16/lb (spot price on 21 November 2008). There will be some minor differences in cobalt recovery; however, at the level of accuracy considered above this is regarded as being negligible.

NPV calculation

The NPV model shows an overwhelming financial benefit for production of cobalt metal. Figure 2 summarizes the values.

From Figure 2 it can be seen that the NPV for the cobalt metal options is far larger than for the intermediate option. This is due to the large increase in revenue associated with metal production, as only 70% of the cobalt value has been assigned to the intermediate product, and this is believed to be an optimistic view. The NPV for the intermediate option is MUS$647 as opposed to MUS$894 for the cathode from intermediate option and MUS$905 from the cathode via re-solution option.

The relative change in NPV between the two metal production options is small.

Conclusions and recommendations

Numerous factors have been discussed, which affect the type of cobalt product that is produced. It is often environmental considerations and operational complexity that determine the optimum route.
It most cases copper is the major contributor to project revenue but cobalt often makes a significant contribution. In the early years of operation it is wise to follow the lowest complexity route possible, and thus the production of cobalt hydroxide intermediate is recommended. This also reduces the upfront capital required and lowers overall project risk.

Once operational experience has been obtained and cash flow realized, the conversion from the intermediate product to metal should be considered as this returns greater value. The complexity of cobalt metal production should not be underestimated, particularly in the DRC where experienced operators are scarce. Cobalt metal production should not be undertaken before understanding the complexity and risk.

Figure 2. Product NPV values

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Graduated from Wits University in 1988 with a BSc Hons degree in Applied Chemistry. As a JCI bursar I was assigned to Randfontein Estates Gold Mine for in service training. After completion of this training I was transferred to the JCI Minerals Processing Research Labs (later Anglo Platinum Research Centre), where I was involved in development work for the Anglo Platinum Precious Metals Refinery, Anglo Platinum Base Metals Refinery and managed the pilot plant campaign for the ACP project (Anglo Platinum Converting Process).

After this I was seconded to ATD (Anglo American Technical Division) where I was the Senior Process Engineer on the Konkola High Pressure Leach Project, amongst other studies, and worked for a short period at KCM in Zambia as Metallurgical Projects Manager. I returned to Anglo Platinum Process Technology Division in early 2002 as a Senior Process Engineer and worked projects and optimisation work associated for the Anglo Platinum smelting and converting operations in Rustenburg and Polokwane, including the ACP project, Waterval smelter upgrade, Waterval Slag Cleaning Furnace and Polokwane smelter project.

Joined GRD Minproc as a Senior Process Engineer in January 2006 and was promoted to Process Engineering Manager in September of that year. I am currently responsible for all process engineering activities in the South African office and have managed a number of bankable feasibility studies.