

## **COMPARISON OF SOLVENT EXTRACTION AND SELECTIVE PRECIPITATION FOR THE PURIFICATION OF COBALT ELECTROLYTES AT THE LUILU REFINERY, DRC**

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### **ABSTRACT**

The classic technique to remove impurities in a cobalt stream is selective precipitation. This is usually followed by re-leaching to produce an advanced electrolyte for cobalt electrowinning. Although this technique ensures an adequate cobalt concentration in the electrowinning circuit, some of the consequences include unwanted losses through the numerous selective precipitation steps resulting in lower recoveries. One plant that currently utilizes selective precipitation is the Luilu Metallurgical Plant (Luilu) in the Democratic Republic of the Congo.

This paper evaluates the benefits of using solvent extraction technology as a purification step in the Luilu Process with the aim of further increasing the overall cobalt recovery and to ensure that a suitable cobalt concentration in the advanced electrolyte is achieved.

The implementation of a solvent extraction-electrowinning process is found to be beneficial in terms of achieving improved recoveries and cobalt cathode quality, however there are many other factors that impact the success of converting from the classic precipitation process used at Luilu.

### **INTRODUCTION**

The Luilu metallurgical plant is located approximately 18km northwest of Kolwezi in the Democratic Republic of the Congo (DRC). It was originally constructed in 1960 and in 1972 it was expanded to a nameplate of 175,000 tonnes of copper and 8,000 tonnes of cobalt per annum. The site consists of roasters, leaching, precipitation circuits and electrolytic cells for copper and cobalt production. From 1984 through 1989, the annual production at Luilu averaged 173,000 tonnes of copper and 5,900 tonnes of cobalt. The highest production year was 1986 with 177,500 tonnes of copper and 7,800 tonnes of cobalt. By 1996, production had fallen to an estimated 27,000 tonnes of copper and 1,200 tonnes of cobalt, and has continued to decline since then<sup>(1)</sup>.

A rehabilitation feasibility study was conducted in 2005 for the Kamoto Mine and related infrastructure, including the Kamoto Concentrator and Luilu Refinery, to restore production of copper and cobalt to the original design capacity. The production will be increased in phases over a period of four years allowing time for the ore tonnages that are fed to the

concentrator to increase, and for the refurbishment of the concentrator and refinery sites to take place.

The traditional processing methods at the Luilu Refinery, as originally constructed in 1960 will initially be refurbished to restore production as quickly as possible. However the implementation of newer advanced technologies is currently being investigated.

Cobalt rich solution is bled from the copper spent electrolyte, and undergoes selective precipitation steps to purify the solution before electrowinning. The cobalt stream is then subjected to final precipitation and re-leaching with sulphuric acid in aim to obtain a suitable cobalt concentration in the advance electrolyte.

Many cobalt plants still follow this classic technique of selective precipitation as the purification method. Although this technique ensures an adequate cobalt concentration in the electrowinning circuit, some of the consequences include unwanted losses through the numerous precipitation steps, resulting in lower recoveries. One of the proposed process improvements, which aim at increasing the overall metal recoveries, include the implementation of a solvent extraction circuit as a purification method for the cobalt rich solution.

A preliminary trade-off study<sup>(2)</sup> was conducted to compare the capital and operating cost of one of the optional solvent extraction circuits versus the refurbishment cost of rehabilitating a section of the current cobalt purification circuit. Based on solvent extraction batch test work results performed by Mintek<sup>(3)</sup>, a preliminary flowsheet was developed and costed.

This paper discusses the following:

- A general discussion on the implementation of different solvent extraction pilot plant tests and circuits in Africa;
- Current process flowsheet for purification of cobalt at Luilu;
- Solvent Extraction test work results done at Mintek on Luilu feed – the test work is based on only one of the possible solvent extraction circuit options;
- Solvent Extraction circuit process description based on the test work done by Mintek which forms the basis for the cost estimate specific for Luilu;
- Summary of Cost Estimate;
- Further Considerations;
- Additional Challenges;
- Current Status of Luilu.

## **BACKGROUND**

The Democratic Republic of the Congo (DRC), formerly known as Zaire, is Africa's third-largest country, and approximately the size of Western Europe. All of its 11 provinces are rich in minerals, which makes it potentially one of the richest mining countries in Africa<sup>(4)</sup>. The copper, cobalt and diamond mining industries have the potential to be the largest on the

continent, while the gold mining industry also has excellent potential. The DRC's main copper and cobalt interests are dominated by Gécamines, the State-owned mining giant<sup>(4)</sup>.

The Copperbelt extends from Angola through the DRC into Zambia, and reserve estimates from the copperbelt total 55.5 Mt of copper and 3.6 Mt of cobalt. It has long been an area of mining exploitation by major international companies from colonial times to the present resurgence of interest, driven by an international metals boom. Kolwezi is a mining town located on the Congolese Copperbelt approximately 300 km north-west of Lubumbashi, the capital of the Katanga Province<sup>(4)</sup> in the DRC and is shown in Figure 1.

Since about 1993 production in the DRC has stagnated with total capacity utilisation estimated at less than 10 percent. The major contributor to this is political instability in the area. However, the DRC government has attempted to improve its declining production by promoting several ailing mines and projects to foreign investors<sup>(4)</sup>.

New mining contracts have recently been approved, which combined with high mineral and metal prices<sup>(6, 7)</sup> (Figure 2), could improve the DRC's fiscal position and GDP growth. The renewed interest by the international mining industry in the DRC has been as a result of the 2003 implementation of the new DRC Mining Code, which was drafted in conjunction with the World Bank<sup>(4, 8)</sup>.



Figure 1 – Democratic Republic of the Congo<sup>(1)</sup>

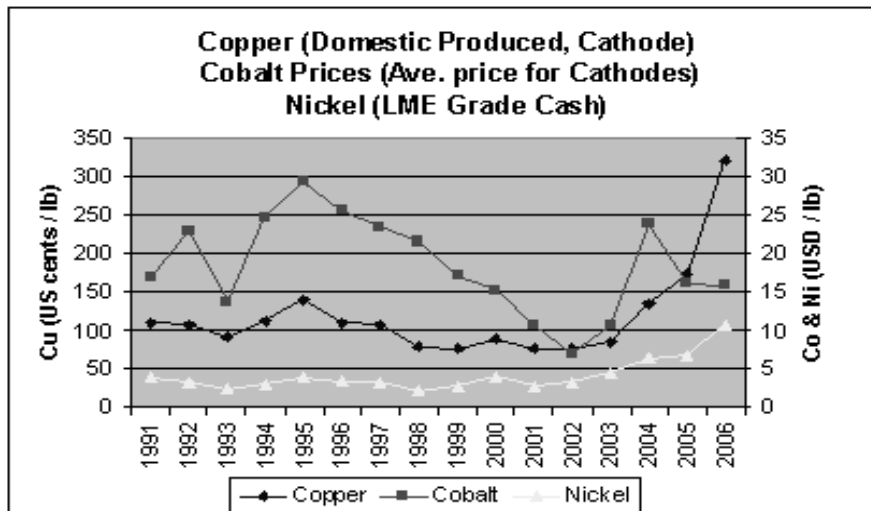


Figure 2 – Commodity prices for Copper, Cobalt and Nickel<sup>(7)</sup>

A feasibility study was commissioned to develop a comprehensive plan for the rehabilitation and redevelopment of the Kamoto mine and related infrastructure located near Kolwezi in the Democratic Republic of the Congo (DRC).

The Kamoto Copper Company feasibility study was commissioned by Kinross Forrest Limited (KFL), the owner of a 75% interest in the Kamoto Joint venture. The other 25% of the Kamoto Joint venture is owned by Gécamines. Katanga Mining Limited has purchased one hundred percent of Kinross Forrest Limited in the Joint Venture in June 2006<sup>(1, 9, 10)</sup> (Figure 3). After successful completion of the feasibility study, Katanga Mining Limited (KML) was listed on the Toronto Stock Exchange with effect from June 2006.

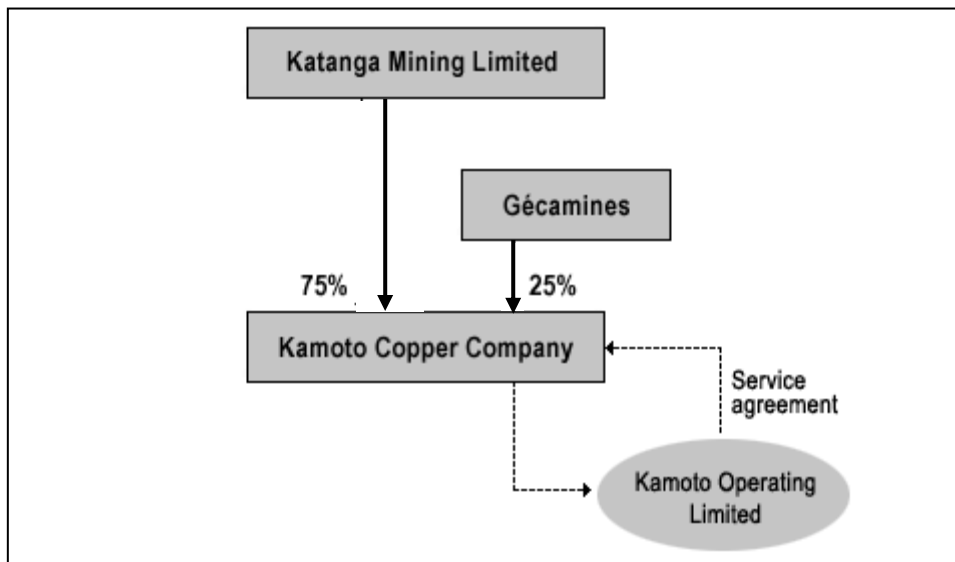


Figure 3 – Company Structure<sup>(9)</sup>

The re-establishment of operations back to the original design capacity is based on a phased approach over a four-year period, which was based on an assessment of the condition of the plant sections, the capacity constraints of the facilities and the condition of the mines.

The results of the feasibility study (2005) demonstrated that economically viable operations can be restored within a relatively short time frame. A copper price of \$1.10/lb and a cobalt price of \$10/lb were used in the cost estimate, which allowed for a conservative approach to revenue generated<sup>(1)</sup>. Capital costs, on a comparative basis for the restored production capacity, are very low. On an annual basis based on current information, the operation is predicted to become cash neutral in year four and cash positive in year eight.

This gives an illustration of the high potential of the project by utilizing the existing process. The implementation of newer, more efficient technologies can only add to the already strong case for the rehabilitation of the Kamoto Mine and surrounding infrastructures.

### CURRENT SOLVENT EXTRACTION TECHNOLOGIES IN AFRICA

During the last two decades, many diverse hydrometallurgical solvent extraction circuit installations applied for the recovery of zinc, cobalt and nickel have been commissioned in Southern Africa. Many flowsheets that use solvent extraction processes are currently under feasibility study, pilot plant, development, commissioning, or re-commissioning for implementation in this part of the world<sup>(11, 12)</sup>.

The use of precipitation steps are still widely used as the choice for separating metals, however incorporating solvent extraction together with precipitation can provide the required recoveries and purities for today's industry. Table 1 summarizes some of the successful impurity removal and cobalt solvent extraction implementations and studies currently in Africa<sup>(12)</sup>.

Company	Zinc SX (Impurity Removal SX)	Cobalt SX	Nickel SX	Precipitation
<b>Kasese Cobalt Company Ltd (KCCL), Uganda</b>	Zinc and manganese removed with 2 vol.% D2EHPA.	Selective extraction of cobalt from nickel and magnesium with CYANEX 272.		Iron is precipitated in a two-stage neutralization circuit. Copper, cobalt and nickel are precipitated with caustic soda as hydroxides.
<b>Chambishi Metals Plc, Zambia</b>	Zinc removed with 2.5 vol.% D2EHPA.	Exploring the option of cobalt SX to improve overall plant performance by using 30 vol.% CYANEX 272.		Iron removal with precipitation. Cobalt purification with precipitation.
<b>Knightsbridge Cobalt in South Africa treats oxide ore originating in the DRC</b>	Zinc, manganese and calcium removed with 20 vol.% D2EHPA.	Cobalt SX with 15 vol.% CYANEX 272.		Iron, copper and cobalt carbonate precipitation with CaCO <sub>3</sub> .
<b>Kolwezi Tailings, DRC</b>	Zinc is removed with SX using CYANEX 272.	Cobalt SX with CYANEX 272.		Iron, aluminium and manganese impurities

Company	Zinc SX (Impurity Removal SX)	Cobalt SX	Nickel SX	Precipitation
		Pilot plant and study only.		are removed through precipitation with air/SO <sub>2</sub> .
<b>Kakanda Tailings, DRC</b>	Zinc and manganese is removed by using D2EHPA.	Cobalt SX with CYANEX 272. Pilot plant and study only.		Iron precipitation.
<b>Tati Nickel, Botswana (currently in construction)</b>		Cobalt SX with 5 vol.% CYANEX 272.	Nickel purification with 20 vol.% Versatic Acid followed by EW.	Iron precipitation with limestone, might produce CoCO <sub>3</sub> as final product with EW later.
<b>Nkomati, South Africa (Pilot plant only)</b>		Cobalt SX with 7 vol.% CYANEX 272.	Nickel purification with 30 vol.% Versatic Acid followed by EW.	Iron precipitation.

Company	Zinc SX (Impurity Removal SX)	Cobalt SX	Nickel SX	Precipitation
<b>Anglo Platinum Rustenburg Base Metals Refinery, South Africa</b>		Cobalt SX with 15 vol.% D2EHPA.		Lead removal with Ba(OH) <sub>2</sub> before cobalt removal using the Outokumpu nickelic hydroxide process. Residue iron precipitation with NaOH, and copper removal using BaS.
<b>Hartley Platinum, Zimbabwe</b>		Cobalt SX with 3 vol.% CYANEX 272.		CoCO <sub>3</sub> precipitation as final product.
<b>Scorpion Zinc, Namibia</b>	Zinc removal with 40 vol.% D2EHPA.			Iron, aluminium and silica removal with precipitation.

Table 1 – Summary of some of the successful Solvent Extraction implementations in Africa<sup>(12)</sup>

The Kasese Cobalt Company Ltd (KCCL) in Uganda, removes iron in a neutralization circuit. The iron free solution is processed through a first solvent extraction circuit where zinc and some manganese are removed using D2EHPA. After treatment of the raffinate with caustic soda to remove the copper as hydroxide, the solution is transferred to a second solvent extraction circuit in which cobalt is selectively extracted from nickel and magnesium using CYANEX 272. This produces an advanced electrolyte that can be fed to the cobalt electrowinning circuit.

The raffinate from the cobalt solvent extraction uses precipitation to recover nickel as a hydroxide, and cobalt is recovered from the cobalt electrowinning bleed stream also with precipitation.

Chambishi Metals Plc in Zambia do iron removal in a neutralization circuit followed by impurity removal with D2EHPA, then precipitating and re-dissolving the cobalt to increase

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the concentration of the electrolyte feeding cobalt electrowinning. Nickel removal is done with ion exchange prior to electrowinning. Chambishi is also exploring the possibility of cobalt solvent extraction to improve overall plant performance.

Knightsbridge Cobalt Corporation of South Africa utilizes precipitation steps to remove iron and copper from the pregnant leach solution (PLS). Impurity removal is done with a solvent extraction circuit using D2EHPA to remove zinc, magnesium and calcium. Cobalt solvent extraction using, CYANEX 272, is used to increase the cobalt concentration before final precipitation.

The Kolwezi Tailings project (pilot plant campaign by AARL) uses a two-stage solvent extraction circuit. A CYANEX 272 circuit is used to remove impurities followed by a copper/zinc ion exchange circuit. A second separate CYANEX 272 solvent extraction circuit is incorporated to recover cobalt and concentrate the solution before electrowinning. This process is developed on a pilot plant scale only.

Kakanda (pilot plant campaign by Mintek) does an impurity removal solvent extraction circuit with D2EHPA followed by a separate solvent extraction for the cobalt concentration using CYANEX 272. Trace amounts of copper and zinc are removed with ion exchange ahead of cobalt electrowinning. This process is still in a study phase.

The Nkomati project (constructed in Botswana, pilot plant stage only) uses precipitation to remove iron from solution. Cobalt solvent extraction with CYANEX 272 is used and nickel, magnesium and calcium reports to the raffinate. Nickel solvent extraction with Versatic Acid is used to increase the nickel concentration in the advanced electrolyte feeding nickel electrowinning.

Hartley Platinum Base Metal Refinery (Zimbabwe, operation discontinued) utilizes cobalt solvent extraction to remove cobalt before nickel electrowinning. Copper, lead, zinc and iron are all extracted and cobalt is precipitation from this solution to produce an impure cobalt carbonate final product.

Skorpion Zinc (Namibia) process uses an atmospheric leach with sulphuric acid. The iron, aluminium and silica are removed from the solution by precipitation. Zinc is then extracted with D2EHPA to produce the correct concentration for the advanced electrolyte feeding zinc electrowinning.

## **PROCESS DEVELOPMENT**

Efficient separation of cobalt, nickel and to some extent zinc has always been a problem in hydrometallurgy. Their adjacent positions in the transition metal series in the periodic table results in aqueous chemical behavior that is too similar for development of easy separation routes. Traditionally these metals were separated by processes based on selective oxidation and/or precipitation and such processes are still in use today<sup>(13)</sup>.

In Luilu, the removal of metals such as iron, aluminium, copper, zinc and nickel from solution is required to obtain a purified cobalt solution feeding the cobalt electrowinning circuit. Iron, aluminium and copper can easily be removed with selective hydroxide

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precipitation techniques<sup>(14)</sup> (Figure 4), however zinc and nickel removal is done with sulphide precipitation techniques<sup>(14)</sup> (Figure 5).

Using sulphide precipitation for the removal of zinc and nickel from a cobalt stream, results in high cobalt losses to the precipitate. The use of solvent extraction to separate these metals has become the process of choice these days to provide the high degree of separation of metals and yields demanded by today's industry<sup>(13)</sup>.

Figures 4 and 5 provide a broad indication of the degree of separation that can be expected with hydroxide and sulphide precipitation.

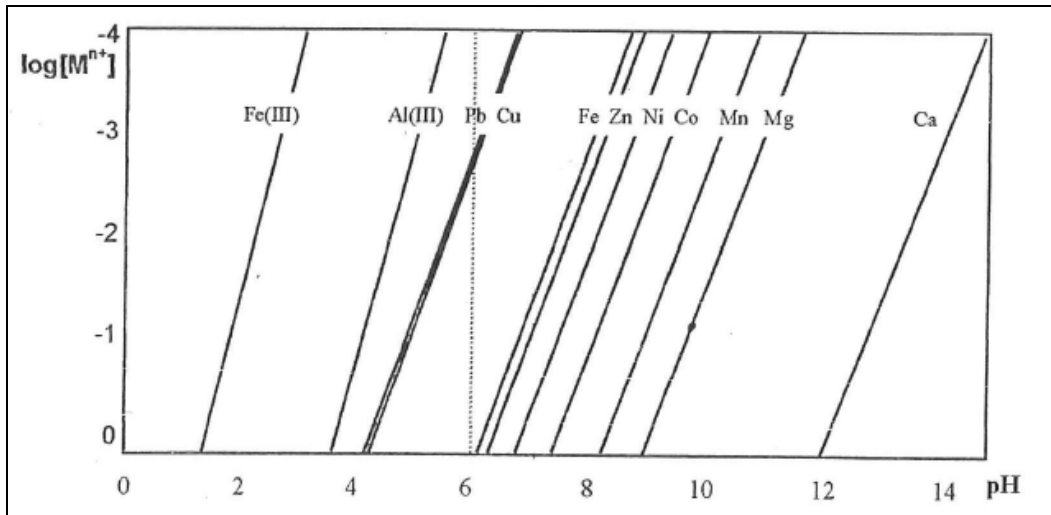


Figure 4 – Equilibrium Concentration of Metal Hydroxides<sup>(14)</sup>

In Figure 4, solution compositions that lie to the left of the vertical line will quantitatively be precipitated. In this case Fe(III) and Al(III), Cu(II) and Pb(II) will partially be precipitated with residual ionic concentrations of about  $10^{-3}$  mol per litre and all other metal ions to the right will be soluble.

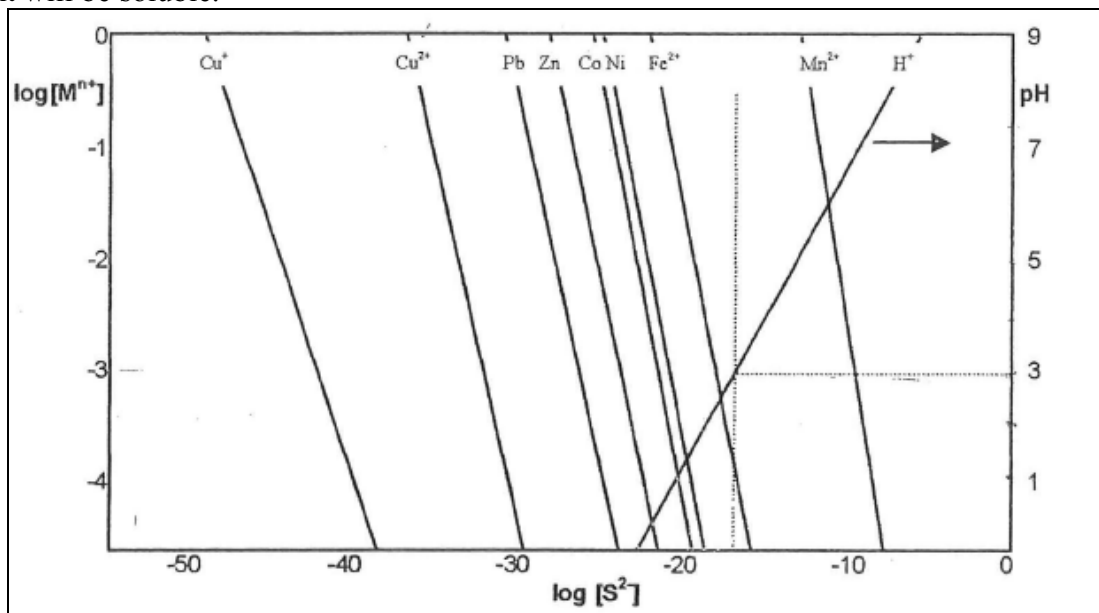


Figure 5 - Equilibrium Concentration of Metal Sulphide Precipitates<sup>(14)</sup>

The most common way to use the diagram in Figure 5 is with a horizontal dotted line at the solution pH, as illustrated. For a solution saturated with H<sub>2</sub>S at a pH of 3 (horizontal dotted line from right axis), the equilibrium is [S<sup>2-</sup>] is 10<sup>-17</sup> mol per litre, as shown by the vertical dotted line. This line intercepts that for Fe<sup>2+</sup> at a metal concentration a little above 10<sup>-4</sup> mol per litre (left axis), which is the solubility of Fe<sup>2+</sup> under these conditions. All of the metals ions to the left of the vertical line, such as Co<sup>2+</sup> and Ni<sup>2+</sup>, will be more or less insoluble at pH 3 while those to the right, such as Mn<sup>2+</sup> will be soluble.

### CURRENT PROCESS AT LUILU REFINERY:

#### SELECTIVE PRECIPITATION OPTION – PROCESS DESCRIPTION

Figure 6 illustrates the basic flow of the purification steps of the cobalt rich solution for the current Luilu flowsheet. A basic process description follows thereafter.

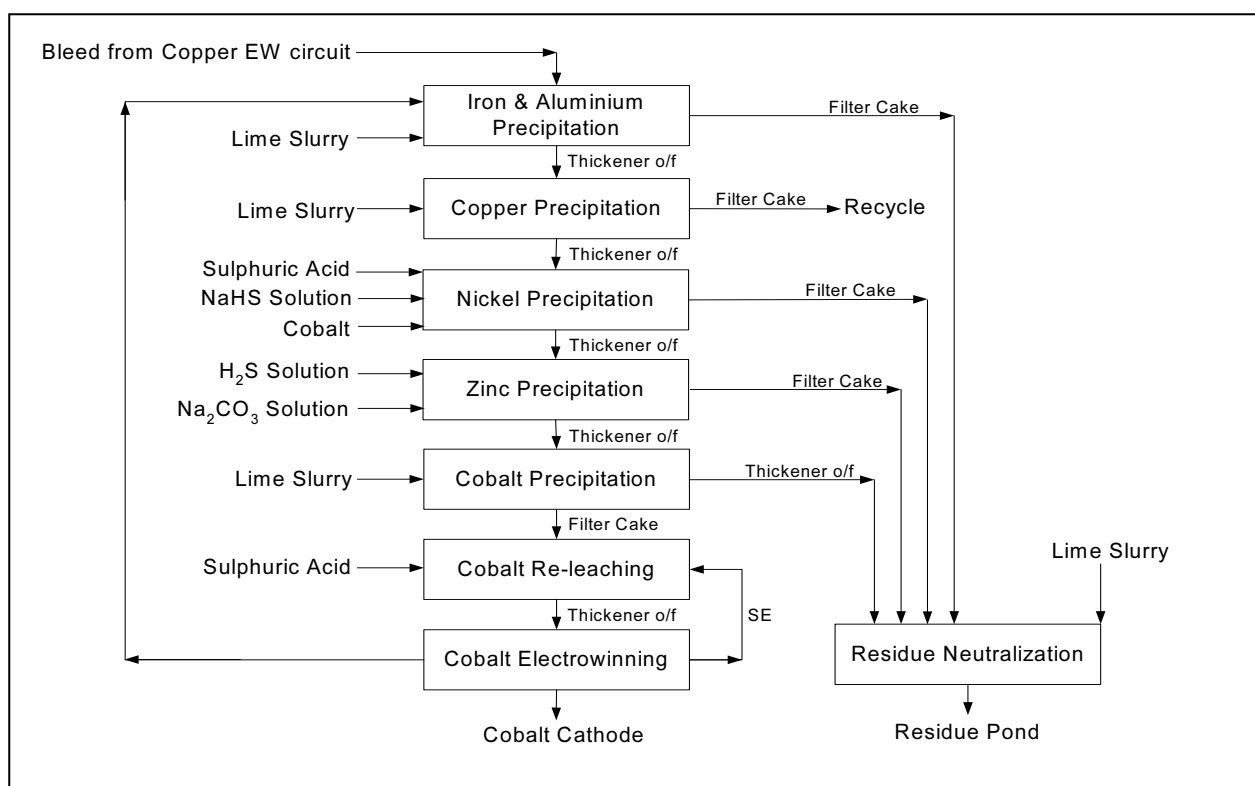


Figure 6 – Current cobalt recovery process at the Luilu<sup>(1)</sup>

#### Iron And Aluminium Precipitation

Spent electrolyte bleed from the copper electrowinning circuit is transferred to the iron and aluminium precipitation circuit. The process liquor contains mostly cobalt (25 to 30 g/L) and copper (5 to 10 g/L), which is removed in subsequent purification steps.

Lime slurry is added to precipitate the iron and aluminium as hydroxides. The slurry from the reactors is discharged to thickening and filtration.

### **Copper Precipitation**

Process liquor containing minimal amounts of iron and aluminium is fed to the copper precipitation circuit.

Copper removal is done in two and lime slurry is added to precipitate the copper as a hydroxide. The final copper concentration is about 1,5 to 1,0 g/L. The discharge slurry reports to thickening and filtration.

### **Nickel Sulphide Precipitation**

The copper precipitation discharge solution is fed to the nickel sulphide precipitation circuit.

Recycled cobalt powder and sodium hydrogen sulphide solution is used to precipitate the nickel as nickel sulphide. Acid is added to control the pH during the precipitation. The discharge slurry reports to thickening and filtration.

### **Zinc Sulphide Precipitation**

Process liquor containing minimal amounts of iron, aluminium, copper and nickel is fed to the zinc sulphide precipitation circuit.

Zinc removal is done with dissolved hydrogen sulphide gas. Sodium carbonate solution is added for pH neutralization and to prevent the reverse reaction of re-dissolving the zinc. After precipitation, the discharge slurry reports to filtration.

### **Cobalt Precipitation**

Cobalt precipitation facilitates the removal of soluble sodium and magnesium. The precipitation also delivers a high cobalt content intermediate that is re-dissolved in downstream processing to produce a higher cobalt tenor in the advance electrolyte.

Lime slurry is added to precipitate the cobalt as a hydroxide. The discharge slurry reports to thickening and filtration.

### **Cobalt Hydroxide Leaching**

Repulped cobalt hydroxide filter cake is transferred to the cobalt leaching reactors. Acid is added to leach all the cobalt into solution after which the discharge slurry reports to thickening and filtration.

### **Cobalt Electrowinning**

The purified advanced cobalt electrolyte reports to the tankhouse where the cobalt in solution is recovered by electrowinning. Electrowinning is done by using lead based anodes and stainless steel blank cathodes. Electrolyte depletion is avoided by plating cobalt to a final spent electrolyte concentration of about 30 g/L. The required current density is 310 A/m<sup>2</sup>.

## SOLVENT EXTRACTION – TEST WORK

Batch test work to investigate the allocation of solvent extraction to replace part of the cobalt purification circuit was conducted by Mintek<sup>(3)</sup>. The choice of the solvent extraction circuit used in the test work is only one of many possible circuits that could be used at the Luilu.

A sample of the cobalt circuit feed liquor was provided by Luilu. To obtain a solution suitable for performing solvent extraction test work, the first precipitation steps as found in the cobalt plant had to be performed, including iron and aluminium precipitation and copper precipitation with lime slurry.

Bench scale solvent extraction test work was performed on the resulting iron and copper free solution. The extraction and stripping isotherms were determined and used to establish the required number of extraction and stripping stages. This information was used to perform the trade off study, comparing the refurbishment cost and operating cost of the existing circuit versus the capital and operating costs required to operate the solvent extraction circuit.

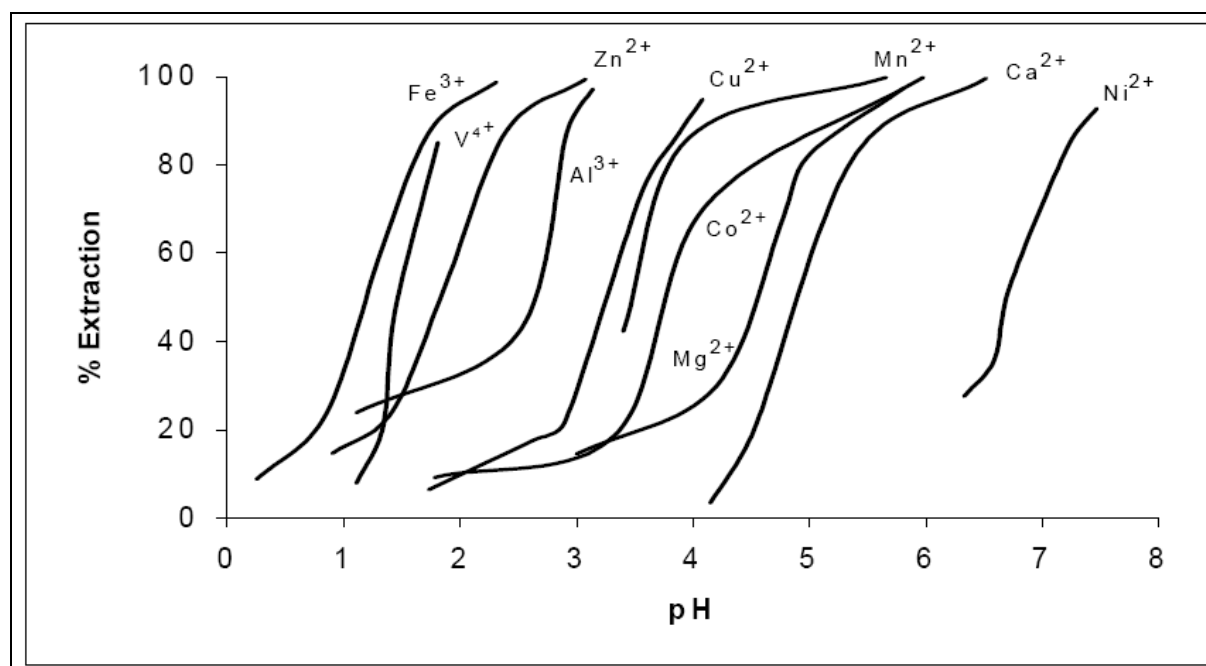


Figure 7 – Typical Cyanex 272 Extraction curves at different pH values<sup>(19)</sup>

Thirty percent (v/v) Cyanex 272 in SSX 210 diluent was used to load cobalt and zinc, with nickel and other impurities reporting to the raffinate at 35°C. The extraction isotherm (Figure 7) indicated that extraction could be conducted within a pH range of 4.8 to 5.1. An extraction isotherm was conducted at a pH of 5 (Figure 8) and the McCabe-Thiele diagram indicated that about 20 g/L Co could be extracted in five stages, to a concentration of less than 10 mg/L in the raffinate. An O:A ratio of 1.33:1 was used.

Batch counter-current extraction test work was then conducted and confirmed that five extraction stages are required to produce a raffinate of less than 10 mg/L.

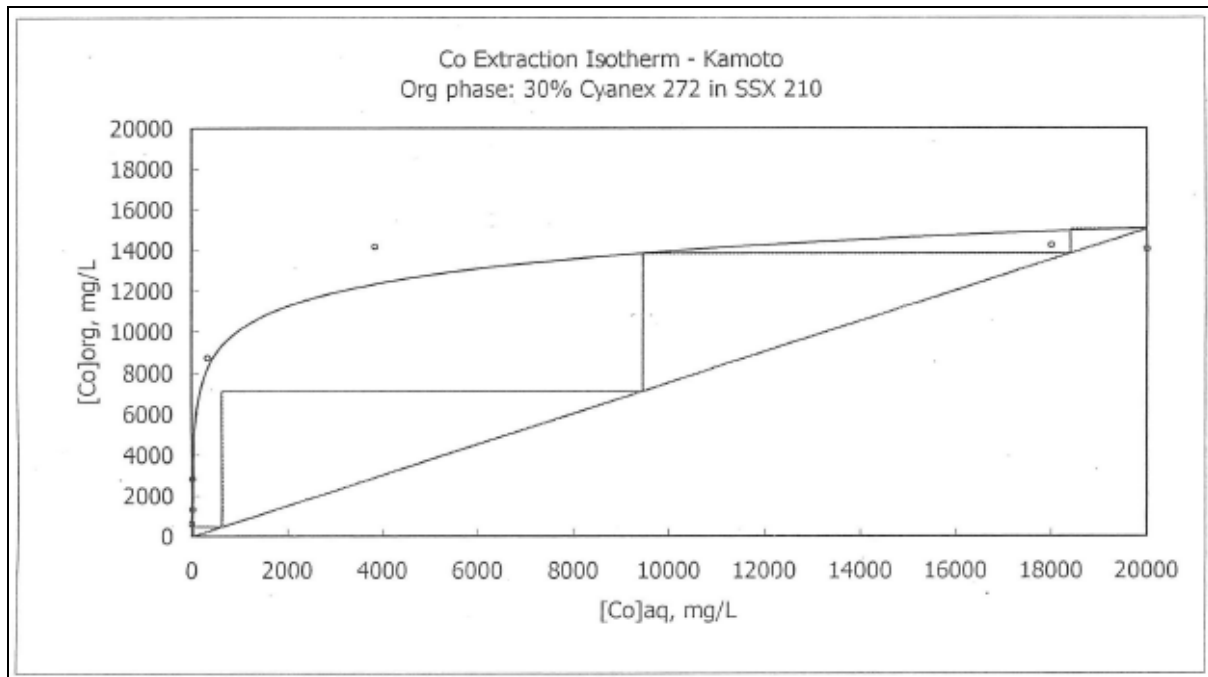


Figure 8 – Cobalt Extraction Isotherm (O:A = 1.33)<sup>(3)</sup>

Scrubbing was not investigated due to the high O:A ratios required. This can only be optimised on the full-scale pilot plant.

Stripping test work was conducted with the loaded organic generated from the bench scale counter-current extraction test. A stripping isotherm (Figure 9) indicated that selective stripping of Co would be optimal at a pH of about 2.8 in every stage. A McCabe-Thiele diagram (Figure 10) showed that three strip stages would be required for the stripping of Co from a scrubbed organic containing about 15 g/L Co. An additional Zn stripping stage is required to strip the remaining Zn from the organic at a much lower pH of 1.0.

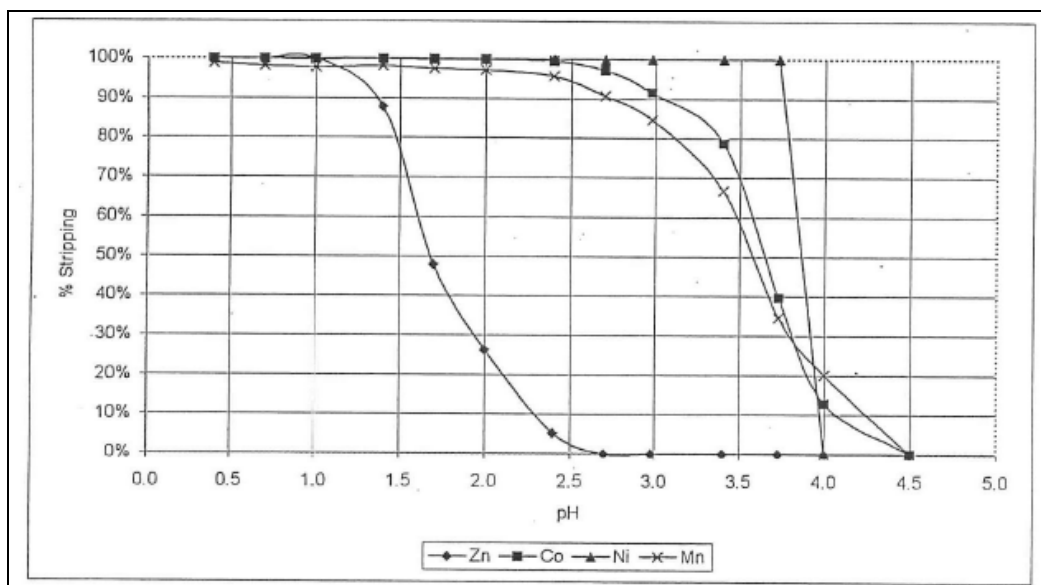


Figure 9 – Stripping Curve<sup>(3)</sup>

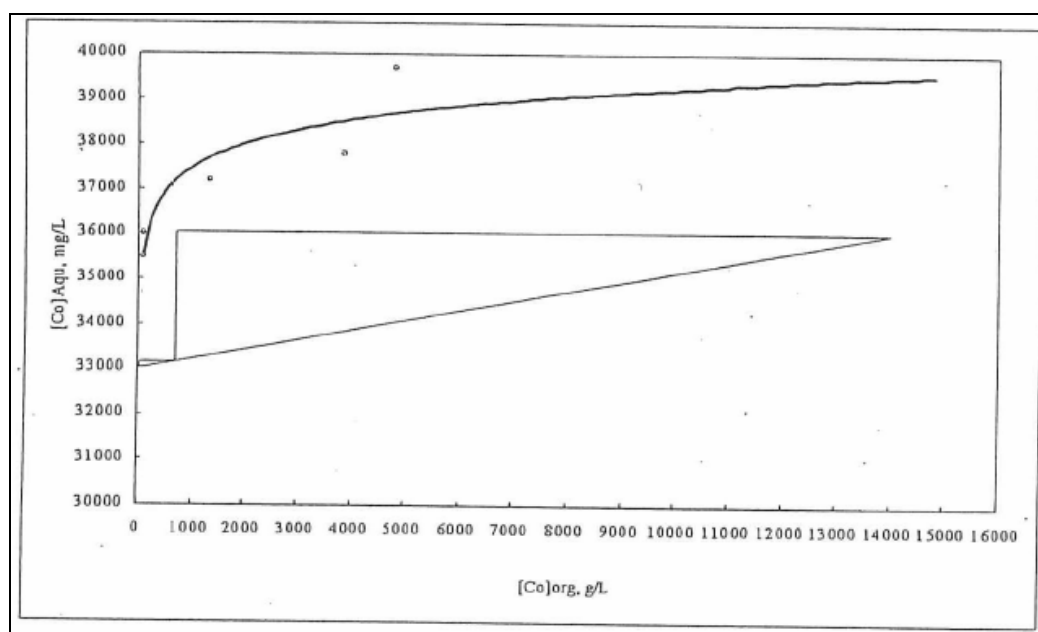


Figure 10 – Cobalt Stripping Isotherm<sup>(3)</sup>

### SOLVENT EXTRACTION OPTION – ONE OF THE PROPOSED CIRCUITS

Figure 11 illustrates the basic flow logic of one of the proposed solvent extraction circuits to purify the cobalt rich solution stream<sup>(2)</sup>. This concept was used as the basis for the cost estimate.

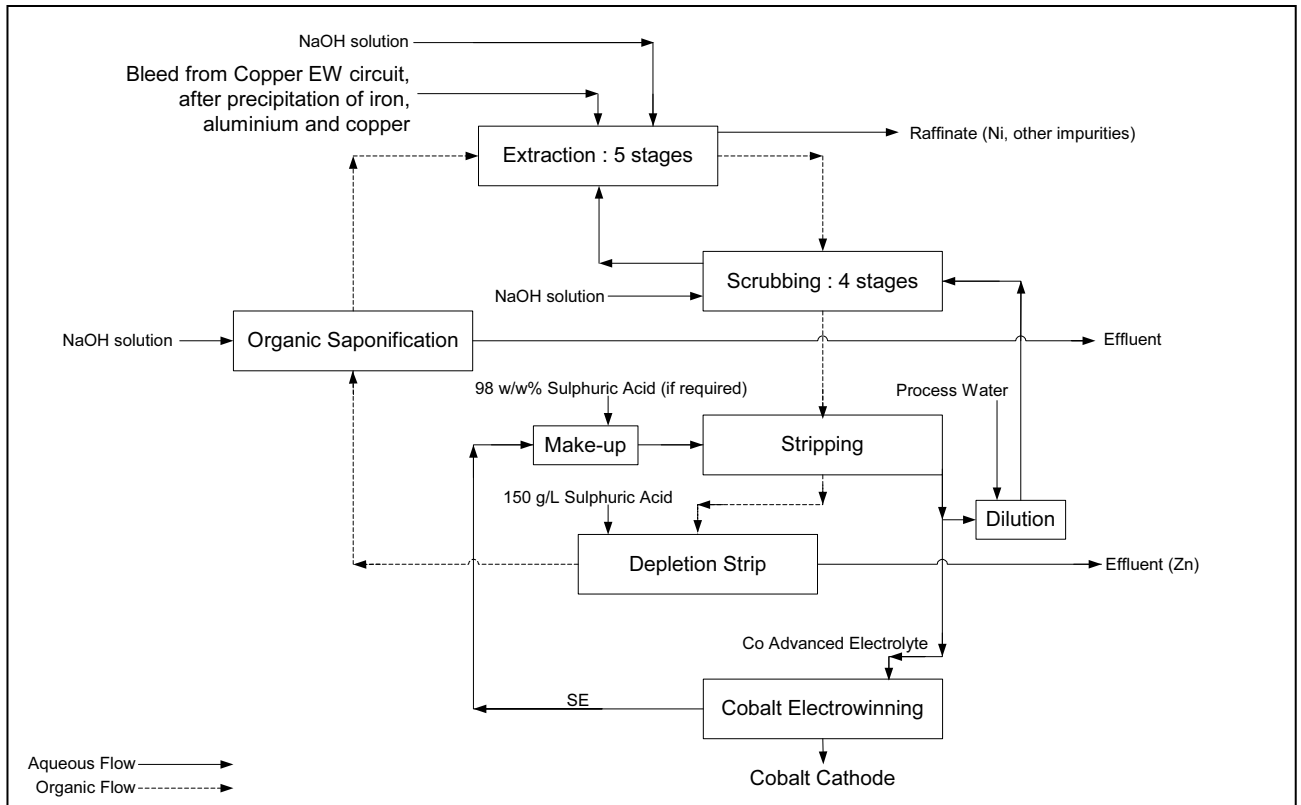


Figure 11 – One of the proposed Solvent Extraction circuits for the purification of cobalt<sup>(2)</sup>

### Saponification

Barren organic is contacted with sodium hydroxide solution in the saponification mixer/settler to exchange hydrogen ion (from the organic phase) with sodium cation. This saponification step is incorporated in the design to simplify pH control in the extraction stages, which operate in the tight range of pH 4.8 to 5.1.

### Extraction

Iron, aluminium and copper free solution reports to the cobalt solvent extraction circuit where the pH is adjusted with sulphuric acid to the required set point.

The saponified organic reports to the cobalt extraction circuit where it is contacted counter-currently with the aqueous solution in five extraction stages. The cobalt and zinc in the aqueous solution is transferred to the organic phase, whilst the nickel and other impurities remain largely in the aqueous phase. The aqueous phase leaving the last extraction stage (termed the cobalt SX raffinate) contains virtually no cobalt and zinc and reports to residue waste disposal.

### Scrubbing

The organic phase from the extraction circuit, now loaded with cobalt, zinc and other impurities, passes to a scrubbing circuit. Here it is contacted counter-currently with diluted cobalt advanced electrolyte solution in four stages, in order to displace any co-extracted nickel and other impurities from the loaded organic phase with cobalt.

### **Stripping**

The scrubbed, loaded organic phase now passes to the first of three stripping stages. The required acid solution for cobalt stripping is made up from spent cobalt electrolyte. In these stages, the cobalt is selectively stripped of the organic. The remainder of the impurities (any zinc) are removed in the depletion strip stage to prevent a build up that will cause lower organic extraction efficiencies. The depletion strip solution consists of about 150 g/L sulphuric acid. The cobalt rich solution from the stripping stages reports to the organic removal circuit. The barren organic from the depletion strip stage reports to the barren organic tank that feeds the saponification stage. The depletion strip end solution has to be bled off to residue waste disposal.

### **Organic Removal and Crud Treatment**

The cobalt rich strip solution contains entrained and dissolved organic content, which must be removed prior to entering the cobalt electrowinning circuit. Provision is made in the cost estimate for the equipment required for the removal of entrained and dissolved organic. The organic free solution reports to the cobalt electrowinning circuit.

Spillage from the cobalt solvent extraction circuit will be captured in various sumps and report to a crud treatment circuit. The main purpose of the crud treatment circuit will be to recover organic and aqueous solution from spillage, as well as remove any entrained solid crud and separate the solutions into two distinct phases for re-introduction into the process.

## **COST ESTIMATE**

Based on in-house data and design criteria derived from the Mintek test work, a solvent extraction circuit for the selective stripping option was sized and an equipment list was compiled. A capital estimate<sup>(2)</sup> based on in-house data was developed to an accuracy of  $\pm 20\%$  (Table 3). The estimated capital required for the solvent extraction facility was compared to the capital cost required to refurbish the following circuits during all four phases as estimated during the feasibility study.

- Nickel and Zinc Sulphide Precipitation
- Cobalt precipitation
- Cobalt Leaching
- Clarification

The direct and indirect costs are summarized in Table 2, and a contingency of 15% was applied:

<b>SUMMARY OF CAPEX FOR TOTAL PRECIPITATION CIRCUIT</b>	<b>TOTAL COST</b>
Nickel and Zinc Removal	\$1,350,000
Cobalt Precipitation	\$5,320,000
Cobalt Leaching	\$1,250,000
Clarification	\$310,000
<b>SUBTOTAL (Direct and Indirect Costs)</b>	<b>\$8,230,000</b>
Contingency (15 %)	\$1,234,500
<b>TOTAL CAPEX</b>	<b>\$9,464,500</b>

Table 2 – Summary of Capital Cost per Area required for refurbishing the Existing Precipitation Processes

For the solvent extraction capital cost an allowance for EPCM with additional test work is included. The organic first fill was estimated and a contingency of 15% was applied:

<b>SUMMARY OF CAPEX FOR SOLVENT EXTRACTION CIRCUIT</b>	<b>TOTAL COST</b>
<b>SUBTOTAL (Direct and Indirect Costs)</b>	<b>\$4,320,000</b>
Organic First Fill	\$6,370,000
Contingency (15 %)	\$1,603,500
<b>TOTAL CAPEX</b>	<b>\$12,293,500</b>

Table 3 – Summary of Capital Cost required for Solvent Extraction

The operating costs<sup>(2)</sup> related to the precipitation process versus the solvent extraction option were developed using the feasibility phase 4 mass balance information (Tables 4 and 5).

<b>SUMMARY OF OPEX FOR TOTAL PRECIPITATION CIRCUIT</b>	<b>TOTAL COST / YEAR</b>
Nickel Removal	\$60,000
Zinc Removal	\$5,860,000
Cobalt Precipitation	\$2,540,000
Cobalt Leaching and Clarification	\$1,460,000
<b>SUBTOTAL OPEX</b>	<b>\$9,920,000</b>
Labour	\$310,000
Maintenance	\$330,000
Contingency (15%)	\$1,584,000
<b>TOTAL OPEX</b>	<b>\$12,144,000</b>

Table 4 – Summary of Operational Costs for the Precipitation Processes

<b>SUMMARY OF OPEX FOR SOLVENT EXTRACTION CIRCUIT</b>	<b>TOTAL COST / YEAR</b>
<b>SUBTOTAL OPEX</b>	<b>\$23,500,000</b>
Labour	\$270,000
Maintenance	\$130,000
Extra costs related to increased production	\$790,000
Contingency (15%)	\$3,703,500
<b>TOTAL OPEX</b>	<b>\$28,393,500</b>

Table 5 – Summary of Operational Costs required for Solvent Extraction Option

Based on calculations and solvent extraction operational experience, it is expected that a 9% increase in cobalt recovery can be obtained when using the selective stripping solvent extraction circuit option versus the current precipitation circuit<sup>(2)</sup>. An IRR calculation was completed (Table 6) using the expected increased cobalt production numbers, the increase in operating costs, and the estimated solvent extraction circuit capital cost.

ITEMS	VALUE
Increased Cobalt Revenue / year	\$26,740,000
Net Present Value (NPV)	\$41,530,000
Discount Rate	15%
Internal Rate of Return (IRR)	85%

Table 6 – Summary of IRR Calculation Results based on a 9% increase in Cobalt recovery

Comparing the estimated capital and operating costs for the two process options, it can be concluded from Table 6 that it is economically attractive to replace the existing precipitation circuits with a solvent extraction circuit.

Comparing Tables 2 and 3 it can be concluded that the capital cost required to refurbish the precipitation circuits is lower than the capital expenditure required for implementing a solvent extraction circuit. One major contributor to the higher capital required for the solvent extraction circuit is the cost related to the first fill of organic.

From Tables 4 and 5 it can be observed that the operating cost of the solvent extraction circuit is higher than the existing precipitation process. This is mainly attributed to high reagent cost. However, when the increased cobalt recovery that can be obtained with the implementation of solvent extraction technology is considered, it is clear that a swift return on investment can be achieved. With the anticipated increase in the impurity removal efficiency, an improvement in the overall cathode quality can also be expected thus possibly a higher return for cobalt cathode.

Other solvent extraction circuits should be evaluated and additional cost estimates and test work should be developed for the various options available. Some of these various options include the following possibilities<sup>(12)</sup>:

- Selective extraction of zinc and cobalt with an impurity removal solvent extraction circuit with D2EHPA, followed by a cobalt solvent extraction circuit with CYANEX 272 (Kasese Cobalt Company Ltd. (KCCL) in Uganda, Knightsbrigde Cobalt in South Africa, Kakanda Tailings Project in the DRC);
- Selective extraction of zinc and cobalt with an impurity removal solvent extraction circuit with CYANEX 272, followed by a cobalt solvent extraction circuit with CYANEX 272 (Kolwezi Tailing Project in the DRC);
- Impurity removal solvent extraction circuit with D2EHPA followed by cobalt precipitation technology (Chambishi Metals Plc in Zambia);
- Incorporating ion exchange in addition to solvent extraction as impurity removal steps (Chambishi Metals Plc in Zambia, Kolwezi Tailing Project in the DRC, Kakanda Tailings Project in the DRC);

Evaluating other solvent extraction circuits will ensure that the selected circuit will be robust enough to produce on spec metal at a high recovery. It should be noted that a continuous pilot plant campaign is required on the most likely option to confirm the assumptions made.

### **FURTHER CONSIDERATIONS**

Any new technology that is implemented requires extensive training and initial technical and operational support. With the implementation of new technologies in Africa a few additional aspects should be considered which could influence the final success. Some of these include the availability of consistent technical support and spares that could be challenging in these remote areas. Another aspect, which is not only an issue in Africa, is the pressure placed on production performance due to tight ramp-up schedules and mining contracts that allow less time for trial periods.

Selective stripping is a technology that requires very strict control. In the discussed case where zinc and cobalt are selectively stripped from the loaded organic, extensive pilot runs are required to determine the correct number of stripping stages needed to ensure an efficient separation of the two metals. According to the stripping curve (Figure 9), cobalt will typically be stripped first at a lower acid concentration, followed by the stripping of zinc at a higher acid concentration. Very strict volume and pH control over the stages are required to avoid the formation of localized low pH areas that will strip the zinc prematurely into the aqueous phase instead of cobalt. In addition, sufficient stripping stages should be allowed for any stripped zinc to be loaded back onto the organic phase.

There are a few pilot plant campaigns that have successfully run selective stripping as a separation method for zinc and cobalt. Some of these include the Demonstration Phase 2 Pilot Plant campaign at SGS Lakefield Research that treated ore from Baja Mining's Boleo property, Baja California Sur, Mexico. Positive results for the DSX (Direct Solvent Extraction)<sup>(15)</sup> were obtained from their pilot plant campaign. Also the Platsol Pilot Plant campaign showed promising results<sup>(16)</sup>. These pilot plants had a feed matrix that may have been different from the DRC originated Luilu matrix and it may not be directly applicable.

Solvent Extraction is a feasible improvement in terms of enhancing the cobalt recovery and purity, however the use of separate zinc and cobalt solvent extraction circuits might give this process the robustness required to operate in remote areas.

Should the sensitivities surrounding selective stripping amount to an unacceptable risk, it is possible to consider a flowsheet that incorporates selective extraction of zinc and cobalt. It is acceptable, even conservative, to assume a flowsheet of this nature requires double the capital investment of the selective stripping flowsheet because:

- The size of the mixer-settlers is normally directly proportional to the aqueous feed flow, resulting in the mixer-settlers in both options being more or less the same in size.
- Selective extraction requires that independent extraction, scrubbing and stripping steps are employed for each metal.

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- The number of stages in each step will be the same or slightly less than that required in each of the selective stripping steps.
- Therefore, the number of mixer-settlers in the selective extraction flowsheet will be the same or slightly less than that required in the selective stripping flowsheet.

Should a conservative approach be taken to re-assess the financial analysis, by assuming the capital costs will be double, and the same cobalt recovery is obtained, the internal rate of return is found to be 42%. This implies that even if a conservative design is employed, solvent extraction will likely offer a significant financial benefit to the refinery.

## **ADDITIONAL CHALLENGES**

The majority of the research done for the implementation of cobalt solvent extraction circuits up to date is based on typical nickel laterite hydrometallurgical applications. These solvent extraction circuits are designed to separate cobalt and nickel, and the streams typically contain high concentrations of both cobalt and nickel in solution.

The successful implementation of cobalt solvent extraction circuits to separate cobalt from nickel is dependent on a few factors and some of these include the type of impurities present in the solution such as cadmium, zinc, nickel, calcium and iron<sup>(17, 18)</sup>. This will have a impact on the operating conditions and overall design of the solvent extraction circuit. Cawse, Bulong and Murrin-Murrin in Western Australia have faced many challenges involving the successful long-term operation of their cobalt solvent extraction circuits, but only limited lessons from this can be carried over to the Luilu application<sup>(17, 18)</sup>.

The cobalt rich solution at Luilu contains a high cobalt concentration but lower impurity concentrations such as nickel and zinc. This is typical to the Copperbelt district, however very different from the applications within the typical nickel laterite circuits. Typical Copperbelt applications such as Chambishi Metals Plc in Zambia, Kolwezi Tailings and Kakanda in the DRC are examples of solvent extraction circuits that underwent extensive pilot plant campaigns to fully understand the complexity that exists in the successful implementation of the solvent extraction circuits.

Chambishi operates a full-scale impurity removal solvent extraction circuit using D2EHPA to remove zinc and other impurities. The purified solution undergoes cobalt precipitation to concentrate the cobalt in solution before electrowinning. A nickel ion exchange operation is also implemented before cobalt electrowinning. Kolwezi Tailings utilizes two solvent extraction circuits, both using CYANEX 272. The first circuit is used for zinc removal and the second circuit for cobalt concentration enhancement. A copper/zinc ion exchange operation is implemented between the two solvent extraction circuits to ensure proper removal of these metals before cobalt electrowinning. Kakanda incorporates an impurity removal solvent extraction circuit with D2EHPA and a cobalt solvent extraction circuit with CYANEX 272 together with a copper/zinc ion exchange operation to ensure a suitable cobalt advance electrolyte feeding cobalt electrowinning.

From these examples it can be concluded that even though the flowsheets are in the same general area on the Copperbelt, the circuits are different and totally customized for each application to ensure adequate quality cobalt cathodes are produced as the final product.

### **CURRENT STATUS**

The project is currently finalising the implementation of Phase 1 of the rehabilitation programme and is due to start up in the second half of 2007. It was decided to defer the implementation of a solvent extraction circuit to a later phase in the rehabilitation programme due to sampling and timing issues. The sampling that was done for this study can only be considered as preliminary and indicative, since sampling was done just prior to shutting down of facility. The process was not running at steady state and the matrix of materials received was not representative of the expected material during operations. This provides a risk when conducting solvent extraction test work to design an appropriate circuit. Additional pilot scale testing will commence with the Phase 1 start-up, and further detailed trade-off studies, test work and design for the implementation of solvent extraction will follow. The timing requirement is approximately 18 to 24 months to finalize an appropriate circuit, to conduct a detailed design for implementation followed by construction and commissioning.

With the benefits of a solvent extraction circuit being incorporated into the project plan, the investment scenario could be improved even further. Hence, Katanga Mining Limited is considering the implementation of a solvent extraction circuit at the Luilu Refinery in the near future.

### **CONCLUSIONS**

Katanga Mining Limited has undertaken to rehabilitate Luilu in the Democratic Republic of the Congo over a period of phases. The cobalt circuit will be rehabilitated to its full nameplate capacity during this time. The current flowsheet employs the use of selective precipitation to refine the cobalt electrolyte, and selective precipitation is found to result in less than optimal cobalt recoveries.

Solvent extraction is a technology that has the ability to produce a pure cobalt electrolyte, with improved cobalt recoveries compared to selective precipitation. Although the capital and operating costs associated with a solvent extraction circuit are considerably more than is required to rehabilitate and operate a selective precipitation circuit, the potential improvement in cobalt recovery and cathode quality is a significant factor and presents an attractive financial proposition.

As a result of these findings, we recommend that Katanga Mining Limited should at a later stage consider the implementation of a solvent extraction circuit to refine the Luilu cobalt electrolyte.

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