



Refinery Process and Plant Description

KCR-1000-PR-CRT-002 | October 2024

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DOCUMENT DESCRIPTION

This document defines the process description for the Kwinana Cobalt Refinery project for the purpose of detailed engineering study.

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1 PROCESS DESCRIPTION

1.1 Feed Material

The refinery can process a range of cobalt intermediates. In the current market (2024), supply of precipitated hydroxide and sulphide intermediates is available, with the expectation that in the longer-term products from battery recycling facilities (e.g. black mass) may also be supplied. It may also be possible to process metals into sulphates, if the economic conditions are favourable.

Mixed Hydroxide Precipitates (MHP)

The refinery will purchase cobalt hydroxide or cobalt-nickel hydroxides. These can be supplied from copper operations (e.g. DRC) or from nickel operations (e.g. Indonesia, Philippines). The cobalt hydroxides typically contain 35-45% cobalt. If the BHCP project is developed, then COB will supply a cobalt-nickel hydroxide to the refinery containing 25-35% cobalt and 5-10% nickel.

Mixed Sulphide Precipitates (MSP)

The refinery will purchase cobalt sulphides. These are produced as by-products from nickel (e.g. BHP) and copper refining operations. The cobalt content can be 25-45%, and the nickel content 10-25%.

General Feed Characteristics

The feed material has the following characteristics:

- Originate as precipitated artificial minerals with a small particle size ($p_{80} < 10 \text{ um}$), however these may cake into larger lumps during transport and storage
- Bagged as damp filter cake (10-50% free moisture), but can be delivered with much lower free moisture if evaporation has occurred during transport and storage
- Generally contain trapped process liquor due to poor washing (e.g. the process liquor may contain ammonium sulphate)
- Generally contain unreacted precipitation reagents (e.g. magnesia, lime, sodium sulphide)

COB purchase 5 tonne lots of MHP and MSP, for evaluation in the demonstration plant. The feed grades are listed in Table 1.

Table 1 - Typical Feed Grade Received of MHP & MSP

Feed	Unit	MSP	MHP
Cobalt	%	35	40
Nickel	%	22	0
Manganese	%	3	2
Zinc	%	0.5	0.5
Calcium	%	-	-
Sodium	%	-	-
Magnesium	%	-	-



Figure 1 - Showing received MHP Bulka Bag and Feed Sample



Figure 2 - Showing received MSP Bulk Bag and Feed Sample

1.2 Process Flowsheet

The refinery block flow diagram is shown below in Figure 3.

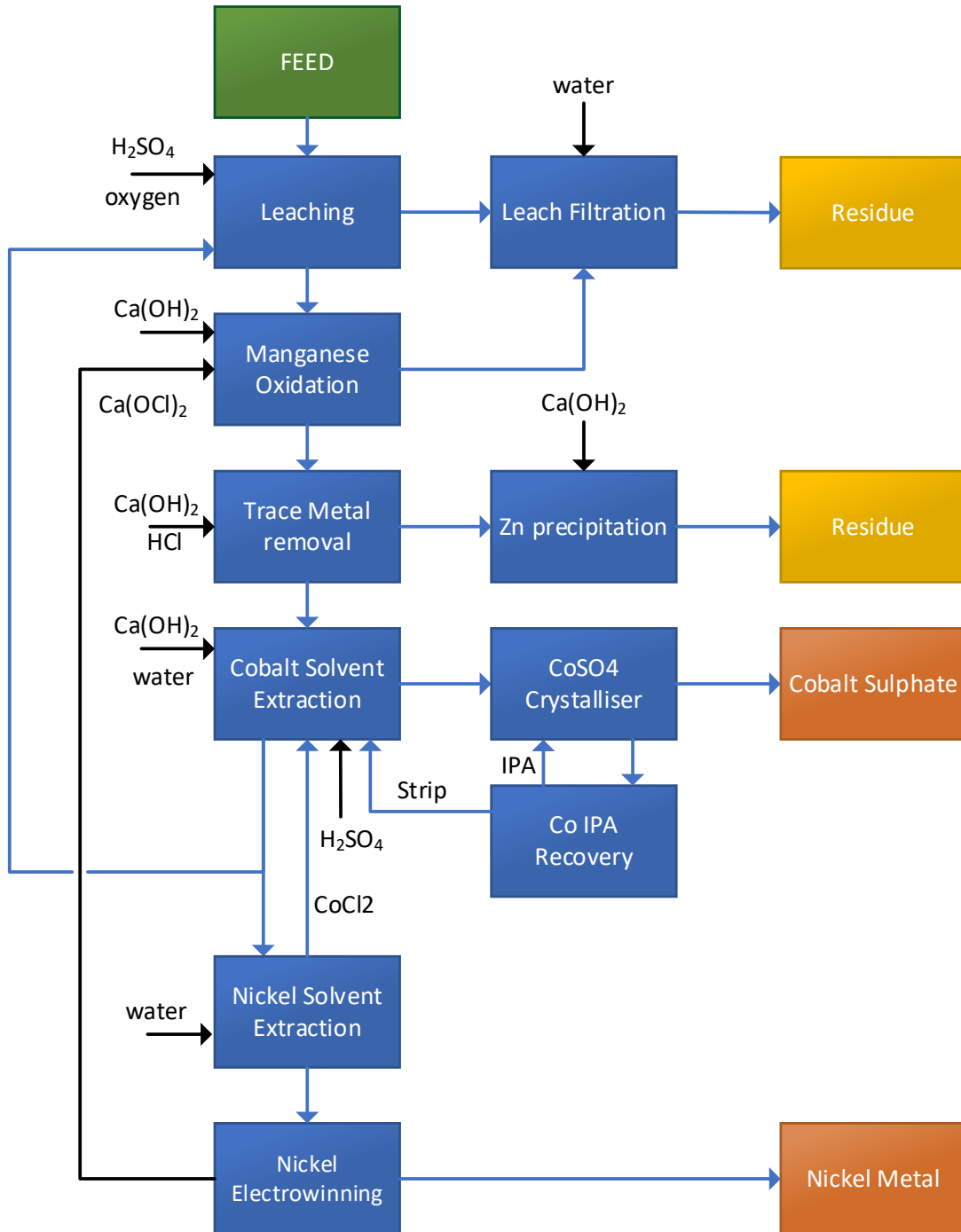


Figure 3 - Shows the KCR Block Flow Diagram

1.3 Area 100 Leaching

1.3.1 Area 110 Materials Handling and Storage

Feed will be delivered to site in 1 t bulka bags. This will contain at least 600 kg of dry feed, but may be up to 1 t wet weight. Imported material will be in sea containers, and local supply will be on flat-bed truck or similar (from BHP).

Onsite storage for containers and racks for unloaded bags will be included in the plant footprint. A 30 t forklift will unload/load containers from trucks, and bring containers to the edge of the plant building for unloading. A 3 t forklift will unload bags into the plant, and store them on racking for the operations.

Bags will be retrieved from storage and moved to the feed station. Bags will be lifted using an overhead hoist, and discharged into the respective entry points. A single hopper with four entry points will hold approximately 12 m³ of storage, which is equivalent to around 24 t of dry feed hoppers. The nominal total storage after hopper filling will be 24 hour of operation. Fresh material can be added to the hoppers on a regular basis, such as once per 12 hour shift.

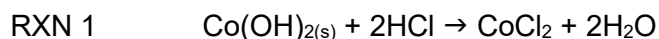
The hopper will have a single discharge system in the base. A series of parallel screws will meter the solids into the plant. The feed rate will be monitored using a weightometer (or similar). A swirl hopper (FRP) will combine feed solids with process liquor to the desired density. The resulting slurry discharges through the base of the swirl hopper into a pump for transfer to the first leach reactor.

The hopper feed entry points and the swirl hopper will be connected to a scrubber to minimise dust emissions. Any spilt feedstock will be washed to a sump and returned to the plant.

Empty bags will be disengaged from the hoist and placed directly into a chute connecting to a bag compactor. The compacted bags will be periodically removed from the plant and sent to landfill.

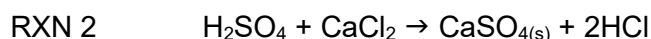
1.3.2 Area 120 Acid Leach

The Acid Leach circuit consists of two 10 m³ agitated tanks (FRP) in series for the partial dissolution of MHP and MSP to extract cobalt and nickel into solution. MHP is leached using hydrochloric acid, with a generic reaction given in RXN 1:



The leaching of the MHP is rapid and near complete dissolution of the hydroxide is achieved at a pH of 0.5-1, and temperatures >70 °C (in 120-TK-002). Minor metal content such as iron, manganese, magnesium and calcium will be solubilised along with cobalt and nickel. A portion of the iron and manganese may precipitate, as the preferred operating pH is close to the stability pH for ferric ions and manganese is often in an elevated oxidation state in the MHP. Remaining iron is removed in the subsequent oxidative leach step. An intertank recycle stream is used to send the slurry from Acid Leach Tank 2 (120-TK-002) back to Acid Leach Tank 1 (120-TK-001), where the pH is nominally 2.5-3.5. This system provides a high level of acid utilisation. The recycle rate is adjusted to achieve the desired pH differential between the two tanks.

Technical grade sulphuric acid is dosed into the Acid Leach Tank 2 (120-TK-002) to generate in-situ hydrochloric acid, according to RXN 2:

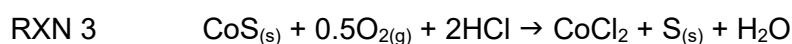


The precipitated calcium sulphate, and unreacted feed material is stored in a stock tank (120-TK-003), and then advanced to the oxidation leach reactors (130-PU-001).

Heating is provided in the leach tanks using titanium coils, fed with hot glycol solution (120 °C).

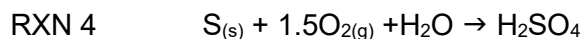
1.3.3 Area 130 Oxidation Leach

In the Oxidation Leach, the solids are leached using oxygen, with a generic reaction shown in RXN 3:



The mechanism of oxygen transfer is via $\text{Fe}^{2+}/\text{Fe}^{3+}$ couples. Typically the iron concentration is <5 g/L.

While the oxidation reactions can occur at atmospheric conditions, it is preferable to operate just above the melting point of elemental sulphur (operating at 130-140 °C versus sulphur melting temperature of 118 °C) to enhance the oxidation of elemental sulphur into sulphate. This generates in-situ acid, along with significant heat which is used to help maintain the plant wide water balance. The elemental sulphur oxidation is given in RXN 4:

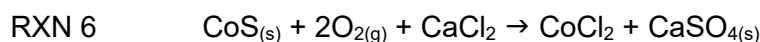


In turn, the sulphuric acid is converted into hydrochloric acid, due to the presence of calcium in the background brine, as given in RXN 5:



Importantly, the precipitated calcium sulphate is anhydrite (no waters of crystallisation) as opposed to gypsum. This occurs since the reaction is operated at temperatures > 100 °C. The anhydrite is readily filterable.

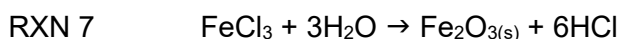
Hence the overall reaction for the leaching of MSP at temperatures of 130-140 °C, and with complete oxidation of the elemental sulphur to sulphate, is given by RXN 6:



The Oxidation Leach is preceded by the Acid Leach for two reasons. Firstly, due to variations in feed grades or to minimise oxygen consumption, not all of the sulphur will be oxidised to sulphate. The acid short fall (for the transfer of oxygen via the iron couple) can be made up with residual acid from the Acid Leach. Secondly, it is thermodynamically more efficient for the plant water and heat balance to utilise the heats of reaction in the Oxidation Leach to increase the process slurry from 80-85 °C exiting the Acid Leach, up to 130-140 °C in the Oxidation Leach, than applying external heat (e.g. steam/electrical).

The Oxidation Leach circuit consists of four 2 m³ reactors (Ti grade 2), which are operated in series with slurry cascading by gravity. The natural pressure of the slurry at 130-140 °C is approximately 2 bar. Hence, the total operating pressure of 6-8 bar is achieved by supplying overpressure from the oxygen gas. A VPSA plant will supply 92-93% oxygen and 7-8% nitrogen. Each reactor is connected to a common gas vent line, where the overall pressure is controlled by a single valve.

The leach is operated within a pH range of 0.5-2. This is controlled by varying the oxygen addition and thus extent of reactions. At the pH > 1.5 a large portion of the iron will precipitate according to RXN 7:



After the four reactors, the slurry is cooled down below the process liquor boiling point of < 108 °C via an inline shell and tube heat exchanger (130-HX-001, Ti Gr2). This reduces the amount of 'flashing' during discharge of the slurry back to atmospheric pressures, and also ensures that any elemental sulphur droplets are frozen and cannot accrete in valves or pipes.

The cooled slurry is discharged from the autoclave using a stepper pinch valve (130-CV-001), which controls the weight of a buffer vessel between the heat exchanger and the valve. The valve is mechanically controlled via a SCADA control loop, and the lining is EPDM (or SBRT). The discharged slurry is collected in a 2 m³ storage tank (130-TK-001).

The overall leach solids and liquor rates are controlled to target a cobalt tenor of 30-40 g/L in the solution advancing to Area 150. The expected nickel concentration will be approximately 6-8 g/L. If the Refinery cobalt production rate is increased, then the target metal concentrations in the leach circuit will be increased, and thus reduce the need for additional equipment to be installed.

The background salt concentrations are maintained around 150 g/L NaCl and a minimum of 150 g/L CaCl₂ in the leach. Make up salt can be added directly into the Acid Leach tanks, using the same system as the Feed solids.

1.3.4 Area 140 Leach Filtration

The slurry is forwarded from the storage tank (130-TK-001) to a 20 m³ Leach Thickener (140-TH-001). The slurry will pass through an inline mixing tank for the addition of floc. Flocculant (0.1% solution of Rheomax DR1010) is dosed at 5-10 g per tonne of leach residue. The thickener underflow is continuously transferred to a 20 m³ (FRP) Filter Feed Tank (140-TK-001), with the thickener overflow transferred by gravity to the 5 m³ Leach Thickener O/F Tank (140-TK-002). The thickened slurry is approximately 45-55% wt/wt density.

Slurry from the Leach Filter Feed Tank (140-TK-001) is fed to a horizontal belt filter with a four-stage counter-current wash to be applied. This reduces losses of cobalt and nickel, along with salt brine, to the residue. Hot water (60-80 °C) is used as primary wash solution. A cobalt free brine wash may also be applied to dilute the incoming mother liquor, and further reduce soluble losses to the cake.

The belt filter cake discharge, retaining 15-20% wt/wt moisture, is further dried in a heated screw to remove moisture to a target of < 3-5% wt/wt. The 'dry' solids are then stored in skip bins for disposal. The main component of the residue is anhydrite, with minor amounts of hematite (this is the main outlet for iron from the plant). Filter wash waters are collected and blended back into the plant via reagent make-up (hydrated lime), or other duties via the Leach Wash Water Recycle Tank (140-TK-004). The filtrate is returned back to the acid leach.

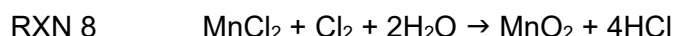
1.3.5 Area 150 Manganese Oxidation

The leach liquor stored in Leach Thickener O/F Tank (140-TK-002) is transferred to a series of reactors for manganese oxidation and precipitation as manganese dioxide. If present in the leach liquor, iron and copper are also precipitated in this step. Hypochlorite is used to oxidise the manganese. This is sourced from the nickel electrowinning circuit, or a standalone chlorinator unit.

The circuit consists of two 5 m³ (FRP) reactors in series. These are fitted with small disc and donut columns for scrubbing of the gases in the reactor headspaces. A recirculation loop on each reactor provides the scrubbing solution. The exit gas from the second reactor moves counter-currently to the first reactor. The exit gas from the first reactor goes to the main plant scrubber. In this fashion, the potential for release of free chlorine to the main scrubber is minimised.

The circuit is operated at a pH of ~3 (by addition of excess lime with the hypochlorite) to ensure oxidised iron and copper are also precipitated, and to remove excess acid advancing to the ion-exchange resin circuit. The temperature is maintained at 85-95 °C, with natural heat loss minimised by vessel insulation.

The manganese reaction is given as follows RXN 8, and the hydrochloric acid is neutralised with hydrated lime:



The slurry contains ~ 2-4% wt/wt precipitated solids. A decanter centrifuge (150-CF-001) is used a high density slurry (paste) to send to the Leach Filter Feed Tank (140-TK-001). The clarified liquor stream from the decanter, is passed through a carbon filter to remove any organic or dissolved chlorine species that could foul the ion-exchange resins, and then collected in a stock tank (210-TK-001).

1.4 Area 200 Trace Metals Removal

1.4.1 Area 210 Stage 1

An anionic resin (Lewatit MP64 from Lanxess) is used to collect zinc, copper, iron and some manganese. These load as anionic chloro-complexes. The metals are eluted with water. The resin is periodically regenerated with 6M HCl to clean iron or manganese fouling.

Three 2 m³ fixed-bed columns operate in a cyclic operation with a Lead, Lag, Elution configuration. This format allows three identical columns operating in a sequential cycle where two columns operate as the 'lead' and 'lag' for continuous trace metal removal while the third

column undergoes 'elution' for resin stripping. The nominal loading capacity has been determined to be 7 g zinc per litre of resin.

Each column (FRP) will be ~ 4.9m high active length, and ~0.7 m diameter. This provides a H:D ratio of 7. The columns will have three entry/exit points at the base, and three entry/exit points at the top. Pneumatically operated open/shut valves will be installed, to control the direction of flow. The top and base of the column will have a ring of titanium mesh covering the inlet/outlets, supporting 200-400 um aperture PP cloth to retain the resin. Intermediate storage tanks will be part of the circuit to hold: solution ahead of the lag column, elution feed solution, elution exit solution.

The feed flowrate is 5-6 BV/hr, and the elution flowrate is at 2-3 BV/hr. The first BV of the elution stream is separated from the rest of the elution, as this contains residual mother liquor, and co-loaded cobalt from the resin. The remaining elution stream contains up to 3-6 g/L zinc as zinc chloride.

The typical sequence time will be: 2 hours of loading, 20 mins of washing, and 40 mins of stripping.

The temperature is maintained at 70-80 °C, with natural heat loss minimised by vessel insulation.

The zinc bearing solution (eluant) will be forwarded to a 5 m³ tank for precipitation with hydrated lime. The precipitated zinc hydroxychloride ($Zn_5OH_8Cl_2$) is recovered using a plate and frame filter press. The remaining solution is recycled back to the Acid Leach Buffer Tank (120-TK-003). The zinc hydroxychloride filter cake will be sent for disposal, or to an industrial recycler. There will be capacity to handle 50 tpa of zinc.

1.4.2 Area 220 Stage 2 Polishing

Cationic resins are used for polishing duties. Two resins (sourced from Purolite) are used simultaneously in the same resin-in-pulp stirred contactor. A D2EHPA functional group resin (MTX7010) is used to collect iron, zinc, manganese, and an aminophosphinic function group resin (MTS 9500) is used to collect copper. Due to the similar loading and elution characteristics of the cationic resins, the active bed in each reactor will be a mixture of 50% vol/vol MTX7010 and 50% vol/vol MTS9500 resin.

Three 2 m³ reactors are used in parallel. This allows the head liquor to be alternated between reactors for loading, while a reactor goes through elution. The reactors are sized for a nominal flowrate of 5-6 BV/hr, and are identical dimensions to the other ion-exchange columns used for trace metal removal. The only difference is the inclusion of a ribbon mixer used to gently (slowly) mix the resin. This helps facilitate the use of hydrated lime as the neutralising reagent, which is added to maintain a pH of 3.5-4.5. The agitation is essential, otherwise the addition of solids could result in blockages.

Once the resin is fully loaded, the solution is drained from the reactors through the bottom exit. Elution acid (100 g/L HCl) is applied at a rate of 3 BV/hr. The eluant is reused, until the pH is 1. At that point, the eluant is returned to the leach circuit, and fresh (100 g/L HCl) elution acid is prepared.

The typical sequence time is: 40 hours of loading, 20 mins of stripping. The nominal loading capacity is 7 g/L Zn (MTX) and 7 g/L Cu (MTS).

After trace metal removal, the liquor is stored for cobalt solvent extraction (310-TK-001).

The temperature is maintained at 60-70 °C, with natural heat loss minimised by vessel insulation.

1.5 Area 300 Cobalt Recovery

1.5.1 Area 310 Cobalt Solvent Extraction

Cobalt is separated from nickel using Cyanex 272 (Cytec) (or Ionquest 220). The organic phase is 30% vol/vol Cyanex 272 diluted in 70% vol/vol GS80 (or Isopar L). The solvent extraction circuit consists of a calcium pre-load, three cobalt extraction stages, one scrub stage, three water wash stages, and two stripping stages. A third stripping stage is included for periodic regeneration of the organic using more concentrated acid, than is used in the primary stripping stages.

In the first step, the organic is 'pre-loaded' with calcium. This is achieved by mixing the organic with a lime slurry. The target calcium loading is 8-9 g/L. A continuous flow column contactor is used for the loading of calcium. The flowrate of organic is ~25-30 m³/hr, requiring the addition of ~350-450 kg/hr of lime. The loading occurs at a pH of 6-6.5.

The contactor has organic and aqueous streams added in counter current fashion, with organic added 1/3 from the base and aqueous added 1/3 from the top. The organic stream flows upwards, and clear organic exits from the top of the unit, and advances to a static coalescer column with a drain port for any accumulated aqueous droplets. The aqueous stream flows downwards and exits from the base of the unit. The aqueous phase is recycled to a second tank, for addition of dry lime (by conveyor from the hopper). Periodically the aqueous phase is withdrawn to control water volumes, and forwarded to the leach circuit.

The organic is then split into three fractions and sent in parallel to the extraction stages. The trace metal free liquor from 310-TK-001, is passed through the reactors in series. This allows a high aqueous to organic ratio to be selected for the extraction reactors, which improves the exchange of cobalt for calcium on the organic.

The loaded organic phases are collected, and sent to the scrubbing stage. A cobalt chloride solution from the nickel solvent extraction circuit is used as the scrubbing solution. This significantly reduces the concentration of chloride salts in the entrained aqueous droplets, ahead of the water wash stages.

The resulting aqueous raffinate from the extraction and scrubbing stage are collected in 310-TK-008. Approximately 50% of the flow is progressed to nickel solvent extraction, and the other 50% is returned to the acid leach feed system (swirl hopper).

The cobalt loaded organic is forwarded to the three wash stages. High purity water is applied counter-currently from wash stage 3, to wash stage 2, to wash stage 1, a 10:1 O:A ratio. The

wash waters are collected in Cobalt SX Wash Water Collection Tank (310-TK-009) to be used back in Area 100 Leaching.

The washed cobalt-loaded organic is forwarded to the stripping stages. A solution of acidified (~100 g/L free H₂SO₄) cobalt sulphate is used as the stripping solution at a pH of 1-2. This is operated counter-currently. The stripped organic is recycled to the calcium pre-loading stage, and the strip solution is stored for crystallisation (410-TK-001).

The typical metal concentrations are given in Table 2.

Table 2 - Typical concentrations in cobalt solvent extraction circuit

Feed	Aqueous	Organic
Calcium pre-load		8-9 g/L calcium
Cobalt extraction	Input 30-40 g/L Co Output 5-10 g/L Co	12-13 g/L cobalt loading
Scrubbing	30-40 g/L Co as CoCl ₂ from Ni SX	
Water washing	Wash 1 <3 g/L Co, < 3 g/L Ca Wash 2 <1 g/L Co, < 1 g/L Ca Wash 3 <0.2 g/L Co, < 0.2 g/L Ca	
Stripping	Input 40-50 g/L Co, 100 g/L free H ₂ SO ₄ Output 80-90 g/L Co	

The use of a calcium pre-load has the following two critical advantages:

- No pH control during cobalt extraction is required
- Minimal nickel co-loading occurs
- Lime can be used as the neutralisation agent

The cobalt solvent extraction is maintained at 50-60 °C to reduce organic viscosity and improve cobalt loading and stripping kinetics. Heating is provided in the aqueous and organic stock tanks using titanium coils, fed with hot glycol solution (120 °C).

As appropriate, the organic can be fully stripped (regenerated) using 1 M H₂SO₄ via Cobalt SX Organic Clean-up Cell (310-CF-011) if there is a build-up of zinc or iron, as these metals are not stripped during the primary cobalt stripping at pH 1-2.

Centrifugal mixers have been selected as a package plant. These provide advantages over conventional mixer-settlers as follows: lower footprint, lower entrainment, faster phase disengagement times, and much lower overall organic inventory (small fire risk). Importantly, they allow for improved crud separation into the aqueous phase, due to the centrifugal action on particles in the mixer.

1.6 Area 400 Crystallisation

The stripped cobalt sulphate solution from the solvent extraction circuit is forwarded to a single stage 5 m³ crystalliser (410-TK-002) reactor (PP). Isopropanol (90% IPA / 10 % water) is added to the reactor, decreasing cobalt sulphate solubility from 80-90 g/L to 30-40 g/L. The target final isopropanol concentration is ~30%. The reactor is operated at a temperature of < 35 °C, and

preferably at 30 °C. The low temperature is to ensure production of heptahydrate crystals, and no hexahydrate or monohydrate.

The reactor uses a halar coated agitator to minimise possible trace metal contamination. The reactor has a constant flowing bottom outlet, controlled by a stepper pinch valve. The valve is also linked to level control in the reactor. The outlet discharges directly to the pusher centrifuge. The crystals are filtered, washed, and then dried in a fluidised bed dryer to a final moisture content of 0.1-0.3 wt%. The fluidised bed dryer offgas goes to an intermediate water scrubber, and then to main plant lime scrubber. The crystals are transferred from the dryer to storage in a hopper (20-25 m³) using an aero-mechanical conveyor. A second conveyor is used to transfer the crystals to a bag filling station, where 1t bulka bags are filled. The bags are packed into sea containers using a 2-3 t forklift. Any spillage will be washed to a sump and returned to the process.

Fines (<150 um) passing through the centrifuge screen, are recycled back to the crystalliser reactor using a cyclone. The cyclone overflow (free of solids) advances to IPA separation. The target particle size distribution of the recovered crystals is 250-600 um, p80 400-500 um.

A fractionation distillation column is used to separate an IPA azeotrope (90% IPA, 10% water). The top section of the column operates at 81 °C, with a portion of the condensed offgas returned to the column. The bottom section operates at 100 °C (or equivalent boiling point) of the IPA-free liquor. The cooled bottoms are recycled back to the Cobalt SX Strip Acid Feed Tank (310-TK-007). The majority of heating is applied in the boiler using a glycol passing through titanium heating coils. Various heat exchangers are employed across the circuit to: increase the temperature of the incoming stream (target to heat to > 60 °C), decrease the temperature of the exit IPA and IPA-free streams (target to cool to < 60 °C).

1.7 Area 500 Nickel Recovery

1.7.1 Area 510 Nickel Solvent Extraction

The aqueous raffinate from the cobalt solvent extraction process contains < 5 g/L cobalt, and 3-7 g/l nickel. The liquor stream will be split, with approximately half returning to the leach circuit, and the rest progressing to the Nickel SX Feed Tank (510-TK-001).

The residual cobalt is separated from nickel by selectively extracting as a chloro-complex using undiluted Alamine 336. The cobalt is extracted across two extraction stages operating in counter-current mode. The cobalt loaded organic is stripped across two stages operating in counter-current mode with water (e.g. the cobalt SX wash waters), producing a solution containing 20-30 g/L cobalt and 30-40 g/L chloride. This stream is used for scrubbing in the cobalt solvent extraction circuit.

The cobalt-free raffinate is stored for nickel recovery by electrowinning. A bleed is periodically taken from this tank for magnesium control.

The nickel solvent extraction circuit utilises centrifugal separators as per the cobalt circuit.

The circuit operates at a temperature of 70-80 °C to promote phase disengagement and reduce organic viscosity. Heating is provided in the aqueous and organic stock tanks using titanium coils, fed with hot glycol solution (120 °C).

1.7.2 Area 520 Nickel Electrowinning

Nickel is electrowon from the nickel chloride solutions using high-flow tubular cells (similar to EMEW cells). The anode and cathode compartments are divided using a bag.

The cell operates at a cathode current density of 500-1000 A/m², and the nickel is plated from 5-10 g/L down to 2-3 g/L. Titanium (Gr2) starter sheets are used. The nickel plates are periodically harvested from the cells. Power is supplied by a rectifier operating at 3-5 V and 225,000 A.

Catholyte is recirculated at a flowrate of ~ 3 m³/hr per each cell, returning back to the feed storage tank. The pH is maintained at ~ 1, through the addition of lime to the storage tank. The temperature is maintained at ~60 °C, using titanium coils in the storage tank, fed with hot glycol (120 °C).

A portion of the catholyte passed through the dividing bag into the anode chamber. At the anode, chlorine is produced. The anolyte (liquor and chlorine gas) is withdrawn from the cells and contacted with lime. The lime maintains a pH of >10, and the chlorine gas is dissolved and converted into hypochlorite. The hypochlorite slurry is used for oxidation in the manganese circuit, and any excess is sent to the acid leach reactors.

The total number of cells is 225. These are divided into banks of 20, in four rows of five. Each row is connected to a single pump, with the flow passing through four rows in sequence, before returning to the storage tank. During nickel plate harvesting, a bank of cells (4 x 5) are taken offline.

Where market conditions are favourable, cobalt metal could be produced in the electrowinning circuit using the same equipment and minor process control changes.

1.7.3 Area 530 Magnesium Recovery

Magnesium throughout the plant circuit is periodically controlled by a bleed stream.

First, the pH is raised to 7-8 with lime to precipitate the contained nickel and cobalt. The solids are separated using a plate and frame filter. The solids are recycled to the acid leach reactors.

Secondly, the pH of the nickel-free liquor is raised to ~10 with lime to precipitate magnesium. The solids are separated by a plate and frame filter. The solids are sent for disposal. The remaining liquor is recycled back to the acid leach reactors.

During the second step at pH 10, some ammonia may be evolved from the liquor. This will be captured in a dedicated droplet scrubber to form ammonium hydroxide/sulphate (ammonia-water-sulphate). An acidified (pH 2-4) water-sulphuric acid solution will be recirculated through a packed bed unit to capture the ammonia gas. The ammonium sulphate will be periodically

sent for disposal (ideally reused industrially). The residual ammonia free gas will continue to the main plant scrubber.

1.8 Process Plant Heating

Heat management in the plant will be achieved using three discrete glycol systems:

- System one will cover the leach and manganese circuits.
- System two will cover the solvent extraction circuits.
- System three will cover the IPA distillation circuit.

Glycol will be heated in three separate holding tanks using electric elements (or gas heating?) to 120 °C. The hot glycol is then recirculated through titanium coils, jackets, tube and shell heat exchangers as required, using magnetic drive pumps. The cold glycol then returns to the holding tanks. Glycol has been selected to offer better performance and lower operating risks and costs, compared to steam systems.

1.9 Water Management

Water is consumed in various parts of the plant: filter washing, ion-exchange stripping, and solvent extraction washing/stripping.

The highest purity water must be used in the cobalt solvent extraction wash stages. This is to minimise any contamination of the final cobalt sulphate crystals. The wash water is then reused for stripping cobalt from the nickel solvent extraction circuit.

Mains water can be used for washing the leach residue, zinc precipitate, and the magnesium precipitate on the respective filters.

Where appropriate, the used water is collected in a central tank. The water is then re-used for reagent make-up (lime, floc, acid dilution, etc).

The overall plant water management, needs to balance inputs with outputs. The only outputs are: evaporation, free moisture in filter cakes, and bound water in filter cakes (e.g. the seven waters of crystallisation in cobalt sulphate heptahydrate). Hence evaporation is critical to maintaining the plant water balance

The primary water evaporation unit in the plant, is the oxidation leach heat exchanger. The hot slurry at 140-150 °C is passed through a titanium tube and shell heat exchanger. Wash waters (from the central storage tank) are heated, with some evaporating as steam. The remaining water is then added to the acid leach tanks, where further evaporation occurs through the addition of external heat (glycol heater coils) to those tanks.

Steam is released from the plant via the main plant scrubber.

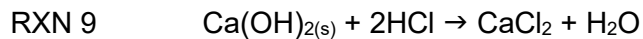
1.10 Plant Scrubber

The plant off-gas will be scrubbed using a lime scrubber. A venturi eductor design will be used. This provides slight negative pressure on the connected tanks, and thus pulling the offgas

through the ducting to the recirculating lime slurry. The pressure is controlled using a “false-air” inlet valve to balance the system.

The scrubber operates at a pH of 10-11, with excess lime. The temperature is typically 60-80 °C due to the steam contained in the tank offgas.

The efficiency of HCl scrubbing is >99%, as per RXN 9:



Periodically, a bleed of the scrubber slurry can be recycled to the acid leach reactors.

The scrubber ducting is 300 mm diameter FRP. This connects to the respective vessels in 150 mm diameter FRP. Drain points will be installed to recover condensed water.

There are three specific inter-stage scrubbers in the plant:

1. Dust scrubbing around the feed section.

A recirculating water, disc and donut scrubber will be used to capture dust. This will discharge to the leach circuit. The residual dust-free gas will go the main plant scrubber.

The bag unloading on the feed hopper, bag compactor chute, and the swirl hopper will be connected to this unit.

2. Cobalt sulphate scrubbing around the product section.

A recirculating water, packed bed scrubber will be used to capture cobalt sulphate droplets or dust. This will discharge to the leach circuit. The residual dust-free gas will go the main plant scrubber.

The cobalt solvent extraction strip storage feed and exit tanks, crystal conveyors, crystal hopper, and bagging station will be connected to this unit.

The IPA distillation unit top fraction will be ducted directly to the IPA storage tank. This tank will then be ducted to the crystalliser. The crystalliser will be ducted to the cobalt solvent extraction strip exit storage tank (feed aqueous stream to the crystalliser). This sequence will minimise the losses of IPA into the main ducting system.

3. Solvent extraction organic scrubbing.

Various packed bed carbon filters will be used to capture organic vapours. The residual organic free gas will go the main plant scrubber.

The cobalt solvent extraction organic storage tanks, and nickel solvent extraction organic storage tanks will be connected to these units.

1.11 Area 600 Reagents

1.11.1 Reagent List

The following is a list of reagents utilised in the process:

- Lime delivered by tankers and stored in a silo
- Technical Grade Sulphuric Acid delivered by tankers and stored in isotainers.
- Hydrochloric acid delivered by tankers and stored in isotainers.
- Flocculant (Rheomax DR1010) delivered as drums
- Water by mains connection, and high purity water by tanker

Make-up requirements:

- Cyanex 272
- Alamine 336
- GS80 (diluent)
- MP64 resin
- MTX7010 resin
- MTS9500 resin
- NaCl
- CaCl₂
- Isopropanol
- Glycol

Onsite generation

- DI Water Package (if required)
- Oxygen (VPSA plant)

1.12 Area 700 Utilities

A list of ancillaries are required as follows:

- Air Compressor