



**AWARUITE REFINERY, British Columbia, Canada**

## **SCOPING STUDY REPORT**

**Issued: 26 March 2025, Revision A**

**Project No.: 263522**

**Document Number: 263522-0000-BA00-RPT-0001**

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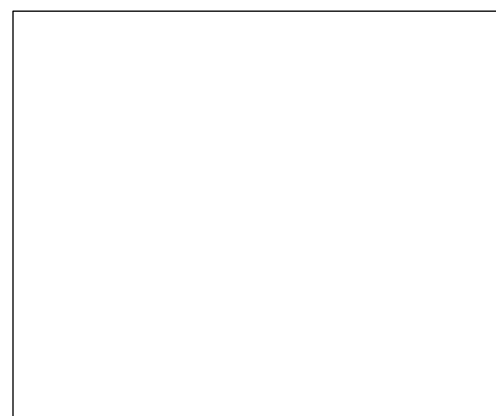
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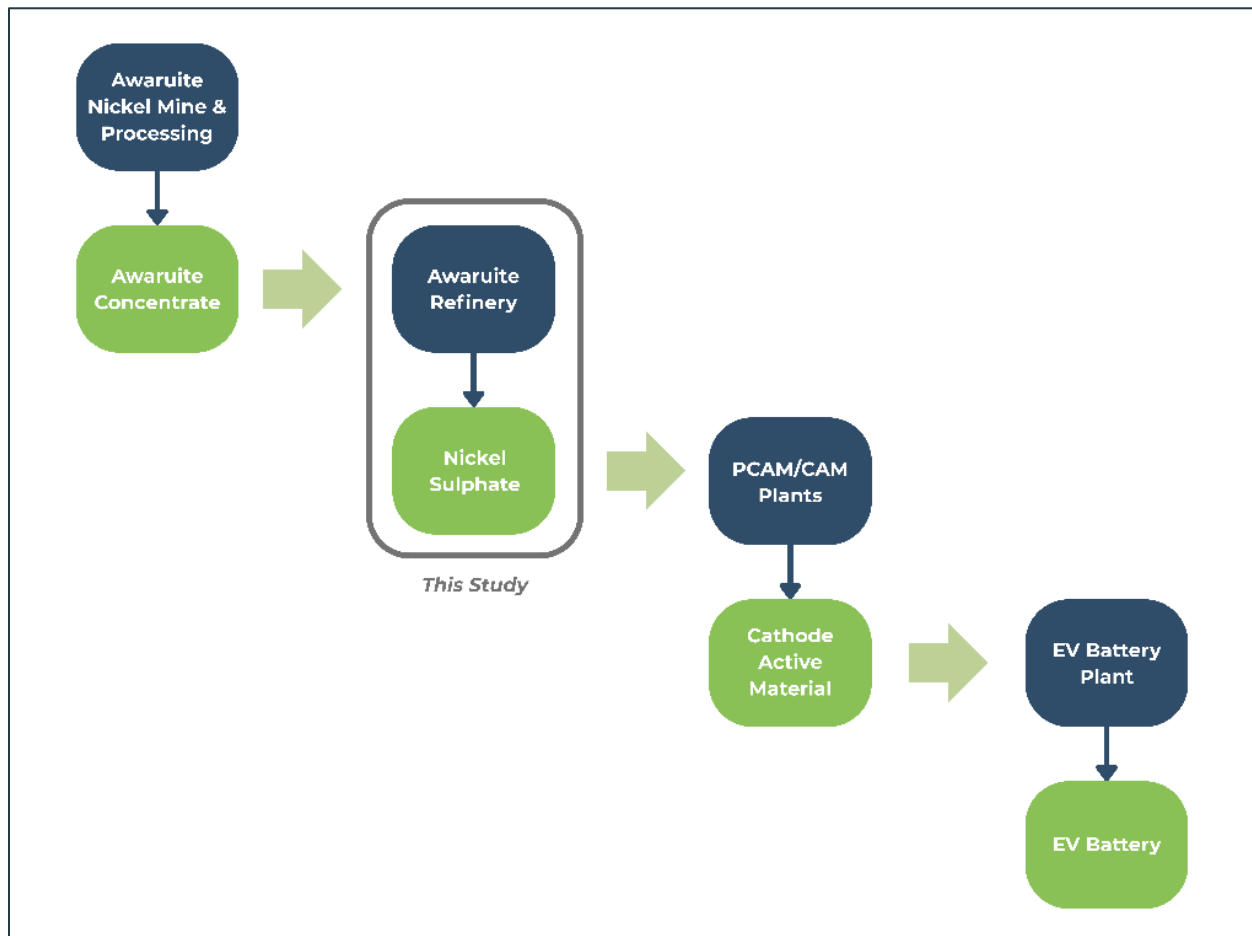
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## Executive Summary

FPX commenced development of this awaruite refinery scoping study in October 2024 to demonstrate the economic and strategic opportunity to refine awaruite concentrates to battery-grade nickel sulphate and other valuable byproducts. This study incorporates the flowsheet advances from FPX's 2024 pilot-scale hydrometallurgical testwork and value engineering studies.

The mine-to-battery pathway for awaruite mineralization is presented in Figure ES-1. At a mine site, awaruite mineralization can be subjected to a simple mineral processing flowsheet to produce a high-grade awaruite concentrate, which can then be marketed either to the stainless steel or electric vehicle (EV) battery supply chains. This study envisions purchasing of such awaruite concentrate and refining to battery-grade nickel sulphate, which would then be marketed to producers of precursor cathode active material (PCAM) and cathode active material (CAM), to further process the nickel sulphate into CAM, a direct input in EV battery cell fabrication.



**Figure ES-1. Path of Awaruite Nickel Units from Mine to EV Battery**

This study outlines a mid-stream refinery located in an industrial location in central British Columbia, which will be fed with awaruite concentrate and produce battery-grade nickel sulphate. The study considers a refinery capable of producing 32,000 tpy of contained nickel in battery-grade nickel sulphate. In addition to nickel sulphate, the refinery will produce three byproducts, including cobalt carbonate, copper cement, and ammonium sulphate (a valuable fertilizer product).

## Study Scope

In tandem with advancing its awaruite nickel projects, FPX has advanced the concept of an off-site refinery to process awaruite concentrate into products for the EV battery supply chain, incorporating the latest pilot-scale hydrometallurgical testwork and value engineering studies. In July 2024, FPX issued a request for proposal for a scoping study for an off-site awaruite refinery.

Wood began work on the scoping study in October 2024, to develop and refine a design and cost estimate for the awaruite refinery. The study addresses a stand-alone industrial project and anticipates that it would refine awaruite concentrate from multiple sources.

The study comprised several tasks:

- Build on the results of metallurgical testing and previous studies.
- Conduct trade-off studies.
- Develop a basis of design.
- Progress the process design to approximate preliminary feasibility study definition
- Progress functional engineering to scoping study definition.
- Estimate capital and operating costs at an accuracy level of  $\pm 35\%$ .
- Identify data gaps and opportunities to be addressed during the next stage (PFS).

This report provides a preliminary engineering and economic evaluation of the awaruite concentrate refinery project. Over its envisioned 40-year life, the refinery will produce 32,000 tpy of battery-grade nickel contained in nickel sulphate hexahydrate crystals, as well as co-products including:

- 570 tpy of contained cobalt in cobalt carbonate
- 240 tpy of contained copper in copper cement
- 87,400 tpy of ammonium sulphate, a valuable fertilizer product.

FPX used this information to determine the economic viability of the awaruite refinery project. All costing and financial information in the study is presented in Q4-2024 US dollars.

## Metallurgy & Process Facilities

The metallurgical testwork program involved multiple bench- and a pilot-scale campaigns. The latest testwork campaign involved continuous pilot-scale testing of leaching unit operations, and confirmed the leaching flowsheet. nickel leach extractions greater than 99% and the production of low-impurity leach solution, suitable for downstream purification and crystallization, were simultaneously achieved during piloting. In addition, bench-scale testing of solution purification and crystallization unit operations demonstrated the ability to produce battery-grade nickel sulphate crystals using the new ammonia-based flowsheet. Incorporating testwork results, the refining strategy takes advantage of awaruite's characteristics in a simple flowsheet utilizing well-proven unit operations.



In the leaching area, awaruite concentrate is first subjected to an atmospheric leach, which serves the dual purpose of 1) commencing awaruite dissolution, and 2) using awaruite as a reagent to neutralize free acid and precipitate remaining iron from the pressure leach solution. Any unleached awaruite is then further leached in a mild pressure oxidation circuit, where full awaruite dissolution is achieved in tandem with initial iron precipitation. Pressure leach solution then reports to the atmospheric leach circuit, and pressure leach residue is dewatered for disposal. A slipstream of pressure leach solution is processed in a copper removal circuit, where a copper cement grading approximately 70 to 80% copper is produced.

The final leach solution, grading 100 g/L nickel, is first processed in a solvent extraction (SX) circuit to extract cobalt. Extracted cobalt is then precipitated from the cobalt-rich solution as a carbonate product grading approximately 50% cobalt. While a cobalt carbonate product was selected for the study, FPX testwork has also demonstrated the ability to produce cobalt-rich mixed hydroxide precipitate (MHP).

Nickel is then extracted away from the cobalt-depleted leach solution to produce a purified and concentrated stream of nickel sulphate, which is then crystallized into battery-grade nickel sulphate crystals suitable for use in the EV supply chain.

The nickel-depleted leach solution is then treated to sequentially removal trace levels of nickel and magnesium. This purified stream is then crystallized into ammonium sulphate crystals, a widely used industrial fertilizer. Miscellaneous minor process streams are also processed in the ammonium sulphate crystallizer which enables the refinery to operate as a zero liquid discharge facility.

## Utilities and Infrastructure

This study assumes a project site will be in central British Columbia, in proximity to existing utilities and sources of skilled labour. The project is envisioned to connect to the BC Hydro grid.

Infrastructure will include a perimeter access road, internal roads, staging areas, gatehouse, office complex, laboratory, warehouse, maintenance shop, electrical sub-station, acid plant, oxygen plant, and railroad spurs to receive commodities for refinery operations.

Due to assumed proximity to an urban center, no construction or operational camps are required.

## Capital Cost Estimate

Initial capital costs were estimated in alignment with AACE (Association for the Advancement of Cost Engineering) Class 5 standards, while sustaining and closure capital costs were estimated on an order-of-magnitude basis. All estimates are presented in Q4-2024 US dollars. The total initial capital cost for the project is estimated to be \$424 million, with no expansion considered. Total sustaining capital cost is estimated to be \$40 million, and total closure capital cost is estimated to be \$42 million. No salvage value is considered due to the 40-year operation life.

**Table ES-1. Total Estimated Capital Costs**

Cost Type	Category	Total, US\$ millions
Initial Capital Costs	Refinery Process	\$152
	Reagents	\$45
	Utilities, Services and Infrastructure	\$40
	<b>Total Direct Costs</b>	<b>\$237</b>
	Indirect Costs	\$81
	Contingency	\$89
	Owner's Costs	\$18
	<b>Total Initial Capital</b>	<b>\$424</b>
Total Sustaining Capital Cost (estimated by FPX)		\$40
Total Closure Capital Cost (estimated by FPX)		\$42
<b>Total Capital Cost</b>		<b>\$506</b>

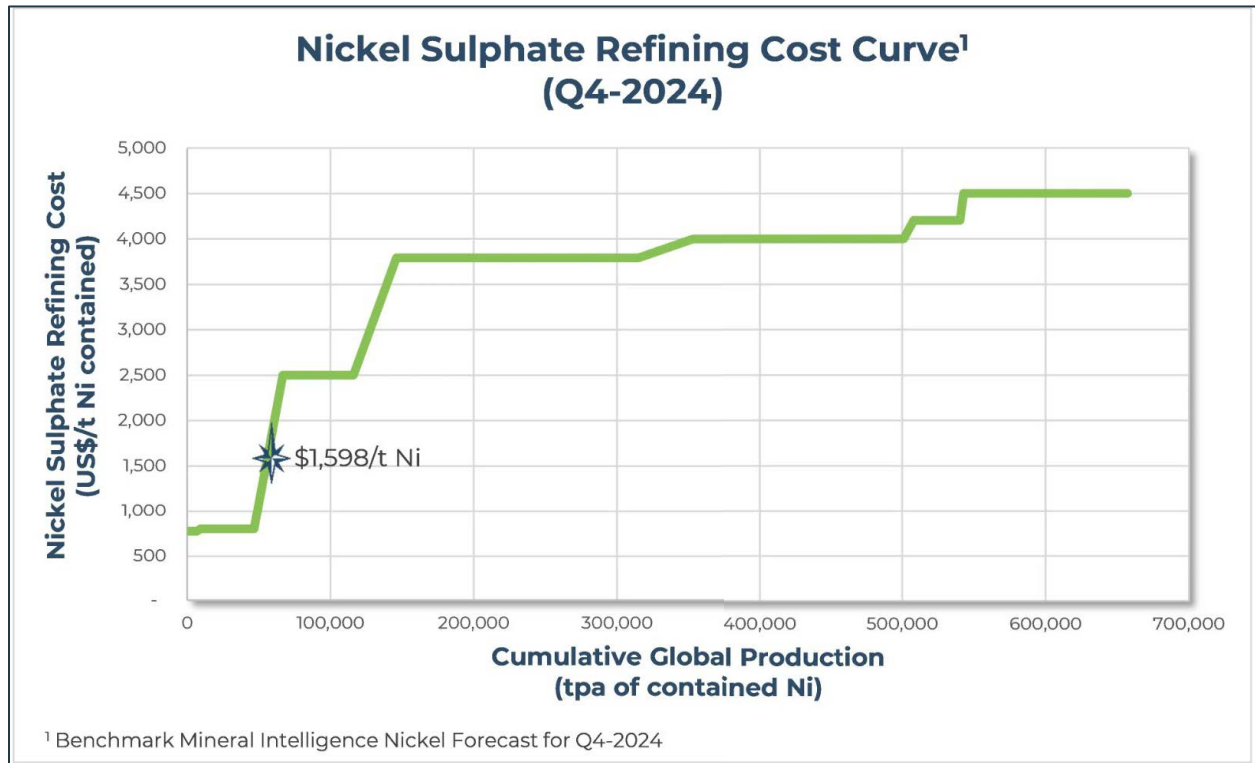
## Operating Costs

Total operating costs are estimated to average \$1,598 per tonne of nickel produced before byproduct credits, with a breakdown of these costs by cost centre presented in Table 2. This equates to approximately \$51 million per year in total operating costs. The net operating cost inclusive of byproduct credits for cobalt, copper, and ammonium sulphate is \$133/t Ni (\$0.06 /lb Ni).

**Table ES-2. Estimated Operating Costs**

Category	Cost (US\$/t Ni Produced)
Reagents	\$757
Consumables	\$264
Labour	\$258
Maintenance	\$136
Power	\$69
General & Administrative	\$114
Total (excluding byproduct credits)	\$1,598
Total (including byproduct credits)	\$133

According to Benchmark Mineral Intelligence's (Benchmark's) database, the estimated operating costs for the refinery are in the lowest decile of current global production, as presented in Figure ES-2.



**Figure ES-2. Nickel Sulphate Refining Cost Curve**

## Economic Analysis

At an assumed nickel price of \$8.50/lb (\$18,738/t) and a USD:CAD exchange rate of 0.74, the refinery generates an after-tax NPV8% of \$445 million, an after-tax IRR of 20%, and an after-tax payback of 4.0 years. Table ES-3 provides further details on study economics.

Benchmark maintains a comprehensive database of nickel production statistics, and forecasts long-term pricing premiums relative to the LME nickel price for battery-grade nickel sulphate. While the nickel sulphate market is currently small with inherent pricing volatility, even the most conservative of EV adoption rates will see a significant increase in nickel sulphate requirements. As the nickel sulphate market grows in coming years and preferred feedstocks are established, it is expected that a more consistent premium basis will be established based on typical upgrading costs. Benchmark's forecast nickel sulphate premium basis for the year 2030 is \$1,575/t nickel (\$0.71/lb nickel), which has been applied in the economic analysis. This premium is based on Benchmark's "base case" forecast of EV adoption and battery chemistry trends.

The study models taxes in accordance with provincial and federal legislation. It reflects the impact of the federal government's refundable critical minerals investment tax credit, announced in the 2023 federal budget, which is proposed to be equal to 30% of the capital cost of eligible property for the extraction and processing of certain critical minerals, including nickel. The study estimates total LOM taxes paid of C\$1,000 million, including C\$520 million to the Province of British Columbia and C\$480 million to the Government of Canada.

According to Benchmark’s database, the 2024 annualized nickel sulphate production market size was approximately 657,000 tonnes per year of contained nickel as of the fourth quarter of 2024, with production heavily dominated by China at 76%. Additionally, less than 1% of current nickel sulphate production is North American (3,300 tonnes per year of contained nickel). As such, the 32,000 tonnes per year of high-quality nickel sulphate produced by the refinery in Canada would represent an approximate tenfold increase in current North American nickel sulphate production.

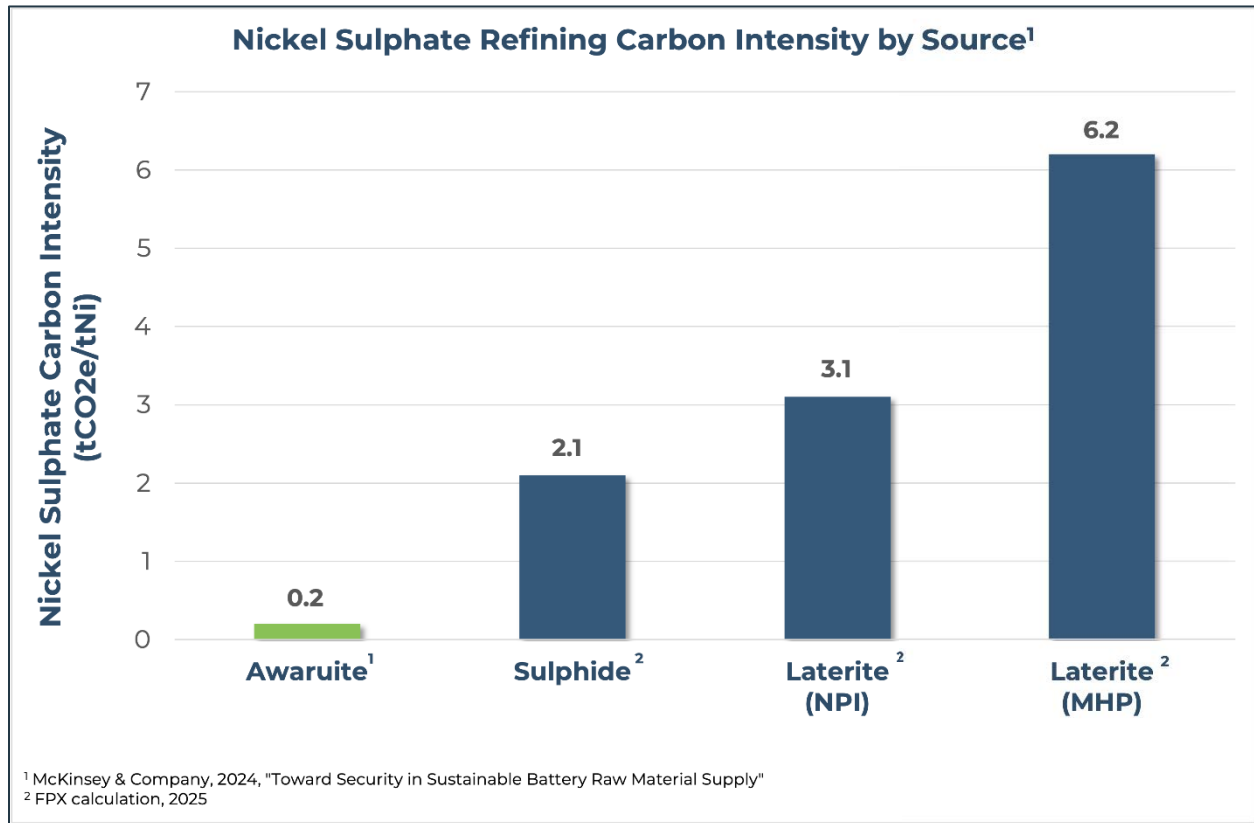
**Table ES-3. Study Economics**

Economic Basis/Results		Units	Value
Payability, Awaruite Concentrate		% of LME Ni	92
Price	Nickel	US\$/lb	8.50
	Cobalt	US\$/lb	15.00
	Copper	US\$/lb	4.00
	Ammonium Sulphate	US\$/t	330
Payability	Cobalt	%	85
	Copper	%	95
After-Tax	NPV <sub>8%</sub>	US\$, millions	\$445
	Payback	years	4.0

## Environment and Community

This study assumes the refinery will be in central BC, with the selected location ultimately subject to community consultation, environmental characterization and baseline studies. The *BC Environmental Assessment Act Reviewable Projects Regulation* specifies that a new non-ferrous metal refinery would require assessment under the established provincial EA process.

Refining operations are projected to have a carbon intensity of 0.2 tonne of CO<sub>2</sub> per tonne of Ni produced. This is magnitudes lower than current nickel sulphate production routes, as presented in Figure ES-3.



**Figure ES-3. Nickel Sulphate Refining Carbon Intensity by Source**

# 1. Metallurgical

## 1.1 Introduction

Metallurgical testwork on awaruite ( $\text{Ni}_3\text{Fe}$ ) nickel ores has shown that a relatively simple flowsheet using magnetic separation, followed by conventional froth flotation, can produce a highly desirable awaruite concentrate that provides flexibility for downstream consumption and/or further processing. Due to awaruite's very high nickel tenor (approximately 76%), very high-grade mineral concentrates of 60% nickel can be produced.

The high-nickel content and low-impurity grades of awaruite concentrate make it a suitable candidate for hydrometallurgical refining to produce battery-grade nickel sulphate. FPX's metallurgical testwork campaigns to refine awaruite concentrate are summarized in Table 1-1, and form the basis of design for the hydrometallurgical refinery. Testwork completed in 2023 and 2024 testwork is summarized in this section. All testwork was conducted by Sherritt Technologies in Fort Saskatchewan.

**Table 1-1: Metallurgical Testwork Summary**

Year	Major Testwork Performed	Metallurgical Sample	Sample Size
2020	Bench-scale pressure leaching	FPX 1	130 g
2023	Bench-scale pressure and atmospheric leaching, solution purification (with NaOH), nickel sulphate crystallization	FPX 2	6 kg
2024	Pilot-scale pressure and atmospheric leaching Bench-scale solution purification (with $\text{NH}_3$ ), nickel sulphate crystallization, ammonium sulphate purification	FPX 3	50 kg

## 1.2 Sample Selection

The composition of the awaruite concentrates used in the three testwork programs and the refinery design grade are presented in Table 1-2.

**Table 1-2: Composition of Samples Used in Refining Metallurgical Testwork Programs**

Metallurgical Sample	Ni (%)	Fe (%)	Co (%)	Mg (%)	Cu (%)	S (%)	Si (%)	Cr (%)	Mn (%)	Zn (%)	Al (%)	Ca (%)
FPX 1	59.0	23.8	0.90	0.62	0.75	0.70	0.51	0.22	0.03	-	0.01	-
FPX 2	66.3	24.6	1.08	0.56	0.43	0.44	0.49	0.21	0.02	<0.03	<0.04	-
FPX 3	61.5	25.4	1.16	1.49	0.44	1.66	1.26	0.22	0.02	<0.01	<0.04	<0.01
Refinery Design	60	25	1.1	1.7	0.5	1.0	3.2	0.22	0.02	0.01	0.05	0.01

### 1.3 Summary of Metallurgical Test Results

Counter-current leach conditions, consisting of a pressure oxidative leach (POX) and an atmospheric leach, were optimized in batch testing. During continuous piloting, these demonstrated overall nickel leach recovery of greater than 99% while producing a low-impurity pregnant leach solution (PLS). Awaruite concentrate was demonstrated to be a sufficiently strong neutralization agent to neutralize above pH 4.5; this permitted complete removal of free acid and iron in the atmospheric leach, and direct feeding of PLS to downstream SX operations.

Batch cobalt solvent extraction (CoSX) and nickel solvent extraction (NiSX) testwork provided proof-of-concept of a scheme to produce a purified and concentrated nickel sulphate solution. Different test series were completed, using either sodium hydroxide or ammonium hydroxide as the SX neutralizing reagent. In both cases, CoSX achieved sufficient Co over Ni selectively, and NiSX achieved sufficient Ni over Mg selectively, resulting in nickel sulphate crystals sufficiently low in these impurities. Other impurities in the PLS, present only in trace quantities, include copper, zinc, manganese and calcium; all of which were effectively managed in the purification scheme.

Nickel sulphate crystals were produced at purity levels meeting current battery-grade specifications. This confirms that awaruite concentrate is a suitable nickel intermediate for refining nickel sulphate to feed the battery electric vehicle industry.

CoSX produces a cobalt-rich solution with minor impurities, from which a high-grade cobalt intermediate can be produced. Cobalt hydroxide and carbonate precipitates have been successfully produced.

Trace levels of nickel and magnesium in the NiSX raffinate were successfully removed to produce a concentrated ammonium sulphate solution suitable for production of fertilizer-quality ammonium sulphate crystals.

The proposed block flow diagram for refining awaruite concentrate is shown in Figure 1-1, and is based on the metallurgical results presented in this section.

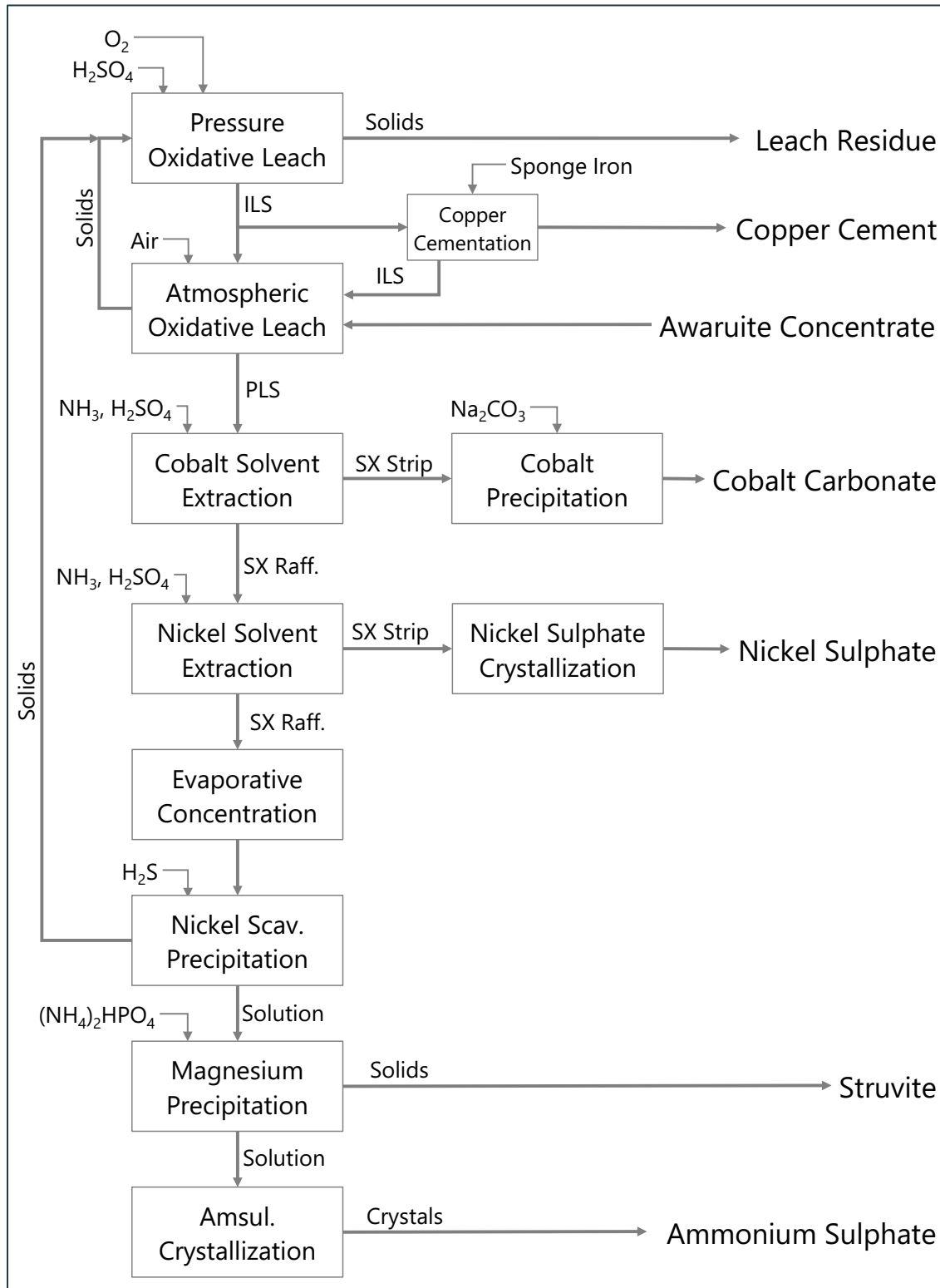


Figure 1-1: Refinery Block Flow Diagram



## 1.4 Mineralogy

Dedicated mineralogy work was not completed on the awaruite concentrate used in the hydrometallurgical testwork programs. However, mineralogy has been inferred based on concentrate elemental grades (see Table 1-2), and in publicly available mineralogy work completed on ultramafic ores from which awaruite concentrates have been produced. Inferred mineralogy is presented in Table 1-3. The concentrate is dominated by awaruite, with lesser concentrations of magnetite and serpentine. All sulphur is suspected to be present as nickel sulphides, predominantly pentlandite, though a lesser amount of heazlewoodite is expected as well. The unidentified chromium spinel has been assigned chromite stoichiometry for calculation purposes. Copper and cobalt are associated with awaruite.

**Table 1-3: Inferred Mineralogy of Awaruite Concentrate**

Mineral	Chemical Composition	Normalized Concentration (%)
Awaruite	Ni <sub>3</sub> Fe	79.7
Magnetite	Fe <sub>3</sub> O <sub>4</sub>	9.4
Serpentine	(Mg,Fe) <sub>3</sub> Si <sub>2</sub> O <sub>5</sub> (OH) <sub>4</sub>	7.3
Pentlandite	Fe <sub>4.5</sub> Ni <sub>4.5</sub> S <sub>8</sub>	3.1
Unidentified Cr-Spinel	FeCr <sub>2</sub> O <sub>4</sub>	0.5

## 1.5 Pressure Oxidative Leach

### 1.5.1 2023 Testwork Summary

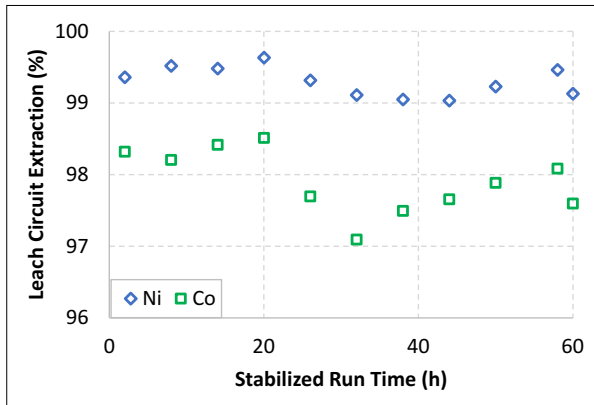
The 2023 bench testwork program established optimized POX conditions of 150°C, 375 kPa oxygen partial pressure, and a one-hour residence time. These conditions yielded nickel and cobalt extractions greater than 99% while effectively managing terminal acid (20 to 25 g/L H<sub>2</sub>SO<sub>4</sub>) and iron (4-8 g/L) concentrations. Similar results were achieved using fresh awaruite concentrate, and a blend of fresh awaruite concentrate and partially leached awaruite concentrate from atmospheric leach.

### 1.5.2 2024 Pilot Plant

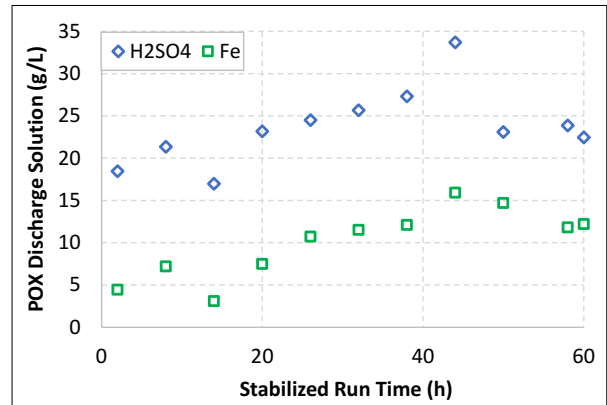
The leaching pilot plant used POX conditions comparable to the 2023 work, and operated with POX solids feed rates of ~0.6 kg/h atmospheric leach residue. Both pressure and atmospheric leach stages were operated continuously, while solid-liquid-separation was testing was completed on a batch basis.

Pilot-plant leach circuit extractions of nickel and cobalt are summarized in Figure 1-2 over a 60-hour run time, following an initial period of circuit stabilization. Nickel extractions were consistently above 99%, and averaged 99.3%. Cobalt extractions ranged between 97.0% and 98.5%, and averaged 97.9%.

Free acid and iron concentrations in the POX discharge solution are summarized in Figure 1-3. Target free acidity, controlled by acid additional rate, was 25 g/L while iron concentrations expectedly varied as a function of free acidity. The sum of free acidity and iron in the POX discharge solution determines the neutralization load in the downstream atmospheric leach. Nickel concentrations ranged between 40 and 70 g/L.



**Figure 1-2: Pilot Plant POX Nickel and Cobalt Leach Extractions**



**Figure 1-3: Pilot Plant POX Discharge Solution-Free Acid and Iron Concentrations**

In addition to batch filtration of the POX discharge slurry, bench-scale solid-liquid-separation characterization testing was completed. This included static settling, vacuum filtration, and pressure filtration. Acceptable thickening and pressure filtration rates were achieved, while vacuum filtration rates proved too slow to be commercially viable. Based on these results, a leach discharge thickener, three-stage counter-current decantation circuit followed by pressure filtration, was chosen to recover leach solution from the POX discharge slurry and generate washed leach residue.

## 1.6 Atmospheric Leach

The objective of atmospheric leach is to maximize the removal of sulphuric acid and iron from the POX intermediate leach solution (ILS) by using awaruite concentrate as the neutralizing agent. This simultaneously:

- purifies the leach solution to reduce the cost of downstream purification
- effectively uses the contained acid in POX discharge, including acid generated through iron precipitation, to dissolve nickel
- partially dissolves the awaruite to reduce the need for dissolution in the POX leach, which in turn reduces the autoclave cooling load (awaruite dissolution is exothermic)
- increases the nickel concentration of the leach solution, thereby reducing the size of downstream equipment (nickel sulphate solubility is greater under atmospheric leach conditions than under pressure leach conditions).

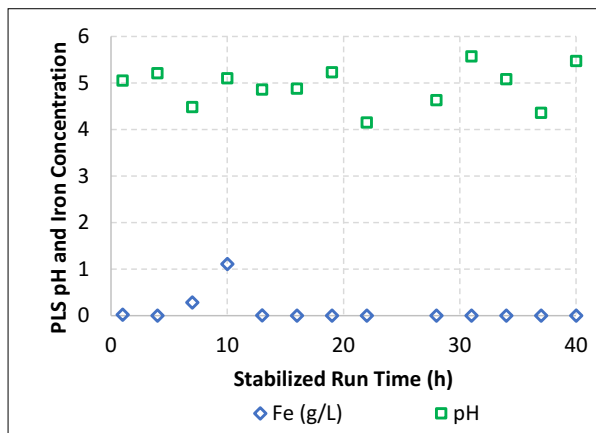
### 1.6.1 2023 Testwork Summary

The 2023 bench testwork program established optimized atmospheric leach conditions of 85°C and a residence time of four to six hours. This resulted in complete removal of iron, copper, aluminum and chromium with a resultant leach solution pH greater than 4.5. Concentrated oxygen was used for oxidation. Awaruite concentrate addition rates were set based on those expected in a counter-current leach circuit with 100% of the fresh awaruite concentrate reporting to atmospheric leach.

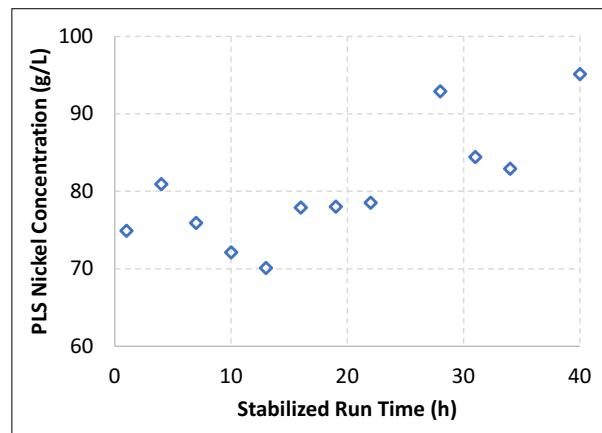
### 1.6.2 2024 Pilot Plant

Comparable atmospheric leach conditions were used in pilot plant, which operated with an awaruite concentrate fresh feed rate of 0.6 kg/h. Pilot plant performance is summarized in Figure 1-4. This showed that awaruite concentrate was consistently able to neutralize the solution above pH 4, averaging pH 4.9 during the stabilized run time. At this elevated pH, ferrous iron oxidation, followed by ferric iron precipitation, readily occur as demonstrated by low iron concentrations presented in Figure 1-4.

Nickel concentrations in the PLS are summarized in Figure 1-5. Nickel concentrations of 80 g/L were achieved early in the pilot run. At approximately hour 28, the water addition rate to the POX was reduced to increase atmospheric leach nickel concentrations. The pilot plant did not run for sufficient time for this change to be fully observed in the atmospheric leach, but 95 g/L Ni in the final data point demonstrates its trending success.



**Figure 1-4: Pilot Plant Atmospheric Leach Discharge Solution pH and Iron Concentrations**

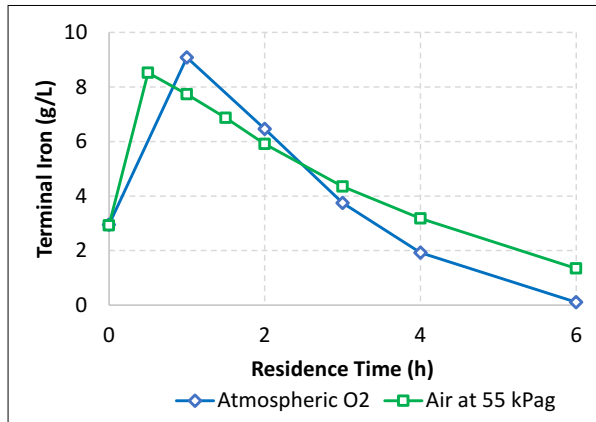


**Figure 1-5: Pilot Plant Atmospheric Leach Discharge Solution Nickel Concentrations**

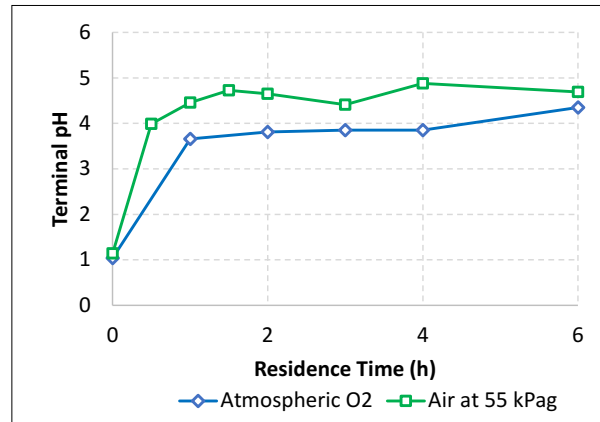
In addition to batch filtration of the atmospheric leach discharge slurry, bench-scale solid-liquid-separation characterization testing was completed, including static settling, vacuum filtration, pressure filtration, PLS clarification. These tests achieved acceptable thickening rates, vacuum/pressure filtration rates and PLS clarity. Based on the results, an atmospheric leach discharge thickener and PLS polishing filter was chosen to generate PLS for downstream SX and recycle thickened solids back to POX.

### 1.6.3 2024 Bench Testwork

The 2023 bench testwork and the 2024 pilot plant both used concentrated oxygen as the oxidant. Concentrated oxygen, rather than air, was selected due to bench-scale oxygen mass transfer limitations. To assess the potential for using air in industrial operations, additional testing was completed using air at the average pressure expected in a commercial-height atmospheric leach tank. Figure 1-6 and Figure 1-7 compare these results to atmospheric testing using oxygen. While oxidation using oxygen leads to a higher solution pH and more rapid iron removal, the oxidation rates achieved with air indicate that its commercial use is technically viable.



**Figure 1-6: Iron Removal in Atmospheric Leach using Air vs. Oxygen**



**Figure 1-7: Solution pH in Atmospheric Leach using Air vs. Oxygen**

## 1.7 Copper Cementation

Near-complete copper extraction is achieved in POX, while complete copper precipitation, via cementation with awaruite, is achieved in atmospheric leach. This creates a circulating load of copper that is managed by treating a portion of the ILS through a cementation operation.

### 1.7.1 2023 Testwork Summary

The 2023 testwork program successfully demonstrated copper removal using awaruite concentrate as a reducing agent. This produced a copper cement grading 66% Cu, 24% Ni with a residual copper concentration in solution of 0.8 g/L.

### 1.7.2 2024 Pilot Plant

In the 2024 testwork program, the cementation reagent was changed to iron dust to eliminate the potential for nickel loss in the copper cement from unreacted awaruite. In the pilot plant, copper was removed from the circuit by batch-treating the ILS using iron dust. A stoichiometric addition rate of iron dust precipitated greater than 99.9% of the copper down to 0.001 g/L. In the commercial design, copper concentrations feeding atmospheric leach will be controlled to 0.5 g/L by treating approximately 50% of the ILS with sponge iron. Higher concentrations of copper have been shown to reduce atmospheric leach neutralization kinetics. Sponge iron was selected for the refinery study because of its lower cost and greater availability than iron dust. While there is precedent for the use of sponge iron for copper cementation, additional testing is recommended to confirm the required dosage rates and the resulting copper cement grade.

## 1.8 Aluminum and Chromium Removal

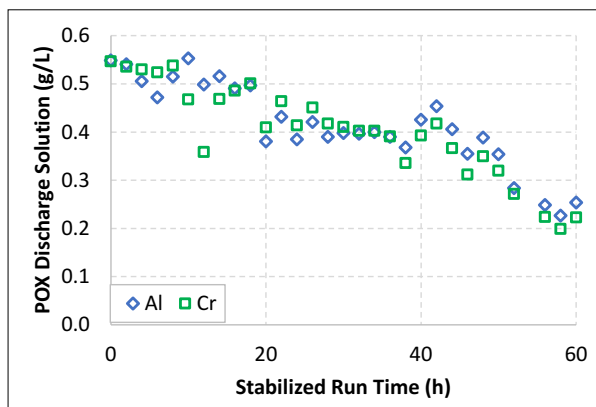
Trace quantities of aluminum and chromium are extracted in POX, with Ni/Al and Ni/Cr ratios in pressure leach solutions of 3,600:1 and 1,800:1, respectively. These elements fully precipitate in atmospheric leach and are recycled back to POX, creating a circulating load.

### 1.8.1 2023 Testwork Summary

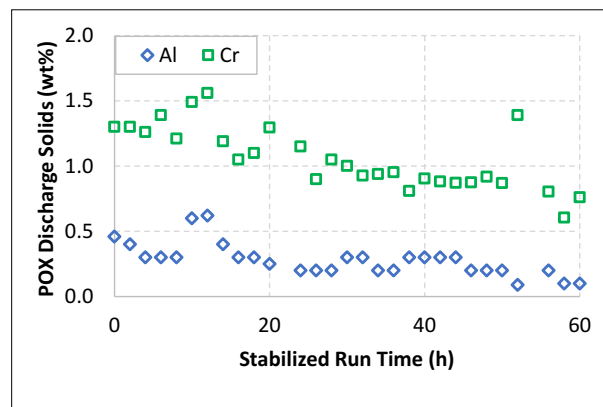
Testing in 2023 envisioned aluminum and chromium removal by selectively leaching these elements from a small percentage of the atmospheric leach residue. This testwork is no longer relevant, as it was superseded by pilot plant testwork that demonstrated that these unit operations are not required.

### 1.8.2 2024 Pilot Plant

To assess the potential for aluminum and chromium to build up in the leaching circuit, the pilot plant initially ran using synthetic ILS solutions, spiked with ~1 g/L each of aluminum and chromium. The behaviour of these elements in POX are summarized in Figure 1-8 and Figure 1-9. These show concentrations of both elements decreasing over time in both the POX discharge solution and solids, as the spiked quantities of these elements are removed from the circuit. This demonstrates that, under steady-state conditions, these elements concentrate only to a limited extent and ultimately report to the POX residue.



**Figure 1-8: Aluminum and Chromium Concentrations in the POX Discharge Solution**



**Figure 1-9: Aluminum and Chromium Concentrations in the POX Discharge Solids**

## 1.9 Cobalt Solvent Extraction

### 1.9.1 2023 Testwork Summary

The 2023 bench CoSX program was completed using a phosphinic acid extractant and sodium hydroxide as the neutralizing reagent. Testing included successive batch organic contacts of the PLS to produce cobalt-free raffinate for NiSX tests, and extractions at various aqueous-to-organic (A:O) ratios to generate

an isotherm and engineering data. Three successive organic contacts (10% v/v Cyanex 272, 55°C, 1:1 A:O, pH 5.4) of the PLS produced a NiSX feed containing <1 mg/L Co.

## 1.9.2 2024 Bench Testwork

A portion of the pilot-plant PLS was purified through bench-scale CoSX and NiSX to produce nickel sulphate crystals. The composition of the PLS used is presented in Table 1-4.

**Table 1-4: PLS Solution Composition**

Al (g/L)	Ca (g/L)	Co (g/L)	Cr (g/L)	Cu (g/L)	Fe (g/L)	Mg (g/L)	Mn (g/L)	Ni (g/L)	Si (g/L)	Zn (g/L)	pH
<0.003	0.019	2.30	<0.0006	0.001	<0.001	1.61	0.02	111.0	<0.002	0.01	5.0

CoSX testing was completed in a similar manner to the 2023 testing, but used ammonium hydroxide instead of sodium hydroxide as the neutralizing reagent. Three successive organic contacts (10% v/v Cyanex 272, 55°C, 1:1 A:O, pH 5.4) of the PLS produced a NiSX feed with <1 mg/L Co. These conditions resulted in 1.5% co-extraction of nickel and 77% co-extraction of magnesium. These co-extractions are expected to drop significantly during continuous CoSX operation, which would include an organic scrubbing stage. Modelling by the Cyanex 272 provider on a 3E-2Sc-1S circuit indicates that net nickel extraction can be virtually eliminated, while net magnesium extraction can be limited to 0.4 to 1.4%, depending on the quantity of cobalt recycled to scrubbing.

## 1.10 Cobalt Precipitation

Cobalt precipitation from concentrated CoSX strip solution has been demonstrated, using either sodium hydroxide or sodium carbonate. The 2023 testwork program demonstrated >99% cobalt recovery using sodium hydroxide to produce a 40% Co precipitate. The 2024 testwork program demonstrated >99% cobalt recovery using sodium carbonate to produce a 50% Co precipitate. Neither reagent schemes are selective against zinc, copper, or manganese, all of which co-precipitate with the cobalt. Both reagent schemes are partially selective over magnesium.

Either reagent scheme is technically acceptable. Sodium carbonate was selected for this study due to reduced reagent safety considerations, an expectation of improved solid-liquid-separation properties, and the higher cobalt grade of the precipitate.

## 1.11 Nickel Solvent Extraction

### 1.11.1 2023 Testwork Summary

The 2023 bench NiSX program was completed using a neodecanoic acid extractant and sodium hydroxide as the neutralizing reagent. Testing included a single-stage extraction with CoSX raffinate, to produce loaded organic. Extractions at various A:O ratios were also completed to generate an isotherm and engineering data. A single-stage extraction (20%v/v Versatic 10, A:O 0.2, 55°C, pH 6.6) achieved approximately 80% nickel extraction with a Ni:Mg ratio on the loaded organic of 68,000:1.

The loaded organic from the single-stage extraction was washed with water to remove entrained soluble impurities, and then acid-stripped to produce a concentrated nickel sulphate solution for crystallization.

### 1.11.2 2024 Bench Testwork

NiSX testing was completed in a similar manner to the 2023 testing, but used ammonium hydroxide instead of sodium hydroxide as the neutralizing reagent. Using CoSX raffinate, a single-stage extraction (20%v/v Versatic 10, A:O 0.2, 55°C, pH 6.6) achieved approximately 67% nickel extraction with a Ni:Mg ratio on the loaded organic of 64,000:1.

Continuous NiSX testing is recommended in the next phase of study, to confirm the circuit configuration and conditions required for acceptable magnesium rejection. Additional isotherm work is also recommended using ammonium hydroxide.

The change in SX reagent from sodium hydroxide to ammonium hydroxide introduces the potential for formation of double salts of ammonium sulphate and nickel sulphate. To investigate this, cooling crystallization tests were conducted on solutions of NiSX feed and NiSX intermediate raffinate. Ammonium sulphate concentrations were spiked to levels higher than expected industrially, to represent worst-case conditions. The results, summarized in Table 1-5, indicate the potential for crystallization in the intermediate NiSX raffinate at 40°C. This is well below the NiSX operating temperature of 55°C, and thus provides a safe operating window to avoid crystallization.

**Table 1-5: Double Salt Crystallization Test Results**

Solution	Ni (g/L)	Mg (g/L)	(NH <sub>4</sub> ) <sub>2</sub> SO <sub>4</sub> (g/L)	Crystallization Onset (°C)
NiSX Feed	95	2.2	12	<20
NiSX Intermediate Raffinate	38	1.5	176	40

## 1.12 Nickel Sulphate Crystallization

### 1.12.1 2023 Testwork Summary

Strip solutions from NiSX were used in batch crystallization tests. A crystallization temperature of 50°C was used to ensure formation of alpha nickel sulphate hexahydrate, the preferred form of nickel sulphate for battery applications. Successive crops of nickel sulphate crystals were obtained in a semi-continuous manner. Crystals meeting target nickel sulphate specifications were successfully produced.

### 1.12.2 2024 Bench Testwork

Crystallization testing was completed using feed solutions generated from the pilot-plant PLS treated through CoSX and NiSX (using ammonium hydroxide). Testing was conducted in a similar manner to the 2023 testwork. An elevated zinc concentration of 1.8 mg/L was noted in the NiSX strip solution. Complete zinc removal is expected in CoSX, and it was not evident if the zinc occurred as a result of sample contamination or another transfer mechanism. Zinc concentration was successfully reduced to <0.5 mg/L through batch treatment with a D2EHPA-impregnated ion-exchange resin.

The specifications of crystals from two crops of crystallization are presented in Table 1-6, and show that both crops meet target specifications for battery-grade nickel sulphate.

**Table 1-6: Nickel Sulphate Crystallization Results Summary**

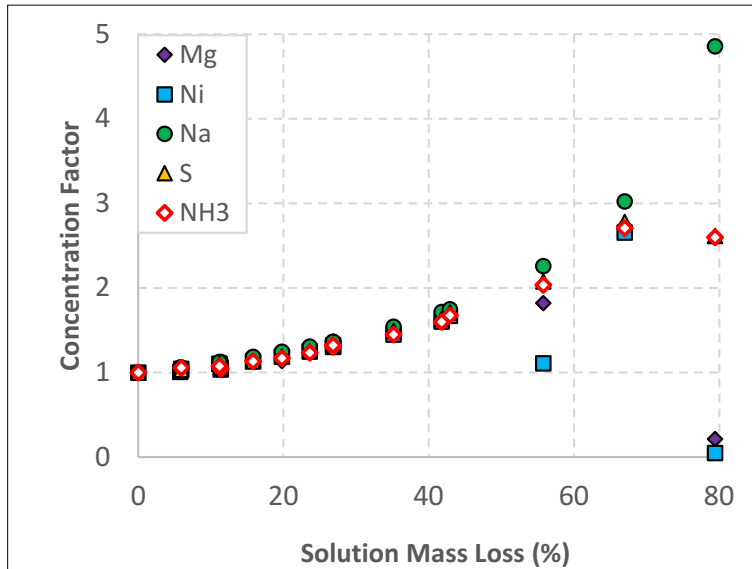
Element	Unit	Crop 1	Crop 2	Target Specification
Nickel	%	23.1	22.7	>22
Sulphur	%	12.4	12.8	N/A
Aluminum	ppm	0.3	1.3	<5
Arsenic	ppm	<0.1	<0.1	<2
Calcium	ppm	1	1	<5
Cadmium	ppm	<0.1	<0.1	<1
Cobalt	ppm	1.2	0.9	<50
Chromium	ppm	<0.1	<0.1	<3
Copper	ppm	0.6	0.6	<3
Iron	ppm	<1.0	<1.0	<3
Potassium	ppm	1.0	1.0	<10
Magnesium	ppm	3.0	4.0	<5
Manganese	ppm	0.3	0.5	<5
Sodium	ppm	15	17	<20
Lead	ppm	<0.1	<0.1	<2
Silicon	ppm	<5.0	<5.0	<10
Zinc	ppm	0.1	0.1	<5

### 1.13 Evaporative Concentration

Evaporation concentration of the NiSX raffinate was introduced in the 2024 bench testwork program. This unit operation does not increase the total evaporation load of the circuit, but shifts evaporation load away from the ammonium sulphate crystallization. Evaporation concentration improves the performance of the downstream nickel scavenging and magnesium precipitation unit operations by increasing the concentrations of nickel and magnesium, respectively.

To determine the desired evaporation extent, evaporative tests were completed to determine the onset of crystallization. Figure 1-10 shows the solution concentration factors during evaporation concentration. A low concentration of sodium was added to the solution to serve as a tracer element. Figure 1-10 indicates that no crystallization occurs up to 43% solution mass loss. This equates to 28 wt.%  $(\text{NH}_4)_2\text{SO}_4$  in solution. Crystallization was observed at 56% mass loss or 34 wt.%  $(\text{NH}_4)_2\text{SO}_4$  in solution.





**Figure 1-10: Solution Concentration Factors During Evaporative Concentration**

## 1.14 Nickel Scavenging Precipitation

### 1.14.1 2023 Testwork Summary

Residual nickel in the NiSX raffinate, containing elevated concentrations of sodium sulphate, was recovered using sodium hydroxide. Greater than 99% nickel recovery, down to a residual solution concentration of <3 mg/L, was achieved while limiting magnesium co-precipitation to 6%.

### 1.14.2 2024 Bench Testwork

The presence of elevated ammonium sulphate concentrations in the NiSX raffinate preclude the use of sodium hydroxide to precipitate nickel, due to significant neutralization of ammonium to ammonia. Instead, nickel scavenging precipitation tests were conducted using ammonium sulphide.

Precipitation testing at 80°C on concentrated NiSX raffinate, grading 0.95 g/L Ni, 3.4 g/L Mg and 28 wt.%  $(\text{NH}_4)_2\text{SO}_4$ , achieved terminal nickel concentrations of 0.02 g/L at 1.3 times the stoichiometric dosage of  $(\text{NH}_4)_2\text{S}$ . The resultant precipitate graded 53% Ni. As expected for sulphide precipitation, no co-precipitation of magnesium was observed. Re-leaching of the precipitate to recover nickel values was not tested, but these values can be reasonably expected to be high when recycled to POX.

## 1.15 Magnesium Removal

### 1.15.1 2023 Testwork Summary

Magnesium removal using sodium hydroxide was considered for a flowsheet using sodium hydroxide in SX. No testing was completed, because removal of magnesium down to <0.05 g/L through hydroxide precipitation is well-established industrially and the exact performance of this unit operation had limited impact on the overall circuit.

### 1.15.2 2024 Bench Testwork

The presence of elevated ammonium sulphate concentrations in the NiSX raffinate precludes the use of sodium hydroxide to precipitate magnesium, due to significant neutralization of ammonium to ammonia. Instead, magnesium removal tests were conducted using diammonium phosphate (DAP) to precipitate a monohydrate variant of struvite ( $\text{NH}_4\text{MgPO}_4 \cdot \text{H}_2\text{O}$ ).

Precipitation testing conducted at 80°C on concentrated nickel scavenging discharge solution, grading 3.6 g/L Mg and 33 wt.%  $(\text{NH}_4)_2\text{SO}_4$ , achieved terminal magnesium concentrations of 0.16 g/L at a stoichiometric dosage of DAP. The terminal phosphorous concentration was 0.4 g/L, and the resulting precipitate graded 16% Mg, 19% P, and 8.4% N. A marketing study is recommended to investigate the potential to valorize the struvite precipitate as a fertilizer product.

## 1.16 Ammonium Sulphate Crystallization

### 1.16.1 2023 Testwork Summary

The 2023 testwork is not relevant to the current refinery flowsheet, because it used sodium hydroxide as the SX neutralizing reagent, resulting in sodium sulphate as the salt byproduct.

### 1.16.2 2024 Bench Testwork

No crystallization testing was completed, because crystallization of ammonium sulphate is well-established industrially. The expected composition of the ammonium sulphate crystals can be estimated based on the resulting composition produced in the magnesium removal testwork, assuming that all elements in the discharge solution report to the ammonium sulphate crystals. The estimated composition of the ammonium sulphate crystals is >21.0 % N, 0.1% P, <0.05% Mg, and <0.01% Ni.

Additional nickel scavenging and magnesium removal testwork, followed by ammonium sulphate crystallization, is recommended to confirm the expected grade of the ammonium sulphate crystals and the associated valorization of such as a fertilizer product.

## 2. Refinery Design

### 2.1 Overview

This study addresses the processing of awaruite concentrate in a dedicated refinery to produce 32,000 metric tonnes per year (tpy) of nickel contained in nickel sulphate hexahydrate crystals. The refinery will also produce several valuable co-products, including cobalt carbonate, copper cement, and ammonium sulphate crystals.

It is assumed that awaruite concentrate will arrive at the refinery by highway truck. At the refinery, the concentrate will be processed to extract the nickel, cobalt, and copper through atmospheric and pressure leach circuits. Iron will be rejected with the pressure leach residue, and copper will be recovered in a cementation process. The pregnant leach solution (PLS) will be processed through sequential cobalt and nickel solvent extraction (SX) to produce concentrated and low-impurity liquors for cobalt carbonate precipitation and nickel sulphate hexahydrate crystallization, respectively. Magnesium will be recovered

from the remaining leach solution to form a struvite precipitate, and ammonium sulphate will be recovered as crystals.

The leach residue, consisting primarily of hematite/goethite, and other precipitate species will be transported as a wet filter cake for off-site disposal at a regulated facility.

## 2.2 Process Design Basis

FPX conducted a series of laboratory tests, at a bench and pilot scale (see Section 1) to develop a sulphuric acid leach system that could achieve high nickel recoveries while producing a PLS with low free-acid and iron concentrations. Purification of the PLS will use established and industrially proven technologies to produce battery-grade nickel sulphate crystals from awaruite concentrate. The testwork demonstrated the technical viability of using an awaruite-rich concentrate as feedstock in the EV battery material supply chain. Wood's scoping study used an ammonia-based neutralization scheme instead of caustic; this generates a valuable ammonium sulphate byproduct instead of sodium sulphate, which is a potential waste disposal liability.

The testwork demonstrated the process steps to produce nickel sulphate hexahydrate crystals from the awaruite concentrate, and established the base-case flowsheet for this scoping study.

The plant design basis is summarized in Table 2-1.

**Table 2-1: Refinery Plant Design Basis**

Parameter	Units	Value
Design capacity – Ni in NiSO <sub>4</sub> ·6H <sub>2</sub> O	tpy	32,000
Refinery operating hours per day	hours	24
Product packaging operating hours per day	hours	12
Refinery availability	%	92
Refinery operating hours per year	hours	8,059
Design feed rate (dry)	tph	6.69

## 2.3 Process Design Criteria

Comprehensive process design criteria were developed for the study. These were based predominantly on results from metallurgical testing, supplemented with project database information, industry standards and vendor recommendations. The following metallurgical testwork was used as the basis of the process design criteria:

- atmospheric leach and pressure leach
- atmospheric leach and pressure leach solid-liquid separation
- copper cementation
- cobalt and nickel SX
- cobalt carbonate precipitation
- nickel sulphate hexahydrate crystallization

- nickel SX raffinate concentration
- nickel scavenging precipitation
- magnesium (struvite) precipitation.

Selected process design information is summarized in Table 2-2.

**Table 2-2: Summary of Major Process Design Criteria**

Parameter	Units	Value
<b>Awaruite Feed Composition</b>		
Nickel	% w/w	60
Iron	% w/w	25
Cobalt	% w/w	1.1
Copper	% w/w	0.5
Magnesium	% w/w	1.7
<b>Atmospheric Leach</b>		
Leach residence time	h	5
Oxygen source		Air
Discharge pH	pH	>4
<b>Leach extraction extent</b>		
Ni	%	~42
Co	%	~50
Atmospheric leach thickener diameter	m	22
<b>Pressure Leach</b>		
Number of autoclaves	#	1
Number of compartments	#	5
Leach residence time	min	60
Operating temperature	°C	150
Oxygen over pressure	kPa	375
Discharge flash stages	#	1
<b>Combined atmospheric and pressure leach extraction</b>		
Ni	%	99.3
Co	%	98.0
Cu	%	92.0
Mg	%	75.0
<b>Pressure Leach Residue Solid Liquid Separation</b>		
Residue and CCD circuit – High-rate thickeners installed	#	4
Thickener diameter	m	7

Parameter	Units	Value
<b>Residue Filtration – Plate and Frame Pressure Filter</b>		
Cake moisture	% w/w	26
Cake solids dry bulk density	kg/m <sup>3</sup>	1,595
<b>Copper Cementation</b>		
Target ILS Cu concentration	g/L	0.01
Cementation Reagent		Sponge Iron
Product Cu Concentration	% w/w	70-80
<b>Cobalt Solvent Extraction</b>		
Plant configuration: extraction-scrub-strip	-	4 - 2 - 2
Settler specific throughput	m <sup>3</sup> /h.m <sup>2</sup>	5
Extractant used		Cyanex 272
Extractant concentration	% v/v	10
Strip feed solution		Dilute sulphuric acid
Strip feed solution acid concentration	g/L	90
<b>Cobalt Recovery</b>		
Reactant		Sodium carbonate
Cobalt recovery filter		Horizontal vacuum belt filter
Filter cake moisture content	% w/w	20
<b>Nickel Solvent Extraction</b>		
Plant configuration: extraction-scrub-strip-versatic recovery		4 - 1 - 3 - 1
Settler specific throughput	m <sup>3</sup> /h.m <sup>2</sup>	5
Extractant used		Versatic 10
Extractant concentration	% (v/v)	40
Strip feed solution		Dilute sulphuric acid
Strip feed solution acid concentration	g/L	250
<b>Nickel Sulphate Crystallization</b>		
Nickel concentration in final product - minimum	% w/w	22.3
Nickel in final product as NiSO <sub>4</sub> .6H <sub>2</sub> O - minimum	% w/w	99.87
Type		Forced circulation
Nickel sulphate dryer		Fluidized bed
Product target free moisture content	% w/w	<0.2
<b>Nickel Concentration and Scavenging</b>		
Concentration method		Falling film evaporator
Ammonium sulphate concentration	% w/w	32
Nickel scavenging precipitation reactant		Hydrogen sulphide

Parameter	Units	Value
<b>Magnesium (Struvite) Precipitation</b>		
Reactant	-	Diammonium phosphate
<b>Ammonium Sulphate Crystallizer</b>		
Ammonium sulphate concentration in final product - minimum	% w/w	>98
Type		Draft tube baffle with MVR
Operating pressure		Atmospheric
Ammonium sulphate dryer		Fluidized bed dryer
Product target free moisture content	% w/w	<0.1

## 2.4 Block Flow Diagram

Wood developed a block flow diagram of the major process steps. This is presented in Figure 2-1.

## 2.5 Refinery Layout

Wood developed a site layout for the refinery and required infrastructure. This is presented in Figure 2-2.

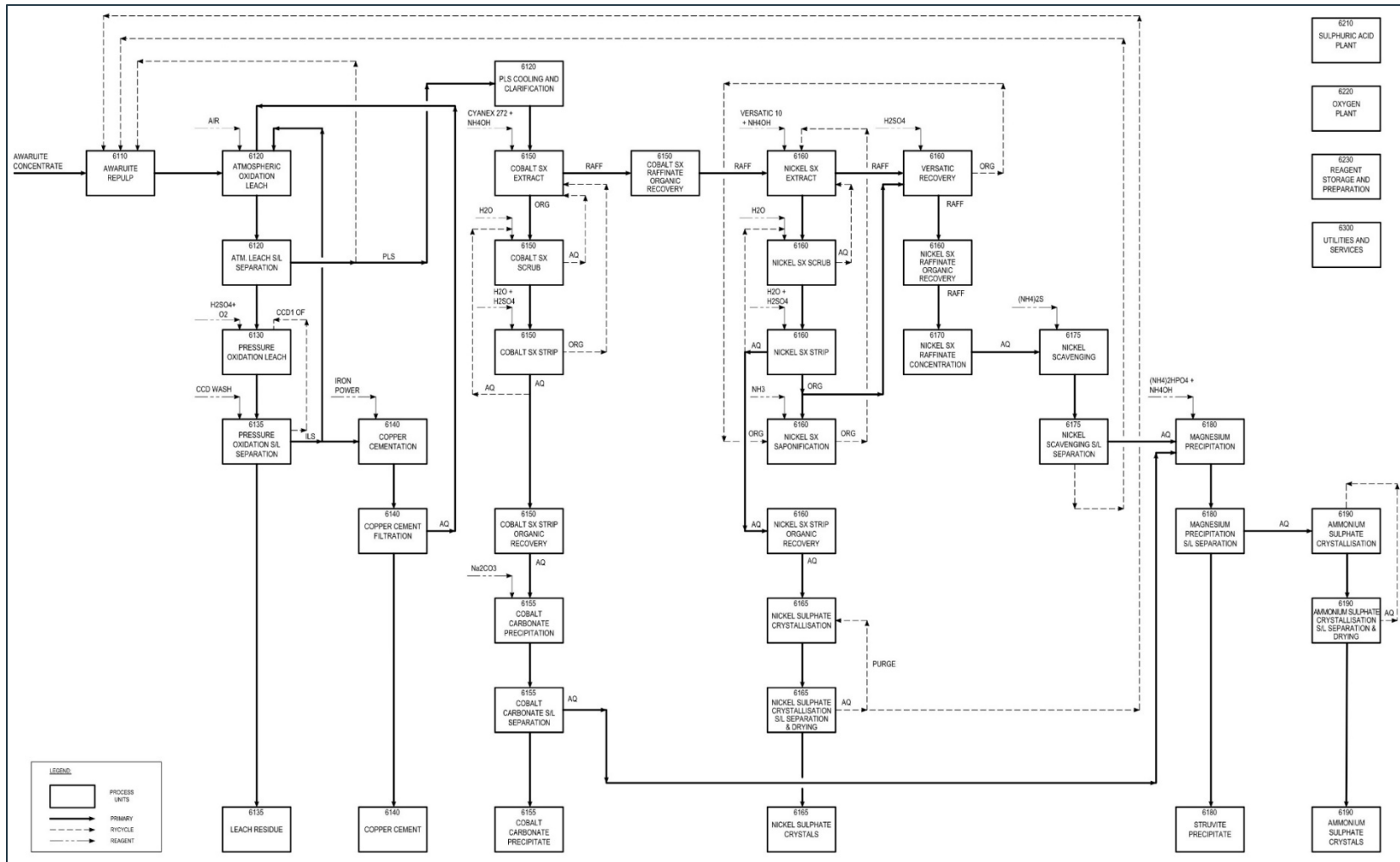


Figure 2-1: Process Block Flow Diagram

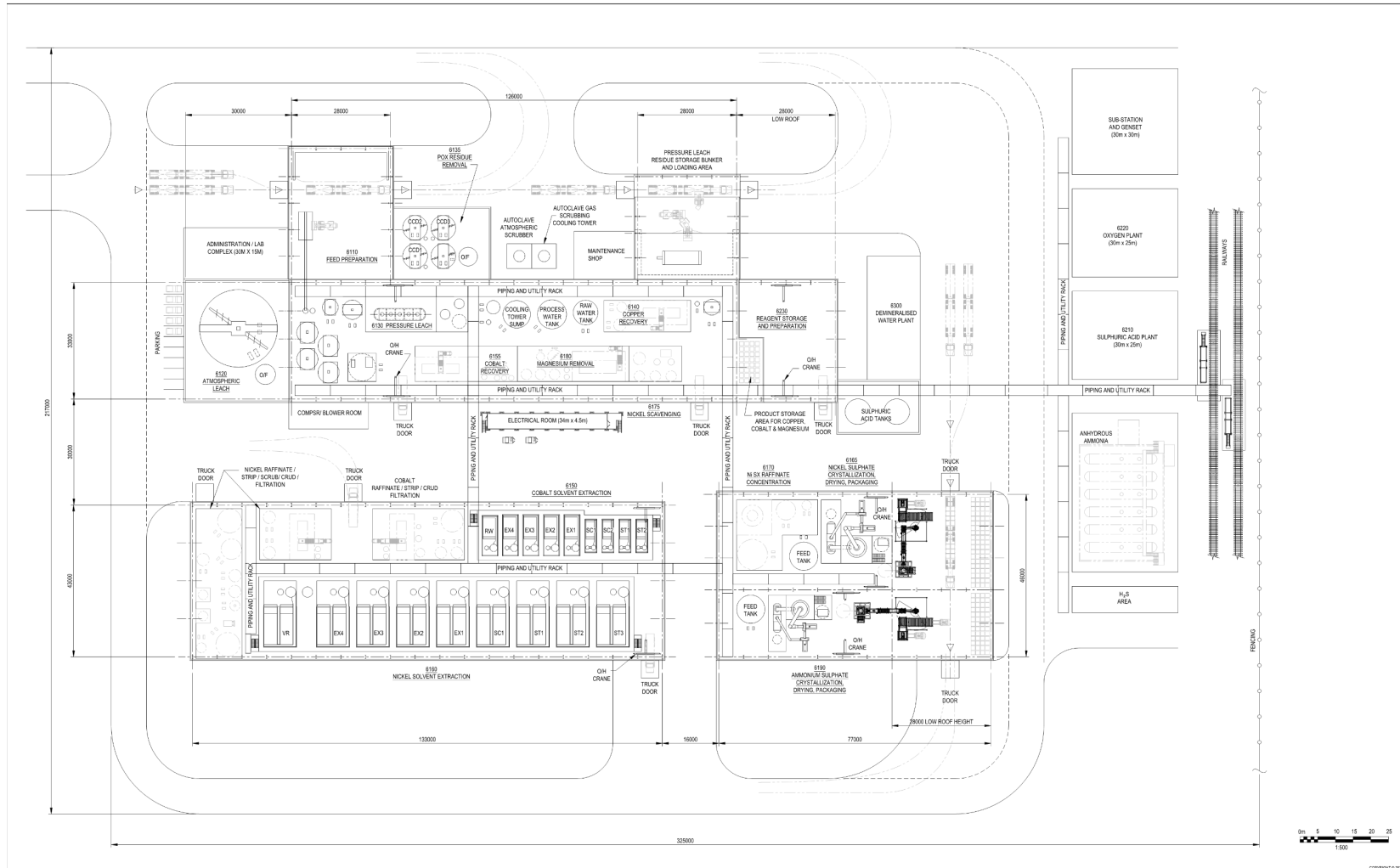


Figure 2-2: Refinery Layout



## 2.6 Process Description

### 2.6.1 Feed Preparation

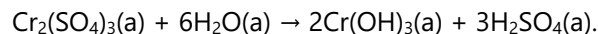
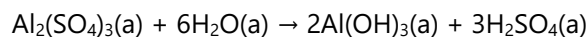
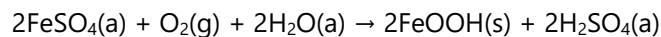
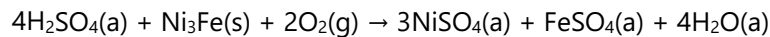
Awaruite concentrate will arrive at the refinery in highway trucks, and be tipped directly into a covered stockpile area. The concentrate will be reclaimed from the stockpile with a front-end loader, and loaded into a concentrate bin. A screw conveyor will convey the concentrate to an attrition scrubber at a feed rate of 6.69 dry tonnes per hour (tph).

A recycle stream of PLS, nickel crystallizer bleed and nickel scavenging precipitate will be mixed with the concentrate in an attrition scrubber to repulp the concentrate. The repulped slurry will overflow the attrition scrubber into a concentrate slurry tank, where the slurry will be conditioned to 30% solids concentration. The concentrate slurry tank will provide surge capacity for the downstream process.

### 2.6.2 Atmospheric Leach

The awaruite concentrate slurry will be partially leached in a series of four agitated leach tanks, with intermediate leach solution (ILS) recycled from the pressure leach. Air will be sparged into the leach tanks to provide oxygen for oxidation. Under the atmospheric leach conditions, awaruite leaching will continue up to pH 4-5. The awaruite will thus function as an effective neutralizing agent to neutralize free acid and then precipitate iron, aluminum and chrome as hydroxides. Awaruite will also reduce copper in solution to form copper cement.

The main reactions in the atmospheric leach are:

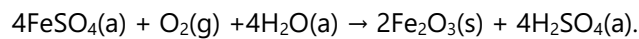
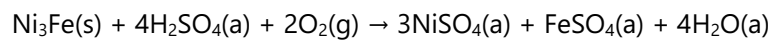


Slurry discharge from the final leach tank will flow by gravity to the atmospheric leach thickener for solid-liquid separation. The thickener will overflow into a thickener overflow tank. A portion of the PLS will be recycled back to the concentrate repulp circuit. The balance of the PLS will be pumped through a polishing filter to remove suspended solids, before discharging into the cobalt SX feed tank. The thickener underflow slurry will be pumped to the autoclave feed tank.

### 2.6.3 Pressure Leach

The atmospheric leach discharge slurry will be pumped from the autoclave feed tank to a five-compartment autoclave. Concentrated sulphuric acid will be added into the autoclave to target a residual acid concentration of 25 g/L in the autoclave discharge slurry. Oxygen will be sparged into each compartment at an overpressure of 375 kPa. The autoclave will operate at a temperature of 150°C and a pressure of 930 kPa, with a total retention time of 60 minutes.

The nickel, cobalt, copper, iron, chromium, aluminum and magnesium will leach into solution. The precipitated iron, aluminum and chrome hydroxides will re-dissolve and re-precipitate in the autoclave as goethite or hematite species. Under steady-state conditions, all dissolved iron, aluminum and chrome will eventually report to the pressure leach residue. The main reactions in the autoclave are:



The autoclave will discharge to a flash vessel, where the resulting pressure drop will flash water to steam, and the slurry will collect in the autoclave seal tank. The slurry in the autoclave seal tank will be pumped to the pressure leach discharge thickener.

The flash vessel vapour will go through a gas cyclone to remove entrained solids. The cyclone overflow gas will pass through a quench vessel, where process water will quench the hot gas. The quench vessel will vent to the atmospheric scrubber, which will use process water to scrub the gas before finally venting to the atmosphere.

The cyclone underflow solids will report to the autoclave seal tank. The quench tank and atmospheric scrubber will discharge the liquid phases to the autoclave cooling tower, where the liquid will be cooled before being recycled back. Accumulated solids will be bled off to the autoclave cooling water circuit. Make-up process water will be supplied to the cooling tower sump.

Cooling water for the autoclave will be wash solution from the pressure leach residue counter-current decantation (CCD) circuit, with make-up process water as required.

### 2.6.4 Pressure Leach Residue

Underflow from the pressure leach residue thickener will be washed in a series of three CCD thickeners to recover dissolved nickel values. Process water will be used as CCD wash water, pumped to CCD 3, and cascaded down to CCD 1. The CCD 1 overflow will discharge into the autoclave cooling water tank, and be recycled back to the autoclave as cooling water. The underflow slurry from CCD 3 will be pumped to a residue filter.

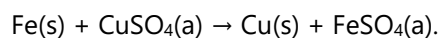
ILS overflow from the pressure leach residue thickener will discharge into the pressure leach thickener overflow tank. The ILS will be split into two streams, at nominally equal flow rates. One stream will be pumped directly back to the atmospheric leach circuit. The second stream will advance through the copper cementation process to remove copper from solution, after which that stream will also return to the atmospheric leach circuit. The combined ILS copper concentration should not exceed 0.5 g/L. The ILS proportion to copper cementation can be varied to control the copper concentration.

Underflow slurry from CCD 3 will be transferred to a residue filter feed tank. The slurry will be pumped through a plate and frame pressure filter, to produce a residue filter cake. The filter cake, which is mostly an iron-based residue, will be temporarily stored in a bunker onsite before being loaded onto trucks for haulage off-site for permanent storage in a regulated facility.

The filter filtrate will be recycled back to CCD 3 feed.

### 2.6.5 Copper Recovery

ILS will be pumped from the pressure leach discharge thickener overflow tank into a reactor tank. Sponge iron or iron powder will be added to the reactor where it will react with the copper in solution to reduce the copper to a copper cement. The main reaction is:

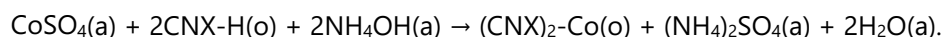


Slurry from the in-line reactor will advance to the copper cementation filter press. The copper and residual iron solids will be retained, and the cementation reaction will continue in the filter press. When a filter-press pressure setpoint is reached, the press will discharge the filter cake into a screw conveyor, which will in turn discharge the copper cement product through a discharge chute into a bulk bag. The bulk bag will be manually removed and replaced periodically when it has been filled to capacity. The target copper tenor in the ILS advancing to atmospheric leach is 0.01 g/L Cu.

### 2.6.6 Cobalt Solvent Extraction

PLS from the atmospheric leach thickener overflow tank will advance through a heat exchanger to cool ahead of SX, then through the PLS polishing filter to remove any remaining suspended solids before discharging into the cobalt SX feed tank.

The cobalt SX circuit will comprise nine mixer-settlers arranged in series: four extraction, two scrub, two strip, and one diluent wash mixer-settler. The PLS will contact the stripped organic phase, which consists of Cyanex 272 extractant at a concentration of 10% by volume in Shellsol 2046 diluent, in counter-current flow to transfer the cobalt from the aqueous to the organic phase. The pH in each extraction step will be controlled at 5.4, using dilute ammonium hydroxide. The cobalt loads on the extractant and the acid released by the extractant will react with the ammonium hydroxide to form ammonium sulphate. Minor amounts of magnesium, manganese, copper, nickel and zinc will also be extracted by the organic. The main cobalt extraction reaction is as follows:



The cobalt-loaded organic will advance to the two scrub stages. A scrub solution, consisting of a recycle stream of loaded strip liquor with a water make-up, will be contacted with the loaded organic in the scrub mixer-settlers. Nearly all the co-extracted nickel and a portion of the magnesium will be scrubbed from the organic, and any ammonium contained in the entrained aqueous will be washed out. Some cobalt will also be scrubbed during this stage, so the scrub raffinate will be recycled back in to the extraction stages to recover the cobalt.

The scrubbed organic will then advance to the two strip stages. In these stages, the loaded organic will be contacted with 90 g/L sulphuric acid strip solution to remove the cobalt from the organic and transfer it to the loaded strip liquor. The loaded strip liquor will also contain minor amounts of manganese, zinc, and copper originally present in the PLS in trace concentrations.

The loaded strip liquor will be pumped through a multi-stage organic recovery system to recover any entrained organic that could contaminate the cobalt product. The strip liquor will advance to the cobalt recovery circuit.

Cross-contamination of nickel SX extractant with Cyanex 272 must be prevented to maintain nickel over magnesium selectivity in the nickel SX, so the extraction raffinate will be scrubbed of entrained organic through multiple steps.

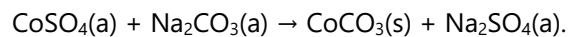
The extraction raffinate will discharge from the E4 extraction mixer-settler to the diluent wash mixer-settler. Diluent, without any extractant, will be contacted with the raffinate and effectively dilute the entrained extractant to a very low concentration. Any entrained organic remaining in the diluent scrubbed raffinate will contain only trace amounts of Cyanex 272.

The diluent scrubbed raffinate will be pumped through a coalescer and multimedia filter to recover entrained organic. To further reduce the risk of organic carry-over, the raffinate will also pass through an activated carbon filter before discharging into the nickel SX feed tank.

Any crud generated in the cobalt SX circuit will be collected and treated in a crud tank. The organic, aqueous and solid fractions will be separated in a crud filter press. The recovered aqueous and organic phases will be returned to the process, and the remaining solids collected for disposal.

### **2.6.7 Cobalt Recovery**

The cobalt SX strip liquor will discharge into the first of three precipitation tanks arranged in series. The strip solution will be treated with sodium carbonate to pH 6.5 in three precipitation tanks. This will cause the cobalt to precipitate as cobalt carbonate. The primary reaction is as follows:



The cobalt precipitate will be thickened in the cobalt precipitation thickener. A portion of the underflow will be recycled back to the first precipitation tank, to provide seed for the cobalt precipitation. The remainder of the thickener underflow will be pumped to a cobalt precipitation filter tank that feeds a horizontal vacuum belt filter. The precipitate will be washed and dewatered on the belt filter. The filter cake will discharge into a packaging system that loads the cobalt carbonate into lined bulk bags, ready for shipping off-site for sale.

The belt filter filtrate and wash solution will be returned to the cobalt precipitation thickener. The cobalt precipitation thickener will overflow into a tank, and will then be pumped to the magnesium precipitation circuit.

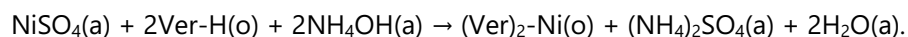
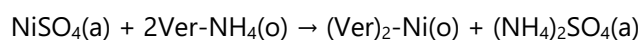
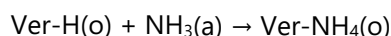
### **2.6.8 Nickel Solvent Extraction**

The organic depleted cobalt SX raffinate will discharge into the nickel SX feed tank. The nickel SX feed liquor will be pumped to the first extraction mixer-settler.

The nickel SX circuit will comprise nine mixer-settlers arranged in series: four extraction, one scrub, three strip, and one versatic acid recovery mixer-settler. The nickel SX feed liquor will contact the stripped and saponified organic phase, which consists of Versatic 10 at a concentration of 40% by volume in Shellsol 2046 diluent, in counter-current flow to transfer the nickel from the aqueous to the organic phase. The pH in each extraction step will be controlled at 6.5 using dilute ammonium hydroxide. The nickel loads on the

extractant, and the acid released by the extractant, will react with the ammonium hydroxide to form ammonium sulphate. Versatic acid is highly selective for nickel over any impurities that may be in the nickel SX feed liquor.

The Versatic 10 will be saponified with ammonia. Approximately 70% of the nickel equivalent neutralization will be achieved, with the remaining 30% nickel equivalent neutralized by a dilute ammonium hydroxide solution in the extraction mixer-settlers, which allows for accurate pH control. To reduce water consumption and downstream evaporation requirements, anhydrous ammonia will be used for saponification rather than ammonium hydroxide. The saponification and primary extraction reactions are as follows:



The nickel-loaded organic will advance to the scrub mixer-settler to scrub co-extracted magnesium and trace levels of calcium, using a scrub solution prepared by diluting a recycle stream of loaded strip liquor. The scrubbed organic will be stripped using 250 g/L sulphuric acid strip solution to generate a near-saturated nickel-sulphate-loaded strip liquor with 160 g/L nickel.

The nickel-loaded strip liquor will be pumped through a coalescer and multi-media filter to remove entrained organic, and then polished in activated carbon. The polished loaded strip liquor will advance to the nickel crystallization process.

The final pH for nickel extraction with versatic acid will be ~6.5. At this pH, versatic acid is slightly soluble in aqueous solutions. To recover the versatic acid from the extraction raffinate, a stripped organic stream from the strip circuit will be diverted to the versatic recovery mixer-settler, where it will contact the extraction raffinate. Sulphuric acid will be added to reduce the versatic acid solubility in the aqueous phase. The organic phase from the settler will go to the saponification tank. The scrubbed raffinate will discharge into the raffinate tank, and then be pumped through a coalescer and multi-media filter to remove any entrained organic. The filtered raffinate will advance to the nickel raffinate evaporation circuit.

Any crud generated in the nickel SX circuit will be collected and treated in a crud tank. The organic, aqueous and solid fractions will be separated in a crud filter press. The recovered aqueous and organic phases will be returned to the process, and the remaining solids collected for disposal.

### **2.6.9 Nickel Sulphate Crystallization, Drying, Packaging**

The polished nickel SX strip liquor will discharge into the nickel crystallizer feed tank. The loaded strip liquor will be pumped through a pre-heating heat exchanger into the centrate tank, combining with centrifuge centrate, which will then be pumped into the forced-circulation crystallizer. The crystallizer will operate under vacuum to maintain a temperature of approximately 50°C, to produce the preferred alpha-hexahydrate nickel sulphate crystal product. The expected particle size distribution has a d50 of 800 µm. The crystallizer will be driven by a single stage of mechanical vapour recompression (MVR).

The crystal slurry will be discharged from the crystallizer, and washed and dewatered using a pusher-type centrifuge. Centrifuge centrate will discharge into the centrate tank, and be recycled back into the crystallizer.

Process vapour generated by the evaporation of water will evolve in the crystallizer vapour body. This vapour will flow through mist eliminators to separate liquid from vapour. The vapour will then exit the vapour body and enter the MVR. The compressed vapour will be de-superheated by direct contact with process condensate, and returned to the shell side of the main heat exchanger, providing the driving force for evaporation. Make-up steam from the sulphuric acid plant will be added when necessary. The process vapour will condense and discharge to the condensate tank. Non-condensable gases will be vented.

A portion of the condensate will be recycled for de-superheating the compressor discharge. The remaining condensate will be pumped to the process water circuit.

Washed crystals from the centrifuge will be conveyed from the centrifuge to a closed-cycle fluidized bed dryer. The dryer will use steam to dry the crystalline product to <0.2% moisture content. The nickel sulphate product will be discharged from the dryer to a packaging plant, where the product will be bagged in lined bags and prepared for shipping offsite for sale.

### 2.6.10 Nickel SX Raffinate Concentration

The scrubbed nickel SX raffinate will discharge into the nickel raffinate evaporator feed tank. The evaporator will concentrate the nickel sulphate and ammonium sulphate, bringing the ammonium sulphate close to saturation. This will improve the downstream process performance and reduce the required hydraulic capacity of the downstream nickel scavenging and magnesium precipitation circuits. It will also reduce the evaporative load on the ammonium sulphate crystallizer.

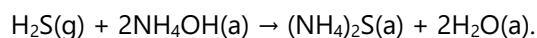
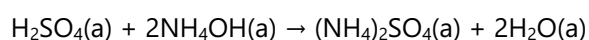
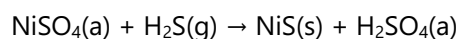
The evaporator feed will be pumped through a pre-heating heat exchanger and into the evaporator sump. The solution will then be recirculated through the internal recirculation system and over the evaporation tubes of the falling film evaporator. The evaporator will operate under a partial vacuum at approximately 60°C to evaporate water. The evaporator is steam-driven using a thermocompressor (TVR). The steam will be sourced from the sulphuric acid plant.

The evaporator brine will be pumped to the nickel scavenging circuit to recover nickel in solution.

The process vapour will exit the heater tubes and flow through de-misters to separate liquid from the vapour. The vapour will then flow to the TVR, where it will be compressed and then returned to the evaporator heater. Some make-up steam may be required to maintain the heat balance as non-condensable gases are vented. Process condensate will be collected in a condensate tank. A portion of the condensate will be used for de-superheating, and the remainder transferred into the process water circuit.

### 2.6.11 Nickel Scavenging

The raffinate evaporator brine will discharge into the first of three nickel scavenging tanks arranged in series. Hydrogen sulphide gas (H<sub>2</sub>S) will be added to the tanks to react with the nickel sulphate and precipitate nickel sulphide. The reaction will produce sulphuric acid, which will be controlled with the addition of ammonium hydroxide solution. The primary reactions are as follows:

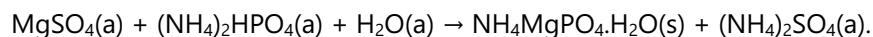


The precipitate slurry will overflow the third precipitation tank into the nickel scavenger thickener where the slurry is thickened. A portion of the thickener underflow slurry is recycled back to the first nickel scavenger tank to provide seeding for precipitation. The balance is pumped back to the concentrate repulp tank. The nickel sulphide will re-leach in the pressure leach circuit.

The nickel scavenger thickener overflow will be pumped through a polishing filter to remove suspended solids. The polishing filter will be periodically backwashed to the scavenger tanks. The filtered liquor will react with hydrogen peroxide to oxidize residual ammonium sulphide, before being pumped to the magnesium removal circuit.

### 2.6.12 Magnesium Removal

Filtrate from the nickel scavenging filter will advance to the magnesium precipitation circuit, which comprises three tanks arranged in series. The magnesium will be precipitated from the solution as a struvite species, by reacting the solution with diammonium phosphate (DAP). The primary reaction is as follows:



The slurry from the third magnesium precipitation tank will overflow to the magnesium precipitation thickener. A portion of the thickener underflow slurry will be recycled back to the first precipitation tank, to provide seeding for precipitation. The balance of the underflow slurry will be pumped to a magnesium filter feed tank that feeds a horizontal vacuum belt filter. The precipitate will be dewatered on the magnesium precipitate filter and discharged into a packaging system, where the struvite will be loaded into lined bulk bags, ready for shipping offsite for sale.

The magnesium precipitation filter filtrate will be recycled back to the magnesium precipitation thickener. The magnesium thickener overflow will advance to the ammonium sulphate crystallization circuit.

### 2.6.13 Ammonium Sulphate Crystallization, Drying, Packaging

The ammonium sulphate crystallizer feed tank will receive process liquor from the magnesium precipitation thickener. The process liquor will be pumped through a pre-heating heat exchanger and into the centrate tank, combining with centrifuge centrate, which will then be pumped into the draft tube baffle (DTB) crystallizer. The crystallizer will operate under atmospheric conditions to produce ammonium sulphate crystals. The expected particle size distribution has a d50 of 2000 µm, making this product suitable for sale to the fertilizer industry. The crystallizer will be driven by two stages of MVR.

The crystal slurry will be discharged from the crystallizer and dewatered using a pusher-type centrifuge. The centrifuge centrate will discharge into the centrate tank and be recycled back into the crystallizer.

The process vapour generated by the evaporation of water will flow through demisters to separate liquid from vapour. The vapour will then exit the crystallizer and enter the MVR stages arranged in series. The compressed vapour will be de-superheated by direct contact with process condensate and returned to the shell side of the heat exchanger, providing the driving force for the evaporation. Steam from the sulphuric acid plant will be necessary only for startup. The process vapour will condense and discharge to the condensate tank. Non-condensable gases will be vented.

A portion of the condensate will be recycled for de-superheating the compressor discharge. The remaining condensate will be pumped to the process water circuit.

The crystals from the centrifuge will be conveyed from the centrifuge to a closed-cycle fluidized bed dryer. The dryer will use steam to dry the crystalline product to less than 0.2% moisture. The ammonium sulphate product will be discharged from the dryer to a packaging plant, where the product will be bagged in lined bags and prepared for shipping offsite for sale.

## **2.6.14 Reagents**

### **2.6.14.1 Sulphuric Acid Plant**

Concentrated sulphuric acid (98%) from a molten sulphur feedstock will be produced onsite by an acid plant, which will be sized for 330 tonnes per day (tpd) on a 100% H<sub>2</sub>SO<sub>4</sub> basis. The sulphuric acid plant will also produce medium-pressure (20 bar) saturated steam that will be routed to the crystallizers, evaporators, product dryers and other areas of the process plant for heating. Wood selected an acid plant technology that uses concentrated oxygen, rather than air, for clean sulphur combustion and minimal footprint.

### **2.6.14.2 Oxygen**

Oxygen will be supplied to the plant by an over-the-fence supply agreement with a vendor. The vacuum pressure swing adsorption (VPSA) plant will be constructed and operated onsite by the vendor. The plant and liquid back-up system will be sized for 215 tpd of 93% purity oxygen, which will meet the oxygen requirements of the pressure leach and sulphuric acid plant.

### **2.6.14.3 Anhydrous Ammonia**

Liquid anhydrous ammonia will be delivered to site by road or rail tanker, and stored in six 30,000-gal ammonia bullets. The ammonia will be transferred to a vaporizer to convert it to ammonia gas before being distributed to ammonium hydroxide make-up, and to nickel SX for Versatic 10 saponification.

### **2.6.14.4 Ammonium Hydroxide**

Ammonium hydroxide will be prepared onsite from anhydrous ammonia, and transferred to the ammonium hydroxide dosing tank, from where it will be dosed to the cobalt SX, nickel SX, nickel scavenging and magnesium precipitation circuits as a pH modifier.

### **2.6.14.5 Flocculant**

Flocculant will be delivered to site as a liquid in IBC containers. The flocculant will be transferred to dosing tanks and will then be dosed to the process thickeners.

### **2.6.14.6 Shellsol 2046**

Shellsol 2046 will be used as the SX diluent in both the cobalt and nickel SX circuits. The diluent will be delivered to site in a bulk tanker and transferred to the diluent storage tank, from where it will be transferred to the cobalt SX and nickel SX organic make-up tanks.



#### **2.6.14.7 CYANEX 272**

Cyanex 272 will be used as the extractant in the cobalt SX circuit. The Cyanex 272 will be delivered to site in IBC containers, from where it will be transferred to the cobalt SX organic make-up tank.

#### **2.6.14.8 Versatic 10**

Versatic 10 will be used as the extractant in the nickel SX circuit. The Versatic 10 will be delivered to site in IBC containers, from where it will be transferred to the nickel SX organic make-up tank.

#### **2.6.14.9 Sodium Carbonate**

Anhydrous sodium carbonate will be used in the cobalt carbonate precipitation circuit as a pH modifier. The sodium carbonate will be delivered to site in a bulk tanker and transferred to the sodium carbonate storage silo. The sodium carbonate will be dissolved in water in the make-up system and transferred to the sodium carbonate dosing tank, from where it will be dosed to the cobalt carbonate precipitation circuit.

#### **2.6.14.10 Sponge Iron**

Sponge iron will be delivered to site in bulk bags by truck. The sponge iron will be used in the copper cementation circuit.

#### **2.6.14.11 Diammonium Phosphate**

Diammonium phosphate (DAP) will be delivered to site in a bulk tanker, and transferred to the DAP storage silo. The DAP will be dissolved in water in the make-up system and transferred to the DAP dosing tank, from where it will be dosed to the magnesium precipitation circuit.

#### **2.6.14.12 Hydrogen Sulphide**

Hydrogen sulphide (H<sub>2</sub>S) gas will be supplied to site in pressurized 1 ton cylinders. The H<sub>2</sub>S will be used in the nickel scavenging circuit for nickel sulphide precipitation.

#### **2.6.14.13 Hydrogen Peroxide**

Hydrogen peroxide will be delivered to site in a bulk tanker and transferred to the hydrogen peroxide storage tank, from where it will be dosed to the nickel scavenging filtrate tank.

### **2.7 Cobalt Carbonate vs Cobalt Sulphate Heptahydrate Trade-Off**

The base case design for the awaruite refinery will recover cobalt as a high-purity cobalt carbonate exceeding 48% Co. During the previous study phase, the option of producing high-purity battery grade cobalt sulphate heptahydrate was identified as a value-add opportunity.

The expected impurity levels in the Co SX strip will result in the cobalt sulphate not meeting minimum specifications, so additional purification will be necessary. The additional purification steps required to produce cobalt sulphate are outlined below:

- additional CoSX scrubbing stages to increase magnesium rejection

- addition of an impurity SX (ISX) circuit, using D2EHPA as the extractant, treating the CoSX strip solution to remove the majority of zinc and manganese impurities
- addition of an ion exchange (IX) circuit, using a D2EHPA impregnated resin, treating the ISX raffinate to further remove zinc and manganese impurities
- addition of a manganese precipitation step, using Caro's acid, treating the IX discharge solution for final manganese impurity removal.

The purified cobalt sulphate liquor containing only trace amounts of impurities, advances to a forced circulation crystallizer operating under vacuum to produce cobalt sulphate hexahydrate crystals that will meet EV battery manufacturer specifications.

Table 2-3 summarizes the capital and operating costs for the two evaluated options. For the purposes of this study, the lower-cost option of cobalt carbonate was selected. However, depending on refinery location, market conditions, strategic objectives, etc. the preferred cobalt product should be selected as part of the next study stage.

**Table 2-3: Cobalt Product Capital and Operating Cost Comparison**

Description	Unit	Cobalt Carbonate	Cobalt Sulphate
Total Capital Cost	\$	\$3,933,300	\$15,633,000
Operating Cost	\$/a	\$853,700	\$154,200

## 2.8 Opportunities and Recommendations

Several value-add opportunities have been identified during the course of the study, some of which were formally investigated and included. Other opportunities that are being carried forward, as well as other possible value-add initiatives, are as follows:

- An evaluation of autoclave materials of construction, based on the composition of the process fluids from testwork and simulation, supported the use of UNS S32750 instead of a brick lined autoclave. This resulted in a substantial capital cost saving. Corrosion testing should be conducted to confirm the selected materials, as well as investigate whether a lower grade alloy such as UNS S32205 would be acceptable.
- Corrosion testing should be conducted on the balance of plant to confirm the selected materials of construction.
- The onsite production of ammonium sulphide from ammonia, and hydrogen sulphide as a nickel precipitation reagent, should be considered to reduce the risks associated with hydrogen sulphide use within the process building.
- Depending on market conditions, selected site, strategic objectives, etc, the preferred cobalt product should be selected as part of the next study stage.
- Depending on the testwork results and flowsheet implications, the potential to market struvite as an additional byproduct.

### **3. Project Infrastructure**

#### **3.1 Site Preparation and Development**

Wood's cost estimate includes allowances for site preparation and development based on the mechanical general arrangement (GA) layout (see Figure 2-2).

As a specific site in central British Columbia has not been selected, Wood assumed a reasonably flat site in proximity of the required infrastructure. To estimate order-of-magnitude costs, it was assumed that clearing and grubbing will be required within the footprint of the developed site, followed by topsoil stripping and stockpiling. The site will be then graded to create a pad for the facility, while also addressing the site drainage requirements. All processing facilities will have internal spill containment areas, while the yard surfaces (non-contact water) will drain via ditches and culverts directly to the environment, assuming suitable water quality. This will be re-visited once a specific site and layout are defined.

Once the site has been selected, a geotechnical site investigation will be required to define the foundation subgrade preparation and the foundation type. For the Study, it was assumed that structural fill will be required from the foundations subgrade to the underside of slab-on-grade.

The facility will have a perimeter access road, internal roads and staging areas to provide access to all loading and unloading points, complete with drive-through arrangements for increased safety. All roads will be gravel-surfaced and will consist of a road subbase and road base, installed on a prepared subgrade.

Railroad spurs with unloading platforms and associated spill containments will be constructed at the edge of the refinery to facilitate the supply of various commodities for the operation.

The facility will be fenced off by security fencing, and access will be controlled via a gatehouse.

#### **3.2 Utilities**

Potable water, fire water and sanitary sewage piping will be installed in piping corridors between the buildings, and will be buried below the frost line. Process piping will run on pipe racks as much as practicable.

##### **3.2.1 Raw Water**

Raw water will be sourced from the existing municipal system and stored in the raw water tank. Raw water will be used and distributed throughout the plant as for gland/seal water, cooling tower make-up, and reagent mixing. The other water systems are described in the sections that follow.

##### **3.2.2 Process Condensate**

Process condensate from the nickel raffinate evaporator, nickel sulphate crystallizer and ammonium sulphate crystallizer processes will report to individual condensate tanks. The condensates will be used in specific applications in the process dependant on their purity (for example nickel crystallizer process condensate will be used to wash nickel sulphate crystals). Excess process condensate will report to the process water tank.

### **3.2.3 Process Water**

Excess process condensate and cooling-tower and boiler blowdown water will report to the process water tank, and will be distributed throughout the plant excluding SX/crystallizers areas where higher purity process condensate is used.

### **3.2.4 Demineralized Water / Boiler Feed Water**

Steam condensate will be returned to the boiler feed water tank in the sulphuric acid plant. A demineralization plant will produce demineralized water as make-up boiler feed water. Boiler feed water feeds the sulphuric acid plant boiler and the start-up boiler

### **3.2.5 Cooling Water**

Cooling water will be used in the process and condensate heat exchangers. Return cooling water will be cooled in forced-ventilation, direct-air-contact cooling towers. Cooling tower blowdown water will be sent to the process water tank.

### **3.2.6 Fire Water**

Fire water will be sourced from the local municipal supply.

### **3.2.7 Potable Water**

Potable water from the existing municipal system will be used for the administration and process areas.

### **3.2.8 Sanitary Sewage**

Sanitary (domestic) sewage will be collected throughout the site and conveyed via gravity piping and force mains, where required, to the local municipal sewage system for treatment and disposal.

### **3.2.9 Compressed Air**

Air compressors will provide high-pressure compressed air for use as plant air, air squeeze and cake blow on pressure filters. Compressed air will be dried and distributed as instrument air throughout the plant.

### **3.2.10 Blower Air**

Air blowers will provide low-pressure air for use in the atmospheric leach circuit as an oxygen source.

### **3.2.11 Steam**

Medium pressure (20 bar), saturated steam is produced from the sulphur-burning acid plant and provides motive steam for the NiSX raffinate evaporator, make-up steam for the MVR-driven crystallizers, and heating steam for product drying and the sulphuric acid plant. A start-up boiler will provide partial steam flow when the sulphuric acid plant is offline.

### 3.3 Power Supply and Distribution

#### 3.3.1 Power Supply Overview

Incoming power is assumed at 25 kV from the existing grid serviced by BC Hydro.

The main substation will step down the transmission line voltage from 25 kV to 4.16 kV using a single 10 MVA main power transformer complete with automatic tap changer. The tap changer will maintain the transformer's secondary voltage level at 4.16 kV. The main power transformer will feed power to a 4.16 kV switchgear line-up, located in a prefabricated e-house within the main substation area.

Anticipated electrical loads for the refinery are 9.8 MW connected load, 7.7 MW peak load, and 5.8 MW average load.

#### 3.3.2 Main Substation

The main substation will include the following equipment:

- incoming dead-end structure
- structures and bus system
- main incoming circuit breaker
- high-voltage isolation switch
- Single 10 MVA 25 kV to 4.16 kV power transformer
- Harmonic and power factor correction equipment
- 4.16 kV, 1.5 MW standby power diesel generator
- prefabricated e-house with BC hydro revenue metering, protective relaying system, and 4.16 kV distribution switchgear.

#### 3.3.3 Process Plant and Ancillary Services Power Supply

From the main 4.16 kV distribution switchgear, site power will be distributed as required utilizing the utility (pipe) racks. The voltage will be stepped down to 600 V via two 1.5 MVA transformers to supply 600 V power to the refinery.

#### 3.3.4 Electrical Houses

One e-house will be inside the main substation area, and a second will be centrally located within the plant area and will supply 600 V power to the process and ancillary loads.

The e-houses will be modular units assembled off-site and installed outdoors on raised steel structures. The e-houses will be self-supporting, designed and packaged for road shipment to site. All electrical distribution equipment, controls, and instrumentation equipment will be installed, wired, and completely tested before shipment.

The e-houses will be built to meet a one-hour fire rating. All openings will be sealed and made water- and dust-tight by using approved fire-retardant materials.

Each e-house will have two means on ingress/egress by man-doors at each end, complete with panic buttons for emergency exit. Each e-house will also have an equipment door, sized to permit the largest piece of equipment to be installed/removed without removing the door from its hinges.

The e-houses will be pressurized, air conditioned, and designed in accordance with occupancy regulations.

### **3.3.5 Standby Power**

A 1.5 MW modular standby diesel power plant will be provided, rated for the maximum power required in the event of a utility power failure. The power plant will consist of a single 1.5 MW generator connected to the main 4.16 kV distribution switchgear.

Emergency power loads will be controlled through the process control system, which will stagger starts, automatically start and stop loads to keep process tanks properly agitated, and other critical equipment operational. In addition to the critical process loads, the power plant will supply power to the building heat systems and heat tracing circuits.

Uninterruptible power supplies (UPSs) will provide backup power to critical control systems. The UPSs will be sized to permit operations to shut down and back-up the computer and control systems for start-up on restoration of normal (utility) power.

## **3.4 Buildings and Services**

Three major process buildings will house most of the process equipment:

- The main process building will house the feed preparation, POX, reagents, leach residue loadout, magnesium removal, copper and cobalt recovery. This building will have two overhead cranes servicing the main process and reagent areas. A dedicated maintenance shop will abut the main process building.
- The SX building will house nickel / cobalt solvent extraction complete with raffinate strip and crud filtration. The SX building will have an overhead crane to service the entire area.
- The crystallizer building will house the NiSX raffinate evaporator, nickel sulphate and ammonium sulphate crystallizers, complete with drying and packaging lines and product load-out facilities. This building will have two overhead cranes servicing the different products.

An administration complex will house the offices, analytical laboratory, and change rooms.

## **4. Marketing**

This section outlines the marketing aspects of refining an awaruite concentrate into four saleable products including:

- battery-grade nickel sulphate
- cobalt carbonate
- copper cement
- ammonium sulphate.

This section describes key considerations regarding the feedstock and products from the refinery, as well as market conditions that complement the engineering and design basis in the project’s economic model.

## 4.1 Refinery Feedstock

Metallurgical testwork on awaruite ( $\text{Ni}_3\text{Fe}$ ) nickel ores has shown that a relatively simple flowsheet using magnetic separation followed by conventional froth flotation can produce a highly desirable awaruite concentrate that provides flexibility for downstream consumption and/or further processing. Due to awaruite’s very high nickel tenor (approximately 76%), very high-grade mineral concentrates can be produced, where testwork has consistently demonstrated the production of 60% nickel concentrates. Awaruite concentrates can be used either directly in stainless steel production, without the need for intermediate smelting/upgrading, or subjected to a relatively simple hydrometallurgical process to produce battery-grade nickel sulphate, as presented in this study.

FPX is advancing several awaruite nickel projects in Canada. Other companies are also in the early stages of advancing awaruite nickel projects in North America and Europe. Some ultramafic nickel sulphide projects in Canada have minor awaruite nickel content and could potentially produce an awaruite concentrate. Considering this emerging production landscape, a 40-year operating life is envisioned in this study.

A representative specification for an awaruite concentrate that could be procured to feed the refinery is presented in Table 4-1.

**Table 4-1: Awaruite Concentrate Specification**

Element	Content (%)
Nickel	60
Iron	25
Cobalt	1.1
Copper	0.5
Magnesium	1.7
Sulphur	1.0

## 4.2 Refinery Products

### 4.2.1 Nickel Sulphate

The chemistry of the nickel sulphate crystals produced by the refinery, as demonstrated during FPX’s metallurgical testwork, is presented in Table 4-2, along with an amalgamation of typical target specifications as published by major battery manufacturers. As shown, FPX’s refinery product meets or exceeds typical target specifications for all elements of interest.

Through testwork supporting this study, FPX has again demonstrated the ability to produce battery-grade nickel sulphate, first with the flowsheet using sodium hydroxide ( $\text{NaOH}$ , or caustic) as a neutralizing agent, and now with this study’s flowsheet using ammonia ( $\text{NH}_3$ ) as a neutralizing agent. To confirm, the nickel

sulphate specifications presented in Table 4-2 were achieved using the new ammonia-based neutralization flowsheet.

**Table 4-2: FPX Nickel Sulphate Crystal Quality vs. Target Specification<sup>1</sup>**

Element	Units	Nickel Sulphate Crystals	
		FPX Testwork <sup>1</sup>	Target Specification <sup>2</sup>
Nickel	Wt%	>22	>22
Aluminum	ppm	<1	<5
Arsenic	ppm	<1	<2
Calcium	ppm	<1	<5
Cadmium	ppm	<1	<1
Cobalt	ppm	1	<50
Chromium	ppm	<1	<3
Copper	ppm	<1	<3
Iron	ppm	<1	<3
Potassium	ppm	<1	<10
Magnesium	ppm	3	<5
Sodium	ppm	<10	<20
Lead	ppm	<1	<2
Silicon	ppm	<5	<10
Zinc	ppm	3	<5

<sup>1</sup> Based on the ammonia-based neutralization flowsheet as presented in this study

<sup>2</sup> Based on review of industrially available products and battery producers' requirements

#### 4.2.2 Cobalt Intermediate

Cobalt carbonate produced by FPX testwork supporting this study assays at 51% cobalt, 0.3% magnesium, 0.6% manganese, <0.01% copper, 0.1% nickel, 0.2% zinc and 0.3% sulphur. This is slightly higher than the cobalt hydroxide precipitate from testwork supporting the previous study, which ranged from 38% to 40% cobalt. Both products are deemed to be of relatively high quality compared to existing cobalt products, most notably with respect to the low manganese and magnesium impurity levels relative to cobalt hydroxide from the Southern African copper belt. As such, both products are expected to be readily saleable.

Internal estimates indicate that the choice to produce either a cobalt carbonate or a cobalt hydroxide precipitate does not materially affect refinery economics; therefore, selection of the preferred precipitate type will likely be governed by the reagent market at the finalized refinery location.



### 4.2.3 Copper Cement

The copper cement from iron precipitation is expected to be 70% to 80% copper, with iron as a major diluting element. This product is expected to be readily saleable to any copper smelter.

### 4.2.4 Ammonium Sulphate

As ammonium sulphate crystallization is a mature industrial process, no crystallization testing was conducted. However, as the ammonium sulphate crystallizer is an “end-of-the-line crystallizer” with no solution bleed, the composition of the ammonium sulphate crystals can be reasonably estimated using the composition of the crystallization feed liquor. The estimated ammonium sulphate crystal grade is >21.0% nitrogen, 0.1% phosphorus, 0.05% magnesium, and 0.01% nickel. Based on discussions with ammonium sulphate industry participants, this is deemed to be of sufficient quality to be marketable as a mid-range fertilizer product. Future testwork will test the potential to produce a higher quality product, which could improve the pricing basis.

## 4.3 Pricing

### 4.3.1 Nickel, Cobalt, and Copper

The refinery will procure nickel units in the form of awaruite concentrate. Given the suitability of awaruite concentrate to be directly used in stainless steel production, the pricing basis for this concentrate will be similar to currently available ferronickel products. As such, the nickel unit procurement pricing basis for the refinery will be a function of the LME nickel price, with a discount/premium (payability) applied based on prevailing market conditions.

Table 4-3 presents a summary of projected long-term nickel, cobalt, and copper prices from reputable analysts as of December 2024.

**Table 4-3: Projected Long-Term LME Metal Prices by Analysts, US\$/lb**

Analyst	Nickel	Cobalt	Copper
Analyst 1	\$8.62	\$16.00	\$4.16
Analyst 2	\$7.48	\$13.15	\$4.54
Analyst 3	\$9.07	\$15.00	\$4.08
Analyst 4	\$8.35	\$14.00	\$4.32
Analyst 5	\$9.00	\$20.64	\$3.45
Analyst 6	\$9.50	\$15.00	\$4.25
<b>Average</b>	<b>\$8.67</b>	<b>\$15.63</b>	<b>\$4.13</b>

For the purposes of this study, the following long term metal prices (\$US) were used:

- Nickel \$8.50/lb (\$18,738/t)
- Cobalt \$15.00/lb (\$33,068/t)
- Copper \$4.00/lb (\$8,818/t).

Table 4-4 presents a summary of payability by year for ferronickel products, as reported by Anglo American from the sale of its ferronickel products.

**Table 4-4: Ferronickel Payability Reported by Anglo American**

Year	Payability (% of LME Nickel Price)
2017	101
2018	99
2019	99
2020	90
2021	92
2022	88
2023	79
<b>Average</b>	<b>92.6%</b>

For the purposes of this study, FPX assumed that awaruite concentrate will be procured on a 92% payability basis in relation to the LME nickel price. This was applied to the assumed LME nickel price of \$8.50 per pound (\$18,738 per tonne), resulting in a procurement basis for awaruite concentrate to feed the refinery of \$7.82 per pound of contained nickel (\$17,239 per tonne).

### 4.3.2 Nickel Sulphate

Nickel-based batteries make up a significant proportion of the market share for electric vehicle (EV) batteries due their role in enhancing battery performance. Nickel-based batteries have improved energy density, allowing for longer driving ranges, faster charging, and better overall performance. Nickel sulphate is the preferred feedstock for nickel units in current precursor cathode active material (PCAM) production, with PCAM then feeding cathode active material (CAM) production, followed by EV battery cell production itself.

Benchmark Mineral Intelligence (Benchmark) maintains a comprehensive database of nickel production statistics, and forecasts long-term pricing premiums relative to the LME nickel price for battery-grade nickel sulphate. While the nickel sulphate market is currently small with inherent pricing volatility, even the most conservative EV adoption rates predict a significant increase in nickel sulphate requirements. As the nickel sulphate market grows in the coming years and preferred feedstocks are established, it is expected that a more consistent premium basis will be established based on typical upgrading costs. Benchmark is well-positioned to provide a nickel sulphate premium forecast.

Benchmark's forecast nickel sulphate premium basis for the year 2030, around the time when the refinery would enter production, is \$0.72/lb (\$1,575/t), which has been applied in FPX's economic analysis. This premium is based on Benchmark's "Base Case" forecast of EV adoption and battery chemistry trends, as seen in Table 4-5.

**Table 4-5: Benchmarked Nickel Sulphate Premium – Demand Scenario**

Nickel Sulphate Demand Scenario	Nickel Sulphate Premium (US\$/t above LME Ni Price)
Low Case	\$220
Base Case	\$1,575
High Case	\$2,305

While the high quality, very low carbon intensity, and favourable production jurisdiction of the FPX refinery's nickel sulphate product may be able to attract a preferential pricing basis, such an improved pricing basis was not assumed in this study.

### 4.3.3 Cobalt Carbonate

Cobalt is used in a range of applications, but the largest single market is lithium-ion batteries, which is forecast to have a strong growth profile over the coming years. Cobalt use is generally stable in a variety of other applications, including alloying, colour pigment, catalysts, hard facings, and magnets.

Cobalt in awaruite concentrate has no value to the stainless-steel industry, as producers can neither separate out cobalt nor significantly benefit by having it in their nickel feedstock. However, the refinery will be able to separate and produce a saleable cobalt carbonate product, with the contained cobalt value thereby improving project economics.

Cobalt will make up approximately 11% of refinery profit, so pricing assumptions for cobalt have a limited impact on project economics. For this study, a payability of 85% of LME cobalt price was assumed, which is within the typical payability range for cobalt precipitates. This is deemed conservative given the product's high quality, very low carbon intensity, and favourable production jurisdiction.

### 4.3.4 Copper Cement

The demand for copper is experiencing significant growth due to its essential role in the transition to green energy and electric mobility. As the world moves towards more renewable energy sources, copper is increasingly vital in the manufacturing of EVs, wind turbines, solar panels, and energy storage facilities. The global shift toward electrification and sustainability, coupled with the increasing need for infrastructure and technology upgrades, is expected to drive copper demand well into the next decade.

As with cobalt, copper in awaruite concentrate has no value to the stainless-steel industry, as producers can neither separate out copper nor significantly benefit by having it in their nickel feedstock. However, the refinery will be able to separate and produce a saleable copper cement product, with the contained copper value therein thereby improving project economics.

Copper will make up approximately 1% of refinery profit, so pricing assumptions have minimal impact on project economics. For this study, a payability of 95% of LME copper price was assumed, which is the middle of typical payability range for copper concentrates. This is deemed conservative given the product's high quality, very low carbon intensity, and favourable production jurisdiction.

### 4.3.5 Ammonium Sulphate

Ammonium sulphate is a widely used nitrogenous fertilizer. It is most typically produced as either a byproduct of the petroleum refining process or as a primary product through the reaction of ammonia with sulphuric acid. In addition to its use as a fertilizer, ammonium sulphate has applications in industries like water treatment, pharmaceuticals, and food processing.

The Asia-Pacific region dominates the ammonium sulphate market. China's ammonium sulphate production is large, and India's growing demand for fertilizers ensures robust market growth. The United States and Canada are anticipated to see stable growth in ammonium sulphate consumption due to steady agricultural activity.

Although the ammonium sulphate market presents growth opportunities, inherent challenges include fluctuations in raw material prices, environmental/regulatory constraints, and competition from alternative fertilizers. These may constrain new investment in the sector, which could potentially increase prices.

The current ammonium sulphate market is estimated to be 30,000,000 tpy worldwide and 3,000,000 tpy in the United States. The refinery would produce an average of 87,400 tpy of ammonium sulphate, which would represent a little less than 3% of the current US market.

Ammonium sulphate will make up approximately 20% of refinery profit, so pricing assumptions have moderate impact to project economics. For this study, an ammonium sulphate price of US\$330/t was assumed, which is more conservative than the International Raw Materials report for October 2024 (listed as \$320/st, or \$353/t, for sales in western Canada). According to the Imarc Group, pricing for ammonium sulphate averaged \$360/t over 2023 and 2024.

### 4.3.6 Other Products

Based on flowsheet optimizations during this study, and supported by FPX's initial testwork, there may be an opportunity to potentially valorize the struvite product produced from the magnesium precipitation circuit. Further testwork is required to confirm if fertilizer-grade struvite could be produced, which would then become a fifth revenue-generating product; however, this study assumes that the struvite will be disposed, not sold.

## 5. Environmental Studies, Permitting & Community

### 5.1 Environmental Studies

The refinery site, assumed to be in central BC, will be subject to environmental characterization and baseline studies. Environmental studies will be based on the required permitting processes such as the provincial Project Environmental Assessment (EA), and applicable stipulations will be listed in the Application Information Requirements (AIR) document. Siting of the refinery within or adjacent to municipal boundaries will likely require additional studies to meet bylaw requirements (e.g., evaluation of soil and groundwater from historical land uses, comprehensive air emissions quality modelling, noise modelling, human health, transportation, and waste management). Environmental studies will inform mitigation plans and refinery-specific management plans.

## 5.2 Permitting

Permitting of a metals refinery in BC is subject to existing, well-defined processes. The BC Environmental Assessment Act Reviewable Projects Regulation (BC. Reg. 32/2023) specifies thresholds for the evaluation of a “metals refining” facility under Part 2 – Industrial Projects, Table 2, Primary Metals Industry. If the refinery is integrated into an application for a mine undergoing permitting, in the vicinity of the mine site and dedicated to that mine, then it does not require a separate EA as per Column 2, criterion 2 of the regulation, and would be included in the mining project EA.

The federal Impact Assessment Act (IAA) Physical Activities Regulation (SOR/2019-285) Schedule 2 does not identify a metals refinery as a designated project.

In addition to an EA Certificate, the refinery will require provincial and potentially federal and local government permits for construction, operation and ultimately closure. Permits for the discharge of air and effluent to the environment will require detailed design and modelling. If municipal bylaws do not include a refinery as a listed industrial use, then re-zoning, development permits, and permits to connect to municipal utilities may be required.

## 5.3 Community

FPX recognizes that engagement and support of the project from Indigenous peoples is critical to its success. Engagement and consultations will be driven by best practices, company policies, and federal and provincial requirements. FPX will seek to engage based on transparency, accountability, mutual understanding, respect for Indigenous rights and title, and a long-term commitment to shared values. As an example of an opportunity for engagement driven by this long-term commitment to shared values, the BC First Nations Energy and Mining Council March 2024 First Nations Critical Minerals Strategy calls on proponents, Canada, BC, and First Nations to maximize the economic value in the entire spectrum of critical minerals industry, including spin-offs that could include the refining of critical minerals.

Provisions for consultations with First Nations, local communities and the public are a component of the provincial and federal legislation for both the EA and permitting. FPX will develop and implement an engagement plan for the project, as required by the EA and other regulatory processes.

## 6. Capital Cost

### 6.1 Estimate Summary

The total capital cost estimate is summarized in Table 6-1, in United States dollars (US\$) with a base date of Q4-2024 and no provision for forward escalation. This estimate falls under the AACE Class 5 Estimate classification and its accuracy is expected to be within  $\pm 35\%$  of final project cost.

**Table 6-1: Total Capital Cost Estimate Summary**

Cost Type	Category	Total, US\$ millions
Initial Capital Costs	Refinery Process	\$152
	Reagents	\$45
	Utilities, Services and Infrastructure	\$40
	<b>Total Direct Cost</b>	<b>\$237</b>
	Indirect Costs	\$81
	Contingency	\$89
	Owner's Costs	\$18
	<b>Total Initial Capital Cost</b>	<b>\$424</b>
Total Sustaining Capital Cost (estimated by FPX)		\$40
Total Closure Capital Cost (estimated by FPX)		\$42
<b>Total Capital Cost</b>		<b>\$506</b>

## 6.2 Direct Costs

Direct costs are based on sources such as equipment lists, MTOs, factors and allowances for similar projects in the same region. Total direct costs are summarized in Table 6-2.

**Table 6-2: Total Direct Cost Estimate Summary**

Cost Type	Category	Total, US\$ millions
Direct Capital Costs	Earthworks	\$5
	Concrete	\$13
	Structural Steel	\$8
	Architectural	\$37
	Platework	\$32
	Mechanical Equipment	\$75
	Pipework	\$34
	Electrical	\$28
	Instrumentation	\$7
<b>Total Direct Capital Cost</b>		<b>\$237</b>

### 6.2.1 Civil Works

Wood developed civil MTOs at a conceptual level to support the capital cost estimate. Quantities are included for site clearing and grubbing, topsoil stripping, bulk excavation, and backfill. Quantities for unpaved roads are included, along with allowances for tie-ins for the rail spur and turnouts.

Unit rates and manhours were applied based on Q4-2024 contractor pricing from similar projects in the region.

### **6.2.2 Concrete**

Concrete pricing was factored as 17% of the estimated installed mechanical equipment cost. Wood benchmarked this blended factor against our in-house data for hydromet plants.

### **6.2.3 Structural Steel**

Structural steel pricing was factored as 10% of the estimated installed mechanical equipment cost. Wood benchmarked this blended factor against our in-house data for hydromet plants. Additional structural steel quantities and costs are allocated to certain vertical packages, such as pre-engineered buildings, the acid plant, and the oxygen plant; these are therefore included within other packages.

### **6.2.4 Architectural**

Wood calculated architectural quantities, such as areas and volumes, from the conceptual and GAs. Pre-engineered buildings were specified for the POX, crystallization and SX areas. Other structures include a fabric maintenance/warehouse building and a modular office building/complex. Unit rates and manhours are based on Q4-2024 contractor pricing from similar projects in the region.

Wood's discipline teams provided the costs of detailed earthworks, concrete foundations, internal steel for equipment supports, building services, lighting, and grounding for the buildings.

### **6.2.5 Mechanical**

Wood listed and sized the project's mechanical equipment with the guiding principle of selecting proven designs used on similar projects, to ensure equipment selection was robust. E-mail requests were issued for budgetary quotations for major equipment, including all technical requirements and performance criteria.

Over 90% of major equipment prices are based on Q4-2024 budgetary quotations, with the remainder priced using Wood's in-house database and escalated to 2024 dollars. Ancillary mechanical equipment was factored at 2.5% of the cost of the major mechanical equipment. Installation costs are based on unit manhours from Wood's database. Base hours were adjusted for productivity and site-specific conditions.

### **6.2.6 Platework (Tanks)**

Platework items are identified on the mechanical equipment list. Major tanks were priced using a Q1-2024 budgetary quotation (past project pricing). while minor items were quantified by weight and square meters of liner.

Pricing is based on recent in-house data for each material of construction. Back-calculated total platework costs are 43% of the estimated installed mechanical equipment costs. Additional platework quantities and costs are allocated within certain vertical packages such as the acid plant and the oxygen plant and are therefore included within other packages.

### 6.2.7 Piping

Piping cost was factored as 45% of the estimated installed mechanical equipment cost. Again, Wood benchmarked this blended factor against our in-house data for hydromet plants. The factor includes an allowance for site services and utilities. Additional piping quantities and costs are allocated within certain vertical packages such as the acid plant and the oxygen plant, and are therefore included within other packages.

### 6.2.8 Electrical

The electrical estimate was based on the preliminary overall single-line diagram, process connected loads detailed in the mechanical equipment list, and the site layout. Major electrical equipment was priced based on historical data. Additional electrical quantities and costs are allocated within certain vertical packages such as the acid plant and the oxygen plant, and are therefore included within other packages.

### 6.2.9 Instrumentation

Instrumentation pricing was factored as 9% of the estimated installed mechanical equipment cost. Wood benchmarked this blended factor against our in-house data for hydromet plants. Additional instrumentation quantities and costs are allocated within certain vertical packages such as the acid plant and the oxygen plant, therefore are included within other direct cost disciplines or packages.

## 6.3 Indirect Costs

Indirect costs were factored from total direct costs where applicable, based on Wood's benchmarks for projects of similar size, complexity and location. The total indirect estimate costs are shown in Table 6-3.

**Table 6-3: Total Indirect Cost Estimate Summary**

Cost Type	Category	Total, US\$ millions
Indirect Capital Costs	Freight & Logistics	\$12
	EPCM (Project Delivery)	\$43
	Field Indirects	\$19
	First Fills	\$2
	Spares	\$3
	Commissioning Operations Readiness	\$1
	Vendor Representatives	\$1
<b>Total Indirect Capital Cost</b>		<b>\$81</b>

## 6.4 Contingency

Contingency was factored as 28% of total direct and indirect costs, or \$88.9M. Wood determined this factor based on the level of engineering completed to date, and on the quality of equipment and material pricing.



## 6.5 Owner's Costs

Owner's costs were factored as 5% of total direct and indirect costs, plus an allowance for surface mobile equipment, for a total of \$15.9M. Contingency was not applied to owner's costs.

## 6.6 Sustaining Capital Costs

FPX provided a sustaining capital cost allowance of \$1M per operating year for inclusion in the economic model, for a total of \$40M.

## 6.7 Closure Capital Costs

FPX provided a closure capital cost allowance of 10% of initial capital costs for inclusion in the economic model, for a total of \$42M.

## 6.8 Exclusions

The following items are excluded from the capital cost estimate:

- financing and interest during construction
- currency fluctuations
- sunk costs
- working capital
- changes to design criteria
- time and material contracts
- scope changes or accelerated schedule
- modifications after hand-over
- taxes and duties
- piling or additional soil reinforcement
- any provision for force majeure events
- schedule delays caused by external influences.

## 7. Operating Costs

### 7.1 Summary

Wood calculated the average annual operating costs for the refinery, in US\$ with an accuracy of  $\pm 35\%$  and a base date of Q4 2024. Total operating costs can be divided into fixed and variable costs as shown in Table 7-1. Reagents constituted almost 50% of the total cost, followed by consumables and labour at 33%.

**Table 7-1: Summary of Total Operating Costs**

<b>Cost Centre</b>	<b>US\$/year</b>	<b>Production Basis \$/t Ni</b>
Labour	\$8,252,443	\$258
Miscellaneous	\$3,647,126	\$114
<b>Subtotal, Fixed Costs</b>	<b>\$11,899,569</b>	<b>\$372</b>
Reagents	\$24,235,802	\$757
Maintenance	\$4,358,705	\$136
Power	\$2,192,125	\$69
Consumables	\$8,451,815	\$264
<b>Subtotal, Variable Costs</b>	<b>\$39,238,447</b>	<b>\$1,226</b>
<b>TOTAL</b>	<b>\$51,138,016</b>	<b>\$1,598</b>

## 7.2 Estimate Basis

The estimated operating cost to produce 32,000 tpy of Ni contained in nickel sulphate hexahydrate crystals is based on study deliverables including the process design criteria and mass balance. This totals \$51.1 million/year and includes the following:

- reagents
- labour
- miscellaneous
- maintenance
- power
- consumables.

### 7.2.1 Reagents

Reagents account for nearly 50% of the overall operating costs, as determined by the consumption and unit costs of various reagents. The total amounts to \$24.2 million/year, with anhydrous ammonia representing almost 60% of this cost. Budgetary quotations or reputable market forecasts were obtained for all reagents.

### 7.2.2 Process Plant Labour

Labour costs are based on published 2024 rates for British Columbia. Around-the-clock operation with four shift crews was assumed, each working a four-days-on/four-days-off schedule, as well as day staff.

### 7.2.3 Process Miscellaneous

Miscellaneous process costs include general allowances, such as training expenses, office supplies, HSSE (Health, Safety, Security, and Environment), laboratory expenses, and mobile equipment maintenance. Laboratory costs were determined by the number of samples needed for metallurgical accounting and

process control. Mobile equipment costs cover operations, maintenance, and safety vehicles. Estimates for maintenance and fuel are based on the expected operating hours per year and published hourly maintenance rates.

#### **7.2.4 Maintenance**

Wood applied a factor of 4% to the direct capital cost for instrumentation, mechanical and electrical facilities to cover maintenance requirements.

#### **7.2.5 Electrical Power**

Power requirements are based on the mechanical load list for the refinery, which is estimated to have an installed load of 9 MWh and average consumption of 5.9 MWh. Average power consumption was determined using the installed power, with power utilization and power loss factors applied to each electrical drive on the mechanical equipment list. Costs are based on a unit power cost of US\$0.0446 per kWh.

#### **7.2.6 Consumables**

The estimate of process plant consumables is based on mass balance data, vendor recommendations, and in-house experience with comparable plants or unit operations. Oxygen supply accounts for ~42% of the total consumables cost and is assumed to be sourced through an agreement for an over-the-fence supply. Note that a budget quote was obtained for the oxygen supply.

### **7.3 Exclusions**

The following items are excluded from the operating cost estimate:

- import duties, withholding taxes and other in-country taxes
- impact of foreign exchange rate fluctuations
- contingency allowance
- rehabilitation or closure costs
- licence fees or royalties
- government monitoring/compliance costs
- all owner's costs and risk allowances, other than specified.

## **8. Economic Analysis**

### **8.1 Forward-Looking Cautionary Statements**

The results of the economic analysis discussed in this section represent forward-looking information as defined under Canadian securities law. The results depend on inputs that are subject to a number of known and unknown risks, uncertainties and other factors that may cause actual results to differ materially from those presented herein. Information that is forward-looking includes commodity pricing, ferronickel payability, nickel sulphate upgrade premium, exchange rates, metallurgical recovery rates, cost estimates, environmental risks, permitting, social risks, availability of utilities, labour availability, and taxation basis.

## 8.2 Methodologies Used

An economic model was developed to estimate annual pre-tax and post-tax cash flows and sensitivities of the project based on an 8% discount rate. A detailed tax build-up was conducted by a third-party tax expert, however tax estimates involve many complex variables that can only be accurately calculated during operations and, as such, the after-tax results are only approximations. A sensitivity analysis was performed to assess the impact of variations in discount rate, nickel sulphate premium, operating cost and initial capital cost. The capital and operating cost estimates were developed specifically for this Study and are summarized in Sections 6 and 7 of this report, respectively. The estimates are prepared in fourth quarter 2024 United States Dollars (Q4-2024 US\$).

## 8.3 Financial Model Parameters

The economic analysis was performed using the following assumptions:

- construction period of 2 years
- operating life of 40 years
- long-term nickel price of US\$18,738/t (US\$8.50/lb)
- ferronickel payability rate of 92% of LME nickel price
- nickel sulphate upgrade premium of US\$1,575/t of contained nickel (\$0.72/lb)
- United States-to-Canadian dollar exchange rate of 1.00 CA\$ : 0.74 US\$
- cost estimates in constant Q4-2024 US\$ with no inflation or escalation factors considered
- no salvage value
- results based on 100% ownership
- no royalties applicable
- capital costs funded with 100% equity (i.e., no financing costs assumed)
- all cash flows discounted to the beginning of construction
- all metal products sold in the same year they are produced.

## 8.4 Taxes

A detailed tax build-up was conducted by a third-party tax expert; however tax estimates involve many complex variables that can only be accurately calculated during operations and, as such, the after-tax results are only approximations. The study models taxes in accordance with provincial and federal legislation. The Study reflects the impact of the federal government's refundable critical minerals investment tax credit, announced in the 2023 federal budget, which is proposed to be equal to 30% of the capital cost of eligible property for the extraction and processing of certain critical minerals, including nickel. The Study estimates total LOM taxes paid of C\$1,000 million including C\$520 million to the Province of British Columbia and C\$480 million to the Government of Canada.

## 8.5 Economic Analysis

The economic analysis was performed assuming an 8% discount rate in Q4-2024 United States dollars. On a pre-tax basis, the net present value (NPV) is US\$542 M, the internal rate of return (IRR) is 19%, and the payback is 5.0 years. On an after-tax basis, the NPV is US\$445 M, the IRR is 20%, and the payback is 4.0

years. The after-tax payback period is shorter than the pre-tax payback period due to the inclusion of Canadian investment tax credit refunds in Year 1 to Year 2 of the Project.

A summary of project economics is presented in Table 8-1.

**Table 8-1: Summary of Project Cashflow Assumptions and Results**

<b>General</b>	<b>Units</b>	<b>Project Total / Average</b>
Nickel Price	US\$/t	18,738
Nickel Payability for Awaruite Concentrate (% of LME)	%	92
Nickel Sulphate Premium (in addition to nickel price)	US\$/t	1,575
Cobalt Price	US\$/t	33,068
Cobalt Payability (% of LME)	%	85
Copper Price	US\$/t	8,818
Copper Payability (% of LME)	%	95
Ammonium Sulphate Price	US\$/t	330
Ammonium Sulphate Payability (% of market)	%	100
Exchange Rate	C\$/US\$	0.74
Discount Rate	%	8
Project Life	years	40
<b>Production (excluding first year)</b>		
Average Annual Nickel Production	tpy	32,000
Average Annual Cobalt Production	tpy	570
Average Annual Copper Produced	tpy	240
Average Annual Ammonium Sulphate Produced	tpy	87,400
<b>Operating Costs</b>		
Total (excluding byproduct credits)	US\$/t produced	1,598
<b>Capital Costs</b>		
Initial Capital	US\$M	424
Sustaining Capital	US\$M	40
Closure Capital	US\$M	42
<b>Financials – Pre-Tax</b>		
Pre-Tax NPV <sub>8%</sub>	US\$/M	542
Pre-Tax IRR	%	19
Pre-Tax Payback	years	5.0
<b>Financials – After-Tax</b>		

General	Units	Project Total / Average
After-Tax NPV <sub>8%</sub>	US\$/M	445
After-Tax IRR	%	20
After-Tax Payback	years	4.0

## 8.6 Sensitivity Analysis

An after-tax sensitivity analysis was performed to assess impact of variations in nickel sulphate premium, initial capital cost, and operating cost. Sensitivity results are summarized in Table 8-2, Figure 8-1 and Figure 8-2. Analysis reveals that the project's NPV and IRR are most sensitive to changes in nickel sulphate premium, and due to the 40-year project life, the project's NPV is more sensitive to operating cost than initial capital cost.

**Table 8-2: After-Tax Sensitivity Analysis Results**

After-Tax NPV (US\$M) Sensitivity to Discount Rate		After-Tax IRR Sensitivity to Discount Rate	
5%	\$793M	5%	19.9%
8%	\$445M	8%	19.9%
10%	\$304M	10%	19.9%
After-Tax NPV (US\$M) Sensitivity to Nickel Sulphate Premium		After-Tax IRR Sensitivity to Nickel Sulphate Premium	
(20%)	\$368M	(20%)	18.1%
(10%)	\$407M	(10%)	19.0%
-	\$445M	-	19.9%
10%	\$484M	10%	20.8%
20%	\$522M	20%	21.6%
After-Tax NPV (US\$M) Sensitivity to Initial Capital		After-Tax IRR Sensitivity to Initial Capital	
(20%)	\$493M	(20%)	23.7%
(10%)	\$469M	(10%)	21.6%
-	\$445M	-	19.9%
10%	\$422M	10%	18.4%
20%	\$397M	20%	17.2%
After-Tax NPV (US\$M) Sensitivity to Operating Cost		After-Tax IRR Sensitivity to Operating Cost	
(20%)	\$524M	(20%)	21.7%
(10%)	\$484M	(10%)	20.8%
-	\$445M	-	19.9%
10%	\$406M	10%	19.0%
20%	\$369M	20%	18.0%

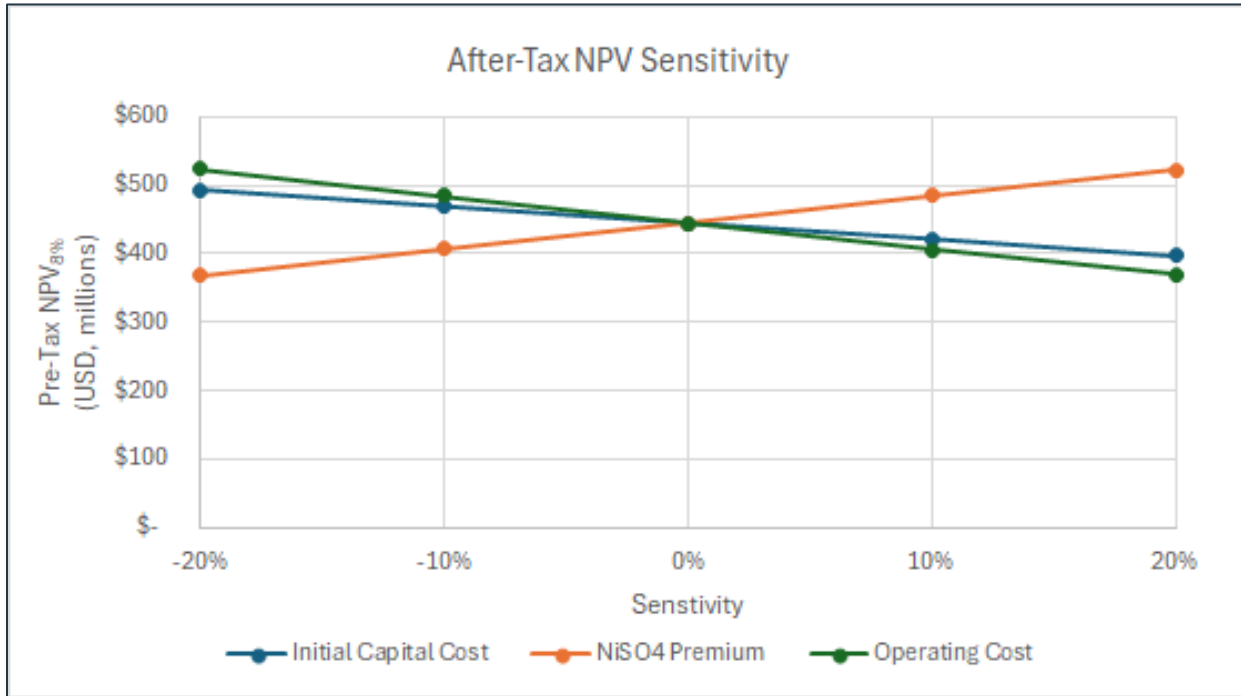


Figure 8-1: After-Tax NPV Sensitivity Chart (US\$M)

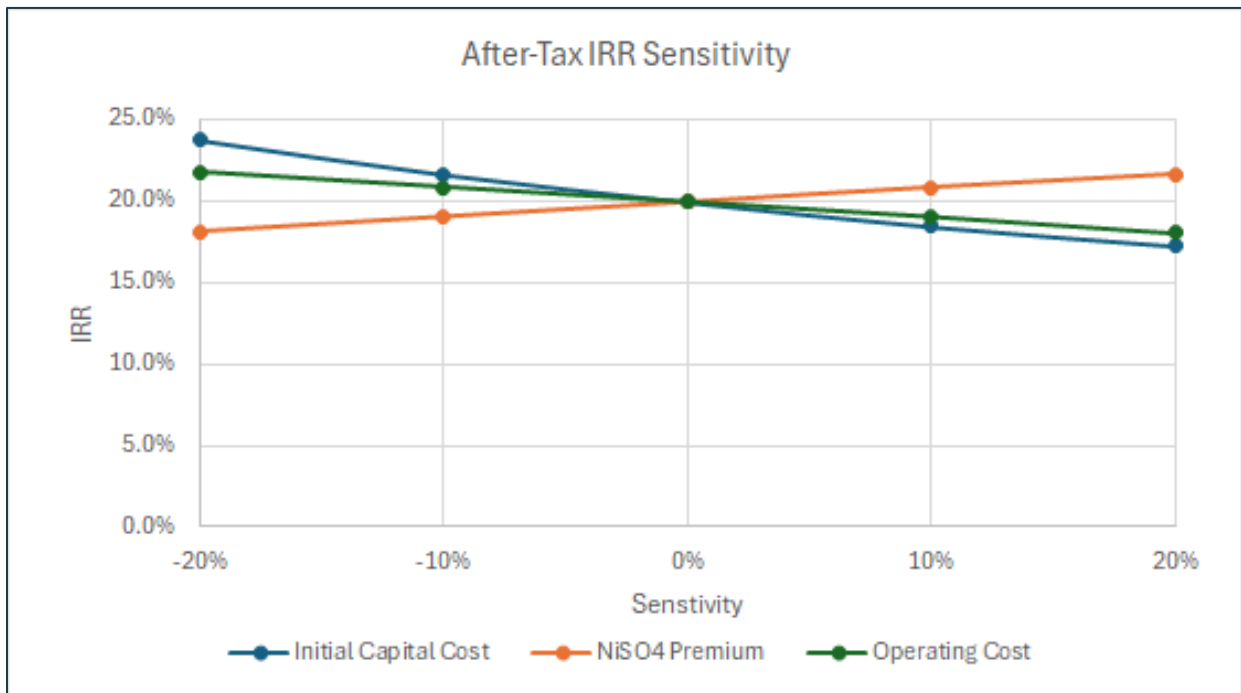


Figure 8-2: After-Tax IRR Sensitivity Chart (US\$M)



## 9. Project Execution

### 9.1 Overview

Project implementation includes engineering, procurement, construction and commissioning of facilities to generate saleable products including nickel sulphate (primary), cobalt carbonate, copper cement, and ammonium sulphate.

The project is assumed to be located in central BC for this study, which offers the following execution advantages:

- close to an urban centre with a capable trades workforce
- well established services industry
- well established construction contractors
- well established vendor support network
- no construction camp required
- availability of required reagents
- utilities available over the fence.

The Study assumes construction of the process plant, associated facilities and infrastructure on an engineering, procurement and construction management (EPCM) basis, using a strategy of multiple horizontal, discipline-oriented construction contract packages. In addition, several vertical packages are considered, such as pre-engineering buildings, the acid plant, the oxygen plant and other select process and/or ancillary facilities.

The preliminary execution strategy presented in this section will be refined during the next phase of the study, and will be used as the basis for execution plans that will be used to manage and control the project.

### 9.2 Scope

The project scope covers the refining of awaruite concentrate to nickel sulphate hexahydrate crystal as a primary product. In addition to the process plant and associated infrastructure, the project includes non-process infrastructure within the process plant, and tie-ins to local access roads, water and sewer, and the high-voltage (HV) power grid.

### 9.3 Execution Strategy

The project involves engineering, procurement, construction and commissioning of equipment, plant and facilities for refining awaruite concentrate to produce battery grade nickel sulphate. It also includes construction of the infrastructure and services.

The EPCM contractor will design and manage the construction of the processing plant and associated infrastructure. The EPCM contractor's work includes:

- engineering and procurement of the refinery facilities, infrastructure, and services
- contracting and management of construction contractors.

FPX will manage the EPCM contractor, and manage or execute all other activities needed to complete the project, including:

- finance, insurance, governmental approvals, environmental approvals and licences
- land permits, security, medical, taxes and duties
- engagement of specialist consultants and contractors for community engagement, environmental approvals and other specialist scopes
- contracts for water, power and reagent supply, transport of product including customs and port handling to free on board, etc
- engagement and training of key operations personnel
- operational readiness.

## 9.4 Schedule

Based on the facility design and Wood's experience in construction of facilities of similar scale and complexity in the project region, it is envisioned that a two-year construction duration will be required. This duration covers the period from start of construction early works through to completion of pre-commissioning activities, at which time the facility will be handed over to FPX's operations group. Note that a ramp-up of production during the first year of operation is considered, with full production only considered during year two of operations.

The critical path to construction is anticipated to run through the environmental assessment (EA) and permitting work streams. A typical project development sequence of preliminary feasibility study (PFS), feasibility study (FS), and basic engineering will be integrated within and fully support the EA and permitting critical path. The PFS, FS, and basic engineering will be supported by metallurgical testwork, engineering field investigations, baseline studies, etc.

A stage gate approach will be taken to project development to ensure the project is suitably well defined to progress to the next stage of project development, including a final investment decision (FID). The FID will consider successful receipt of EA approval, receipt of key permits, and completion of basic engineering which would confirm the economic business case.

## 10. Conclusions and Recommendations

This awaruite refinery scoping study demonstrates a compelling business case for the development of a stand-alone refinery to refine awaruite concentrate into battery-grade nickel sulphate for the electric vehicle industry, along with producing valuable cobalt, copper, and ammonium sulphate byproducts.

The project is large-scale and long-life, producing 32,000 tpy of nickel contained in battery-grade nickel sulphate over a 40-year duration. This production rate would represent an approximate tenfold increase to current North American nickel sulphate production capacity.

The study presents capital and operating cost estimates either aligned with or exceeding typical engineering and cost definition for scoping studies. The initial capital cost estimate is \$424M, and total capital cost, inclusive of sustaining and closure costs, totaling \$506M. Operating costs are estimated at \$1,598/t of nickel produced, exclusive of byproduct credits. Inclusive of byproduct credits, net operating costs are reduced to \$133/t of nickel produced.

The refinery would produce nickel sulphate with a carbon intensity of 0.2 tCO<sub>2</sub>/t Ni for refining operations, which is magnitudes lower than current nickel sulphate production routes.

Wood recommends the next steps as outlined below:

- Work to be completed before the PFS:
  - Identify proposed refinery site.
  - Complete the pilot-scale purification and crystallization testwork using the PLS generated during the already completed pilot-scale leach testwork.
  - Complete corrosion testing to confirm the selection of materials of construction.
- Work to be completed as part of PFS:
  - Undertake a project workshop to establish the Class of Facility for the project, with a focus on identifying the best value design basis for the project.
  - Conduct trade-off studies to confirm reagent availability and cobalt product type based on selected refinery location.
  - Evaluate the economics of over-the-fence oxygen supply.
  - Confirm evaluation for availability of major utilities.