APPENDIX A – ORE PROCESSING DETAIL

The following is a detailed description of the proposed ore processing steps.

CRUSHING

Coarse Crushing

The objective of the Coarse Crushing area is to reduce ore in two stages to a size suitable for further size reduction in Fine Crushing.

Operating approximately 17 hours a day, the coarse crushing circuit comprises Primary and Secondary Crushers that deliver crushed ore at a P_{80} of 3.4 in to the Coarse Ore Bins in the Fine Crushing building.

Trains deliver ore from the open pit to the Coarse Crusher Dump Pocket. A remotely controlled jack empties each ore car sequentially into the Primary Crusher (60"gyratory) that is capable of crushing ore at an average rate of 1852 short tons/hr, with peak loadings up to 2150 short ton/hr.

The Primary Crusher discharges into a Primary Crusher Product Pocket from which crushed ore is withdrawn under gravity into four parallel Secondary Crushers (36" gyratory) that operate in a choke fed mode.

A conveyor train (1A/2A), is fed by Apron Feeders under each Secondary Crusher. The 2A conveyor delivers coarse ore to the Coarse Ore Bin via tripper.

The Coarse Ore Bin has a live capacity of approximately 4 hours and supplies feed to Fine Crushing.

Emission control is provided by a separate collection system within the Coarse Crushing building. The emission controls consist of a single dry bag house for the Primary and Secondary Crushers. The material collected is slurried and added to the milling circuit as Ball Mill Feed.

Fine Crushing

The objective of the Fine Crushing area is to further reduce the coarse crushed product in two stages of reduction to a size suitable for feed to Milling. There are three Fine Crushing Trains.

Coarse ore is withdrawn from the Coarse Ore Bin to one of three Fine Crushing Trains. Each train consists of a Tertiary Crushers (7' Standard cone), two Vibrating Screens and Two Quaternary Crushers (7' Shorthead cone). For each train, a set of six feeders discharges to three conveyors (two Cross Conveyors and one Retract Conveyor) that in turn discharge into each Tertiary Crusher. The Retract Conveyor that feeds the Tertiary Crusher is retractable to allow for crusher maintenance. Each Retract Conveyor is fitted with a metal detector to prevent tramp metal from entering the Tertiary Crushers.

The discharge product of each Tertiary Crusher then passes via vibrating feeder to two screens per crusher, six in total. The oversize material from each screen discharges to a matched Quaternary Crusher operating in open circuit on the screen oversize. The screen undersize material passes directly to a conveyor below the fine crushing circuit that collects all screen undersize and Quaternary Crusher products. The final crushed product conveyor 3A discharges to a second conveyor 4A or 4B that transfers the crushed product to the Fine Ore Bins where conveyor 4A or 4B discharges to conveyors 5N or 5S. A tripper on each conveyor 5N and 5S distributes the crushed ore across the Fine Ore Bin. A transfer point at the discharge of conveyor 3A allows crushed ore to be distributed to conveyors 4A or 4B that in turn allows the crushed ore to be distributed to the Fine Ore Bins ahead of the North and South Milling Trains.

The Fine Ore Bin has a live capacity of approximately 13 hours and feeds eleven groups of 12 vibrating feeders, one group for each of eleven Milling Trains. Each group of 12 feeders then discharges to a single Rod Mill Feed Conveyor. Not all of the feeders are required to run to supply the required milling feed rate to each Rod Mill. Each Rod Mill Feed Conveyor is fitted with a weightometer that is used to adjust the vibrating feeders for controlled delivery of crushed fine ore to the Rod Mills.

Emission control is provided by wet venturi scrubbers on each Fine Crushing Train plus the conveyor transfer points Fine Crushing Building. The material collected is added to the milling circuit as Ball Mill Feed.

MILLING

The objective of the Milling area is to liberate the sulfide minerals contained in the ore through the milling process. Once liberated, the base metal sulfide minerals are amenable to separation and upgrade by froth flotation.

The milling circuit comprises eleven parallel trains of Rod/Ball Mills, which each take an equal proportion of the overall plant feed.

Each Rod Mill is fed at an average rate of 135 short tons/hr on an annualized basis with a crushed product with a P_{80} of 0.4 in and produces a milled product with a nominal P_{80} of 0.03 in. Short-term throughputs may rise to a maximum of 156 short tons/hr.

Each Rod Mill operates in open circuit and delivers a ground product to the feed end of a matched down stream Ball Mill. The Ball Mill operates in closed circuit with a bank of Primary Cyclones, with a circulating load of 350%; and the Primary Cyclone Overflow produced from this circuit is suitable for subsequent flotation.

Grinding media in the form of rods and balls is added to each mill to sustain a load of 35% and 40% by volume respectively to maintain the mill power draw. Additionally, water is added at sufficient rate to each mill feed to maintain the mill density of each mill at about 70 - 75% solids by weight. Grinding media is added periodically to the Ball Mill via a port on the mill feed chute. Rods are added to each Rod Mill using a dedicated rail mounted rod charger that can be located in front of each Rod Mill as required to charge rods through the rod mill discharge feed trunnion.

A Ball Mill Discharge Pump Box (not shown on Figure 6-6) receives ball mill discharge product with a P_{80} of 0.00787 in (65mesh), and dilution water to generate slurry of approximately 63% solids by weight. The resultant slurry is pumped by the duty Primary Cyclone Feed Pump (not shown on Figure 6-6) to a bank of three 26-in diameter Primary Cyclones. The overflow from this cyclone bank from each mill is combined in a launder and then divided into two streams of approximately equal volume, which report under gravity to the Rougher Conditioning Tank in each of two parallel Flotation Trains.

The Primary Cyclone Underflow is combined with the respective Rod Mill Discharge and flows under gravity to the feed spout of the associated Ball Mill.

FLOTATION

The objective of the Flotation area is to recover a bulk sulfide product containing the majority of the base and precious metals, rejecting largely siliceous tailings. Flotation of the sulfide minerals is achieved using a simple collector/frother combination. The frother is an organic compound that allows the formation of small bubbles when air is blown into the flotation cell. The collector is an organic chemical containing the xanthate ion, which is a long-chain hydrocarbon. One end of the ion is polar and will adhere selectively to sulfide minerals. The other end is non-polar and will adhere to the bubbles. This results in the sulfide minerals being selectively floated to the top of the tank.

Two Flotation Trains are provided, each matched to sets of Rod/Ball Mill Trains; in one set five trains are operating, in the other set six.

The Rougher Conditioning Tank Discharge flows under gravity to two Rougher Flotation Cells in each train, the residence time of the Rougher Flotation Cell is 7 minutes at an operating density of 30% by weight solids and at a design feed flow rate.

The collector will be a commercial preparation (Flex 31) of sodium isopropyl xanthate (SIPX). Methyl isobutyl carbonate (MIBC) will be used as the frother at addition rates as required to achieve the target flotation recoveries. Dextrin (a simple starch) is also added as a silicate depressant in the cleaner flotation circuit.

The Rougher Flotation Concentrate from each set of Rougher Flotation Cells is pumped to the cleaner circuit and combined in a single Rougher Cleaner Conditioning Tank for each cleaner flotation circuit. Additional frother, collector and depressant are added before the slurry flows by gravity to Cleaner 1 Flotation. The Rougher Flotation Tailings from each train flows under gravity to six Scavenger Flotation Cells.

Scavenger Flotation operates at a 30% by weight pulp density with a residence time of 20 minutes, and the resulting Scavenger Flotation Concentrate from Scavenger Flotation flows under gravity to a dedicated regrind milling circuit. Scavenger Flotation Concentrates are combined in each Regrind Hopper where they are combined with Cleaner 1 Flotation Tailings and Regrind Mill Discharge. The Scavenger Flotation Tailings from each train, at a flow rate of 838t/hr solids and a density of 40% by weight flow under gravity to a dedicated high rate Flotation Tails Thickener for each Flotation Train. With the aid of flocculant, the Flotation Tails Thickeners produce a thickened underflow at 60% by weight solids, that is then pumped to the existing Flotation Tailings Basin. The composition of Flotation Tailings is discussed in response to Question 20.

Two stages of concentrate cleaning are provided. Overall, the cleaner flotation circuit has a residence time of 10 minutes at a design flow rate and density of 15t% by weight solids.

In each train, stage 1 cleaner flotation is conducted in four Cleaner 1 Flotation Cells and the resultant Cleaner 1 Flotation Concentrate is pumped to two Cleaner 2 Flotation Cells. The final Cleaner 2 Flotation Concentrate stream from each train is then pumped to a single Concentrate Thickener at an average flow rate of 37.8 short tons/hr solids on an annualized basis with a solids density of approximately 12% by weight. The Cleaner 1 Flotation Tailings are directed to the Regrind Hopper and Cleaner 2 Flotation Tailings are recycled back to the Rougher Cleaner Conditioning Tank.

The Concentrate Thickener produces a thickened concentrate underflow containing approximately 65% by weight solids, which is pumped by Concentrate Thickener Underflow Pumps (not shown on figure 6-6) to the one of two Final Concentrate Surge Tanks, which provide surge capacity between the concentrate production circuits and the concentrate leaching circuits. The Final Concentrate Surge Tanks provide approximately twelve hours of surge capacity ahead of the Autoclave Leach area.

Each regrind mill classification circuit comprises a Regrind Cyclone and Regrind Mill to ensure Scavenger Flotation Concentrate and Cleaner 1 Flotation Tailings from each train are ground to a suitable size to allow higher final concentrate grades to be achieved in the subsequent cleaning circuits. The combined streams in each Regrind Hopper are pumped to the Regrind Cyclone from which overflow recycles back to the Rougher Conditioning Tank and underflow discharges to the Regrind Mill.

All pumps in the flotation circuit are organized with a duty and standby pump arrangement, and water spray is added to the concentrate launders to prevent sanding and assist the flow of concentrate into the Regrind Hopper.

Table 6-14 shows the assumed compositions of flotation concentrate and tailings based on the available metallurgical test work.

Table 6-14
Target Flotation Products

Stream	Assays (wt%)					Distribution (%)				
	Cu %	Ni %	Au ppm	Pt ppm	Pd ppm	Cu	Ni	Au	Pt	Pd
Feed	0.39	0.11	0.14	0.09	0.3	100	100	100	100	100
Concentrate	13.5	3.11	1.48	2.33	9.99	94.8	70.1	66.5	71.7	83.8
Tailings	0.02	0.037	0.02	0.02	0.06	5.2	29.9	33.5	28.3	16.2

AUTOCLAVE LEACH

The objective of the Autoclave Leach area is to convert the metal sulfides in the flotation concentrate into a soluble form for as a preliminary step to further processing.

The metal values of the copper, nickel and cobalt sulfides, PGMs and pyrrhotite (iron sulfide) contained in the flotation concentrate are thereby liberated. Conversion of the metal sulfide species into soluble metal species is achieved by the use of high temperature, high pressure leaching conditions in an acidic liquor in three Autoclaves, which is injected with oxygen gas, supplied from an 882 short tons or 1,763,678 lbs per day maximum capacity cryogenic oxygen plant, to oxidize the sulfides. The nominal oxygen consumption of the primary autoclaves is expected to be 781 short tons/hr.

The Autoclave Feed Tank for each Autoclave receives flotation concentrate and copper sulfide precipitate recycled from the downstream processing. The combined product is then pumped to the Autoclave. Hydrochloric acid is added to the Autoclave Feed Tank to maintain 10,000 ppm chloride concentration in the leach solution to enable leaching of the PGM's in the Autoclave. Raffinate (the barren solution remaining after removal of the copper) from Copper Solvent Extraction is also recycled to the Autoclave. The Autoclave Feed Tank slurry and raffinate are pumped by separate pumping systems to each of the three Autoclaves operating in parallel. Raffinate is added to each Autoclave at a rate designed to maintain Autoclave operation at 437°F; temperature control is maintained by adjusting the flow of copper raffinate relative to the flotation concentrate.

Each Autoclave is made up of five agitated compartments with the dividing wall removed between compartment 1 and 2. The Autoclaves are constructed from carbon steel and are lined with a corrosion resistant membrane overlain with two layers of acid resistant bricks. Autoclave internals, agitators, baffles and cell dividers are all of titanium construction to resist corrosion by the acidic leach liquor.

Oxygen is fed to each Autoclave at a controlled rate to ensure complete oxidation of all of the sulfide sulfur in the autoclave feed.

The following reactions are expected to pass to completion in the Autoclave.

Chalcopyrite Dissolution
$4CuFeS_{2(s)} + 4H_2O_{(l)} + 17O_{2(g/aq)} \rightarrow CuSO_{4(aq)} + 2Fe_2O_{3(aq)} + 4H_2SO_{4(aq)}$

Cubanite Dissolution
$2CuFe_2S_{3(s)} + 4H_2O_{(l)} + 13O_{2(g/aq)} \rightarrow 2CuSO_{4(aq)} + 2Fe_2O_{3(aq)} + 4H_2SO_{4(aq)}$

Pentlandite Dissolution
$NiFeS_{2(s)} + 4O_{2(g/aq)} \rightarrow NiSO_{4(aq)} + FeSO_{4(aq)}$

Pyrhotite Dissolution
$2sFe_7S_{8(s)} + 2H_2O_{(l)} + 31O_{2(g/aq)} \rightarrow 14FeSO_{4(aq)} + 2H_2SO_{4(aq)}$

Linnaeite Dissolution
$2Co_3S_{4(s)} + 2H_2O_{(l)} + 15O_{2(g/aq)} \rightarrow 6CoSO_{4(aq)} + 2H_2SO_{4(aq)}$

Sphalerite Dissolution
$ZnS_{(s)} + 2O_{2(g/aq)} \rightarrow ZnSO_{4(aq)}$

Gold Dissolution
$4Au_{(s)} + O_{2(g/aq)} + 2H_2SO_{4(aq)} + 16NaCl_{(aq)} \rightarrow 4Na_3AuCl_{4(aq)} + 2Na_2SO_{4(aq)} + 2H_2O_{(l)}$

Palladium Dissolution
$2Pd_{(s)} + O_{2(g/aq)} + 2H_2SO_{4(aq)} + 8NaCl_{(aq)} \rightarrow 2Na_2PdCl_{4(aq)} + 2Na_2SO_{4(aq)} + 2H_2O_{(l)}$

Platinum Dissolution
$2Pt_{(s)} + O_{2(g/aq)} + 2H_2SO_{4(aq)} + 8NaCl_{(aq)} \rightarrow 4Na_2PtCl_{4(aq)} + 2Na_2SO_{4(aq)} + 2H_2O_{(l)}$

Rhodium Dissolution
$4Rh_{(s)} + O_{2(g/aq)} + 2H_2SO_{4(aq)} + 24NaCl_{(aq)} \rightarrow 4Na_3RhCl_{6(aq)} + 6Na_2SO_{4(aq)} + 6H_2O_{(l)}$
Ferrous Oxidation
$2aFe(OH)_{3(aq)} + 3H_2SO_{4(aq)} \rightarrow Fe_2(SO_4)_{3(aq)} + 6H_2O_{(l)}$

Jarosite Precipitation
$Na_2SO_{4(aq)} + 3Fe_2(SO_4)_{3(aq)} + 12H_2O_{(l)} \rightarrow 2NaFe_3(SO_4)_2(OH)_{6(s)} + 6H_2SO_{4(aq)}$

The generation of acid from the oxidation of major sulfide minerals results in partial leaching of the silicate, hydroxide and carbonate minerals present in the flotation concentrate. The leaching process also generates substantial heat, which will subsequently be converted to steam in each Autoclave Flash Vessel and used in winter to heat the Process Plant buildings. Leached slurry exiting the Autoclaves is brought to atmospheric pressure using a dedicated Autoclave Flash Vessel for each Autoclave, which allows the removal of excess heat through the release of steam from the slurry.

A dedicated Autoclave Gas Scrubber is provided for each Autoclave Flash Vessel to deliver an initial scrub of the excess flash vessel gas streams to remove the majority of the entrained solids and process liquor in the vapor phase exiting the Autoclave Flash Vessel. This is achieved using raw water sprays to remove the entrained salts, solids and acid from the gas flow. The majority of the steam is deliberately not condensed in this first scrub stage. Scrubbed steam then passes through a Heat Exchanger to recover heat by condensing the steam. The hot water produced by this indirect heat exchange process is then used as required for building heating; any remaining hot water not required is cooled in a Cooling Tower and recycled to the Heat Exchanger. Any steam not condensed in the indirect heat recovery process then passes to a Final Gas Scrubber before discharge to atmosphere. The material scrubbed from the gas stream reports to the Leach Residue Thickener as a slurry.

PLATINUM GROUP METALS (PGM) PRECIPITATION

PGM Concentration

The objective of the PGM Precipitation area is to recover PGMs and Gold -- from the pregnant leach liquor using sodium hydrosulfide (NaHS) as a precipitant. This circuit produces a mixed PGM sulfide precipitate with a relatively large proportion (to the precipitated PGM's) of co-precipitated base metal sulfides.

The PGM precipitate is then repulped treated by a selective base metal leach. This removes base metals and produces a PGM concentrate of higher purity for shipment to an off-site refinery. Prior to NaHS addition to precipitate the PGM's and base metals, sulfur dioxide is introduced to the process slurry to reduce all of the contained ferric ions to ferrous ions in order to minimize side reactions of the NaHS precipitant with the dissolved iron values.

Discharge from each of three Autoclave Flash Vessels flows by gravity to matched Autoclave Let Down Tanks (not shown on Figure 6-7) where the slurry is further cooled to 185°F via recycling of cooled leach slurry. The Let Down Tank Discharge is then pumped to the Leach Residue Thickener through a Shell and Tube Heat Exchanger (not shown on Figure 6-7), which is supplied with non-contact cooling water from a Cooling Tower to further reduce the slurry temperature to 140°F. The heated water exiting the Heat Exchanger is recycled back to the Cooling Tower.

Air is drawn into the base of the Cooling Tower counter currently to the hot water sprayed into the top of the towers. Cooling of the heated water exiting the Heat Exchangers is effected by evaporation of water to the counter current air stream. Cooled water exiting the Cooling Tower is collected in a common sump where makeup water is added, in addition to an anti bacterial reagent to maintain a safe working environment.

The cooled discharge from all three Autoclave Flash Vessels is then pumped to the Leach Residue Thickener. The contained solids are then settled with the aid of flocculant; the thickener produces a dewatered underflow containing 40% by weight solids. The underflow is then pumped to the Leach Residue Surge Tank (not shown on Figure 6-7), which has a retention time of 8 hours, thus providing surge capacity between the Leach Residue Filter and thickener. The Leach Residue Filter separates the barren autoclave residue solids from the process liquor containing the metal values of interest. Residual entrained metal values are recovered by washing the leach residue with HM Process Water to recover the remaining dissolved metal values.

A Leach Residue Filter Cake of 75% solids is produced by the filter. The cake is discharged to the Residue Re-pulp Tank (not shown on Figure 6-7) where the cake is repulped and then pumped for disposal in the lined Reactive Residue facility. Wash liquor and primary filtrate from the Leach Residue Filter are recycled back to the Leach Residue Thickener.

The Leach Residue Thickener overflows at a flow rate of 1427 gpm and gravitates to the Thickener Overflow Tank (not shown on Figure 6-7) from which it is pumped at a controlled rate to the Iron Reduction Tank.

Iron Reduction

In the Iron Reduction Tank, ferric sulfate is almost completely reduced to ferrous sulfate using 0.07t/hr sulfur dioxide gas sparged into the tank.

$$Fe_2(SO_4)_{3(aq)} + SO_{2(g)} + 2\ H_2O_{(l)} \rightarrow 2\ FeSO_{4(aq)} + 2\ H_2SO_{4(aq)}$$

The Iron Removal Tank is covered and kept under a negative pressure to minimize fugitive gaseous SO_2 emissions from the tank. The offgas from the tank passes to the Final Gas Scrubber where the sulfur dioxide is recovered to a liquor stream.

The discharge from the Iron Reduction Tank is pumped directly to the PGM Precipitation Pipe Reactor where 1.4t/hr sodium hydrosulfide (NaHS) is added and metal and metal sulfide precipitation occurs via the following reactions.

$Na_3AuCl_{4(aq)}$ + $NaHS_{(aq)}$ $\rightarrow$ $2Au_{(s)}$ + $7NaCl_{(aq)}$ + $HCl_{(aq)}$ + $S_{(s)}$

$Na_2PdCl_{4(aq)}$ + $NaHS_{(aq)}$ $\rightarrow$ $sPd_{(s)}$ + $3NaCl_{(aq)}$ + $HCl_{(aq)}$ + $S_{(s)}$

$Na_2PtCl_{4(aq)}$ + $NaHS_{(aq)}$ $\rightarrow$ $sPt_{(s)}$ + $3NaCl_{(aq)}$ + $HCl_{(aq)}$ + $S_{(s)}$

$Na_3RhCl_{6(aq)}$ + $3NaHS_{(aq)}$ $\rightarrow$ $2sRh_{(s)}$ + $9NaCl_{(aq)}$ + $3HCl_{(aq)}$ + $3S_{(s)}$

$2FeSO_{4(aq)}$ + $2NaHS_{(aq)}$ $\rightarrow$ $2FeS_{(s)}$ + $Na_2SO_{4(aq)}$ + $H_2SO_{4(aq)}$

$2NiSO_{4(aq)}$ + $2NaHS_{(aq)}$ $\rightarrow$ $2NiS_{(s)}$ + $Na_2SO_{4(aq)}$ + $H_2SO_{4(aq)}$

$2CoSO_{4(aq)}$ + $2NaHS_{(aq)}$ $\rightarrow$ $2CoS_{(s)}$ + $Na_2SO_{4(aq)}$ + $H_2SO_{4(aq)}$

$2CuSO_{4(aq)}$ + $2NaHS_{(aq)}$ + $\rightarrow$ $2CuS_{(s)}$ + $Na_2SO_{4(aq)}$ + $H_2SO_{4(aq)}$

The PGM Precipitation Pipe Reactor discharges slurry at an average flow rate of 1294 gpm, comprising the process liquor mixed PGM sulfide precipitates with some base metal impurity. This slurry is directed to the Clarifier and with aid of flocculant an underflow of 80% solids is achieved.

The Clarifier Underflow is then pumped to the PGM Precipitate Surge Tank (not shown on Figure 6-7), which has a retention time of 8 hours, thus providing surge capacity between the PGM Filter and Clarifier. The PGM Filter separates the PGM precipitate solids from the process liquor containing the metal values of interest. Residual entrained metal values are recovered by washing the PGM precipitate with Service Water and recycling to the Clarifier.

A PGM Precipitate Filter Cake of 80% solids is produced by the filter. The cake is discharged to the Releach Autoclave Feed Tank (not shown on Figure 6-7) where the cake is repulped to achieve 60% by weight solids and then fed to the Releach Autoclave.

The Clarifier Overflow, 441t/hr, contains the remaining base metals that are then recovered in downstream process stages, the copper by solvent extraction and the nickel and cobalt by precipitation.

PGM Releach

The Releach Autoclave removes base metal sulfides from the PGM precipitate. The Releach Autoclave operates at a temperature of 239°F with a design residence time of 120 minutes, at a pressure of approximately 102 psi to preferentially leach base metal sulfides. Oxygen gas is injected into the autoclave and additional sulfuric acid is added to the autoclave feed to maintain 5000 ppm H_2SO_4 in the Autoclave Discharge. There is also a small net generation of heat from the dissolution of base metal sulfides.

Leached slurry (containing base metals) exiting the Releach Autoclave is brought to atmospheric pressure through the Releach Autoclave Flash Vessel, which allows the removal of excess heat through the release of steam. The autoclave vent and flash steam flows are small and so these flows are combined with a number of other offgas streams and passed through the autoclave gas scrubbing systems (two stages of gas cleaning and one of heat recovery as described above). The two offgas lines from the Releach Autoclave will be tied into the autoclave gas scrubbing systems in such a manner as to allow the Releach Autoclave to remain operational when a primary autoclave is offline for maintenance.

The Releach Autoclave Flash Vessel underflow then discharges to a Releach Autoclave Let Down Tank (not shown on Figure 6-7) where it is further cooled to 140°F by cooling coils located within the tank. Cooling water is supplied as required to maintain the target temperature. The letdown tank has a residence time of 8 hours to provide surge capacity between the releach circuit and the PGM Concentrate Filter. Cooling water is supplied from the Cooling Tower to the letdown tank at 86°F and returned at 113°F.

The cooled slurry, 12.6 short tons/hr and 0.4% by weight solids, is then pumped to the PGM Concentrate Filter to produce a PGM Concentrate Filter Cake of 80% solids, which is washed in the filter with Service Water. The PGM Concentrate is a final product for sale to a refinery. The wash filtrate liquor is recycled to the Iron Reduction Tank.

PRE NEUTRALIZATION

The aim of the Pre Neutralization area is to remove acid from the copper extraction circuit feed using limestone to precipitate acid as gypsum, thus reducing the process liquor stream acidity to a value appropriate for copper solvent extraction, avoiding parallel copper and iron precipitation.

The feed to the pre neutralization circuit is a liquor stream of 390 short tons/hr of overflow from the Clarifier following PGM precipitation with NaSH. The Clarifier Overflow is pumped to a Pre Neutralization Tank that is followed by three Gypsum Removal Tanks (not shown on Figure 6-7). Each tank is agitated and baffled.

All of the reaction tanks in this area, comprising the Pre Neutralization Tank and the three Gypsum Removal Tanks are covered and kept under a negative pressure to minimize fugitive emissions from the tanks. The offgas streams from the tanks, which are predominantly CO_2 evolved by the reaction of limestone and acid, are combined and pass to the Final Gas Scrubber.

Limestone slurry is added in part to the Pre Neutralization Tank along with recycled gypsum slurry, from the underflow of the Pre Neutralization Thickener, to seed gypsum precipitation.

The following neutralization/precipitation reactions occur in the pre neutralization circuit:

$$CaCO_{3(s)} + H_2SO_{4(aq)} \rightarrow CaSO_{4\,(s)} + CO_{2(g)}$$

$$CaCO_{3(s)} + 2HCl_{(aq)} + Na_2SO_{4(aq)} \rightarrow CaSO_{4(aq)} + 2NaCl_{(aq)} + CO_{2(g)} + H_2O_{(l)}$$

$$CaSO_{4(aq)} + 2H_2O_{(l)} \rightarrow CaSO_4.2H_2O_{(s)}$$

The neutralized slurry from the gypsum precipitation circuit is pumped by the duty Transfer Pump (not shown on Figure 6-7) to the Pre Neutralization Thickener. The Pre Neutralization Thickener produces a dewatered underflow containing 40% by weight solids that is predominantly gypsum.

Approx 70% of the underflow is recycled to the Pre Neutralization Tank for gypsum seeding and the remainder is pumped to the Gypsum Filter press to separate gypsum residue. A final Gypsum Filter Cake of 60% by weight solids is produced, which is then washed with HM Process Water, and then repulped and pumped to the Reactive Residue Facility. Wash water filtrate is recycled back to the Pre Neutralization Thickener.

The overflow from the Pre Neutralization Thickener is sent to the Copper Solvent Extraction (SX) area.

COPPER SOLVENT EXTRACTION (SX)

Overview

Copper is selectively removed from the process liquor stream by the process of solvent extraction. In turn, the copper is removed from the solvent by a strip liquor solution at a suitable copper concentration for the subsequent Copper Electrowinning area.

The copper recovery process takes place in three extraction stages and two strip stages, with a single wash stage in between to remove entrained aqueous phase liquor from the loaded (copper-containing) organic phase prior to stripping. The solvent extraction plant operates in closed circuit with the electrowinning plant, acid from the latter circuit is used to strip the extracted copper in the two stripping stages.

Copper Extraction

Pre Neutralization Thickener Overflow gravitates to a Copper Extraction Feed Tank (not shown on Figure 6-7). The thickener overflow is then pumped to the copper extraction circuit for copper recovery through the SX Feed Cooler (not shown on Figure 6-7) which cools the copper process liquor stream from ~140 °F to 104°F. The SX Feed Cooler is a plate and frame heat exchanger supplied with cooled water from the Cooling Tower. The cooled process liquor stream is then retained in a PLS Tank (not shown on Figure 6-7) ahead of the SX circuit; the PLS Tank has a four hour residence time at design process liquor stream flows.

From the PLS Tank, liquor is pumped to the copper solvent extraction stages, at a rate approx 2,100 gpm and 14,000 ppm copper. Each extraction stage includes two mixer tanks providing a total of 3 minutes retention time during which the organic and aqueous phases are intimately mixed. During mixing, the extraction of copper occurs into the organic phase according to:

$$Cu^{+2}_{(aq)} + 2R\text{-}H_{(org)} \rightarrow CuR_{2(org)} + 2H^{+}_{(aq)}$$

The organic phase solvent consists of a commercial chelating agent (to assist in dissolution and precipitation) mixed in a high-flashpoint hydrocarbon. Further description of reagents is given in response to Question 20.

The aqueous / organic mixture flows from the second mixer tank into a Reverse Flow Settler, where the two phases are allowed to separate. “Picket fences” located at the top of each settler distribute the flow across the width of the settler. The two phases are collected in separate launders at the end of the settler. The aqueous and organic streams flow counter current through the SX extraction circuit. Process liquor stream enters the first extraction stage and flows sequentially through the other two stages. Raffinate (the aqueous solution remaining after extraction of the copper) leaves the last stage at 1780 gpm and gravitates to the Raffinate Holding Tank (not shown on Figure 6-7). The duty Raffinate Pump (not shown on Figure 6-7) is used to pump the raffinate in part to the initial concentrate leach Autoclave Cooling Water Tank (not shown on Figure 6-7), at approximately 75% of the total raffinate flow, with the remainder pumped to a bleed circuit, for nickel and cobalt recovery.

Stripped organic from the first stage of copper stripping gravitates to the last extraction stage and flows counter to the organic stream through the second and first extraction stages. Loaded organic from the first extraction stage gravitates to a Loaded Organic Tank which feeds a coalesceor wash stage to remove entrained aqueous to reduce carry-over of entrained aqueous solution. The loaded organic is then pumped to the first strip stage. Spent electrolyte from the copper tankhouse is used to strip the loaded organic in two stages. Loaded electrolyte leaving the second stage strip settler flows by gravity to the Electrolyte Filter Feed Tank (not shown on Figure 6-7).

Copper Stripping Stages

Contacting the loaded organic stream with highly acidic aqueous spent electrolyte (containing about 180,000 ppm sulfuric acid) during the strip mixing reverses the copper loading reaction thus consuming acid and increasing the copper content of the electrolyte stream while simultaneously depleting the organic phase of copper. During mixing the stripping of copper occurs in exchange for acidity from the spent electrolyte, thus regenerating the organic for further copper extraction.

$$\mathrm{CuR}_{2(\mathrm{org})} + 2\mathrm{H}^{+}_{(\mathrm{aq})} \rightarrow \mathrm{Cu}^{+2}_{(\mathrm{aq})} + 2\mathrm{R\text{-}H}_{(\mathrm{org})}$$

The barren organic is recycled to the extraction circuit. Following stripping, the loaded electrolyte passes to Copper Electrowinning (EW) where the copper is recovered as metal from the solution. The copper-loaded electrolyte is referred to as “advance electrolyte”.

Stripping of the loaded organic is carried out by counter-current flow in two mixer/settler stages. The aqueous phase, consisting of spent electrolyte is pumped to the first stage primary mix tank and then flows by gravity through to the second stage. As the quantity of spent electrolyte is considerably smaller than the organic input, the stripping stages operate with a large internal electrolyte recycle to maintain an organic/aqueous ratio of 1:1 internal to the settler. Concentrated sulfuric acid is added as required to compensate for that lost in the EW bleed. The stripped organic is transferred from the second stage settler directly to the third extraction stage.

Advance electrolyte is discharged from the second stage settler by gravity to the Electrolyte Filter Feed Tank (not shown on Figure 6-7) and pumped to a coalescing filter with anthracite/garnet dual media. The filter traps organic droplets and any solids remaining entrained in the electrolyte. Periodically the filter is drained and backwashed with water. The resultant liquor is then held in a storage tank and bled at a controlled rate to the Raffinate Tank (not shown on Figure 6-7). Loss of organic (entrained and dissolved) will be between 10 and 50 ppm in the aqueous stream leaving the stripping circuit. From the Electrolyte Filter (not shown on Figure 6-7), clean advance electrolyte is discharged into the Advance Electrolyte Tank (not shown on Figure 6-7), providing about one hour of surge between the Electrolyte Filter and the electrowinning circuit.

"Crud" is a colloquial term used in the industry to describe the result of contamination of the organic/aqueous reagents by dust, silica and even insects. Crud formation at the organic/aqueous interfaces in the settlers inhibits solvent extraction and contributes to organic loss. Crud is removed from the settlers by decanting, drainage and routinely by use of a portable, air operated crud pump.

Crud is pumped to a Crud/Spillage Holding Tank (not shown on Figure 6-7) to allow crud to be accumulated and then treated on a batch basis. From the Crud/Spillage Holding Tank (not shown on Figure 6-7), the crud is pumped to the Organic Recovery Tank (not shown in Figure 6-7). This tank is fitted with several off-takes at different depths. In general, fairly clear aqueous solution can be drained from the bottom of the tank while a clean organic layer should form on the surface. However, a mixed layer, consisting mostly of frothy crud, will remain in the transition layer in the middle of the tank and may require filtration to recover the remaining organic.

The clean organic material from the top of the Crud/Spillage Holding Tank (not shown on Figure 6-7) reports to the Loaded Organic Tank while the aqueous solution and transitional phase material from the bottom and middle of the tank report to a small filter press. To enable filtration of the crud, a pre-coat/filtration medium (clay) is added to a Clay Makeup Tank (not shown on Figure 6-7) by hand and gravitates to the Crud Spillage/Holding Tank (not shown on Figure 6-7) prior to being mixed with the other constituents before pumping to a small filter press. The filter cake containing unwanted solids and residual organic material is discarded to the Reactive Residue Facility, while filtrate is delivered to the Loaded Organic Tank.

COPPER ELECTROWINNING (EW)

From the Filtered Electrolyte Tanks (not shown on Figure 6-7) in the Copper SX area, copper-rich advance electrolyte solution is pumped to the Electrolyte Re-circulating Tank (not shown on Figure 6-7) in which the advance electrolyte is mixed with spent electrolyte recycle from electrowinning, water make-up, spillage (if free of solids), as well as plating agents (guar gum). Cobalt sulfate is also added to maintain the required concentration of 100ppm cobalt in the electrolyte.

From the Electrolyte Re-circulation Tank (not shown on Figure 6-7) the combined electrolyte flow is pumped through strainers to remove tramp material before being distributed to the electrowinning cells. Tramp will be only minor trash that will be cleaned out and properly disposed annually. A duty/standby transformer rectifier arrangement provides the cell electrical supply.

A ventilation system comprising covers for each electrowinning cell and an associated EW Mist Scrubber will be provided for the electrowinning cells. This system collects gases vented from the electrowinning cells and then scrubs the off-gas to remove entrained acid mist before venting to atmosphere. The EW Mist Scrubber water discharges back to the Pre Neutralization Thickener.

Copper is plated onto stainless steel cathode blanks over a cycle of approximately seven days. When the full copper deposit has plated, an overhead crane is used to remove one third of the cathodes from each cell in sequence. The cathodes are then water washed to remove the copper-bearing electrolyte, and immediately stripped in an automatic stripping machine. Wash liquor is recycled to the SX circuit to recover the contained copper. Stripped cathodes are bundled sampled and weighed in the stripping machine and the bundles of cathodes are then removed by forklift to a laydown area prior to shipping.

Spent electrolyte leaving the production cells gravitates to the Electrolyte Re-circulation Tank (not shown on Figure 6-7). The majority of the electrolyte is re-circulated to the electrowinning cells but sufficient spent electrolyte is recycled back to stripping to balance the advance electrolyte flow entering electrowinning from stripping. Spent electrolyte passes through strainers before being pumped through a plate heat exchanger using cooling water to maintain the spent electrolyte temperature below approximately 104°F. Tramp will be only minor trash that will be cleaned out and properly disposed annually.

A small amount of electrolyte is bled out of the SX/EW circuit back to the solvent extraction circuit in order to prevent iron and impurity build-up in the electrolytic circuit.

A plating agent (guar gum) is made up in an agitated tank and pumped to the circuit using a dosing pump at an addition rate proportional to cathode production. There are two spillage pumps in the circuit, each of which pumps back to the Electrolyte Re-Circulation Tank (not shown on Figure 6-7).

BLEED STREAM PURIFICATION

Iron/Aluminum Removal

A stream comprising 25% of the primary raffinate flow from Copper Extraction is removed as a bleed stream. The bleed stream is first treated to remove iron and aluminum and subsequently remove nickel and cobalt, and finally to precipitate magnesium before the remaining solution is recycled to the process. The first stage of the bleed circuit involves neutralization of excess acid and precipitation of iron.

The bleed stream, 470 gpm, at 104°F and ~20,000 ppm H_2SO_4, is pre-heated with steam in the Preheat Tank (not shown on Figure 6-7) to increase the stream temperature to 176°F. The Preheat Tank discharge is then pumped to a series of five cascaded Iron (Fe) Removal Tanks into which air, O_2 and limestone slurry are added.

All of the reaction tanks in this area, comprising the Preheat Tank, and the five Iron (Fe) Removal Tanks are covered and kept under a negative pressure to minimize fugitive emissions from the tanks. The offgas from the tanks is combined and passes to the Final Gas.

The underflow from the Aluminum (Al) Removal Thickener, the next downstream purification stage, is also recycled to the head of the Iron (Fe) Removal Tanks in order to combine the aluminum and iron residues for separation in a single filter step. The main reactions occurring in this circuit are listed below:

Acid Neutralization:
$CaCO_{3(s)} + H_2SO_{4(aq)} + H_2O_{(l)} \rightarrow CaSO_4.2H_2O_{(s)} + CO_{2(g)}$

Iron Oxidation:
$2FeSO_{4(aq)} + H_2SO_{4(aq)} + 1/2O_{2(aq)} \rightarrow Fe_2(SO_4)_{3(aq)} + H_2O_{(l)}$

Iron Precipitation:
$Fe_2(SO_4)_{3(aq)} + 3CaCO_{3(s)} + 9H_2O_{(l)} \rightarrow 2Fe(OH)_{3(s)} + 3CaSO_4.2H_2O_{(s)} + 3CO_{2(g)}$

Slurry gravity overflows from the fifth Iron (Fe) Removal Tank to a high rate Iron (Fe) Removal Thickener where flocculant is added to aid solids settling. The thickener feed stream enters the thickener with a solids density of approximately 8% by weight. The thickener produces a thickener underflow stream at 55% by weight solids, which is pumped by Iron Removal Thickener Underflow Pumps (not shown on Figure 6-7) to the Iron/Aluminum (Fe/Al) Removal Filter Surge Tank (not shown on Figurer 6-7). The combined Iron/Aluminum (Fe/Al) Removal Filter is fed from the Iron/Aluminum (Fe/Al) Removal Filter Surge Tank (not shown on Figure 6-7). The surge tank has a residence time of 8 hours to allow for online filter maintenance.

A final Fe/Al Removal Filter Cake of 80% by weight solids is produced, which is then washed with HM Process Water, to remove soluble metal values and the washed cake is then repulped and pumped to the Reactive Residue Facility. Wash water filtrate is divided and recycled to both the Aluminum (Al) Removal Tanks and Aluminum (Al) Removal Thickener.

Iron (Fe) Removal Thickener Overflow, containing only ppm levels of iron, is directed to two Aluminum (Al) Removal Tanks to which limestone is added. The pH is further raised to ensure aluminum precipitation but to minimize co-precipitation of copper, nickel, cobalt and zinc via the following reactions:

$CuSO_4 + 2aH_2O = sCu(OH)_2 + aH_2SO_4$

$NiSO_4 + 2aH_2O = sNi(OH)_2 + aH_2SO_4$

$ZnSO_4 + 2aH_2O = sZn(OH)_2 + aH_2SO_4$

$CoSO_4 + 2aH_2O = sCo(OH)_2 + aH_2SO_4$

$Al_2(SO_4)_3 + 6aH_2O = 2sAl(OH)_3 + 3aH_2SO_4$

$H_2SO_4 + sCaCO_3 = aCaSO_4 + aH_2O + gCO_2$

$2aHCl + sCaCO_3 = aCaCl_2 + aH_2O + gCO_2$

$CaCl_2 + aNa_2SO_4 = aNaCl + aCaSO$

$CaSO_4 + 2aH_2O = sCaSO_4.2H_2O$

The slurry from the Aluminum (Al) Removal Tanks, 555 gpm at 0.9% solids, is combined with wash water filtrate from the Iron/Aluminum (Fe/Al) Removal Filter and directed to the Aluminum (Al) Removal Thickener. With the aid of flocculant an underflow of 50% by weight solids is achieved and recycled back to the first Iron (Fe) Removal Tank, via a Re-pulp Tank (not shown on Figure 6-7). The Aluminum (Al) Removal Thickener Overflow, at a flow rate of 533 gpm and ~140°F, contains approximately 700 ppm Cu, 500 ppm Co, 1,000 ppm Zn, and 10,000 ppm Ni that is first treated to recover the contained copper separately prior to the subsequent Co/Ni recovery stage.

Copper Removal

The objective of the copper removal stage is to remove soluble copper from the process liquor stream before nickel and cobalt precipitation. This is achieved by precipitating the copper as a copper sulfide. Sodium Hydrosulfide (NaHS) is used to precipitate the copper sulfide along with sulfides of minor amounts of other heavy metals present.

The overflow from the Aluminum (Al) Removal Thickener gravitates to the Aluminum Overflow Tank (not shown on Figure 6-7), which provides surge capacity between the Aluminum (Al) Removal Thickener and the copper removal circuit. The liquor is pumped to a Preheating Tank (not shown on Figure 6-7) where steam is directly sparged into the tank to increase the stream temperature to 158 °F. The stream then gravitates to a De-aeration Tank (not shown on Figure 6-7) where nitrogen is introduced to remove any dissolved oxygen from the solution prior to NaHS addition. The overflow from the De-aeration Tank then gravitates to the Copper Precipitation Tanks where NaHS is introduced on a staged basis as required by the reaction extents.

The following precipitation reactions occur:

$$2CuSO_{4(aq)} + 2NaHS_{(aq)} \rightarrow 2CuS_{(s)} + Na_2SO_{4(aq)} + H_2SO_{4(aq)}$$

$$6CuS_{(s)} + 2CuSO_{4(aq)} + 2H_2O \rightarrow 4Cu_2S_{(s)} + 2S_{(s)} + 2H_2SO_{4(aq)} + O_{2(g)}$$

$$2NaHS_{(aq)} + 2ZnSO_{4(aq)} \rightarrow 2ZnS_{(s)} + Na_2SO_{4(aq)} + H_2SO_{4(aq)}$$

$$2NaHS_{(aq)} + 2CoSO_{4(aq)} \rightarrow 2CoS_{(s)} + Na_2SO_{4(aq)} + H_2SO_{4(aq)}$$

$$2NaHS_{(aq)} + 2NiSO_{4(aq)} \rightarrow 2NiS_{(s)} + Na_2SO_{4(aq)} + H_2SO_{4(aq)}$$

All of the reaction tanks in this area, comprising the Preheating Tank, the De-aeration Tank, and the three Copper Precipitation Tanks are covered and kept under a negative pressure to minimize fugitive emissions from the tanks. The off-gas from the tanks is combined and passes to the Final Gas Scrubber.

The Copper (Cu) Precipitation Tank Discharge gravitates to the CuS Filter Surge Tank (not shown on Figure 6-7) that provides buffer storage between the copper removal circuit and the subsequent CuS Filter. The precipitated solids are separated from the stream using a plate and frame CuS Filter that produces a filter cake of 80% solids containing a mixed sulfide precipitate. No washing of the solids occurs on the filter. The filter cake intermittently discharges from the filter into a hopper and is then conveyed into a re-pulp tank where the filter cake is repulped using Service Water to a density of 65% solids. A proportion of the slurry, ~75%, is recycled to the Copper (Cu) Precipitation Tank to provide a seed for the precipitation of sulfides, while the remaining 25% is recycled back to the Autoclave Feed Tank, to releach and recover the contained metals, predominantly copper. The CuS Filter Filtrate is collected in a Filtrate Tank (not shown on Figure 6-7).

The CuS Filter Filtrate contains minimal copper levels and is pumped to the final purification area, which involves Hydroxide Precipitation of remaining heavy metals.

HYDROXIDE PRECIPITATION

Hydroxide Product Precipitation

The objective of Hydroxide Precipitation is to remove remaining heavy metal species from solution in two hydroxide precipitation stages. The first stage utilizes magnesium oxide slurry (30%) and air to precipitate a mixed Ni/Zn/Co/Mg hydroxide, which is separated from the process slurry in a filter press as a saleable mixed hydroxide product for off-site refining. In the second stage precipitation, lime is added to further raise the pH in order to precipitate the majority of the remaining heavy metal hydroxides for recycle back to the post autoclave leach slurry via the Autoclave Let Down Tank.

Filtrate from the CuS Filter is pumped at a controlled rate from the Filtrate Tank (not shown on Figure 6-7) to the first of a series of three 1st Stage Hydroxide Precipitation Tanks. Air is sparged to these tanks and staged addition of magnesium oxide is provided to facilitate precipitation of a range of metal hydroxides.

The major reactions include:

$4FeSO_{4(aq)} + O_{2(g)} + 2H_2SO_{4(aq)} \rightarrow 2Fe_2(SO_4)_{3(aq)} + 2H_2O_{(l)}$

$NiSO_{4(aq)} + Mg(OH)_{2(s)} \rightarrow Ni(OH)_{2(s)} + MgSO_{4(aq)}$

$CoSO_{4(aq)} + Mg(OH)_{2(s)} \rightarrow Co(OH)_{2(s)} + MgSO_{4(aq)}$

$ZnSO_{4(aq)} + Mg(OH)_{2(s)} \rightarrow Zn(OH)_{2(s)} + MgSO_{4(aq)}$

$H_2SO_{4(aq)} + Mg(OH)_{2(s)} \rightarrow 2H_2O_{(l)} + MgSO_{4(aq)}$

$2HCl_{(aq)} + Mg(OH)_{2(s)} \rightarrow 2H_2O_{(l)} + MgCl_{2(aq)}$

The resulting discharge from the first stage of hydroxide precipitation, at a flow rate of approximately 552 gpm and 1.5% by weight solids, gravitates to the 1st Stage Hydroxide Thickener and, with the aid of flocculant, an underflow of 40% by weight solids is achieved containing the precipitated heavy metals. This underflow is pumped to a Hydroxide Filter Surge Tank (not shown on Figure 6-7) that has a residence time of 8 hours to allow for filter maintenance. The slurry is then pumped at a controlled rate from that tank into the Hydroxide Filter to produce a filter cake of 75% by weight solids. The filter cake is washed with Service Water to remove entrained process solution, and the filter wash water is reports to the 2nd Stage Hydroxide Precipitation Tank.

The final Mixed Hydroxide Product has an approximate composition of 80% $Ni(OH)_2$, 9% $Zn(OH)_2$, 4% $Co(OH)_2$, and 7% $Mg(OH)_2$. The filter cake is intermittently discharged into a hopper beneath the filter and is conveyed to a storage hopper ahead of a bagging plant that packages the filter cake for shipment to a refiner.

Second and Third Stage Hydroxide Precipitation

The 1st Stage Hydroxide Thickener Overflow is pumped to the first of three 2nd Stage Hydroxide Precipitation Tanks. Lime is added to these tanks to raise the pH to a higher level than achieved in the first stage precipitation process to ensure precipitation of all remaining heavy metal species resulting in a mixed hydroxide/gypsum precipitate. Discharge from the tanks gravitates to the 2nd Stage Hydroxide Thickener. Flocculent is added to settle the hydroxide precipitates and produce an underflow product at a density of 40% by weight solids. The underflow product is then recycled back to the Autoclave Let Down Tank (not shown on Figure 6-7) where the high acidity will ensure that the metals contained in the precipitate will be redissolved, and to aid cooling of the leach slurry to a temperature suitable for pumping.

The 2nd Stage Hydroxide Thickener Overflow is then pumped to a final purification stage in the Magnesium Precipitation area for the removal of remaining magnesium to allow the solution to be recycled to the process as HM Process Water.

MAGNESIUM PRECIPITATION

The objective of Magnesium Precipitation is to remove the majority of the remaining magnesium from the process stream exiting the Hydroxide Precipitation area to prevent magnesium build-up in the process liquors throughout the entire hydrometallurgical leach circuit, and to minimize the concentration of other metals present in the recycled process liquor to avoid impurity build-up in the process liquor.

The feed to this circuit is overflow from the 2nd Stage Hydroxide Thickener, at a flow rate of approx 553 gpm and 158°F, which then gravitates through a series of four cascaded Magnesium (Mg) Removal Tanks. Lime slurry is stage added to each reactor as required to facilitate magnesium precipitation via the following reaction:

$$MgSO_{4(aq)} + Ca(OH)_{2(s)} \rightarrow Mg(OH)_{2(s)} + CaSO_{4(aq)}$$

These reaction tanks will be covered due to the elevated temperature but as no gas evolution is expected the tanks will not be connected to the plant gas scrubbing system.

The resulting slurry exiting the Magnesium (Mg) Removal Tanks gravitates to the Mg Removal Thickener where, with the aid of flocculant, an underflow product of 50% by weight solids of magnesium hydroxide is produced. This underflow is then pumped to the Reactive Residues Facility. The thickener overflow gravitates to the Magnesium Removal Thickener Overflow Tank (not shown on Figure 6-7) and is pumped back to the HM Process Water Tank for reuse.