Scoping Study for a Recommended Grassroots Acid Mine Drainage Treatment and Rare Earth Element Extraction Facility

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Dr. Wayne Seames

Dr. Ian Foerster

Scoping Study

Group 10D

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Dr. Wayne Seames & Dr. Ian Foerster May 15th, 2025

Project Supervisors
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Title: Scoping Study to Recommend a Grassroots Acid Mine Drainage Treatment Facility with Rare Earth Element Extraction

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Dear Dr. Seames and Dr. Foerster,

Please see the attached revised scoping study for a grassroots acid mine drainage (AMD) treatment and rare earth element extraction facility. The recommended proposal is a profitable alternative to the costly, on-going maintenance of the Berkley Pitt in Butte, Montana. The proposal will remediate 8.5 MT/year of AMD annually while generating 31 tons/year of mixed rare earth oxides (REOs), 360 tons/year of cooper sulfide and 870 tons/year of zinc sulfides. The project has an NPV@24 of $160 million +/-40% at 20 years of operation and a DCFROR of 37%. The proposal will contribute to the US rare earth supply chain, a strategic priority of the Department of Energy, while reducing the ecological burden of the Berkley Pit.

This report provides the present situation at the Berkley Pit, an overview of the rare earth metal supply chain and demand drivers, as well as process design information at the input and output (I/O), block flow diagram (BFD), and process flow diagram (PFD) levels of resolution. A rough project schedule and environmental and safety statement are included as well. The economic analysis of this report includes a budget brief, capital and operating cost summaries, cashflow summary, and multiple economic sensitivity analyses. This report also includes a raw materials list, chemicals and catalysts list, products and by-products list, waste composition list, utilities list, and major equipment list. An alternative process analysis, assumptions list, sample calculations for equipment specifications, intermediate calculations for REE revenue estimates, unit operation simulation results, as well as vendor correspondences are included as appendix items.

The design involves the staged neutralization of AMD with hydrated lime. A slurry enriched in REEs and byproduct metals is settled and withdrawn at pH 4.5 and subjected to acid leaching. The leachate is passed through semi-batch operation D2EHPA and IDA ion exchange (IX) resin beds, which selectively adsorb heavy rare earths & Zn and light rare earths & Cu, respectively. Adsorbed metals are stripped from the resins by a HCl regenerate and sent to downstream selective precipitation. REE precipitates are calcined into mixed rare oxide product streams while byproduct metals are sold as their precipitated sulfide species.

Preliminary design revisions are listed in the report introduction. We request you review this scoping study report and contact our group with any questions prior to our upcoming review meeting.

Best Regards,

Cameron Cooper Olivia DeLong Craig Morton Andrea Williams

Budget Brief **May 15th, 2025**

Title: Grassroots Acid Mine Drainage Treatment & Rare Earth Elements Extraction Facility
Funding Request: $200 million (Basis Date: Feb. 2025)
Project Duration: 30 months

Proposal: The proposed project will treat 2.1 billion gallons of acid mine drainage (AMD) from the Berkley Pit site in Butte, Montana. It will produce 75 tons/year of mixed rare eath oxides, 6 tons/year of light rare earth oxide product, 1000 tons/year of ZnS byproduct and 400 tons/year of CuS byproduct. The design involves the staged neutralization of AMD with hydrated lime. At the pH=4.5 stage, precipitates are settled to a slurry enriched in REEs and byproduct metals. The REE-enriched slurry is subjected to acid leaching. The leachate is passed through semi-batch operation ion exchange (IX) resin beds. Heavy rare earths (HREEs) and Zn are adsorbed in the D2EHPA IX beds, while light rare earths (LREEs) and Cu are adsorbed in the IDA resin beds. The metals are stripped from the resins by hydrochloric (HCl) acid and selectively precipitated. REE precipitates are calcined into mixed rare earth oxide (REO) product streams while byproduct metals are sold as their precipitated species.

Present Situation: The Berkley Pit is a former open-pit copper mine and current Superfund site. The pit requires costly, on-going maintenance to prevent ground water contamination and reduce ecological burden. The water contains ~40 ppm of REEs among other metals. REEs are non-substitutable inputs to the US advanced manufacturing technology. China controls over 70% of REE supply chain. Increasing the domestic REO production capacity is a strategic priority of the United States.

Qualitative Justification: The proposal reduces hazards to human life and ecosystems posed by Berkley Pit AMD. It also contributes to the national effort to secure a US economy based on advanced manufacturing capability, where defense and clean energy technologies are critical areas of focus. The flowsheet developed in this proposal may also be applicable to remediation and value recovery from other AMD sources beyond the Berkley Pit.

Quantitative Justification: The NPV@24% is $160 million +/- 40% over the 20-year lifespan of the project. The project has a DCFROR of 37%. The NPV@24% suggests the project is economically worthwhile

Alternatives: The best alternative design was completed at the BFD resolution level for an additional scandium oxide-enriched product stream. The Sc-recovery flowsheet requires a 30% increase of all major equipment. As Sc^{3+} is the most difficult REE in the site AMD to recover, this alternative was considered uncompetitive.

Hazards: 95% of revenue comes from water treatment, making the NPV@24% insensitive to REE market volatility and REE and byproduct metal recovery rates. Assuming the existing cost-to-treat of the Berkley Pit site remains $0.02/L, the economic risk of the required $200 TCI is 13%. The NPV@24% remains positive for a cost-to-treat greater than $0.013/L.

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Scoping Study for a Recommended Grassroots Acid Mine Drainage Treatment and Rare Earth Element Extraction Facility

1. Introduction

The purpose of this document is to make a recommendation on whether the proposed process to recover rare earth elements (REEs) from AMD is worth pursuing. The information in the report covers the process design for the extraction of rare earth elements (REEs) from acid mine drainage (AMD) through staged neutralization, acid leaching, ion exchange with D2EHPA (di-(2-ethylhexyl) phosphoric acid) and IDA (iminodiacetic acid) resins, and oxalate precipitation. Recovered REEs are calcined into saleable oxides. Our team reccomends proceeding with this project as it indicates a positive net present value over its lifespan of 20 years

The treatment will remediate contaminated water while producing saleable rare earth oxides (REOs) as well as byproduct ZnS and CuS. The process is intended to serve as a cost-covering method of treating AMD by adding revenue sources to existing environmental cleanup efforts. The proposed process is beneficial because it bypasses roughly 75% of overall process costs associated with the initial removal of hard rock deposits and subsequent grinding and leaching.[1]

The objectives of this opportunity are to treat AMD so that it does not have a negative impact on the environment and surrounding communities and to recover REEs that are a product of AMD pollution. The proposed process is designed to be a world scale facility for AMD treatment and will process 24 million liters of fluid daily. At this flow rate, the facility would process 93% of the current volume of the Berkeley Pit over 20 years.[2] This will have a positive impact on the environment and create a market for REEs in the USA. This will decrease the necessity for foreign trade with our competitors.

The AMD is treated by staged precipitation with hydrated lime. The REE-rich precipitate is then leached using sulfuric acid. The REE-rich leachate which is passed through ion exchange columns containing VP OC 1026 D2EHPA and TP 207 IDA resins. These resins have high selectivity for REEs, zinc, and copper over major competing ions. The reason these two were considered is because after leaching you have a relatively high amount of zinc and copper and very low concentrations of REEs. VP OC 1026 resin is known to have a high affinity for zinc and trivalent metal ions, particularly heavy rare earth elements (HREEs), even at low pH values.

[1] Theaker, N. (2025, January 22). *Group meeting with Dr. Nolan Theaker, Research Subject Matter Expert* [Meeting].
[2] WSP. (n.d.). *Berkeley Pit*. WSP. Retrieved April 7, 2025, from https://www.wsp.com/en-us/projects/berkeley-pit#:~:text=At%20one%20mile%20long%20and,arsenic%2C%20cadmium%2C%20and%20zinc

TP 207 resin displays the same characteristics but it has a relatively higher affinity for copper at low pH values.

Oxalic acid is used to selectively precipitate REE oxalates (REOx). This acid precipitates REEs and little else. The remaining liquids zinc/copper-rich and can be precipitated out using sodium sulfide. The REOx are converted to rare earth oxide (REO) and are the primary product. All wastes are directed to a waste pond and disposed of by outside sources.

1.1 ***Market & Geopolitical Background***

REEs are important raw materials for many products including fiber optics, specialty alloys used in aerospace applications, and high-performance magnets used in electric generators, electric drives and control systems using magnetic actuators. [1,2] They have become synonymous with technical innovation and there is rapidly increasing global demand. Demand drivers include the expanding EV market as well as the increasing investment in advanced defense systems around the world. The key REEs for these sectors are praseodymium (Pr), neodymium (Nd), terbium (Tb), and dysprosium (Dy).[4]

Rare earths are traditionally mined from bauxite ore and undergo benefaction into mixed oxides. These oxides are then sold to refineries which produce high purity reduced element metal products which can then be sold to manufacturers. REEs production has become a significant geopolitical issue. China controls roughly 70% of REO production from mined ore and 90% of REO refinement and magnet production.[3] This monopoly of the value chain allows for strategic price manipulation over the entire value chain to discourage to market entry.[4] China's domination of the market has caused concern in the US and its allied countries because REEs are both necessary for military advantage and for survival of economies based on advanced manufacturing technology.

All REEs were listed as critical minerals by the US Geological Survey due to their strategic importance and supply risk.[5]. Pr, Nd, Tb, and Dy were listed as critical materials for energy on the 2023 DOE Critical Materials List.[7] This has led to the development of U.S. DOE-NETL's Feasibility of Recovering Rare Earth Elements program. The implementation of this program demonstrates the growing interest of the U.S. to develop potential processes for the recovery of 90-99 wt% rare earth element oxides.[8]

1.2 Technology Background: REEs from Secondary Sources

AMD treatment facilities are a standard part of current tailings management programs. Hundreds of AMD treatment facilities exist globally. However, the technology for the REE recovery at AMD treatment facilities is only recently being scaled and commercialized. The University of West Virgina has constructed a pilot plant to produce and REE concentrate from AMD.[6] Related technology is being scaled for the extraction of REEs from other secondary sources. The Energy Institute of North Dakota has constructed a pilot plant for the recovery of REEs from waste coal using ion exchange technology.[7] Rivalia Chemical is developing technology for the recovery of REEs from coal ash ponds using solvent extraction with ionic liquids.[8] Phoenix Tailings is constructing a pilot refinery in Massachusetts for processing mining waste into pure REE metals through a novel pyrometallurgical process.[9] UCORE is currently building a pilot facility in Louisiana for refinement of mixed REOs using solvent exchange technology.[10]

1.3 Summary of Changes to Preliminary Design

The following design changes were made after a **second** review of the preliminary design:

i. The pressure profile with more conservative estimates of pumping losses. Pump motors were resized accordingly.
ii. All solid conveyors greater than 3 meters in length were changed from screw conveyors to belt conveyors. Screw conveyors greater than 3 meters in length would require constant maintenance and are much more expensive than belt conveyors.
iii. All vacuum drum rotary filters (RDVFs) were given installed spares. This is necessary because RDVFs are complex rotating equipment and require frequent maintenance.
iv. In Process Area 1, sludge recirculation in paste thickener H-102 and clarifier H-103 was increased to be 3/100ths of total solid formed. This facilitates faster settling rates and bring H-102 and H-103 in closer alignment with literature flowsheets.[11]
v. In Process Area 3, a pH control tank was added in between the D2EHPA and IDA IX columns F-301 and F-303. The IX regenerate was changed from H_2SO_4 to HCl. This prevents precipitation within the columns due to excessive sulfate levels. Precipitation would result in resin fouling and system upsets.
vi. Also in Process Area 3, nitric acid dosing tanks for HCl regenerate streams were removed after it was found that pH would already be sufficiently low for the upstream selective precipitation of Process Area 04.
vii. In Process Area 4, the recovery rates of HREE and LREE in oxalic acid was increased to 100%. Given that three times the stoichiometric demand of oxalic acid is used, 100% REE recovery can be assumed.

viii. Finally, all equipment specifications in Process Area 03 were brought to design conditions. Preliminary design operating parameters were scaled up using safety factors found in Seames (2026).[12] The volume of ion exchange columns in Process Area 3 was increased by 10% retaining their original height:diameter ratio and the power requirements of the pumps were increased by 10% as well.

2. Present Situation

The Berkely Pit in Butte, Montana is a former open-pit copper mine and current Superfund site. AMD formed in the pit due to the natural oxidation of sulfide minerals into sulfuric acid when exposed to water and air. The acidic waters leach heavy metals from the surrounding earth into solution, increasing its toxicity. Pit water contains roughly 40 ppm of REEs.[13]

Groundwater veins and rainfall supply water to the pit, necessitating constant dewatering to prevent overflow into the water table and nearby bodies of water, as show in Figure 1.[14] In 2002, the U.S. Environmental Protection Agency and state of Montana reached an $87 million settlement with responsible mining entities for the remediation of the Berkley Pit. Dewatering and AMD neutralization is currently being carried out by Montana Resources. The Montana Resource has submitted a $75 million grant proposal to the Department of Defense for the construction of a REE extraction facility adjacent to the current dewatering & neutralization facility.[15]

3. Scope of Work

This scoping study includes a breakdown of the present situation and background of the technical factors involved. This includes a layered description of work at the input/output (I/O) level, the block flow diagram (BFD) level and the process flow diagram (PFD) level. An economic analysis was done to analyze the feasibility and profitability of the project. The report also has a major equipment list, a raw materials/byproducts list, a utility requirements list, chemicals and catalysts list, capital cost summary, operating cost summary, and cash flow summary. It includes a fully developed alternative analysis in Appendix A, an assumptions list in Appendix B, sample calculations in appendix C, simulation input/output in appendix D, revenue sensitivity in appendix E, and vendor correspondence in appendix F.

4. Overview Description

The following description is based off IO Drawing 00-A-100/1, which displays overall stream compositions and major reactions.

The system inputs consist AMD, raw materials, and air. AMD is transferred from the Berkley Pit at 1,000,000 kg/h. The inlet stream 1 contains 630 kg/hr of Zn^{2+}, 140 kg/hr of Cu^{2+}, 3.7 kg/h of REEs, 990,000 kg/hr of water, and a total of 11,000 kg of contaminants (Fe, Al, Mn, alkali metals and anions). 5800 kg/hr of air enter the system for the oxidation of Fe^{2+} in stream 2. Raw material in stream 3 consists of 9800 kg/hr of all the chemicals associated with the reactions to form final products (Na_2S, NaOH, H_2SO_4, $Ca(OH)_2$, HCl and Oxalic acid). Raw material in stream 2 consists of all the consumable chemicals used in the process. These are H_2SO_4, Na_2S, $Ca(OH)_2$, oxalic acid, and NaOH and total 9,800 kg/h. Stream 3 consists of the incoming air needed to oxidize Fe^{2+} to Fe^{3+}.

Gas emissions in stream 4 are produced during the calcination of metal oxalates and the unreacted air from the oxidation of iron at a rate of 5,800 kg/h.

Alkali metals consist of Mg^{2+} , Ca^{2+}, Mg^{2+}, Na^{+} and trace K^{+}. Anions consist of $SiO4^{4-}$, PO_4^{3-} , Cl^{-} and F^{-} ions. Waste precipitates are associated with any precipitate that ends up in waste streams with the majority representative of gypsum, $Fe(OH)_2$, $Ca(OH)_2$, and $CaSO_4$. Waste materials of 56,000 kg/hr exit the system in stream 5.

Byproduct metals consist of copper sulfide and zinc sulfide. Byproduct metals exit the system in stream 6 at a total rate of 160 kg/h.

There is a greater partitioning of HREOs than LREOs in the product streams. Stream 7 contains 3.9 kg/h of HREO product. This stream is 97 wt% REOs, and 54 wt% HREOs. Stream 8 contains 1.7 kg/h of LREO product. This stream is 85 wt% REOs and 84 wt% LREOs. These are representative of the final saleable product.

Stream 9 consist of all the treated water that comes from the system of 980,000 kg/hr. All treated water emissions are compliant with the regulation and standards of the National Pollutant Discharge Elimination System.

5. General Description

The REE recovery from AMD consists of 5 process areas. In Process Area 01, raw AMD is neutralized in stages. Purified water is returned to the environment, and a REE-enriched slurry is produced. In Process Area 02, acid leaching and neutralization operations concentrate REEs in a neutralized leachate and produce a high-purity aluminum byproduct stream. In Process Area 03, HREEs and LREEs are concentrated and separated via ion exchange. In Process Area 04, REEs are precipitated out as oxalates by introducing oxalic acid. This area is also the location for precipitation of byproduct sulfides with Na_2S. The rare earth oxalates (REOx) are then oxidized in a rotary vacuum kiln in Process Area 05.

5.1 Process Area 1 – Staged Precipitation

A general description of staged precipitation operations is shown in Drawing 00-A-200/1. In this area, AMD is neutralized in two continuous operation clarifiers. Treated water is returned to the environment. An iron hydroxide (Fe(OH)3) and gypsum ($CaSO_4 \cdot 2H_2O$) rich waste stream and an REE-enriched intermediate product stream are produced.

In Reactor #1, AMD pH is brought from 2.5 to 4.5 by the addition of hydrated lime ($Ca(OH)_2$) in stream 2. Air is fed to the reactor in stream 3 to oxidize all Fe^{2+} to Fe^{3+} .The air also strips dissolved CO2 dissolved in the AMD. Stripped CO2 and excess air leave Reactor #1 in stream 4. The following reactions occur in Reactor #1:

$$4Fe^{2+}(aq) + O_2(aq) + 4H^{+}(aq) \rightarrow 4Fe^{3+}(aq) \qquad \textbf{(1)}$$ [3]

$$Fe^{3+}(aq) + 3(OH)^{-}(aq) \rightarrow 2Fe(OH)_3(s) \qquad \textbf{(2)}$$

$$CO_2(aq) \rightarrow CO_2(g) \qquad \textbf{(3)}$$

$$Mn^{2+}(aq)+2OH^{-}(aq)\rightarrow Mn(OH)_2\,(s) \qquad \textbf{(4)}$$

$$Ca^{2+}(aq)+SO_4^{2-}(aq)\rightarrow CaSO_4{\cdot}2H_2O \qquad \textbf{(5)}$$ [4]

Oxidizing all Fe^{2+} to Fe^{3+} in Eq. 1 allows for the formation of $Fe(OH)_3$ shown in Eq. 2. The saturation index of $Fe(OH)_3$ is much higher than any Fe^{2+} mineral precipitates. Oxidation in

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[3] Eq 1.1 equilibrium data is calculated from Yuan, F. (2018). Study on kinetics of Fe (II) oxidized by air in FeSO4-H2SO4 solutions. *Minerals Engineering, 121*, 164-168. https://doi.org/10.1016/j.mineng.2018.03.013

[4] Eq 1.2 – 1.5 equilibrium data is taken from AMDTreat simulation results, as shown in Appendix D.1.

Reactor #1 enables the removal of 99% of total Fe ions at pH = 4.5. The stripping of dissolved CO_2 in Eq. 3 prevents $Ca(OH)_2$ consumption by carbonate complexes and precipitates.[5]

Precipitation reactions must be carried out in Reactor #1 because the rate of Fe^{2+} oxidation is negligible at low pH values.

Stream 5 enters Separator #1 at pH = 4.5. Mineral precipitates are settled to a slurry, which is sent to Separator #2 in stream 6.

Excess $Ca(OH)_2$ is added to Separator #2 in stream 8 to bring the pH to 7, at which the majority of all remaining MMC ions precipitate out of solution. Flocculant is added to Separator #2 in stream 7 to increase solid-settling rate. The following precipitation reactions occur in Separator #2:

$$Al^{3+}(aq)+2OH^{-}(aq)\rightarrow Al(OH)_3(s) \qquad \textbf{(6)}^{6}$$

$$Zn^{2+}(aq)+2OH^{-}(aq)\rightarrow Zn(OH)_2(s) \qquad \textbf{(7)}$$

$$Cu^{2+}(aq)+2OH^{-}(aq)\rightarrow Cu(OH)_2(s) \qquad \textbf{(8)}^{7}$$

The clarified overflow leaves Separator #2 in stream 9. NPDES permit limits for this specific proposal would determine if stream 9 can be directly discharged to the environment or to an intermediary settling pond. Precipitated solids are settled to a paste and leave Separator #2 in stream 10. Stream 10 paste is sent to tailings storage.

The overflow from Separator #1 is sent to Separator #3 in stream 11. Excess $Ca(OH)_2$ is added to Separator #3 in stream 13 to bring pH to 7. Flocculant is added in stream 12 to increase solids settling rate. Precipitation reactions Eq. 6 – Eq. 8 occur in Separator #3. The clarified overflow leaves Separator #3 in stream 14. Stream 14 is either returned directly to the environment or to an intermediary settling pond. Mineral precipitates are settled to a slurry, which is sent to Separator #4 in stream 15.

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[5] Office of Surface Mining Reclamation and Enforcement. (n.d.). *Help instruction file: Decarbonation module overview*. Retrieved May 5, 2025, from https://www.osmre.gov/programs/reclaiming-abandoned-mine-lands/amdtreat

[6] Eq 1.6 equilibrium data is taken from AMDTreat simulation data, as shown in Appendix D.1.

[7] Eq 1.7 and 1.8 equilibrium data is taken from **Wang, L., et al. (2014).** Selective precipitation of copper and zinc over iron from acid mine drainage by neutralization and sulfidization for recovery. *International Journal of the Society of Materials Engineering for Resources, 20*, 136–140. https://doi.org/10.5188/ijsmer.20.136).)

5.2 Process Area 2 – Acid Leaching & Neutralization

A general description of acid leaching and neutralization operations is also shown in BFD Drawing 02-A-200/2.

Stream 15 solids slurry enters Separator #4 at a pH of 7. H_2SO_4 is added to separator #4 to bring pH to 1. REEs are desorbed from anionic sites the solids slurry and transferred to the aqueous phase. The primary dissociation reaction that occurs in separator #4 is:

$$CaSO_4{\cdot}2H_2O\ (s) \rightarrow Ca^{2+}(aq)+SO_4^{2-}\ (aq) \qquad \textbf{(5)}$$

Undissolved solids leave Separator #4 in stream 17 and are sent to tailings storage. The liquid overflow leaves Separator #4 in stream 18 and enters Separator #5. Metal ion recoveries into stream 18 leachate are taken from Ziemkiewicz, P. (2023).[8]

NaOH is added to Separator #5 in stream 19 to bring the pH to 4.5. The following reactions occur in Separator #5:

$$Al^{3+}(aq)+2OH^{-}\ (aq)\rightarrow Al(OH)_3\ (s) \qquad \textbf{(6)}$$

$$Zn^{2+}(aq)+\ 2OH^{-}\ (aq)\rightarrow Zn(OH)_2\ (s) \qquad \textbf{(7)}$$

$$Cu^{2+}(aq)+2OH^{-}(aq)\rightarrow Cu(OH)_2\ (s) \qquad \textbf{(8)}$$

$$4Al^{3+}(aq)+SO_4^{2-}\ (aq)+10OH^{-}(aq)\rightarrow Al_4(SO_4)(OH)_{10}\ {\cdot}xH_2O(s) \qquad \textbf{(9)}$$

Eq. 9 shows the formation of hydrated basaluminite ($Al_4(SO_4)(OH)_{10}\ {\cdot}xH_2O$). *x* values may range from 5 to 36.[9] The potential adsorption of REEs onto basalumite is one of the major process risks of this proposal.[10] Basalumite formation is driven by the excessive sulfate concentration in stream 18.

Settled solids leave Separator #5 in stream 20 and are sent to tailings storage. Clarified liquid leaves Separator #5 in stream 21 and is sent to Separator #6 in BFD Drawing 02-A-200/1.

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[8] Ziemkiewicz, P. (2023). *Development and testing of an integrated acid mine drainage (AMD) treatment and rare earth/critical mineral plant: Final scientific/technical report*. West Virginia University Research Corporation.

[9] Caraballo, M. A., Wanty, R. B., Verplanck, P. L., Navarro-Valdivia, L., Ayora, C., & Hochella, M. (2019). Aluminum mobility in mildly acidic mine drainage: Interactions between hydrobasaluminite, silica and trace metals from the nano to the meso-scale. *Chemical Geology, 519*, 72–84. https://doi.org/10.1016/j.chemgeo.2019.04.013

[10] Lozano, A., Ayora, C., & Fernández-Martínez, A. (2019). Sorption of rare earth elements onto basaluminite: The role of sulfate and pH. *Geochimica et Cosmochimica Acta, 258*, 50–62. https://doi.org/10.1016/j.gca.2019.05.016

5.3 Process Area 3 - Ion Exchange via Chelating Resins

The ion exchange process functions on principles of selective adsorption to ion exchange resins and desorption regulated by acids which regenerate the resin and open adsorption sites for reuse. Initial concentration, contact time, and pH impact selectivity. At high feed rates, a component with high initial concentration will be adsorbed. At low feed rates, however, increased contact time allows for more competition between ions based on selectivity. The pH of solution impacts the ability of certain ions to adsorb to resin surface. Additionally, the pH can be used to control the adsorption of desired ions relative to undesired ions.

Different functional groups for resins allow for adsorption of desired high affinity ions. Di-2-ethyhexyl phosphoric acid (D2EHPA) functional groups have a high affinity for zinc and trivalent ions. Selectivity is shown as follows: $VO^{2+} = UO^{2+} = Fe^{3+} > In^{3+} > Al^{3+} > Pb^{2+} = Zn^{2+} > Ca^{2+} > Cd^{2+} = Mn^{2+} > Mg^{2+} > Co^{2+}$. [11] Iminodiacetic Acid (IDA) functional groups have high affinity for copper and trivalent ions, particularly at pH of 3 or lower. At pH ~ 1, the extraction percentages for REEs and Copper can exceed values of 88%.[12] At a pH ~ 2 the resin has very low affinity for alkali elements.[13] The selectivity is shown as follows: $Fe^{3+} > Cu^{2+} > VO^{2+} > UO^{2+} > Pb^{2+} > Ni^{2+} > Zn^{2+} > Cd^{2+} > Co^{2+} > Fe^{2+} > Mn^{2+} > Ca^{2+} > Mg^{2+} >>> Na^{2+}$.[14] Resins Lewatit VP OC 1026 and TP 207 have been chosen for their HREE/Zinc and LREE/Copper selectivity, respectively.

The model for mass balances was based off equilibrium data from studies on the effect of pH and adsorption for each. A conservative approach was taken and components that did not adsorb were modeled as 1% adsorbed to resin. The wash step was assumed to rid the entire column of unabsorbed metallic cations. The elution with HCl was assumed to elute 100% of the adsorbed ions. Equilibrium data suggests for D2EHPA resin, low pH promotes the selective extraction of zinc and REEs. Extraction percentages under these conditions for D2EHPA resins were up to 88% for Zinc and 90% for certain REEs based on data from recent studies.[15,16] On the

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[11] Lewatit. (2024). *Lewatit VP OC 1026 Product Information Sheet*. https://lanxess.com/en-us/products-and-brands/products/l/lewatit--vp-oc-1026. *Accessed May 2025*.
[12] Amphlett, J., Sharrad, C., & Ogden, M. D. (2018). Extraction of uranium from non-saline and hypersaline conditions using iminodiacetic acid chelating resin Purolite S930+. *Chemical Engineering Journal*, 342, 133-141. https://doi.org/10.1016/j.cej.2018.01.090
[13] Xie, Y., Li, M., Liu, K., Xu, F., Wang, X., & Xun, Y. (2022). Removal of Alkali and Alkaline Earth Elements from Cobalt Solutions Using Iminodiacetic Acid Chelating Resin. *SSRN Electronic Journal.*
[14] Lewatit. (2024). *Lewatit TP 207 Product Information Sheet*. https://lanxess.com/en-us/products-and-brands/products/l/lewatit--tp-207. *Accessed May 2025*.
[15] Cortina, J. L. (2021). Valorisation options for Zn and Cu recovery from metal influenced acid mine waters through selective precipitation and ion-exchange processes: promotion of on-site/off-site management options. Journal of environmental management, 283, 112004. https://doi.org/10.1016/j.jenvman.2021.112004
[16] Hermassi, M. & Granados, Mercè & Valderrama, César & Ayora, C. & Cortina, Jose. (2021). Recovery of Rare Earth Elements from Acidic Mine Waters by integration of a Selective Chelating Ion-Exchanger and a Solvent Impregnated Resin. Journal of Environmental Chemical Engineering. 9. 105906. 10.1016/j.jece.2021.105906.

other hand, IDA resins are highly selective for Fe^{3+} and Cu^{2+} at a pH of 3 and lower.[17] Extraction percentages for IDA resins at low pH values were up to 90% and 99% for copper and REEs, respectively.[18]

A general description of ion exchange operations is shown in Drawing: 03-A-200/1.

Two separation unit operations in series are utilized to selectively adsorb desired feed components. Stream 21 from acid leaching, process area 2, is rich in desired components (zinc, copper, HREEs, & LREEs) and has an incoming pH ~ 3. Zinc and HREEs are adsorbed by the D2EHPA resin in separator #6. The effluent, stream 26, is sent from separator #6 to separator #7 (IDA resin) to adsorb the remaining REEs, copper, and zinc ions that were not adsorbed in separator #6. The residual fluid in separator #6 is washed with deionized water, stream 23 to rid the system of contaminants. Separator #7 is washed with deionized water in stream 33. The discarded residual fluid leaves as stream #24, and stream #36 for separator #8, and goes to waste storage for disposal. 10 wt% HCl is fed in stream 22 to separator #7, and in stream 34 to separator #8, to regenerate the resins via elution. Resins are regenerated by the displacement of adsorbed ions by hydrogen ions (H^+) of the regenerating acid. **The pH of the resin bed decreases and this causes decreased retention of adsorbed ions.** The eluent from separator #7 is concentrated zinc & HREEs and exits as stream 25 and is sent to process area 4. The eluent from separator #8 is concentrated copper & LREEs and exits as stream 37 and is sent to process area 4.

5.4 Process Area 4 – Oxalate & Sulfide Precipitation

Due to the selectivity of the REEs, oxalic acid was chosen to precipitate out the ions. One study shows that the oxalates will precipitate out as fast as 2 mins[23]. The pH is also dropped due to the conversion of REEs to REOx. This must be avoided because at lower pH values oxalates are more soluble. Oxalates are precipitated out in reactor 6 where **100% conversion can be assumed**. The chemical equation for this can be seen in Eq 10.

$$2REE^{+3} + 3C_2H_2O_4 + 10H_2O \rightleftharpoons REE_2(C_2O_4)_3 \cdot 10H_2O_{(s)} + 6H^+ \qquad \textbf{\textit{(10)}}$$

Zinc and copper are large byproducts that are readily precipitated by the addition of Na_2S. Precipitation kinetics have yielded high recoveries and reaction times for the process. The

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[17] Lenntech. (2020). AMBERLITE IRC 748, Industrial Grade Chelating Resin for Metals Removal. https://www.lenntech.com/Data-sheets/Rohm-&-Haas-Amberlite-IRC-748-L.pdf

[18] Roa, A., López, J., & Cortina, J. L. (2024). Selective separation of light and heavy rare earth elements from acidic mine waters by integration of chelating ion exchange and ligand impregnated resin. *The Science of the Total Environment*, 954, 176700. https://doi.org/10.1016/j.scitotenv.2024.176700

reaction rates (k) for the formation of sphalerite (ZnS) and covellite (CuS) k values were found to be 0.02470 s^{-1} and 0.0217 s^{-1}, respectively.

High recoveries of sphalerite and covellite are governed primarily by solubility and sulfide to metal molar ratio ($n_{s/m}$). Studies suggest that optimal recoveries of sphalerite are when the $n_{s/m}$ is greater than 2. The maximum precipitation rate was found to be 99%.[19] The optimum $n_{s/m}$ for the formation of covellite is greater than 2.5 yielding a precipitation rate of 99%. [20] At increased concentrations of S^{2-} in solution the conversions of Zinc and Copper can reach up to 99%.[21] The reaction is shown below:

$$Zn^{2+}(aq) + S^{2-}(aq) \rightarrow ZnS\ (s) \tag{11}$$

$$Cu^{2+}(aq) + S^{2-}(aq) \rightarrow CuS\ (s) \tag{12}$$

The sphalerite and covellite precipitation was assumed to coincide with the data from Q. Li et al. and Mei-qing SHI et al., respectively. Utilizing a conservative approach a recovery rate of 80% was assumed since the data suggested 100% recovery for both zinc and copper within the first 5 minutes.[19,22]

A general description of oxalate and sulfide precipitation operations is shown in Drawing 00-A-200/3. Stream 25 enters the oxalate acid precipitation reactor 2. Stream 28 doses the vessel with three times the stoichiometric amount of oxalic acid. The precipitation is expected to happen within minutes and can be modeled continuously. The precipitates are sent in stream 27 to separator 8 in drawing 00-A-200/5.

The liquids from reactor 2 in stream 29 enter the sulfide precipitation vessel, reactor 3, and is dosed with 50 wt% Na_2S from stream 30. Stream 32 exits the sulfide precipitation vessel where the solids, ZnS precipitates, are sent to product storage and the liquids exit in stream 32 to waste storage for disposal.

Stream 37 undergoes the exact same process description as stream 25 and is shown in Drawing 00-A-200/4. After the absorption column, stream 37 is then sent to a separate

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[19] Meiqing, S., Min, X., Shen, C.H., Chai, L., Ke, Y., Yan, X., & Liang, Y. (2021). Separation and recovery of copper in Cu−As-bearing copper electrorefining black slime by oxidation acid leaching and sulfide precipitation. *Transactions of Nonferrous Metals Society of China*, 31, 1103-1112.

[20] SHI, Mei-qing & MIN, Xiao-bo & SHEN, Chen & CHAI, Li-yuan & Ke, Yong & YAN, Xu & LIANG, Yan-jie. (2021). Separation and recovery of copper in Cu−As-bearing copper electrorefining black slime by oxidation acid leaching and sulfide precipitation. *Transactions of Nonferrous Metals Society of China*. 31. 1103-1112. 10.1016/S1003-6326(21)65564-4.

[21] Van Hille, Rob & Peterson, Karen & Lewis, Alison. (2005). Copper sulfide precipitation in a fluidised bed reactor. *Chemical Engineering Science - CHEM ENG SCI*. 60. 2571-2578. 10.1016/j.ces.2004.11.052.

[22] Li, Qi, Zhongqing Xiao, and Wencai Zhang. "Sulfide Precipitation Characteristics of Mn, Ni, Co, and Zn in the Presence of Contaminant Metal Ions." Minerals engineering 215.C (2024): 108814-. Web.

precipitation reactor 4. Stream 38 doses the vessel with three times the stoichiometric amount of oxalic acid. The solids are then sent to the next process area.

Stream 40 is then sent to reactor 5. Here stream 41 doses the vessel with sodium sulfide to precipitate out copper. Stream 42 is our product stream that gets sent to storage. All of the remaining material in the vessel is sent to wastewater treatment in stream 43.

5.5 Process Area 5 – Oxide Formation

A general description of oxide formation operations is shown in Drawing 00-A-200/5. Streams 27 and 39 are sent to Reactor #7 and #8, respectively, where the oxalates are converted to oxides. This is done in a kiln at 800° C. There have been some studies to see the effect of temperature and residence time on yield, but after one hour at this temperature, we can assume 100% conversion of oxalates to oxides. The chemical equation for this is show below:

$$REE_2(C_2O_4)_3 = REE_2O_3 + 6CO_2 \qquad \textbf{\textit{(13)}}$$

6. Detailed Description

6.1 Software Justification

AMD has complex solution chemistry. Many possible species may precipitate given the addition of a base. The base chosen also impacts precipitate species. AMDTreat Titration Module, a simulation software developed by the US Geological Survey, was used to simulate the solid precipitate quantity and composition. It was also used to calculate how much $Ca(OH)_2$ would be required in the staged neutralization and tailings thickening in Process Area 01. Simulation I/O for these mass balances can be found Appendix D.1.

Total $Ca(OH)_2$ used is reported in Table 3. AMDTreat was also used to simulate the capital costs of Sparging Vessel F-101, Stage 1 Clarifier H-101 and Stage 2 Clarifier H-103. Simulation I/O for these capital costs can be found in Appendix D.2.

Aspen Plus was used to simulate requirements for the compressors. Isentropic compressors were used and assumed to have 70% efficiency. Molar flow rates were determined from hand calculations and were specified as functional parameters. Inlet flows were assumed to be from storage at 101 kPa and room temperature. Sample calculations for finding molar flow required and screenshots of Aspen results can be found in Appendix D.3.

6.2 Process Area 01: Staged Precipitation with Hydrated Lime

The staged precipitation of AMD with $Ca(OH)_2$ is shown in Drawing 01-A-300. All slurry % solids in the following description of wt%, free water basis. 1,000,000 kg/h of AMD feed is pumped by L-101A/B in stream 15 at the Berkley Pit the annual average temperature 10 ºC. This is the annual average temperature in the top 15m of water. The seasonal fluctuation over this depth is +/- 5 ºC.[30,31] **Pump L-101A/B is sized to overcome the total head pressure of 10 m of water at the discharge into F-101**. F-101 is an open-air vessel operating at 101 kPa. Blower G-202 delivers air to the sparging system of F-101 in stream 2 at **230 ºC and 230 kPa**. G-202 is sized to bring Montana summer temperature air to a pressure of 220 kPa. This pressure is sufficient to overcome the pressure of 10m of water at its discharge as well as the pressure drop of the sparging system. The sparging system is a fine bubble plate diffusor, with a negligible fouled pressure drop of 11 kPa.[32] Excess air and stripped CO_2 leave F-101 in stream 5 at 30 ºC and **110kPa**. Excess 50% $Ca(OH)_2$ slurry is fed to F-101 in stream 4 at 20 ºC and **150 kPa**. **Chemical addition and aeration lead to a temperature rise of 5 ºC.** Aerated solution and < 1wt% solid precipitates are pumped in stream 5 by L-102A/B at 15 ºC and **150 kPa** to Stage 1 Clarifier H-101. H-101 is open to the atmosphere and is sized for 4 hours of residence time, which allows precipitates to settle to a 2% solids slurry. Negligible heat loss in Stage 1 Clarifier is assumed. The solid slurry leaves H-101 in stream 5 and is pumped to 103 A/B/C to Paste Thickener H-102 at 15 ºC and **150 kPa**. Excess 50% $Ca(OH)_2$ slurry and flocculant are fed to H-102 in stream 8 and 9, respectively. Flocculant is fed 46:1 ratio of TSS ppm following literature recommendations.[33] The pH of H-102 is maintained at 7, which leads at which nearly all Al^{3+} and Mn^{2+} ions precipitate out of solution as $Al(OH)_3$ and $Mn(OH)_2$. Solids settle to a 50% solids paste in 1 h of residence time. Steady-state solids accumulation in H-101 generate 50 Pa of yield pressure, allowing for the dewatering of bottom-layer sludge.[34] Sludge recirculation pump L-106A/B/C **recycles 80 kg/h of sludge** to the top of H-102 in stream 10 to increase settling rates. **The clarified overflow from H-102, stream 11, is discharged to the environment or settling pond per NDPES permit requirements by L-104 at a pH of 7, 15 ºC and 150 kPa.** The 50% solids paste is transported by Gypsum/Fe **Belt** Conveyor J-101A/B.

In Sheet 2 of Drawing 01-A-200, the overflow from Stage 1 Clarifier H-101 in stream 7 is sent to Stage 2 Clarifier by pump L-105A/B at a pH of 4.5 and 15 ºC, **Stream 7 overcomes a total head pressure of 7m and enters H-103 at 150 kPa.** Flocculant and excess 50% $Ca(OH)_2$ slurry is fed to H-103 in streams 13 and 14, respectively. H-103 is open to the atmosphere and is sized for 4 hours of residence time, which allows precipitates settle to a 17% solids slurry. The settling rate in Stage 2 Clarifier is substantially higher than Stage 1 Clarifier due to its higher fraction of solid formation per unit volume of solution. Simulation I/O for Stage 1 and Stage 2 clarifiers can be found in Appendix D.

The clarified overflow from H-103, stream 15, is discharged to the environment or settling pond per NDPES permit requirements by L-108 at a pH of 7, 15 ºC and 150 kPa.

The underflow from H-103 is pumped by L-109A/B/C in stream 17 to Acid Leaching Vessel F-301 at 15 °C and **300 kPa.**

6.3 Process Area 02: Acid Leaching and Neutralization

Stream 17 enters Acid Mixing Vessel F-301 in Drawing 02-A-300/1 at **150 kPa** and 15 °C. 730 kg/h of 95wt% H2SO4 is dosed to F-201 at 20 °C in stream 18 to bring slurry pH to 1. The slurry leaves the bottom F-201 as a 10% solids in stream 19 at 16 °C and **160 kPa. Stream 19 overcomes 4m of vertical rise to enter the horizontal Leachate Settling Vessel H-201 at 120 kPa.** H-201 is sized for a residence time of 6 hours.[35] A 50% solid slurry underflow from H-201 is sent to Leachate Rotary Drum Vacuum filter (RDVF) H-220 by pump L-202A/B/C at 16 °C and **220 kPa** in stream 20. H-201 dewaters the slurry to an 80% solids gypsum stream. The solids are transported at **atmospheric pressure** to tailings storage by **Belt** Conveyor J-201A/B in stream 23. The filtrate from H-201 is sent to Leachate Neutralization Vessel F-401 in stream 22 by L-204A/B at 16 C and **202 kPa**. The clarified overflow from Leachate Settling Vessel H-201 is sent to F-401 in stream 21 by **L-203A/B** at 16 °C and **202 kPa**. All equipment in contact with leachate and precipitated solids are and vessels are Alloy 20 or Alloy 20 clad.

Drawing 02-A-300/2, clarified leachate enters F-401 after overcoming 5m of vertical rise at pH = 1, 16 °C and 150 kPa. 450kg/h of 50% NaOH solution at **20 °C** is dosed in stream 24 to bring the pH to 4.5 **and cause 1 °C of temperature rise**. A 5% solid slurry leaves the bottom of F-401 in stream 25 at 17 °C and **160 kPa**. **Stream 25 overcomes 4m of vertical rise to enter the horizontal Leachate Settling Vessel H-203 at 120 kPa.** H-203 is sized for a residence time of 6 hours. A 10% solid slurry underflow from H-203 is sent to Neutralized Leachate RDVF H-204 in stream 27 by **L-205A/B/C** at 17 °C and **220 kPa**. H-204 dewaters the slurry to a 90% solids gel consisting primarily of hydrated aluminum sulfates, ($Al_4(SO_4)(OH)_{10} \cdot xH_2O$), and $Al(OH)_3$. The solids are transported to tailings storage by **Belt** Conveyor J-202A/B in stream 28. **The liquid filtrate leaves H-204 at 110 kPa and is sent IX Dosing Tank F-203 in stream 29 by L-206A/B**. **Stream 29 overcomes 7m of vertical rise and enters F-203 at 17 °C and 120 kPa.** The overflow from H-203 is sent to F-203 in stream 26 **by L-206A/B** at 17 °C and 120 kPa. Sufficient H_2SO_4 is dosed to F-203 in stream 30 to bring pH to 3. F-203 is sized for 4 residence time to allow for pH control and flow rates dynamics **before being pumped by L-209A/B at 17 °C and 350 kPa to D2EHPA IX Column F-301 in Process Area 03.**

6.4 Process Area 03: REE Concentration via Ion Exchange

Shown in Drawing 03-A-300/1, IX column D-301A/B separates zinc and HREE uses the selective adsorption properties of D2EHPA resin. Stream 31 from F-203 comes in at a pH ~ 3 and 290 kPa to overcome pressure drop from inlet piping, sprayer/nozzles, and liquid head pressure. The effluent (lean leachate) from D-301 as stream 32 is at a pressure 150 kPa and enters D-302, the pH control tank, to maintain a steady incoming pH of 2.5 to D-303A/B. The lean leachate exits the pH control tank (F-302) as stream 36 where L-301A/B increases the pressure to overcome the pressure drop associated with D-303A/B inlet piping, sprayers/nozzles, and liquid head pressure. The effluent (depleted leachate) from F-303A/B is stream 39, which goes to pump L-303A/B where the pressure is increased to 250 kPa and is sent to wastewater storage/treatment.

Streams 33 and 40 are deionized water streams from storage. The incoming pressure is 290 kPa to overcome pressure drop of the IX columns D-301A/B and D-303A/B, respectively. The DI water and contaminants exit the IX columns in stream 34 and stream 39, respectively. L-302A/B increases the pressure of stream 34 to 250 kPa and L-303 A/B increases the pressure of stream 39 to 250 kPa. Both DI wash discharge streams are sent to wastewater treatment/storage.

Streams 35 and 42 are 10 wt% HCl streams from chemical storage and have an incoming pressure of 290 kPa to overcome pressure drop of the IX columns D-301A/B and D-303A/B, respectively. The eluent from D-301A/B exits the column bottom as stream 38 and the pressure is increased to 250 kPa. The eluent from D-303A/B exits the column bottom as stream 43 and the pressure is increased to 250 kPa. Pump L-302A/B sends stream 38 to R-401 for oxalate precipitation and zinc sulfide formation in process area 04 (DWG #: 04-A-200/1). Pump L-303A/B sends stream 43 to reactor R-403 for oxalate precipitation and copper sulfide formation in process area 04 (DWG #: 04-A-200/2).

D-301A/B and D-303A/B operate in a semi-batch process sequence. While D-301A and D-303A are in the adsorption phase (Step #1), D-301B and D-303B are undergoing the DI wash and resin regeneration phases (Step #2 & Step #3). This allows for continuous operation.

In Drawing 03-A-300/2 the semi-batch operation is shown and is sequenced as follows:

1. Step 1 (Adsorb): Leaching from F-203 in 02-A-300/2 comes into D-301A in stream 31 at a flowrate of 3.2 BV/hr (1 BV ~ 4500 L). The inlet pressure of 290 kPa allows for stream 31 to overcome pressure drop associated with the column. As the fluid runs through the resin bed Zn and HREEs are adsorbed onto the functional group of the resin (D2EHPA). The effluent exits the column bottom in stream 32 at a pressure of 190 kPa and enters F-302 for pH adjustment.
2. Step 2 (Wash): Deionized water is pumped into D-301A in stream 33 at a flow rate of 3.2 BV/hr from storage at a pressure of 290 kPa. The residual fluid, deionized water, and

contaminant metals (MMC) exit the column bottom in stream 34 at a pressure of 100 kPa. It is assumed that all aforementioned contaminants are discarded in wash phase.
3. Step 3 (Regenerate): 10 wt% HCl from product storage enters D-301A in stream 35 for 4.5 hours at a pressure of 290 kPa. The outlet stream 36 is rich with Zn and HREEs. Stream 36 is routed by L-302A/B to R-401 for oxalate precipitation (DWG #: 04-A-400/1).

Drawing 02-A-300/3 shows the batch operation for the separation of components in stream 36, the effluent from Step 1 of IX column D-301A that has been adjusted to a pH ~ 2.5. Stream 36 is rich in copper, zinc and remaining REEs, primarily LREEs. In order to collect copper and LREEs, stream 36 is fed to IDA chelating resin IX column (D-303A/B).

The semi-batch operation for D-303A/B is sequenced as follows:

1. Step 1 (Adsorb): Stream 36, lean leachate from F-302, enters column D-303A at a flow rate of 3.2 BV/h (1 BV ~ 4500 L). The inlet pressure of stream 36 is 290 kPa to overcome any pressure drop associated with the column (piping, nozzles, liquid head). During this time, the copper and LREEs are adsorbed onto the column. Stream 39, the depleted leachate, exits the column where pump L-303A/B increases the pressure to 250 kPa and is sent to wastewater treatment/storage.
2. Step 2 (Wash): Deionized water in stream 40 enters the D-303A at a pressure of 290 kPa and a flow rate of 3.2 BV/hr. The residual fluid, contaminants (MMC) and DI water exit the column bottom in stream 41 at a pressure of 190 kPa. Pump L-303A/B increases the pressure to 250 kPa where it is sent to wastewater storage.
3. Step 3 (Regenerate): 10 wt% HCl is sent to the D-303A at 3.2 BV/h in stream 42. The inlet pressure is 290 kPa. The copper and LREE rich outlet, stream 43, is pumped by L-303A/B to increase the pressure to 290 kPa. Stream 43 is sent to R-403 for oxalate precipitation (DWG #: 04-A-300/2).

6.5 Process Area 04: REE and Sulfide Precipitation

Due to the selectivity of the REEs, oxalic acid was chosen to precipitate out the ions. . One study shows that the oxalates will precipitate out as fast as 2 mins[23]. With oxalic acid, we can assume an almost perfect recovery of the oxalate precipitates in the vessels. The oxalates then need to be converted into oxides. This is done in a kiln at 800° C. There have been some studies to see the effect of temperature and residence time on yield, but after one hour at this temperature, we can assume 100% conversion of oxalates to oxides.°

The detailed description for Process Area 04: REE and Sulfide Precipitation is described below. Major equipment and mass balances are shown in Drawing 04-A-300/1 & 04-A-300/2.

Stream 36 is carried from Process Area 03 to Process Area 04 at 130 kPa. The stream is dumped into R-401, where the oxalate precipitation reaction occurs. Here three times the stoichiometric amount of oxalic acid is dosed from storage as seen on stream 46. When calculating the amount of oxalic acid needed, both the molar basis of REEs and full contaminants were used so there would be no competition for precipitation from contaminants. Precipitation is expected to happen within five minutes, making sure this can be modeled as a continuous process. Oxalic acid is a weak acid and will not fully dissolved, whatever was not consumed in the precipitation reaction does remain in the process stream. The vessel was sized based on the volumetric amount of flow from stream 47, the outlet.

Stream 47 is then sent to a rotary vacuum filter, H-401, by L-401A/B. L-401A/B increases the pressure from 140 kPa to 220 kPA to achieve the pressure difference needed for the rotary drum vacuum filter. The filter attains a 10% moisture content in the solids stream 48. The rest of the fluid is sent to the zinc sulfide precipitation tank, as seen in stream 49. The precipitates are carried by conveyor J-401 to the kiln in process area 5. The conveyor was based off of mass flow rate of the precipitate stream as seen in stream 48 and is assumed to be atmospheric pressure.

Stream 49 is pumped by L-402A/B to the ZnS reaction vessel R-402. L402A/B increases the pressure from 110 kPa to 190 kPa to overcome the height difference to charge into the vessel. Sodium sulfide is charged from storage into R-402 in stream 54. The ZnS (sphalerite) precipitation reaction was modeled looking at three different parameters: percent conversion, column volume, and residence time. The optimal reaction vessel was found to be 1000 L with a residence time of 5.3 minutes. The outlet slurry, stream 55, is sent to the rotary vacuum drum filter H-402A/B through pump L-403A/B. L-403 A/B increase the pressure of stream 55 from 140 kPa to 220 kPa. Dried sphalerite cake is continuously scraped onto conveyor J-402 and transported to ZnS byproduct storage as seen in stream 57 at 101 kPa. Its purity is 99.95 wt%. The remaining liquid is sent to wastewater storage through pump L-404A/B in stream 56. L-404A/B increases pressure from 110kPa to 250 kPa.

Stream 43 is carried from Process Area 03 to Process Area 04 at 130 kPa. The stream is dumped into R-403, where the oxalate precipitation reaction occurs. Here three times the stoichiometric amount of oxalic acid is dosed from storage as seen on stream 58. When calculating the amount of oxalic acid needed, both the molar basis of REEs and full contaminants were used so there would be no competition for precipitation from contaminants. Precipitation is expected to happen within five minutes, making sure this can be modeled as a continuous process. Oxalic acid is a weak acid and will not fully dissolved, whatever was not consumed in the precipitation reaction does remain in the process stream. The vessel was sized based on the volumetric amount of flow from stream 59, the outlet.

Stream 59 is then sent to a rotary vacuum filter, H-403A/B, by L-405A/B/C. L-405A/B/C increases the pressure from 140 kPa to 220 kPA to achieve the pressure difference needed for the rotary drum vacuum filter. The filter attains a 10% moisture content in the solids stream 60. The rest of the fluid is sent to the zinc sulfide precipitation tank, as seen in stream 61. The precipitates are carried by conveyor J-403 to the kiln in process area 5. The conveyor was based off of mass flow rate of the precipitate stream as seen in stream 60 and is assumed to be atmospheric pressure.

Stream 61 is pumped by L-406A/B to the CuS reaction vessel R-404. L406A/B increases the pressure from 110 kPa to 190 kPa to overcome the height difference to charge into the vessel. Sodium sulfide is charged from storage into R-404 in stream 66. The CuS (sphalerite) precipitation reaction was modeled looking at three different parameters: percent conversion, column volume, and residence time. The optimal reaction vessel was found to be 1000 L with a residence time of 5.3 minutes. The outlet slurry, stream 67, is sent to the rotary vacuum drum filter H-404A/B through pump L-407A/B. L-407 A/B increases the pressure of stream 67 from 140 kPa to 220 kPa. Dried sphalerite cake is continuously scraped onto conveyor J-404 and transported to CuS byproduct storage as seen in stream 69 at 101 kPa. Its purity is 99.95 wt%. The remaining liquid is sent to wastewater storage through pump L-408A/B in stream 68. L-408A/B increases pressure from 110kPa to 250 kPa.

6.6 Process Area 05: REE Oxidation

Process Area 06 is the last step in the process and uses a fired heater to convert oxalates to oxides. Drawing 00-A-300/9 shows streams 48 and 60 where they are set into electric kilns B-501 and B-502. The temperature is raised from 12°C to 800°C. Within the electric kilns, oxalates are converted to oxides and carbon monoxide is a by-product as seen in streams 72 and 76. Carbon monoxide is released at 150°C and at 130 kPa. It is assumed that all the remaining water in the stream is flashed off and 100% conversion of oxalates to oxides. The fired heater is assumed to have a 75% efficiency. Due to the small amount of mass flow of REEs, the fired heater does not have a large duty requirement, it only requires 1.1 kW and 2.1 kW.

The electric kilns are modeled as a batch process. The oxalates set in the reactor and then heated up to temperature. It's assumed that the oxalates will reach 100% conversion after an hour of being in the kiln. Once the hour is done, the REEs can be removed and sent to storage. Stream 73 exits the HREE kiln at 3.9 kg/hr of product and stream 77 exits the LREE kiln at 1.7 kg/hr of product. The kilns are assumed to run at atmospheric pressure.

7. Process Lifecycle Outline

The project lifecycle is defined by seven different stages including scoping, project definition, detailed specification, procurement, implementation, documentation, and commissioning and testing. The scoping study phase includes the scoping study and budget brief as deliverables. This phase is considered to be a preliminary design and includes a broad cost estimate (with 40% error) and a summary schedule. The purpose of this phase is to determine if the opportunity is worthwhile to pursue.

The project definition phase has the project proposal, project execution plan, and a process hazards analysis as the deliverables. This is the conceptual design phase and includes a budgetary cost estimate (within 20% error) and a project coordination schedule. The purpose of this phase is to determine the entire scope of work that must be completed.

The detailed specification phase includes design packages, design reviews, and a hazards and operability study. This is the detailed design phase and includes a control cost estimate (within 10% error) and an execution schedule. The purpose of this phase is to give enough detail to allow procurement to gather the resources needed.

The rest of the phases are not considered to have any design phases. The procurement phase has purchase requisitions as a deliverable and requires an updated control cost estimate and a working level schedule. The implementation phase has physical facilities and systems as a deliverable and includes updated control budget and detailed implementation schedule. The documentation phase includes as-built drawings and operating and maintenance manuals as a deliverable. The commissioning and testing phase has the operating plant and performance test reports as the deliverable. This includes a commissioning and test schedule.

8. Economic Assessment

8.1 Broad Cost Estimate

A project economic lifespan of 20 years was assumed. Table 7 outlines the capital cost summary, including unadjusted cost estimates for each piece of equipment from vendor quotes, simulation, and literature values.[23,24] Vendor price books were used to price all pumps. Direct vendor quotes were obtained for ZnS RVDF H-402, Paste Thickener H-102, and Stage 1 Tailings Screw Convery J-101. Capacity adjustments were applied to the RVDF and screw conveyor for

[23] Ulrich, G. D., & Vasudevan, P. T. (2004). *Capital cost estimation in chemical engineering—Process design and economics: A practical guide* (2nd ed.). Process Publishing.
[24] Seames, W. (2026). Process Design, Economics, and Project Engineering (1st ed.). CRC Press. https://doi.org/10.1201/9781003509882

equipment use elsewhere in the flowsheet. Vendor correspondences can be found in Appendix E.

Base costs from vendor, simulation and literature price charts were multiplied by basis date, material of construction (MOC), pressure and equipment estimating factors to calculate total line item (TLI) costs. TLIs were summed to find a total direct process cost (TDPC). A basis date of February 2025, corresponding to the most recent CPI index available, was assumed. Additional direct costs were estimated at 30% of TDCP and indirect costs 25% of TDC. The fixed capital investment (FCI) of $110 million was calculated as the sum of the TDPC, additional direct and indirect costs. The initial charge of chemicals and catalysts required for commissioning was calculated assuming a 120 day supply was needed prior to commissioning. The working capital to cover expenses over non-revenue generating years of operation were estimated at 10% of FCI. The working capital was estimated to be $18 million dollars.

The cost for the initial charge of chemical and catalysts and working capital were summed with the FCI to find the total capital investment (TCI). The TCI of the project was estimated at $200 million.

8.2 Operating Cost Summary

Operating costs consist of operating labor, maintenance, utilities, chemicals & catalysts, operating materials, lab charges, raw materials, and indirect costs. Operating labor cost was estimated following guidelines in Seames (2026). The proposal consists of 9 major unit operations and 12 minor unit operations, corresponding to a total of 12 outside operators, 1 board operator and 1 supervisor per shift. A 95% operating factor was assumed, giving 4.5 shifts per week. Loaded operator salaries were assumed to be 130% Bureau of Labor Statistics salaries for water plant operators in Montana. Supervisor salary was estimated at 15% of total operator costs.[25] Total operating labor totals out to $7.4 million.Supervisor salary was estimated at 15% of total operator costs.[26]

Maintenance costs were assumed to be 8% of FCI due to the large presence of complex rotating equipment such as RDVFs. This totals $14 million a year.

Utility costs were dominated by electricity costs. Electricity costs were based off of Montana industrial user rates, which are trending flat at $0.083 USD/kWh.[27] This totals $950,000 a year.

Chemicals and catalysts consisted of make-up resin for IX columns, valued at $90,000/yr.

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[25] U.S. Bureau of Labor Statistics. (2023, May). *Occupational employment and wages, May 2023: 51-8092 Gas plant operators*. U.S. Department of Labor. https://www.bls.gov/oes/2023/may/oes518092.htm#st

[26] U.S. Bureau of Labor Statistics. (2023, May). *Occupational employment and wages, May 2023: 51-8092 Gas plant operators*. U.S. Department of Labor. https://www.bls.gov/oes/2023/may/oes518092.htm#st

[27] U.S. Energy Information Administration. *Electric Power Monthly: Table 5.6.A. Average price of electricity to ultimate customers by end-use sector, by state, 2024 and February 2025*. U.S. Department of Energy. https://www.eia.gov/electricity/monthly/epm_table_grapher.php?t=epmt_5_6_a

Laboratory costs were assumed to be 15% of operating labor costs. Materials were assumed to be 5% of maintenance costs. These both total $1.8 million a year.

Raw material costs were the sum of hydrated lime, caustic, sulfuric acid, nitric acid, sodium sulfide, hydrochloric acid and oxalic acid. Total annual manufacturing costs were found from the sum of labor, maintenance, utility and chemicals and catalysts. Finally, annual indirect costs were assumed to be 35% of the total manufacturing costs. Total operating costs were calculated as $50 million/year. This was done for the duration of the project economic lifetime of 20 years. Operating costs summary can be referenced in Table 8.

8.3 Revenues

Revenues are generated through the sale of HREO and LREO products as well as CuS and ZnS bypdroducts. Revenues were also generated on the basis of cost to treat AMD per liter. Historical biannual (Q1 and Q3) prices from 2022 to 2024 were individually trended to the project basis date of February 2025. All REO prices were taken from the Shanghai Metals Market and converted to USD at the current exchange rate. Post-Covid historical data was not available for Sm_2O_3, Eu_2O_3, Gd_2O_3, Ho_2O_3, Er_2O_3, Yb_2O_3, Lu_2O_3, and Y_2O_3. Average prices for these oxides over 2020-2021 were trended to Q3-2024 using overall REO market trend.[28][29]

The REO revenue was found by the summing the annual revenues of the LREO and HREO product stream. A reduction factor of 60% was applied to account for the fact that the metals will be sold to refineries as mixed products, following the guidelines of Ziemkiewicz, P. (2023).

The byproducts generate $5 million/yr of revenue. The rare earth oxides generate $6 million dollars a year of revenue. The primary revenues are generated per the cost to treat the AMD. At $0.07/L of AMD the annual revenues amount to $160 million dollars annually.

8.4 Taxes & Depreciation

The federal income tax rate of 21% and a Montana state corporate tax rate of 6.75% were used. An 11-year MACRS depreciation schedule was applied to the FCI. Federal and state taxes and the depreciation schedule can be seen in the cash flow summary of Table 9.

8.5 Overall Profitability

The overall profitability of the project is found in the cashflow sheet of Table 9. The proposal has a net present value (NPV) of $160 million ± 40% at a hurdle rate of 24% and a

[28] Institute for Rare Earth Elements and Strategic Metals. "Institute for Rare Earth Elements and Strategic Metals." *Institute for Rare Earth Elements and Strategic Metals*, Rare Earths Institute. Accessed 8 May 2025.
[29] Statista. *Statista*. Statista, https://www.statista.com/. Accessed 8 May 2025.

DCFROR of 37%. The FCI was spread over 30 months. The project schedule was 30 months on the basis that a compressor has the longest lead time of 12 months. This was based off of 40% procurement for most complex piece of equipment, 30% design time, and 30% implementation. . The preliminary economics suggest that the project is financially viable and offers economic sustainability.

9. Environmental and Safety Statement

No process areas have any rated areas for fire and safety.

The facility has 4 planned emissions: the sparging tank off-gas stream in Process Area 01, 2 treated water streams in Process Area 01, the solids tailings discharges of Process Area 01 and 02, the combined the gas stream in Process Area 05, and the collected wastewater stream of Process Areas 03 and 04. Gas emission rates and composition can be found in Table 2. Treated water and wastewater emission rates and compositions can be found in Table 3.

Throughout the process, strong and weak acids and bases are used to adjust pH. Besides potential releases of oxalic acid, nitric acid, sulfuric acid, sodium hydroxide from storage, the process fluid itself ranges from pH 1 – 10. All tanks are sized with excess mixing volume per Seames (2026). Secondary containment around chemical storage and low pH unit operations would be necessary.

10. Economic Hazard Analysis

The key economic hazard of the proposal is the cost to treat AMD. The projects primary revenue stream comes from what the facility charges for water treatment. The costs to treat AMD can range and vary dramatically. Ultimately, the US government pays approximately $0.02 USD/L of AMD treated.

The sensitivity of the NPV with respect to the cost to treat showed that there is a 13% financial risk against the initial capital investment. At the upper bound cost to treat AMD, the NPV@24% is $370 million. At the lower bound cost to treat AMD, the NPV@24% is ($56 million). This shows that there is a slight risk associated with fluctuations in the amount the facility can charge to treat the AMD. Also note the highly conservative hurdle rate of 24%, the hurdle rate accounts for uncertain times yet still gives a positive NPV @24%. Please see Figure 3 for details.

Uncertainty analysis were ran for total direct process costs and operating costs. The upper and lower bounds showed that even at +/-40% of these values there is no risk associated with these parameter. Please see Figure 4 and Figure 5 for details.

The largest revenue driver for product revenues was Sc2O3. The large quantity of Sc2O3 calculated in the product stream was however the greatest departure from literature expectations.

This is due to the high affinity of Sc3+ to Fe and Al precipitates, and basaluminite in particular. To account for this, a basis revenue assuming 50% of modeled Scandium was present in REO product streams. REO trend price calculations as well as revenue sensitivity to scandium recovery and REO market volatility can be found in Appendix E. NVP@24 was insensitive to 0% Sc2O3 production. As before, these effects were overshadowed by the FCI.

Given the positive NPV@24% of $160 million and the 13% risk assessment. The recommendation is to proceed and invest the TCI of $200 million. The conservative assumption of 24% hurdle rate and profits driven by cost to treat give additional economic "cushion" in the estimated values.

11. Broad Comparison of Process Alternatives

Three additional process alternatives were considered. The best alternative process was to produce a mixed REO product stream enriched in Scandium Oxide (ScO_3). This process is shown in Drawing 00-A-500 of Appendix A.

Sc^{3+} is the most valuable metal ion contained in the AMD feed. Scandium recovery requires an additional 4000 m^3 clarifier and creates an additional intermediate product stream that runs through all process areas. This intermediate product stream requires a doubling of all process equipment.

The scandium capture process relies on the higher affinity of the Sc^{3+} ion to anionic sites precipitating species. This high affinity increase the likelihood of Sc^{3+} loss in tailings waste stream, and in particular, the basalumite tailings discarded in Process Area 02. Even if Sc^{3+} remained in the neutralized leachate that is fed to the IX columns of Process Area 03, Leipke (2022) found that Sc^{3+} has such a high affinity for IDA resins they suggested burning the resin itself to recover it.[30] This option may be viable for secondary REE sources that contain a higher fraction of $Sc^{3+.}$ Given the low concentration of Sc^{3+} in the AMD of Berkley Pit, however, economic justification at this site is unlikely.

The additional process alternatives of Solvent Extraction and Membrane Separation can be found in Appendix A.

Lepke, J. P. (2021). *Upgrade of rare earth element concentrate by selective dissolution and ion exchange* (Master's thesis). University of North Dakota. UND Scholarly Commons.

12. Conclusion

The purpose of this report was to provide process information so a recommendation of the REE recovery process capable of recovering 97 wt% pure HREO and 96 wt% LREO streams from a raw AMD feed was worth pursuing. This process has been described in which the AMD is pretreated with staged precipitation and acid leaching and neutralization. The stream is then passed through a series of ion exchange columns using VP OC 1026 D2EHPA and TP 207 IDA resins were used for selective REE recovery. The REE rich stream was then precipitated out as oxalates and then converted to oxides through a fired heater. Copper and zinc sulfide were created from this process as byproducts as well. The proposed process satisfies the objective of making saleable REOs at a respectable purity, driving up the price of the product while retaining a positive NPV over the lifetime of the project. Therefore, it is our recommendation to proceed.

Table 1. Raw Materials List

Raw Material	Amount (kg/h)	
Acid Mine Drainage	1000000	
Components	**Amount (kg/h)**	**Wt. %**
CO2	56	5.6E-05
H2O	990000	9.9E-01
Fe2+ / Fe3+	820	8.2E-04
Zn2+	630	6.3E-04
Al3+	290	2.9E-04
Mn2+	230	2.3E-04
Cu2+	140	1.4E-04
Mg2+	510	5.1E-04
Ca2+	470	4.7E-04
Na+	92	9.2E-05
K+	7.7	7.7E-06
SiO44-	76	7.6E-05
F-	32	3.2E-05
Cl-	12	1.2E-05
PO43-	0.72	7.2E-07
S2-	0	0.0E+00
SO42-	8600	8.6E-03
Sc	0.034	3.4E-08
Y	0.88	8.8E-07
La	0.28	2.8E-07
Ce	1.1	1.1E-06
Pr	0.11	1.1E-07
Nd	0.49	4.9E-07
Sm	0.12	1.2E-07
Eu	0.033	3.3E-08
Gd	0.16	1.6E-07
Tb	0.028	2.8E-08
Dy	0.18	1.8E-07
Ho	0.037	3.7E-08
Er	0.11	1.1E-07
Tm	0.016	1.6E-08
Yb	0.1	1.0E-07
Lu	0.015	1.5E-08

Table 1. Raw Materials List Continued

Raw Material	Amount (kg/h)	
50% Ca(OH)2	3000	
Components	**Amount (kg/h)**	**Wt. %**
Ca(OH)2	1500	50
Water	1500	50
Raw Material	**Amount (kg/h)**	
HCl	30000	
Components	**Amount (kg/h)**	**Wt. %**
H2SO4	24000	25
Water	6000	75
Raw Material	**Amount (kg/h)**	
50% NaOH	300	
Components	**Amount (kg/h)**	**Wt. %**
NaOH	150	50
Water	150	50
Raw Material	**Amount (kg/h)**	
35% NaCO3	150	
Components	**Amount (kg/h)**	**Wt. %**
NaCO3	40	35
Water	110	65
Raw Material	**Amount (kg/h)**	
90% Oxalic Acid	53	
Components	**Amount (kg/h)**	**Wt. %**
Oxalic Acid	48	90
Water	5.3	10
Raw Material	**Amount (kg/h)**	
50% Na2S	280	
Components	**Amount (kg/h)**	**Wt. %**
Na2S	140	50
Water	140	50

Table 2. Products/Byproducts List

Product	Amount (kg/hr)	
HREO Product	3.9	
Components	**Amount (kg/hr)**	**Weight %**
Sc2O3	0.89	22.82%
Y2O3	0.023	0.59%
La2O3	0.47	12.05%
Ce2O3	0.026	0.67%
Pr2O3	0.015	0.38%
Nd2O3	0.0072	0.18%
Sm2O3	0.0041	0.11%
Eu2O3	0.021	0.54%
Gd2O3	0.0069	0.18%
Tb2O3	0.09	2.31%
Dy2O3	0.016	0.41%
Ho2O3	0.062	1.59%
Er2O3	0.0098	0.25%
Tm2O3	0.054	1.38%
Yb2O3	0.0043	0.11%
Lu2O3	1.5	38.46%
MMC	0.0912	2.34%

Table 2. Products/Byproducts List Continued

Product	Amount (kg/hr)	
LREE Oxide Product	1.7	
Components	**Amount (kg/hr)**	**Weight %**
Sc2O3	0.000075	0.00%
Y2O3	0.038	2.24%
La2O3	0.096	5.65%
Ce2O3	0.024	1.41%
Pr2O3	0.014	0.82%
Nd2O3	0.27	15.88%
Sm2O3	0.062	3.65%
Eu2O3	0.018	1.06%
Gd2O3	0.079	4.65%
Tb2O3	0.0052	0.31%
Dy2O3	0.023	1.35%
Ho2O3	0.0038	0.22%
Er2O3	0.0067	0.39%
Tm2O3	0.0005	0.03%
Yb2O3	0.0027	0.16%
Lu2O3	0.00022	0.01%
MMC	0.1113	6.55%

Table 2. Products/Byproducts List Continued

Product	Amount (kg/hr)	
Copper Sulfide	48	
Components	**Amount (kg/hr)**	**Weight %**
CuS	45	93.75%
Product	**Amount (kg/hr)**	
Zinc Sulfide	110	
Components	**Amount (kg/hr)**	**Weight %**
ZnS	110	99.97%
Byproduct	**Amount (kg/h)**	
Treated Water	980000	
Components	**Amount (kg/hr)**	**Weight %**
O2	13	0.001%
H2O	970000	98.98%
Zn2+	360	0.04%
Al3+	96	0.01%
Mn2+	2	0.00020%
Cu2+	30	0.003%
Mg2+	500	0.05%
Ca2+	380	0.04%
Na+	91	0.01%
K+	0	0.0000%
SiO44-	25	0.0026%
F-	11	0.0011%
Cl-	12	0.0012%
SO42-	4400	0.45%

Table 3. Discharged Water Emissions

Component	**Waste Water**	**Treated Water**
Oxalic Acid	34	0
Flocculant	45	0
O2	0.071	13
H2O	46000	970000
Fe2+ / Fe3+	0.011	1.3
Zn2+	32	360
Al3+	16	96
Mn2+	1.4	2
Cu2+	15	30
Mg2+	3.4	500
Ca2+	250	380
Na+	420	91
K+	0	0
SiO44-	0.41	25
F-	0.18	11
Cl-	0.075	12
PO43-	0	0
S2-	63	0
SO42-	2900	4400
Ca(OH)2	120	0
CaSO4:2H2O	2700	6.4
Zn(OH)2	250	1
Fe(OH)3	1300	3.4
Al(OH)3	630	2.2
Mn(OH)2	5.8	0.02
Al4(SO4)(OH)10·4H2O	270	0
Cu(OH)2	110	0.44
Sc	0.14	0.0074
Y	0.42	0.18
La	0.19	0.13
Ce	0.33	0.41
Pr	0.039	0.048
Nd	0.23	0.14
Sm	0.06	0.035
Eu	0.017	0.0065
Gd	0.078	0.039
Tb	0.031	0.0068
Dy	0.077	0.04

Table 3. Discharged Water Emissions Continued

Component	**Waste Water**	**Treated Water**
Ho	0.023	0.0076
Er	0.042	0.021
Tm	0.016	0.0031
Yb	0.038	0.022
Lu	0.33	0.0074
Total	***56000***	***980000***

Table 4. Chemicals, Catalysts, & Adsorbents

Adsorbents	
Initial Requirements	
Resin Type	**Amount (kg)**
Lanxess VP OC 1026	5000
Lanxess TP 207	5000
Makeup Requirements	
Resin Type	**Amount (kg/yr)**
Lanxess VP OC 1026	1000
Lanxess TP 207	1000

Table 5. Utilities List

Utility Description	Utility Conditions	Utility Requirements
Electricity	-	1100 kWh
Air	30 °C	36,000 kg/hr
	101 kPa	

Table 6. Major Equipment List

Equipment Number	Equipment Name/Description	Equipment Specifications
Process Area 01		
F-101 A/B/C	Sparging Vessel	Height = 11 m
		Width = 15 m
		Length = 50 m
		Temperature = 15 C
		Pressure = 101 kPa
		MOC = Epoxy-lined concrete
G-101	Sparging Air Blower	Stages = 1
		MOC = Carbon Steel
		Inlet Pressure = 101 kPa
		Outlet Pressure = 220 kPa
		Inlet Temperature = 30 C
		Outlet Temperature = 134.7 C
		Fluid Components = Air
		Volumetric Flow Rate = 28,000 m3/h STP
		Power = 950 kW
H-101	Stage 1 Clarifier	Height = 10 m
		Diameter = 35 m
		Temperature = 15 C
		Pressure = 101 kPa
		Rake Power = 10 kW
		MOC = Epoxy-lined concrete
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 4.5)
H-102	Tailings Paste Thickener	Height = 9 m
		Diameter = 8 m
		Temperature = 15 C
		Pressure = 101 kPa
		Rake Power = 10 kW
		MOC = Ally 20 Clad Carbon Steel

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
H-103	Stage 2 Clarifier	Height = 6 m
		Diameter = 34 m
		Temperature = 15 C
		Pressure = 101 kPa
		Rake Power = 10 kW
		MOC = Epoxy-lined concrete
J-101A/B	Iron Tailings Belt Conveyer	Mass Flow Rate = 5,500 kg/h
		Length = 30 m
		Arpon ID = 30 cm
		Power = 7.5 kW
		MOC = Stainless Clad Carbon Steel
L-101 A/B	AMD Feed Pump	Power = 65 kW
		Volumetric Flow Rate = 1,000,000 L/h
		Inlet Pressure = 101 kPa
		Outlet Pressure = 250 kPa
		Temperature = 10 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC (pH 2.4)
L-102 A/B/C	Aerated AMD Transfer Pump	Power = 7.9 kW
		Volumetric Flow Rate = 1,000,000 L/h
		Inlet Pressure = 202 kPa
		Outlet Pressure = 220 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 4.5)
L-103 A/B/C	Stage 1 Underflow Pump	Power = 2.6 kW
		Volumetric Flow Rate = 330,000 L/h
		Inlet Pressure = 202 kPa
		Outlet Pressure = 202 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 4.5)

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-104 A/B	Stage 1 Water Pump	Power = 5.6 kW
		Volumetric Flow Rate = 320,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 150 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, (pH 7)
L-105 A/B	Overflow Transfer Pump	Power = 32 kW
		Volumetric Flow Rate = 670,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 250 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 4.5)
L-106 A/B/C	Stage 1 Sludge Recirculation Pump	Power = 0.30 kW
		Volumetric Flow Rate = 80 L/h
		Inlet Pressure = 250 kPa
		Outlet Pressure = 290 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 7)
L-107 A/B/C	Stage 2 Clarifier Sludge Recirculation Pump	Power = 0.27 kW
		Volumetric Flow Rate = 66 L/h
		Inlet Pressure = 250 kPa
		Outlet Pressure = 290 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 7)

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-108 A/B	Stage 2 Water Pump	Power = 12 kW
		Volumetric Flow Rate = 660,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 150 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, (pH 7)
L-109 A/B/C	Preconcentrate Pump	Power = 1.2 kW
		Volumetric Flow Rate = 17,000 L/h
		Inlet Pressure = 220 kPa
		Outlet Pressure = 300 kPa
		Temperature = 15 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings, (pH 7)
Process Area 02		
F-201	Acid Leaching Vessel	Height = 3.8 m
		Diameter = 3.4 m
		Temperature = 16 C
		Pressure = 101 kPa
		MOC = Alloy 20 Clad Carbon Steel
F-202	Leachate Neutralization Vessel	Height = 2.1 m
		Diameter = 2.3 m
		Temperature = 25 C
		Pressure = 101 kPa
		MOC = Alloy 20 Clad Carbon Steel
F-203	Acid Dosing Vessel	Height = 3.5 m
		Diameter = 3.2 m
		Temperature = 17 C
		Pressure = 101 kPa
		MOC = Alloy 20 Clad Carbon Steel
H-201	Leachate Solids Settler	Length = 14 m
		Diameter = 3.7 m
		Temperature = 16 C
		Pressure = 101 kPa
		MOC = Alloy 20 Clad Carbon Steel

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
H-202	Leachate Underflow Rotary Vacuum Drum Filter	Area = 140 m2
		Pore Size = 50 µm
		Air Flow Rate = 353 m3/h (STP)
		MOC = Alloy 20 Clad Carbon Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings, (pH 1)
		Power = .35 kW
H-203	Neutralized Leachate Solids Settler	Length = 12 m
		Diameter = 3.7 m
		Temperature = 16 C
		Pressure = 101 kPa
		MOC = Alloy 20 Clad Carbon Steel
H-204	Neutralized Leachate Underflow Rotary Drum Vacuum Filter	Area = 40 m2
		Pore Size = 50 µm
		Air Flow Rate = 135 m3/h (STP)
		MOC = Alloy 20 Clad Carbon Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings, (pH 4.5)
		Power = .20 kW
J-201 A/B	Leachate Tailings Screw Conveyer	Mass Flow Rate = 3,100 kg/h
		Length = 30 m
		Auger OD = 20 cm
		Power = 4.5 kW
		MOC = Alloy 20 Clad Carbon Steel
J-202 A/B	Neutralized Leachate Tailings Screw Conveyer	Mass Flow Rate = 890
		Length = 30 m
		Flight OD = 15 cm
		Power = 0.75 kW
		MOC = Stainless Clad Carbon Steel

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-202 A/B/C	Gypsum Slurry Pump	Power = 0.20 kW
		Volumetric Flow Rate = 4,000 L/h
		Inlet Pressure = 130 kPa
		Outlet Pressure = 220 kPa
		Temperature = 16 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings, (pH 1)
L-203A/B	Leachate Overflow Pump	Power = 0.47 kW
		Volumetric Flow Rate = 13,000 L/h
		Inlet Pressure = 120 kPa
		Outlet Pressure = 202 kPa
		Temperature = 16 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC (pH 1)
L-204A/B	Gypsum Filtrate Pump	Power = 0.20 kW
		Volumetric Flow Rate = 930 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 202 kPa
		Temperature = 16 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC (pH 1)
L-205A/B/C	Basaluminite Slurry Pump	Power = 0.26 kW
		Volumetric Flow Rate = 6,500 L/h
		Inlet Pressure = 120 kPa
		Outlet Pressure = 220 kPa
		Temperature = 17 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC (pH 1)

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-206 A/B	Neutralized Leachate Overflow Pump	Power = 0.27 kW
		Volumetric Flow Rate = 8,700L/h
		Inlet Pressure = 120 kPa
		Outlet Pressure = 190 kPa
		Temperature = 17 C
		MOC = Alloy 20 Clad Carbon Steel
		Fluid Components = Water, Sulfates, MMC, Suspended Tailings (pH 4.5)
L-207 A/B	Basaluminite Filtrate Pump	Power = 0.20 kW
		Volumetric Flow Rate = 7,800 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 190 kPa
		Temperature = 16 C
		MOC = Alloy 20 Clad Carbon Steel
		Fluid Components = Water, Sulfates, MMC (pH 4.5)
L-208 A/B	IX Feed Pump	Power = 0.56 kW
		Volumetric Flow Rate = 14,000 L/h
		Inlet Pressure = 120 kPa
		Outlet Pressure = 350 kPa
		Temperature = 17 C
		MOC = Alloy 20 Stainless Steel
		Fluid Components = Water, Sulfates, MMC (pH 3)
Process Area 03		
D-301 A/B	VP OC 1026 Ion Exchange Column	Total Volume = 2,000 L
		Diameter = 1 m
		Height = 6 m
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 290 kPa
		Outlet Pressure = 190 kPa
		Temperature = 20 C
		Resin Amount = 5000 kg
		Pressure Drop = 100 kPa
		MOC = Nickel Clad onto Carbon Steel
		Fluid Components = REE, MMCs, Water, Sulfates, HCl

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
F-302	Lean Leachate pH Control Tank	Height = 2.0 m
		Diameter = 3.0 m
		Temperature = 20 C
		Pressure = 150 kPa
		MOC = Nickel Clad onto Carbon Steel
D-303 A/B/C	TP 207 Imino Diacetic Ion Exchange Column	Total Volume = 20000 L
		Diameter = 1 m
		Height = 6 m
		Volumetric Flow Rate = 15000 L/h
		Inlet Pressure = 290 kPa
		Outlet Pressure = 190 kPa
		Resin Amount = 5000 kg
		Temperature = 20 C
		Pressure Drop = 100 kPa
		MOC = Nickel Clad onto Carbon Steel
		Fluid Components = REE, MMCs, Water, Sulfates, HCl
L-301 A/B	Lean Leachate Pump	Power = 0.93 kW
		Volumetric Flow Rate = 14,000 L/h
		Inlet Pressure = 150 kPa
		Outlet Pressure = 350 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REE, MMCs, Water, Sulfates, HCl
L-302 A/B	D2EHPA Discharge Pump	Power = 0.4 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 190 kPa
		Outlet Pressure = 250 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REE, MMCs, Water, HCl

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-303 A/B	IDA Discharge Pump	Power = 0.4 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 190 kPa
		Outlet Pressure = 250 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REEs, MMC, Water, HCl
Process Area 04		
J-401 A/B	HREE/MMC Screw Oxalate Conveyor	Mass Flow Rate = 7 kg/h
		Length = 10 m
		Auger OD = 2.5 cm
		Power = 0.20 kW
		MOC = Stainless Clad Carbon Steel
J-402 A/B	Zinc Sulfide Screw Conveyor	Mass Flow Rate = 110 kg/h
		Length = 30 m
		Auger OD = 10
		Power = 0.20 kW
		MOC = Stainless Clad Carbon Steel
J-403 A/B	LREE/MMC Screw Conveyor	Mass Flow Rate = 1.6 kg/h
		Length = 10 m
		Auger OD = 2.5 cm
		Power = 0.20 kW
		MOC = Stainless Clad Carbon Steel
J-404 A/B	Copper Sulfide Screw Conveyor	Mass Flow Rate = 48 kg/h
		Length = 30 m
		Auger OD = 10
		Power = 0.20 kW
		MOC = Stainless Clad Carbon Steel

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
H-401A/B	HREE/MMC Rotary Drum Vacuum Filter	Area = 0.5 m2
		Pore Size = 4µ
		Air Flow Rate = 1.1 m3/h (STP)
		MOC = Nickel Clad onto Stainless Steel
		Fluid Components = MMC, Water, Oxalic Acid, Oxalates
		Power = 1.07 kW
H-402A/B	Zinc Sulfide Rotary Drum Vacuum Filter	Area = 6 m2
		Pore Size = 50 µm
		Air Flow Rate = 18.8 m3/h (STP)
		MOC = Nickel Clad onto Stainless Steel
		Fluid Components = REE, MMC, Water, Sulfides
		Power = .03 kW
H-403A/B	LREE/MMC Oxalate Rotary Drum Vacuum Filter	Area = 0.1 m2
		Pore Size = 4 µm
		Air Flow Rate = 0.3 m3/h (STP)
		MOC = Alloy 20 Stainless Steel
		Fluid Components = MMC, Water, Oxalic Acid, Oxalates
		Power = 0.25 kW
H-404A/B	Copper Sulfide Rotary Drum Vacuum Filter	Area = 2.4 m2
		Pore Size = 50 µm
		Air Flow Rate = 8.14 m3/h (STP)
		MOC = Nickel Clad onto Stainless Steel
		Fluid Components = REE, MMC, Water, Sulfides
		Power = .08 W
L-401 A/B/C	HREOx Transfer Pump	Power = 0.53 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 140 kPa
		Outlet Pressure = 220 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REEs, MMC, Water, HCl, Oxalic Acid

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-402 A/B	REOx Filtrate Pump	Power = 0.55 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 190 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REEs, MMC, Water, HCl, Oxalic Acid
L-403 A/B/C	ZnS Transfer Pump	Power = 0.57 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 140 kPa
		Outlet Pressure = 220 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = Sulfides, MMC, Water, HCl, Oxalic Acid
L-404 A/B	ZnS Filtrate Pump	Power = 0.93 kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 250 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = Sulfides, MMC, Water, HCl Oxalic Acid
L-405 A/B/C	LREOx Transfer Pump	Power = 0.53kW
		Volumetric Flow Rate = 15,000 L/h
		Inlet Pressure = 140 kPa
		Outlet Pressure = 220 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = MMC, Water, Oxalic Acid

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
L-406 A/B	LREOx Filtrate Pump	Power = 0.55 kW
		Volumetric Flow Rate = 16,000 L/h
		Inlet Pressure = 110 kPa
		Outlet Pressure = 190 kPa
		Temperature = 20 C
		MOC = Alloy 20
		Fluid Components = REE, MMC, Water, Sulfides
R-401	HREOx Precipitation Reactor	Height = 2.0 m
		Diameter = 3.0 m
		Temperature = 12 C
		Pressure = 150 kPa
		MOC = Nickel Clad onto Carbon Steel
R-402	ZnS Precipitation Reactor	Height = 12.7 m
		Diameter = 1 m
		Temperature = 12 C
		Pressure = 150 kPa
		MOC = Nickel Clad onto Carbon Steel
R-403	LREOx Precipitation Reactor	Height = 2.2 m
		Diameter = 3.0 m
		Temperature = 12 C
		Pressure = 150 kPa
		MOC = Nickel Clad onto Carbon Steel
R-404	CuS Precipitation Reactor	Height = 12.7 m
		Diameter = 1 m
		Temperature = 12 C
		Pressure = 150 kPa
		MOC = Nickel Clad onto Carbon Steel
Process Area 05		
B-501	HREO Kiln	Duty = 1.1 kW
		Pressure = 130 kPa
		Power = 1.5 kW
		MOC = Carbon Steel

Table 6. Major Equipment List (Continued)

Equipment Number	Equipment Name/Description	Equipment Specifications
B-502	LREO Kiln	Duty = 1.6 kW
		Pressure = 130 kPa
		Power = 2.1 kW
		MOC = Carbon Steel

Table 7. Capital Cost Summary

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
B-501	HREO Kiln[U]	Duty = 1.1 kW Pressure = 101 kPa Power = 1.5 kW MOC = Carbon Steel	$1,000	1.96525	1.00	1	$1,965	1	$1,965	3.2	$6,289
B-502	LREO Kiln[U]	Duty = 1.6 kW Pressure = 101 kPa Power = 2.1 kW MOC = Carbon Steel	$1,200	1.96525	1.00	1	$2,358	1	$2,358	3.2	$7,547
F-101	Sparging Vessel[S]	Height = 11 m Width = 15 m Length = 50 m Pressure = 101 kPa MOC = Epoxy-lined concrete	$360,000	1	1.00	1	$360,000	1	$360,000	3.2	$1,152,000
F-201	Acid Leaching Vessel[U]	Height = 3.8 m Diameter = 3.4 m Pressure = 101 kPa MOC = Alloy 20 Clad Carbon Steel	$12,500	1.96525	4.50	1	$110,545	1	$110,545	3.2	$353,745
F-202	Leachate Neutralization Vessel[U]	Height = 2.1 m Diameter = 2.3 m Temperature = 7 C Pressure = 101 kPa MOC = Alloy 20 Clad Carbon Steel	$10,000	1.96525	4.50	1	$88,436	1	$88,436	3.2	$282,996
F-203	Acid Dosing Vessel[U]	Height = 3.5 m Diameter = 3.2 m Temperature = 7 C Pressure = 101 kPa MOC = Alloy 20 Clad Carbon Steel	$12,000	1.96525	4.50	1	$106,124	1	$106,124	3.2	$339,595
F-301 A/B/C	VP OC 1026 Ion Exchange Column[U]	Total Volume = 20000 L Diameter = 1 m Height = 6 m Resin Amount = 10000 kg MOC = Nickel Clad onto Stainless Steel	$100,000	1.96525	4.50	1	$884,363	2	$1,768,725	4.4	$7,782,390

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
F-302	Nitric Acid Dosing Tank[U]	Height = 2.0 m Diameter = 3.0 m Temperature = 10 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$50,000	1.96525	2.80	1	$275,135	1	$275,135	3.2	$880,432
F-303 A/B/C	TP 207 Imino Diacetic Ion Exchange Column[U]	Total Volume = 20000 L Diameter = 1 m Height = 6 m Volumetric Flow Rate = 14000 L/hr Resin Amount = 10000 kg Temperature = 10 C MOC = Nickel Clad onto Stainless Steel	$100,000	1.96525	4.50	1	$884,363	2	$1,768,725	4.4	$7,782,390
F-304	Nitric Acid Dosing Tank[U]	Height = 2.0 m Diameter = 3.0 m Temperature = 10 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$50,000	1.96525	2.80	1	$275,135	1	$275,135	3.2	$880,432
G-101	Sparging Air Compressor	MOC = Carbon Steel Inlet Pressure = 101 kPa Outlet Pressure = 220 kPa Inlet Temperature = 30 C Outlet Temperature = 134.7 C Volumetric Flow Rate = 28,000 m3/h STP Power = 950 kW	$1,245,400	1	1.00	1	$1,245,400	1	$1,245,400	2.2	$2,739,880
H-101	Stage 1 AMD Clarifier[S]	Height = 10 m Diameter = 35 m Temperature = 5 C Pressure = 101 kPa Rake Power = 10 kW MOC = Epoxy-lined concrete	$2,500,000	1	1.00	1	$2,500,000	1	$2,500,000	3.2	$8,000,000

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
H-102	Tailings Paste Thickener[V]	Height = 9 m Diameter = 8 m Temperature = 5 C Pressure = 101 kPa Rake Power = 10 kW MOC = Ally 20 Clad Carbon Steel	$750,000	1	4.50	1	$3,375,000	1	$3,375,000	3.2	$10,800,000
H-103	Stage 2 AMD Clarifier[S]	Height = 6 m Diameter = 34 m Temperature = 7 C Pressure = 101 kPa Rake Power = 10 kW MOC = Epoxy-lined concrete	$1,500,000	1	1.00	1	$1,500,000	1	$1,500,000	3.2	$4,800,000
H-201	Leachate Solids Settler[U]	Length = 14 m Diameter = 3.7 m Temperature = 7 C Pressure = 101 kPa MOC = Alloy 20 Clad Carbon Steel	$110,000	1	4.50	1	$495,000	1	$495,000	3.2	$1,584,000
H-202	Leachate Underflow Rotary Vacuum Drum Filter[V]	Area = 140 m2 Pore Size = 50 µm Air Flow Rate = 353 m3/hr (STP) MOC = Alloy 20 Clad Carbon Steel Power = .35 kW	$2,600,000	1	1.00	1	$2,600,000	1	$2,600,000	3.4	$8,840,000
H-203	Neutralized Leachate Solids Settler[U]	Length = 12 m Diameter = 3.7 m Temperature = 7 C Pressure = 101 kPa MOC = Alloy 20 Clad Carbon Steel	$110,000	1	4.50	1	$495,000	1	$495,000	3.2	$1,584,000
H-204	Neutralized Leachate Underflow Rotary Drum Vacuum Filter[V]	Area = 40 m2 Pore Size = 50 µm Air Flow Rate = 135 m3/hr (STP) MOC = Alloy 20 Clad Carbon Steel Power = .20 kW	$1,200,000	1	4.50	1	$5,400,000	1	$5,400,000	3.4	$18,360,000

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
H-401	HREE/MMC Rotary Drum Vacuum Filter[V]	Area = 0.5 m2 Pore Size = 4µ Air Flow Rate = 1.1 m3/hr (STP) MOC = Nickel Clad onto Stainless Steel Power = 1.07 kW	$90,000	1	4.50	1	$405,000	2	$810,000	3.4	$2,754,000
H-402	Zinc Sulfide Rotary Drum Vacuum Filter[V]	Area = 6 m2 Pore Size = 50 µm Air Flow Rate = 18.8 m3/hr (STP) MOC = Nickel Clad onto Stainless Steel Power = .03 kW	$400,000	1	4.50	1	$1,800,000	2	$3,600,000	3.4	$12,240,000
H-403	LREE/MMC Oxalate Rotary Drum Vacuum Filter[V]	Area = 0.1 m2 Pore Size = 4 µm Air Flow Rate = 0.3 m3/hr (STP) MOC = Alloy 20 Stainless Steel Power = 0.25 kW	$35,000	1	4.50	1	$157,500	2	$315,000	3.4	$1,071,000
H-404	Copper Sulfide Rotary Drum Vacuum Filter[V]	Area = 2.4 m2 Pore Size = 50 µm Air Flow Rate = 8.14 m3/hr (STP) MOC = Nickel Clad onto Stainless Steel Power = .08 W	$230,000	1	4.50	1	$1,035,000	2	$2,070,000	3.4	$7,038,000
J-101A/B	Stage 1 Tailings Screw Conveyer[V]	Mass Flow Rate = 5,500 kg/h Length = 30 m Arpon ID = 30 cm Power = 7.5 kW MOC = Stainless Clad Carbon Steel	$85,000	1	2.50	1	$212,500	2	$425,000	3.2	$1,360,000
J-201A/B	Leachate Tailings Screw Conveyer[V]	Mass Flow Rate = 3,100 kg/h Length = 30 m Auger OD = 20 cm Power = 4.5 kW MOC = Alloy 20 Clad Carbon Steel	$60,000	1	4.50	1	$270,000	2	$540,000	3.2	$1,728,000

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
J-202A/B	Neutralized Leachate Tailings Screw Conveyer[V]	Mass Flow Rate = 890 kg/hr Length = 30 m Flight OD = 15 cm Power = 0.75 kW MOC = Stainless Clad Carbon Steel	$36,500	1	2.50	1	$91,250	2	$182,500	3.2	$584,000
J-401A/B	HREE/MMC Screw Oxalate Conveyor[V]	Mass Flow Rate = 7 kg/h Length = 10 m Auger OD = 2.5 cm Power = 0.20 kW MOC = Stainless Clad Carbon Steel	$2,200	1	2.80	1	$6,160	2	$12,320	3.2	$39,424
J-402A/B	Zinc Sulfide Screw Conveyor[V]	Mass Flow Rate = 110 kg/h Length = 30 m Auger OD = 10 Power = 0.20 kW MOC = Stainless Clad Carbon Steel	$23,000	1	2.80	1	$64,400	2	$128,800	3.2	$412,160
J-403A/B	LREE/MMC Screw Conveyor[V]	Mass Flow Rate = 1.6 kg/h Length = 10 m Auger OD = 2.5 cm Power = 0.20 kW MOC = Stainless Clad Carbon Steel	$2,200	1	2.80	1	$6,160	2	$12,320	3.2	$39,424
J-404A/B	Copper Sulfide Screw Conveyor[V]	Mass Flow Rate = 48 kg/h Length = 30 m Auger OD = 10 Power = 0.20 kW MOC = Stainless Clad Carbon Steel	$23,000	1	2.80	1	$64,400	2	$128,800	3.2	$412,160
L-101 A/B	AMD Feed Pump[V]	Power = 78 kW Volumetric Flow Rate = 1,000,000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 220 kPa Temperature = 5 C MOC = Alloy 20 Stainless Steel MOC = Alloy 20 Stainless Steel	$10,335	1	9.00	1	$93,015	2	$186,030	3.2	$595,296

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-102 A/B/C	Sparging Pool Transfer Pump[V]	Power = 15.25 kW Volumetric Flow Rate = 1,000,000 L/hr Inlet Pressure = 202 kPa Outlet Pressure = 220 kPa Temperature = 5 C MOC = Alloy 20 Stainless Steel	$3,730	1	9.00	1	$33,570	3	$100,710	3.2	$322,272
L-103 A/B/C	Stage 1 Clarifier Underflow Pump[V]	Power = .5 kW Volumetric Flow Rate = 330,000 L/hr Inlet Pressure = 202 kPa Outlet Pressure = 202 kPa Temperature = 5 C MOC = Alloy 20 Stainless Steel	$521	1	9.00	1	$4,690	3	$14,070	3.2	$45,025
L-104 A/B	Paste Thickener Overflow Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 320,000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 101 kPa Temperature = 5 C MOC = Alloy 20 Stainless Steel	$411	1	9.00	1	$3,699	2	$7,398	3.2	$23,674
L-105 A/B	Stage 1 Clarifier Overflow Pump[V]	Power = 30 kW Volumetric Flow Rate = 670,000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 220 kPa Temperature = 5 C MOC = Alloy 20 Stainless Steel	$6,489	1	9.00	1	$58,401	2	$116,802	3.2	$373,766
L-106 A/B/C	Paste Thickener Sludge Recirculation Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 15 L/hr Inlet Pressure = 202 kPa Outlet Pressure = 202 kPa Temperature = 20 C MOC = Alloy 20 Stainless Steel	$200	1	9.00	1	$1,804	3	$5,413	3.2	$17,322

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-107 A/B/C	Stage 2 Clarifier Sludge Recirculation Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 34 L/hr Inlet Pressure = 150 kPa Outlet Pressure = 202 kPa Temperature = 6 C MOC = Alloy 20 Stainless Steel	$200	1	9.00	1	$1,804	3	$5,413	3.2	$17,322
L-108 A/B	Stage 2 Clarifier Overflow Pump[V]	Power = 0.5 kW Volumetric Flow Rate = 660,000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 101 kPa Temperature = 6 C MOC = Alloy 20 Stainless Steel	$521	1	9.00	1	$4,690	2	$9,380	3.2	$30,017
L-109 A/B/C	Stage 2 Clarifier Underflow Pump[V]	Power = 0.52 kW Volumetric Flow Rate = 17,000 L/hr Inlet Pressure = 161.6 kPa Outlet Pressure = 202 kPa Temperature = 7 C MOC = Alloy 20 Stainless Steel	$521	1	9.00	1	$4,690	3	$14,070	3.2	$45,025
L-202 A/B/C	Leachate Underflow Pump[V]	Power = 0.5 kW Volumetric Flow Rate = 4,200 L/hr Inlet Pressure = 140 kPa Outlet Pressure = 220 kPa Temperature = 8 C MOC = Alloy 20 Stainless Steel	$521	1	4.50	1	$2,345	3	$7,035	3.2	$22,512

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-203 A/B	Leachate Overflow Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 13,000 L/hr Inlet Pressure = 140 kPa Outlet Pressure = 202 kPa Temperature = 8 C MOC = Alloy 20 Stainless Steel MOC = Alloy 20 Stainless Steel	$200	1	4.50	1	$902	2	$1,804	3.2	$5,774
L-204 A/B	Leachate Filtrate Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 930 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 202 kPa Temperature = 8 C MOC = Alloy 20 Stainless Steel MOC = Alloy 20 Stainless Steel	$200	1	4.50	1	$902	2	$1,804	3.2	$5,774
L-207A/B/C	Neutralized Leachate Underflow Pump[V]	Power = 1 kW Volumetric Flow Rate = 8,400L/hr Inlet Pressure = 140 kPa Outlet Pressure = 220 kPa Temperature = 8 C MOC = 316 stainless steel MOC = 316 stainless steel	$631	1	1.00	1	$631	3	$1,893	3.2	$6,058
L-208 A/B	Neutralized Leachate Filtrate Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 7,700 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 140 kPa Temperature = 8 C MOC = 316 stainless steel MOC = 316 stainless steel	$411	1	1.00	1	$411	2	$822	3.2	$2,630

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-209 A/B	Acid Dosing Vessel Discharge Pump[V]	Power = 0.57 kW Volumetric Flow Rate = 14,000 L/hr Inlet Pressure = 126 kPa Outlet Pressure = 180 kPa Temperature = 8 C MOC = Alloy 20 Stainless Steel MOC = Alloy 20 Stainless Steel	$521	1	9.00	1	$4,690	2	$9,380	3.2	$30,017
L-301 A/B	F-301 Outlet Pump[V]	Power = 1 kW Volumetric Flow Rate = 14000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 190 kPa Temperature = 10 C MOC = Nickel Clad onto Stainless Steel	$631	1	4.50	1	$2,839	2	$5,679	3.2	$18,173
L-302 A/B	F-302 Outlet Pump[V]	Power = 0.55 kW Volumetric Flow Rate = 14000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 101 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$521	1	4.50	1	$2,345	2	$4,690	3.2	$15,008
L-303 A/B	Nitric Acid Dosing Tank Pump[V]	Power = 0.42 kW Volumetric Flow Rate = 14000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 140 kPa Temperature = 10 C MOC = Nickel Clad onto Stainless Steel	$521	1	4.50	1	$2,345	2	$4,690	3.2	$15,008

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Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-304 A/B	Nitric Acid Dosing Tank Pump[V]	Power = 0.55 kW Volumetric Flow Rate = 14000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 101 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel MOC = Nickel Clad onto Stainless Steel	$521	1	4.50	1	$2,345	2	$4,690	3.2	$15,008
L-401 A/B	Oxalate Precipitation Pump[V]	Power = 1.2 kW Volumetric Flow Rate = 15000 L/hr Inlet Pressure = 121 kPa Outlet Pressure = 220 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel MOC = Nickel Clad onto Stainless Steel	$691	1	4.50	1	$3,112	2	$6,223	3.2	$19,914
L-402 A/B	Zinc Recovery Feed Pump[V]	Power = 1.32 kW Volumetric Flow Rate = 13000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 235 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$691	1	4.50	1	$3,112	2	$6,223	3.2	$19,914
L-404 A/B	H-402 Liquid Outlet Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 15000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 115 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$200	1	4.50	1	$902	2	$1,804	3.2	$5,774

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[V] Denotes Quote from Vendor

Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
L-405 A/B	Oxalate Precipitation Pump[V]	Power = 1.2 kW Volumetric Flow Rate = 15000 L/hr Inlet Pressure = 121 kPa Outlet Pressure = 220 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$691	1	4.50	1	$3,112	2	$6,223	3.2	$19,914
L-406 A/B	Zinc Recovery Feed Pump[V]	Power = 1.53kW Volumetric Flow Rate = 15000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 235 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$691	1	4.50	1	$3,112	2	$6,223	3.2	$19,914
L-408 A/B	H-404 Liquid Outlet Pump[V]	Power = 0.20 kW Volumetric Flow Rate = 14000 L/hr Inlet Pressure = 101 kPa Outlet Pressure = 115 kPa Temperature = 12 C MOC = Nickel Clad onto Stainless Steel	$200	1	4.50	1	$902	2	$1,804	3.2	$5,774
R-401	Oxalate Precipitation Reactor[U]	Height = 2.0 m Diameter = 3.0 m Temperature = 12 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$50,000	1.96525	4.50	1	$442,181	1	$442,181	3.2	$1,414,980

[U] Denotes Quote from Ulrich Costing Charts
[A] Denotes Quote from Aspen Plus
[V] Denotes Quote from Vendor

Table 7. Capital Cost Summary Continued

Equipment Tag	Equipment Description	Key Specifications	Unadjusted Unit Equipment Supplier Cost	Basis Date Adjustment Index	MOC A_F	Pressure Adjustment Factor	Adjusted Basis Date Unit Equipment Cost	Number of Units	Line Item Equipment Cost	Estimating Factor, F_T	Total Line Item Cost
R-402	Zinc Precipitation Reactor[U]	Height = 12.7 m Diameter = 1 m Temperature = 12 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$69,000	1.96525	2.80	1	$379,686	1	$379,686	3.2	$1,214,996
R-403	Oxalate Precipitation Reactor[U]	Height = 2.2 m Diameter = 3.0 m Temperature = 12 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$50,000	1.96525	4.50	1	$442,181	1	$442,181	3.2	$1,414,980
R-404	Copper Precipitation Reactor[U]	Height = 12.7 m Diameter = 1 m Temperature = 12 C Pressure = 101 kPa MOC = Nickel Clad onto Stainless Steel	$69,000	1.96525	2.80	1	$379,686	1	$379,686	3.2	$1,214,996

[U] Denotes Quote from Ulrich Costing Charts
[A] Denotes Quote from Aspen Plus
[V] Denotes Quote from Vendor

February 2025
BASIS DATE

TOTAL DIRECT PROCESS COST	$110,000,000
ADDITIONAL DIRECT COSTS	$33,000,000
TOTAL DIRECT COSTS	$140,000,000
INDIRECT COSTS	$36,000,000
FIXED CAPITAL INVESTMENT	$180,000,000
INITIAL CHAGRE OF CHEMICAL & CATALYST	$5,400,000
WORKING CAPITAL	$18,000,000
TOTAL CAPITAL INVESTMENT	$200,000,000

Table 8. Operating Cost Summary

A. Time Since Project Commisioning	B. Operating Labor	C. Maintenance	D. Utilities	E. Chemicals & Catalysts	F. Other Taxes	G. Operating Materials & Lab Charges	H. Other Direct Costs	I. Total Manufacturing Costs (sum of B-H)	J. Raw Materials	K. Total Direct Operating Costs (sum of I & J)	L. Indirect Costs	M. Total Operating Costs (sum of K&L)
1	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
2	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
3	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
4	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
5	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
6	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
7	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
8	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
9	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
10	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
11	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
12	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
13	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
14	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
15	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
16	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
17	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
18	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
19	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
20	$ 7,400,000	$ 14,000,000	$ 950,000	$ 90,000	$ -	$ 1,800,000	$ -	$ 40,000,000	$ 16,000,000	$ 40,000,000	$ 8,600,000	$ 50,000,000
Basis Date	February 2025											

Table 9. Cash Flow Summary

Year	Revenues	Operating Costs	Gross Profit	Depreciation	Federal Taxable Profit	Federal Income Tax	State Taxable Profit	State Income Tax	Nontaxable Charges	Net Profit	Present Value @ HR
-2	-	-	-	-	-	-	-	-	($36)	($36)	($55)
-1	-	-	-	-	-	-	-	-	($72)	($72)	($89)
0	-	-	-	-	-	-	-	-	($92)	($92)	($92)
1	$ 170	($50)	$120	($33)	$90	($19)	$71	($5)	$0	$99	$80
2	$ 170	($50)	$120	($27)	$96	($20)	$76	($5)	$0	$97	$63
3	$ 170	($50)	$120	($22)	$100	($21)	$80	($5)	$0	$96	$50
4	$ 170	($50)	$120	($18)	$100	($22)	$83	($6)	$0	$95	$40
5	$ 170	($50)	$120	($15)	$110	($23)	$85	($6)	$0	$94	$32
6	$ 170	($50)	$120	($12)	$110	($23)	$87	($6)	$0	$94	$26
7	$ 170	($50)	$120	($11)	$110	($23)	$88	($6)	$0	$93	$21
8	$ 170	($50)	$120	($11)	$110	($23)	$88	($6)	$0	$93	$17
9	$ 170	($50)	$120	($11)	$110	($23)	$88	($6)	$0	$93	$13
10	$ 170	($50)	$120	($11)	$110	($23)	$88	($6)	$0	$93	$11
11	$ 170	($50)	$120	($11)	$110	($23)	$88	($6)	$0	$93	$9
12	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$7
13	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$6
14	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$4
15	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$4
16	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$3
17	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$2
18	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$2
19	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$0	$90	$2
20	$ 170	($50)	$120	$0	$120	($26)	$97	($7)	$18	$110	$2

Basis Date for Estimate = February 2025 HR = 24% NPV@HR = $160

CECPI Index 786.1

**** All values are in $ million**

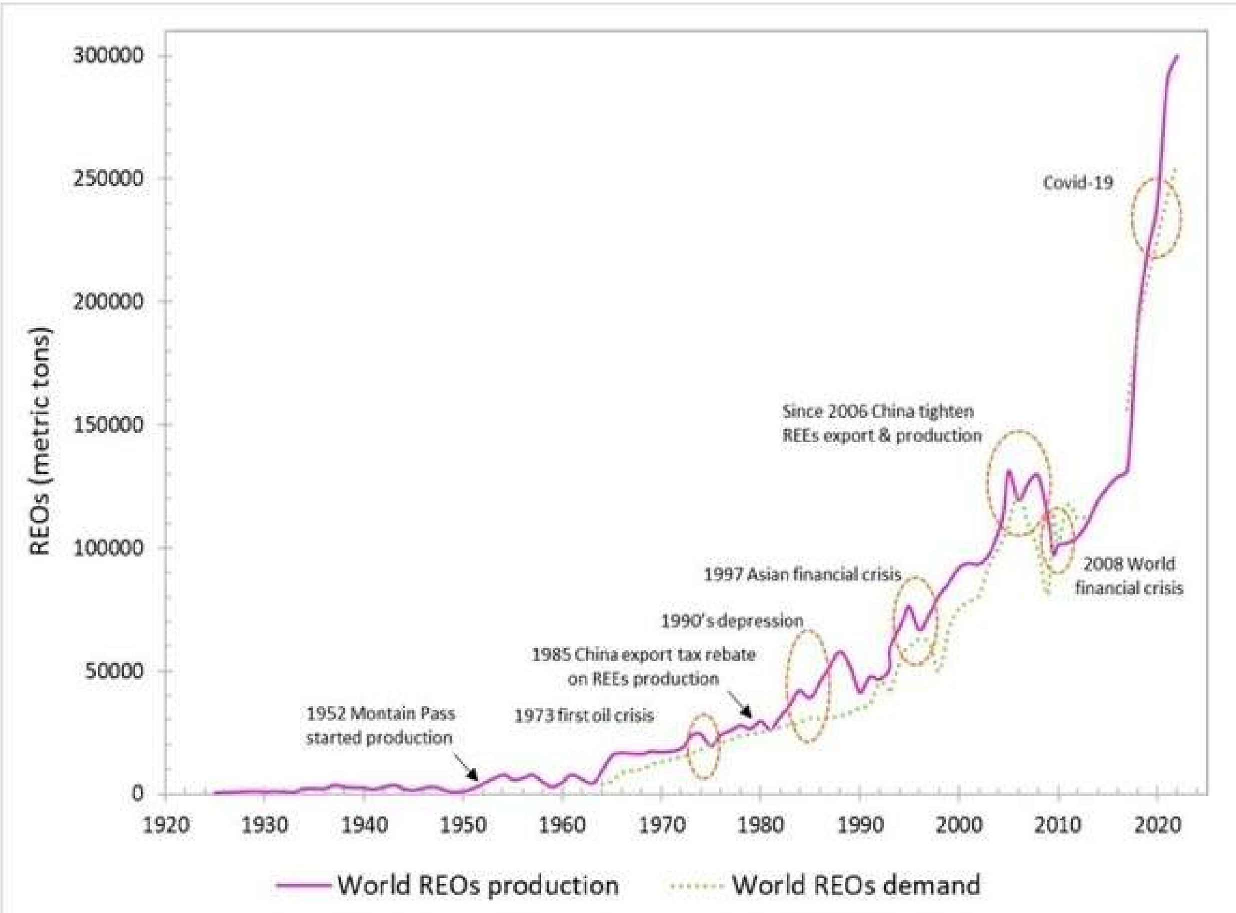


Figure 1. Production and Demand of Rare Earth Oxides[32]

[32] Merroune, A., Ait Brahim, J., Berrada, M., Essakhraoui, M., Achiou, B., Mazouz, H., & Beniazza, R. (2024). A comprehensive review on solvent extraction technologies of rare earth elements from different acidic media: Current challenges and future perspectives. *Journal of Industrial and Engineering Chemistry, 139*, 1-17. https://doi.org/10.1016/j.jiec.2024.04.042

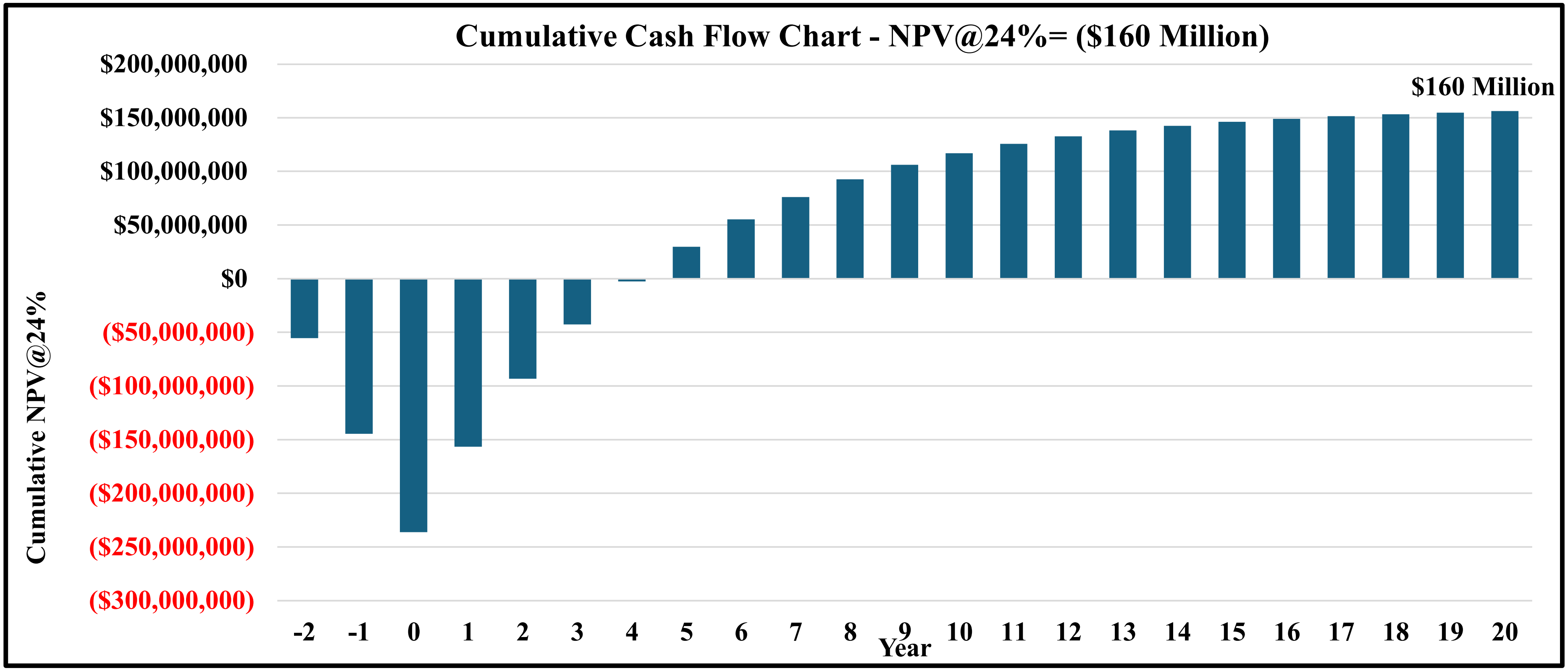


Figure 2. Cumulative Cash Flow Chart for Project REE Recovery from Acid Mine Drainage @ 24% Hurdle Rate

** $0.07 / gal represented as revenue for treating AMD - DCFROR - 37%

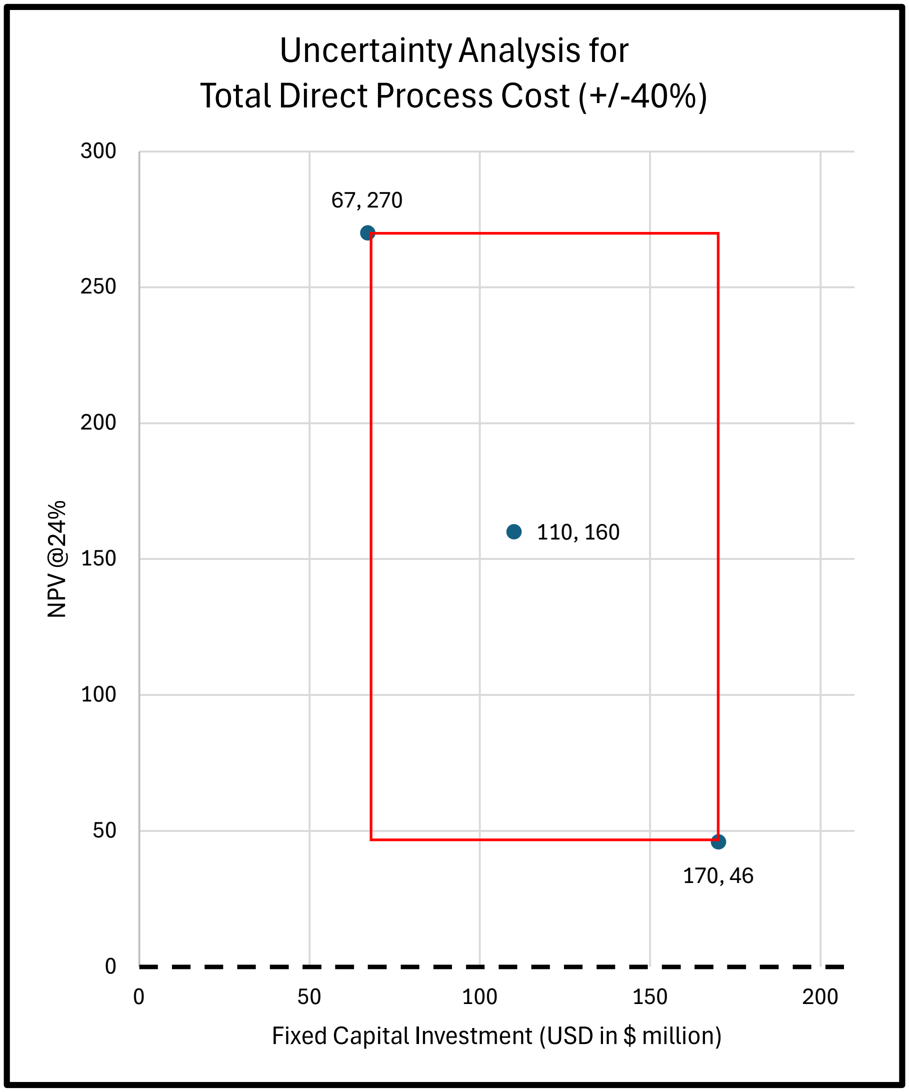


Figure 3. Uncertainty Analysis for Total Direct Process Costs (+/-40%)

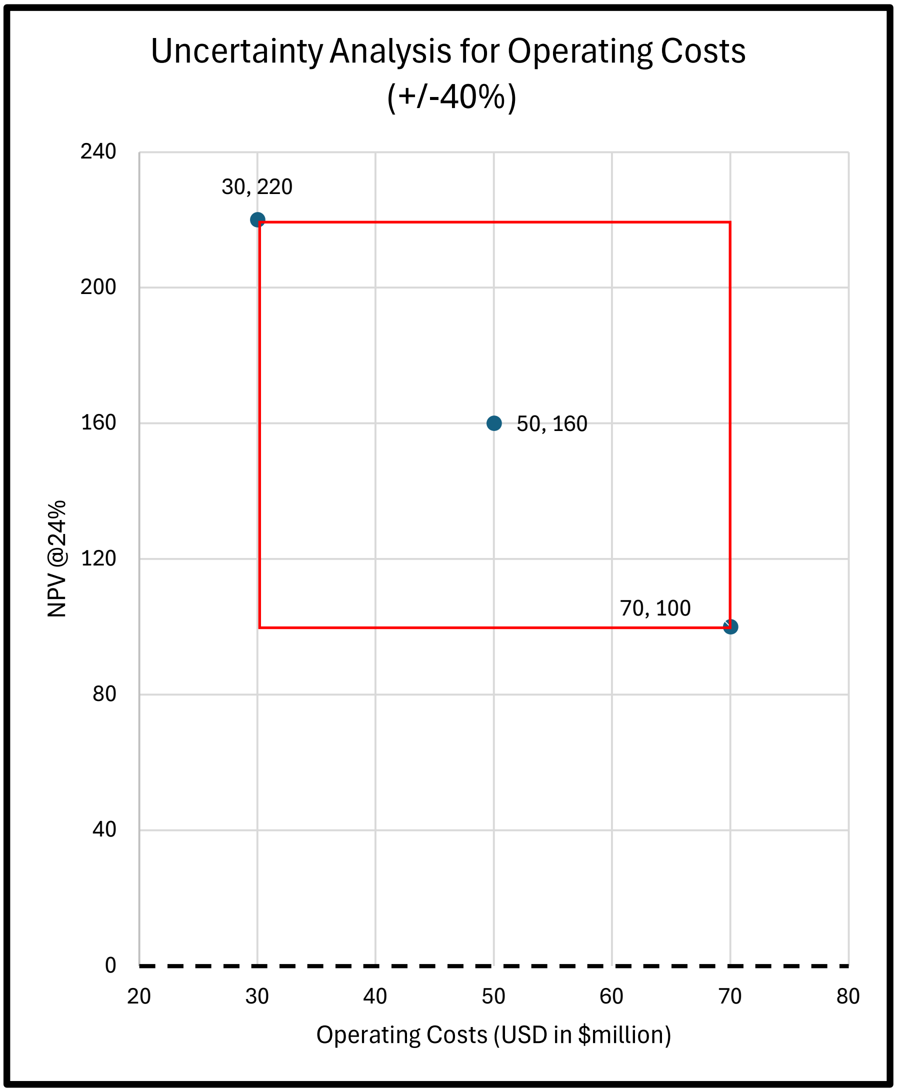


Figure 4. Uncertainty Analysis for Operating Costs (+/-40%)

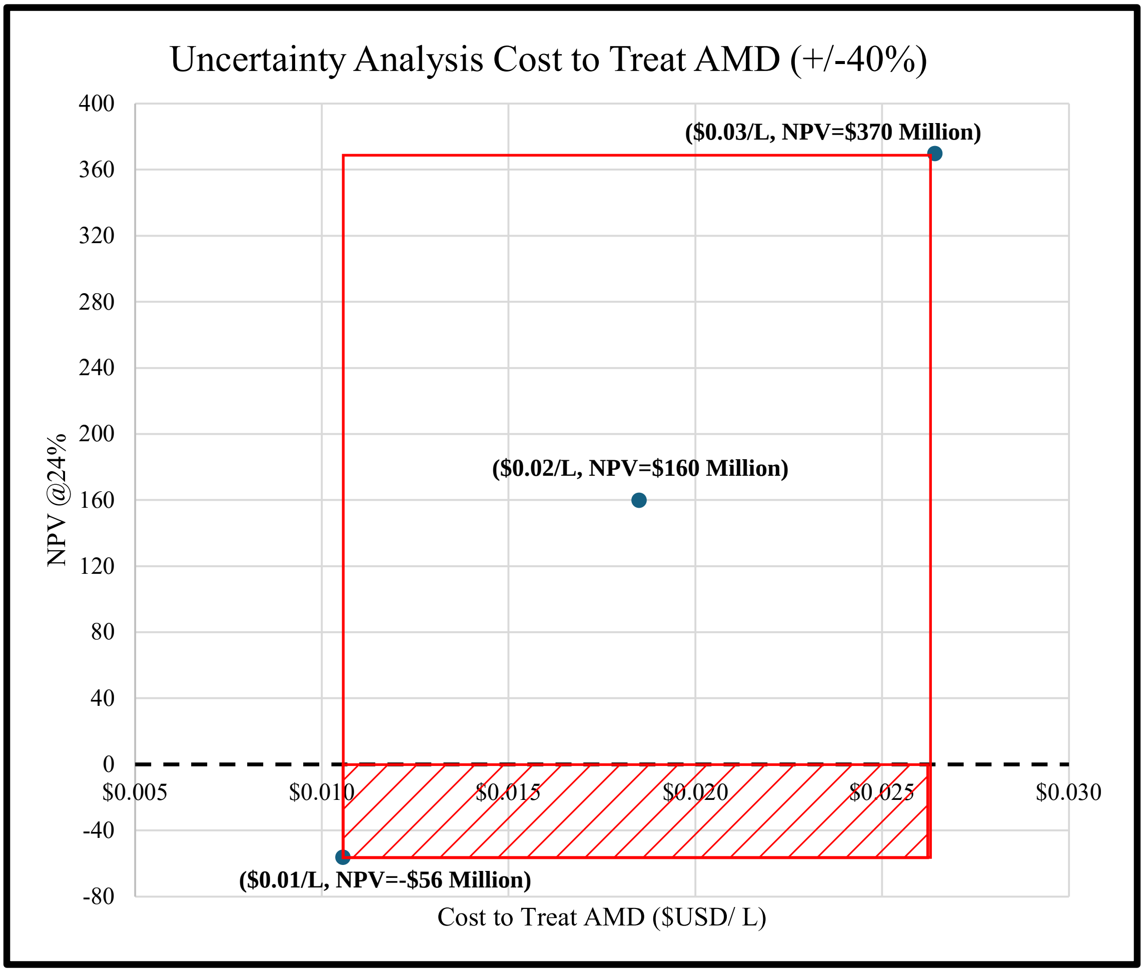


Figure 5. Uncertainty analysis for cost to treat acid mine drainage per L. (13% Risk)

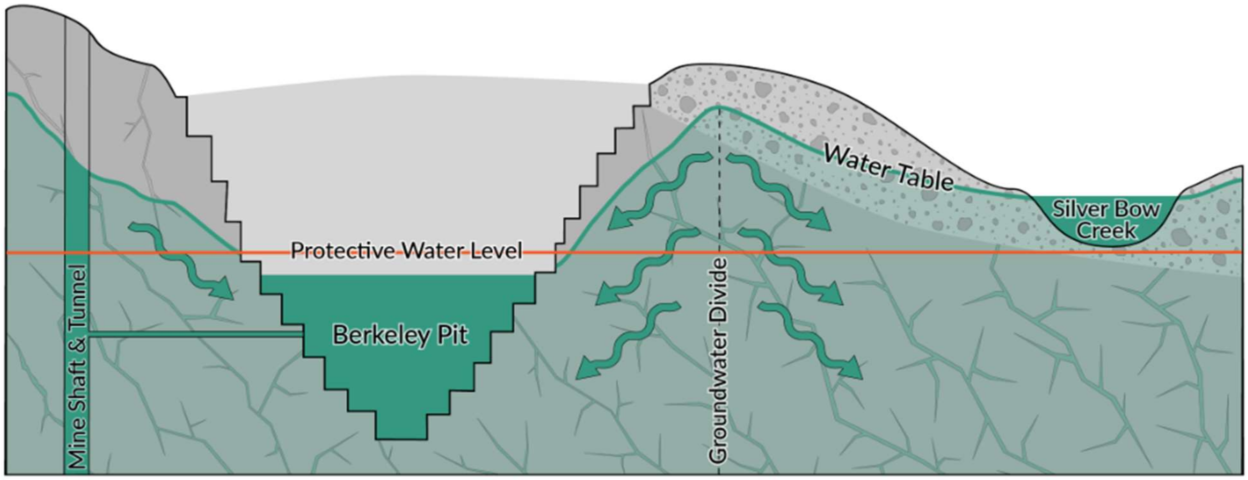
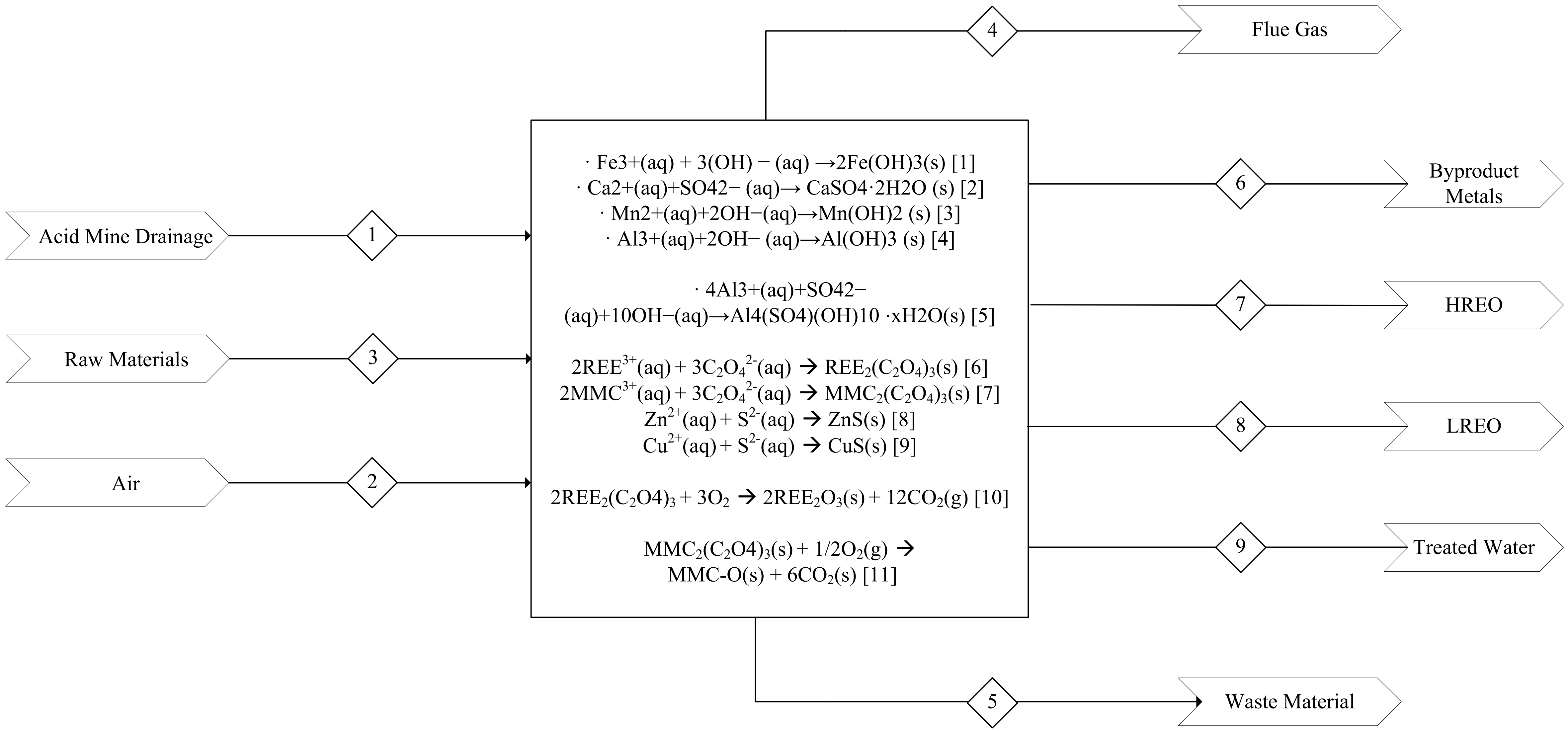


Figure 6: Diagram of Berkley Pit and Water Table[33]

[33] PitWatch. (n.d.). *FAQs*. https://pitwatch.org/faqs/ (Accessed June 4, 2025)



Component/Stream	1	2	3	4	5	6	7	8	9
Acids & Bases	0	0	9800	0	80	0	0	0	0
Gases	56	5800	0	5800	0.071	0	0	0	13
H_2O	990000	0	0	1.9	46000	0	0	0	970000
Fe^{2+} / Fe^{3+}	820	0	0	0	0.011	0	0	0	1.3
Zn^{2+}	630	0	0	0	32	0	0	0	360
Al^{3+}	290	0	0	0	16	0	0	0	96
Mn^{2+}	230	0	0	0	1.4	0	0	0	2
Cu^{2+}	140	0	0	0	15	0	0	0	30
MMC	10000	0	0	0	3700	0	0	0	5400
Tailings	0	0	0	0	5400	0	0	0	14
REE	3.7	0	0	0	1.1	0	0	0	1.1
ZnS	0	0	0	0	0	110	0	0	0
CuS	0	0	0	0	0	48	0	0	0
HREO	0	0	0	0	0	0	3.9	0	0
LREO	0	0	0	0	0	0	0	1.7	0
Total	1000000	5800	9800	5800	56000	160	3.9	1.7	980000

Notes

1. Flow rates are in kg/hr
2. Flow rates are rounded to two significant figures
3. Chemicals in components list consist of NaCO3, HCl, H2SO4, NaOH, Ca(OH)2, Na2S, and Oxalic Acid
4. MMCs in component list consist of Mg2+ , Ca2+, Mg2+, Na+, trace K+, SiO44-, PO43- , Cl- and F- ions
5. Stream 7 is 97wt% rare earth oxide, and 54 wt% heavy rare earth oxide.
6. Stream 8 is 85 wt% rare earth oxide, and 84 wt% light rare earth oxide.

Special Symbols / Nomenclature

Stream Number

Input/Output Information

Reference Drawings

BFD: 00-A-200
PFD: 01-A-300

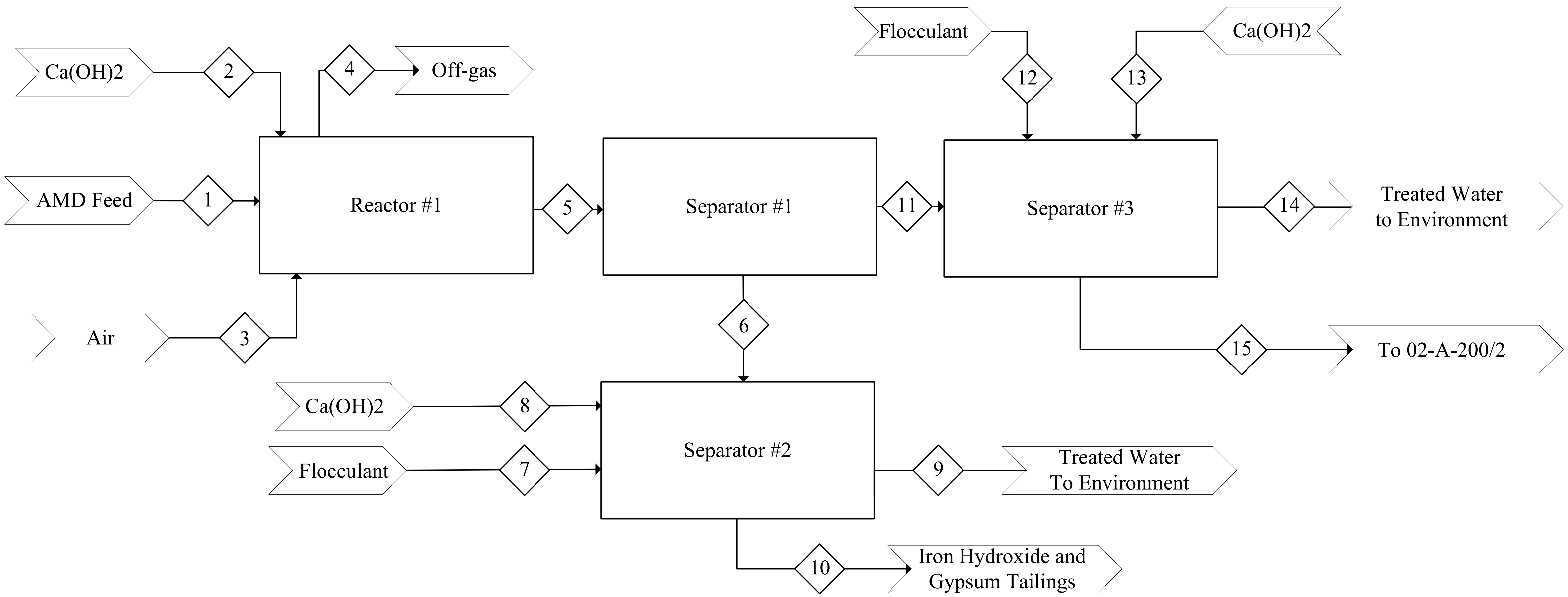
Drawing Type

I/O Diagram

Title

Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	00-A-100/1	0



Component/Stream	1	2	3	4	5	6	7	8	9	10	11	12	13
N_2	0	0	28000	28000	0	0	0	0	0	0	0	0	0
O_2	0	0	8300	8200	13	4.2	0	0	4.1	0.069	8.6	0	0
CO_2	56	0	0	55	0.88	0.29	0	0	0.28	0.0047	0.59	0	0
H_2O	990000	1800	0	0	990000	320000	0	390	320000	2700	670000	0	810
SO_4^{2-}	8600	0	0	0	5800	1900	0	0	1400	12	3900	0	0
Fe^{2+}	400	0	0	0	4	1.3	0	0	1.3	0.011	2.7	0	0
Fe^{3+}	420	0	0	0	0	0	0	0	0	0	0	0	0
Zn^{2+}	630	0	0	0	630	210	0	0	120	1	420	0	0
Cu^{2+}	140	0	0	0	140	47	0	0	10	0.085	95	0	0
MMC	1700	0	0	0	1400	460	0	0	460	7.5	930	0	0
$Ca(OH)_2$	0	1800	0	0	0	180	0	390	0	39	0	0	810
$Zn(OH)_2$	0	0	0	0	0	0	0	0	0.34	130	0	0	0
$Cu(OH)_2$	0	0	0	0	0	0	0	0	0.14	56	0	0	0
Tailings	0	0	0	0	6700	6600	0	0	6.3	2500	12	0	0
REE	3.7	0	0	0	3.7	0.14	0	0	0.048	0.088	3.5	0	0
Flocculant	0	0	0	0	0	0	15	0	0	15	0	30	0
Total	1000000	3700	36000	36000	1000000	330000	15	790	320000	5400	670000	30	1600

Notes

1. Drawing not to scale

2. Discharge in streams 9 and 14 are in line with NPDES standards

3. Reactor 1 – Primary precipitation reactions:
4Fe2+(aq) + O2(aq) + 4H+(aq) →4Fe3+(aq)
Fe3+(aq) + 3(OH) − (aq) →2Fe(OH)3(s)

Reactor 1 pH 2.5 → pH 4.5

4. Separator 1 – Settles solids existing in stream 5. pH = 4.5

5. Separator 2 – Primary precipitation reactions:
Ca2+(aq)+SO42− (aq)→ CaSO4·2H2O (s)
Al3+(aq)+2OH− (aq)→Al(OH)3 (s)

6. Separator 3 –Primary precipitation reactions:
Al3+(aq)+2OH− (aq)→Al(OH)3 (s)
Ca2+(aq)+SO42− (aq)→ CaSO4·2H2O (s)

Special Symbols / Nomenclature

◇ Stream Number

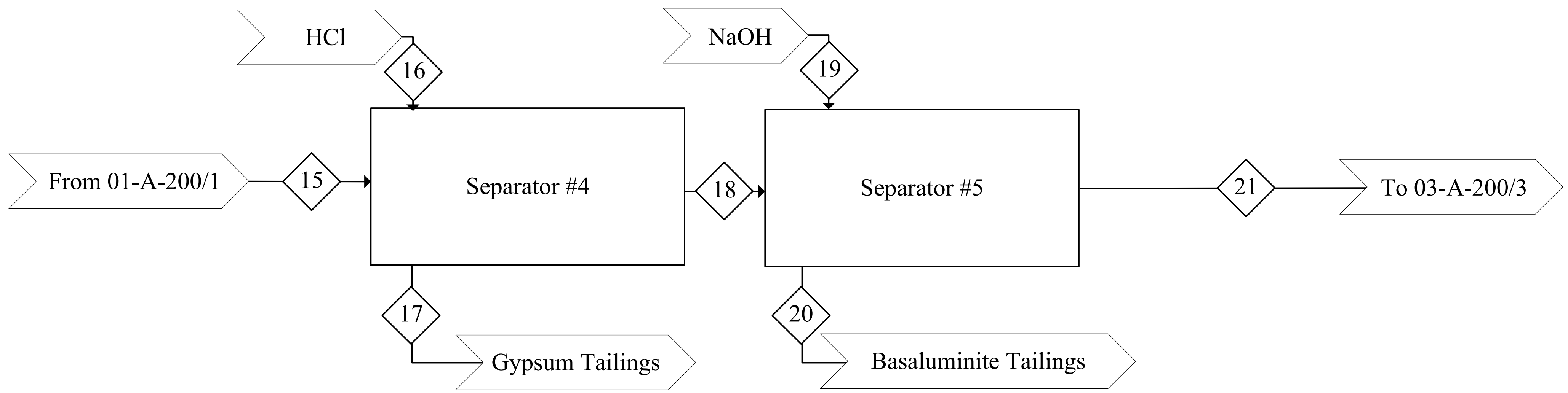
Input/Output Information

Reference Drawings:
I/O 00-A-100
PFD 01-A-300

Drawing Type
Block Flow Diagram

Title
Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
OD 6-10-25	01-A-200/1	0



Component/Stream	15	16	17	18	19	20	21
O2	0.13	0	0.0018	0.126	0	0.00087	0.13
HCl	0	690	0	0	0	0	0
NaOH	0	0	0	0	300	0	0
H2O	0	35	1100	13000	150	88	13000
SO42-	49	0	55	670	0	4.2	610
Zn2+	4	0	8.2	100	0	0.63	93
Cu2+	0.34	0	3.2	39	0	0.26	38
MMC	11	0	34	420	0	3.3	490
Ca(OH)2	81	0	81	0	0	0	0
CaSO4:2H2O	1600	0	1500	0	0	340	0
Fe(OH)3	2.3	0	2.3	0	0	0	0
Al(OH)3	570	0	170	0	0	180	0
Al4(SO4)(OH)10·4H2O	0	0	0	0	0	0.47	0
Mn(OH)2	4.8	0	2.3	0	0	270	0
Zn(OH)2	270	0	110	0	0	10	0
Cu(OH)2	110	0	50	0	0	0.59	0
REE	2.6	0	0.17	2.1	0	0.017	2.1
Flocculant		0	30	0	0	0	0
Total	*2800*	*730*	*3100*	*1400*	*450*	*900*	*1400*

Notes

1. Drawing not to scale

2. Separator 4 – $CaSO_4 \cdot 2H_2O$ (s) → Ca^{2+}(aq)+SO_4^{2-} (aq)

Separator 4 pH = 1

3. Separator 5- Primary precipitation reactions:

Al^{3+}(aq)+2OH^-(aq)→$Al(OH)_3$ (s)

4Al^{3+}(aq)+SO_4^{2-}(aq)+10OH^-(aq)→$Al_4(SO_4)(OH)_{10} \cdot xH_2O$(s) [5]

Separator 5 pH = 4.5

Special Symbols / Nomenclature

Stream Number

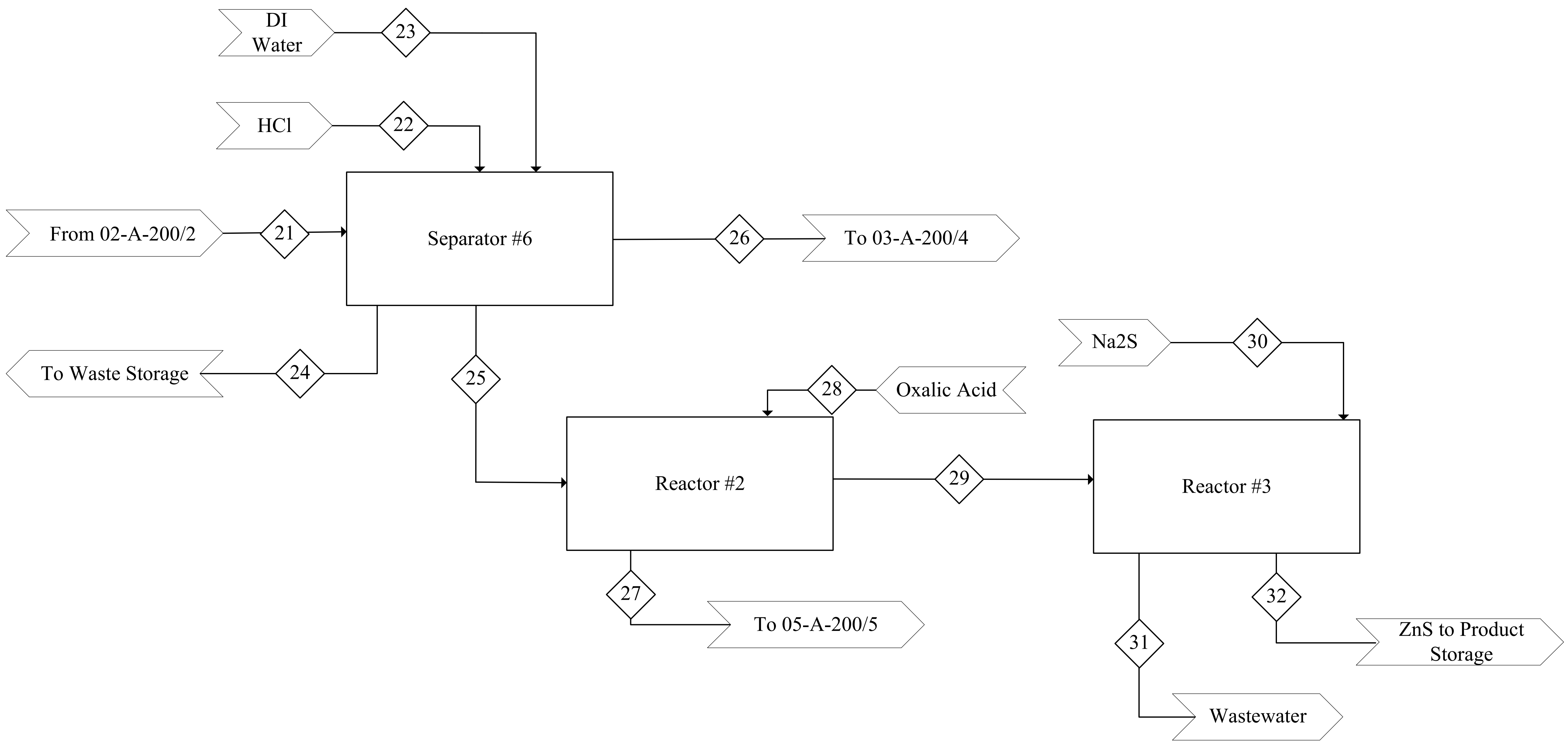
Input/Output Information

Reference Drawings:
I/O 00-A-100
PFD 01-A-300

Drawing Type
Block Flow Diagram

Title
Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
OD/ 06-10-25	02-A-200/2	0



Component / Stream	21	22	23	24	25	26	27	28	29	30	31	32
HCl	0	3000	0	0	0	0	0	0	0	0	0	0
N2S	0	0	0	0	0	0	0	0	0	85	0	0
Oxalic Acid	0	0	0	0	0	0	22	0	15	0	14	0
Zn^{2+}	93	0	0	42	93	2.1	0	0	2	0	0.27	0
Cu^{2+}	38	0	0	26	0.58	37	0	0	37	0	4.9	0
MMC	330	0	0	33	2.8	2.8	0	0	2300	0	2300	0
HREE	1.1	0	0	0.45	1	0.19	0	0	0	0	0	0
LREE	0.98	0	0	0.26	0.56	0.5	0	0	0	0	0	0
ZnS	0	0	0	0	0	0	0	0	0		0	2.5
CuS	0	0	0	0	0	0	0	0	0	0	0	45
SO4	290	0	0	0	290	290	0	0	0	0	290	0
MMCs	0	0	0	0	0	0	0	0	0	0	0	0
REEOx	0	0	0	0	0	0	0	3.3	0	0	0	0
MMCOx	0	0	0	0	0	0	0	0	0	0	0	0
H2O	14000	12000	14000	29000	14000	14000	2.2	0.15	14000	270	13000	0
Total	15,000.00	15,000.00	14,000.00	29,000.00	14,000.00	14,000.00	24	3.50	15,000.00	350.00	14,000.00	48.00

Notes

1. Drawing not to scale
2. Discharge in stream 8 is in line with NPDES standards
3. Separator 6 – Zinc and HREE separation
4. Reactor 2 – Oxalate precipitation pH 1.5
5. Reactor 3 – Zinc sulfide precipitation, 80% conversion

Special Symbols / Nomenclature

Stream Number

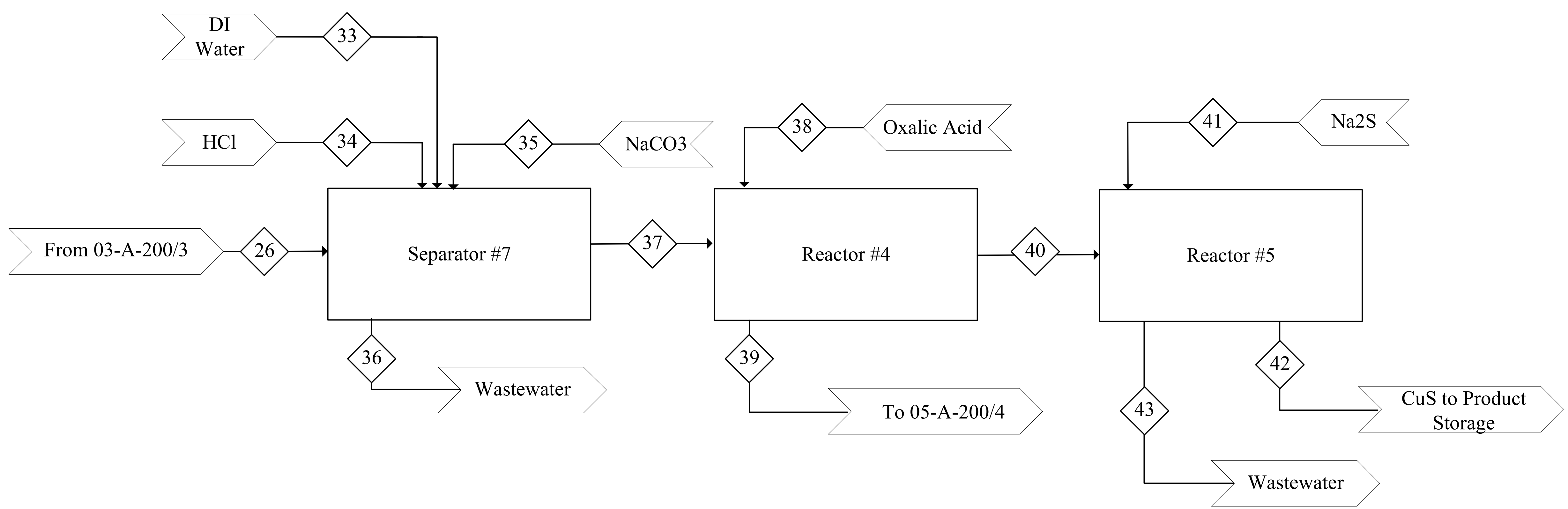
Input/Output Information

Reference Drawings:
I/O 00-A-100
PFD 01-A-300

Drawing Type
Block Flow Diagram

Title
Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
OD 6-10-25	03-A-200/3	0



Components/Stream	25	32	33	34	35	36	37	38	39	40
H2O	13000	14000	13000	14000	2.2	0.16	14000	0	0	13000
SO42-	290	1300	300	1300	0	0	1300	0	0	1200
N2S	0	0	0	0	0	0	0	260	0	70
Oxalic Acid	0	0	0	0	22	0	15	0	0	14
Zn	9.2	0	7.4	2.1	0	0	2	0	0	0.27
Cu	35	0	1.8	37	0	0	37	0	0	4.9
MMC	330	0	480	2.8	0	0	0.63	0	0	0.53
REE	0.63	0	0.02	0.69	0	0	0.14	0	0	0.13
ZnS	0	0	0	0	0	0	0	0	2.5	0
CuS	0	0	0	0	0	0	0	0	45	0
MMCS	0	0	0	0	0	0	0	0	0.01	0
REEOx	0	0	0	0	0	1.1	0	0	0	0
MMCOx	0	0	0	0	0	0.36	0	0	0	0
Total	14000	15000	14000	15000	24	1.6	15000	260	48	14000

Notes

1. Drawing not to scale
2. Discharge in stream 8 is in line with NPDES standards
3. Reactor 4 – Copper and LREE separation
4. Separator 7 – Oxalate precipitation 80% conversion
5. Reactor 5 – Copper sulfide 80% conversion

Special Symbols / Nomenclature

Stream Number

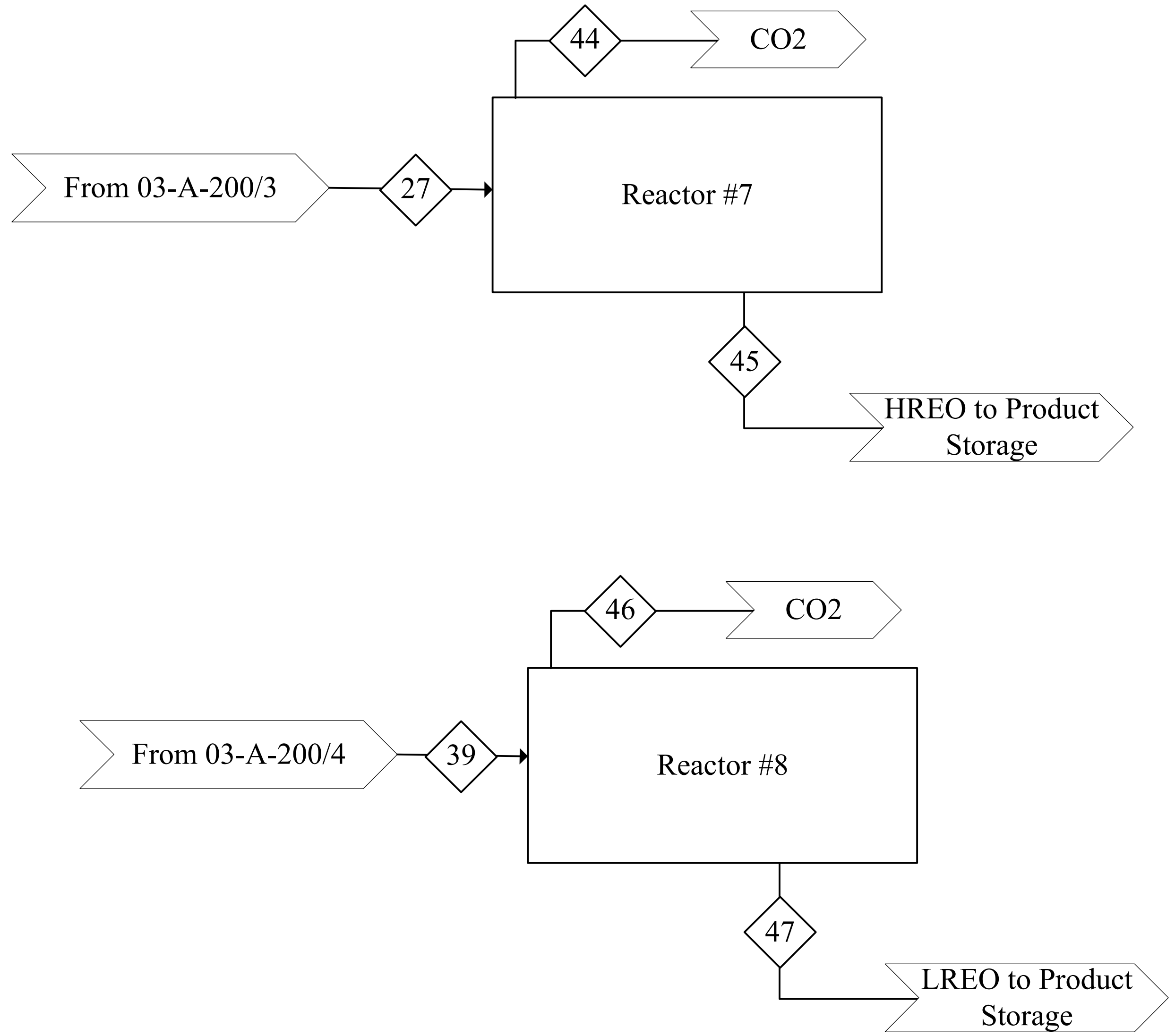
Input/Output Information

Reference Drawings:
I/O 00-A-100
PFD 01-A-300

Drawing Type
Block Flow Diagram

Title
Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
OD 06-10-25	03-A-200/4	0



Components/Stream	26	41	42	36	43	44
H2O	0.71	0	0	0.16	0	0
REOx	6	0	0	0	0	0
MMCOx	0.29	0	0	0	0	0
REO	0	0	3.2	1.1	0	0.64
MMCO	0	0	0.09	0.36	0	0.11
Total	7	0	3.3	1.6	0	0.75

Notes

1. Drawing not to scale
2. Reactor 6 – Heavy rare earth oxide formation. 800C
3. Reactor 7 – Light rare earth oxide formation. 800C
4. Chemical equation for both reactors:

$$REE_2(C_2O_4)_3 = REE_2O_3 + 6CO_2$$

5. Conversion is assumed to be 100%

Special Symbols / Nomenclature

Stream Number

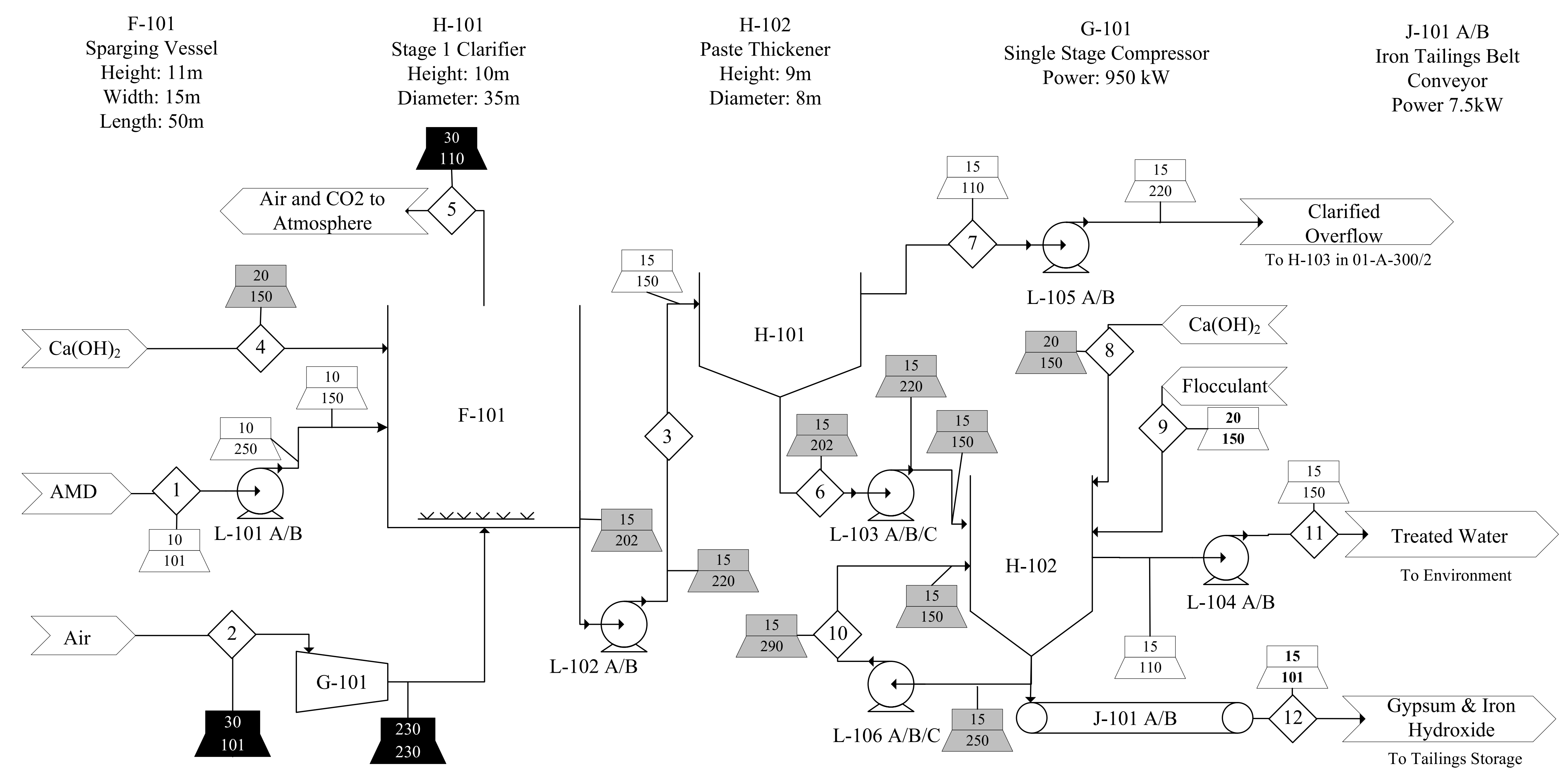
Input/Output Information

Reference Drawings:
I/O 00-A-100
PFD 01-A-300

Drawing Type
Block Flow Diagram

Title
Ion Exchange Separation of REE from AMD

By/Date	DWG #/ Sheet #	Rev
OD 6-10-25	05-A-200/5	0



Component / Stream	1	2	3	4	5	6	7	8	9	10	11	12
N_2	0	28000	0	0	28000	0	0	0	0	0	0	0
O_2	0	8300	13	0	8200	4.2	8.6	0	0	0.001	4.1	0.068
CO_2	56	0	0.88	0	55	0.29	0.59	0	0	0.000069	0.28	0.0046
H_2O	990000	0	990000	1800	0	320000	670000	390	0	40	310000	9100
SO_4^{2-}	8600	0	5800	0	0	1900	3900	0	0	0.181	1400	12
Fe2+	400	0	4	0	0	1.3	2.7	0	0	0	1.3	0.011
Fe3+	420	0	0	0	0	0	0	0	0	0	0	0
Zn2+	630	0	630	0	0	210	420	0	0	0.015	120	0.99
Cu2+	140	0	140	0	0	47	95	0	0	0.0013	10	0.084
MMC	1700	0	1400	0	0	460	930	0	0	0.11	460	7.4
$Ca(OH)_2$	0	0	0	1800	0	180	0	390	0	0.59	0	220
$Zn(OH)_2$	0	0	0	0	0	0	0	0	0	2	0.34	130
$Cu(OH)_2$	0	0	0	0	0	0	0	0	0	0.85	0.14	55
Tailings	0	0	6700	0	0	6600	12	0	0	37	6.3	9100
REE	3.7	0	3.7	0	0	0.14	3.5	0	0	0.0013	0.048	0.23
Flocculant	0	0	0	0	0	0	0	0	15	0.22	0	15
Total	1,000,000	36,000	1,000,000	3,700	36,000	330,000	670,000	790	15	80	320,000	19,000

L-101 A/B
AMD Feed Pump
Power 65 kW

L-102 A/B
Aerated AMD Transfer Pump
Power 7.9 kW

L-103 A/B
Stage 1 Underflow Pump
Power 2.6 kW

L-104 A/B
Stage 1 Water Pump
Power 5.6 kW

L-105 A/B
Overflow Transfer Pump
Power 32 kW

L-106 A/B
Stage 1 Sludge Recirculation Pump
Power: 0.3 kW

Notes

1. AMD enters at pH ~2.5
2. The discharge of F-101 maintained at pH~4.5
3. Pumps have an assumed efficiency of 70%
4. G-101 has a polytropic efficiency of 80%
6. MMC consists of Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+} and trace $SiO4^{4-}$, F^-, Cl^-, K^+, $PO4^{3-}$.
7. Stream tables are in kg/h
8. Streams may not balance due to rounding
9. Drawing not to scale
10. Discharge in stream 16 is discharged to surface water or an intermediate settling pond per NPDES permit requirements

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX XXX Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type

Process Flow Diagram

Title

Staged Neutralization of AMD

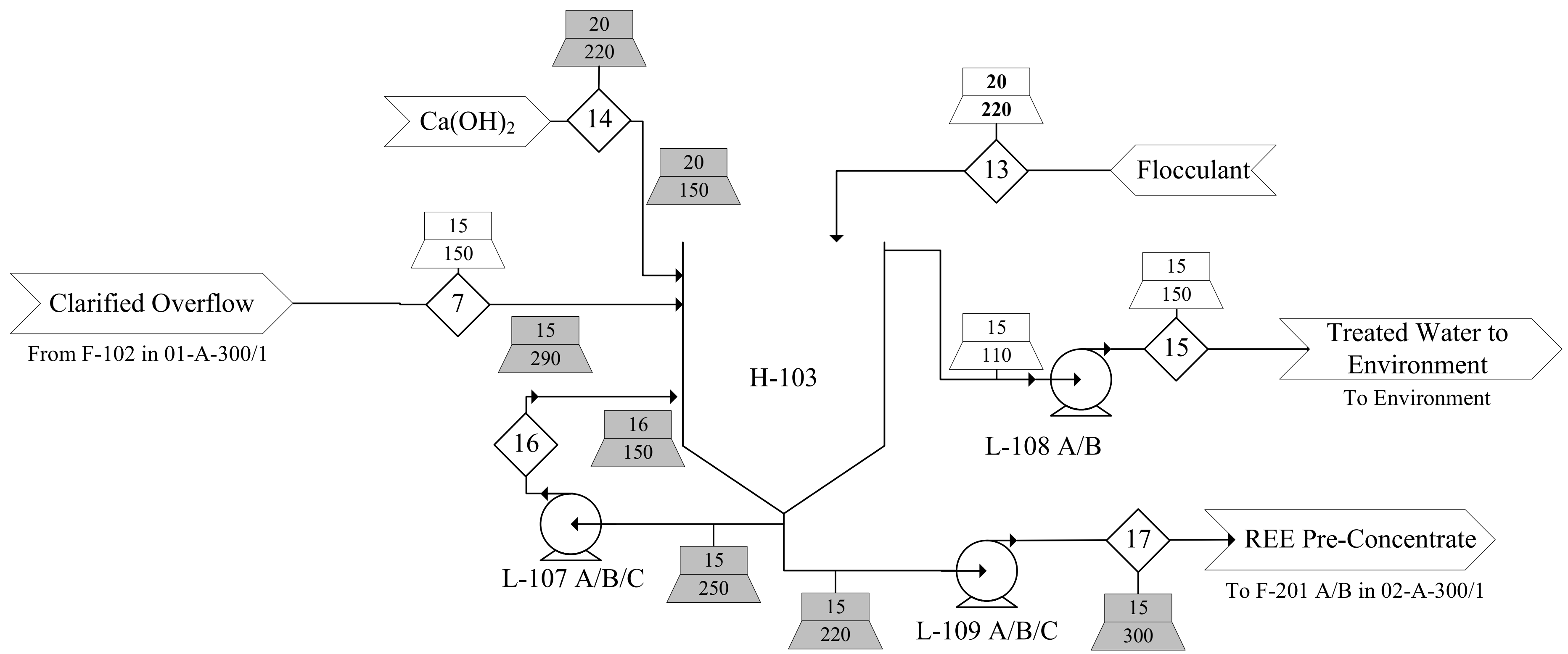
By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	01-A-300/1	0

H-103
Stage 2 Clarifier
Height: 6m
Diameter: 34m

L-107 A/B/C
Stage 2 Sludge Recirculation Pump
Power 0.27 kW

L-108 A/B
Stage 2 Water Pump
Power 12 kW

L-109 A/B/C
Preconcentrate Pump
Power 1.2 kW



Component / **Stream**	7	13	14	15	16	17
O_2	8.6	0	0	8.5	0.00062	0.13
CO_2	0.59	0	0	0	0	0
H_2O	660000	0	810	650000	51	13000
SO_4^{2-}	3900	0	0	2900	0.23	49
Fe2+	2.7	0	0	2.7	0.00019	0.041
Zn2+	420	0	0	240	0.019	4
Cu2+	95	0	0	20	0.0016	0.34
MMC	930	0	0	660	0.053	11
$Ca(OH)_2$	0	0	810	0	0.38	81
$Zn(OH)_2$	0	0	0	0.69	1.3	270
$Cu(OH)_2$	0	0	0	0.29	0.55	110
Tailings	12	0	0	5.7	11	2200
REE	3.5	0	0	1.1	0.012	2.5
Flocculant	0	30	0	0	0.15	30
Total	670000	30	1600	660000	66	16000

Notes

1. H-103 is an upflow clarifier maintained at a pH of 7.
2. Discharge in stream 16 is discharged to surface water or an intermediate settling pond per NPDES permit requirements.
3. MMC consists of Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+} and trace Fe^{2+}, $SiO4^{4-}$, F^-, Cl^-, K^+, $PO4^{3-}$.
4. Streams may not balance due to rounding
5. Stream tables are in kg/h
6. Pumps have an assumed efficiency of 70%
7. Drawing not to scale

Special Symbols / Nomenclature

Input/Output

Stream Number

Temperature °C (gas)
Pressure kPa (gas)

XX
XXX
Temperature °C (liquid)
Pressure kPa (liquid)

XX
XXX
Temperature °C (solid)
Pressure kPa (solid)

Temperature °C (mixed)
Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type

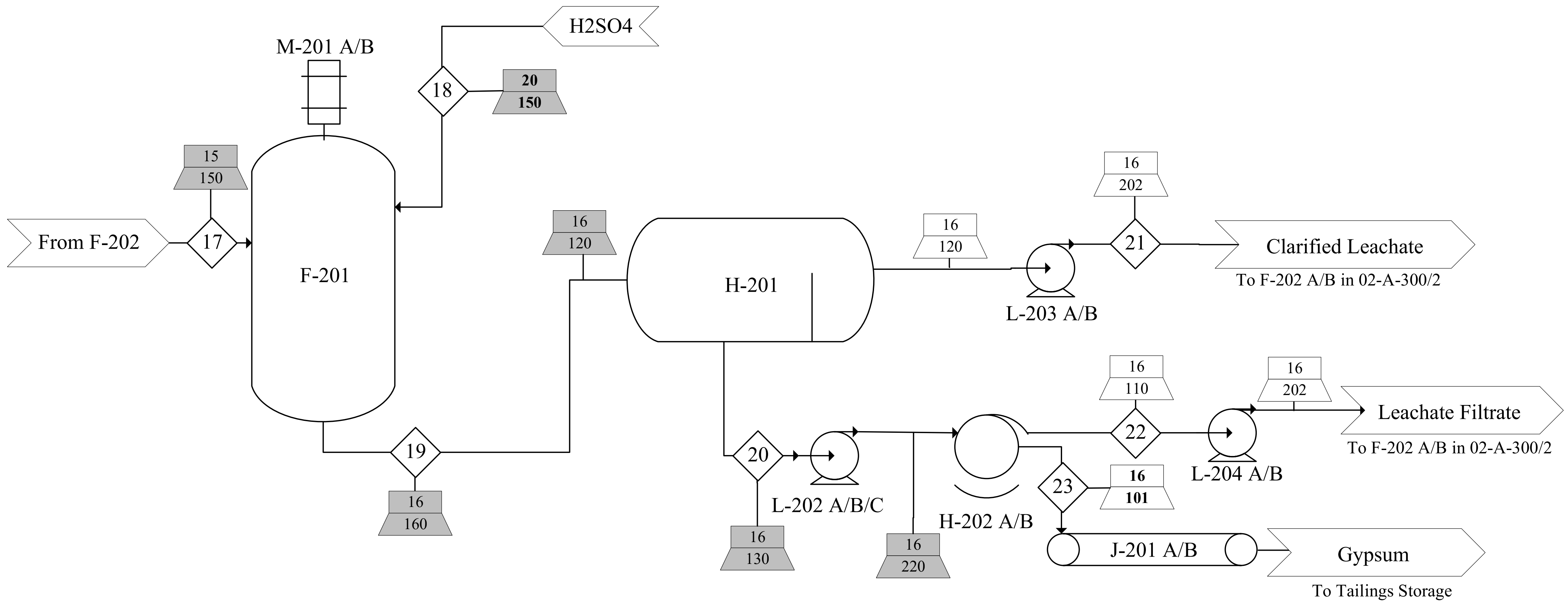
Process Flow Diagram

Title

Staged Neutralization of AMD

By/Date	**DWG #/ Sheet #**	**Rev**
CEC/ 05-08-25	01-A-300/2	0

F-201	H-201	H-202 A/B	L-202 A/B/C	L-203 A/B	L-204 A/B	J-201 A/B	M-301 A/B
Acid Leaching Vessel Height: 3.8m Diameter: 3.4m	Gypsum Settler Length: 14m Diameter: 3.7m	Gypsum Rotary Vacuum Drum Filter Power: 83kW Area: 140 m^2	Gypsum Slurry Pump Power: 0.2 kW	Leachate Overflow Pump Power: 0.47 kW	Gypsum Filtrate Pump Power: 0.2 kW	Gypsum Tailings Belt Conveyor Power: 4.5kW	Acid Leaching Agitation Motor 25 kW

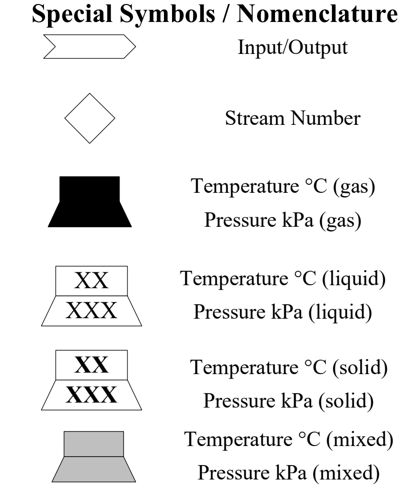


Component / Stream	17	18	19	20	21	22	23
O2	0.13	0	0.13	0.018	0.11	0.016	0.0018
H2SO4	0	690	0	0	0	0	0
H2O	14000	35	14000	1900	12000	850	1100
$SO4^{2-}$	49	0	720	99	630	44.5	555
Zn^{2+}	4	0	110	15	9.3	6.5	8.2
Cu^{2+}	0.34	0	42	5.7	36	2.5	3.2
MMC	11	0	450	62	390	27	34
Ca(OH)2	81	0	81	81	0	0	0
CaSO4·2H2O	1600	0	1500	1500	0	0	1500
Fe(OH)3	2.3	0	2.3	2.3	0	0	2.3
Al(OH)3	570	0	170	170	0	0	170
Mn(OH)2	4.8	0	2.3	2.3	0	0	2.3
Zn(OH)2	270	0	0	0	0	0	170
Cu(OH)2	110	0	50	50	0	0	50
REE	2.6	0	2.6	0.31	2	0.14	0.17
Flocculant	30	0	30	30	0	0	30
Total	16,000	730	17,000	4,000	13,000	930	3,100

Notes

1. Drawing not to scale.
2. Streams may not balance due to rounding.
3. Stream tables are in kg/h.
4. Pumps have an assumed efficiency of 70%.
5. MMC consists of Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+} and trace F^- Cl^{-}.

Special Symbols / Nomenclature



Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type

Process Flow Diagram

Title

REE Preconcentrate Acid Leaching

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	02-A-300/1	0

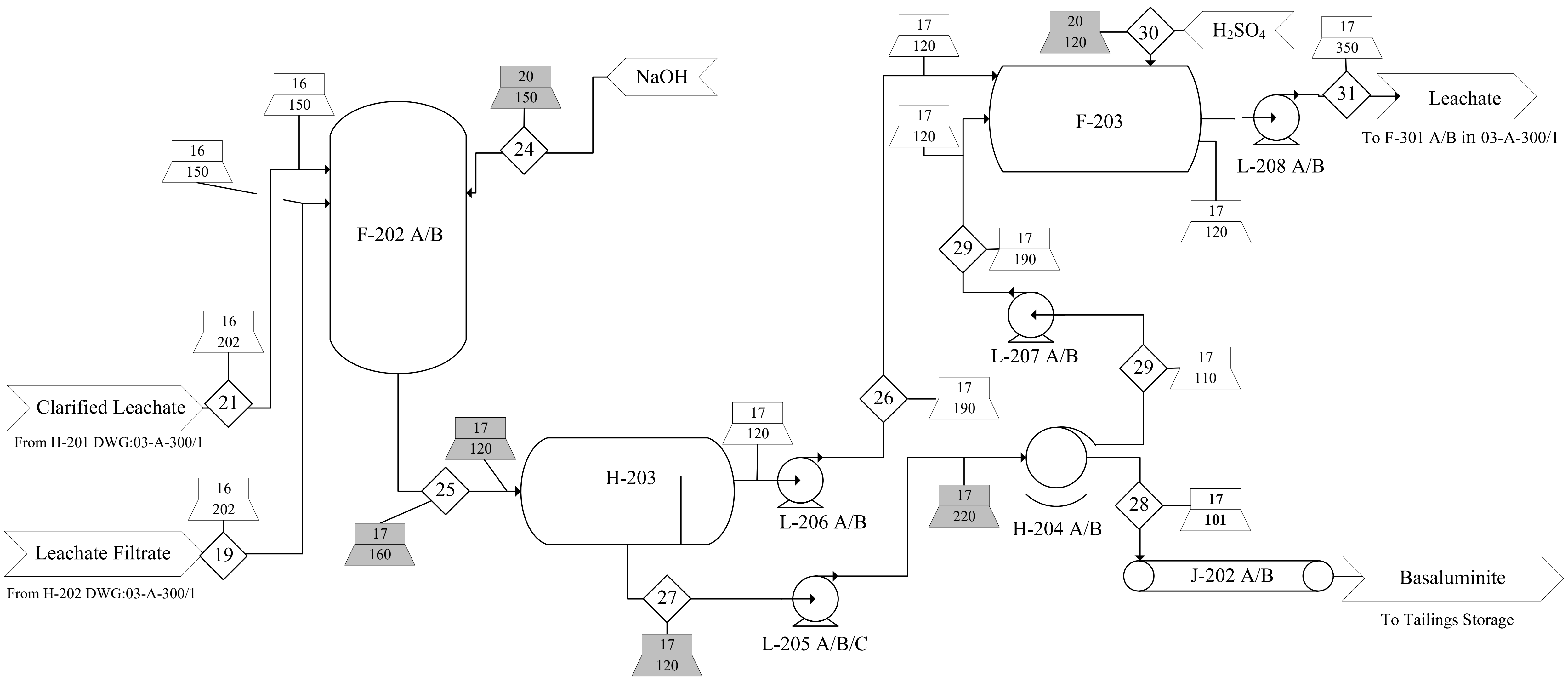
H-203
Basaluminite Settler
Length: 12m
Diameter: 3.7m

F-202 A/B
Neutralization Vessel
Height: 2.1m
Diameter: 2.3m

F-203
Leachate pH Control Tank
Height: 2.8m
Diameter: 2.6m

H-204 A/B
Basaluminite Rotary Vacuum Drum Filter
Power: 83 kW
Area: 140 m^2

J-202 A/B
Basaluminite Belt Conveyor
Power: 0.75kW



Component / **Stream**	**21**	**22**	**24**	**25**	**26**	**27**	**28**	**29**	**30**	**31**
O2	0.11	0.016	0	0.13	0.07	0.058	0.00087	0.07	0	0.13
H2SO4	0	0	0	0	0	0	0	0	0.95	0
NaOH	12000	0	0	0	0	0	0	0	0	0
H2O	630	850	150	13000	7200	5900	88	7100	0.05	13000
SO4²⁻	93	45	0	610	340	280	4.2	330	0	610
Zn²⁺	0.62	6.5	0	82	51	42	0.63	51	93	38
Cu²⁺	0.86	2.5	0.6	38	21	17	0.26	21	0	3.2
MMC	430	27	0	490	270	220	3.3	270	0	490
CaSO4·2H2O	0	0	0	340	340	340	0	0	0	0
Fe(OH)3	0	0	0	0	0	0	0	0	0	0
Al(OH)3	0	0	0	180	180	180	0	0	0	0
Mn(OH)2	0	0	0.47	0.47	0.47	0.47	0.347	0	0	0
Al4(SO4)(OH)10·4H2O	0	0	0	270	270	270	0	0	0	0
Zn(OH)2	0	0	0	1	1	1	0	0	0	0
Cu(OH)2	0.003	0.00021	0	0.59	0.59	0.59	0.59	0.59	0	0
REE	2	0.14	2.1	2.1	1.4	0.74	0.017	1.3	0	2.1
Total	13,000	930	450	15,000	8,700	6,500	900	7,800	1	14,000

L-205 A/B
Basaluminite Slurry Pump
Power: 0.26 kW

L-206 A/B
Neutralized Leachate Overflow Pump
Power: 0.27 kW

L-207 A/B/C
Basaluminite Filtrate Pump
Power: 0.20 kW

L-208 A/B
IX Feed Pump
Power: 0.56 kW

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Pumps have an assumed efficiency of 70%
5. MMC consists of Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+} and trace F^- Cl^{-}.

Special Symbols / Nomenclature

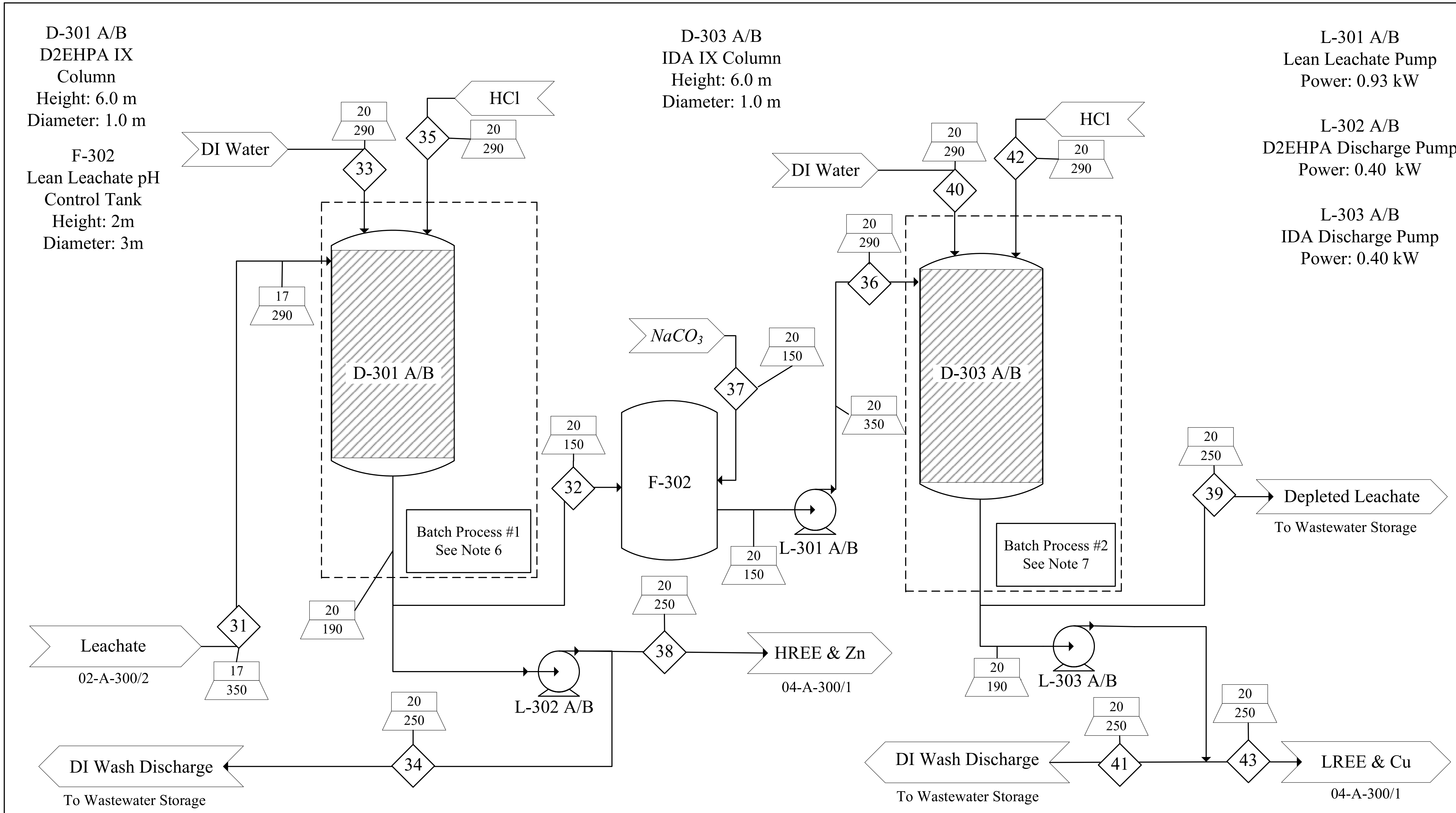
Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX **XXX** Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:
I/O Diagram
Block Flow Diagram

Drawing Type
Process Flow Diagram

Title
REE Preconcentrate Acid Leaching

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	02-A-300/2	0



Component / **Stream**	31	32	33	34	35	36	37	38	39	40	41	42	43
HCl	0	0	0	0	3000	0	0	370	0	0	0	3000	0
NaCO3	0	0	0	0	0	0	38	0	0	0	0	0	0
Zn^{2+}	93	93	0	42	0	93	0	93	7.4	0	0.93	0	2.1
Cu^{2+}	38	35	0	26	0	0.58	0	0.58	1.8	17	0	0	37
CO_3^-	0	0	0	0	0	22	0	0	20	0	22	0	0
MMC	330	300	0	33	0	300	0	2.8	270	0	31	0	2.8
HREE	1.1	0.17	0	0.45	0	1	0	1	0.0052	0	0.084	0	0.19
LREE	0.98	0.46	0	0.26	0	0.58	0	0.56	0.014	0	0.22	0	0.5
SO4	290	290	0	0	290	290	0	0	290	0	0		0
Cl^-	0	0	0	0	0	0	0	0	0	0	0	0	0
Tailings	140	70	0	70	0	70	0	0	35	0	35	0	0
H2O	13000	13000	14000	14000	12000	14000	110	14000	13000	14000	14000	12000	14000
Total	14,000	14,000	14,000	14,000	15,000	15,000	150	14,000	14,000	14,000	14,000	15,000	14,000

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
5. MMC consists of Cl-, Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+}
6. For batch steps of F-301 see 03-A-300/2
7. For batch steps of F-303 see 03-A-300/3
8. Na+ from stream 37 grouped in MMC.
9. Stream 31 contains 1% suspended tailings. Assume 50% of this stays in resin bed and that all removed by wash.
10. 10% residual liquid in beds after adsorption cycle. 100% removed in wash cycles..

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX XXX Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

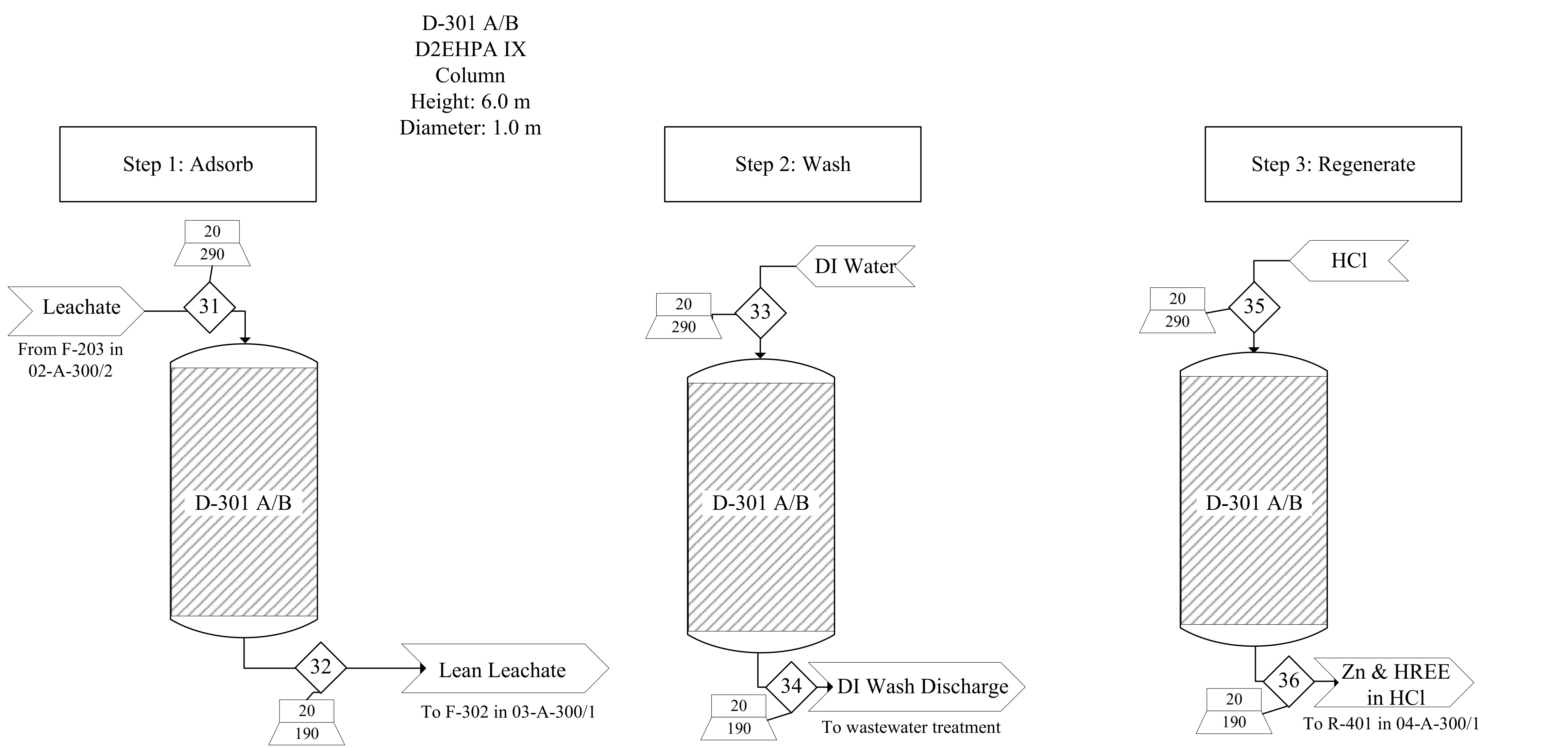
Drawing Type

Process Flow Diagram

Title

Concentration of REEs via Ion Exchange

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	03-A-300/1	0



Component / **Stream**	**31**	**32**
HCl	0	0
NaCO3	0	0
Zn^{2+}	93	93
Cu^{2+}	38	35
$CO3^-$	0	0
MMC	330	300
HREE	1.1	0.17
LREE	0.98	0.46
SO4	290	290
Cl^-	0	0
Tailings	140	70
H2O	13000	13000
Total	14,000	14,000

Component / **Stream**	**33**	**34**
HCl	0	0
NaCO3	0	0
Zn^{2+}	0	42
Cu^{2+}	0	26
$CO3^-$	0	0
MMC	0	33
HREE	0	0.45
LREE	0	0.26
SO4	0	0
Cl^-	0	0
Tailings	0	70
H2O	14000	14000
Total	14,000	14,000

Component / **Stream**	**35**	**36**
HCl	3000	0
NaCO3	0	0
Zn^{2+}	0	93
Cu^{2+}	0	0.58
$CO3^-$	0	22
MMC	0	300
HREE	0	1
LREE	0	0.58
SO4	290	290
Cl^-	0	0
Tailings	0	70
H2O	12000	14000
Total	15,000	15,000

Step	**Event**	**Termination Criteria**
1	Adsorption under leachate feed.	$C_{Zn,OUT} > C_{Zn,MIN}$
2	Wash under DI water feed.	$C_{TSS,OUT} < C_{TSS,MAX}$
3	Regeneration under HCl feed.	$C_{Zn,OUT} = 0$
4	Stand-by	$C_{Zn,OUT} > C_{Zn,MIN}$

*The stand-by termination condition is that duplicate column has completed its adsorption cycle.

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
5. MMC consists of Cl-, Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+}
6. For batch steps of F-301 see 03-A-300/2
7. For batch steps of F-303 see 03-A-300/3
8. Na+ from stream 37 grouped in MMC.
9. Stream 31 contains 1% suspended tailings. Assume 50% of this stays in resin bed and that all removed by wash.
10. 10% residual liquid in beds after adsorption cycle. 100% removed in wash cycles..

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX **XXX** Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

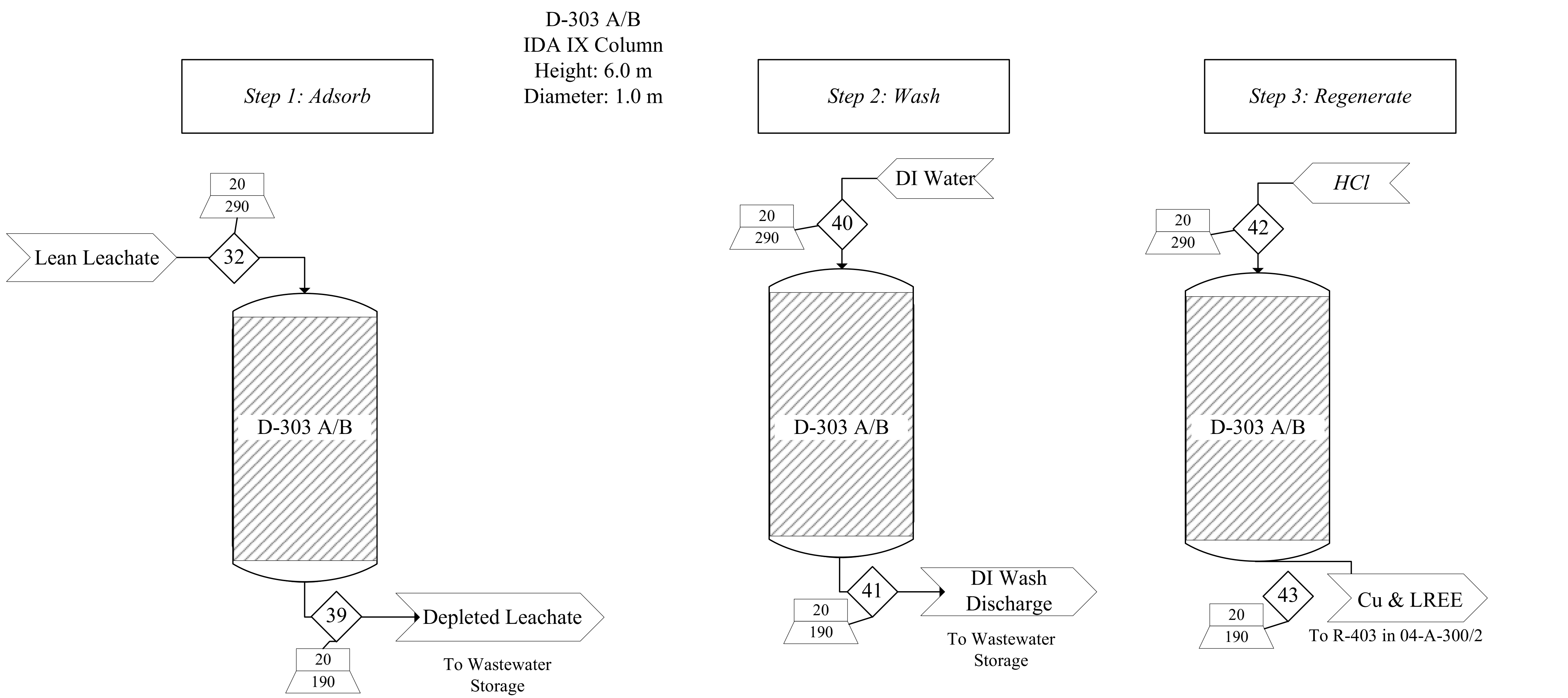
Drawing Type

Process Flow Diagram

Title

Concentration of REEs via Ion Exchange

By/Date	**DWG #/ Sheet #**	**Rev**
CEC/ 05-08-25	03-A-300/2	0



Component / **Stream**	**32**	**39**
HCl	0	0
N_2CO_3	0	0
Zn2+	9.3	7.4
Cu2+	35	1.8
MMC	330	480
HREE	0.17	0.0052
LREE	0.46	0.014
SO_4	290	290
Cl-	0	0
H_2O	13000	13000
Total	14,000	14,000

Component / **Stream**	**40**	**41**
HCl	0	0
N_2CO_3	0	0
Zn2+	0	0.93
Cu2+	0	17
MMC	0	31
HREE	0	0.084
LREE	0	0.22
SO_4	0	0
Cl-	0	0
H_2O	14000	14000
Total	14,000	14,000

Component / **Stream**	**42**	**43**
HCl	370	0
N_2CO_3	0	0
Zn2+	0	2.1
Cu2+	0	37
MMC	0	2.8
HREE	0	0.19
LREE	0	0.5
SO_4	290	290
Cl-	1010	2020
H_2O	14000	14000
Total	16,000	15,000

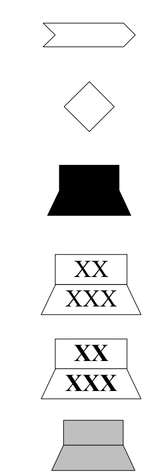
Step	**Event**	**Termination Criteria**
1	Adsorption under leachate feed.	$C_{Cu,OUT} > C_{Cu,MIN}$
2	Wash under DI water feed.	$C_{TSS,OUT} < C_{TSS,MAX}$
3	Regeneration under HCl feed.	$C_{Cu,OUT} = 0$
4	Stand-by	$C_{Cu,OUT} > C_{Cu,MIN}$

*The stand-by termination condition is that duplicate column has completed its adsorption cycle.

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
5. MMC consists of Cl-, Mg^{2+}, Ca^{2+}, Al^{3+}, Mn^{2+}, Na^{2+}
6. For batch steps of F-301 see 03-A-300/2
7. For batch steps of F-303 see 03-A-300/3
8. Na+ from stream 37 grouped in MMC.
9. Stream 31 contains 1% suspended tailings. Assume 50% of this stays in resin bed and that all removed by wash.
10. 10% residual liquid in beds after adsorption cycle. 100% removed in wash cycles..

Special Symbols / Nomenclature



Reference Drawings:

I/O Diagram

Block Flow Diagram

Drawing Type

Process Flow Diagram

Title

Concentration of REEs via Ion Exchange

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	03-A-300/3	0

R-401
HREOx Precipitation Reactor
Height: 3m
Diameter: 2m

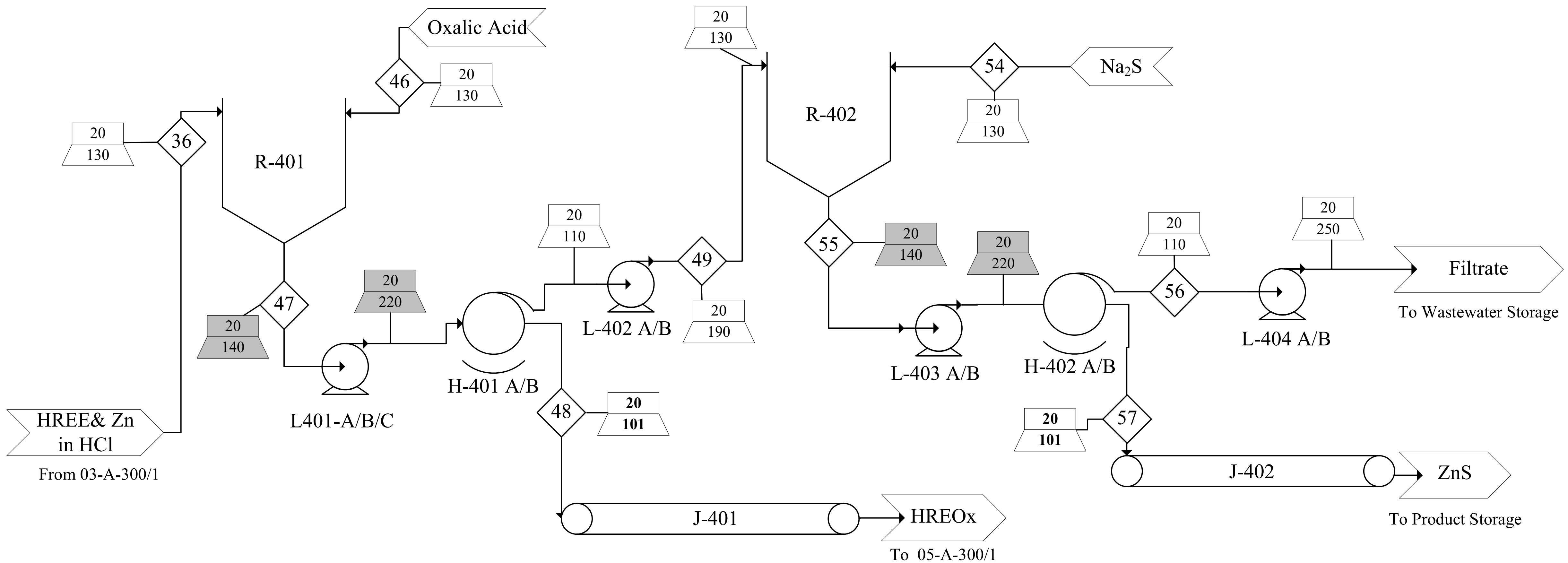
R-402
ZnS Precipitation Reactor
Height: 4 m
Diameter: 3 m

J-401 A/B
HREOx Screw Conveyor
Power: 0.2kW

J-402 A/B
ZnS Screw Conveyor
Power: 0.2kW

H-401 A/B
HREOx RDVF
Area: 0.5m2
Power: 1.07kW

H-402 A/B
ZnS RDVF
Area: 6m2
Power: 30kW



L-401 A/B/C
HREOx Transfer Pump
Power: 0.53 kW

L-402 A/B
REOx Filtrate Pump
Power: 0.55 kW

L-403 A/B
ZnS Transfer Pump
Power: 0.57 kW

L-404 A/B
ZnS Filtrate Pump
Power: 0.93 kW

Component / **Stream**	**36**	**46**	**47**	**48**	**49**	**54**	**55**	**56**	**57**
Oxalic Acid	0	30	20	0	20	0	20	20	0
H2O	14000	3.4	14000	0.64	14000	390	14000	14000	0
Zn2+	93	0	84	0.0039	84	0	10	10	0
Cu2+	0.58	0	0.52	0	0.52	0	0.062	0.062	0
MMC	1300	0	1300	0	1300	80	1300	1300	0
ZnS	0	0	0	0	0	0	110	0	110
REE	1.6	0	0	0	0	0	0	0	0
REOx	0	0	7.8	7.8	0	0	0	0	0
Total	15,000	33	15,000	9	15,000	580	15,000	15,000	110

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Pumps have an assumed efficiency of 70%
5. MMC consists of Cl-, SO42-, and trace Mg2+, Ca2+, Al3+, Mn2+, Na^{2+}.

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX XXX Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type
Process Flow Diagram

Title
Oxalate Precipitation & Sulfide Precipitation

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	04-A-300/1	0

R-403
LREOx Precipitation Reactor
Height: 3m
Diameter: 2m

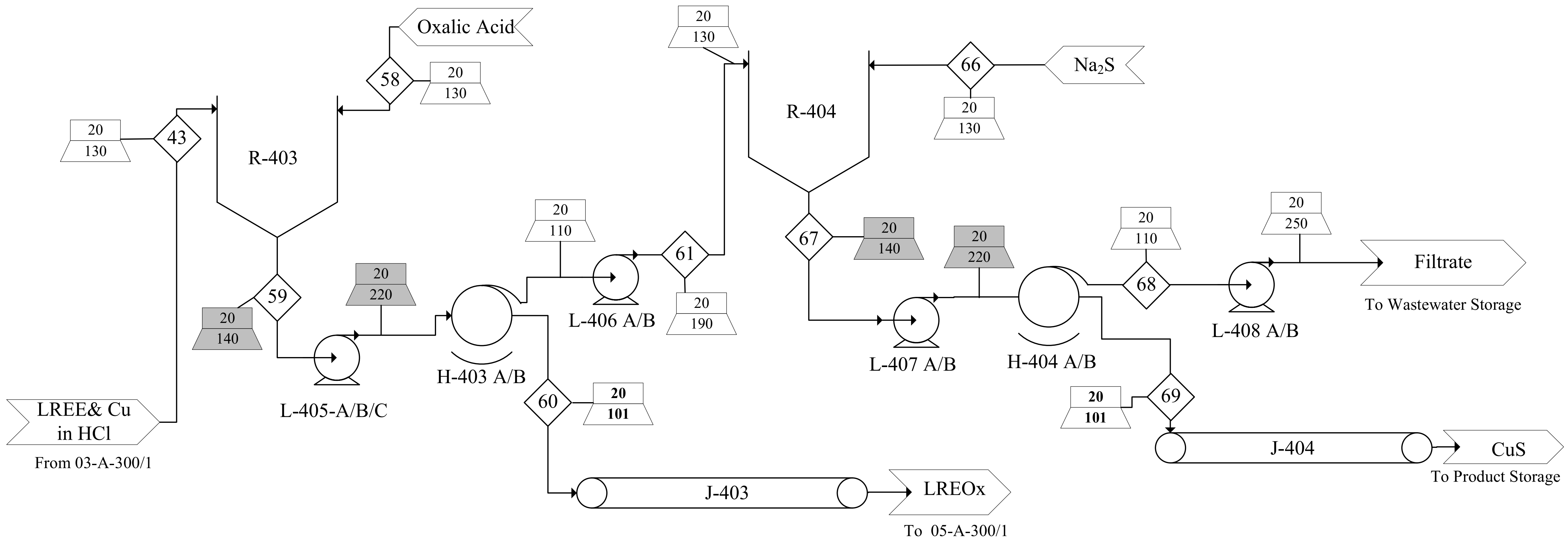
R-404
CuS Precipitation Reactor
Height: 4 m
Diameter: 3 m

J-403 A/B
LREOx Screw Conveyor
Power: 0.2kW

J-404 A/B
CuS Screw Conveyor
Power: 0.2kW

H-403 A/B
LREOx RDVF
Area: 0.5m2
Power: 1.07kW

H-404 A/B
CuS RDVF
Area: 6m2
Power: 30kW



Component / Stream	43	58	59	60	61	66	67	68	69
Oxalic Acid	0	22	15	0.00016	15	0	14	14	0
H_2O	14000	2.2	14000	0.15	14000	170	13000	13000	0
Na_2S	0	0	0	0	0	85	0	0	0
Zn2+	2.1	0	2	0	2	0	0.27	0.27	0
Cu2+	37	0	37	0	37	0	4.9	4.9	0
MMC	2300	0	0	0	2300	0	2300	2300	0
ZnS	0	0	0	0	0	0	2.5	0	2.5
CuS	0	0	0	0	0	0	45	0	45
REE	0.69	0	0	0	0	0	0.13	0.13	0
REOx	0	0	3.3	3.3	0	0	0	0	0
Total	15,000	24	15,000	3	16,000	260	14,000	14,000	48

L-405 A/B/C
LREOx Tansfer Pump
Power: 0.53 W

L-406 A/B
LREOx Filtrate Pump
Power: 0.55 kW

L-407 A/B
CuS Transfer Pump
Power: 0.57 kW

L-408 A/B
CuS Filtrate Pump
Power: 0.93 kW

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Pumps have an assumed efficiency of 70%
5. MMC consists of Cl-, SO42-, and trace Mg2+, Ca2+, Al3+, Mn2+, Na^{2+}.

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX XXX Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

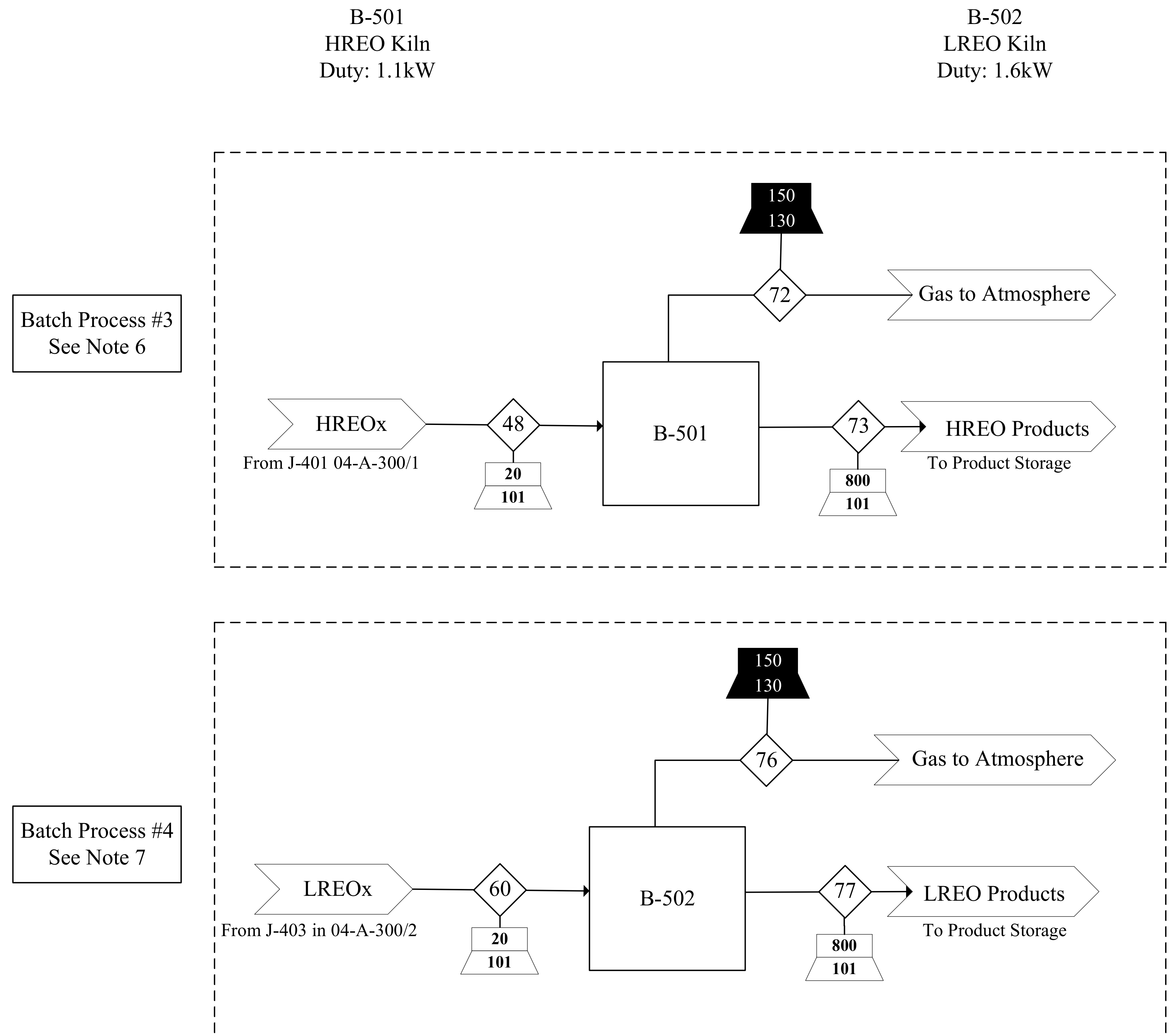
Drawing Type

Process Flow Diagram

Title

Oxalate Precipitation & Sulfide Precipitation

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	04-A-300/1	0



Component / **Stream**	**48**	**72**	**73**
SO_4	0.061	0	0
N_2	0	1.9	0
O2	0	0.13	0
CO_2	0	4.1	0
H_2O	0.64	0.86	0
REOx	7.8	0	0
REO	0	0	3.9
Total	8.5	7	3.9

Component / **Stream**	**60**	**76**	**77**
SO_4	0.014	0	0
N_2	0	1.3	0
O2	0	0.09	0
CO_2	0	0.54	0
H_2O	0.15	1	0
REOx	3.3	0	0
REO	0	0	1.7
Total	3.5	2.9	1.7

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Compressors have a polytropic efficiency of 80%
6. For batch steps of B-501 see 05-A-300/2
7. For batch steps of B-502 see 05-A-300/3

Special Symbols / Nomenclature

Input/Output
Stream Number
Temperature °C (gas) Pressure kPa (gas)
XX XXX Temperature °C (liquid) Pressure kPa (liquid)
XX **XXX** Temperature °C (solid) Pressure kPa (solid)
Temperature °C (mixed) Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

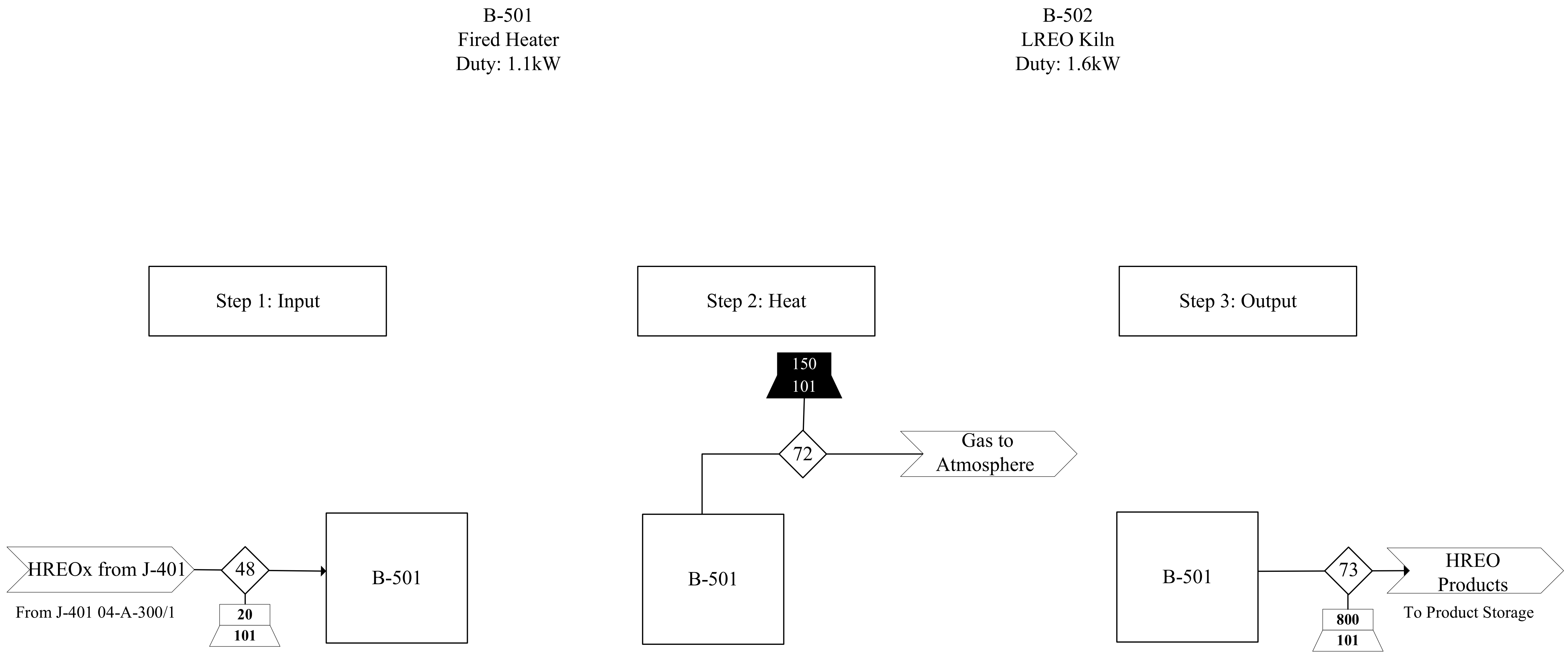
Drawing Type

Process Flow Diagram

Title

Calcination to REO Products

By/Date	**DWG #/ Sheet #**	**Rev**
CEC/ 05-08-25	05-A-300/1	0



Component / Stream	48
SO_4	0.061
N_2	0
O2	0
CO_2	0
H_2O	0.64
REOx	7.8
REO	0
Total	8.5

Component / Stream	72
SO_4	0
N_2	1.9
O2	0.13
CO_2	4.1
H_2O	0.86
REOx	0
REO	0
Total	7

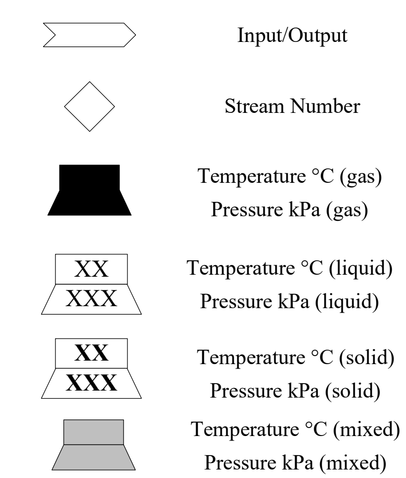
Component / Stream	73
SO_4	0
N_2	0
O2	0
CO_2	0
H_2O	0
REOx	0
REO	3.9
Total	3.9

Step	Event	Termination Criteria
1	Feed oxalates to kiln	$M_{ox,KILN} > M_{ox,MIN}$
2	Bring oxalates to 800 °C	T_{ox} = 800 °C
3	Remove oxide products from kiln	$M_{OXIDE,KILN} = 0$

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Compressors have a polytropic efficiency of 80%

Special Symbols / Nomenclature



Reference Drawings:

I/O Diagram
Block Flow Diagram

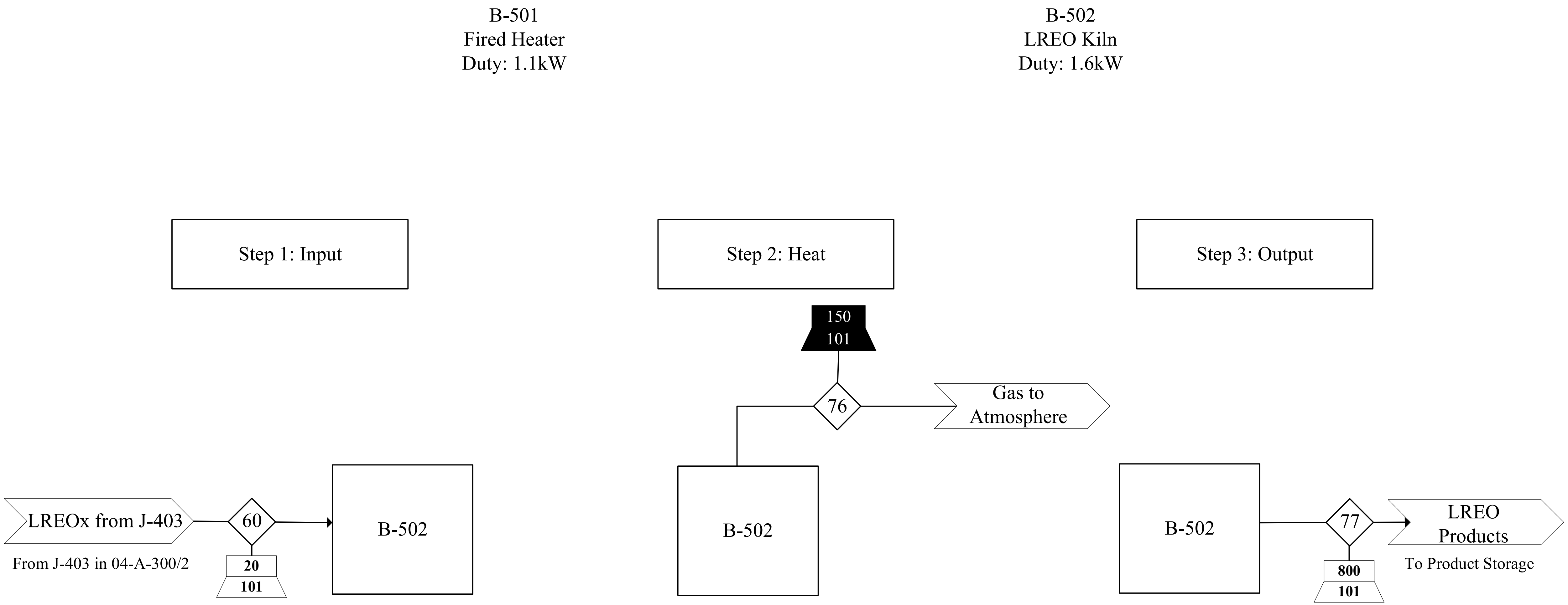
Drawing Type

Process Flow Diagram

Title

Calcination to REO Products

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	05-A-300/2	0



Component / Stream	60
SO_4	0.014
N_2	0
O2	0
CO_2	0
H_2O	0.15
REOx	3.3
REO	0
Total	3.5

Component / Stream	76
SO_4	0
N_2	1.3
O2	0.09
CO_2	0.54
H_2O	1
REOx	0
REO	0
Total	2.9

Component / Stream	77
SO_4	0
N_2	0
O2	0
CO_2	0
H_2O	0
REOx	0
REO	1.7
Total	1.7

Step	Event	Termination Criteria
1	Feed oxalates to kiln	$M_{ox,KILN} > M_{ox,MIN}$
2	Bring oxalates to 800 °C	$T_{ox} = 800$ °C
3	Remove oxide products from kiln	$M_{OXIDE,KILN} = 0$

Notes

1. Drawing not to scale
2. Streams may not balance due to rounding
3. Stream tables are in kg/h
4. Compressors have a polytropic efficiency of 80%.

Special Symbols / Nomenclature

Input/Output

Stream Number

Temperature °C (gas)
Pressure kPa (gas)

XX
XXX
Temperature °C (liquid)
Pressure kPa (liquid)

XX
XXX
Temperature °C (solid)
Pressure kPa (solid)

Temperature °C (mixed)
Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type

Process Flow Diagram

Title

Calcination to REO Products

By/Date	DWG #/ Sheet #	Rev
CEC/ 05-08-25	05-A-300/3	0

Component / **Stream**	**48**	**72**	**73**	**60**	**76**	**77**
SO4	0.061	0	0	0.014	0	0
N2	0	1.9	0	0	1.3	0
O2	0	0.13	0	0	0.09	0
CO2	0	4.1	0	0	0.54	0
CH4	0	0.015	0	0	0.011	0
H2O	0.64	0.86	0	0.15	1	0
Mg Oxalate	0.006	0	0	0.012	0	0
Ca Oxalate	0.0063	0	0	0.016	0	0
Al Oxalate	0.28	0	0	0.32	0	0
Mn Oxalate	0.0028	0	0	0.0054	0	0
Sc Oxalate	2.4	0	0	0.0002	0	0
Y Oxalate	0.047	0	0	0.075	0	0
La Oxalate	0.81	0	0	0.16	0	0
Ce Oxalate	0.045	0	0	0.041	0	0
Pr Oxalate	0.025	0	0	0.023	0	0
Nd Oxalate	0.012	0	0	0.45	0	0
Sm Oxalate	0.0067	0	0	0.1	0	0
Eu Oxalate	0.034	0	0	0.03	0	0
Gd Oxalate	0.011	0	0	0.13	0	0
Tb Oxalate	0.15	0	0	0.0085	0	0
Dy Oxalate	0.025	0	0	0.037	0	0
Ho Oxalate	0.1	0	0	0.0061	0	0
Er Oxalate	0.016	0	0	0.011	0	0
Tm Oxalate	0.085	0	0	0.00079	0	0
Yb Oxalate	0.0068	0	0	0.0043	0	0
Lu Oxalate	2.3	0	0	0.00035	0	0
Mg Oxide	0	0	0.002151	0	0	0.007886
Ca Oxide	0	0	0.002748	0	0	0.012618
Al Oxide	0	0	0.102759	0	0	0.225316
Mn Oxide	0	0	0.001314	0	0	0.004957
Sc Oxide	0	0	0.531717	0	0	0.000169
Y Oxide	0	0	0.027482	0	0	0.08562
La Oxide	0	0	0.561588	0	0	0.216304
Ce Oxide	0	0	0.031067	0	0	0.054076
Pr Oxide	0	0	0.017923	0	0	0.031544
Nd Oxide	0	0	0.008603	0	0	0.608354
Sm Oxide	0	0	0.004899	0	0	0.139696
Eu Oxide	0	0	0.025092	0	0	0.040557
Gd Oxide	0	0	0.008245	0	0	0.178
Tb Oxide	0	0	0.107538	0	0	0.011716
Dy Oxide	0	0	0.019118	0	0	0.051823
Ho Oxide	0	0	0.074082	0	0	0.008562
Er Oxide	0	0	0.01171	0	0	0.015096
Tm Oxide	0	0	0.064523	0	0	0.001127
Yb Oxide	0	0	0.005138	0	0	0.006084
Lu Oxide	0	0	1.792304	0	0	0.000496
Total	8.5	7	3.4	3.5	2.9	1.7

Notes

1. Extended stream tables for calcination of REOx to HREO product streams.

2. Stream 77 is the LREO product stream.

3. Stream 73 is the HREO product stream.

Special Symbols / Nomenclature

Input/Output

Stream Number

Temperature °C (gas)
Pressure kPa (gas)

XX
XXX
Temperature °C (liquid)
Pressure kPa (liquid)

XX
XXX
Temperature °C (solid)
Pressure kPa (solid)

Temperature °C (mixed)
Pressure kPa (mixed)

Reference Drawings:

I/O Diagram
Block Flow Diagram

Drawing Type

Process Flow Diagram

Title

Calcination to REO Products

By/Date	**DWG #/ Sheet #**	**Rev**
CEC/ 05-08-25	05-A-300/4	0

Appendix A: Alternatives Analysis Report Regarding Rare Earth Element Recovery from Acid Mine Drainage

Revisions

The best alternative process was changed from Solvent Extraction to Scandium Recovery. A BFD for this process has been created (Drawing 00-A-500/4). A general description of the Scandium recovery process has been updated to reflect the Drawing 00-A-500/4. Grammar and wording have been revised in line with markups from previous submission. An expanded economic analysis has also been added.

Introduction

The purpose of this report is to present the alternative processes to recover rare earth elements (REEs) from acid mine drainage (AMD) and propose a recommendation and best alternative. The project is to be implemented at the Berkley Pit site in Butte, Montana. The objective of the report is to propose a recommended process and best alternative to recover and purify acid mine drainage at an industrial scale and receive approval to proceed with the scoping study.

Four processes were developed and analyzed. Process #1 utilizes treated silica-chitosan ion exchange and mesoporous silica beads for the selective extraction of heavy rare earth elements (HREEs) and light rare earth elements (LREEs) from AMD. Process #2 utilizes acid leaching with HNO_3 and solvent extraction using D2EHPA with subsequent selective precipitation using oxalic acid and dilute hydrochloric acid/water. Process #3 is a membrane separation that utilizes supported liquid membranes embedded with (D2EHPA). Process #4 is the recovery of scandium by the addition of a separator to pretreatment processes. After analysis, Process #1 is the recommended process due to its greater operability, despite producing products of relatively low purity.

The scope of work included is the analysis of four alternative processes to recover and purify REEs from AMD at the input/output (I/O) level to determine two most viable options, an analysis of the two most viable processes at the qualitative block flow diagram (BFD) level, and a proposal of the recommended process and best alternative.

Present Situation

REEs are an important raw material for many products including fiber optics, specialty alloys used in aerospace applications, and permanent magnets used in turbines and electrical drives.[1] They have become synonymous with technical innovation, increasing world demand. Due to expansion of the wind turbine and EV markets, recovery of REEs is a significant topic in scientific communities. The key REEs for this sector are praseodymium (Pr), neodymium (Nd), terbium (Tb), and dysprosium (Dy).[2] Work will be conducted at the Berkley Pit site in Butte, Montana.

Since 1998, China has become the leading supplier in REEs accounting for over 70% of global supply.[3] Chinese producers are state-owned conglomerates, which has given China strategic control over global REE prices. Due to this Chinese-concentrated supply chain there is increased volatility in the REE price and REEs are at risk for supply chain disruption.[4] This has caused concern in the global community. All REEs were listed as critical minerals and Pr, Nd, Tb, and Dy were listed as critical materials for energy on the 2023 DOE Critical Materials List.[4] This has led to the development of U.S. DOE-NETL's Feasibility of Recovering Rare Earth Elements program. The implementation of this program demonstrates the growing interest of the U.S. to develop potential processes for the recovery of 90-99 wt% rare earth element oxides.[5]

Background

REE extraction from AMD is a relatively new area of research. One benefit is that it bypasses the costs associated with removal of rock deposits and subsequent grinding and acid leaching.[6] The Rare Earth Extraction Facility, a pilot plant operated by the West Virginia Water Research Institute, is the only operating facility producing REE concentrates from AMD. Phoenix Tailings is constructing a pilot refinery in Massachusetts for processing mining waste and/or mixed REE concentrates into pure REE metals through a novel pyrometallurgical process. UCORE is currently building a pilot facility for REE mixed oxide purification from an upstream

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[1] National Energy Technical Laboratory. "Rare Earth Elements - A Subset of Critical Minerals".https://www.netl.doe.gov/resource-sustainability/critical-minerals-and-materials/rare-earth-elements

[2] Theaker, N. (Producer). (2025, February 2). *North Dakota lignite rare earth element (REE) exploration and extraction review* [Video]. Bureau of Economic Geology.

[3] Imholte, D. Nyugen, R. (2016, March). "China's rare earth supply chain: illegal production, and response to new Cerium demand". *Idaho National Laboratory*.

[4] Department of Energy. (2024, July). "Notice of Final Determination of 2023 DOE Critical Materials List".

[5] National Energy Technology Laboratory. "REE Program Overview - Program Mission, Objectives, and Goals". https://www.netl.doe.gov/coal/rare-earth-elements/program-overview/mission-goals

[6] Theaker, N. (2025, January 22). *Group meeting with Dr. Nolan Theaker, Research Subject Matter Expert* [Meeting].

concentrate using solvent exchange technology.[7] Beyond commercialized activities, there is a wealth of lab-scale projects specific to REE recovery from AMD.[8,9,10]

AMD is prevalent in the United States due to legacy mining activity. There is a growing concern of the sustainability of mining production in the U.S. and this presents an opportunity for resource utilization of AMD.[11] Berkeley Pit, an AMD site in Butte, Montana, is a critical site that poses a threat to the ecosystem and human health due to its high concentration of heavy metals and strong acidity (pH~2-3). Feed composition data was based on data from 2003 and 2005.[12]

The alternatives produced in this report are an adaptation to recent technologies developed by West Virginia Water Research Institute's pilot plant. Modifications have been made to analyze the different pretreatment processes and extraction methods available and how they affect REE recovery. The operation is an adaptation of known hydrometallurgical extraction methods. Integral technologies to this approach involve AMD neutralization through pretreatment processes or novel membrane separation to produce preconcentrated REE, acid leaching, an adaptation of solvent extraction (SX) or selective extraction of REEs in ion exchange (IX) resin beads, precipitation of REEs as oxalates, and subsequent calcination into rare earth oxide (REO) products.

Upstream processes include the pretreatment of AMD which is an essential step to produce saleable REEs. One process is precipitation of gangue elements by adding a base and flocculation agent. The base maintains a constant pH to hold REEs in solution and the flocculation agent aids in the settling of waste precipitate.[13] The process is repeated at a higher pH to precipitate out REEs to optimize recovery. Another option is to use supported liquid membranes impregnated with a REE selective extractant dissolved in an organic phase. The

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[7] Ucore Rare Metals Inc. (2025, January 22). *Ucore Rare Metals Inc. | Webinar Replay* [Video]. Red Cloud TV. HYPERLINK "https://www.youtube.com/watch?v=C1Yr3njdOgw"https://www.youtube.com/watch?v=C1Yr3njdOgw
[8] Hermassi, M et al. "Recovery of Rare Earth Elements from Acidic Mine Waters: An Unknown Secondary Resource." *The Science of the total environment* 810 (2022): 152258–152258. Web.
[9] Depp, Charles T et al. "Potential for High-Grade Recovery of Rare Earth Elements and Cobalt from Acid Mine Drainage via Adsorption to Precipitated Manganese (IV) Oxides." *Chemosphere (Oxford)* 364 (2024): 143144-. Web.
[10] Wang, Qiuming et al. "Amine Sorbents for Selective Recovery of Heavy Rare-Earth Elements (Dysprosium, Ytterbium) from Aqueous Solution." *ChemPlusChem (Weinheim, Germany)* 85.1 (2020): 130–136. Web.
[11] Yanan, J. et al. "A review of acid mine drainage: Formation mechanism, treatment technology, typical engineering cases and resource utilization". ***Process Safety and Environmental Protection***, Vol. 170, (2023) pgs. 1240-1260.
[12] Pellicori, C. H. et al. "Geochemistry of the rare earth elements and uranium in the acid Berkeley Pit lake, Butte, Montana". ***Applied Geochemistry,*** 20 (2005) pgs. 2116-2137.
[13] Ziemkiewicz et al. "*SYSTEMS AND PROCESSES FOR RECOVERY OF HIGH-GRADE RARE EARTH CONCENTRATE FROM ACID MINE DRAINAGE*". Pub. No.: US 2023/0039988 A1, ***United States Patent Application Publication***.

REEs are transferred across the membrane and stripped with acid.[14] This produces preconcentrated REEs that can be used for further purification methods.

Midstream processes involve further purification of REEs. Two processes were evaluated for viability, IX with impregnated resin beads and batch solvent extraction. IX uses impregnated resin beads that are selective for REEs. Different resins have differing affinities for metallic ions. This allows for the selective removal of the contaminates using highly selective resins. After contaminates are removed the REEs in solution can be passed through IX resin beads that are selective for light REEs and heavy REEs. The resin is washed with oxalic acid to produce light rare earth oxalates (LREOx). The heavy REEs are mixed with oxalic acid to produce the heavy rare earth oxalates (HREOx). SX is a process that utilizes an extractant with high affinity for REEs. The extractant is then selectively stripped to produce concentrated LREOx and HREOx from the column.

The downstream processes recover concentrated REE ions in solution into saleable solid mixed oxides. This is done via oxalate precipitation and conversion to oxides in a furnace. The oxides are washed in water or mildly acidic acid to remove any unwanted contaminates. They undergo calcination to produce final product streams of light rare earth oxides (LREO) and heavy rare earth oxides (HREO). Desired product yield is 90-99% light rare earth oxides (LREOs) and 90-99% heavy rare earth oxides (HREOs) or minimum of 60% mixed rare earth oxides (MREOs).

Economics risks associated include the use of consumable chemicals such as acids and bases that may not be recoverable. The development of optimal conditions for solvent extraction is of primary concern for economic analysis.

Demand for rare earth oxides year over year is increasing with an article in 2024 putting the global demand at over 300,000 metric tons per year. China is the main exporter of rare earth oxides, producing 240,000 tons in 2023, while the United States only produced 43,000 tons in the same year.[15]

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[14] Middleton, Andrew et al.”Recovery of Rare Earth Elements from Acid Mine Drainage with Supported Liquid membranes: Impacts of Feedstock Composition for Extraction Performance”. ***Environmental Science and Technology*** S8 (2024), 2998-3006.
[15] U.S. Geological Survey. (2024). *Rare earths*. In *Mineral commodity summaries 2024* (pp. 134-135). Reston, VA: U.S. Geological Survey. Retrieved February 26, 2025, from https://pubs.usgs.gov/periodicals/mcs2024/mcs2024-rare-earths.pdf

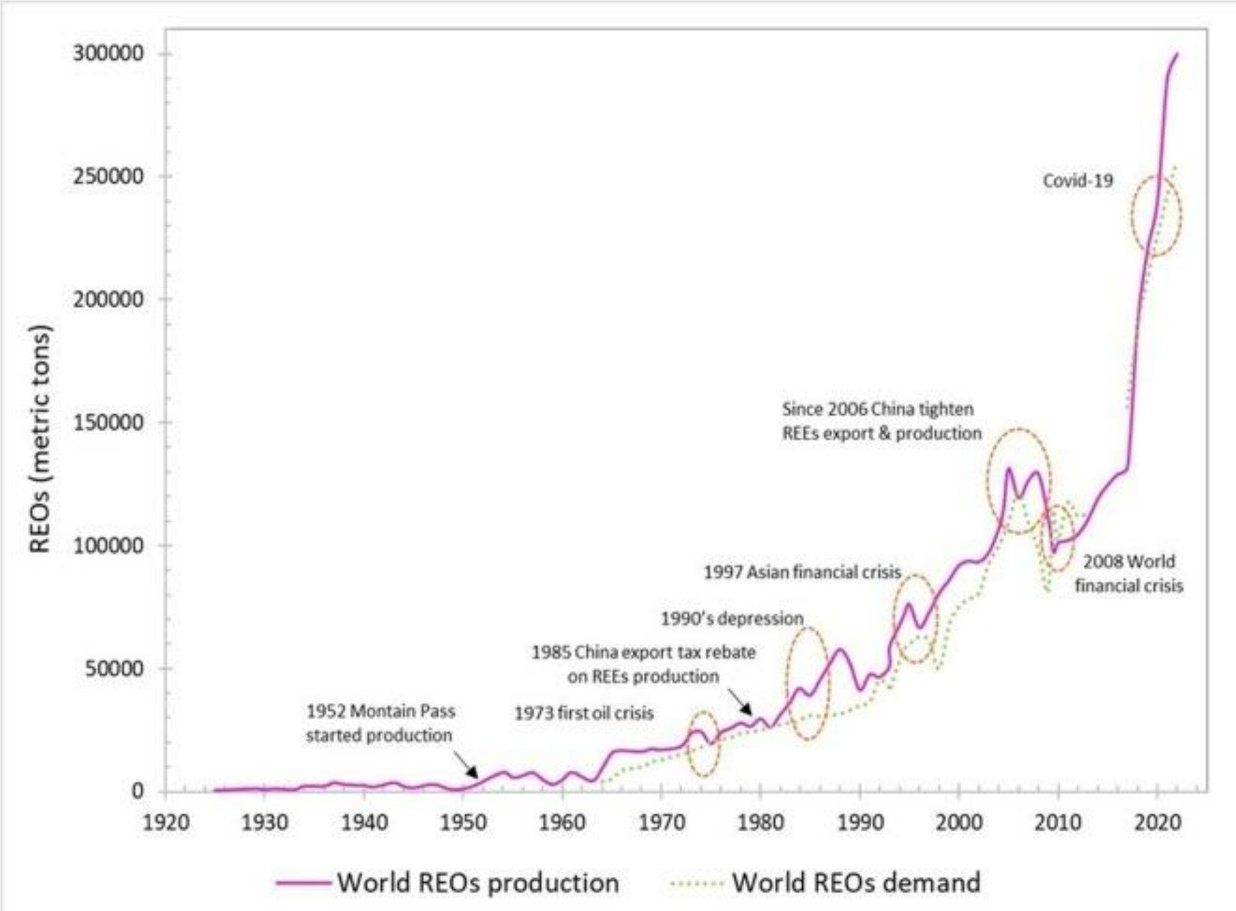

Figure 1: Production and Demand of Rare Earth Oxides.[16]

The most profitable REE's for this project are dysprosium, scandium, neodymium, and terbium. This profitability is based on market pricing and REE concentration at the Berkley Pit site. Table 1 below details the potential yearly economic outlook of this project with perfect separation of each REE. Average REE concentration values from the Berkley Pit are used with a flowrate of 1,000,000 L/hr. Final pricing will change based on flowrate, purity, and recovery values.

[16] Merroune, A., Ait Brahim, J., Berrada, M., Essakhraoui, M., Achiou, B., Mazouz, H., & Beniazza, R. (2024). A comprehensive review on solvent extraction technologies of rare earth elements from different acidic media: Current challenges and future perspectives. *Journal of Industrial and Engineering Chemistry, 139*, 1-17. https://doi.org/10.1016/j.jiec.2024.04.042

Oxide	USD/kg	Concentration (kg/L)	USD/year
Lanthanum oxides	3.7	2.75E-07	$8,926.77
Cerium oxides	3.2	1.10E-06	$30,586.79
Praseodymium oxides	57	1.07E-07	$53,304.94
Neodymium oxides	49	4.92E-07	$211,389.14
Samarium oxides	1.8	1.24E-07	$1,952.99
Europium oxides	33	3.32E-08	$9,609.29
Gadolinium oxides	28	1.65E-07	$39,655.19
Terbium oxides	590	2.76E-08	$142,792.61
Dysprosium oxides	280	1.78E-07	$435,893.56
Holmium oxides	59	3.71E-08	$19,178.48
Erbium oxides	30	1.08E-07	$28,306.12
Ytterbium oxides	16	1.01E-07	$14,165.86
Lutetium oxides	615	1.54E-08	$82,777.49
Scandium oxides	1060	3.38E-08	$314,321.92
Yttrium oxides	3.4	8.76E-07	$25,727.38

Table 1: Yearly potential profit per oxide at 1,000,000 L/hr

Based on information from this table, with such a high flowrate the yearly profits for REE's could total higher than $1,000,000, but these will likely be diminished by operating costs and capital cost expenditure. Further research into the sale of byproducts and methods of increasing the efficiency of the process will be necessary.

Both the proposal and alternative will contain NFPA Division 1 process areas due to the use of higher temperature furnaces. Further research into NFPA division classification for organic reagents is needed.

Recommended Process Description

Ion exchange is the recommended process using precipitation preconcentration. It utilizes sorbents to recover an HREE stream and an LREE stream from an AMD feed. The I/O diagram (01-A-001/1) gives the compositions and flow rates of the raw materials, products, and waste streams. The process takes 1000 kg/h of AMD as a feed and produces approximately 900 kg/h of treated water, 1.1 kg/h of LREO, 2.6 kg/h of HREO, and 100 kg/h of waste. The LREO product has 1% purity, total rare earth oxide basis. The HREO product has 5% purity, total rare earth oxide basis. The waste consists mainly of a gangue mineral slurry, which is 2-6 % solids.[17,18]

The major unit operations of the IX recovery route can be found in the BFD 02-A-001. The process begins with staged precipitation to remove gangue minerals. AMD is fed to a clarifier (Separator #1) where a base is added until the solution reaches a pH of 4.0 to 4.5. At this pH, gangue minerals are precipitated out primarily as solid hydroxides. A polymer flocculant is added to decrease settling time. The settled solids are discarded as a waste slurry at 2-6% solids. The liquid overflow is sent to a second clarifier (Separator #2) where the pH is brought to 8.0-8.5. The REE ions in solution precipitate out as hydroxides and are recovered in a slurry concentrate. 90% of REEs are recovered in the slurry concentrate, with the balance leaving in the gangue mineral slurry.

The REE concentrate slurry is then redissolved and brought to the optimal pH of the first of three fixed IX resin beds in Separator #3. All resin beds operate semi-batch, alternating between adsorption and desorption phases. The first resin bed set (Separator #4) removes approximately 92% of remaining Mn and Cu using silica-chitosan beads treated with acetic acid and acetyl acetone and mixed with methanol. The beads are stripped with hydrochloric acid, which takes the adsorbed Mn and Cu to be recovered.[19]

The second fixed resin bed (Separator #5) removes approximately 90% of the remaining Al and Fe using KIT-6 mesoporous silica treated with 1,4-phthaloyl diamido-propyltriethoxysilane and 1,4-phenylenedioxydiamido-propyltriethoxysilane. The beads are stripped with phosphoric acid, which takes the Al and Fe to be recovered.[20]

The third fixed resin bed (Separator #6) would use Santa Barbara Amorphous-15 silica treated with diglycholchloride resin to selectively adsorb (L or H) REEs. The sorbents collect approximately 70% of the LREEs (Tb, Dy, Ho), 5% of the HREEs (Y, La, Ce, Pr, Nd) and 30% of mixed REEs (Sm, Eu, Gd, Er, Tm, Yn, Lu) in the LREE stream. The remaining is the HREE

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[17] Gangue minerals refer to low value minerals such as iron, aluminum, silicates and zinc. We use the term to refer to both ions in solution and their corresponding precipitates.

[18] Mass balance for the staged precipitation step used data in: Zhang, W., & Honaker, R. Q. (2018). Rare earth elements recovery using staged precipitation from a leachate generated from coarse coal refuse. *International Journal of Coal Geology, 195*, 189–199. https://doi.org/10.1016/j.coal.2018.06.008

[19] Deepika Lakshmi Ramasamy, Ville Puhakka, Sidra Iftekhar, Anna Wojtuś, Eveliina Repo, Samia Ben Hammouda, Evgenia Iakovleva, Mika Sillanpää, N- and O- ligand doped mesoporous silica-chitosan hybrid beads for the efficient, sustainable and selective recovery of rare earth elements (REE) from acid mine drainage (AMD): Understanding the significance of physical modification and conditioning of the polymer,Journal of Hazardous Materials, Volume 348 ,2018, Pages 84-91, ISSN 0304-3894, https://doi.org/10.1016/j.jhazmat.2018.01.030.

[20] Hu Y, Misal Castro LC, Drouin E, Florek J, Kählig H, Larivière D, Kleitz F, Fontaine FG. Size-Selective Separation of Rare Earth Elements Using Functionalized Mesoporous Silica Materials. ACS Appl Mater Interfaces. 2019 Jul 3;11(26):23681-23691. doi: 10.1021/acsami.9b04183. Epub 2019 Jun 18. PMID: 31117444.

stream, which passes through the sorbent. The LREE is extracted from the sorbent using ammonium oxalate.[21]

The REE ions are converted to solid precipitates for further processing in separator #7. Following addition of certain reagents, REE ions can be extracted from their solutions into oxalic acid. The LREEs and HREEs can then be precipitated out as oxalates. In a final step (separator #8), the REE oxalates are oxidized in Reactor #1 and Reactor #2. LREO and HREO are the final product streams.

Best Alternative: Scandium Recovery

Sc^{3+} is the most valuable metal ion contained in the AMD feed. To design a scandium recovery process, the main points of Sc loss in the proposal are modified. In the proposal, 14% of Sc^{3+} co-precipitates with the iron and gypsum waste material at pH = 3 and is lost. The remaining Sc^{3+} co-precipitates with the gypsum, major metals and REEs at pH = 7. This intermediate product stream undergoes acid leaching. When the acid leachate is filtered, 85% Sc adsorbs to the alumina silicates and gypsum solid waste and is lost. The remaining Sc^{3+} contained in the leachate is transferred to neutralization. When the neutralized leachate is filtered, 33% of the Sc^{3+} is adsorbed to the aluminum hydroxide byproduct and is lost in this byproduct stream.

The remove these losses, Clarifier 1 is titrated only to pH = 3, where 98% of iron species are removed. Clarifier 2, the Sc-recovery Clarifier, is titrated to pH = 6, where 80% of Sc^{3+} is captured. Clarifier 3 is then titrated to pH = 7.5, where the balance of major metals, Sc^{3+} and $REEs^{3+}$ are captured. The Sc-enriched slurry from Clarifier 2 is sent to a designated acid leaching vessel, where prolonged leaching at a lower pH allows 90% of Sc^{3+} to be recovered into the leachate. This Sc^{3+}-enriched leachate is now sent to an alternative Gypsum & Aluminum removal process. In this process, 85% is Sc adsorbs to the alumina silicates and gypsum solid precipitates. Instead of being discarded, this slurry stream is then sent to an additional acid leaching step. pH is brought extremely low, and the adsorbed Sc^{3+} is liberated from the surface of the gypsum and aluminum silicates at a faster rate than the solid species themselves redissolve. The purified Sc^{3+}-enriched leachate is then sent to IX and oxalate precipitation processes.

The scandium capture process relies on the higher affinity of the Sc^{3+} ion to anionic sites on whatever solids species it is around. This high affinity, however, creates problems in the IX columns. Research by Leipke (2022) found that Sc^{3+} has such a high affinity for IDA resins, that the only method of recovering it is burning the resin itself after loading.[9] This option may be

[21] Support effects in rare earth element separation using diglycolamide-functionalized mesoporous silica
E. Juère, J. Florek, D. Larivière, K. Kim and F. Kleitz, New J. Chem., 2016, 40, 4325 DOI: 10.1039/C5NJ03147H

viable for secondary REE sources that contain a higher fraction of Sc^{3+}) such as coal ash, however in the AMD feed of Berkley Pitt, this could not be justified.

Additional Alternatives

A. Solvent Extraction

The recovery of a mixed heavy and light REE concentrate through solvent extraction is the best alternative to the proposed process. As shown in 01-A-002, the process takes 1000 kg/h of AMD as feed and produces 4.9 g/h of pure LREO product, 2.1 g/h of HREO product, 900 kg/h of treated water, and 100 kg/h waste. Both REO product streams have purities of 90% total rare earth oxide content. The waste is primarily a gangue mineral tailings slurry.

The major unit operations of the solvent extraction recovery route can be found in the block flow diagram of 02-A-002. A staged precipitation pretreatment of AMD is performed to remove gangue minerals. AMD is fed to a clarifier (Seperator #1) where a base is added until the solution reaches a pH of 4.0 to 4.5. At this pH, gangue minerals are precipitated out as solid hydroxides and other more complex minerals.[22] A polymer flocculant is added to decrease settling time. The settled solids are discarded as a waste slurry at 2-6% solids. The liquid overflow is sent to a second clarifier (Seperator #2) where the pH is brought to 8.0-8.5. The REEs ions in solution precipitate out as hydroxides and are recovered in a as slurry concentrate (Stream 9). 90% of REEs are recovered in the slurry concentrate, with the balance leaving in the gangue mineral slurry.

The concentrated slurry of REE hydroxides is then redissolved in acid and separated from any insoluble components (Separator #3). The remaining REE concentrate solution is fed to Separator #4, where the REE ions are extracted into the organic phase. An organic: aqueous ratio of 1:1 is used. Two separate stripping processes extract the LREEs, and then the HREEs, into oxalic acid (Separators #5 and #6, respectively). The organic solvent is then sent to solvent recovery which recycles back to separator #4 (Separator #7).

The LREE and HREE organic are stripped into Stream 20 and Stream 21 and are fed to Separators #8 and Separator #9, where the REEs are precipitated out as oxalates and

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[22] The mass balance assumed that all gangue metal ions precipitate out as hydroxides, $MCC(OH)_3$. In reality, many different precipitates are possible. This is reviewed in: Torres-Rivero, K., Bastos-Arrieta, J., Florido, A., & Martí, V. (2023). Potential Use of Precipitates from Acid Mine Drainage (AMD) as Arsenic Adsorbents. *Water*, *15*(18), 3179. https://doi.org/10.3390/w15183179

filtered. In a final step, the REE oxalates are converted to oxides in Reactor #1 and Reactor #2. LREO and HREO are the final product streams.

B. Membrane Separation

An additional alternative is the use of supported liquid membranes to create a concentrated solution of REEs. Membrane separation would replace the staged precipitation used in the pretreatment step of both the proposal and the best alternative. The I/O diagram for this additional alternative is shown in 01-A-003. Note that the product stream for this alternative is REE ions in solution.

The use of membranes to recover REE from AMD is not well researched. A study published in January 2024 reported an REE recovery from 60% to 83% with a final dry weight of 0.2% to 8.5% REE depending on feed composition and pH. The final composition of the product reported is the average of the REE yield across the AMD samples at 3% with the remaining 97% mostly consisting of iron and calcium. Exact product composition depends on the feedstock conditions which is site dependent.[23]

Membranes could prove to be a useful tool for initial cleanup comparable to staged precipitation but cost of production and maintenance would need to be investigated. Membranes also carry the drawback of easily getting fouled and may need to be changed out regularly. Ideally AMD would be separated through multiple membranes in parallel, but at current recovery rates this method would prove non-viable to produce a complete product.

Justification

The focus of this project is to clean up AMD sites while yielding the most REEs. Of the four alternatives explored in this report, ion exchange and solvent extraction can produce 90+% MREOs which could be sold. The other alternatives, membrane separation and precipitation with

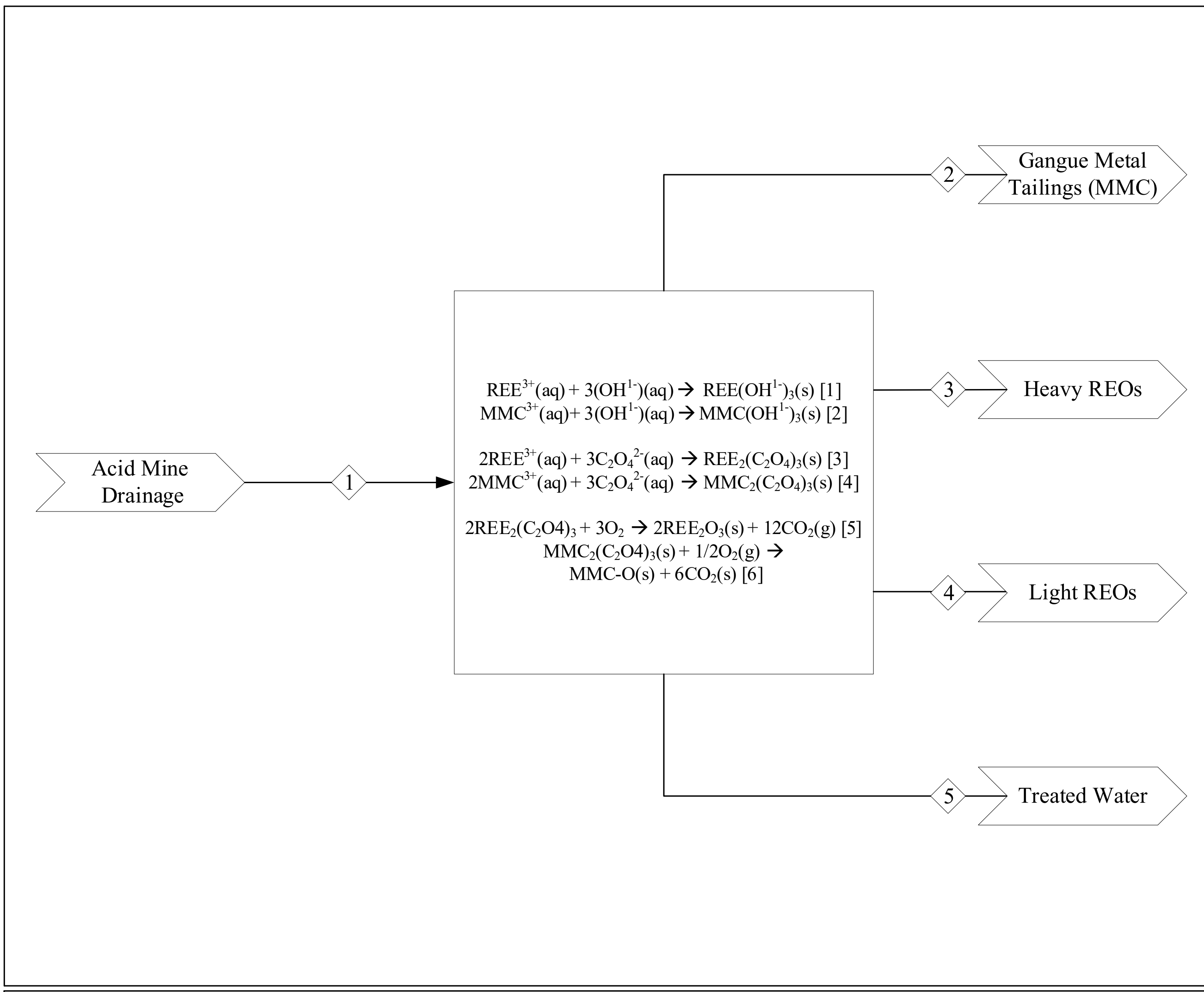
[23] Middleton, A. Hedin, B. Hsu-Kim,H. (2024). "Recovery of Rare Earth Elements from Acid Mine Drainage with Supported Liquid Membranes: Impacts of Feedstock Composition for Extraction Performance". *Environmental Science and Technology.* 202 *58* (6), 2998-3006.

scandium recovery are valuable pretreatment methods but are not capable of being implemented as a complete process at this time.

Ion exchange is a relatively straightforward process that focuses on separating REEs from gangue minerals and other non-desired contaminants through adsorption using a variety of sorbents that are regenerated using acid. The bulk of the process takes place in three fixed resin beds which have excess capacity to handle changes in feed variations allowing for a stable process.

Solvent extraction functions by the affinity of REEs to bind to the extractant and modifier in the organic phase. This process has similar yields to ion exchange but requires a larger amount of equipment and is more prone to upsets due to fluctuating precipitation amounts. These fluctuations result from the source being open to the environment. Excess and or a lack of rain needs to be accounted for in the instrumentation of the process. The correct pH needs to be maintained in each separator which requires more monitoring resulting in higher operating costs.

With operability and maintenance in mind, ion exchange is the recommended process moving forward and solvent extraction is the best alternative. Currently, the ion exchange process described in this report does not reach the correct purity specifications, but it has potential to be a more stable process. If configured correctly, ion exchange could prove to be the most sensible alternative to implement.



Notes

1. Eq. [1] and [2] show the precipitation of REEs and MMCs from AMD feed as hydroxide solids. This occurs with the addition of a base.

2. Eq [3] and [4] show REE and MMC precipitation as oxalates. This occurs with the addition of oxalic acid.

3. Eq [5] and [6] show REEs and MMC oxidation. This occurs in a furnace.

4. Waste water in stream 2 is based on 2-6% solids in slurry waste streams.

5. Top section is an atomic balance and do not sum. Bottom section is in kg/hr

Special Symbols / Nomenclature

Stream Label

Material or Product Stream

Reference Drawings:

Drawing Type

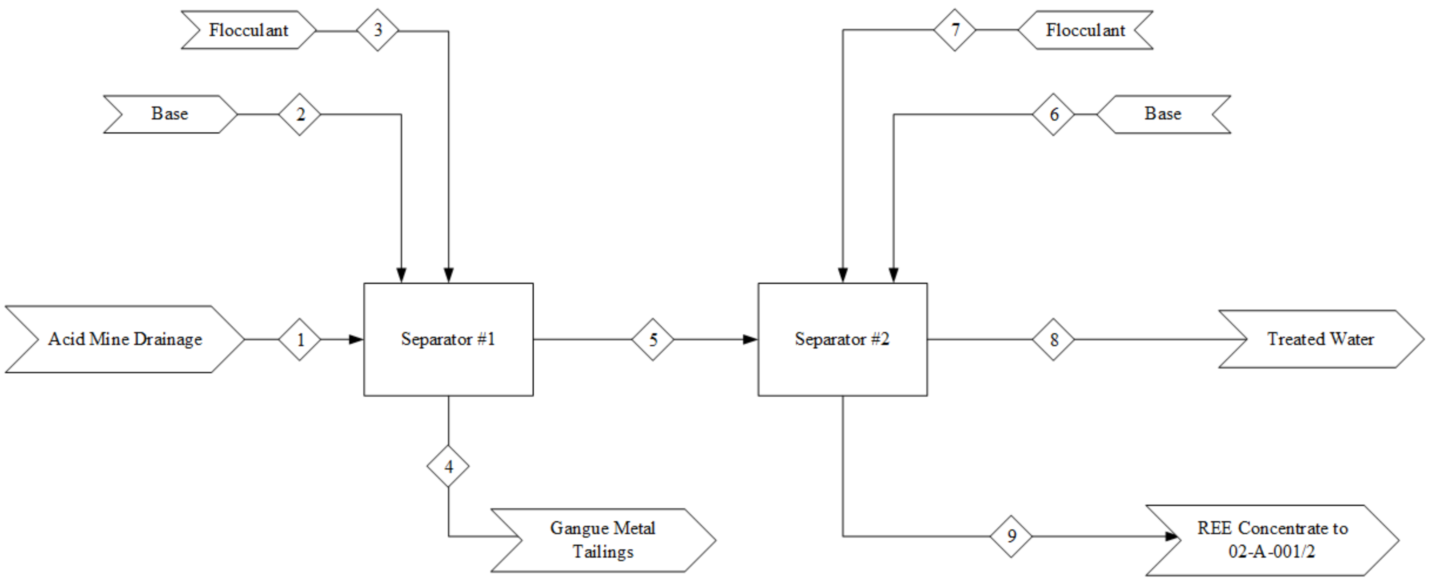
Input/Output Diagram

Title

Alternative: Solvent Extraction

By/Date	DWG #/ Sheet #	Rev	Page
AW/FEB25	01-A-002/1	0	1

Component/Stream	1	2	3	4	5
Sc	3.4E-05	1.4E-05	4.75E-07	2.04E-07	9.0E-07
Y	8.8E-04	1.9E-05	3.57E-05	1.88E-06	5.0E-05
La	2.8E-04	1.8E-05	2.79E-05	1.47E-06	3.9E-05
Ce	1.1E-03	6.2E-05	7.45E-05	3.92E-06	1.0E-04
Pr	1.1E-04	3.6E-05	4.82E-06	2.54E-07	6.8E-06
Nd	4.9E-04	1.6E-05	2.84E-05	1.50E-06	4.0E-05
Sm	1.2E-04	1.9E-05	5.56E-06	1.72E-07	7.6E-06
Eu	3.3E-05	1.0E-05	1.15E-06	3.57E-08	1.6E-06
Gd	1.6E-04	1.3E-05	7.59E-06	2.35E-07	1.0E-05
Tb	2.8E-05	1.3E-05	8.49E-08	1.98E-07	3.8E-07
Dy	1.8E-04	1.9E-05	1.97E-06	4.61E-06	8.8E-06
Ho	3.7E-05	9.8E-06	4.79E-07	1.12E-06	2.1E-06
Er	1.1E-04	1.1E-05	2.97E-06	1.27E-06	5.7E-06
Tm	1.6E-05	9.6E-06	4.13E-07	1.77E-07	7.9E-07
Yb	1.0E-04	1.5E-05	2.44E-06	1.04E-06	4.6E-06
Lu	1.5E-05	1.5E-05	7.09E-07	3.04E-07	1.3E-06
Fe (III)	8.2E-01	8.2E-01	9.90E-05	9.90E-05	4.0E-04
Al	2.9E-01	5.6E-02	1.4E-03	1.4E-03	5.6E-03
Na	9.2E-02	6.1E-03	1.1E-02	1.1E-02	4.4E-02
Mg	5.1E-01	1.9E-03	1.0E-01	1.0E-01	4.1E-01
Ca	4.7E-01	4.0E-02	7.6E-02	7.6E-02	3.0E-01
Mn	2.3E-01	1.2E-02	4.5E-02	4.5E-02	1.8E-01
Si	7.6E-02	2.0E-02	3.4E-03	3.4E-03	1.4E-02
Zn	6.3E-01	1.5E-03	1.9E-02	1.9E-02	7.5E-02
Cu	1.4E-01	2.2E-03	2.0E-03	2.0E-03	7.9E-03
F	3.2E-02	9.4E-06	0	0	3.2E-02
Cl	1.2E-02	9.4E-06	0	0	1.2E-02
K	7.7E-03	9.4E-06	0	0	7.7E-03
P	7.2E-04	9.4E-06	0	0	7.2E-04
Th + U	9.9E-04	5.2E-04	0	0	3.4E-05
MMC ions (aq)	3.3	1.2E+00	0	0	1.1
REE ions (aq)	3.7E-03	3.0E-04	0	0	2.8E-04
REE(OH)2(s)	0	8.2E-04	0	0	0
MMC(OH)x(s)	0	1.9	0	0	0
REE2O3(s)	0	0	2.8E-02	5.5E-03	0
MMCxOy	0	0	2.0E-03	4.7E-03	0
Water	1000	100	0	0	900
Total	1000	100	5.5E-01	5.5E-01	900



Notes

1. Separator #1 operates at a pH of 3.5. It removes gangue metal (MMC) hydroxide precipitates (Stream 4).

2. Separator #2 operates at a pH of 8.5. It sends an REE hydroxide slurry (Stream 9) to downstream processing.

3. Separator #2 also returns treated water (Stream 8) to the environment.

Special Symbols / Nomenclature

Stream Label

Material or Product Stream

Reference Drawings:

01-A-001/1
02-A-001/2
02-A-001/3

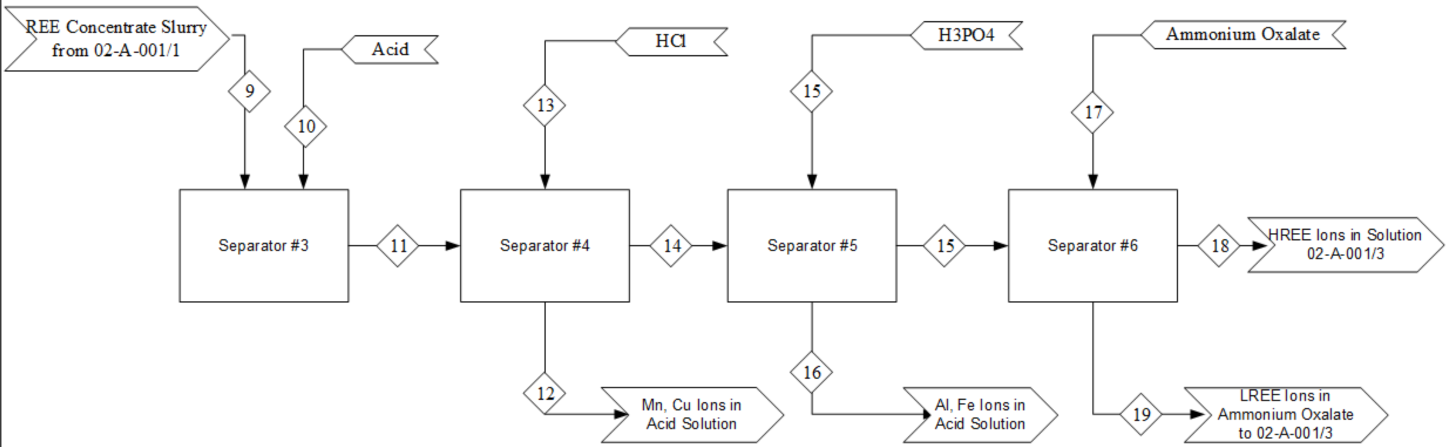
Drawing Type

Block Flow Diagram

Title

Proposal: Ion Exchange:
AMD Pretreatment
Sheet 1 of 3

By/Date	DWG #/ Sheet #	Rev	Page
OD/FE B25	02-A-001/1	1	1



Notes

1. The REE precipitates from pretreatment are separated from insoluble components in the acid leaching of Separator 3.

2. Separator #4 removes approx. 92% of Mn, Cu

3. Separator #5 removes approx. 90% of Al, Fe

4. Separator #6 separates approx. 95% Y, La,Ce,Pr,Nd into HREE stream and 70% Dy,Ho,Er into LREE. 70% of other REEs in HREE, remaining in LREE

Special Symbols / Nomenclature

Stream Label

Material or Product Stream

Reference Drawings:

001-A-001/1
002-A-001/1
002-A-001/3

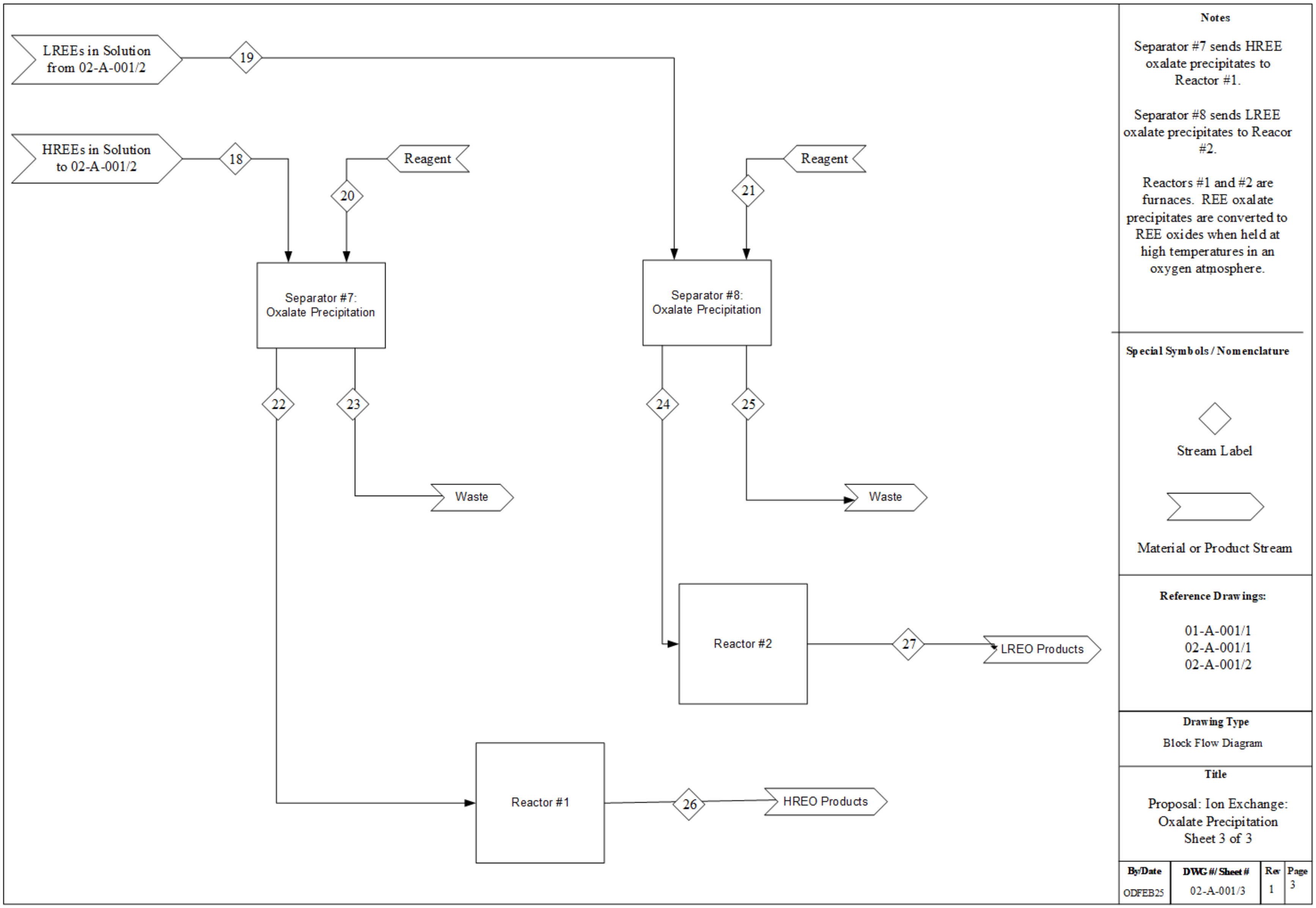
Drawing Type

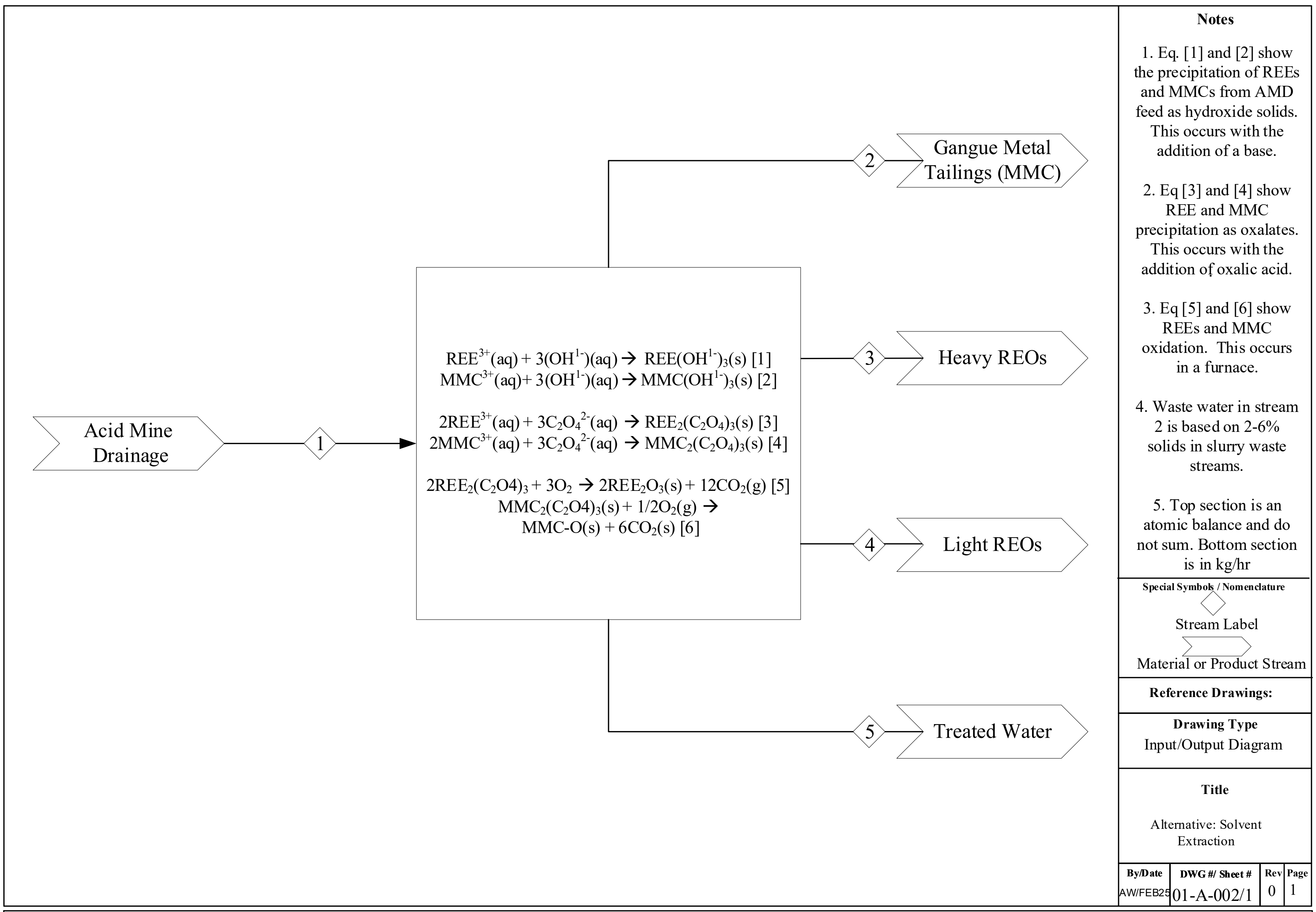
Block Flow Diagram

Title

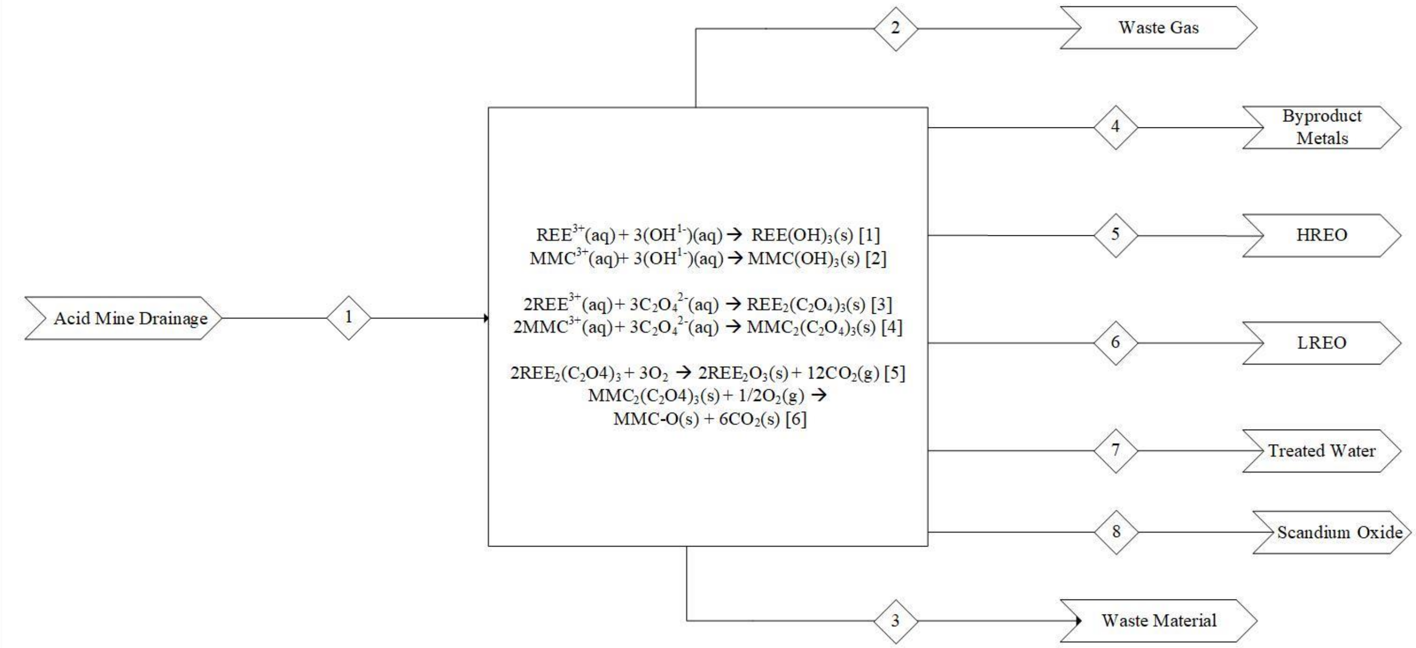
Proposal: Ion Exchange:
Solvent Beds
Sheet 2 of 3

By/Date	DWG #/ Sheet #	Rev	Page
ODFEB25	002-A-001/2	1	2


LREEs in Solution from 02-A-001/2
HREEs in Solution to 02-A-001/2
19
18
20
21
Reagent
Reagent
Separator #7: Oxalate Precipitation
Separator #8: Oxalate Precipitation
22
23
24
25
Waste
Waste
Reactor #2
27
LREO Products
Reactor #1
26
HREO Products
Notes
Separator #7 sends HREE oxalate precipitates to Reactor #1.
Separator #8 sends LREE oxalate precipitates to Reacor #2.
Reactors #1 and #2 are furnaces. REE oxalate precipitates are converted to REE oxides when held at high temperatures in an oxygen atmosphere.
Special Symbols / Nomenclature
Stream Label
Material or Product Stream
Reference Drawings:
01-A-001/1
02-A-001/1
02-A-001/2
Drawing Type
Block Flow Diagram
Title
Proposal: Ion Exchange: Oxalate Precipitation
Sheet 3 of 3
By/Date
ODFEB25
DWG #/ Sheet #
02-A-001/3
Rev
1
Page
3



Component/Stream	1	2	3	4	5
Sc	3.4E-05	1.4E-05	5.6E-06	1.3E-05	9.0E-07
Y	8.8E-04	1.9E-05	1.6E-04	3.8E-04	5.0E-05
La	2.8E-04	1.8E-05	4.5E-05	1.1E-04	3.9E-05
Ce	1.1E-03	6.2E-05	1.9E-04	4.4E-04	1.0E-04
Pr	1.1E-04	3.6E-05	1.5E-05	3.4E-05	6.8E-06
Nd	4.9E-04	1.6E-05	8.9E-05	2.1E-04	4.0E-05
Sm	1.2E-04	1.9E-05	2.1E-05	5.0E-05	7.6E-06
Eu	3.3E-05	1.0E-05	6.2E-06	1.5E-05	1.6E-06
Gd	1.6E-04	1.3E-05	3.0E-05	7.1E-05	1.0E-05
Tb	2.8E-05	1.3E-05	4.7E-06	1.1E-05	3.8E-07
Dy	1.8E-04	1.9E-05	3.2E-05	7.5E-05	8.8E-06
Ho	3.7E-05	9.8E-06	6.9E-06	1.6E-05	2.1E-06
Er	1.1E-04	1.1E-05	2.0E-05	4.7E-05	5.7E-06
Tm	1.6E-05	9.6E-06	2.9E-06	6.8E-06	7.9E-07
Yb	1.0E-04	1.5E-05	1.8E-05	4.2E-05	4.6E-06
Lu	1.5E-05	1.5E-05	1.7E-06	4.1E-06	1.3E-06
Fe (III)	8.2E-01	8.2E-01	3.4E-06	7.9E-06	4.0E-04
Al	2.9E-01	5.6E-02	1.7E-04	4.1E-04	5.6E-03
Na	9.2E-02	6.1E-03	3.2E-05	7.4E-05	4.4E-02
Mg	5.1E-01	1.9E-03	7.0E-05	1.6E-04	4.1E-01
Ca	4.7E-01	4.0E-02	9.1E-05	2.1E-04	3.0E-01
Mn	2.3E-01	1.2E-02	2.9E-05	6.9E-05	1.8E-01
Si	7.6E-02	2.0E-02	3.2E-05	7.4E-05	1.4E-02
Zn	6.3E-01	1.5E-03	4.1E-04	9.7E-04	7.5E-02
Cu	1.4E-01	2.2E-03	9.9E-05	2.3E-04	7.9E-03
F	3.2E-02	9.4E-06	0.0E+00	0.0E+00	3.2E-02
Cl	1.2E-02	9.4E-06	0.0E+00	0.0E+00	1.2E-02
K	7.7E-03	9.4E-06	0.0E+00	0.0E+00	7.7E-03
P	7.2E-04	9.4E-06	0.0E+00	0.0E+00	7.2E-04
Th + U	9.9E-04	5.2E-04	8.9E-05	2.1E-04	3.4E-05
MMC ions (aq)	3.3	1.2	0	0	1.1
REE ions (aq)	3.7E-03	3.0E-04	0	0	2.8E-04
REE(OH)2(s)	0	8.2E-04	0	0	0
MMC(OH)x(s)	0	1.9	0	0	0
REE2O3(s)	0	0	1.9E-03	4.4E-03	0
MMCxOy	0	0	2.0E-03	4.7E-03	0
Water	1000	100	0	0	900
Total	1000	100	2.1E-03	4.9E-03	900



Component/Stream	1	2	3	4	5	6	7	8
Water	990000	100	19000	0	0	0	980000	0
Zinc	630	0	72	0	0	0	240	0
Zn Precipitates	0	0	600	240	0	0	210	0
Aluminum	290	0	15	0	0	0	0.25	0
Al Precipitates	0	0	520	690	0	0	0	0
Copper	140	0	1.8	0	0	0	17	0
Cu Precipitates	0	0	90	71	0	0	0	0
REE	3.7	0	1.3	0	0	0	2.1	0
LREE	0	0	0	0	0.31	0.79	0	0
HREE	0	0	0	0	0.43	0.15	0	0
Other	11000	240	8000	0.29	0.032	0.026	3600	0
Total	1000000	340	28000	1000	0.77	0.97	980000	0

Notes

1. Flow rates are in kg/day

2. Flow rates are rounded to two significant figures

Special Symbols / Nomenclature

Stream Number

Input/Output Information

Reference Drawings:

N/A

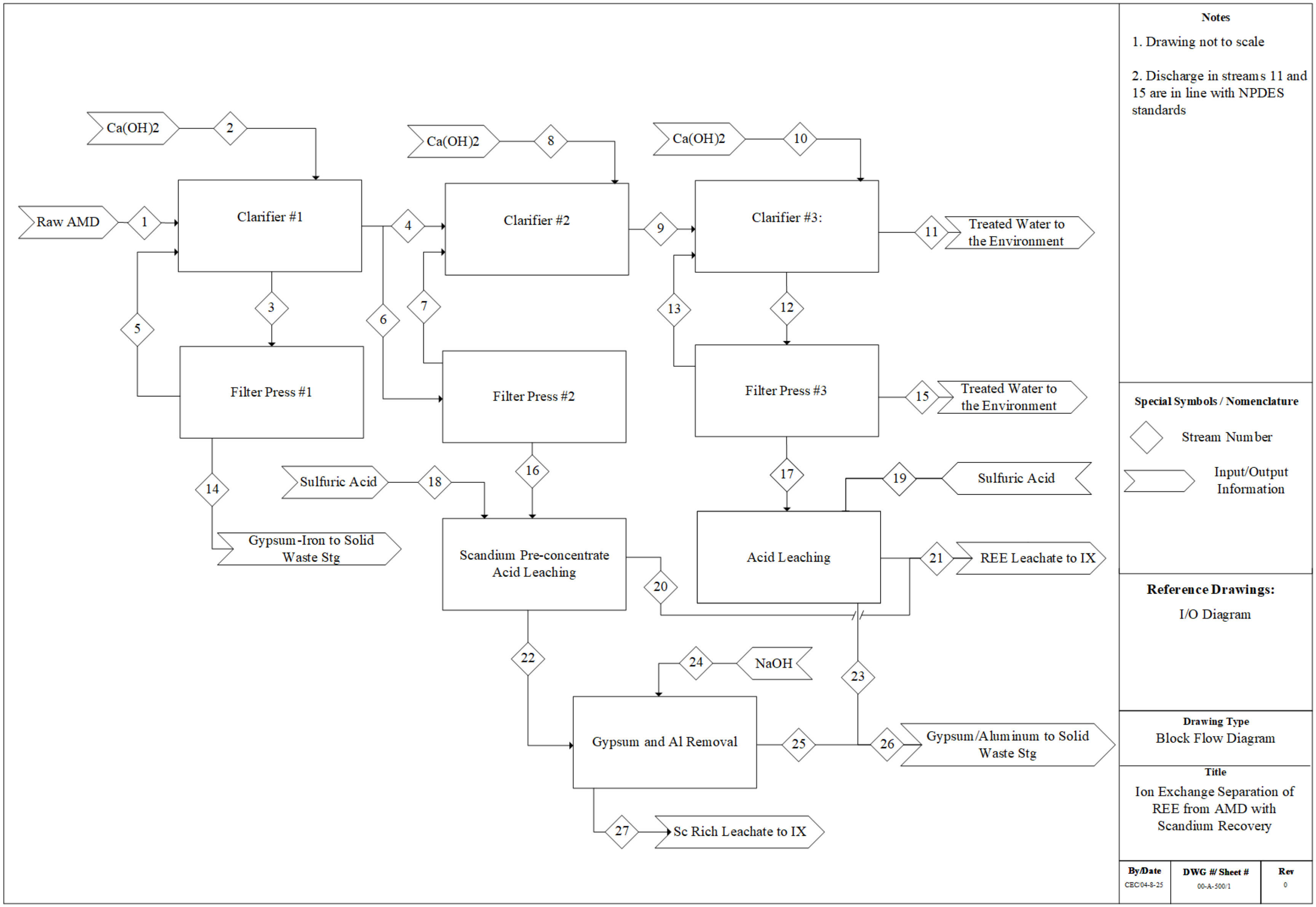
Drawing Type

I/O Diagram

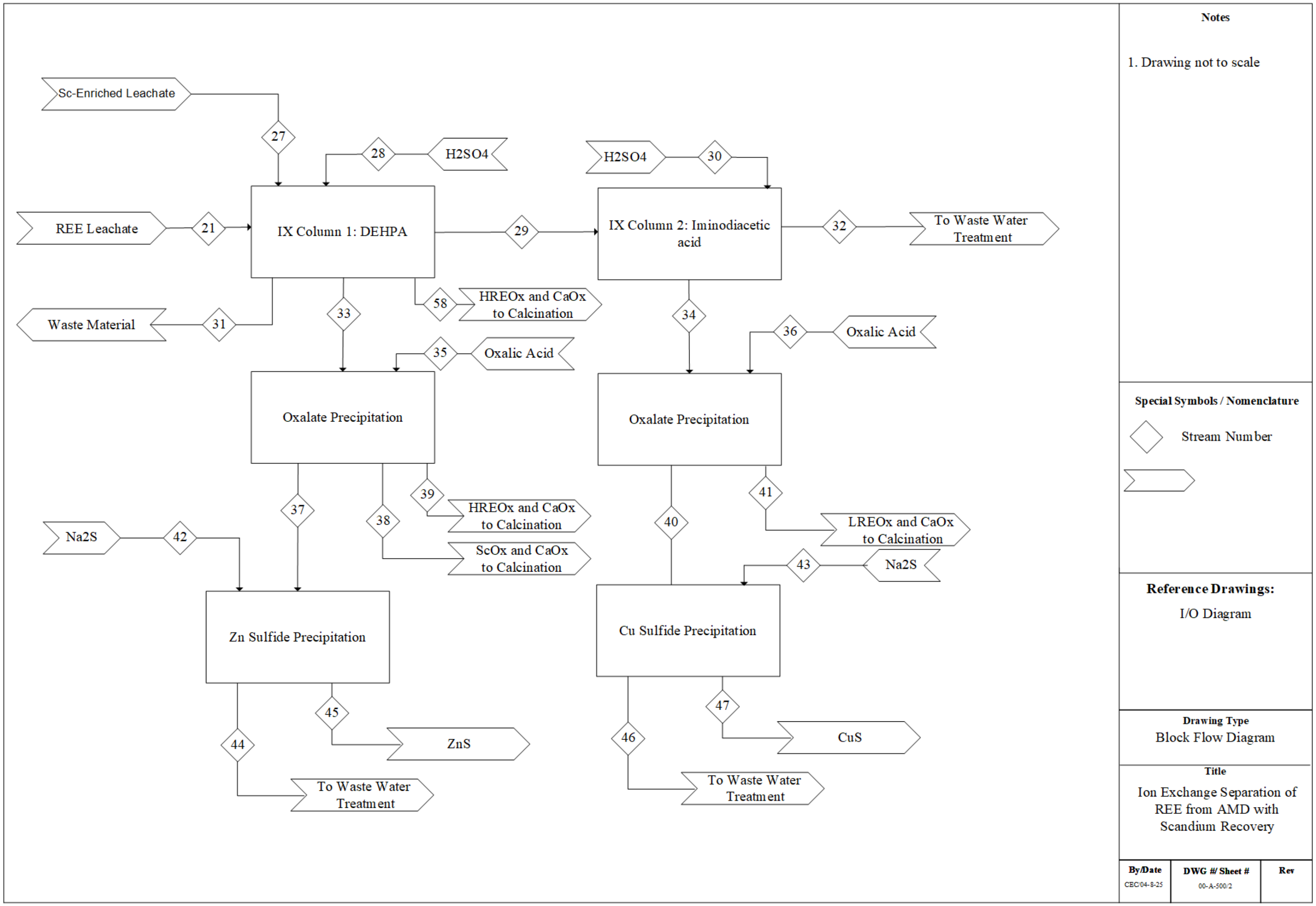
Title

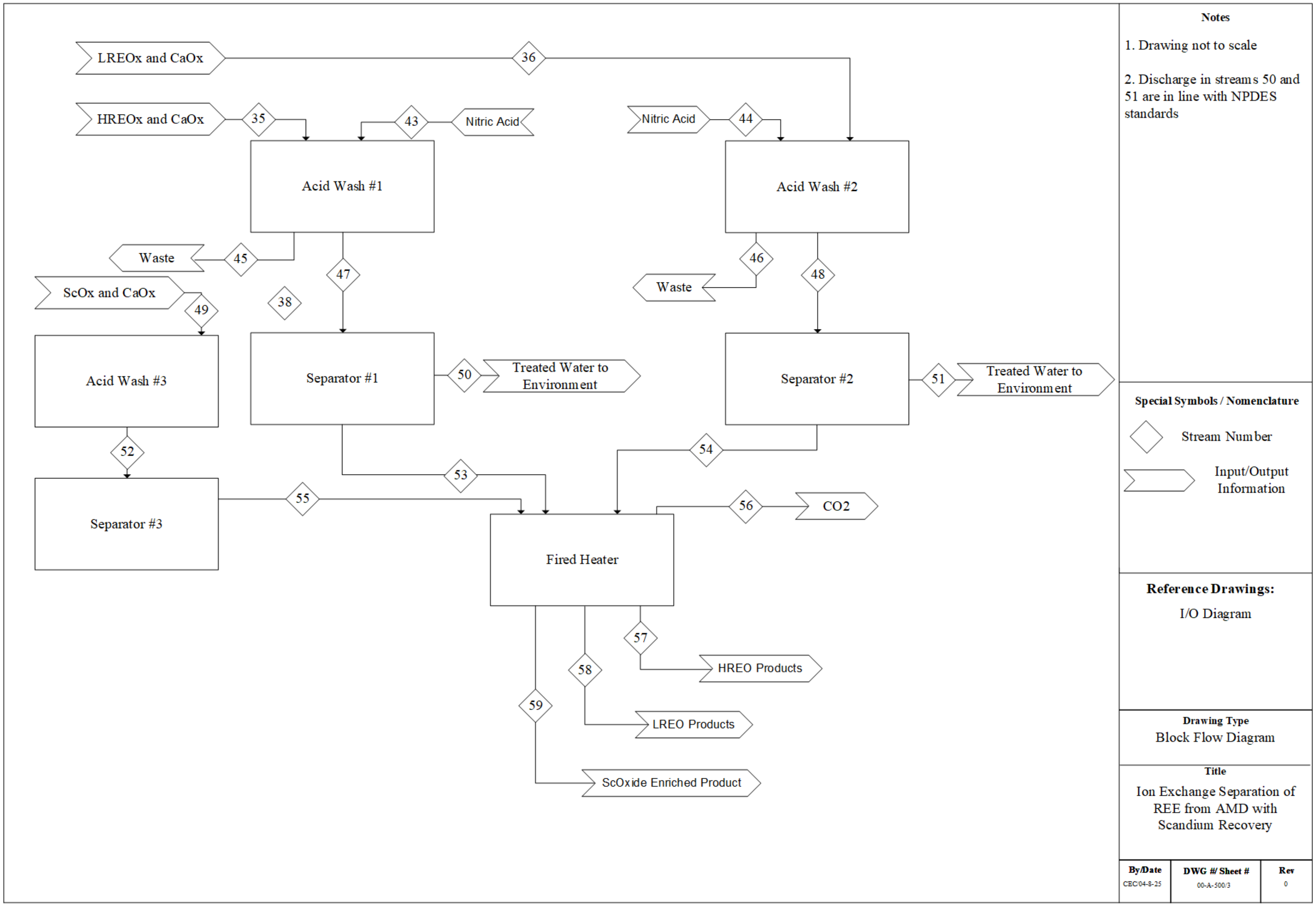
Ion Exchange Separation of REE from AMD with Scandium Recovery

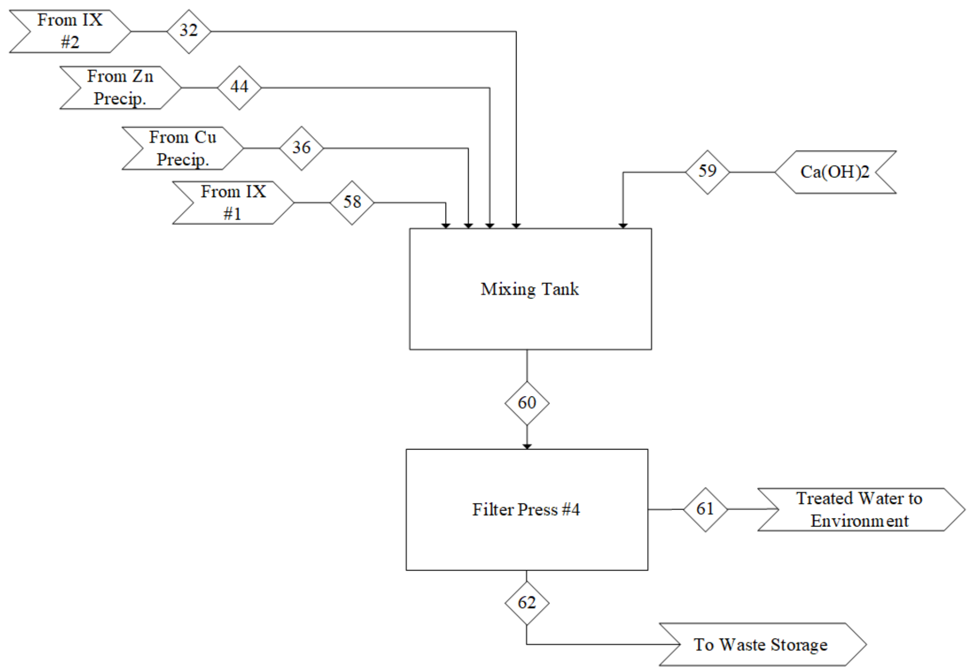
By/Date	DWG #/ Sheet #	Rev
CEC/04-8-25	00-A-600/1	0



Ca(OH)2
2
Ca(OH)2
8
Ca(OH)2
10
Raw AMD
1
Clarifier #1
4
Clarifier #2
9
Clarifier #3:
11
Treated Water to the Environment
3
5
6
7
13
12
Filter Press #1
Filter Press #2
Filter Press #3
15
Treated Water to the Environment
14
Sulfuric Acid
18
16
17
19
Sulfuric Acid
Gypsum-Iron to Solid Waste Stg
Scandium Pre-concentrate Acid Leaching
Acid Leaching
21
REE Leachate to IX
20
22
24
NaOH
23
Gypsum and Al Removal
25
26
Gypsum/Aluminum to Solid Waste Stg
27
Sc Rich Leachate to IX
Notes
1. Drawing not to scale
2. Discharge in streams 11 and 15 are in line with NPDES standards
Special Symbols / Nomenclature
Stream Number
Input/Output Information
Reference Drawings:
I/O Diagram
Drawing Type
Block Flow Diagram
Title
Ion Exchange Separation of REE from AMD with Scandium Recovery
By/Date
CEC/04-8-25
DWG #/ Sheet #
00-A-500/1
Rev
0


Notes
1. Drawing not to scale
Sc-Enriched Leachate
27
28
H2SO4
H2SO4
30
REE Leachate
21
IX Column 1: DEHPA
29
IX Column 2: Iminodiacetic acid
32
To Waste Water Treatment
Waste Material
31
33
58
HREOx and CaOx to Calcination
34
36
Oxalic Acid
35
Oxalic Acid
Oxalate Precipitation
Oxalate Precipitation
37
38
39
HREOx and CaOx to Calcination
ScOx and CaOx to Calcination
41
40
LREOx and CaOx to Calcination
Na2S
42
43
Na2S
Zn Sulfide Precipitation
Cu Sulfide Precipitation
45
44
ZnS
To Waste Water Treatment
47
46
CuS
To Waste Water Treatment
Special Symbols / Nomenclature
Stream Number
Reference Drawings:
I/O Diagram
Drawing Type
Block Flow Diagram
Title
Ion Exchange Separation of REE from AMD with Scandium Recovery
By/Date
CEC/04-8-25
DWG #/ Sheet #
00-A-500/2
Rev


LREOx and CaOx
36
HREOx and CaOx
35
43
Nitric Acid
Nitric Acid
44
Acid Wash #1
Acid Wash #2
Waste
45
47
46
48
Waste
ScOx and CaOx
49
38
Acid Wash #3
Separator #1
50
Treated Water to Environment
Separator #2
51
Treated Water to Environment
52
54
53
Separator #3
55
56
CO2
Fired Heater
57
HREO Products
58
LREO Products
59
ScOxide Enriched Product
Notes
1. Drawing not to scale
2. Discharge in streams 50 and 51 are in line with NPDES standards
Special Symbols / Nomenclature
Stream Number
Input/Output Information
Reference Drawings:
I/O Diagram
Drawing Type
Block Flow Diagram
Title
Ion Exchange Separation of REE from AMD with Scandium Recovery
By/Date
CEC/04-8-25
DWG #/ Sheet #
00-A-500/3
Rev
0



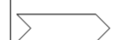
Notes

1. Drawing not to scale

2. Discharge in stream 61 is in line with NPDES standards

Special Symbols / Nomenclature

Stream Number



Input/Output Information

Reference Drawings:

I/O Diagram

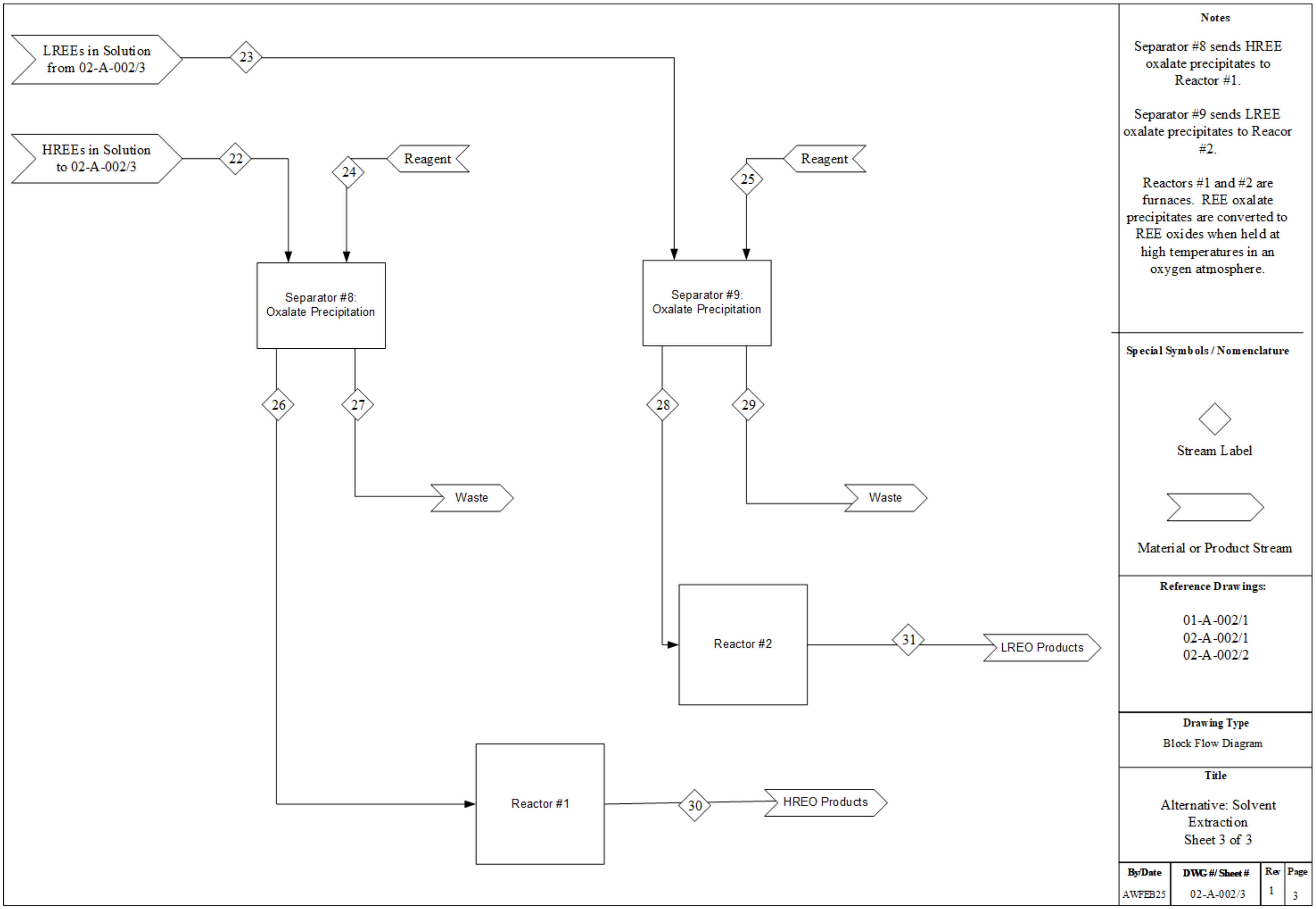
Drawing Type

Block Flow Diagram

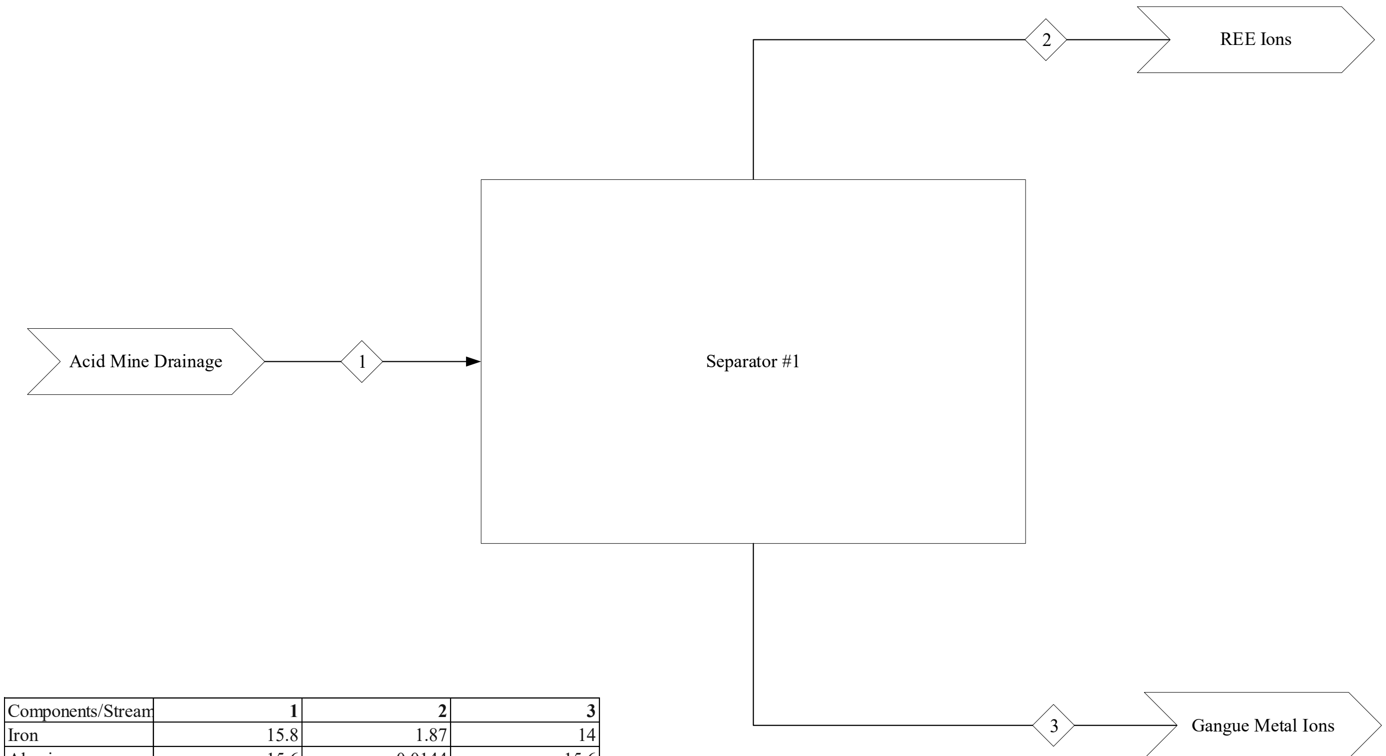
Title

Ion Exchange Separation of REE from AMD with Scandium Recovery

By/Date	DWG #/ Sheet #	Rev
CEC/04-8-25	00-A-500/4	0



LREEs in Solution from 02-A-002/3
HREEs in Solution to 02-A-002/3
23
22
24
25
Reagent
Reagent
Separator #8: Oxalate Precipitation
Separator #9: Oxalate Precipitation
26
27
28
29
Waste
Waste
Reactor #2
31
LREO Products
Reactor #1
30
HREO Products
Notes
Separator #8 sends HREE oxalate precipitates to Reactor #1.
Separator #9 sends LREE oxalate precipitates to Reacor #2.
Reactors #1 and #2 are furnaces. REE oxalate precipitates are converted to REE oxides when held at high temperatures in an oxygen atmosphere.
Special Symbols / Nomenclature
Stream Label
Material or Product Stream
Reference Drawings:
01-A-002/1
02-A-002/1
02-A-002/2
Drawing Type
Block Flow Diagram
Title
Alternative: Solvent Extraction
Sheet 3 of 3
By/Date
AWFEB25
DWG #/ Sheet #
02-A-002/3
Rev
1
Page
3



Components/Stream	1	2	3
Iron	15.8	1.87	14
Aluminum	15.6	0.0144	15.6
Manganese	12.8	0.0118	12.8
Calcium	83	5.74	77
Magnesium	43.3	0.0312	43.3
Sodium	26.1	0.0188	26.1
Potassium	2.18	0.00157	2.18
Chlorine	29.7	0.0214	29.7
Sulfate	722	0	722
REE	0.337	0.239	0.098
Total	951	7.95	943

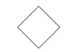
Notes

1. Flow rates are in kg/day.

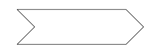
2. Flow rates are rounded to 2 significant figures.

.

Special Symbols / Nomenclature



Stream Label



Material or Product Stream

Reference Drawings:

01-A-001/1
02-A-001/2
02-A-001/3

Drawing Type

Block Flow Diagram

Title

Additional Alternative: Membrane Separation

By/Date	DWG #/ Sheet #	Rev	Page
CCFEB25	02-A-001/1	1	1

Appendix B. List of Assumptions

B.1 Facility-Wide Assumptions

1. All pumps are sized with Aspen assuming 15m of piping, a 10m rise between pieces of equipment, and an efficiency of 75%
2. All compressors assume a polytropic efficiency of 70%.
3. All screw conveyers are purely horizontal and assume a drive efficiency of 88%. Empty conveyor friction is based specified length along with vendor-recommend blade pitch, diameter, loading, and RPM. All conveyors assume a material friction factor of 2. [1]
4. All rotary drum vacuum filters assume the following rules of thumb, independent of solids particle side distribution:
    a. Operating at a pressure of 28 kPa for 10 wt% water carry over in solids scrapings.
    b. 3.4 W/m^2 used as calculation for power needed.
    c. Rules of thumb in Appendix B (Seames 2024) used for required area for rotary vacuum drum filter.

B.2 Feed Conditions

1. REE and major ion composition is averaged from published analysis in 1998, May 2003 and October 2003. Average composition data for the top 15m of Berkley Pitt are used. The feed is assumed to have zero suspended solids.
2. Feed is pumped 100 horizontal meters and 20 vertical meters to reach the sparging vessel.

B.3 Process Area 01: Staged Precipitation

1. Only abiotic oxidation of Fe^{2+} is considered.
2. Zn and Cu are not included in the simulator software package. Mass balances assume the following:
    a. 0% of Zn leaves solution at pH = 4.5 and 78% of Zn precipitates out as $Zn(OH)_2$ at pH 7,
    b. 0% of Cu leaves solution at pH = 4.5 and 42% of Cu precipitates out at pH 7.[2]
3. The sparging tank uses a plate-type fine bubble diffuser. There is 5.5 kPa pressure drop across new diffuser of air sparging tank for air inlet, increases to 11 kPa drop when fouled.[3]

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[1] KWS Conveying Solutions. (n.d.). *Screw conveyor horsepower*. Screw Conveyor Engineering Guide. https://www.kwsmfg.com/engineering-guides/screw-conveyor/screw-conveyor-horsepower/

[2] **Wang, L., et al. (2014).** Selective precipitation of copper and zinc over iron from acid mine drainage by neutralization and sulfidization for recovery. *International Journal of the Society of Materials Engineering for Resources, 20*, 136–140. https://doi.org/10.5188/ijsmer.20.136

[3] **Krampe, J.** (2011). Assessment of diffuser pressure loss on WWTPs in Baden-Württemberg. *Water Science & Technology, 63*(12), 2855–2861

B.4 Process Area 02: Acid Leaching & Neutralization

1. Agitator power consumption assumes an impeller power factor of 1.
2. Density in leach tank, density in neutralization tank.
3. All gypsum, aluminum hydroxide, and alumina silicate precipitates are assumed to be larger than 50 um.[1][1] Frédéric Bard, Essaid Bilal. Semi-batch precipitation of calcium sulfate dihydrate from calcite and sulfuric acid. Carpathian Journal of Earth and Environmental Sciences, 2011, 6 (1), pp.241-250. hal-00542983
4. Solids leaving filter presses J-303 and J-403 contain a negligible amount of liquid and are modeled as bone dry.

B.5 Process Area 03: Ion Exchange

1. Losses discarded for high affinity molecules is 5% for IX 1
2. Holding tank is used between pretreatment and IX to standardize flow rate.
3. Assumed that ions that can't be absorbed by resins at pH~2.7 lost in step 3, DI wash.
4. Density of all inlet and outlet streams assumed to be 1 g/ml.
5. Data for functional groups of the same type are applicable for modeling adsorption for both VP OC 1026 and TP 207 resins.

B.6 Process Area 04: Oxalate Precipitation

1. Assume 80% recovery of REE and 5% recovery of Ca at pH 1.5.
2. If no data available for fouling metals, take average of other class ions for recovery (excluding Ca)
3. In drying step (before reactor) 95% of water is removed.
4. 3X stoichiometric amount of Oxalic acid used.
5. All REOx precipitates aggregate to particle sizes greater than 4 um. Filter presses H-701 and H-702 have 4 um pores and 100% REOx recovery.
6. 99% of calcium reabsorbed at pH 10.
7. Dewatering in filter press assumes 95% of liquid is removed.
8. ZnS and CuS particle size. Make sure particle size distribution corresponds to the process conditions (pH, sulfate system) and agrees with your reaction rate. Please provide references.

B.7 Fired Heater

1. 100% conversion rate of oxalate to oxide.
2. Heat of formation based off of oxalate to calcium oxide for all materials.

3. Fired heater has 75% efficiency.
4. 25% excess air is used.
5. All oxalates remain solid, no pressure difference, allowing for isobaric specific heat capacities.
6. All REEs assumed to have specific heat capacities of bismuth.
7. All remaining water is flashed off into flue gas stream.
8. J-901 and J-902 modelled as tubes for heat transfer purposes.
9. Aspen was used to model flue gas outlet temperature at the acceptable 150 °C

Appendix C. Sample Calculations

C.1 Sample Calculation for Sparging Vessel F-101

The sparging tank is responsible for the oxidation of 400 kg/h of Fe2+ to Fe^{3+}. The sparging tank is defined as an isothermal plug flow reactor (PFR). The oxidation of Fe to Fe3+ has beens shown to follow second-order kinetics. The PFR mass balance for Fe2+ in the sparging tank can therefore be written as:

$$\frac{v_o}{A}\frac{d[Fe^{2+}]}{dx} = -k[Fe^{2+}]^2 \qquad \text{[Eq. 1.1]}$$

Where, $[Fe^{2+}]$ is the concentration of Fe(II) in mols/L,
v_o is the inlet liquid feed rate in L/min,,
A is cross-sectional area of the reactor in dm^2,
x is the length of the reactor in m , and
k is the reaction rate constant in L/mol-min.

This integrates to:

$$\frac{1}{[Fe^{2+}]_{out}} - \frac{1}{[Fe^{2+}]_{in}} = -\frac{k(x \cdot A)}{v_o} = -k\left(\frac{V}{v_o}\right) \qquad \text{[Eq. 1.2]}$$

Yaun (2018) gives an empirical correlation for oxidation rate in sulfate systems assuming second order reaction kinetics.[1] The correlation is based on sparging air flow rate, pH and temperature.

$$\frac{d[Fe^{2+}]}{dt} = -1.48[Fe^{2+}]^2(Q^s)^{0.42}[H_2SO_4]^{-0.36}\exp\left(-\frac{31670}{RT}\right) \qquad \text{[Eq. 1.3]}$$

Where, Q^s represents the maximum flow of sparging air before saturation,

R is 8.314 J/mol-K,

T is defined in K.

[1] Yuan, F. (2018). Study on kinetics of Fe (II) oxidized by air in FeSO4-H2SO4 solutions. *Minerals Engineering, 121*, 164-168. https://doi.org/10.1016/j.mineng.2018.03.013

$[H_2\ SO_4]$is approximated from the pH of in the sparger. The pH of the sparger goes from 2.5 at the entrance to 4.5 at the exit. An average pH value of 3.5 is assumed to calculate *k*.[2]

$[H_2SO_4] = \frac{10^{-3.5}}{2} = 0.0002\ \frac{mol}{L}$ [Eq. 1.3]

Finding Qs is more complicated. Yaun (2018) find Qs as a function of vessel volume in their batch experimental set up:

$Q^s = 0.056V$ [Eq. 1.4]

To translate this to a PFR system, an initial guess for total volume was made and then iterated upon until it converged to the volume produced from the PFR design equation. Assuming a residence time of 11 hours gives:

$$V = 10^6\ \frac{L}{h} \cdot 8.2\ h\ =\ 8.2 \cdot 10^6\ L \qquad \text{[Eq. 1.4]}$$

Plugging this into Eq. 1.4 gives a sparging air saturation flow rate of:

$Q^s = 0.056 \cdot 11 \cdot 10^6 = 460{,}000\ \frac{L[STP]}{min}$ [Eq. 1.5]

Eq. 1.1 can be rearranged with all variable substitutions made to solve for *k*:

$k = -1.48 \cdot 460{,}000^{0.42} \cdot (0.0002)^{-0.36} \cdot \exp\left(-\frac{31670}{8.314 \cdot 177}\right) = -0.010\ \frac{L}{mol-min}$ [Eq. 1.5]

Assuming 99% of the 400 kg/h of Fe2+ is oxidized gives an outlet of 4kg/h of Fe2+. Converting each of these values to a molar concentration basis gives:

$[Fe^{2+}]_{in} = 22\ \frac{mol}{L}$ and $[Fe^{2+}]_{out} = 0.22\ \frac{mol}{L}$.

An inlet flowrate of 1,000,000 L/h is 17,000 L/m.

[2] This justifies introducing Ca(OH)2 directly into the sparging vessel at the entrance of the raw feed. Reducing the acidity of the AMD as quickly as possible increases oxidation rates of Fe(II) to Fe(III) and decreases residence time in the sparging vessel.

Finally, these all substitutions can be made into Eq. 1.2 to solve for sparging vessel volume:

$$\frac{1}{0.22} - \frac{1}{22} = 0.01\left(\frac{V}{10^6}\right)$$

$V = \frac{\left[\frac{1}{0.22} - \frac{1}{22}\right]}{0.01} \cdot 17{,}000 = 8{,}200{,}000\ L$ [Eq 1.6]

The pressure at which sparging air must be delivered is set by the head pressure of AMD in the sparging tank:

$$P_{m-H_2O} = \frac{V}{length \cdot width}$$

[Eq. 1.7]

Assuming a tank length of 50 m and a width of 15 meters:

$P_{m-H_2O} = \frac{8{,}200\ m^3}{50\ m \cdot 15\ m} = 11\ m_{H_2O}$ [Eq. 1.8]

The sparging air compressor must then meet the following functional requirements:

- Discharge pressure > 11m-H2O
- Flow rate > 460,000 L(STP)/min

C.2 Fired Heater Calculations

The sum of LREO and HREO streams are combined for total flow rates. Total duty needed for conversion found from equation below: $Q = \Delta H = \Delta H_f + mC_p\Delta T$ [Eq. 2.3]

Example for La:

$$825.1J - 635090J - 3 \times 393509J + \left(22\frac{kg}{hr} \times 123\frac{J}{kg \times K} \times 780\right) = 3944902\,J$$

[Eq. 2.4]

Once all required energies were calculated per element, everything was summed together. The amount of fuel (methane) was calculated through the following.

$$Q_{tot} = \eta \times \Delta H_c^\circ$$

[Eq. 2.5]

The mass flow of methane is converted to molar flow and then multiplied by the molecular weight and 25% excess air and converted to kilograms. The amount of air was then calculated through the following:

$$\frac{\dot{m} \times MW \times .25_{excess}}{1000} = \dot{M}\frac{kg_{O_2}}{hr}$$

[Eq. 2.6]

Air was then calculated as a percentage of oxygen in the air, assuming 21%.

The composition of the flare stream was based off the amount of water flashed off and the stoichiometric value of the combustion reaction. The amount of water was calculated as follows:

$$H_2O\ \frac{kg}{hr} = \frac{2\ mol \times 18\frac{g}{mol}}{1000\frac{g}{kg}} + 25\ \frac{kg}{hr}$$

[Eq. 2.7]

C.3 Sample Calculation for Agitator Power Consumption

Power consumption by the acid leaching vessel agitator M-301 and by the neutralization vessel agitator M-401 are calculated from the following equation:

$$P = N_p \cdot \rho \cdot N^3 \cdot D^3 \quad \text{[Eq. 4.1]}$$

Where,

P is power consumption in Watts,
N_p is a dimensionless factor correlated to flow regime and impeller geometry,
ρ is fluid density in kg/m^3,
N is impeller RPM, and
D is impeller diameter.

N_p is assumed to take the value of 10 based on an average value for laminar flow[34],
N is assumed to be 1 RPM, and D is assumed to be 1 m less than the vessel diameter.

Example calculation for M-301:

$$P = 10 \cdot 1{,}500 \cdot 1^3 \cdot 2^3 = 120{,}000\ W$$

C. 5 Sample Calculation for Bed Volume (BV) for a Fixed Bed Adsorption Column

The bed volumes for a fixed bed adsorption column are defined by the void space that fluid can flow through due to porosity of resin bed. BV is defined by the equation:

$$BV = \pi \cdot r_{ID}^2 \cdot \ \varepsilon \cdot h_B$$

[Eq. 19]

Where, BV is bed volume in m^3,

π is the pi constant (3.14159265),

r_{ID} is the internal diameter of the column,

ε is the porosity associated with resin of choice, and

h_B is the bed height.

Eg. Calculation for F-301A/B:

BV is bed volume in m^3,

π is the pi constant (3.14159265),

r_{ID} is the internal diameter of the column (ID=1m) ,

ε is the porosity associated with resin of choice ($\varepsilon = .55$), and

h_B is the bed height ($h_B = 3m$).

$$BV = 3.14 \cdot (1m)^2 \ \cdot \ 0.55 \cdot 3m \quad = 5m^3$$

C. 6 Sample Calculation for Pressure Drop Across Fixed Bed Adsorption Column

The pressure drop for a fixed bed adsorption column can be defined by the Ergun equation. The Ergun equation defines pressure losses by the summation of viscous and kinetic energy losses:

$$\frac{\Delta P}{\Delta z} = \frac{150\mu v_s(1-\varepsilon)^2}{\varphi^2 D_p^2 \varepsilon^3} + \frac{1.75\rho v_s^2(1-\varepsilon)^2}{\varphi D_p \varepsilon^3}$$

[Eq. 20]

Where, ΔP is the change in pressure across column in Pa,

Δz is the height of the resin bed in m,

μ is the viscosity of the fluid in Pa·s^{-1},

v_s is the superficial velocity in m/sec,

ε is the porosity of the resin,

φ is the sphericity of the resin bead,

ρ is the density of the resin kg/m^3, and

D_p is the average diameter of resin particles in m.

Eg. Calculation for Resin VP OC 1026:

ΔP is the change in pressure across column in Pa,

Δz is the height of the resin bed (10m),

μ is the viscosity of the fluid 0.0001002 Pa·s^{-1} – assumed viscosity of water for inlet stream to ion exchange column,

v_s is the superficial velocity in m/sec is the volumetric flow rate divided by the surface area of the resin bed, $v_s = \frac{Q}{m^2}$,

ε is the porosity of the resin – assumed to be 0.55 which is common for macroporous chelating resins,

φ is the sphericity of the resin bead ($\varphi = 1$ for spherical beads),

ρ is the density of the resin, 970 kg/m^3, and

D_p is the average diameter of resin particles, 0.00095 m.

$$\Delta P = \frac{10m}{1}\left(\left(\frac{150 \cdot 1e^{-4}(1.1e^{-3})^2(1-.55)^2 \cdot kg}{1^2(9.5e^{-4})^2 \cdot 0.55^2 m^2 s^2}\right) + \left(\frac{1.75 \cdot 970(1.1e^{-3})^2(1-.55)^2 \cdot kg}{9.5e^{-4} \cdot 0.55^2 \cdot m^2 s^2}\right)\right)$$

$$\Delta P = 139.25\ Pa$$

C.7 Sample Calculation for Pump Sizing

All pumps were sized according to the following equation:

$$P = \frac{(\rho \times g \times Q \times h)}{\eta}$$

Where:

P = Power (in watts, W)

ρ = Density of fluid (kg/m^3)

G = Acceleration due to gravity (approximately 9.81 m/s^2 on Earth)

Q = Flow rate (in m^3/s)

H = Head (m)

η = Pump efficiency (no unit)

Ex. Calc

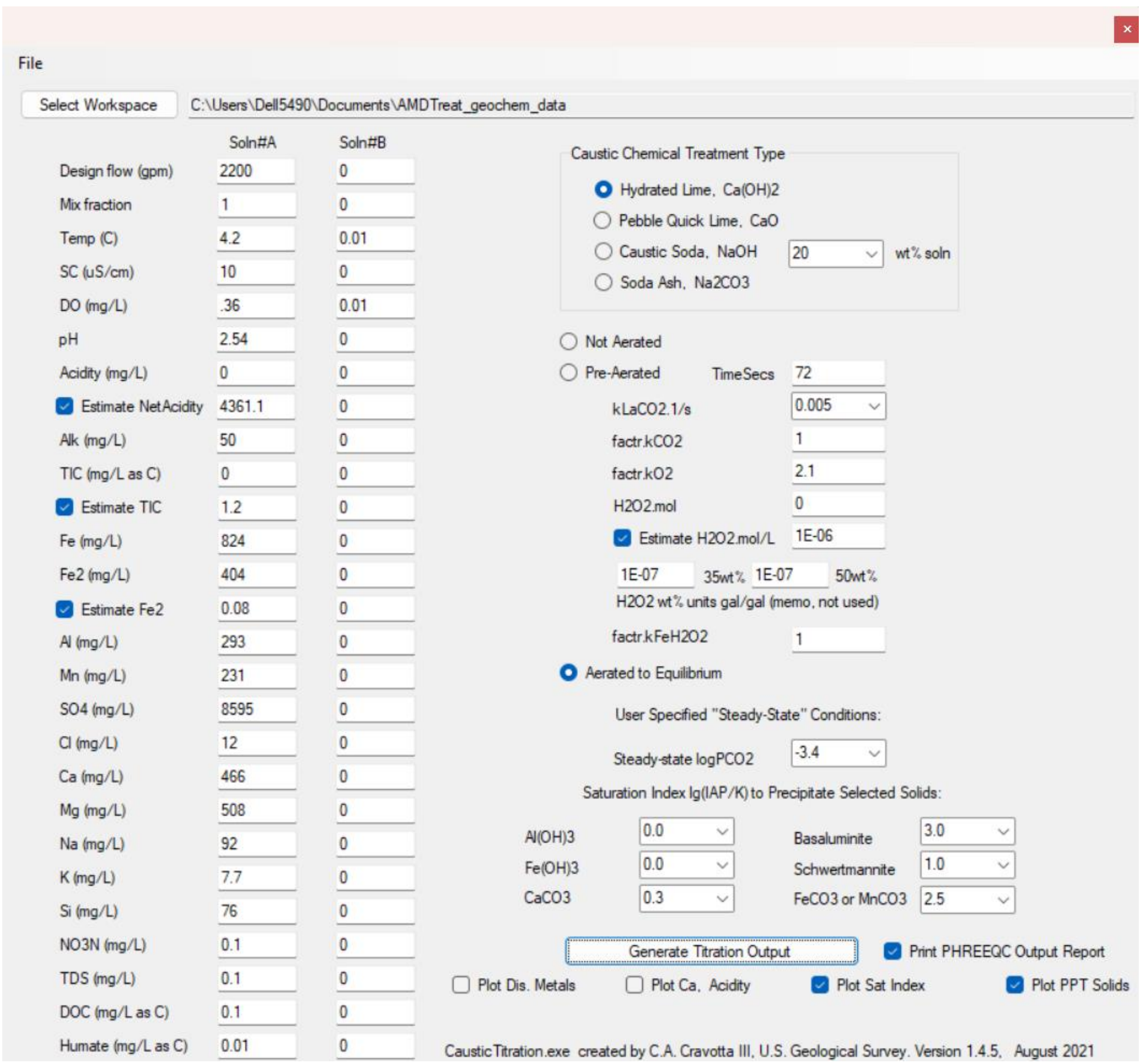
$$Stream\,1 \;=\; \frac{1{,}000{,}000\,\dfrac{kg}{hr} \times 9.81\,\dfrac{m}{s^2} \times 100m}{.7 \times 60\,\dfrac{hr}{min} \times 60\,\dfrac{min}{s} \times 1000\,\dfrac{W}{kW}} = 389.3\,kW$$

Appendix D. Simulation Input/Output

D.1 PHREEQC Simulation I/O for Staged Precipitation Processes

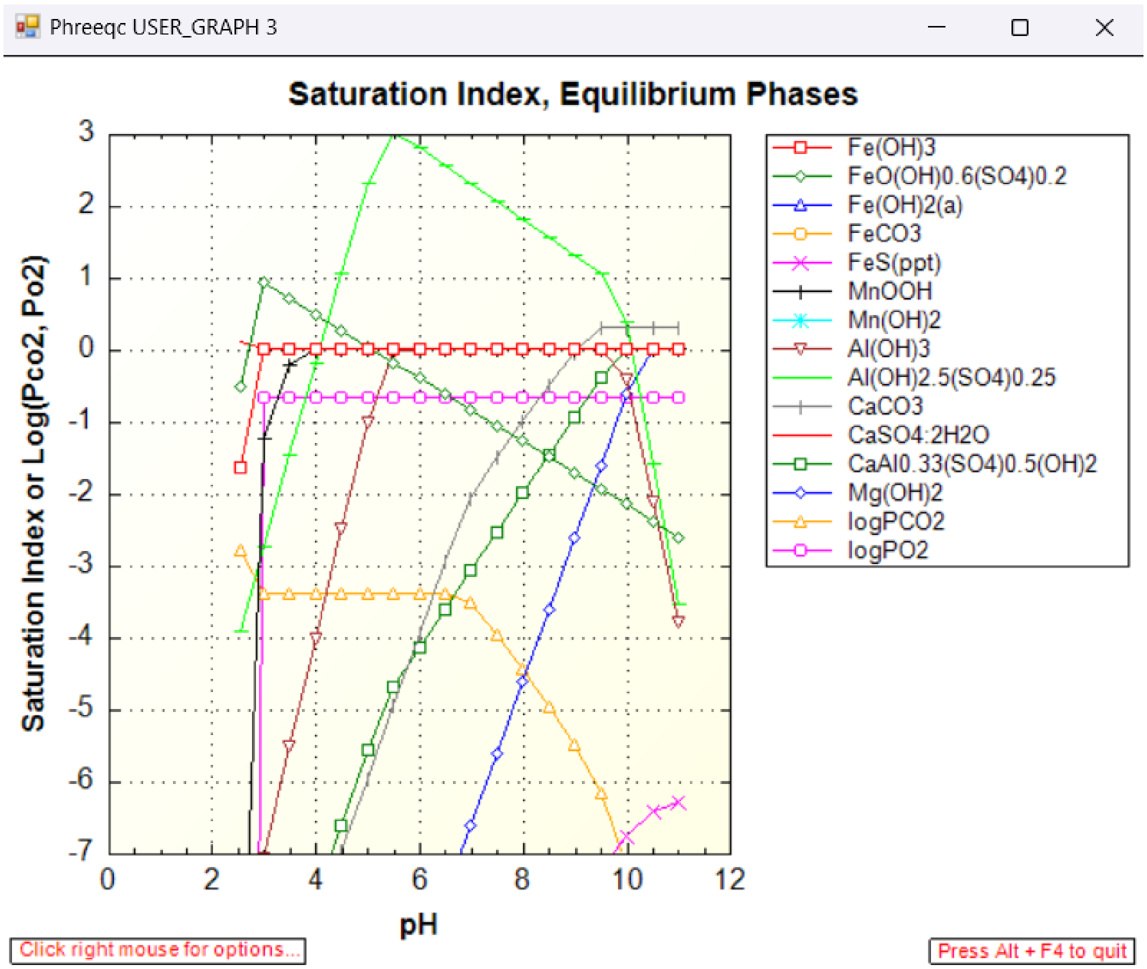
D.1.1 PHREEQC Titration Module Input

PHREEQC Titration Module Input was used to simulate the addition of hydrated lime to AMD feed. Equilibrium aeration is assumed. Default net acidity and alkalinity values are used. Saturation indices set to zero indicate that supersaturation of those species will not be permitted in the simulation: precipitation will occur at saturation.

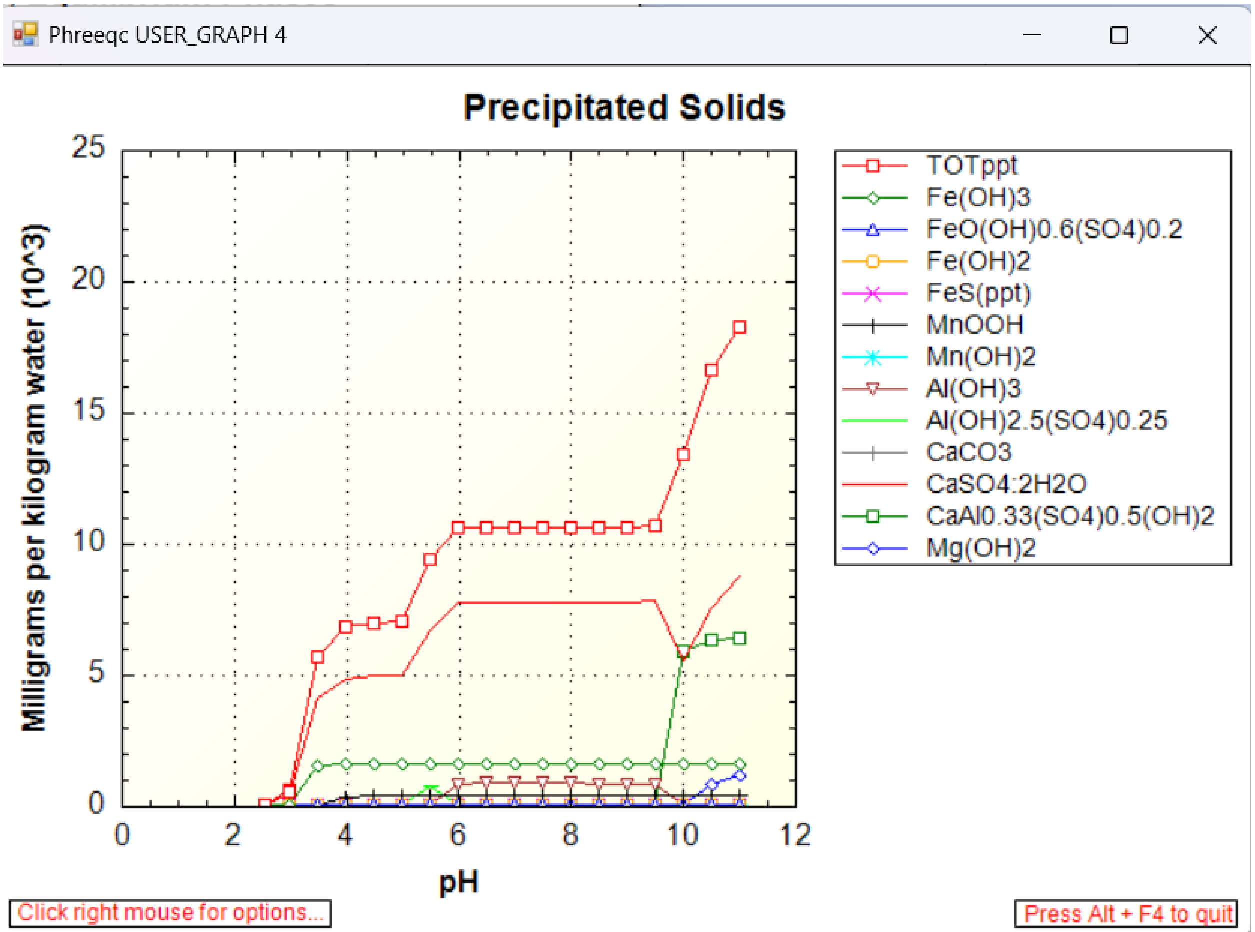


D.1.2 PHREEQC Titration Module Output

Saturation indices can be seen to approach zero for all precipitating species. Negative values represent an undersaturated state, positive values a supersaturated state, and 0 values an equilibrium saturation state at which precipitation is occurring. Basaluminite ($Al(OH)_{2.5}(SO4)_{0.25}$) is the only species that is permitted to reach supersaturation.



In addition, generated plots of soldi precipitates shows that, excluding gypsum, Me(OH)x precipitates dominate all solids formation. Nearly all the Fe present precipitates as Fe(OH)3 at pH < 4. Al(OH)3 precipitation reaches its maximum at pH 6. Note all values are totals, and not cumulative, allowing for the plateau seen in Total Precipitates (TOTppt) 6 < pH < 10.

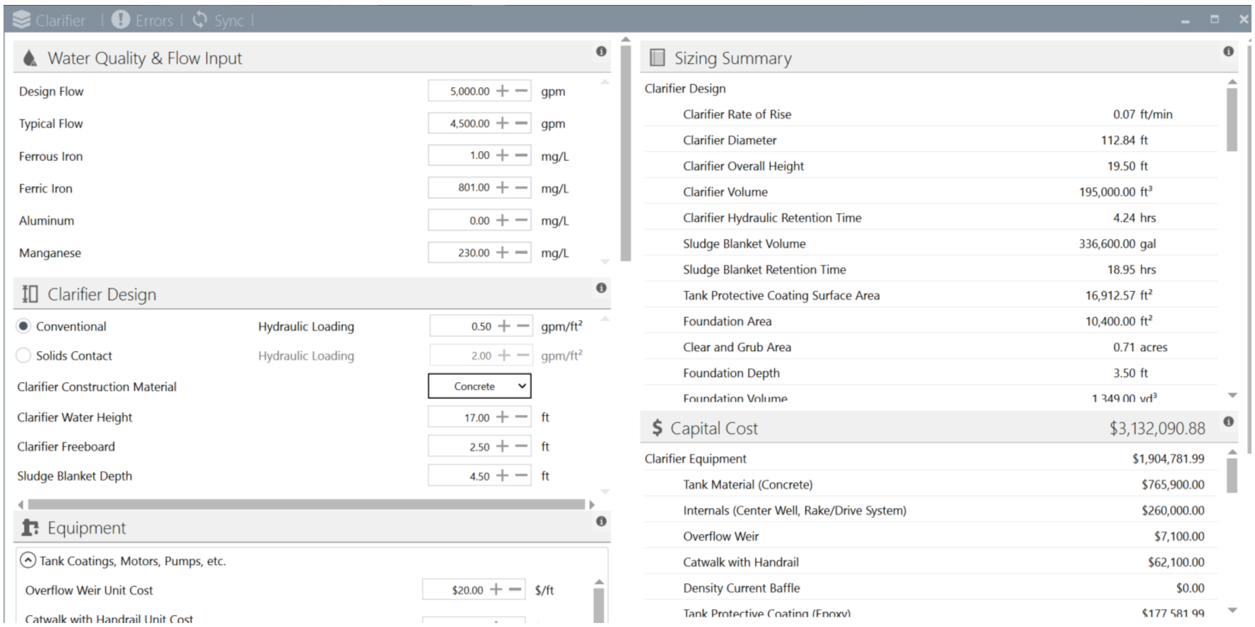


D.2 AMDTreat Simulation I/O for Stage 1 and 2 Clarifiers Capital Cost

AMDTreat6.0 Beta software was published in 2024. Are prices are assumed to have a 2024 basis date.

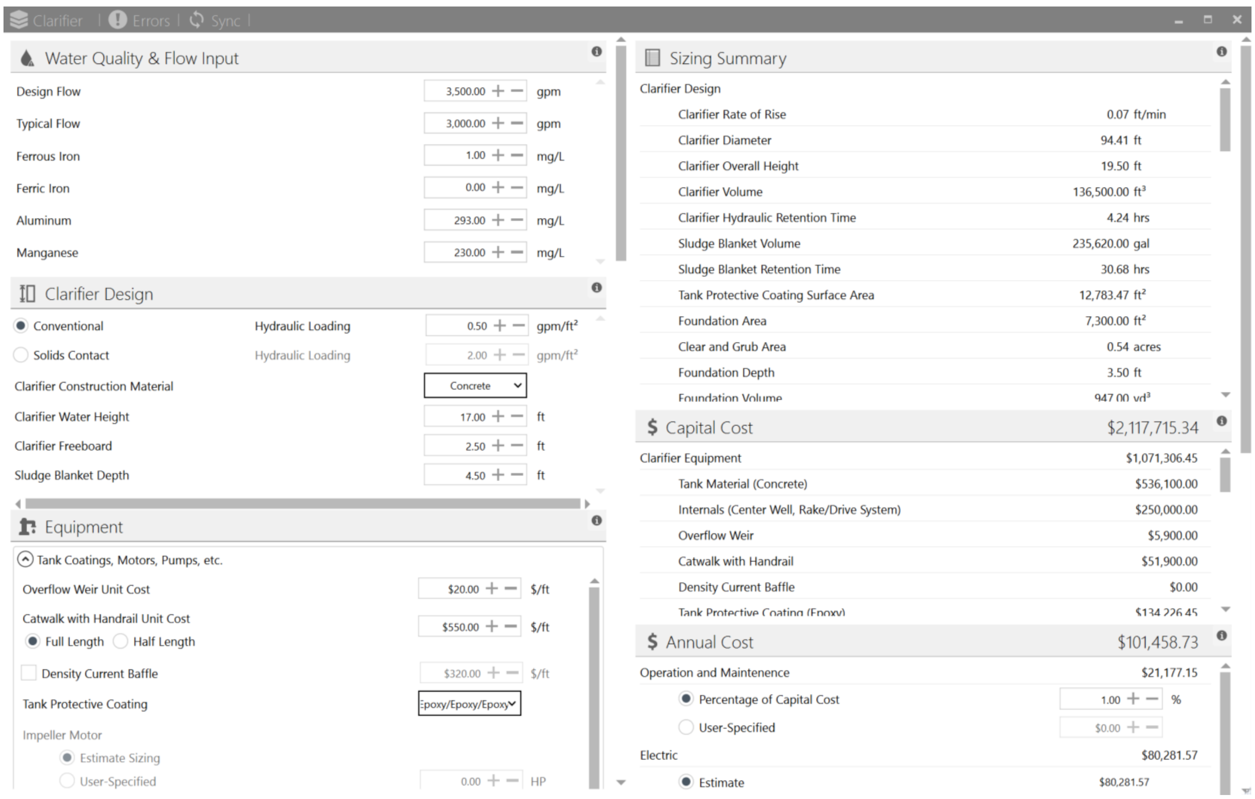
D.2.1 AMDTreat I/O for Stage 1 Clarifier H-101

The simulation input reflects the feed composition into Stage 1 Clarifier H-101, which assumes 99% oxidation of the ferrous ion into the ferric ion in the upstream Sparging Vessel F-101. MOC is input as epoxy-lined concrete. The cost of the sludge recirculation system was subtracted from the total capital cost because this was priced externally based.



D.2.2 AMDTreat I/O for Stage 2 Clarifier H-103

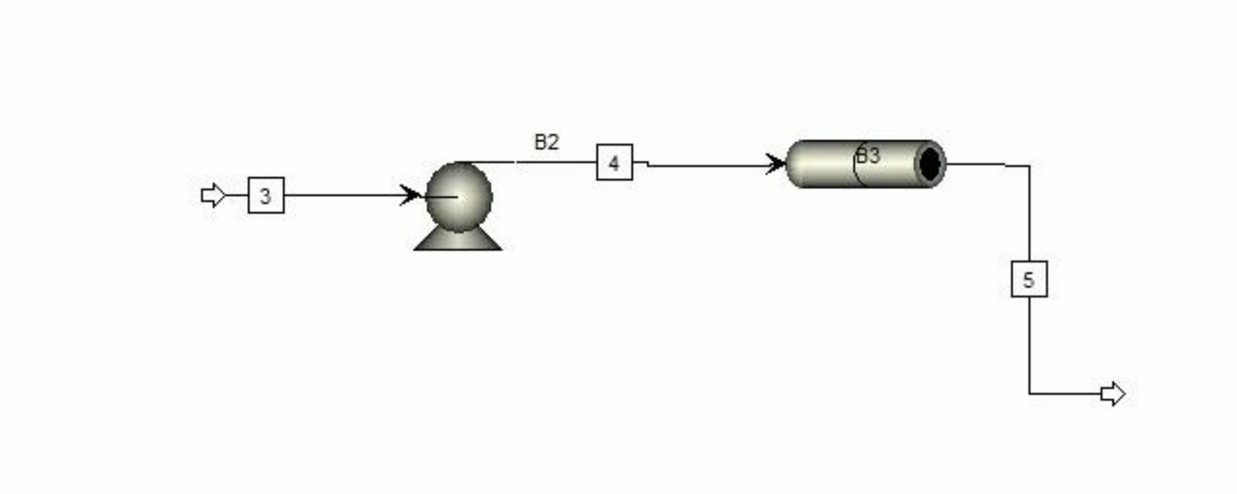
The simulation input reflects the feed composition into H-103, which assumes near complete iron removal in Stage 1 Clarifier H-101. MOC is input as epoxy-lined concrete. The cost of the sludge recirculation system was subtracted from the total capital cost because this was priced externally based.

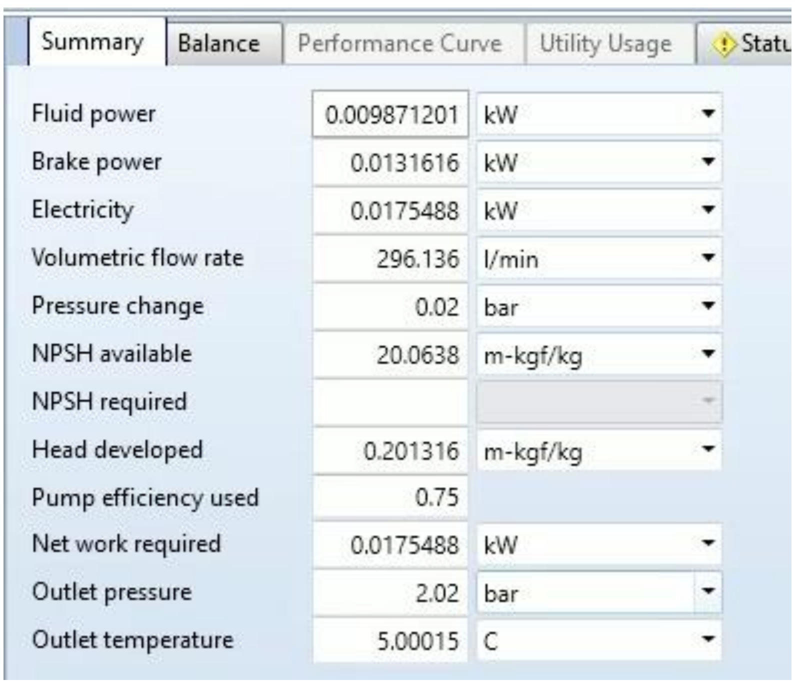


D.3 Aspen Simulation Pumps, Blower and Heat Exhanger

D.3.1 Pumps

Pumps were modeled in aspen based around the requirements of the area: incoming pressure, stream composition, head pressure, height change, and required pressure outlet.



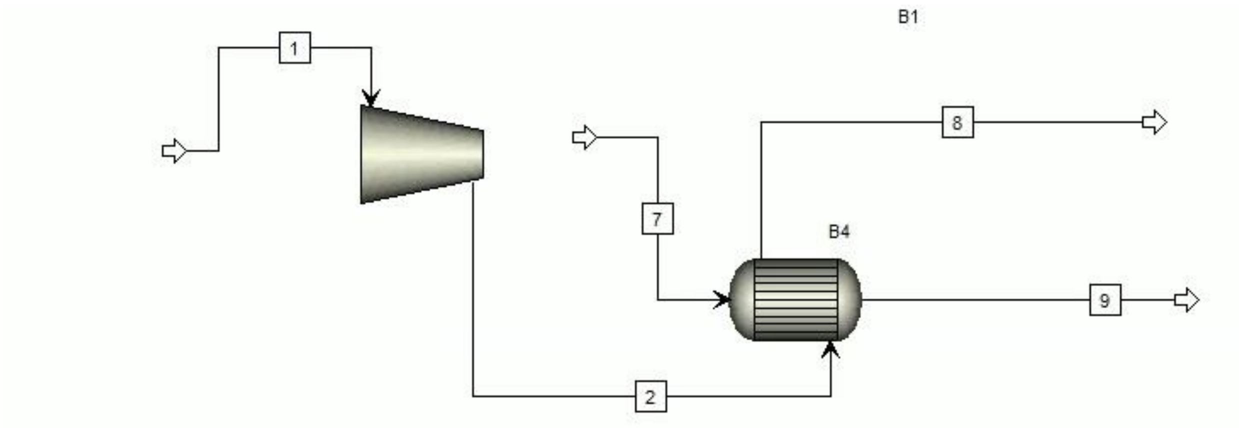


Summary | Balance | Performance Curve | Utility Usage | Statu

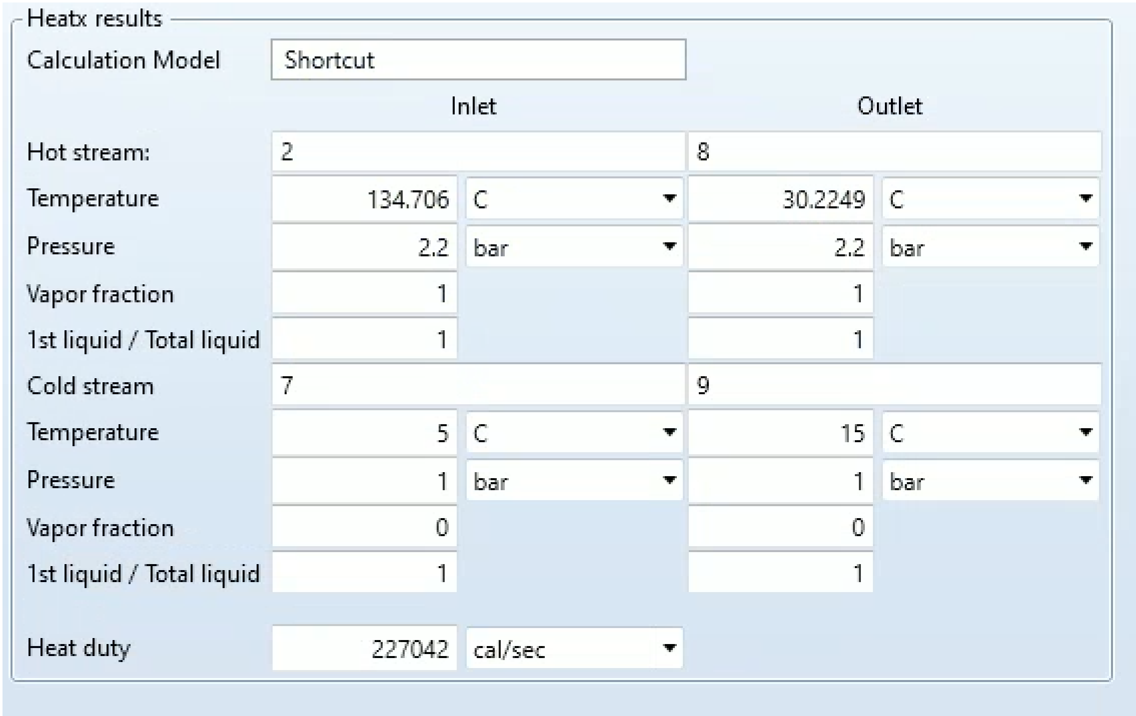
Fluid power	0.009871201	kW
Brake power	0.0131616	kW
Electricity	0.0175488	kW
Volumetric flow rate	296.136	l/min
Pressure change	0.02	bar
NPSH available	20.0638	m-kgf/kg
NPSH required		
Head developed	0.201316	m-kgf/kg
Pump efficiency used	0.75	
Net work required	0.0175488	kW
Outlet pressure	2.02	bar
Outlet temperature	5.00015	C

D.3.1 Blower and Heat Exchanger

The blower for the air sparging unit in the first process area was sized in Aspen using the required flowrate for complete conversion and production of enough pressure to overcome the head pressure in the tank. Thermal transfer between the air and the liquid in the sparging tank was modeled using a heat exchanger.



Compressor model	ASME polytropic
Phase calculations	Vapor phase calculation
Indicated horsepower	949.663 kW
Brake horsepower	949.663 kW
Net work required	949.663 kW
Power loss	0 kW
Efficiency	0.75
Mechanical efficiency	1
Outlet pressure	2.2 bar
Outlet temperature	134.706 C
Isentropic outlet temperature	105.722 C
Vapor fraction	1



Heatx results

Calculation Model	Shortcut			
	Inlet		Outlet	
Hot stream:	2		8	
Temperature	134.706	C	30.2249	C
Pressure	2.2	bar	2.2	bar
Vapor fraction	1		1	
1st liquid / Total liquid	1		1	
Cold stream	7		9	
Temperature	5	C	15	C
Pressure	1	bar	1	bar
Vapor fraction	0		0	
1st liquid / Total liquid	1		1	
Heat duty	227042	cal/sec		

Appendix E. Revenue Sensitivity to Post-Covid Historical Price Volatility

Revenue Upper Bound Calculations: Post-Covid Historical Market Highs									**Max Price**	**HREO Rev, $/h**	**LREO Rev, $/h**
	HREO, kg/h	**LREO, kg/h**	**Jan-22**	**Jul-22**	**Jan-23**	**Jul-23**	**Jan-24**	**Jul-24**	**Feb-25**	**Feb-25**	**Feb-25**
	Numerical Data, x-axis:		**22**	**22.5**	**23**	**23.5**	**24**	**24.5**	**25.1**	**25.1**	**25.1**
Sc Oxide	0.445	0.000075	900	880	860	840	861.11	800	900	400.5	0.0300375
Y Oxide	0.023	0.038	10	9.5	9	8.5	8	7.5	10	0.23	0.00874
La Oxide	0.47	0.096	0.69	0.67	0.65	0.64	0.56	0.55	0.69	0.3243	0.0311328
Ce Oxide	0.026	0.024	1.53	1.67	1.61	1.58	1.44	1.44	1.67	0.04342	0.00104208
Pr Oxide	0.015	0.014	100	90.28	80.56	70.83	53.46	50	100	1.5	0.021
Nd Oxide	0.0072	0.27	120	110.42	95.83	85	56.94	52	120	0.864	0.23328
Sm Oxide	0.0041	0.062	2.5	2.5	2.5	2.5	2.5	2.5	2.5	0.01025	0.0006355
Eu Oxide	0.021	0.018	30	28	27	26	25	24	30	0.63	0.01134
Gd Oxide	0.0069	0.079	40	38	36	34	32	30	40	0.276	0.021804
Tb Oxide	0.09	0.0052	1200	1100	1000	900	759.46	700	1200	108	0.5616
Dy Oxide	0.016	0.023	350	320	300	280	222.22	210	350	5.6	0.1288
Ho Oxide	0.062	0.0038	100	95	90	85	80	75	100	6.2	0.02356
Er Oxide	0.0098	0.0067	50	48	46	44	42	40	50	0.49	0.003283
Tm Oxide	0.054	0.0005	0	0	0	0	0	0	0	0	0
Yb Oxide	0.0043	0.0027	20	19	18	17	16	15	20	0.086	0.0002322
Lu Oxide	1.5	0.00022	600	580	560	540	520	500	600	900	0.198
									95% op factor annual revenue:	**11856802.54**	**10606.28148**
									"Mixed basket" reduction factor of 60%:	**7114081.523**	**6363.768888**
									Total Annual REO Revenue:	***7120445.292***	
									Byproduct Metal Annual Revenue:	***4,583,757.60***	
									Total Annual Revenue:	11,704,202.89	

Appendix E. Revenue Sensitivity to Post-Covid Historical Price Volatility

Revenue Lower Bound Calculations: Post-Covid Historical Market Lows									**Min Price**	**HREO Rev, $/h**	**LREO Rev, $/h**
	HREO, kg/h	**LREO, kg/h**	**Jan-22**	**Jul-22**	**Jan-23**	**Jul-23**	**Jan-24**	**Jul-24**	**Feb-25**	**Feb-25**	**Feb-25**
	Numerical Data, x-axis:		**22**	**22.5**	**23**	**23.5**	**24**	**24.5**	**25.1**	**25.1**	**25.1**
Sc Oxide	0.445	0.000075	900	880	860	840	861.11	800	800	356	0.0267
Y Oxide	0.023	0.038	10	9.5	9	8.5	8	7.5	7.5	0.1725	0.006555
La Oxide	0.47	0.096	0.69	0.67	0.65	0.64	0.56	0.55	0.55	0.2585	0.024816
Ce Oxide	0.026	0.024	1.53	1.67	1.61	1.58	1.44	1.44	1.44	0.03744	0.00089856
Pr Oxide	0.015	0.014	100	90.28	80.56	70.83	53.46	50	50	0.75	0.0105
Nd Oxide	0.0072	0.27	120	110.42	95.83	85	56.94	52	52	0.3744	0.101088
Sm Oxide	0.0041	0.062	2.5	2.5	2.5	2.5	2.5	2.5	2.5	0.01025	0.0006355
Eu Oxide	0.021	0.018	30	28	27	26	25	24	24	0.504	0.009072
Gd Oxide	0.0069	0.079	40	38	36	34	32	30	30	0.207	0.016353
Tb Oxide	0.09	0.0052	1200	1100	1000	900	759.46	700	700	63	0.3276
Dy Oxide	0.016	0.023	350	320	300	280	222.22	210	210	3.36	0.07728
Ho Oxide	0.062	0.0038	100	95	90	85	80	75	75	4.65	0.01767
Er Oxide	0.0098	0.0067	50	48	46	44	42	40	40	0.392	0.0026264
Tm Oxide	0.054	0.0005	0	0	0	0	0	0	0	0	0
Yb Oxide	0.0043	0.0027	20	19	18	17	16	15	15	0.0645	0.00017415
Lu Oxide	1.5	0.00022	600	580	560	540	520	500	500	750	0.165
									95% op factor annual revenue:	**9818134.07**	**6549.152772**
									"Mixed basket" reduction factor of 60%:	**5890880.442**	**3929.491663**
									Total Annual REO Revenue:	***5894809.934***	
									Byproduct Metal Annual Revenue:	***4,583,757.60***	
									Total Annual Revenue:	10,478,567.53	

Appendix E. Revenue Sensitivity to Post-Covid Historical Price Volatility

Revenue Basis Calculations (Trend Price at Basis Date)									**Trend price**	**HREO Rev, $/h**	**LREO Rev, $/h**
	HREO, kg/h	**LREO, kg/h**	**Jan-22**	**Jul-22**	**Jan-23**	**Jul-23**	**Jan-24**	**Jul-24**	**Feb-25**	**Feb-25**	**Feb-25**
	Numerical Data, x-axis:		**22**	**22.5**	**23**	**23.5**	**24**	**24.5**	**25.1**	**25.1**	**25.1**
Sc Oxide	0.445	0.000075	900	880	860	840	861.11	800	795.8894095	354.1707872	0.026562809
Y Oxide	0.023	0.038	10	9.5	9	8.5	8	7.5	6.9	0.1587	0.0060306
La Oxide	0.47	0.096	0.69	0.67	0.65	0.64	0.56	0.55	0.51672381	0.24286019	0.023314578
Ce Oxide	0.026	0.024	1.53	1.67	1.61	1.58	1.44	1.44	1.421314286	0.036954171	0.0008869
Pr Oxide	0.015	0.014	100	90.28	80.56	70.83	53.46	50	35.0539619	0.525809429	0.007361332
Nd Oxide	0.0072	0.27	120	110.42	95.83	85	56.94	52	32.64979048	0.235078491	0.063471193
Sm Oxide	0.0041	0.062	2.5	2.5	2.5	2.5	2.5	2.5	2.5	0.01025	0.0006355
Eu Oxide	0.021	0.018	30	28	27	26	25	24	22.43809524	0.4712	0.0084816
Gd Oxide	0.0069	0.079	40	38	36	34	32	30	27.6	0.19044	0.01504476
Tb Oxide	0.09	0.0052	1200	1100	1000	900	759.46	700	560.3863619	50.43477257	0.262260817
Dy Oxide	0.016	0.023	350	320	300	280	222.22	210	173.2454857	2.771927771	0.063754339
Ho Oxide	0.062	0.0038	100	95	90	85	80	75	69	4.278	0.0162564
Er Oxide	0.0098	0.0067	50	48	46	44	42	40	37.6	0.36848	0.002468816
Tm Oxide	0.054	0.0005	0	0	0	0	0	0	0	0	0
Yb Oxide	0.0043	0.0027	20	19	18	17	16	15	13.8	0.05934	0.000160218
Lu Oxide	1.5	0.00022	600	580	560	540	520	500	476	714	0.15708
									95% op factor annual revenue:	**9386838.18**	**5440.672794**
									"Mixed basket" reduction factor of 60%:	**5632102.908**	**3264.403676**
									Total Annual REO Revenue:	***5635367.312***	
									Byproduct Metal Annual Revenue:	4,583,757.60	
									Total Annual Revenue:	10,219,124.91	

Appendix E. Revenue Sensitivity to Post-Covid Historical Price Volatility

Lanthanum, Cerium, Praseodymium, Neodymium, Terbium, Dysprosium: Prices for 2022–2024 are derived from Shanghai Metals Market (SMM) reports, converted from CNY/mt to USD/kg at current exchange rate.
Scandium: March 2024 price from Statista (6,200 CNY/kg ≈ 861.11 USD/kg).
Samarium, Europium, Gadolinium, Holmium, Erbium, Ytterbium, Lutetium, Yttrium: No pre-Covid historical data available; prices Jan 22 - Jul 24 cells were populated by Statista yrs 2020-2021 average values for these metals and multiplying by a downward trend factor for overall market given by nstitute for Rare Earths and Metals.

Appendix F. Vendor Correspondence

EMAIL CONVERSATION RECORD

ORIGINATOR: Andrea Williams

RECEIVER: COMPANY: WesTech

EMAIL: tdumbaugh@westechwater.com PHONE: 515.268.8549

DATE OF CONTACT: 4/21/2025 TIME OF CONTACT: 9:09 pm MT

REASON FOR CONTACT: Paste Thickener and Vacuum Drum Filter Pricing

>The following email was sent to price H-102 and H-402 in Process Areas 01 and 04:

I am following up on a request for more information about WesTech's Deep Bed Paste Thickener and Rotary Vacuum Drum Filter.

My design team and I are seeking a rough budgetary estimate of these products. This is for a grassroots facility in Butte, Montana that will treat acid mine drainage while extracting critical minerals.

The paste thickener will have 1,000 gpm feed with 20,000 ppm of gypsum and metal hydroxides. We need to settle 13,000 lbs/hour of gypsum, iron and aluminum hydroxides, dry basis. We want to maximum % solids in the underflow.

The vacuum drum filter will have 60 gpm of 10,000 ppm zinc sulfide. We need to settle 500 lbs/hour of zinc sulfide, dry basis. Expected particle size distribution is 1- 20 microns. We also want to maximizing % solids in this stream.

>Tom Dumbaugh, WesTech Regional Engineering Sales Manager, replied:

Here's some info based on your request for the paste thickener and the vacuum drum filter. In our understanding this is for your college project.

PASTE THICKENER

A rough price for a deep bed paste thickener (8m dia.) to handle this would be $1.25 million. That is just the mechanism and tank, no install, concrete work, or freight is included.

We don't think this is a great application for a paste thickener. We've included a Paste Thickener brochure for your reference.

VACUUM DRUM FILTER

WesTech is pleased to offer budget pricing for a 304 SS precoat vacuum drum filter capable of filtering 60 gpm of 10,000 ppm zinc sulfide waste stream.

Drum filter package includes the following.

(1) 6ft diameter x 8ft face drum filter with 151sq ft of filtration area, partially submerged in a zinc sulfate slurry vat.

Swing type agitator in the slurry vat to keep feed solids in suspension.

Automatic Precoat cutting knife advance which cuts off 0.005" of precoat per drum revolution.

Vacuum receiver with filtrate pump.

Vacuum pump.

Precoat mix tank with mixer.

Control panel 480VAC/3Ph/60Hz

All components mounted on a common skid.

Items not Included: Field Erection, Stairs/Ladder, Cover/Roof, Freight and Precoat.

Budget Price (USD): $639,360.00

Also included is a reference drawing and brochure.

>Andrea Williams sent a follow-up question:

Why would this be a poor application for the deep bed paste thickener?

>Tom Dumbaugh replied:

Hi Andrea: Here's some thoughts from a process engineer that is familiar with AMD applications.

A couple items to note:

- *We made a number of assumptions based on the application*
    - *Used an average particle size distribution (68% passing 400 mesh), solids SG (2.71), and liquid SG (1.2) for gypsum. Based on the PSD for the downstream drum filter, they likely have something much finer than what I input.*

- *There would be the potential to make paste from this sample, but given how fine it is, the underflow concentration would likely be <35 wt%*
  - *If it did make paste, the target yield stress would be mild at 30-50 Pa.*
- *I'm not sure what the objective of their project is, but is cost is at all relative, a DeepBed Paste Thickener is as expensive as they come.*

It sounds like a normal basin thickener would be a better option. That price would be around 60% of the cost of a paste thickener.

EMAIL CONVERSATION RECORD

ORIGINATOR: Andrea Williams

RECEIVER: Landon Weeden   COMPANY: KWS

EMAIL: landon.weeden@kwsmfg.com   PHONE: (817) 295-2247

DATE OF CONTACT: 5/04/2025   TIME OF CONTACT: 12:40 am MT

REASON FOR CONTACT: Screw Conveyor Pricing

>The following form was submitted to price screw conveyor J-101 in Process Area 01:

Company: University of North Dakota

Questions: We are seeking a rough estimate of a 100 ft screw conveyor for 60% solids mix of gypsum and iron hydroxide tailings.

Design Conditions: 12,000 lb/hr   Density: 70 lb/ft3

Loading: 30%   Material Factor: 2

>Landon Weeden, KWS Regional Engineering Sales Manager, replied:

Thanks for reaching out to KWS.
I made a few assumptions but please see blow budgetary price. Let me know if you have any questions.

12" Screw Conveyor

- *12" dia. x 100'-0" long screw conveyor (CS) fitted with 10HP screw conveyor drive: designed to convey 12,000 lbs/hr at 70 lbs/cuft of mix of gypsum and iron oxide tailings.*
  Budgetary Cost (+/- 20%): $85,000.00