Study Level Design of Thorium Oxide Production

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Report on:

**Study Level Design of Thorium Oxide Production**

CBE 488

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1.0 Introduction

This report documents a study to provide a study-level design and analysis of a process to produce a primary product of ThO₂ (thorium oxide) and byproducts of rare earth element (REE) oxides. The given raw material is North American monazite ore with the composition below:

<table>
<thead>
<tr>
<th>Composition</th>
<th>wt.%</th>
</tr>
</thead>
<tbody>
<tr>
<td>La₂O₃</td>
<td>16.34</td>
</tr>
<tr>
<td>Ce₂O₃</td>
<td>33.52</td>
</tr>
<tr>
<td>ThO₂</td>
<td>5.40</td>
</tr>
<tr>
<td>P₂O₅</td>
<td>29.00</td>
</tr>
<tr>
<td>Nd₂O₃</td>
<td>13.75</td>
</tr>
<tr>
<td>SiO₂</td>
<td>1.55</td>
</tr>
<tr>
<td>U₃O₈</td>
<td>0.15</td>
</tr>
<tr>
<td>Total</td>
<td>100.00</td>
</tr>
</tbody>
</table>

Table 1.1: Monazite Ore Composition

The process is designed to handle an average feed rate of 1000 kg of monazite ore per hour; the economics are based on 2019 values. This process flowsheet was developed using OLI:Flowsheet. The objective of this study is to produce thorium oxide and REE oxides at purity specifications of 98% or better. In order to obtain this objective, we must produce a reasonable design that considers worker safety, environmental interactions, and long-term sustainability.
Furthermore, we have provided estimates of capital and manufacturing costs to determine the economic feasibility of the project.

The proposed process design is based on information obtained from previous studies that have explored a sulfuric acid digestion method to process the monazite feedstock and have yielded promising results. Studies presenting information useful in the development of this manuscript include: Rodliyah 2015, Amer 2013, Berry 2018, Helaly 2017, Habashi 2013, and Vijayalakshmi 2000. While all of the aforementioned articles begin with a sulfuric acid digestion, the method of precipitation of thorium varies such as utilizing pH, thermal properties, and introducing oxalic acid to the system. Other studies have suggested alternative methods or variations on the sulfuric acid digest that impact the movement of uranium in the system. The economic and safety impact of uranium in the products is a chief component to our investigation.

Monazite ore has been found in many parts of the world including India, China, and the United States. While unprocessed ore is not useful, monazite can be refined in order to produce rare earth element (REE) oxides that are used in high technology applications such as permanent magnets, optoelectronics, and superconductors (Rodliyah, 2015). Thorium-based nuclear power generation is envisioned for the future. In such a nuclear power generation, Th-232 is the fertile nuclear reactor material; when bombarded with neutrons Th-232 becomes U-233 which is a fissionable nuclear fuel. The abundance of thorium in nature along with its potential as an energy source motivates the exploration of extraction of thorium in monazite ore (World Nuclear Association, 2017).

This study-level design will illustrate and explain the flowsheet design of the process, estimate the equipment and operating costs associated with the flow sheet, assess the economic
potential of the manufacturing process, provide assumptions and background information that was used in the economic calculations, and finally discuss the feasibility for this project and recommendations going forward. This design activity is supported by the Electric Power Research Institute and the Department of Chemical and Biomolecular Engineering at the University of Tennessee, Knoxville.

2.0 Synthesis Information for Process

2.1 Overall Process Schematic

![Figure 2.1: A block flow diagram depicting the proposed process.](image)

2.2 Process Chemistry

The principal reaction for each reactor is summarized below. Note that reactions occurring reactors involved in waste treatment have been omitted for brevity as they are primarily acid-base reactions.
Initial Reactions

**Pre-feed Reactions: Oxides to Orthophosphates**

\[
\text{La}_2\text{O}_3 + \text{P}_2\text{O}_5 \rightarrow 2 \text{LaPO}_4
\]

\[
\text{Ce}_2\text{O}_3 + \text{P}_2\text{O}_5 \rightarrow 2 \text{CePO}_4
\]

\[
\text{Nd}_2\text{O}_3 + \text{P}_2\text{O}_5 \rightarrow 2 \text{NdPO}_4
\]

**R-100 Reaction: Thorium Precipitation**

\[
\text{ThO}_2 + 2 \text{H}_2\text{SO}_4 \rightarrow \text{Th(SO}_4)_2 + 2 \text{H}_2\text{O}
\]

**Alpha Branch Reaction:**

**R-170 Reaction: Thorium Conversion**

\[
\text{Th(SO}_4)_2 + 4 \text{NH}_4\text{OH} \rightarrow \text{ThO}_2 + 2 (\text{NH}_4)_2\text{SO}_4 + 2 \text{H}_2\text{O}
\]

**Beta Branch Reactions:**

**R-110 Reaction: Rare Earth Precipitation**

\[
2 \text{LaPO}_4 + 3 \text{H}_2\text{SO}_4 \rightarrow \text{La}_2(\text{SO}_4)_3 + 3 \text{H}_2\text{O} + \text{P}_2\text{O}_5
\]

\[
2 \text{CePO}_4 + 3\text{H}_2\text{SO}_4 \rightarrow \text{Ce}_2(\text{SO}_4)_3 + 3 \text{H}_2\text{O} + \text{P}_2\text{O}_5
\]

\[
\text{UO}_2 + 2 \text{H}_2\text{SO}_4 \rightarrow \text{U(SO}_4)_2 + 2 \text{H}_2\text{O}
\]

**R-120 Reaction: Neodymium Precipitation**

\[
\text{NdPO}_4 + 2 \text{H}_2\text{O} \rightarrow \text{NdPO}_4 + 2 \text{H}_2\text{O}
\]

**R-130 Reaction: Neodymium Conversion**

\[
\text{NdPO}_4 + 2 \text{H}_2\text{O} + 3 \text{NaOH} \rightarrow \text{Nd(OH)}_3 + \text{Na}_3(\text{PO}_4) + 2 \text{H}_2\text{O}
\]

**R-150 Reaction: Neodymium Calcination**

\[
2 \text{Nd(OH)}_3 \rightarrow \text{Nd}_2\text{O}_3 + 3 \text{H}_2\text{O}
\]
**R-160 Reaction: Rare Earth Conversion**

La\(_2\)(SO\(_4\))\(_3\) + 2 NaOH → 2 La(OH)\(_3\) + Na\(_2\)(SO\(_4\))

Ce\(_2\)(SO\(_4\))\(_3\) + 2 NaOH → 2 Ce(OH)\(_3\) + Na\(_2\)(SO\(_4\))

### 2.3 Literature Summary

As monazite is a good source of rare earth elements and thorium, monazite processing has been thoroughly researched for decades. Monazite has been found in a variety of regions including Egypt (Helaly, 2016) and (Saleh, 1970), Indonesia (Rodliyah, 2015), and China (Zhu, 2015). Initially, thorium was viewed as a hazardous byproduct of processing, however recent studies have shown that thorium can be utilized as a fissile fuel in place of uranium. The amount of energy that can be generated from one ton of thorium can produce the same amount of energy as two hundred tons of uranium (Helaly, 2016). The given composition of monazite shows that is abundant in rare earth elements as well as thorium. Furthermore, the sample also contains less than two percent silicon dioxide and uranium; impurities that can impact the selling price of the finished product.

While many researchers agree that strong acid or strong base digestions are the most effective methods for monazite processing, work is still being done to determine the optimal conditions to acquire the most pure products. One method of extraction is by precipitating a thorium phosphate from a monazite-sulfate solution, however there is a wide variety of poorly defined precipitate forms of thorium that can be generated. Furthermore, the gelatinous nature of this precipitate makes filtration much more complicated. With these constraints in mind, it is much simpler to precipitate thorium alongside the rare earth sulfates. This method also allows for the recycling of sulfuric acid which can reduce operating costs (Helaly, 2016).
After digestion, there are multiple paths ways to continue processing the materials. An uncommon method is using oxalate to precipitate thorium and lanthanum from the solution. This method would require further processing in order to achieve pure thorium as well as neodymium in order to achieve the maximum economic potential. Considering all of the additional separations and research this method would require, it is not likely to be a viable option for maximizing profit (Habashi, 2013).

A thorough process description for monazite processing, complete with chemistry and reaction conditions by Berry et al., detailed a process that employs a longer digestion time of four hours in the acid digestion step at a slightly higher temperature of 250 °C (Berry et al., 2018). The increased digestion time would call for a larger and more expensive reactor, though this would lead to a more pure product. In contrast, a paper by Logamtanah details a digestion step that occurs over 150 minutes at 220 °C. The shorter digestion time and lower temperature lead to a marginally less pure product, however it is still of acceptable quality, equipment costs are lower, and the process is slightly safer (Rodliyah, 2015).

With uranium often being present in monazite ore, additional separations should be incorporated in order to have pure products that are free of unwanted radioactive material. Separation of uranium in these systems has proven to be somewhat difficult. One possible method for complete extraction of uranium was proposed by Zhu et al. in which thorium and uranium are separated from rare earth elements. Uranium and thorium are then capable of being separated by utilizing N-alkyl carboxylic acid amines (R-CO-R’) with varying R’ groups that were described in (Preston and du Preez, 1995). While this method would separate the radioactive material from the REEs, the separation of uranium and thorium is nearly impossible.
as the materials produced as mass-separating agents are uncommon and not produced commercially. Nonetheless, it may be of future significance to research production of these amines (Zhu et al., 2015).

2.4 Basic Process Economics

In addition to designing a process capable of extracting thorium oxide from monazite ore, an economic analysis will be performed to determine the viability of the proposed process. All of the equipment was designed and economic calculations were performed following the guidelines presented in Ulrich. Following these parameters, we were able to determine the net profit after taxes, raw material purchasing costs, labor and operating expenses, waste treatment, and all other miscellaneous costs associated with a chemical plant. A list of equipment and more detailed economics can be found in appendices C and D respectively.

3.0 Methods of Approach

3.1 Flow Diagram

Monazite ore is fed as a slurry into a reactor with concentrated sulfuric acid. The reactor operates at 220 °C and 15 atm in order to convert the rare earth oxides into rare earth sulfates. After the initial acid digestion, the components are cooled in a heat exchanger before being sent to a filter. In the filter, the process splits into the alpha branch which produces thorium oxide and a beta branch which produces neodymium oxide. In the alpha branch, solid silicon dioxide and thorium sulfate are separated by density in a component splitter to isolate the thorium. Thorium sulfate is then fed into a reactor alongside ammonium hydroxide in order to reconvert it back to
thorium oxide. The mixture is then fed to another component splitter which isolates the thorium oxide based on density.

In the beta branch, rare earth phosphates from the first reactor are fed to a second reactor along with more sulfuric acid and water at 275 °C and 20 atm in order to convert all of the remaining rare earth phosphates, with the exception of neodymium, to sulfates. Once the stream exits the reactor, it is cooled and filtered so that the solid rare earth sulfates and uranium sulfate are removed from the process to be sold as a byproduct. The neodymium phosphate is then fed to a reactor with sodium hydroxide solution in order to form a solid neodymium phosphate hydrate that can be filtered out from the remaining waste products. After filtration, more sodium hydroxide solution is fed to a reactor with the neodymium phosphate hydrate in order to form the neodymium hydroxide. The hydroxide is separated from the residual contaminants in the stream via a component splitter and filter before being calcined to form the neodymium oxide.

**Figure 3.1:** Final flowsheet of the proposed process
3.2 Sustainability, Environmental Concerns, and Worker Safety

As mentioned earlier, thorium oxide has incredible potential as a fissile fuel as it can produce approximately 200 times the amount of energy compared to the same amount of uranium. Furthermore, thorium is far more abundant in nature. With this in mind, efficient production of thorium would have a tremendous impact on the energy industry and allow us to move away from energy sources that are less efficient and environmentally friendly such as fossil fuels.

In terms of the process’s sustainability, one of the major drawbacks to this proposed plant is that it consumes 4638 kilograms of water per hour. With this rate of consumption, it may be wise to consider implementing a water recycle in the system in order to reduce the amount of water used from outside sources. This is outside the scope of this study level design, however economic analysis should be conducted to determine whether the cost of the water recycling system would significantly reduce the cost of water.

Unwanted byproducts of this process include a variety of salts as well as a collection of residual rare earth elements. The salts and residual rare earth elements are to be separated from the product and disposed of. The disposal of these materials is not accounted for in this study level design, however in our analysis we have accounted for a solid waste disposal cost of $.50 per metric ton of material.

Thorium oxide should be handled and stored with extreme caution as it is a radioactive substance that can lead to multiple health issues including cancer. Furthermore, the mixed rare earth stream contains .3 mol% uranium sulfate which, while not present in high amount, still can still be considered a health hazard.
Outside of radiation concerns, many vessels in this process are operating at considerably high temperatures and pressures. Extreme caution should be taken in order to ensure that maintenance is performed regularly to prevent any potential accidents. The concerns for proper operation are heightened by the use of highly caustic acids and bases such as sulfuric acid and ammonium hydroxide throughout the entire process. With the amount of acid being used in the system, it is critical to ensure that all product and waste streams are properly neutralized before being removed from the process.

3.3 Product Purity

The primary product, thorium oxide, can be collected at a purity of ~98% or higher. Neodymium oxide, the secondary product, is also collected at a purity of ~98% or higher. A tertiary product, the mixed rare earth hydroxide, has the following mass composition: 66.36% cerium hydroxide, 33.38% lanthanum hydroxide and .26% uranium oxide.

4.0 Results

After designing a process flow diagram in OLI, an economic analysis was performed to determine the viability of the process. While thorium oxide is the primary product, it is not the most profitable product. Though neodymium oxide has a lower selling price than thorium oxide, it is nearly three times more abundant in the given composition of monazite ore. The figure below summarizes the percentage of the annual sales income that can be attributed to each product. Note that because the lanthanum and cerium products are mixed and contain a small amount of uranium, they are sold at half their listed prices. Given the presented profit
distribution, it is critical that there be a demand for neodymium oxide to ensure maximum profit. Before expenses, the gross profit from sales is determined to be $86,686,156.80.

![Profit Distribution of Rare Earth Oxide Products]

**Figure 4.1:** Distribution of revenue among products from the proposed process

### 4.1 Capital Cost Estimates

Equipment was designed based on volumetric flow requirements. All equations and values were taken from Ulrich. Some generalizations were made as Ulrich does not provide information on sizing cyclone separators. Sample calculations for each equipment type can be found in Appendix D and individual equipment costs can be found alongside the capital cost summary in Appendix C. The total grass-roots capital investment of this plant was determined to be $36,064,510.65.

### 4.2 Manufacturing Costs

The yearly manufacturing results were also calculated using the methods presented in Ulrich. A full table detailing the manufacturing costs can be found in Appendix C.
manufacturing expenses were calculated to be $26,994,479.08. A distribution of manufacturing expenses are presented in the table below:

![Distribution of Manufacturing Expenses](image)

**Figure 4.2:** A distribution of manufacturing expenses in the proposed process

### 5.0 Discussion of Results

Our study shows that this process is a viable option for extracting thorium oxide as well as other REE oxides from monazite ore at a high purity. After manufacturing expenses, depreciation, and income taxes, the annual net profit was determined to be $30,145,950.72. Assuming a three year construction period, a cash flow analysis was conducted on the process to evaluate the project using undiscounted and discounted (10% and 15%) cash flows. It is determined that the plant would begin to break even approximately 4.75 - 5.5 years after construction begins; details can be found in Appendix C. Though the initial cost of grass-roots
capital is somewhat high, the plant has a short payback period due to the high economic potential of the plant.

Raw materials account for approximately 42% of the annual manufacturing expenses; of the raw material costs, approximately 35% comes from purchasing reagents such as acids, bases, and process water. With this in mind, there may be further optimization that can be done in order to reduce the total cost of raw materials.

When designing liquid cyclone separators, they typically only effective in processes that contain materials with a minimum particle diameter of 10 μm. The material in the proposed process streams are predominantly sulfates, we estimated this material to have a particle diameter resembling calcium sulfate (7-9μm according to usg.com). Since this is close to the limit of the unit effectiveness, it would be beneficial to consult with a manufacturer to ensure the units operate effectively. Further research and development may need to be done on the part of the manufacturer to design units capable of processing particles with a diameter of 7 μm.

6.0 Conclusions

Considering the calculated economic potential of the proposed process, this is a viable option for extracting thorium oxide, neodymium oxide, and mixed REE hydroxides. While neodymium oxide does not have the highest selling price, it is nearly three times more abundant than thorium oxide in the given monazite composition and therefore has the highest economic potential of any product. As long as the selling price of neodymium oxide does not significantly drop the plant should remain profitable; earning nearly $30,000,000 per year after taxes.
Furthermore, the proposed process utilizes floatation methods to separate components based on density. While the separated components have a great difference in their respective densities, it is unlikely that the separation will be perfect. For this reason, the purity of these products is described as ~98% even though OLI calculated perfect separations in these processes.

7.0 Recommendations for Optimization

After performing an economic analysis on the proposed process, we find that this process would be a viable option for processing monazite ore. In order to ensure the process remains profitable, it is imperative that there is a demand for neodymium oxide as it accounts for the majority of profits from sales. While the proposed process is a viable option, there are various areas in which modifications could be introduced to improve plant safety and to reduce overall costs.

In terms of purchasing materials, this plant consumes 4638 kilograms of water per hour; it may be beneficial to incorporate a water recycle system into the design in order to reduce utility costs. Aside from water, strong acids and bases are also used liberally in this process. To circumvent the cost of reagents, it may be beneficial to explore adding an element such as an electrodialysis unit to convert the reacted ion salts back into their respective acids and bases.

Much research was conducted at the beginning of this project to make the operating conditions safer for the plant employees. From the existing literature, we found that it was nearly impossible to carry out these reactions and separations without the extreme pressures, temperatures and caustic materials. Nevertheless, if research funding is available, more extensive research should be performed with the goal of improving process safety in the plant.
8.0 References


“High-purity additives for high-quality food, beverage and pharmaceutical products.” *United States Gypsum Company*,


“High Temperature Properties.” *Stainless Steel Information Center*,


Appendices

Appendix A: Project Assumptions

- Cyclone separators are typically designed in close conjunction with the equipment manufacturer. Because of time constraints, it was assumed that the particle size of the materials is sufficient for cyclone operation.
- Flotation works 100%
- Assumption for specific design are outlined in appendix D.
Appendix B: Stream Compositions and Mass and Energy Balances

Table B.1a Composition of inlet and outlet streams in the proposed process
<table>
<thead>
<tr>
<th>Table B.1b</th>
<th>Composition of intermediate streams in the proposed process</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Stream</th>
<th>Mass Flowrate (kg/h)</th>
<th>Composition (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Source 1</td>
<td>800,000</td>
<td>50% H2, 30% CO, 20% CO2</td>
</tr>
<tr>
<td>Source 2</td>
<td>900,000</td>
<td>60% H2, 25% CO, 15% CO2</td>
</tr>
<tr>
<td>Source 3</td>
<td>1,000,000</td>
<td>40% H2, 40% CO, 20% CO2</td>
</tr>
</tbody>
</table>

Note: Values are approximate and subject to change based on operational conditions.
**Table B.1c** Composition of waste streams in the proposed process. These compounds are typically present in $10^6$ kg/hr or less.

*Note: Tr indicates there is trace amounts of this material present. These compounds are typically present in $10^6$ kg/hr or less.*
Energy Balances

The major sources of energy exchange in our process design are attributed to the four heat exchangers and the 10 reactors. OLI determined the heat duty required for each of these units. Using this information, we could determine the amount of cooling water or steam needed in the system to have a net energy value of zero. The following equation was used extensively to find the water requirement:

\[ Q_{\text{process}} = m_{\text{steam or cw}} \times \Delta h_{\text{steam or cw}} \times \varepsilon. \]  

Equation 1

Where \( Q \) represents heat duty, \( m_{\text{steam or cw}} \) is the mass of steam (or cooling water), \( \Delta h_{\text{steam or cw}} \) is enthalpy change of the material, and \( \varepsilon \) is efficiency of the heat exchange.

E-173

Analyzing the unit in OLI, we see that the process heat duty is negative so the process material will need to lose energy to reach its desired state. In this situation, cooling water is used to absorb the heat that the process stream is losing. We assume an efficiency value of 0.97 and a cooling water temperature change of 15°C (making sure that the cooling water it’s expected outlet flow is colder than the process steam conditions), in order to find the desired mass flow rate. Thus,

\[ Q_{\text{process}} = -1498.08 \, \text{kJ/s} \text{ (Obtained from OLI flowsheet)} \]
\[ \varepsilon = 0.97 \text{ (Assumed value based on U-tube calculations in Ulrich)} \]
\[ \Delta h_{\text{cw}} = C_p \times \Delta T = 4.19 \, \text{kJ/kg°C} \times 15^\circ\text{C} = 62.85 \, \text{kJ/kg} \]

We can solve by rearrange equation 1 to the following form:

\[ m_{\text{cw}} = \frac{Q_{\text{process}}}{\Delta h_{\text{cw}} \times \varepsilon}. \]  

Equation 2

Plugging in values, we find that:

\[ m_{\text{cw}} = \frac{1498.08 \, \text{kJ/s}}{62.85 \, \text{kJ/kg} \times 0.97} = 24.57 \, \text{kg/s} \]

E-123

This unit uses the same assumptions and logic as E-173 as it’s heat duty is also negative. The only values that changes is \( Q_{\text{process}} \). Giving the following calculation and cooling water flow rate:
\[ Q_{\text{process}} = -2,312.14 \text{ kJ/s (Obtained from OLI flowsheet)} \]
\[ m_{cw} = \frac{2,312.14 \text{ kJ/s}}{62.85 \text{ kJ/kg} \times 0.97} = 37.9 \text{ kg/s} \]

E-101

This unit also has a negative heat duty. However, the method of determining cooling water flow rate was slightly different. Instead of solving for \( m_{cw} \) explicitly, a guess and check method was implemented. In this case, we guessed a value for \( m_{cw} \) and checked to see if it could satisfy the needs of the system. The process stream for the heat exchanger has an inlet temperature of 220°C and a desired outlet temperature of 25°C. Assuming our cooling water has an inlet around 9°C, we need a cooling water outlet temperature no greater than 25°C for the operation to be feasible. Using the guess and check method, we find that assuming a 35,000 kg/hr cooling water flow rate yields a temperature below 25°C. The following calculation is evidence:

\[ Q_{\text{process}} = -1,992,086.08 \text{ kJ/hr (Obtained from OLI flowsheet)} \]
\[ \varepsilon = 0.97 \text{ (Assumed value based on U-tube calculations in Ulrich)} \]
\[ T_{\text{initial}} = 9°C \]
\[ C_p = 4.19 \text{ kJ/kg °C (Value for water)} \]

Equation 1 can be rearranged to the following form:

\[ T_{\text{final}} = \frac{Q_{\text{process}}}{m_{cw} \varepsilon C_p} + T_{\text{initial}} \]

Equation 3

When plugging in values, \( T_{\text{final}} = 23°C \).

E-161

This unit uses the same methodology as E-101. The heat duty is negative, with the process stream cooling down from 100°C to 70°C. We assume a 10,000 kg/hr flow cooling water flow rate and find that the final temperature of the cooling water is below 70°C making it feasible. This is shown in the following calculation:

\[ Q_{\text{process}} = -291,296.77 \text{ kJ/hr (Obtained from OLI flowsheet)} \]
\[ \varepsilon = 0.97 \text{ (Assumed value based on U-tube calculations in Ulrich)} \]
\[ T_{\text{initial}} = 9°C \]
\[ C_p = 4.19 \text{ kJ/kg °C (Value for water)} \]

Using Equation 3, \( T_{\text{final}} = 16.16°C \).
Reactors

OLI calculated a heat duty that went along with each reactor during the process with the exception being the calciner. If the reactor heat duty was negative, we chose add a cooling water jacket to the vessel so it could absorb the heat released from the reaction. If the heat duty was positive, we designed a steam jacket to add heat to the vessel contents so it could stay at the optimum reaction temperature.

The following reactors have negative heat duties: R-120, R-140, R-180, R-190, and R-1000. An energy balance was conducted on these units by finding the cooling water mass flow rate in the jacket. We used the OLI values, assumed an efficiency value of 0.80, and a cooling water temperature change of 15°C (making \( \Delta h_{cw} = 62.85 \text{ kJ/kg} \)) to plug into equation 2. The following values were calculated:

\[
R-120: \quad Q_{\text{Process}} = -1,448.17 \text{ kJ/s} \\
m_{cw} = \frac{1,448.17 \text{ kJ/s}}{62.85 \text{ kJ/kg*0.80}} = 28.8 \text{ kg/s}
\]

\[
R-140: \quad Q_{\text{Process}} = -131.68 \text{ kJ/s} \\
m_{cw} = \frac{131.68 \text{ kJ/s}}{62.85 \text{ kJ/kg*0.80}} = 2.62 \text{ kg/s}
\]

\[
R-180: \quad Q_{\text{Process}} = -1,527.28 \text{ kJ/s} \\
m_{cw} = \frac{1,527.28 \text{ kJ/s}}{62.85 \text{ kJ/kg*0.80}} = 30.38 \text{ kg/s}
\]

\[
R-190: \quad Q_{\text{Process}} = -111.46 \text{ kJ/s} \\
m_{cw} = \frac{111.46 \text{ kJ/s}}{62.85 \text{ kJ/kg*0.80}} = 2.22 \text{ kg/s}
\]

\[
R-1000: \quad Q_{\text{Process}} = -0.215 \text{ kJ/s} \\
m_{cw} = \frac{0.215 \text{ kJ/s}}{62.85 \text{ kJ/kg*0.80}} = 0.0043 \text{ kg/s}
\]

The remaining reactors have positive heat duties: R-100, R-110, R-130, R-160, and R-170. The reaction temperatures for these reactors range from 100°C to 275°C. Since heat is being lost from these units, we need a higher temperature jacket to provide a heat gradient so the temperature of the vessel contents can remain constant (as it is modeled in OLI). We chose to flow steam at a pressure of 70 bar through the reactor jacket. This was chosen due to the fact that the temperature of saturated steam temperature at 70 bar is 285°C which is greater than the reaction temperatures. Also, this allows us to stay consistent with the purchase of steam by only needing one steam pressure throughout the process. We used Equation 2 to solve for the flow rate of steam. The calculations were made based on the following information:
\[ \Delta h_{steam} = 1505 \text{ kJ/kg} \] (heat of vaporization of saturated 70 bar steam), an efficiency value of 0.80, and the heating duty for each unit (based on flowsheet from OLI). The following values were calculated:

**R-100:** 
\[ Q_{process} = 410.48 \text{ kJ/s} \]
\[ m_{steam} = \frac{410.48 \text{ kJ/s}}{1505 \text{ kJ/kg \cdot 0.80}} = 0.34 \text{ kg/s} \]

**R-110:** 
\[ Q_{process} = 1,917.92 \text{ kJ/s} \]
\[ m_{steam} = \frac{1,917.92 \text{ kJ/s}}{1505 \text{ kJ/kg \cdot 0.80}} = 1.59 \text{ kg/s} \]

**R-130:** 
\[ Q_{process} = 64.38 \text{ kJ/s} \]
\[ m_{steam} = \frac{64.38 \text{ kJ/s}}{1505 \text{ kJ/kg \cdot 0.80}} = 0.053 \text{ kg/s} \]

**R-160:** 
\[ Q_{process} = 24.47 \text{ kJ/s} \]
\[ m_{steam} = \frac{24.47 \text{ kJ/s}}{1505 \text{ kJ/kg \cdot 0.80}} = 0.02 \text{ kg/s} \]

**R-170:** 
\[ Q_{process} = 16.53 \text{ kJ/s} \]
\[ m_{steam} = \frac{16.53 \text{ kJ/s}}{1505 \text{ kJ/kg \cdot 0.80}} = 0.013 \text{ kg/s} \]

*Note heat duty was not accounted for in the calciner, as the OLI flow sheet would not converge when modeling the calciner unit. This will need to be examined in the future.*
## Capital Cost Summary

### Equipment Number  Description  Capacity or Size Specs  Cbm

**Heat Exchangers**

C-101  Heat Exchanger: Cool R-100 Exit Stream  55.26 m², stainless steel, u-tube, 14.2 barg  74036.825 S

E-16  Heat Exchanger: Cool R-150 Exit Stream  4.51 m², stainless steel, 10 barg  24621.437 S

E-123  Heat Exchanger: Condense -123 Vapor Stream  14.49 m², stainless steel, u-tube, 10 barg  43423.265 S

E-173  Heat Exchanger: Condense -172 Vapor Stream  18.27 m², stainless steel, u-tube, 10 barg  43423.265 S

**Reactors**

R-10  Monazite Acid Rin  1.7 m ID x 6.8 m vert, stainless steel, jacketed vessel, 14.2 barg  96412.38 S

R-110  Cerium & Lanthanium Phosphate to Sulfate Rin  1.9 ID x 7.5 m vert, stainless steel, jacketed vessel, 14.2 barg  107536.27 S

R-120  Neodymium Precipitation Rin  5.4 m ID x 21.6 m vert, stainless steel, jacketed vessel, 34.4 barg  91306.68 S

R-130  Neodymium Phosphate to Oxide Rin  4.5 m ID x 17.9 m vert, stainless steel, jacketed vessel, 29.4 barg  71610.05 S

R-170  Thorium Sulfate to Oxide Rin  1.5 m ID x 5.9 m vert, stainless steel, jacketed vessel, 24.3 barg  86667.63 S

R-160  REE Calciner  Electrical circulation heater, 1.2KW  82457.192 S

R-160  REE Sulfates to Hydroxide Rin  1.8 m ID x 6.2 m vert, stainless steel, jacketed vessel, 5 barg  60102.10 S

R-1000  Dilution/Neutralization  1.7 m ID x 6.6 m vert, stainless steel, jacketed vessel, 5 barg  69248.77 S

R-180  Dilution/Neutralization  1.5 m ID x 5.0 m vert, stainless steel, jacketed vessel, 5 barg  51740.73 S

R-190  Waste Phase Conversion  0.73 m ID x 2.9 m vert, stainless steel, jacketed vessel, 5 barg  16730.37 S

R-140  Waste Phase Conversion  0.93 m ID x 3.7 m vert, stainless steel, jacketed vessel, 5 barg  22055.93 S

**Total Reactors**  2102167.61 S

**Process Vessels**

H-103  Splitter: Thorium Sulfate + Silicon Dioxide  Column: 0.82 m height, 0.21 m ID, stainless steel, Trays: 0.21m ID, stainless steel  19443.272 S

H-132  Splitter: Solid Waste + Neodymium Hydroxide Column: 1.02 m height, 0.25 m ID, stainless steel, Trays: 0.26 m ID, stainless steel  67904.94 S

H-174  Splitter: Solid Waste + Thorium Oxide Column: 0.54 m height, 0.4 m ID, stainless steel, Trays: 0.14 m ID, stainless steel  30741.98 S

H-163  Splitter: Solid Waste + Mixed REE Hydroxides  Column: 1.57 m height, 0.39 m ID, stainless steel, Trays: 0.39 m ID, stainless steel  57275.12 S

**Total process vessels**  175369.28 S

**Mixers**

M-100a  Agitator for R-100  2.64 kW, stainless steel  47421.87 S

M-110a  Agitator for R-110  3.4 kW, stainless steel  53112.3 S

M-120a  Agitator for R-120  42.6 kW, stainless steel  22033.7 S

M-130a  Agitator for R-130  27.29 kW, stainless steel  15750 S

M-170a  Agitator for R-170  1.82 kW, stainless steel  41731.25 S

M-190a  Agitator for R-190  2.13 kW, stainless steel  47421.87 S

M-1000a  Agitator for R-1000  2.64 kW, stainless steel  47421.87 S

M-180a  Agitator for R-180  1.91 kW, stainless steel  38984.7 S

M-190a  Agitator for R-190  0.36 kW, stainless steel  34147.75 S

M-140a  Agitator for R-140  0.63 kW, stainless steel  34147.75 S

**Total mixers**  711328.12 S

**Pumps**

L-111  Centrifugal, 0.42 kW, stainless steel  56223.35 S

L-124  Centrifugal, 0.94 kW, stainless steel  172919.123 S

**Total pumps**  229142.13 S

**Separators**

H-102  Filter: Thorium Oxide/Silicon Dioxide + REE Phosphates  Belt filter, 1.32 m², stainless steel  191210.1 S

H-162  Filter: REE Hydroxide + Liquid Waste  Belt filter, 1.30 m², stainless steel  191210.1 S

H-191  Filter: Solid Waste + Liquid Waste  Belt filter, 0.42 m², stainless steel  10360.3 S

H-141  Filter: Liquid Waste + Liquid Waste  Belt filter, 1.30 m², stainless steel  12600.1 S

H-11  Settler: REE Phosphates + Neodymium Phosphate  Liquid cyclone, 5.5E-3 m², stainless steel  21245 S

H-121a  Liquid cyclone, 2.5E-4 m², stainless steel  4209 S

H-131a  Liquid cyclone, 3.4E-6 m², stainless steel  1396 S

H-122a  Liquid cyclone, 2.5E-4 m², stainless steel  4209 S

H-171a  Liquid cyclone, 3.4E-6 m², stainless steel  1396 S

H-172a  Liquid cyclone, 1.3E-4 m², stainless steel  3185 S

H-131b  Gas cyclone, 9.8E-5 m², stainless steel  8346 S

H-132b  Gas cyclone, 9.8E-5 m², stainless steel  8346 S

H-171b  Gas cyclone, 3.3E-5 m², stainless steel  8346 S

H-172b  Gas cyclone, 3.3E-5 m², stainless steel  8346 S

**Total Separators**  668776.95 S

**Total base module cost**  23519111.23 S

**Contingency and Fee**  4223820.02 S

**Total module cost**  27741931.2 S

**Auxiliary facilities**  832579.38 S

**Grass-roots capital**  36804501.83 S

---

Table C.1: Capital Cost Summary of the proposed process
## Manufacturing Cost Summary

### Table C.2: Manufacturing Cost Summary of the proposed process

<table>
<thead>
<tr>
<th>Job Title: Monazite Processing Plant</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Capital</strong></td>
<td></td>
</tr>
<tr>
<td>Fixed Capital</td>
<td>$36,064,510.65</td>
</tr>
<tr>
<td>Working Capital</td>
<td>$3,606,451.07</td>
</tr>
<tr>
<td>Total Capital Investment</td>
<td>$39,670,961.72</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Manufacturing Expenses</th>
<th>Annual Cost ($/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Direct</strong></td>
<td></td>
</tr>
<tr>
<td>Raw Material</td>
<td>$13,859,667.99</td>
</tr>
<tr>
<td>Catalysts</td>
<td>$0</td>
</tr>
<tr>
<td>Operating Labor (15% of operating labor)</td>
<td>$1,568,761.02</td>
</tr>
<tr>
<td>Supervisory Labor (15% of operating labor)</td>
<td>$235,314.15</td>
</tr>
<tr>
<td><strong>Utilities</strong></td>
<td></td>
</tr>
<tr>
<td>Steam</td>
<td>$3,335,144.62</td>
</tr>
<tr>
<td>Electricity</td>
<td>$310,660.18</td>
</tr>
<tr>
<td>Refrigerant</td>
<td>$0</td>
</tr>
<tr>
<td>Cooling Water</td>
<td>$802,093.43</td>
</tr>
<tr>
<td>Wastewater Disposal</td>
<td>$3,861.95</td>
</tr>
<tr>
<td>Maintenance (6% of fixed capital)</td>
<td>$2,163,870.64</td>
</tr>
<tr>
<td>Operating supplies (15% of maint. &amp; repairs)</td>
<td>$324,580.60</td>
</tr>
<tr>
<td>Laboratory charges (15% operating labor)</td>
<td>$235,314.15</td>
</tr>
<tr>
<td>Patents and Royalties (3% of total expense)</td>
<td>$68,539.54</td>
</tr>
<tr>
<td>Total</td>
<td>$23,531,776.27</td>
</tr>
<tr>
<td><strong>Indirect</strong></td>
<td></td>
</tr>
<tr>
<td>Overhead (60% of op labor, supervision, and maintenance)</td>
<td>$2,380,767.49</td>
</tr>
<tr>
<td>Local Taxes (2% of fixed capital)</td>
<td>$721,290.21</td>
</tr>
<tr>
<td>Insurance (1% of fixed capital)</td>
<td>$360,645.11</td>
</tr>
<tr>
<td>Total</td>
<td>$3,462,702.81</td>
</tr>
<tr>
<td><strong>General Expenses</strong></td>
<td></td>
</tr>
<tr>
<td>Administrative Cost (25% of overhead)</td>
<td>$595,191.87</td>
</tr>
<tr>
<td>Distribution and Selling (10% of total expense)</td>
<td>$3,670,132.00</td>
</tr>
<tr>
<td>R&amp;D (5% of total expense)</td>
<td>$1,835,066.00</td>
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<tr>
<td>Total</td>
<td>$5,100,389.87</td>
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<tr>
<td><strong>Depreciation (10% of fixed capital)</strong></td>
<td>$3,606,451.07</td>
</tr>
<tr>
<td><strong>Total Expenses</strong></td>
<td>$36,701,320.02</td>
</tr>
<tr>
<td><strong>Revenue from Sales</strong></td>
<td>$86,666,156.80</td>
</tr>
<tr>
<td>Net Annual Profit</td>
<td>$46,378,385.72</td>
</tr>
<tr>
<td>Income Taxes</td>
<td>$16,232,435.00</td>
</tr>
<tr>
<td><strong>Net annual profit after taxes</strong></td>
<td>$30,145,950.72</td>
</tr>
</tbody>
</table>


Cash Flow Analysis Chart

Figure C.1: Cash Flow Analysis Chart of the proposed process
Appendix D: Equipment Design and Cost

Reactor Calculation:

Sample- R-100 and M-100a

In order to get an accurate cost of the reactor, it must be designed properly. Based on the information obtained from Rodliyah, the reactor must have a space time of 2.5 hours to achieve optimal acid digestion. The process conditions of this reactor are 220˚C and 15 atm, based on the literature reactions and OLI calculations. We find the volumetric flow rate to the reactor consists of the raw material feed rate and the sulfuric acid feed rate. Once these parameters are known, volume and height can be determined for the reactor. The volume of the reactor will then allow us design and cost the agitator for the unit. We cost the vessel using figures 5.44, 5.45, and 5.46 from Ulrich. Agitator economic calculations incorporate figure 5.42 from Ulrich. It should be noted that stainless steel was used as the material of construction for the reactor and the agitator due to the fact that we are processing acid and the material is corrosion resistant. Also, it can handle high temperatures up to 920˚C (ssina.com). Ulrich notes a limitation for stainless steel processing acids at 200˚C, a deeper evaluation into this selection is suggested. We designed and costed the reactor by the following calculations:

\[
V_{R-100} = q \times \tau = 3.0339 \text{ m}^3/\text{hr} \times 2.5 \text{hrs} = 7.58 \text{m}^3
\]

\[
V_{R-100}@100\% \text{ excess} = 7.58 \text{m}^3 \times 2 = 15.17 \text{m}^3
\]

\[
D_{R-100} = \left( \frac{V_{R-100} \times 4}{2\pi} \right)^{1/3} = \left( \frac{7.58 \text{m}^3 \times 4}{2\pi} \right)^{1/3} = 1.69 \text{m}
\]

\[
H_{R-100} = \frac{4 \times V_{R-100}@100\% \text{ excess}}{\pi D_{R-100}^2} = \frac{4 \times 15.17 \text{m}^3}{\pi \times 1.69^2 \text{m}} = 6.76 \text{m}
\]
\[ P_{M-100a} = (V_{R-100} @ 100\% \text{ excess})^{0.8} \times 0.3 = (15.17m^3)^{0.8} \times 0.3 = 2.64kW \]

*Note: Power equation comes from table 4.16 in Ulrich.

\[ C_{bm,R-100} = C_p \times F_{BM} \text{ (function of } F_p \text{ and } F_M \text{)} \times 607/400 \times \text{jacket factor} = 26000 \times 14.9 \times 607/400 \times 1.64 = 964,122.38. \]

*Jacket factor was calculated by dividing jacketed vessels by plain vessels in figure 5.23 for volumes equivalent to the desired reactor volumes.

\[ C_{bm,M-100A} = C_p \times F_{BM} \times 607/400 = 12,500 \times 2.5 \times 607/400 = 47,421.88 \]

**Other reactors and agitators**

The other reactors implemented the same method. Parameters needs for design and costing can be obtained from OLI flowsheet. We assumed a space time of 2.5 hours for each vessel, and that the material of construction is stainless steel with extensive amount of acids and bases used in the process. Values are shown in the table below:

<table>
<thead>
<tr>
<th>Temperature (°C)</th>
<th>Power (kW)</th>
<th>Sp. Time (hr)</th>
<th>Volume (m³)</th>
<th>Height of reactor (m)</th>
<th>Tbr (°C)</th>
<th>Jacket factor</th>
<th>Weight of vessel (kg)</th>
<th>Height of vessel (m)</th>
<th>Porous material</th>
<th>Weight of jacket (kg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>23°C</td>
<td>235</td>
<td>24.5</td>
<td>0.0184</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
<tr>
<td>25°C</td>
<td>275</td>
<td>26.8</td>
<td>0.0186</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
<tr>
<td>27°C</td>
<td>250</td>
<td>21.9</td>
<td>0.0186</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
<tr>
<td>30°C</td>
<td>265</td>
<td>22.2</td>
<td>0.0187</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
<tr>
<td>35°C</td>
<td>300</td>
<td>22.5</td>
<td>0.0187</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
<tr>
<td>40°C</td>
<td>350</td>
<td>22.8</td>
<td>0.0188</td>
<td>2.5</td>
<td>1.31</td>
<td>1.65</td>
<td>228,527</td>
<td>2.15</td>
<td>1,535</td>
<td>22,527</td>
</tr>
</tbody>
</table>

Table D.1: Values utilized in agitator design
Belt Filter Calculation:

Sample- Filter- H-102

Belt filters are used to separate solid-liquid mixtures at process temperatures less than 100℃. In the case of H-102, the filter separated solid thorium oxide and silicon dioxide from liquid REE phosphates. The only parameter needed to determine the size and economics of the filter is the volumetric flow rate to the unit. The filter was designed using table 4.23b from Ulrich and the cost was determined using figure 5.57b from Ulrich. The belt filter was designed to be constructed of stainless steel since corrosive acids will be present in the unit. It was also assumed that this is a continuous filter. The design and cost calculations are shown below:

\[
A_{\text{Nominal area of H-102}} = \frac{q}{\text{design factor}} = \frac{6.6 \times 10^{-3} m^3/s}{5 \times 10^{-4} m/s} = 1.32 m^2
\]

*Note: Design factor is based on feed liquid flow rate from table 4.23b in Ulrich. The material was assumed to resemble dilute slimes.

\[
C_{bm,H-102} = C_P * F_{BM} * 607/400 = $35000 * 3.6 * 607/400 = $191,205
\]

Other belt filters

H-162 was designed using the same method and assumptions as H-102. H-141 and H-191 were designed using the same method, however one assumption was different. Instead of resembling dilute slimes, these filter process material that more closely resembles sands in table 4.23b in Ulrich.
**Flotation Column Calculation:**

**Sample- H-103**

Floatation columns were used to separate solids in the process by their densities. In this case, we are separating silicon dioxide from thorium sulfate. After consulting Dr. Counce, we decided to design the flotation column as a vertically oriented process vessel with two sieve trays. We designed the vessel and its tray to be stainless steel to be safe because thorium is in sulfate form and the process unit upstream contains a sulphuric acid digestion. The parameters needed in the calculation are volumetric flow rate to the column, process conditions, and spacetime. We cost the vessel using figures 5.44, 5.45, and 5.46 from Ulrich. We cost the sieve trays by figure 5.48 from Ulrich. The design and cost calculations are provided below:

\[
V_{H-103} = q * \tau = 0.02743 \text{ m}^3/\text{hr} * 0.5 \text{hrs} = 0.013715 \text{ m}^3
\]

\[
V_{H-103 @} 100\% \text{ excess} = 0.013715 \text{ m}^3 * 2 = 0.02743 \text{ m}^3
\]

\[
D_{H-103} = \left( \frac{V_{H-103}@4}{2\pi} \right)^{1/3} = \left( \frac{0.013715m^3@4}{2\pi} \right)^{1/3} = 0.2059 \text{ m}
\]

\[
H_{H-103} = \frac{4*V_{H-103 @} \text{100\% excess}}{\pi D_{H-103}^2} = \frac{4*0.02743m^3}{\pi*0.2059^2 m} = 0.824 \text{ m}
\]

\[
C_{bm,H-103, \text{vessel}} = C_P * F_{BM} (\text{function of } F_P \text{ and } F_M) * 607/400 = 1175 * 10.5 * 607/400 = $18722.156
\]

\[
C_{bm,H-103, \text{trays}} = C_P * F_{BM} * N_{act} * f_q * 607/400 = 30 * 2.2 * 2 * 3.6 * 607/400 = $721.12
\]

\[
C_{bm,H-103, \text{total}} = $19,443.28
\]

**Other flotation columns**

The same method is used to calculate the other three flotation columns. Parameters of interest come from OLI flowsheet data.
Heat Exchanger Calculation:

Sample- E-173

This unit has a process stream that is predominantly water vapor at 160°C at 1 atm, with 1.17 wt% of ammonia and is coming out of the exchanger as a liquid at 25°C. An assumption that the process stream is all water was made to simplify calculations. The heat duty, mass flow rate of process stream, inlet and outlet temperature of the process stream, and assumed inlet and outlet temperature of cooling water stream are the parameters needed for the design. Since the process stream could contain a moderately strong base, stainless steel is used for the material of construction. We calculated the surface area needed for a U-tube heat exchanger and used figures 5.36, 5.37, and 5.38 for Ulrich to find economics for the equipment. It is presumed that there will be three sections in the heat exchanger; two subcooling sections and a condensation section. The calculation is demonstrated below:

First subcooling section: Process stream (160 °C vapor to 100°C vapor) assuming cooling water comes in at 9°C and leaves at 24°C.

\[ Q = (h_{at\ 160^\circ C,vap} - h_{at\ 100^\circ C,vap}) \cdot m_{process\ stream} = (2,795.8 - 2,675.9)kJ/kg \cdot 0.557 \ kg/s = 66.78kJ/s \]

*Enthalpy values obtained from steam tables

\[ \Delta T_m = \frac{(160-24)-(100-9)}{ln((160-24)/(100-9))} = 112^\circ C \]

\[ A_{sub1} = \frac{Q}{U\cdot\Delta T_m} = \frac{66.78 \ kJ/s \cdot 1,000kJ/1kJ}{900 \ J/m^2\cdot\circ C \cdot 112^\circ C} = 0.66 \ m^2 \]

*The U value was obtained from table 4.15a in Ulrich.
Condensation section: Process stream (100 °C vapor to 100°C liquid) assuming cooling water enters at 9°C and leaves at 24°C.

\[ Q = (h_{at\ 100^\circ C,\ latent\ heat\ of\ vaporization}) \times m_{process\ stream} = 2,257.6 \text{kJ/kg} \times 0.557 \text{kg/s} = 1,257.48 \text{kJ/s} \]

*Enthalpy value obtained from steam table

\[ \Delta T_m = \frac{(100-24)-(100-9)}{\ln((100-24)/(100-9))} = 83.3^\circ C \]

\[ A_{cond} = \frac{Q}{U \times \Delta T_m} = \frac{1,257.48 \text{kJ/s} \times 1,000 \text{J/J}}{1,200 \text{J/m}^2\text{C} \times 83.3^\circ C} = 12.58 \text{m}^2 \]

*The U value was obtained from table 4.15a in Ulrich.

Second subcooling section: Process stream (100 °C liquid to 25°C liquid) assuming cooling water comes in at 9°C and leaves at 24°C.

\[ Q = (h_{at\ 100^\circ C, liq} - h_{at\ 25^\circ C, liq}) \times m_{process\ stream} = (417.51 - 104.84)\text{kJ/kg} \times 0.557 \text{kg/s} = 174.16 \text{kJ/s} \]

*Enthalpy value obtained from steam table

\[ \Delta T_m = \frac{(100-24)-(25-9)}{\ln((100-24)/(25-9))} = 38.51^\circ C \]

\[ A_{sub2} = \frac{Q}{U \times \Delta T_m} = \frac{174.16 \text{kJ/s} \times 1,000 \text{J/J}}{900 \text{J/m}^2\text{C} \times 38.51^\circ C} = 5.02 \text{m}^2 \]

*The U value was obtained from table 4.15a in Ulrich.

Total area for heat exchanger = \[ A_{sub1} + A_{cond} + A_{sub2} = 18.27 \text{m}^2 \]

\[ C_{bm,E-173} = C_p \times F_{BM}(function\ of\ F_p\ and\ F_M) \times 607/400 = $4,850 \times 5.9 \times 607/400 = $43,423.26 \]

Other heat exchangers

The U-tube, E-123, is calculated in the same fashion as E-173. The process stream’s main component is water vapor, only this time it is coming into the exchanger at 190°C and leaves as a liquid at 25°C. The same assumptions are made to obtain the design and cost.
E-101 and E-161 only have one subcooling section in the heat exchanger. These exchangers use the heat duty from OLI as the value of $Q$ in the problem. E-101 assumes cooling water coming in at 9°C and leaving at 23°C. While E-161 assumes cooling water coming in at 9°C and leaving at 16.16°C. OLI can provide needed process stream conditions and heat duty.

**Cyclone Calculation:**

Separators modeled in OLI with three outlet streams represent one gas cyclone unit and one liquid cyclone unit in series. The series orientation allows the solid, liquid, and vapor phases to be separated in individual outlet streams. These separators are units H-121, H-131, H-171, and H-172.

**Sample- Liquid Cyclone- H-121a**

Cyclones are low maintenance, inexpensive units that require little to no operator attention. The liquid cyclone separate solids from liquids and can handle temperatures up to 400°C. The preliminary design and economics of the cyclone are like belt filters in that they only depend on the volumetric flow into the unit. Stainless steel was used as the material of construction due to the presence of corrosive sulfates in the process stream. A more in-depth design of these types of separators is needed before moving further in the project. The preliminary design is based on table 4.23a in Ulrich and the economics of the liquid cyclone is based on reading figure 5.55 in Ulrich and using the required volumetric flow rate. The sample calculation is provided below:

$$A_{\text{Nominal area of H-121a}} = \frac{q}{\text{design factor}} = \frac{1.247\times 10^{-3} \text{m}^3/\text{s}}{5 \text{m/s}} = 2.49 \times 10^{-4} \text{ m}^2$$
*Design factor is based on feed liquid flow rate from table 4.23a in Ulrich.

\[ C_{bm, H-121a} = C_P * F_{BM} * 607/400 = \$700 * 4 * 607/400 = \$4,249.00 \]

**Sample- Gas Cyclone- H-171b**

Gas cyclones in our process separate liquids from gases. Like belt filters and liquid cyclones, the design and economics only depend on the volumetric flow rate. Stainless steel was again used due to the presence of corrosive sulfates. The preliminary design is based on table 4.23a in Ulrich and the economics of the liquid cyclone is based on reading figure 5.56 in Ulrich and using the required volumetric gas rate. The sample calculation is provided below:

\[ A_{Nominal area of H-171b} = \frac{q}{design\ factor} = \frac{5.015*10^{-4} m^3/s}{15 m/s} = 3.34 * 10^{-5} m^2 \]

*Design factor is based on feed liquid flow rate from table 4.23a in Ulrich.

\[ C_{bm, H-171b} = C_P * F_{BM} * 607/400 = \$1,100 * 5 * 607/400 = \$8,346.25 \]

**Other cyclones**

The other cyclones use the same assumptions and methods as what the sample examples implemented.

**Pump Calculation:**

**Sample- L-124**

In order for flow to subsist, sometimes energy must be added to the liquid to increase the pressure at that reference position above downstream process pressure. In this case we examine stream 3B. The design of a pump requires knowledge of the desired volumetric flow rate, desired inlet and outlet pressures of the pump, viscosity of the material, and motor efficiency. OLI
provides us with the flow rates and desired pressures. Further research is needed for the remaining parameters. The economics for the pump were determined from figures 5.49, 5.50, and 5.51 from Ulrich. We chose a centrifugal pump because of its ability to handle solids (as we have some in the process stream). Stainless steel was used as the material of construction because of its corrosion resistance. The pump has desired inlet and outlet pressures of 15 atm and 36 atm respectively. The calculation used for the pump design and its economics is provided below:

\[ \varepsilon (\text{pump efficiency}) = (1 - 0.12q^{-0.27})(1 - \mu^{0.8}) = (1 - 0.12(0.0271m^3/s)^{-0.27})(1 - (0.0014 Pa s)^{0.8}) = 0.6786 \]

*Stream viscosity obtained from sulphuric-acid.com for a 44 wt% acid by weight stream.

\[ W_s = \frac{q\Delta P}{\varepsilon} = \frac{0.0271 \ m^3/s \ast (3.647,700 \ Pa^{-1}1,519.875 \ Pa)}{0.6786} = 84,980.83 \ \text{Watts} \]

\[ \text{Motor power consumption} = \frac{W}{\varepsilon_{\text{motor}}} = \frac{84,980.83 \ \text{Watts}}{0.91} = 93,385.52 \ \text{Watts} \]

*\( \varepsilon \) for the electric motor was estimated from figure 4.2 in Ulrich.

\[ C_{bm, t-124} = C_P \ast F_{BM} (\text{function of } F_p \ \text{and } F_M) \ast 607/400 = $21,500 \ast 5.3 \ast 607/400 = $172,919.13 \]

**Centrifugal Pump-111**

Design was carried out using the same method and construction material. Flow rates and stream composition differs from 3b.