

5-2019

Extraction of Thorium Oxide from Monazite Ore

Makalee Ruch

University of Tennessee, Knoxville, mruch1@vols.utk.edu

Chloe Frame

University of Tennessee, Knoxville

Molly Landon

University of Tennessee, Knoxville

Ralph Laurel

University of Tennessee, Knoxville

Annabelle Large

University of Tennessee, Knoxville

Follow this and additional works at: https://trace.tennessee.edu/utk_chanhonoproj

Part of the [Environmental Chemistry Commons](#), [Geological Engineering Commons](#), and the [Other Chemical Engineering Commons](#)

Recommended Citation

Ruch, Makalee; Frame, Chloe; Landon, Molly; Laurel, Ralph; and Large, Annabelle, "Extraction of Thorium Oxide from Monazite Ore" (2019). *Chancellor's Honors Program Projects*.

https://trace.tennessee.edu/utk_chanhonoproj/2294

This Dissertation/Thesis is brought to you for free and open access by the Supervised Undergraduate Student Research and Creative Work at Trace: Tennessee Research and Creative Exchange. It has been accepted for inclusion in Chancellor's Honors Program Projects by an authorized administrator of Trace: Tennessee Research and Creative Exchange. For more information, please contact trace@utk.edu.

Extraction of Thorium Oxide from Monazite Ore

Dr. Robert Counce
Department of Chemical and Biomolecular Engineering
University of Tennessee

Chloe Frame
Molly Landon
Annabel Large
Ralph Laurel
Makalee Ruch
CBE 488: Honors Design Internship in Green Engineering
Spring 2019 Group 1

Table of Contents

1.0 Introduction	1
2.0 Synthesis Information for Processes	2
2.1 Overall Process Design	2
2.2 Process Chemistry	2
2.3 Literature Summary	4
2.4 Basic Process Economics	6
4.0 Results	8
4.1 Optimization	8
4.3 Safety, Health, and Environmental Analysis	10
4.4 Capital cost estimates	13
5.0 Discussion of Results	16
6.0 Conclusions	17
7.0 Recommendations	17
8.0 References	18
Appendices	20
Appendix A: Mass and Molar Flow Rates (by stream)	20
Appendix B: Economic Results and Analysis	24
Appendix C: Equipment Design and Costing Examples	26

1.0 Introduction

The purpose of the analysis is to determine the most viable process to extract thorium oxide from crude monazite minerals. The process examined uses sulfuric acid leaching and water to extract the thorium, and rare earth oxides are also extracted as byproducts in the process. The rare earth elements, REEs, are a part of the lanthanide group; specifically, La, Ce, and Nd are the REEs examined here. There are many approaches to processing monazite for extraction of thorium to produce its desired oxide products. One approach to recovering the thorium is through various stripping, filtration, separation, and calcining steps. In this process, thorium is considered a nuclear fuel due to its fertility and relative abundance compared to uranium. Thorium can accept a neutron and transmute to uranium-223, which can then undergo nuclear fission to produce fuel.

This report will start with an overview of the synthesis information for the purification processes, which will vary in scope from chemical-level details, such as reaction parameters, relevant properties, and costs associated with each individual species, up to plant-wide information, including block diagrams and equipment costs in US dollars. Next, the method of approach for this project will be clarified, highlighting what isolation method was chosen and what specific pieces of equipment are included in the design. After detailing the results and making plant-wide comparisons, process flow diagrams will be presented with mass and molar flow rates of each stream. It will end with tables of capital costs and operating costs, a discussion of results and conclusions, and recommendations for further work in isolating thorium oxide. The scale of this process will be determined by an overall feed rate of 1000 kg/hr, and the purity of thorium oxide obtained by this process is anticipated to be 90%. All costs associated with the compounds of interest will be adjusted using a preliminary CE index of 616.4 for October 2018. References and appendices will include precise calculations.

2.0 Synthesis Information for Processes

2.1 Overall Process Design

The process was designed to extract thorium oxide from a given feed constraint of 1000 kg/hr of raw monazite ore.

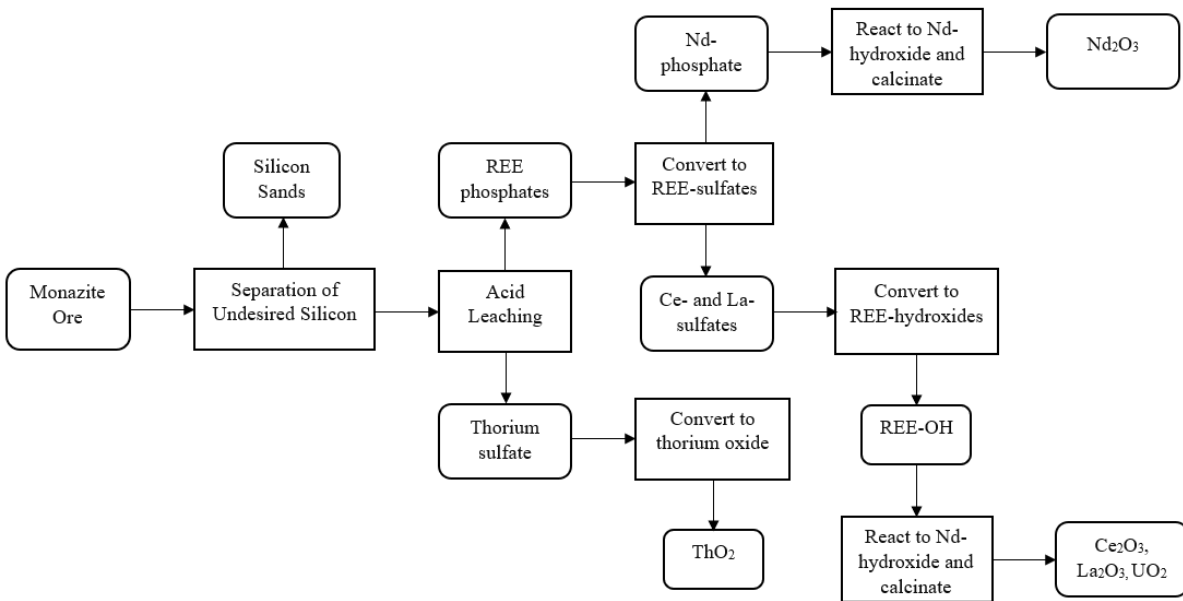
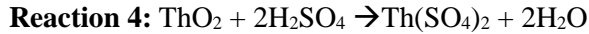


Figure 1. General Monazite Ore Extraction Process Flowsheet

2.2 Process Chemistry

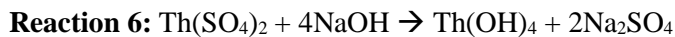
The process begins by separating the silicon from the raw monazite ore before sending the remainder of the monazite into the first reactor. In reactor one, the process continues with the leaching of the REEs from monazite. Inputs of sulfuric acid and water will be used to achieve this leaching through the following reactions:





Next, the resulting thorium and REE sulfates from the first reactor will be cooled in a heat exchanger to lower the temperature to 20°C and crystallize out the thorium sulfate. The cerium and lanthanum sulfates are also converted back to phosphates in this step. A filter is used to separate out the solid thorium sulfate from the REE and remaining compounds.

The crystallized thorium sulfate is then sent into reactor 2, along with an external feed of water and sodium hydroxide. The thorium sulfate reacts with the sodium hydroxide at 150°C to form thorium hydroxide. In the same reactor, the thorium hydroxide is converted to thorium oxide and water.

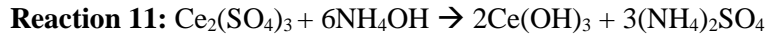


A filter is then used to separate the solid thorium oxide into a purified product stream, producing 200 mol/hr of thorium oxide product. The aqueous stream from the filter is sent through an evaporator to separate off a water vapor stream of 7550 mol/hr from a waste stream of about 455 mol/hr of remaining H₂O, SO₃, Na₂SO₄, and NaOH. Finally, the water vapor stream is sent through a heat exchanger at 25°C and recycled back into reactor 1.

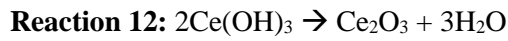
The REE are dealt with separately after the split in filter 1. In reactor 3, the cerium and lanthanum phosphates are converted back to the sulfate form using the sulfuric acid at 300°C. The neodymium does not react under these conditions.



Another filter is used to separate the solid sulfate products, including the cerium, lanthanum, and uranium, from the unreacted neodymium. The sulfates are sent into reactor 4 along with an external feed of NH_4OH at 9000 mol/hr. The cerium and lanthanum are converted to REE-hydroxides at 170°C , while the uranium sulfate reacts to form uranium oxide.



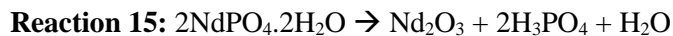
The remainder of the products from reactor 4 are sent into a calciner at 900°C so that the REE-hydroxides are converted to the REE-oxide product form.



The neodymium stream from the filter 3 separation is more difficult to react and must be dealt with under its own conditions; thus, the neodymium phosphate from this separation is sent into reactor 5, in which an external feed of NaOH at 20,000 mol/hr is also added. The streams react at 325°C to convert the neodymium phosphate into its hydrated form.



The stream from reactor 5 is then sent through a filter to separate the aqueous $\text{NdPO}_4 \cdot 2\text{H}_2\text{O}$ from the remaining compounds, which leave as a waste stream. Finally, the neodymium is sent through a calciner at 900°C to convert it to the product stream of neodymium oxide.



2.3 Literature Summary

Thorium is a mildly radioactive element that is more than three times more abundant and half as dense as uranium and is sourced mainly from monazite sands. Thorium has many uses as an

alloy, as addition of thorium in small quantities can vastly improve the strength of the alloy.

Thorium occurs naturally as the stable isotope ^{232}Th , which has a half-life similar to the age of the universe itself. Currently, the Earth contains around 85% of the thorium that was present at the time of the Earth's formation [2]. Thorium has previously been used as a catalyst or in ceramics but has since shifted to use mainly as a power source since the discovery of its radioactivity.

Thorium(IV) oxide, also known as thorium dioxide, is the main product extracted from monazite in this process and is a valuable nuclear fuel source. Because of its extremely high melting point, it is also particularly useful in high-temperature applications such as combustion engines [4].

Rare earth elements (REEs) consist of seventeen elements that are grouped together on the periodic table, fifteen of which are part of the group called lanthanides. While the name suggests that these elements are quite rare, REEs such as cerium, lanthanum, and neodymium are more prevalent in the Earth's crust than more commonly-known elements such as lead and silver [5]. However, these REEs are often found dispersed, rather than in concentrated quantities. As a result, they have been underutilized in industry, even though their diverse chemical, electrical, magnetic, optical, and metallurgical properties make them highly useful. REEs have many uses in applications that require elements of high specificity. Cerium, which is extracted in this process from monazite and is one of the most abundant REEs, has a wide variety of applications. Specifically, cerium oxide is used as a glass polishing agent. Lanthanum and neodymium, both of which are also extracted in this process, are used in the oil refining industry as a cracking catalyst and as a lightweight, high-strength alloy in magnet technology, respectively [3].

REEs exist in many different forms, including oxides, phosphates, and halides. The monazite ore used in this process contains sources of REE phosphates, which will be converted into REE oxides during the extraction and purification process. Monazite also contains varying

amounts of thorium and uranium. Monazite contains, on average, 2.5% thorium, but may contain up to 10% thorium oxide [6]. It is currently the main source for thorium, lanthanum, and cerium, and is most prevalently sourced from India, Indonesia, South Africa, and Madagascar. While monazite will be used only as a source of thorium, uranium, and REEs for the purposes of this project, monazite's radioactivity can also be utilized for geochronology to study geological events [8].

2.4 Basic Process Economics

The main goal of this study is to analyze the economic viability of thorium oxide recovery from monazite ore. Considerations that went into the economic viability of the process include the costs of raw materials, equipment and manufacturing costs, and the recovery of our major product thorium oxide and sellable byproducts SiO_2 and the other REE oxides. Contained in Table 1 is the cost information for the products and raw materials.

Table 1. Raw material, product, and byproduct costs

Species	Cost/Unit
Monazite	\$1.00/kg
H_2SO_4	\$0.04/kg
ThO_2	\$80.00/kg
P_2O_5	\$0.88/kg
La_2O_3	\$2.00/kg
CeO_2	\$2.00/kg
Nd_2O_3	\$60.00
NH_4OH	\$0.20/kg
Natural Gas	\$4.40/GJ

3.0 Method of Approach

Rare earth oxides were separated with the following process. First, silicon is separated from raw monazite ore with a flotation column. Then, the monazite enters a leaching vessel, along with sulfuric acid and water. The products from this reactor flow through a heat exchanger and another flotation column, where thorium sulfate is separated from the rest of the rare earth metals (in their phosphate forms).

The thorium sulfate product is fed into another reactor with sodium hydroxide and water to obtain the final desired product- thorium oxide. The thorium oxide is separated into its own purified product stream via floatation column, and the remaining byproducts are sent to an evaporator. Any water that can be recycled is sent through a heat exchanger and back to the first reactor, where the original acid leaching reaction takes place. The rest of the byproducts are sent off-site through a separate waste stream.

The rest of the rare earth phosphates are sent to a separate reactor to transform them into rare earth sulfates, and the neodymium products are separated from the lanthanum, cerium, and uranium products via floatation column. The lanthanum and cerium are further converted to hydroxides via reaction with ammonium hydroxide, all are crystalized with a calciner, and the final mixed stream of cerium oxide, lanthanum oxide, and uranium oxide are sent offsite to be sold. The neodymium product is treated similarly, except it is reacted with sodium hydroxide, resulting in a sulfite waste stream and a neodymium oxide stream that can also be sold.

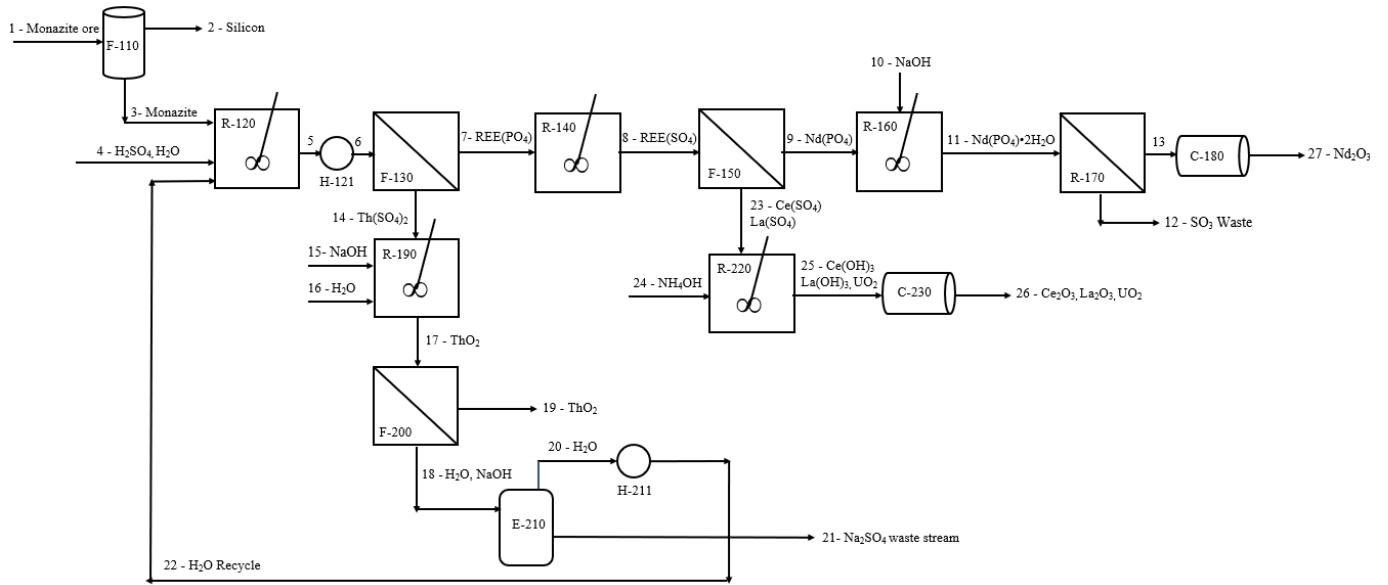


Figure 2. Flow diagram for thorium extraction from monazite ore

4.0 Results

4.1 Optimization

A number of measures were taken in order to ensure optimization of the thorium extraction process, including stream flow reduction and implementation of recycle streams. Originally, the neodymium-phosphate was being converted to its final product state through use of ammonium hydroxide. This process required huge flow rates and thus a very large and expensive reactor in which to sustain the conversion. Once the ammonium hydroxide was switched to sodium hydroxide, however, the required flow rate for conversion was drastically reduced; this allowed for the reactor size to be greatly reduced as well, which reduced the plant's manufacturing cost to about half of what it was before.

After all the expected conversions were reached and the products were isolated, the inlet feed streams were reduced to the greatest extent possible as well. The amount of water and sulfuric

acid, specifically, into reactor 1 were minimized to the point where the least flow rate could still achieve the desired conversions. Finally, a recycle stream was implemented to reuse the water coming out of reactor 2 and recycle it back into reactor 1. This allowed for the feed water stream to be reduced even further, saving in both cost and environmental concerns.

Table 2. Equipment List

Equipment List			
#	Part	Material	Purpose
5	Reactor	stainless steel	Convert raw materials into desired products
2	Calciner	firebrick	Convert REE hydroxides into REE oxides
1	Evaporator	stainless steel	Separate water vapor from waste products
3	Filter	stainless steel	Separate solids; Purify streams
2	Flotation Column	stainless steel	Uranium separation
2	Heat Exchanger	carbon steel / stainless steel	Crystallize La and Ce; Adjust stream temperatures for optimal reaction conditions

4.2 Process Flow Diagram

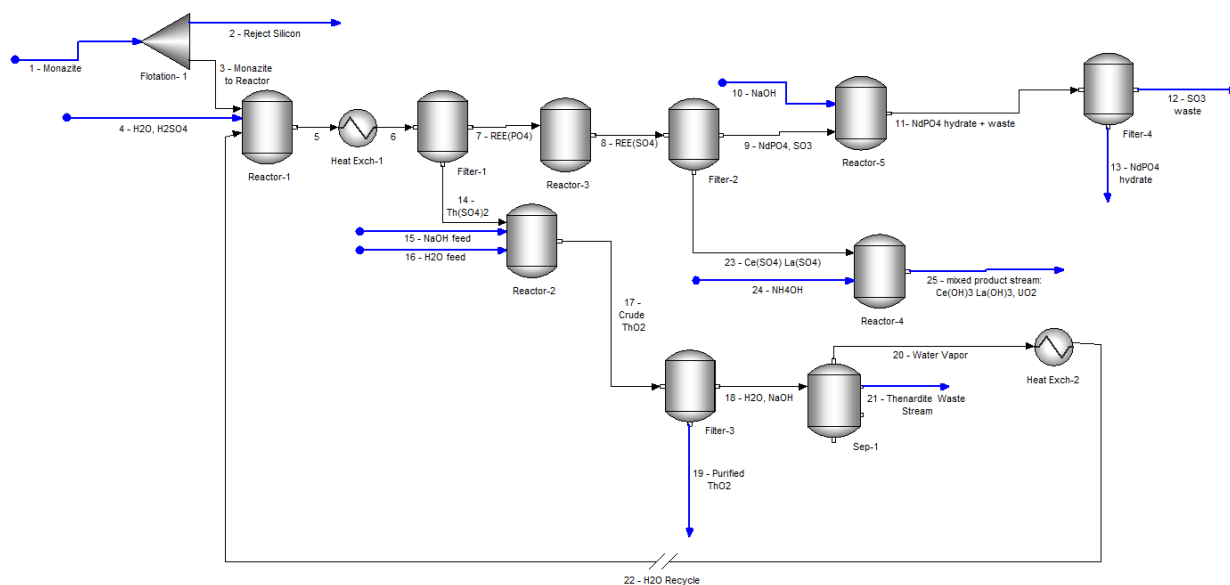


Figure 3. OLI flowsheet of process with stream numbers

4.3 Safety, Health, and Environmental Analysis

The environmental and safety effects of the chemical process were first taken into account when designing the extraction process through inherently safer design. Since the extraction process utilizes acid leaching to isolate the rare earth elements from the monazite ore, certain safety hazards were inevitable. An attempt to choose an acid that possessed the least hazardous properties while maintaining high conversion of the rare earth and thorium phosphates to sulfates was made. After examining several chemicals, sulfuric acid was ultimately used in the leaching process because of its conversion abilities, despite the relatively high corrosion and irritability hazards. The amount of this chemical used in the process was minimized, however, in order to reduce the risk of exposure and accidental release. Moreover, the overall design of the process was simplified to contain as few reactors as possible to reduce risk and ensure process safety. While additional reactors and equipment could be installed to improve conversion, the safety and economic benefit outweighs the potential for higher conversion.

Another major factor that went into the process design was the environmental impact of the extraction, including feed, product, and waste streams. As in any process involving use of chemical reactions, it is important to keep byproducts and their disposal in mind in order to control pollution and keep the environment clean. Pollution can come from many different points in a process, however, not just the end products. Fugitive emissions, for instance, can escape from pipe joints, vents, seals, and other minor areas that are often overlooked and thus make this type of pollution difficult to control. Engineered emissions are typically the most important pollution to control as they stem from the actual reactor and separator processes which can be designed to avoid, reduce, and minimize pollution hazards [2].

One way in which this particular process could be modified to reduce pollution is through implementation of recycle streams so that the byproducts will be reduced. This will reduce waste and ultimately benefit the production economically as well if the resources can be reused properly. The water from the thorium oxide purification stream, for instance, was vaporized off after thorium separation was achieved. This stream was then sent through a heat exchanger to condense the water back to a liquid state before being recycled back to reactor 1. For those streams which a recycle loop was not economically beneficial, proper storage or disposal should be implemented so that they may not pollute the environment. This can be done through waste storage tanks or landfills after the waste is reduced to the smallest volume possible. As such, the uranium oxide stream was separated completely from other waste streams so that the radioactive waste could be disposed of in the smallest amount.

Finally, the health and safety of the workers was also considered in the process design as a critical factor. Because of the potential hazards of the chemicals utilized in the extraction process, which can be seen on the next page in Table 3, various safety precautions and standards should be put in place. Workers dealing with the acid leaching process, for instance, should wear appropriate safety gear, including long sleeves and pants, gloves, and goggles to protect them from dermal exposure. Those workers in areas where potentially hazardous gaseous emissions are present should additionally gas wear masks to protect them from inhalation exposure. Another potential hazard stems from the high temperatures used in the calcination steps at the end of the process. Workers involved in this reaction step should be mindful of temperature and pressure sensors at all times. Since accidental spills and runaway reactions are possible despite the safety factors that went into designing the process, workers should be well trained and prepared to respond with emergency mitigation and response procedures.

Table 3.a. Stream hazards chart

Material	MW (g/mol)	Melting Point (°C)	Boiling Point (°C)	Density (kg/m ³)	Hazards
Air	26.98	-216.2	-194.3	1.225	Hazardous in the case of compression; store away from flammable/combustible materials, corrosive atmospheres
NH ₄ OH	35.05	-69.2	36	880	Very hazardous; skin contact - corrosive, irritant, permeator; eye contact - irritant; extremely corrosive in the presence of zinc, copper, aluminum
NaOH	39.997	318	1388	2130	Corrosive - metals, skin, eye damage; irritant; non-flammable; reactive with metals; chemically stable under normal conditions
Monazite	240.21	~2000	-	5.15	Highly toxic in the case of inhalation and ingestion (kidney, liver, lungs, brain); skin contact - irritant; eye contact - irritant; carcinogen; radioactive
H ₂ SO ₄	98.08	10.31	337	1.8302	Very hazardous; skin contact - corrosive, irritant, permeator; eye contact - irritant, corrosive; ingestion; inhalation; carcinogen
H ₂ O	18.02	0	99.98	0.99	Non-toxic; non-flammable; corrosive to metals
Th(SO ₄) ₂	424.15	-	-	2.8	Highly flammable; low toxicity
Th(P ₂ O ₇)	406	-	-	.00367	Highly flammable; low toxicity; moderate irritant
SiO ₂	60.09	1610	2950	2.65	Slightly hazardous; skin contact - irritant; eye contact - irritant; ingestion; inhalation; non-flammable
Th(OH) ₄	300.07	-	942	-	Low flammability; low toxicity

Table 3.b. Stream hazard flow chart (continued)

Material	MW (g/mol)	Melting Point (°C)	Boiling Point (°C)	Density (kg/m ³)	Hazards
P ₂ O ₅	141.943	340	360	2.39	Extremely hazardous: skin contact - corrosive, irritant; eye contact - irritant; inhalation; very hazardous: ingestion; low flammability; non-corrosive to glass
REE(OH) ₃	192	2315	4200	6.65	Hazardous; skin contact - corrosive, irritant; eye contact - irritant; inhalation; toxic; non-flammable
REE ₂ O ₃	330	2315	4200	6.65	Hazardous in case of ingestion; skin contact - irritant; eye contact - irritant; inhalation; non-flammable
ThO ₂	264	3200	4400	5.15	Toxic - inhalation, ingestion, skin absorption; potential carcinogen; low flammability
U ₃ O ₈	842.1	1150	1300 (decomposes to UO ₂)	8300	Radioactive; toxic - inhalation, ingestion, skin and eye irritant
Na ₂ SO ₄	142.04	884	1429	2660	Hazardous in case of ingestion - may cause gastrointestinal irritation; slightly hazardous in case of skin or eye contact

4.4 Capital cost estimates

All of the equipment was costed via the volumetric flow rates given from OLI flowsheet. When sizing the five reactors, it is necessary to convert the volumetric flow rate to a volume using a residence time and then choosing a diameter to find a height of our desired process vessel. From there using Figure 5.44 in Ulrich [7] to find the Cp value and the pressure of the column in OLI, a stainless-steel reactor can be sized and converted using the current yearly CEE Index. When

designing the two heat exchangers, it is necessary to calculate the log-mean temperature difference from the inlet and outlet flow rates. From there an overall heat transfer coefficient was calculated via a slurry viscosity equation where an example calculation can be seen in Appendix C. Using the heat duty from OLI as well as the overall heat transfer coefficient and the log-mean temperature difference, the area can be calculated and then a Cp value from Figure 5.36 can be gathered. The heat exchangers were designed to be made of carbon steel but have the tubing lining be made of stainless steel. The two heat exchangers in our process vary with respect to what is being carried through them however stainless steel was the most logical choice for most of our equipment due to its lack of corrosion. For the two calciners in our process, the vessels were designed similar to reactors however using Figure 5.33, the calciner was designed using brick lining. The flotation column was sized as stainless-steel process vessels with stainless steel filters included inside and sized similar to the reactors. Four filters were designed using Figure 5.57b in Ulrich as either bag filters to separate out the solids from the liquids or as a liquid cyclone. The separatory column designed at the end of the process was useful to recycle the water to be pumped back into the first reactor and was designed as an evaporator. The evaporator was designed using stainless steel as well. Finally, five pumps were designed using stainless steel in order to account for the pressure increase at two of the reactors in the system. Overall our total cost of all of our equipment sums up to be \$18,418,316. The various percentages of the types of equipment are partitioned out in Figure 4 on the following page.

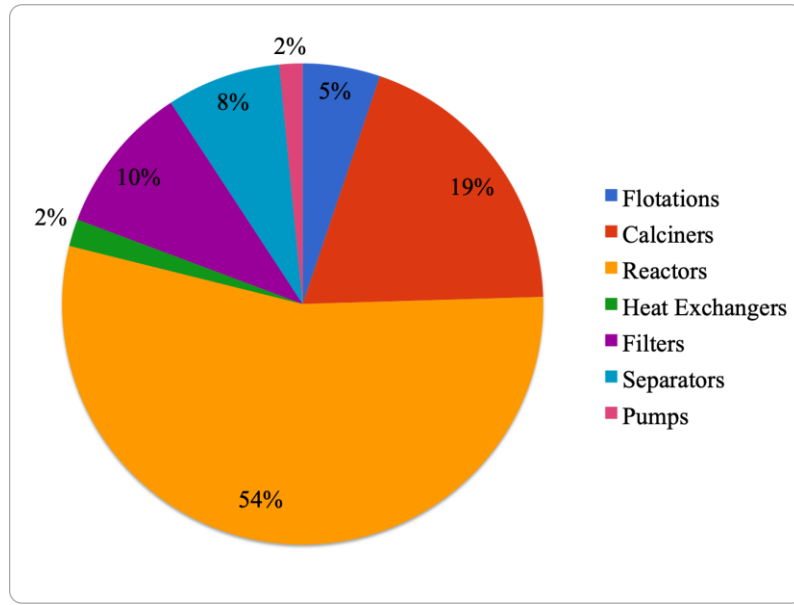


Figure 4. Distribution of equipment costs

4.5 Manufacturing cost estimates

The annual operating costs were calculated in various ways to account for the various expenditures required to maintain and operate a plant this size. The largest expense is the raw materials required to keep the plant running. The costs for these materials are estimated by converting the flowrate input of the initial materials to the price of those materials using Table 1 in the basic economics section above to a dollar amount per year. The next largest expenditure was the indirect manufacturing costs. These costs include overhead charges as in payroll and plant charges. Additionally, indirect costs include local taxes as well as insurance which is about 1% of the fixed capital for each of those. Next is the maintenance costs to keep the plant up and running. We estimated this cost to be about 8% of the fixed capital in order to account for the maintenance made throughout the year. Based on the breakdown of costs in Figure. 5, 8% of the annual manufacturing costs go towards the utilities of the plant. These include steam for processes, electricity of the facility, and cooling water for all of the heating-based processes. Lastly the

operating supplies and labor account for the last 1% which is necessary to fund the supervisory and clerical labor of our plant as well as pay for various supplies necessary for those workers. Overall, the total annual operating costs of this facility is \$19,066,557.

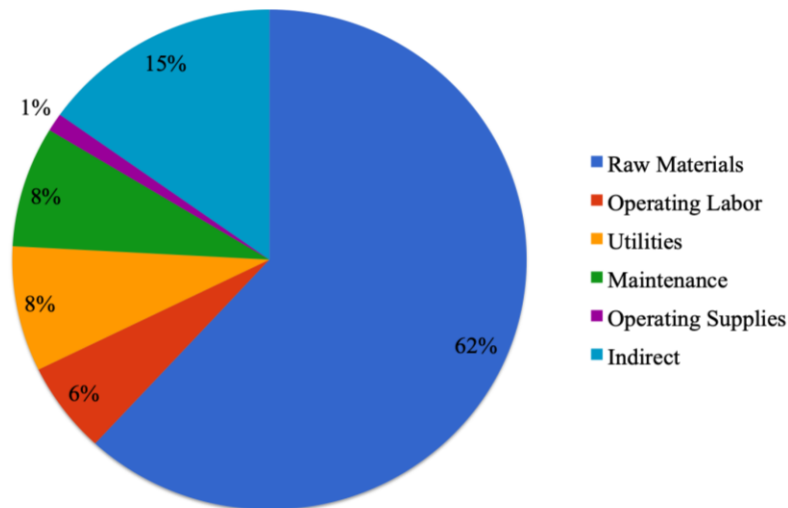


Figure 5. Distribution of manufacturing expenses

5.0 Discussion of Results

The extraction of thorium oxide from monazite ore was found to have a highly favorable economic potential. As detailed in the appendices below, the cost for raw materials will be just under \$11.7 million per year. The fixed capital cost was \$18.4 million, with the amount including 48% extra for contingency due to this plant being grassroots. Annual expenses for the plant, including the \$11.7 million for raw materials, total to \$19 million. Assuming the plant runs at 100% possible efficiency, the revenue would be as high as \$114 million, yielding a return on investment (that is, profit minus expenses, all divided by expenses) of 151%.

6.0 Conclusions

Based on the economic analysis, an estimated \$114 million per year could be earned from this process. It was also shown that, by factoring in expenses such as capital costs (\$18.4 million) and manufacturing costs (\$19 million), the plant could achieve a return on investment of 151%. However, this number comes from many assumptions built into our pricing model and is likely to be less by the end of construction. Most of the profit is driven by the neodymium and thorium oxide products (both selling for \$60/kg and \$80/kg respectively), so any inefficiencies affecting these streams would have a large impact on revenue. The revenue estimates are also based on lanthanum and cerium being pure components; however, they are currently present in a mixed stream (along with uranium oxide). Further separations and handling of any radioactive wastes would present significantly more equipment and manufacturing costs. Additionally, the reactants are highly corrosive; equipment would need regular maintenance and repair, but these costs have not been factored into the annual maintenance cost. Finally, while a thorough ventilation system is highly recommended for our plant, the costs associated with installation and maintenance of such system has not been included. Still, the potential for profit is quite high, and we have confidence in the economic viability of this chemical plant.

7.0 Recommendations

There are several recommendations for the design and operation of this plant that extends beyond the scope of this project. The waste stream containing Uranium will be highly toxic and radioactive and should thus be handled appropriately in order to avoid contamination and ensure the safety of the plant's workers and community. This includes proper isolation and storage of the Uranium waste both on- and off-site. Additionally, the acid used in the leaching process, sulfuric

acid, is highly corrosive and is a strong irritant. The use of this chemical poses potential risks to workers' health, as any skin or eye contact is extremely dangerous, especially in the large amounts used for this process. In order to minimize risks, the amount of sulfuric acid used in the process should be minimized, and a simplified plant layout should be used in order to isolate the areas in which the chemical is used. An extensive ventilation system should also be installed throughout the plant in order to prevent worker inhalation of toxic chemicals, such as sulfuric acid, or dust particles.

There are also several waste streams in the current process that could be further purified or utilized in other industries. As it stands, the process currently produces Ce_2O_3 and La_2O_3 in the same calciner. This stream could be further purified in order to separate the two products to be sold separately. Additionally, sodium Sulfate appears in several of the waste streams, and could be sold off-site for use in the glass or paper industry, where it is used in the manufacturing of detergents or in paper pulping. There are also fairly large quantities of water, NaOH, and NH_4OH being disposed of in several of our waste streams, each of which are used as raw materials or reagents at some point in the process. If economically feasible, these streams could be purified in order to recycle each of these compounds throughout our process to reduce material costs.

8.0 References

- [1] Calculating Physical Properties Of Slurries. (2013). Retrieved from <https://www.cheresources.com/invision/blog/4/entry-340-calculating-physical-properties-of-slurries/>
- [2] Emsley, J. "*Emsley block: an A-Z guide to the elements.*" Oxford University Press, pages 544-548 (2011).

- [3] Haxel, G. Hedrick, J. and Orris, J. “*Rare Earth Elements - Critical Resources for High Technology*,” USGS, (2002).
- [4] Mitchell, Brian S. “*An Introduction to Materials Engineering and Science for Chemical and Materials*,” John Wiley & Sons, page 473 (2004).
- [5] Rodliyah, S. Rochani and T. Wahyudi, “*Extraction of Rare Earth Metals from Monazite Mineral using Acid Method*”, Indonesian Mining Journal 18(1), pages 39 - 45 (2015).
- [6] Saleh, F. A. and Farag, A. K. “*Recovery of Uranium, Thorium and Cerium from Egyptian Monazite Sands*,” Journal of Inorganic and General Chemistry, (1970).
- [7] Ulrich, Gael D. *A Guide to Chemical Engineering Process Design and Economics*. Wiley (1984).
- [8] Zhu, X. K. and O’Nions, R. K. “*Monazite chemical composition: some implications for monazite geochronology*,” Contrib Mineral Petrol, pages 351-363 (1999).

Appendices

Appendix A: Mass and Molar Flow Rates (by stream)

Table A. 1. Mass flow rates (streams 1-14)

Component	MW (kg/kmol)	Streams (kg/hr)													
		1	2	3	4	5	6	7	8	9	10	11	12	13	14
H ₂ O	18.02	-	-	-	2252.5	2925.583	2925.583	2925.583	2925.583	-	2915.167	2915.167	-	-	
CePO ₄	235.086	465.4703	-	465.4703	-	39.64302	371.765	371.765	0.014747	0.014747	0.014747	0.014747	-	-	
LaPO ₄	236.899	234.53	-	234.53	-	190.8293	234.53	234.53	0.041208	0.041208	0.041208	0.041208	-	-	
NdPO ₄	239.213	189.3912	-	189.3912	-	189.3912	189.3912	189.3912	189.3912	189.3912	2.577209	2.577209	-	-	
NH ₄ OH	35.04	-	-	-	-	-	-	-	-	-	-	-	-	-	
NaOH	40	-	-	-	-	-	-	-	-	800	-	-	-	-	
H ₂ SO ₄	98.08	-	-	-	2942.4	-	-	-	-	-	-	-	-	-	
ThO ₂	264.04	52.808	-	52.808	0.00103	7.9E-07	7.9E-07	7.9E-07	2.46E-08	2.46E-08	-	2.46E-08	-	-	
SiO ₂	60.08	47.4632	47.4632	-	-	-	-	-	-	-	-	-	-	-	
UO ₂	270.03	0.54006	-	-	-	7.15E-07	0.54006	0.54006	1.29E-10	1.29E-10	-	-	-	-	
Th(OH) ₄	300.07	-	-	-	-	-	-	-	-	-	-	-	-	-	
Ca(OH) ₂	74.093	-	-	-	-	-	-	-	-	-	-	-	-	-	
SO ₃	80.07	-	-	-	-	2161.762	2401.772	2401.772	2044.764	2044.764	1965.975	1965.975	-	-	
Ce ₂ (SO ₄) ₃	568.42	-	-	-	-	514.8089	-	-	562.7182	-	-	-	-	-	
La ₂ (SO ₄) ₃	565.9987	-	-	-	-	52.20489	-	-	280.1201	-	-	-	-	-	
P ₂ O ₅	141.9445	-	-	-	-	113.5538	-	-	182.6755	182.6755	182.6755	182.6755	-	-	
Th(SO ₄) ₂	424.1633	-	-	-	-	84.83096	84.83266	-	1.23E-06	-	-	-	-	84.83266	
U(SO ₄) ₂	334.085	-	-	-	-	0.66817	-	-	0.66817	-	-	-	-	-	
Ce ₂ O ₃	328.24	-	-	-	-	-	64.9692	64.9692	-	-	-	-	-	-	
NH ₃	17.031	-	-	-	-	-	-	-	-	-	-	-	-	-	
NdPO ₄ ·2H ₂ O	275.242	-	-	-	-	-	-	-	-	-	214.9508	-	214.9508	-	
Na ₂ SO ₄	142.04	-	-	-	-	-	-	-	-	-	139.7682	-	-	-	
Ce(OH) ₃	191.14	-	-	-	-	-	-	-	-	-	-	-	-	-	
La(OH) ₃	189.93	-	-	-	-	-	-	-	-	-	-	-	-	-	
(NH ₄) ₂ SO ₄	132.14	-	-	-	-	-	-	-	-	-	-	-	-	-	
Total Flow Rate (kg/hr)		990.2027	47.4632	942.1994	5194.9	6273.276	6273.383	6188.55	6188.976	5342.469	800	5421.17	5066.451	214.9508	84.83266
Pressure (atm)		1	1	1	1	80	80	80	75	75	1	60	60	60	80
Temperature (°C)		25	25	25	25	220	20	20	300	300	25	295	295	295	20

Table A. 2. Mass flow rates (streams 15-27)

Component	MW (kg/kmol)	Streams (kg/hr)														
		15	16	17	18	19	20	21	22	23	24	25	26	27		
H ₂ O	18.02	-	129,744	129,744	129,744	-	136,051	0.850767	129,744	-	-	1,809064	-	82,08184		
CePO ₄	235.086	-	-	-	-	-	-	-	-	-	-	-	-	-		
LaPO ₄	236.899	-	-	-	-	-	-	-	-	-	-	-	-	-		
NDPO ₄	239.213	-	-	-	-	-	-	-	-	-	-	-	-	-		
NH ₄ OH	35.04	-	-	-	-	-	-	-	-	-	315.36	-	-	-		
NaOH	40	32	-	32	32	-	-	0.223009	-	-	-	-	-	-		
H ₂ SO ₄	98.08	-	-	-	-	-	3.71E-18	-	-	-	-	6.47E-22	-	6.47E-22		
ThO ₂	264.04	-	-	52.808	-	52.808	-	-	-	-	-	7.65E-07	-	7.65E-07		
SiO ₂	60.08	-	-	-	-	-	-	-	-	-	-	-	-	-		
UO ₂	270.03	-	-	-	-	-	-	-	-	-	-	0.54006	-	0.54006		
Th(OH) ₄	300.07	-	-	-	-	-	-	-	-	-	-	-	-	-		
Ca(OH) ₂	74.093	-	-	-	-	-	-	-	-	-	-	-	-	-		
SO ₃	80.07	-	-	32.028	32.028	-	2.27E-22	0.223203	-	-	-	1.095542	-	1.095542		
Ce ₂ (SO ₄) ₃	568.42	-	-	-	-	-	-	-	-	0.001742	-	-	-	-		
La ₂ (SO ₄) ₃	565.9987	-	-	-	-	-	-	-	-	0.000874	-	1.16E-08	-	1.16E-08		
P ₂ O ₅	141.9445	-	-	-	-	-	-	-	-	-	-	-	-	-		
Th(SO ₄) ₂	424.1633	-	-	-	-	-	-	-	-	1.23E-06	-	-	-	-		
U(SO ₄) ₂	334.085	-	-	-	-	-	-	-	-	0.66817	-	-	-	-		
Ce ₂ O ₃	328.24	-	-	-	-	-	-	-	-	-	-	3.09E-06	-	3.09E-06		
NH ₃	17.031	-	-	-	-	-	-	-	-	-	-	1.874687	-	1.874687		
NDPO ₄ ·2H ₂ O	275.242	-	-	-	-	-	-	-	-	-	-	107.4754	-	-		
Na ₂ SO ₄	142.04	-	-	-	-	-	-	56.41999	-	-	-	-	139.7682	-		
Ce(OH) ₃	191.14	-	-	-	-	-	-	-	-	-	-	378.4457	-	189.2229		
La(OH) ₃	189.93	-	-	-	-	-	-	-	-	-	-	187.9977	-	93.99883		
(NH ₄) ₂ SO ₄	132.14	-	-	-	-	-	-	-	-	-	-	587.357	-	587.357		
Total Flow Rate (kg/hr)		32	129,744	246,58	193,772	52,808	136,051	57,71697	129,744	0.670787	315,36	1159,12	247,2436	956,1708		
Pressure (atm)		1	1	50	50	50	13,8425	13,8425	13,8425	75	1	40	40	40		
Temperature (°C)		25	25	150	150	150	200	200	25	300	25	170	900	900		

Table A. 3. Molar flow rates (streams 1-14)

Component	MW (kg/kmol)	Streams (kmol/hr)													
		1	2	3	4	5	6	7	8	9	10	11	12	13	14
H ₂ O	18.02	-	-	-	125	162.352	162.352	162.352	162.352	162.352	-	161.774	161.774	-	-
CePO ₄	235.086	1.98	-	1.98	-	0.168632	1.5814	1.5814	6.27E-05	6.27E-05	-	6.27E-05	6.27E-05	-	
LaPO ₄	236.899	0.99	-	0.99	-	0.80553	0.99	0.99	0.000174	0.000174	-	0.000174	0.000174	-	
NdPO ₄	239.213	0.791726	-	0.791726	-	0.791726	0.791726	0.791726	0.791726	0.791726	-	0.010774	0.010774	-	
NH ₄ OH	35.04	-	-	-	-	-	-	-	-	-	-	-	-	-	
NaOH	40	-	-	-	-	-	-	-	-	-	20	-	-	-	
H ₂ SO ₄	98.08	-	-	-	30	-	-	-	-	-	-	-	-	-	
ThO ₂	264.04	0.2	-	0.2	-	3.9E-06	2.99E-09	2.99E-09	9.31E-11	9.31E-11	-	-	9.31E-11	-	
SiO ₂	60.08	0.79	0.79	-	-	-	-	-	-	-	-	-	-	-	
UO ₂	270.03	0.002	-	-	-	2.65E-09	0.002	0.002	4.77E-13	4.77E-13	-	-	-	-	
Th(OH) ₄	300.07	-	-	-	-	-	-	-	-	-	-	-	-	-	
Ca(OH) ₂	74.093	-	-	-	-	-	-	-	-	-	-	-	-	-	
SO ₃	80.07	-	-	-	-	26.9984	29.9959	29.9959	25.5372	25.5372	-	24.5532	24.5532	-	
Ce ₂ (SO ₄) ₃	568.42	-	-	-	-	0.905684	-	-	0.989969	-	-	-	-	-	
La ₂ (SO ₄) ₃	565.9987	-	-	-	-	0.092235	-	-	0.494913	-	-	-	-	-	
P ₂ O ₅	141.9445	-	-	-	-	0.799987	-	-	1.28695	1.28695	-	1.28695	1.28695	-	
Th(SO ₄) ₂	424.1633	-	-	-	-	0.199996	0.2	-	2.9E-09	-	-	-	-	0.2	
U(SO ₄) ₂	334.085	-	-	-	-	0.002	-	-	0.002	-	-	-	-	-	
Ce ₂ O ₃	328.24	-	-	-	-	-	0.197932	0.197932	-	-	-	-	-	-	
NH ₃	17.031	-	-	-	-	-	-	-	-	-	-	-	-	-	
NdPO ₄ ·2H ₂ O	275.242	-	-	-	-	-	-	-	-	-	-	0.780952	-	0.780952	
Na ₂ SO ₄	142.04	-	-	-	-	-	-	-	-	-	-	0.984006	-	0.984006	
Ce(OH) ₃	191.14	-	-	-	-	-	-	-	-	-	-	-	-	-	
La(OH) ₃	189.93	-	-	-	-	-	-	-	-	-	-	-	-	-	
(NH ₄) ₂ SO ₄	132.14	-	-	-	-	-	-	-	-	-	-	-	-	-	
La ₂ O ₃	325.81	-	-	-	-	-	-	-	-	-	-	-	-	-	
Nd ₂ O ₃	336.48	-	-	-	-	-	-	-	-	-	-	-	-	-	
Total Flow Rate (kmol/hr)	4.754	0.790	3.962	155.000	193.116	196.111	195.911	191.455	189.968	20.000	189.390	187.625	1.765	0.200	
Pressure (atm)	1	1	1	1	80	80	80	75	75	1	60	60	60	80	
Temperature (°C)	25	25	25	25	220	20	20	300	300	25	295	295	295	20	

Table A. 4. Molar flow rates (streams 15-27)

Component	MW (kg/kmol)	Streams (kmol/hr)														
		15	16	17	18	19	20	21	22	23	24	25	26	27		
H ₂ O	18.02	-	7.2	7.2	7.2	-	7.55	0.047212	7.2	-	-	0.100392	-	-		
CePO ₄	235.086	-	-	-	-	-	-	-	-	-	-	-	-	-		
LaPO ₄	236.899	-	-	-	-	-	-	-	-	-	-	-	-	-		
NdPO ₄	239.213	-	-	-	-	-	-	-	-	-	-	-	-	-		
NH ₄ OH	35.04	-	-	-	-	-	-	-	-	9	-	-	-	-		
NaOH	40	0.8	-	0.8	0.8	-	0.005575	-	-	-	-	-	-	-		
H ₂ SO ₄	98.08	-	-	-	-	-	3.78E-20	-	-	-	-	6.6E-24	-	6.6E-24		
ThO ₂	264.04	-	-	0.2	-	0.2	-	-	-	-	-	2.9E-09	-	2.9E-09		
SiO ₂	60.08	-	-	-	-	-	-	-	-	-	-	-	-	-		
UO ₂	270.03	-	-	-	-	-	-	-	-	-	-	0.002	-	0.002		
Th(OH) ₄	300.07	-	-	-	-	-	-	-	-	-	-	-	-	-		
Ca(OH) ₂	74.093	-	-	-	-	-	-	-	-	-	-	-	-	-		
SO ₃	80.07	-	-	0.4	0.4	-	2.83E-24	0.002788	-	-	-	0.013682	-	0.01368		
Ce ₂ (SO ₄) ₃	568.42	-	-	-	-	-	-	-	-	0.989969	-	-	-	-		
La ₂ (SO ₄) ₃	565.9987	-	-	-	-	-	-	-	-	0.494913	-	2.05E-11	-	2E-11		
P ₂ O ₅	141.9445	-	-	-	-	-	-	-	-	-	-	-	-	-		
Th(SO ₄) ₂	424.1633	-	-	-	-	-	-	-	-	2.9E-09	-	-	-	-		
U(SO ₄) ₂	334.085	-	-	-	-	-	-	-	-	0.002	-	-	-	-		
Ce ₂ O ₃	328.24	-	-	-	-	-	-	-	-	-	-	9.42E-09	-	9.4E-09		
NH ₃	17.031	-	-	-	-	-	-	-	-	-	-	0.110075	-	0.11008		
NdPO ₄ ·2H ₂ O	275.242	-	-	-	-	-	-	-	-	-	-	0.390476	-	0.390476		
Na ₂ SO ₄	142.04	-	-	-	-	-	0.397212	-	-	-	-	0.984006	-	0.984006		
Ce(OH) ₃	191.14	-	-	-	-	-	-	-	-	-	-	1.97994	-	0.98997		
La(OH) ₃	189.93	-	-	-	-	-	-	-	-	-	-	0.989826	-	0.49491		
(NH ₄) ₂ SO ₄	132.14	-	-	-	-	-	-	-	-	-	-	4.44496	-	4.44496		
La ₂ O ₃	325.81	-	-	-	-	-	-	-	-	-	-	-	-	-		
Nd ₂ O ₃	336.48	-	-	-	-	-	-	-	-	-	-	-	-	-		
Total Flow Rate (kmol/hr)		0.800	7.200	8.600	8.400	0.200	7.550	0.453	7.200	1.487	9.000	7.641	1.374	10.611		
Pressure (atm)		1	1	50	50	50	13.8425	13.8425	13.8425	75	1	40	40	40		
Temperature (°C)		25	25	150	150	150	200	200	25	300	25	170	900	900		

Appendix B: Economic Results and Analysis

After the necessary calculations to determine fixed capital, the next step was to determine manufacturing costs, annual operating costs, and the final return on investment. The following worksheet, along with the assumptions given in Appendix A, allowed us to determine the first two.

The return on investment was calculated by the following:

$$\text{Return on Investment} = \frac{(\text{Revenue} - \text{Expenses})}{\text{Expenses}} = \frac{114 \text{ mil.} - 45.5 \text{ mil.}}{45.5 \text{ mil.}} = 151\%$$

Appendix C: Equipment Design and Costing Examples

1. Flotation Column (F-110): The sizing and costing of our flotation column is based on SGS Mineral Services on Column Flotation. According to the paper, reasonable dimensions for a flotation column are as follows: diameter within 0.5 m and 5 m, and height within 6 m and 14 m. With these constraints, we determined that a diameter D of 4 m and height H of 8 m would be reasonable for this process. Based on these dimensions and assuming a vertical orientation, the flotation column was priced as a process vessel from figure 5.44 in the Ulrich text, yielding a C_p value of \$70,000. At 0 barg and using Figures 5.45 and 5.46 in the text, a bare module cost C_{bm} of \$431,480 was calculated. Four stainless steel sieve trays were also designed for the flotation column using Figure 5.48 from the Ulrich text [7], contributing a C_{bm} of \$216,972. The total bare module cost for the flotation column and the sieve trays is \$648,453.

$$C_p (D = 4 \text{ m}, H = 8 \text{ m}) = \$70,000$$

$$F_p (P = 0 \text{ barg}) = 1.2; F_m (\text{stainless clad}) = 2.5$$

$$F_p \times F_m = 1.2 \times 2.5 = 3$$

$$F_{bm} (3) = 4$$

$$C_{bm, vessel} = (\$70,000 \times 4 \times (\frac{616.4}{400})) = \$431,480.00$$

$$C_{pss} (D = 4 \text{ m}) = \$10,000$$

$$F_{bm} (\text{stainless steel}) = 2.2$$

Number of trays, $N_{act} = 4$; quantity factor $f_q = 1.6$

$$C_{bm, sieve} = (\$10,000 \times 2.2 \times 4 \times 1.6 \times (\frac{616.4}{400})) = \$216,972.80$$

$$C_{bm, total} = \$431,480.00 + \$216,972.80 = \$648,452.80$$

2. **Reactor (R-120)**: The reactors were priced based on the Ulrich method for costing process vessels, using figures 5.44, 5.45, and 5.46. By obtaining the volumetric flow rates from OLI Flowsheets and setting a residence time, a required volume can be determined. Multiplying the required volume by a factor of 3 will ensure that the reactor can handle any abnormally large volumes during processing. Assuming the reactor is cylindrical in shape, the height can be obtained to determine the price of the reactor using the Ulrich method. Below are the sample calculations for reactor R-120; reactors R-140, R-160, R-190, and R-220 were priced in the same manner.

$$Q \text{ (m}^3\text{/hr)} = 2.08; \text{ residence time (hr)} = 1; V_{req} \text{ (m}^3\text{)} = 2.08 * 1 = 2.08$$

$$V \text{ (m}^3\text{)} = 3 \times V_{req} = 6.24, \text{ set } D = 1 \text{ m}$$

$$V_{cylinder} = (\pi/4)D^2 * H \Rightarrow H = (4/\pi)(V_{cylinder} / D^2)$$

$$H = (4/\pi)(6.24 / 1^2) = 7.9 \text{ m}$$

$$C_p \text{ (H = 8 m)} = \$20,000$$

$$F_p \text{ (P = 80 barg)} = 4.5 ; F_m \text{ (stainless steel)} = 4$$

$$F_p \times F_m = 18$$

$$F_{bm} \text{ (18)} = 33$$

$$C_{bm} = (\$20,000 \times 33 \times (\frac{616.4}{400})) = \$1,017,060.00$$

3. **Heat Exchanger (H-121)**: The heat exchangers were priced based on the Ulrich method for costing shell and tube heat exchangers, using figures 5.36, 5.37, and 5.38. By obtaining the mass

flow rates and heating duty from OLI Flowsheets and setting the inlet and outlet temperatures for the hot and cold streams, a log mean temperature difference can be determined. Assuming a overall heat transfer coefficient U based on the fuel property, the heat exchanger surface area can be determined. From there, finding the purchased equipment cost from figure 5.36 in Ulrich is used to price the exchangers. Heat exchanger H-211 was also priced in the same manner.

$$m \text{ (kg}^3\text{/hr)} = 3,891; \text{ Heat Duty (Q) (kJ/hr)} = -1,871,515; T_{Hi} = 220 \text{ }^\circ\text{C}; T_{Ho} = 20 \text{ }^\circ\text{C};$$

$$T_{Ci} = 10 \text{ }^\circ\text{C}; T_{Co} = 87 \text{ }^\circ\text{C}$$

$$\Delta T_{LM} = \frac{(T_{Hi} - T_{Co}) - (T_{Ho} - T_{Ci})}{\ln\left(\frac{T_{Hi} - T_{Co}}{T_{Ho} - T_{Ci}}\right)} = 47 \text{ }^\circ\text{C}$$

$$U = 500$$

$$A = -\frac{Q}{\Delta T_{LM} * U} = 79.1 \text{ m}^2$$

$$C_p = \$10,000$$

$$F_p \text{ (P = 81 barg)} = 1.2; F_m \text{ (carbon steel and titanium)} = 7.2$$

$$F_p \times F_m = 8.64$$

$$F_{bm} \text{ (Fp x Fm)} = 13$$

$$C_{bm} = (\$10,000 \times 13 \times \left(\frac{616.4}{400}\right)) = \$200,330$$

4. **Filter (F-130):** The filters were priced using the Ulrich method for pricing filters, involving Figure 5.57b. Flow rates were obtained from OLI Flowsheets, and using an estimated residence time, a minimum filter volume was computed. The value was multiplied by a factor of 3 in order to account for any upsets in the process; this way, the actual filter volume could hold larger volumes. By calculating filter volumes and setting a filter diameter $D = 1 \text{ m}$, the area of the filter could be

calculated. With a known area, the C_p and F_{bm} could be found in order to calculate the bare module cost C_{bm} of the filter. Below are the calculations for filter F-130; filters F-150 and F-170 were also priced in this manner. Filter F-200 was priced as a liquid cyclone using Figure 5.55 from the Ulrich text.

$$Q \text{ (m}^3\text{/hr)} = 2.10; \text{ residence time (hr)} = 1; V_{\text{req}} \text{ (m}^3\text{)} = 2.10 * 1 = 2.10$$

$$V_{\text{filter}} \text{ (m}^3\text{)} = 3 \times V_{\text{req}} = 6.31\text{m}^3, \text{ set } D = 1 \text{ m}$$

$$A_{\text{filter}} = (V_{\text{filter}} / D) = 6.31 \text{ m}^3 / 1 \text{ m} = 6.31 \text{ m}^2$$

$$C_p \text{ (Afilter} = 6.31 \text{ m}^2\text{)} = \$80,000$$

$$F_{bm}(\text{polypropylene}) = 3.0$$

$$C_{bm} = (\$80,000 \times 3 \times (\frac{616.4}{400})) = \$369,840$$

5. Evaporator (E-210): The evaporator was priced based on the Ulrich method for costing separation vessels, using figures 5.24, 4.4, and 4.5. By obtaining the mass flow rates and heating duty from OLI Flowsheets and setting the inlet and outlet temperatures for the hot and cold streams, a log mean temperature difference can be determined. Using figures 4.4 and 4.5 from Ulrich, a viscosity of the slurry was calculated and then an overall heat transfer coefficient U was determined assuming forced-convection. From there the surface area can be determined in order to find the purchased equipment cost from figure 5.24 in Ulrich to price the evaporator.

$$m \text{ (kg}^3\text{/hr)} = 193; \text{ Heat Duty (Q) (kJ/hr)} = 2,96,400; T_{Hi} = 150 \text{ } \circ \text{ C}; T_{Ho} = 200 \text{ } \circ \text{ C}; T_{Ci} = 10 \text{ } \circ \text{ C}; T_{Co} = 9 \text{ } \circ \text{ C}$$

$$\Delta T_{LM} = \frac{(T_{Hi} - T_{Co}) - (T_{Ho} - T_{Ci})}{\ln\left(\frac{T_{Hi} - T_{Co}}{T_{Ho} - T_{Ci}}\right)} = 164 \text{ } \circ \text{ C}$$

$$\mu_L = 0.0000147$$

Viscosity of slurry mixture, $\mu_M = \mu_L [1 + 2.5\Phi + 10.05\Phi^2 + 0.00273 \cdot \exp[16.6\Phi]] = 0.00402$

$$U(\mu_M = 0.004026) = 110$$

$$A = -\frac{Q}{\Delta T_{LM} * U} = 16 \text{ m}^2$$

$$C_p = \$100,000$$

$$F_{bm} \text{ (stainless steel forced circulation tubular)} = 6.2$$

$$C_{bm} = (\$100,000 \times 6.2 \times (\frac{616.4}{400})) = \$955,420$$

6. **Calciner (C-180):** The calciners were priced based on the Ulrich method for costing rotary and direct vertical tower contactors, using figure 5.33. By obtaining the volumetric flow rates from OLI Flowsheets and setting a residence time, a required volume can be determined. Multiplying the required volume by a factor of 3 will ensure that the calciner can handle any abnormally large volumes during processing. Assuming the calciner is a rotary dryer/calciner/steam tube in Ulrich 5.33, a C_p value can be found as shown below. Below are the sample calculations for calciner C-180; calciner C-230 was priced in the same manner.

$$Q \text{ (m}^3\text{/hr)} = 0.5566; \text{ residence time (hr)} = 1; V_{\text{req}} \text{ (m}^3\text{)} = 2.08 * 1 = 1.67$$

$$C_p \text{ (Vreq)} = \$90,000$$

$$F_{bm} \text{ (bricklined)} = 4$$

$$C_{bm} = (\$90,000 \times 4 \times (\frac{616.4}{400})) = \$554,760$$

7. **Pumps:** Five pumps were created and priced for this process: three pumps were designed for the first reactor R-120, and two for the neodymium reactor R-170. The Ulrich method and figures 5.49, 5.50, and 5.51 from the text were used in order to price the pumps. In order to determine a C_p value for the pumps, shaft power was calculated for each pump based on work and efficiency.

Work was derived from volumetric flow rates, pressure differences, and efficiencies. Efficiencies were derived from volumetric flow rates, liquid viscosities (using an equation from cheresources [1]), and volume fractions. The sample calculations for one of the pumps for reactor R-170 is shown below; the four other pumps were priced in a similar manner.

$$Q \text{ (m}^3\text{/s)} = 2.92 \times 10^{-5}$$

$$\text{Liquid viscosity, } \mu_L = 0.01$$

$$\text{Liquid volume fraction, } \Phi = 1$$

$$\text{Viscosity of slurry mixture, } \mu_M = \mu_L [1 + 2.5\Phi + 10.05\Phi^2 + 0.00273 \cdot \exp[16.6\Phi]] = 44.22$$

$$\text{Efficiency, } \varepsilon = 0.14$$

$$Q \text{ (m}^3\text{/s)} = 2.92 \times 10^{-5}; \Delta P = 79 \text{ atm}$$

$$\text{Work, } W = Q * (\Delta P / \varepsilon) = (2.92 \times 10^{-5}) * (79 / 0.14) = 0.034$$

$$\text{Power} = W / \varepsilon = 0.24 \text{ kW}$$

$$C_p \text{ (Power = 0.24 kW)} = \$3200$$

$$F_p \text{ (60 barg)} = 2.25 ; F_m \text{ (rotary positive displacement, stainless steel)} = 2.0$$

$$F_p \times F_m = 4.5$$

$$F_{bm} (4.5) = 9$$

$$C_{bm} = (\$3200 \times 9 \times (\frac{616.4}{400})) = \$44,380.8$$