Recovery of Thorium from Monazite Ore for Applications in Clean, Sustainable Nuclear Power

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CBE 488: Honors Design Internship in Green Engineering

Submitted:
May 1, 2020

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1.0 Introduction

This report serves as documentation for the results of a study-level design and economic analysis of a process extracting thorium (Th) and rare earth elements (REEs) from monazite ore. In the process being analyzed, thorium is extracted from monazite ore using sulfuric acid leaching and a series of neutralization, filtration, and drying steps. In addition to thorium, REEs are extracted from monazite ore as byproducts of the process. More specifically, the REEs extracted are cerium (Ce), lanthanum (La), and neodymium (Nd). After extraction from monazite ore, thorium can be used in the generation of thorium-based nuclear power, in which the nucleus of a thorium atom accepts a neutron and beta decays into isotope uranium-233. Then, nuclear fission of the isotope uranium-233 may produce fuel.

This project originated from efforts by the Electric Power Research Institute (EPRI) to design a process for the recovery of thorium and REEs from monazite while maintaining profitability of the process. This report will begin with an outline of the synthesis information for the process, including the chemical reactions, process constraints, block flow diagram, and relevant cost information. Then, the project's method of approach will be outlined, focusing on the equipment details of the process. After outlining the method of approach, the results of the project are analyzed, including the optimization, final process flow diagram, economic viability, capital costs, and manufacturing costs of the process. This report will end with conclusions and further recommendations regarding future design and analysis of thorium-recovery processes.

The scale of the process was determined by setting the monazite ore feed rate to 1000 kg/hr. The monazite ore was considered to have the empirical formula $Ce_{0.5}La_{0.5}Nd_{0.2}Th_{0.05}(PO_4)$ and the following weight percent composition: 22.36% lanthanum as $LaPO_4$, 37.08% cerium as $CePO_4$, 20.64% thorium as $Th(PO_4)_2$, 19.92% neodymium as $NdPO_4$, and trace amounts of uranium and silicon. Uranium (in PO_4 form) and silicon (as SiO_2) were regarded as impurities in low concentration. The disposition and consequences of the presence of these impurities in the feed material were considered in the process design and were further discussed in this report. The product purity is expected to be greater than 90%. All costs associated with the process were adjusted using a Chemical Engineering Index of 596.2 for January 2020.

2.0 Synthesis Information for Processes

2.1 Block Flow Diagram

The thorium-recovery process was designed to extract thorium (Th) and rare earth elements (REEs) from monazite ore, using a monazite ore feed rate of 1000 kg/hr. The recovery of thorium and REEs is accomplished using sulfuric acid leaching and a series of neutralization, filtration, and drying steps, as shown in Figure 1.

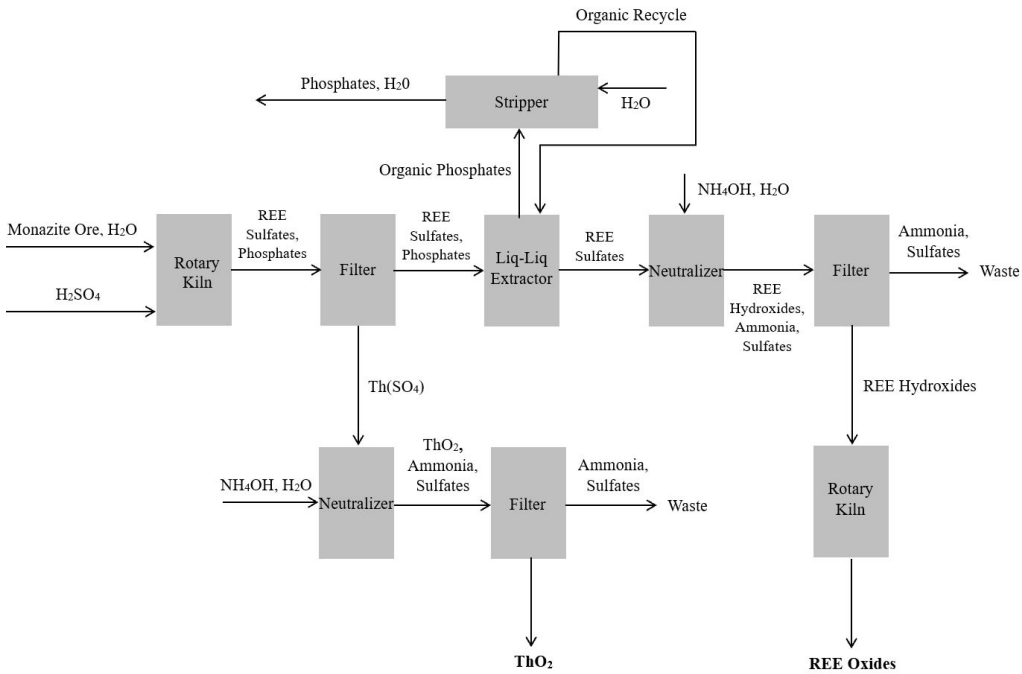

Figure 1. *Block Flow Diagram of the Thorium Recovery Process*

2.2 Process Constraints

To minimize the safety hazards associated with this process, certain safety constraints must be placed on the operation of the process. First, the running temperatures through the system must be maintained below the auto-ignition temperature of all materials used. Second, the handling and disposing of waste streams must be done within federal regulations. Since the process involves acid leaching and neutralization with strong acidic and basic solutions, respectively, special preventative steps must be taken to prevent leakage and protect process employees, the community, and the environment. Additional constraints include monitoring the

radioactivity associated with the thorium extracted from the process to ensure that it does not cause unintended harm to the process employees, the community, and the environment.

2.3 Process Chemistry

The process begins with sulfuric acid leaching of monazite ore in the rotary kiln, as shown in Figure 1. Through sulfuric acid leaching, the thorium and REEs contained within the monazite ore are converted to thorium and REE sulfates. The process chemistry associated with this step is shown in the following reactions:

Reaction 1: $2CePO_4 + 3H_2SO_4 \rightarrow Ce_2(SO_4)_3 + 2H_3PO_4$

Reaction 2: $2LaPO_4 + 3H_2SO_4 \rightarrow La_2(SO_4)_3 + 2H_3PO_4$

Reaction 3: $2NdPO_4 + 3H_2SO_4 \rightarrow Nd_2(SO_4)_3 + 2H_3PO_4$

Reaction 4: $Th(HPO_4)_2 + 2H_2SO_4 \rightarrow Th(SO_4)_2 + 2H_3PO_4$

After sulfuric acid leaching, the products from the rotary kiln are reacted with ammonium hydroxide in two separate neutralization tanks: one for the neutralization of thorium sulfate to thorium oxide, and another for the neutralization of REE sulfates to REE oxides. The process chemistry associated with these steps is shown in the following reactions:

Reaction 5: $Th(SO_4)_2 + 4NH_4OH \rightarrow Th(OH)_4 + 2(NH_4)_2SO_4$

Reaction 6: $La_2(SO_4)_3 + 6NH_4OH \rightarrow 2La(OH)_3 + 3(NH_4)_2SO_4$

Reaction 7: $Ce_2(SO_4)_3 + 6NH_4OH \rightarrow 2Ce(OH)_3 + 3(NH_4)_2SO_4$

Reaction 8: $Nd_2(SO_4)_3 + 6NH_4OH \rightarrow 2Nd(OH)_3 + 3(NH_4)_2SO_4$

After neutralization, the thorium and REE oxides are filtered and dried from aqueous waste. The process chemistry associated with these steps is shown in the following reactions:

Reaction 9: $2Ce(OH)_3 \rightarrow Ce_2O_3 + 3H_2O$

Reaction 10: $2La(OH)_3 \rightarrow La_2O_3 + 3H_2O$

Reaction 11: $2Nd(OH)_3 \rightarrow Nd_2O_3 + 3H_2O$

Reaction 12: $Th(OH)_4 \rightarrow ThO_2 + 2H_2O$

2.4 Literature Summary

Monazite is a rare-earth phosphate, being the second most common mineral used as a source of rare-earth elements. Monazite exists primarily as an ore containing the rare-earth elements of cerium (Ce), lanthanum (La), and neodymium (Nd). In addition to the rare-earth elements, thorium (Th) can be extracted from monazite ore. Thorium is a slightly radioactive metal, identical to uranium. However, it is more abundant and half as dense as uranium. Because of its abundance, the extraction and further use of thorium from monazite ore has proven a beneficial area of study.

Thorium is widely used as an additive in a variety of applications. Most commonly, thorium is used in metal alloys, catalysis, ceramics, glass, and radiocontrast agents, mostly in the form of thorium dioxide (ThO_2) [5]. More recently, however, thorium has been the subject of studies evaluating its potential as a source of nuclear power. Although thorium is not fissile and thus not able to be directly used in most nuclear reactors, it can be paired with a sufficient amount of uranium, allowing the nucleus of the thorium atom to accept a neutron and beta decay into isotope uranium-233 [7]. Then, nuclear fission of the isotope uranium-233 may occur, generating nuclear power. It is for the reasons mentioned previously that the extraction of thorium dioxide from monazite ore and its further uses have been widely considered as a potential source of green energy across the world. In addition to the extracted thorium, the rare-earth elements of cerium, lanthanum, and neodymium also contained within monazite ore can be extracted by the same process. These rare-earth elements also have a variety of applications, from electronics to nuclear reactions. However, they are found sparsely in the environment, making them costly to acquire. Therefore, the extraction of thorium from monazite ore offers an additional opportunity to recover these highly-valuable elements [9].

A majority of the current processes studied to extract rare-earth elements from monazite ore rely on pyro-hydro or a hybrid technique to recover the elements from the ore. The first step in these processes is to isolate monazite ore from other mined ores, and then determine the

concentration of thorium and rare-earth elements in the monazite ore. Naturally-occurring monazite ore deposits are of varying concentration, leading to recovery processes run using a variety of different methods, temperatures, and time steps. Although thorium-recovery processes vary greatly due to the varying concentration of monazite ore, most processes begin with thermally treating the monazite ore prior to rare earth element separation [4]. Typically, the recovery of rare earth elements involves a leaching step in order to produce a salt that can be used in a solvent extraction step to remove the rare earth salts accordingly. One promising monazite extraction process that has been investigated utilizes a sulphuric acid bake-leach process in a rotary kiln for the extraction of thorium and rare earth elements from monazite ore [3]. The economic viability as well as the environmental and safety impacts of this process will be further studied in this report.

2.5 Basic Process Economics

One of the primary objectives of this project is to determine the economic viability of the thorium-recovery process from monazite ore. To do so, the raw materials costs, equipment costs, manufacturing costs, and the profitability of selling the REE byproducts of the process were considered. The cost per unit information of the raw materials, products, and byproducts are given in Table 1.

Table 1. *Prices of Raw Materials, Products, and Byproducts*

Compound	Price
Monazite	\$1.00/kg
H_2SO_4	\$0.04/kg
NH_4OH	\$0.20/kg
ThO_2	\$80.00/kg
Nd_2O_3	\$60.00/kg
La_2O_3	\$2.00/kg
Ce_2O_3	\$2.00/kg
Natural Gas	\$2.81/GJ

3.0 Method of Approach

The thorium-recovery process was designed to extract thorium (Th) and REE oxides from monazite as follows. First, a feed containing monazite ore and water is fed into a rotary kiln. Through a process of sulfuric acid leaching, a feed containing sulfuric acid and water is fed into the rotary kiln, and the thorium and REEs contained within the monazite ore are converted to thorium and REE sulfates. Then, the products from the rotary kiln are fed into a filter, in which the thorium sulfate cake is separated and partially dried from the liquid REE sulfates and phosphates stream.

The stream containing thorium sulfate cake, along with a stream containing ammonium hydroxide and water, are fed into a neutralizer. The neutralization reaction taking place in the neutralizer converts the thorium sulfate cake to thorium oxide, producing ammonia and various sulfates as byproducts. Then, the products from the neutralizer are fed into a filter, in which the thorium oxide is separated and partially dried from the liquid ammonia and sulfates stream. The liquid ammonia and sulfates stream is exited from the process as waste, while the thorium oxide stream is exited from the process as a product.

From the initial filter, the stream containing REE sulfates and phosphates is fed into a liquid-liquid extractor. The liquid-liquid extractor separates the organic phosphates from the REE sulfates. Then, the organic phosphates stream is fed into a stripper, in which a water stream is used to remove the organic components from the organic phosphates stream. As a result of this stripping process, a stream containing phosphates and water exits the stripper. The REE sulfates stream from the liquid-liquid extractor is fed into a neutralizer. The neutralization reaction taking place in the neutralizer converts the REE sulfate cake to REE hydroxides, producing ammonia and various sulfates as byproducts. Then, the products from the neutralizer are fed into a filter, in which the REE hydroxides are separated and partially dried from the liquid ammonia and sulfates stream. The liquid ammonia and sulfates stream is exited from the process as waste, while the REE hydroxide stream is fed into a rotary kiln. The rotary kiln dries the REE hydroxide stream from aqueous waste, producing REE oxides. The REE oxides stream is exited from the process as a product.

The thorium-recovery process outlined above was designed to extract thorium and REE oxides from monazite ore using OLI Flowsheet software. After placing all streams and equipment into the software, the stream flow rates and equipment properties were optimized to result in the process producing the greatest yield of product possible while maintaining a reasonable product purity. The steps taken to optimize the process are discussed in further detail in Section 4.1 of this report. To analyze the economic viability, capital costs, manufacturing costs, and safety of the thorium-recovery process, a number of assumptions were made. The assumptions made during completion of this study-level design and economic analysis are included in Table 2. Then, the economic and safety hazard methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics* were used. All costs associated with the process were adjusted using a Chemical Engineering Index of 596.2 for January 2020. The results of the design, economic viability, capital costs, manufacturing costs, and safety of the thorium-recovery process are discussed in Section 4.0 of this report.

Table 2. *Project Assumptions Chart*

Project Assumptions		
Equipment	**PFD ID**	**Assumption(s)**
Extractor-1	CSplit-1	Assumed an equal volume of organic liquid to aqueous phosphates would be used to separate the phosphates from the aqueous solution; assumed the extractor could be modeled as a verticle process vessel with six trays; assumed 100% conversion
Stripper-1	CSplit-1	Assumed steady state (stripper volumetric flow rates are equal to extractor volumetric flow rates); assumed 100% conversion
Filter-1	Filter-1	Assumed solid phase entering filter could be classified as either a concentrated ore or a concentrated slime/clay (>5% solids)
	Conveyor-1	Assumed a belt width of 1.0 m and a conveying distance of 10.0 m
Filter-2	Filter-2	Assumed solid phase entering filter could be classified as either a concentrated ore or a concentrated slime/clay (>5% solids)
	Conveyor-2	Assumed a belt width of 1.0 m and a conveying distance of 10.0 m
Filter-3	Filter-3	Assumed solid phase entering filter could be classified as either a concentrated ore or a concentrated slime/clay (>5% solids)
	Conveyor-3	Assumed a belt width of 1.0 m and a conveying distance of 10.0 m
Neutralizer-1	Neutrl-1	Assumed to need a residence time of 10.0 min (single-phase liquid mixture); assumed the neutralizer could be modeled as a CSTR
	Agitator-1	Assumed near-vigorous mixing; assumed the agitator could be modeled as a mechanically-sealed agitator in a closed tank
Neutralizer-2	Neutrl-2	Assumed to need a residence time of 30.0 min (two-phase liquid mixture); assumed the neutralizer could be modeled as a CSTR
	Agitator-2	Assumed near-vigorous mixing; assumed the agitator could be modeled as a mechanically-sealed agitator in a closed tank
Rotary Kiln-1	R-1	Assumed a residence time of 1.0 hour; assumed monazite ore had the same specific heat as cerium phosphate; assumed 100% conversion
Rotary Kiln-2	R-2	Assumed a residence time of 1.0 hour; assumed monazite ore had the same specific heat as cerium phosphate; assumed Rotary Kiln-2 had a temperature change equivalent to Rotary Kiln-1 (~25 degC); assumed the inlet water flow rate was zero because a conveyor from Filter-2 was used at the inlet of the rotary kiln; assumed 100% conversion

4.0 Results

4.1 Optimization

In order to optimize the thorium-recovery process, a number of measures were taken. First, the thorium-recovery process was designed to extract thorium and REE oxides from monazite ore using the least number of pieces of equipment to maintain the highest yield of thorium and REE oxides products possible. To accomplish this, the process was designed using OLI Flowsheet software, as outlined in Section 3.0 of this report. The software allowed for simple comparability of product yield as the process equipment was altered. Second, the thorium-recovery process was designed to extract thorium and REE oxides from monazite ore using the smallest stream flow rates to maintain the highest yield of thorium and REE oxides products possible. Identical to the optimization of the process equipment, the stream flow rates were adjusted using OLI Flowsheet software until the highest yield of thorium and REE oxides was produced. Further recommendations for optimization of the thorium-recovery process that were considered after obtaining the results of this study-level design and economic analysis are included in Section 7.0 of this report.

4.2 Process Flow Diagram

The process design results obtained from OLI Flowsheet are shown in Figure 2. The rotary kiln (R-2) used to convert the REE hydroxides contained in Stream 14 to REE oxides could not be modeled in OLI Flowsheet, and thus it is not included in Figure 2. Instead, Microsoft Excel was used to model the second rotary kiln (R-2).

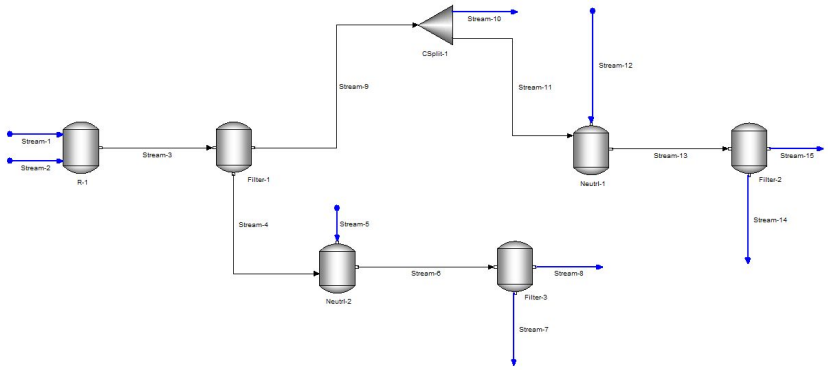

Figure 2. *Process Flow Diagram of the Thorium Recovery Process*

More information on the composition of each stream in the process is included in Tables 3 and 4, based on the mass and molar flow rates of each stream.

Table 3. *Stream Composition Chart with Mass Flow Rates*

		Streams (kg/hr)															
Component	**MW (kg/kmol)**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16 (not on PFD)**
Ce_2O_3	328.23	-	-	-	-	-	-	-	-	-	-	335.20	-	0.03	-	0.03	335.16
$Ce(OH)_3$	191.14	-	-	-	-	-	-	-	-	-	-	-	-	390.36	390.36	-	-
$CePO_4$	235.09	-	480.16	480.16	-	-	-	-	-	480.16	-	-	-	-	-	-	-
H_2O	18.02	1026.87	1000.00	2403.57	-	54.05	66.98	-	66.98	2403.57	95.43	2266.29	-	3373.55	-	3373.55	104.90
H_2SO_4	98.08	2942.39	-	891.57	-	-	-	-	-	891.57	-	-	-	-	-	-	-
La_2O_3	325.81	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	166.39
$La(OH)_3$	189.93	-	-	-	-	-	-	-	-	-	-	-	-	194.00	194.00	-	-
$LaPO_4$	233.88	-	238.89	238.89	-	-	-	-	-	238.89	-	-	-	-	-	-	-
$La(SO_4)_3$	427.09	-	-	-	-	-	-	-	-	-	-	289.07	-	-	-	-	-
Nd_2O_3	336.48	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	137.47
$Nd(OH)_3$	195.26	-	-	-	-	-	-	-	-	-	-	-	-	159.55	159.55	-	-
$NdPO_4$	239.21	-	195.51	195.51	-	-	-	-	-	195.51	-	-	-	-	-	-	-
$Nd_2(SO_4)_3$	576.67	-	-	-	-	-	-	-	-	-	-	235.65	-	0.06	-	0.06	-
NH_3	17.03	-	-	-	-	-	15.83	-	15.83	-	-	-	-	1145.87	-	1145.87	-
NH_4OH	35.05	-	-	-	-	35.05	-	-	-	-	-	-	175.23	-	-	-	-
P_2O_5	141.94	-	14.51	15.81	-	-	-	-	-	15.81	221.88	-	-	-	-	-	-
$(P_2O_5)_2$	283.89	-	-	-	-	-	-	-	-	-	69.39	-	-	-	-	-	-
SO_3	80.06	-	-	1641.37	-	-	32.75	-	32.75	1641.37	-	1420.55	-	1641.35	-	1641.35	-
SiO_2	60.08	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Th(SO_4)_2$	424.16	-	-	86.75	86.75	-	-	-	-	-	-	-	-	-	-	-	-
ThO_2	264.04	-	54.00	-	-	-	54.00	54.00	-	-	-	-	-	-	-	-	-
UO_2	270.03	-	-	3.70	-	-	-	-	-	3.70	-	3.70	-	3.70	3.70	-	-
$U_3(PO_4)_4$	1093.97	-	5.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Flow Rate (kg/hr)		3969.26	1988.07	5957.33	86.75	89.10	169.56	54.00	115.56	5870.58	386.70	4550.46	175.23	6908.47	747.61	6160.86	743.92
Pressure (atm)		1.00	1.00	2.00	2.00	1.00	1.00	1.00	1.00	2.00	2.00	2.00	1.00	4.00	4.00	4.00	-
Temperature (°C)		25.0	25.0	5.0	5.0	25.0	96.3	96.3	96.3	5.0	5.0	5.0	25.0	123.9	123.9	123.9	-

Table 4. *Stream Composition Chart with Molar Flow Rates*

		Streams (kmol/hr)															
Component	**MW (kg/kmol)**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16 (not on PFD)**
Ce_2O_3	328.23	-	-	-	-	-	-	-	-	-	-	1.02	-	0.00	-	0.00	1.02
$Ce(OH)_3$	191.14	-	-	-	-	-	-	-	-	-	-	-	-	2.04	2.04	-	-
$CePO_4$	235.09	-	2.04	2.04	-	-	-	-	-	2.04	-	-	-	-	-	-	-
H_2O	18.02	56.99	55.49	133.38	-	3.00	3.72	-	3.72	133.38	5.30	125.77	-	187.21	-	187.21	5.82
H_2SO_4	98.08	30.00	-	9.09	-	-	-	-	-	9.09	-	-	-	-	-	-	-
La_2O_3	325.81	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.51
$La(OH)_3$	189.93	-	-	-	-	-	-	-	-	-	-	-	-	1.02	1.02	-	-
$LaPO_4$	233.88	-	1.02	1.02	-	-	-	-	-	1.02	-	-	-	-	-	-	-
$La(SO_4)_3$	427.09	-	-	-	-	-	-	-	-	-	-	0.68	-	-	-	-	-
Nd_2O_3	336.48	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0.41
$Nd(OH)_3$	195.26	-	-	-	-	-	-	-	-	-	-	-	-	0.82	0.82	-	-
$NdPO_4$	239.21	-	0.82	0.82	-	-	-	-	-	0.82	-	-	-	-	-	-	-
$Nd_2(SO_4)_3$	576.67	-	-	-	-	-	-	-	-	-	-	0.41	-	0.00	-	0.00	-
NH_3	17.03	-	-	-	-	-	0.93	-	0.93	-	-	-	-	67.29	-	67.29	-
NH_4OH	35.05	-	-	-	-	1.00	-	-	-	-	-	-	5.00	-	-	-	-
P_2O_5	141.94	-	0.10	0.11	-	-	-	-	-	0.11	1.56	-	-	-	-	-	-
$(P_2O_5)_2$	283.89	-	-	-	-	-	-	-	-	-	0.24	-	-	-	-	-	-
SO_3	80.06	-	-	20.50	-	-	0.41	-	0.41	20.50	-	17.74	-	20.50	-	20.50	-
SiO_2	60.08	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
$Th(SO_4)_2$	424.16	-	-	0.20	0.20	-	-	-	-	-	-	-	-	-	-	-	-
ThO_2	264.04	-	0.20	-	-	-	0.20	0.20	-	-	-	-	-	-	-	-	-
UO_2	270.03	-	-	0.01	-	-	-	-	-	0.01	-	0.01	-	0.01	0.01	-	-
$U_3(PO_4)_4$	1093.97	-	0.00	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Total Flow Rate (kmol/hr)		86.98	59.69	167.19	0.20	4.00	5.26	0.20	5.06	166.98	7.10	145.63	5.00	278.89	3.89	275.00	7.76
Pressure (atm)		1.00	1.00	2.00	2.00	1.00	1.00	1.00	1.00	2.00	2.00	2.00	1.00	4.00	4.00	4.00	-
Temperature (˚C)		25.0	25.0	5.0	5.0	25.0	96.3	96.3	96.3	5.0	5.0	5.0	25.0	123.9	123.9	123.9	-

4.3 Safety, Health, and Sustainability Analysis

The safety, health, and environmental impacts of the thorium-recovery process were analyzed by creating a Stream Hazard Chart, as shown in Table 5.

Table 5. *Stream Hazard Chart*

							Toxicity				
Classification	**Compound**	**MW (g/mol)**	**Melting Point (degC)**	**Flash Point (degC)**	**Boiling Point (degC)**	**Liquid Density (g/cm3)**	**Flammability?**	**Deadly Poison?**	**Toxin?**	**Carcinogenic?**	**Mutagenic?**	**Irritating?**	**Corrosive?**
Reactants	Monazite Ore	273.68	3170.0*	-	4400.0*	-	non	no	yes	presumed	yes	yes	no
	H_2SO_4 (sulfuric acid)	98.08	10.0	-	290.0	1.84	non	no	yes	no	no	yes	yes
	NH_4OH (ammonium hydroxide)	35.05	-69.0	-	27.0	0.91	non	no	yes	no	yes	yes	yes
	H_2O (water)	18.02	0.0	-	100.0	1.00	non	no	no	no	no	no	no
Sulfates	$Th(SO_4)_2$ (thorium sulfate)	424.16	200.0	-	-	2.80	non	no	yes	-	-	yes	yes
	$La_2(SO_4)_3$ (lanthanum sulfate)	566.00	1150.0	-	-	2.80	non	no	yes	-	-	yes	yes
	$Nd_2(SO_4)_3$ (neodymium sulfate)	576.67	1176.0	-	-	2.85	non	no	yes	-	-	yes	yes
Phosphates	$CePO_4$ (cerium phosphate)	235.09	-	-	-	5.22	non	no	yes	-	-	yes	no
	$LaPO_4$ (lanthanum phosphate)	233.88	-	-	-	-	non	no	yes	-	-	yes	no
	$NdPO_4$ (neodymium phosphate)	239.21	-	-	-	-	non	no	yes	-	-	yes	no
Hydroxides	$Ce(OH)_3$ (cerium hydroxide)	191.14	~2000.0	-	~4000.0	6.77	non	no	yes	-	-	yes	yes
	$La(OH)_3$ (lanthanum hydroxide)	189.93	~2000.0	-	~4000.0	4.28	non	no	yes	-	-	yes	yes
	$Nd(OH)_3$ (neodymium hydroxide)	195.26	~2000.0	-	~4000.0	-	non	no	yes	-	-	yes	yes
Oxides	ThO_2 (thorium oxide)	264.04	3390.0	-	4400.0	10.00	non	no	yes	presumed	yes	yes	no
	Ce_2O_3 (cerium oxide)	328.23	3730.0	-	-	6.20	non	no	yes	-	-	yes	no
	La_2O_3 (lanthanum oxide)	325.81	2315.0	-	4200.0	6.51	non	no	yes	-	-	yes	no
	Nd_2O_3 (neodymium oxide)	336.48	2270.0	-	-	7.24	non	no	yes	-	-	yes	no
Byproducts	UO_2 (uranium dioxide)	270.03	2868.0	-	-	10.95	yes	no	yes	yes	yes	yes	no
	P_2O_5 (phosphorous pentoxide)	141.94	340.0	-	360.0	2.39	non	no	yes	-	-	yes	yes
	SO_2 (sulfur trioxide)	80.06	16.8	-	44.8	1.92	non	presumed	yes	yes	-	yes	yes
	NH_3 (ammonia)	17.03	-72.0	-	36.0	0.90	non	no	yes	-	-	yes	yes

*properties of thorium phosphate; each chemical component of monazite ore will have different properties
"-" indicates that the property does not apply, or the property is still being investigated

From the prepared Stream Hazard Chart, the primary hazards associated with the thorium-recovery process are the toxicity, carcinogenicity, mutagenicity, corrosivity, and irritating properties of most of the chemical components used in the process. The risk of these hazards to the safety, health, and sustainability of the process employees, community, and environment may be minimized by employing a variety of methods in the process.

One method of reducing the safety and health risks of the hazardous properties associated with the chemical components used in the process to the process employees is to ensure that all employees are informed of the toxicity, carcinogenicity, mutagenicity, corrosivity, and irritating properties of the process chemicals. To accomplish this objective, required chemical safety training should be provided to process employees on a consistent basis in order to promote long-term retention of the stream hazards. Another method of reducing the safety and health risks of the hazardous chemical properties to the process employees is to provide them with the appropriate personal protective equipment. All process employees should wear closed-toe shoes, long sleeves, eye protection, and gloves when working near the process equipment. A final method of reducing the safety and health risks of the hazardous chemical properties to the process employees is to equip the thorium-recovery process plant with the appropriate amount of ventilation, eye wash stations, chemical showers, and first-aid kits to be used in the case of an emergency.

Next, the health and safety of the community surrounding the thorium-recovery process was considered in the design. One method of reducing the safety and health risks of the hazardous chemical properties to the surrounding community is to prevent catastrophic incidents caused by the malfunction of the process equipment before such incidents occur. This may be accomplished by implementing hazard and operability studies (HAZOP), Failure Mode Effect Analysis, or a similar methodology to ensure the thorium-recovery process and its equipment are running as expected. These methods aid in limiting or preventing the exposure of the community to the hazards of the chemical components of the process by drawing attention to equipment malfunctions before a catastrophic incident occurs.

Finally, the sustainability of the thorium-recovery process was considered in the design. One method of improving the sustainability of the process is to ensure that any waste generated from the process is handled properly and according to federal regulations. The primary waste from the thorium-recovery process contains large amounts of ammonia and various sulfates. Because these chemical components are toxic and have irritating properties, the process should be disposed of accordingly. Emissions may emerge from the equipment of the thorium-recovery process, such as the rotary kiln. Therefore, another method of improving the sustainability of the thorium-recovery process is to ensure that any emissions generated from the process are reduced. Following this report, the process design may be further modified with additional recycle streams in order to reduce environmentally-harmful emissions. Additionally, the process design may be later considered with the utilization of more sustainable sources of energy in order to further reduce emissions.

4.4 Capital Cost Estimates

To analyze the capital costs of the thorium-recovery process, the economic methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics* were used for each piece of equipment in the process. A summary of the capital costs associated with our thorium-recovery process is shown in Table 6. For the neutralizers of the process, a reaction occurs in that step of the process. Because a neutralization reaction occurs, the neutralizers were modeled as Continuous Stirred Tank Reactors (CSTRs), using non-jacketed vaporizer vessels to represent the CSTRs. The sizing for each neutralizer was found to be based on the product of the volumetric flow rates and an assumed residence time. Because the first neutralizer (Neutrl-1) contained a single-phase liquid mixture, it was assumed to need an average residence time of 10 minutes. Because the second neutralizer (Neutrl-2) contained a two-phase liquid mixture, it was assumed to need an average residence time of 30 minutes. The material chosen for both neutralizers was carbon steel with no pressure requirements, as OLI Flowsheet reported pH measurements of 7.0 and pressure measurements of less than 5 atm. According to Ulrich, et al., for the agitators of the neutralizers, both neutralizers called for axial turbine impellers. The power requirement of each agitator was different due to the differences in mixture phases present in each neutralizer, assuming near-vigorous mixing. The capital cost of the axial turbine

impellers was estimated as if they were mechanically-sealed agitators in a closed tank. The size and total capital cost of the first neutralizer, including its agitator, was estimated as shown in Appendix A, and were found to be 146 m^3 and $214,632.00, respectively. The size and total capital cost of the second neutralizer, including its agitator, was estimated as shown in Appendix A, and were found to be 3.54 m^3 and $71,171.38, respectively.

For the component splitter (Csplit-1) found in the process flow diagram from OLI Flowsheet, a liquid-liquid extractor and a stripper are used in the real thorium-recovery process to separate the phosphates from the inlet stream. The stripper and the extractor run in tandem to have a continuous phosphate and organic separation, and the organic stream is recycled back into the extractor. The extractor is modeled as a vertical process vessel with six stainless steel trays. The size and total capital cost of the extractor was estimated as shown in Appendix A. The extractor was estimated to have a volume of 0.76 m^3, a diameter of 1 meter, a height of 10 meters, and a capital cost of $204,715.50. Due to the process running at steady state, and the recycling of the organics, the flow rates for the stripper will be the same as the extractor. Thus, the stripper will be sized and priced the same as the extractor. The total capital cost of the stripper was estimated as shown in Appendix A, and was found to be $204,715.50.

For each filter included in the process, the solid phase entering the filter was assumed to be either a concentrated ore or a concentrated slime or clay (>5% solids). Using this classification, the filter area of each filter was found to be proportional to the solids mass feed rate into the filter. The size and total capital cost of each filter were estimated as shown in Appendix A. For the first filter (Filter-1), the filter area was estimated to be 1.20 m^2, while the capital cost was estimated to be $166,339.80. For the second filter (Filter-2), the filter area was estimated to be 10.38 m^2, while the capital cost was estimated to be $547,311.60. For the third filter (Filter-3), the filter area was estimated to be 0.75 m^2, while the capital cost was estimated to be $139,510.80. Because the filters were used to separate a solid-liquid mixture, the solids stream exiting each filter was made mobile and dried using a conveyor. The conveyors were modeled as horizontal belt conveyors with a belt width of 1.0 meters and a minimum conveying distance of 10 meters. The capital cost of each conveyor was estimated as shown in Appendix A.

For all three conveyors (Conveyor-1 through Conveyor-3), the capital cost was estimated to be $53,658.00.

For the first rotary kiln (R-1), a residence time of one hour was assumed, and the rotary kiln was designed using brick lining. The size and capital cost of the first rotary kiln were estimated as shown in Appendix A, and were found to be 3.79 m^3 and $500,000.00, respectively. Similarly, for the second rotary kiln (R-2), a residence time of one hour was assumed, and the rotary kiln was designed using brick lining. The size and capital cost of the second rotary kiln were estimated as shown in Appendix A, and were found to be 0.16 m^3 and $60,000.00, respectively.

Table 6. *Capital Cost Summary*

Capital Cost Summary									
Equipment	**ID**	**Capacity & Size Specifications**	**Base Cost (Cp) (2004)**	**Base Cost (Cp) (2020)**	**Material Factor (FM)**	**Pressure & Other Factor (Fp)**	**Bare Module Factor (FBM)**	**Base Bare Module Cost (CBM)**	**Total**
Neutralizer-1	Neutrl-1	146 m^3, plain vaporizer, CS, 4 atm	$30,000.00	$44,715.00	1	1	3	$134,145.00	$214,632.00
	Agitator-1	16.0 kW, turbine axial, mechanical seal	$27,000.00	$40,243.50	1	1	2	$80,487.00	
Neutralizer-2	Neutrl-2	3.54 m^3, plain vaporizer, CS, 2 atm	$5,250.00	$7,825.13	1	1	3	$23,475.38	$71,171.38
	Agitator-2	5.66 kW, turbine axial, mechanical seal	$16,000.00	$23,848.00	1	1	2	$47,696.00	
Filter-1	Filter-1	1.20 m^2, horizontal belt filter, stainless steel	$31,000.00	$46,205.50	1	1	3.6	$166,339.80	$219,997.80
	Conveyor-1	1.0 m belt width, 10 m conveying distance, belt conveyor	$15,000.00	$22,357.50	1	1	2.4	$53,658.00	
Filter-2	Filter-2	10.4 m^2, horizontal belt filter, stainless steel	$102,000.00	$152,031.00	1	1	3.6	$547,311.60	$600,969.60
	Conveyor-2	1.0 m belt width, 10 m conveying distance, belt conveyor	$15,000.00	$22,357.50	1	1	2.4	$53,658.00	
Filter-3	Filter-3	0.75 m^2, horizontal belt filter, stainless steel	$26,000.00	$38,753.00	1	1	3.6	$139,510.80	$193,168.80
	Conveyor-3	1.0 m belt width, 10 m conveying distance, belt conveyor	$15,000.00	$22,357.50	1	1	2.4	$53,658.00	
Extractor	Csplit-1	10m length, 1m diameter, CS vertical process vessel	$30,000.00	$44,715.00	1	1	4.3	$192,274.50	$210,817.81
	Trays-1	6 stainless steel trays	$650.00	$968.83	6	1.45	2.2	$18,543.31	
Stripper	Csplit-1	10m length, 1m diameter, CS vertical process vessel	$30,000.00	$44,715.00	1	1	4.3	$192,274.50	$210,817.81
	Trays-1	6 stainless steel trays	$650.00	$968.83	6	1.45	2.2	$18,543.31	
Rotary Kiln-1	R-1	3.7897 m^3, brick lining	$100,000.00	$149,050.00	1	1	5	$745,250.00	$745,250.00
Rotary Kiln-2	R-2	0.156982m^3, brick lining	$12,000.00	$17,886.00	1	1	5	$89,430.00	$89,430.00
Total Bare Module Cost									$2,556,255
Contingency and Fees									$460,126
Total Module Cost									$3,016,381
Auxiliary Fees									$904,914
Grass Roots Capital									$3,921,295

4.5 Manufacturing Cost Estimates

To analyze the manufacturing cost of the thorium-recovery process, the economic methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics* were used. The total manufacturing cost for the thorium-recovery process was estimated to be $25,819,873.00, as shown in Appendix B. This estimate was determined by calculating the manufacturing expenses, depreciation, and the general expenses that come from operating this plant year-round. The fixed capital of all the equipment was found to be $3,921,295.00, as shown in Appendix B. The largest financial drawback found in the manufacturing cost comes from the raw materials. The cost of monazite ore is expensive, and using the ore at a feed rate of 1000 kg/hr has caused the price for raw materials to be relatively high.

The money accumulated from sales of the thorium oxide and REE oxides products was found to be $202,454,700.00, as shown in Appendix B. The sales income minus the manufacturing cost gave a net annual profit of $176,634,828.03, as shown in Appendix B. After taxes, the net annual profit from this process for one year was estimated to be $116,578,986.50, as shown in Appendix B. The net annual profit gave an after tax rate of return of 3878%. A summary of all of the manufacturing costs and profit can be found in Table 7.

Table 7. *Manufacturing Cost Summary*

MANUFACTURING COST SUMMARY							
Capital							
Fixed capital, C_{FC}	$3,921,295.47						
Working capital (10-20% of fixed capital), C_{WC}	$588,194.32						
Total capital investment, C_{TC}	$4,509,489.79						
Manufacturing Expenses							
Direct							**$/yr**
Raw materials							$19,756,660.54
By-product credits							$0.00
Catalysts and solvents							$0.00
Operating labor							$552,240.00
Supervisory and clerical labor (10-20% of operating labor)							$82,836.00
Utilities							
Natural Gas	4.9506	GJ/hr	$2.81	$/GJ	$13.91	$/hr	$121,861.99
Electricity	415000	kWh/yr	$0.13	$/kWh	$54,780.00	$/yr	$54,780.00
Process water	3.316	m^3/hr	$1.10	$/$m^3$	$3.65	$/hr	$31,952.98
Waste disposal	6317	kg/hr	$0.00	$/kg	$15.79	$/hr	$138,342.30
Maintenance and repairs (2-10% of fixed capital)							$196,064.77
Operating supplies (10-20% of maint & repairs)							$29,409.72
Laboratory charges (10-20% of operating labor)							$82,836.00
Patents and royalties (0-6% of total expense)							$420,939.69
Total, A_{DME}							$21,467,923.98
Indirect							
Overhead (payroll and plant), packaging, storage (50-70% of op. Labor + supervision + maint.)							$415,570.39
Local taxes (1-3% of fixed capital)							$78,425.91
Insurance (1-2% of fixed capital)							$58,819.43
Total, A_{IME}							$552,815.73
Total manufacturing expense, $A_{ME}=A_{DME}+A_{IME}$							$22,020,739.71
General Expenses							
Administrative costs (25% of overhead)							$103,892.60
Distribution and selling (10% of total expense)							$2,202,073.97
Research and development (5% of total expense)							$1,101,036.99
Total general expense, A_{GE}							$3,407,003.55
Depreciation (approximately 10% of fixed capital), A_{BD}							$392,129.55
Total Expenses, A_{TE}							$25,819,872.81
Revenue from Sales, A_s			11,683,212	kg/yr	Avg ~ $17.33	$/kg	$202,454,700.83
Net annual profit, A_{NP}							$176,634,828.03
Income taxes (net annual profit times the tax rate), A_{IT}							$60,055,841.53
Net annual profit after taxes (A_{NP} -A_{IT}), A_{NNP}							$116,578,986.50
	Aftertax rate of return, i = (1.5 A_{NNP}/C_{TC}) x 100 =						3878%

Table 8. *Raw Material Costs and Outputs*

Coumpound	Price/kg	Input kg/hr	Input mol/hr	Out mol/hr	Out kg/hr	Cost/hr	Value (out/hr)	Cost/yr	Value (out/yr)
Monazite	$1.00	1000				$1,000.00		$8,760,000.00	
H2SO4	$0.04	2942.37	30000			$117.69		$1,031,006.45	
NH4OH	$0.20	5683.133043	71001			$1,136.63		$9,956,849.09	
Methyl isobutyl ketone	$1.50	5870				n/a		$8,805.00	
ThO2	$80.00			204.515	54.0001406		$4,320.01		$37,843,298.53
ND2O3	$60.00			817.088	274.9370386		$16,496.22		$144,506,907.48
La2O3	$2.00			1021.43	332.796194		$665.59		$5,830,589.32
Ce2O3	$2.00			2042.22	670.3382928		$1,340.68		$11,744,326.89
UO2	$78.00			13.71	3.7021113		$288.76		$2,529,578.61
							Total	$19,756,660.54	$202,454,700.83
								Overall Income/yr	$182,698,040.30

5.0 Discussion of Results

The overall process of separating thorium and REE oxides from monazite ore produced a relatively pure stream of the products that can be sold for profit. It was found that separating the thorium from the other rare earth metals and the phosphates before neutralizing the compounds was a viable way to produce the desired products. The first rotary kiln (R-1) served to mix the rare earth phosphates from the raw monazite ore with an acidic solution to produce rare earth sulfates. After separating thorium from the product of the first rotary kiln, the phosphates were removed from the rare earth metals. This extraction of phosphate was performed with an extractor and stripper in tandem using methyl isobutyl ketone. Next, neutralizing the thorium stream, as well as the rare earth element stream, was completed to produce REE hydroxides and thorium oxide. Using a second rotary kiln (R-2), the REE hydroxides stream was converted into the REE oxide products. This process was designed and tested in OLI Flowsheet software, and the data from OLI was used to calculate the economic viability of the thorium-recovery process. The total revenue from products was estimated to be $202,454,700.00, with a raw materials cost of $19,756,660.00, excluding manufacturing costs. The large income associated with the process is due to the purity of the final products as well as the high flow rate at which this process will operate.

One unexpected outcome of the OLI Flowsheet process calculations was that the thorium stream did not need an additional rotary kiln. Based on the literature review performed before the process was modeled, both the thorium and the rare earth elements lines were expected to have their own rotary kiln after sharing the first rotary kiln. Instead, the thorium stream leaving the

first neutralizer (Neutrl-1) was directly converted to thorium oxide. As expected, the rare earth element stream required a rotary kiln to complete the reaction converting REE hydroxides into REE oxides. The discrepancy between the two process lines was attributed to excess heat generated in the first neutralizer (Neutrl-1) due to less material being present that dissipated the heat. It is believed that the thorium stream did not require a rotary kiln because of its purity and the large amount of acid used to neutralize it.

The thorium-recovery process was found to have a favorable economic potential, despite the high cost of monazite ore. Operating the process at a high inlet ore flow rate of 1000 kg/hr caused the price of raw materials to be relatively high, leading to the largest drawback in the manufacturing cost. The overall cost of this process was $25,819,872.00, and the raw materials needed to operate the process at a flow rate of 1000 kg/hr account for 75% of the total manufacturing expense. After taxes, the net annual profit of the process in one year was estimated to be above $116,000,000.00, yielding a return on investment of 3878%. The annual expenses for this process were estimated to be less than $20,000,000.00.

6.0 Conclusions

Based on the economic analysis of the thorium-recovery process, an estimated $182 million per year could be earned. Most of the profit was found to come from the products neodymium and thorium oxide, so it is important that these streams remain as efficient as possible. It is also important to consider that further separations would significantly increase manufacturing costs. Further equipment ventures and other cost considerations that could alter the profitability of the process are discussed in Section 7.0 of this report. It is worth noting that because lower percentages were used in estimating the manufacturing cost of the process, the overall profitability would realistically be not as high as reported. Regardless, this study level design and economic analysis used the data collected from OLI Flowsheet. Thus, based on the economic evaluations for running this process to produce thorium and REE oxides, this design was found to be viable.

7.0 Recommendations

Due to the time constraints of this process design and economic analysis, many aspects of the process were not able to be scrutinized as well as intended. This section of the report serves to explain some shortcomings of the assumptions made during this project, as well as some topics that should be revisited to enhance the reliability of the economic analysis presented in this report. First, for the rotary kilns and neutralizers involved in the process, a standard residence time was assumed based on the reactions occurring and the flow rates involved. This affected the sizing and costing of the equipment. Given more time and resources, a full kinetic analysis of the reactions and their respective rates would allow for a more accurate assumption of residence time and, therefore, sizing and costing of the equipment.

Next, no consideration was taken for any of the potentially valuable by-products of the process, which were simply disposed of instead. This leaves room for further analysis and potential improvement of the process through the addition of auxiliary equipment to separate out valuable by-products in order to either sell or recycle them back into the process. Separating out valuable by-products would lead to higher sales revenue while potentially reducing the cost of utilities and raw materials. More specifically, the phosphates leaving the stripper (CSplit-1) have the potential to be of high value, and could possibly be separated from the water and sold as a product. The water, on the other hand, could potentially be recycled back into the system to further reduce water costs. The other waste streams of the thorium-recovery process contain sulfates and water. At this time, it is not believed that the sulfates hold enough potential value to be worth the capital and manufacturing costs of separation, but the water could be further analyzed for potential process value increases through the addition of recycle streams.

Next, it should be considered that to receive the full value of the REE oxide products in sales revenue, an additional separation subprocess should be considered after the second rotary kiln (R-2) to further purify the products. If this new equipment is considered, the capital cost of this process would increase significantly, reducing the overall value of it. However, unseparated, the REE oxides may not be worth the values they were assumed to be worth in this report. Thus, using lower REE oxide values or considering the additional separation subprocess would produce a more realistic return on investment.

Next, it should be considered that there is the potential to increase the value of the thorium-recovery process by either reducing the excess amount of raw material inputs, namely water, sulfuric acid, and ammonia hydroxide, or changing the chemicals used altogether. First, other acidic and caustic solutions should be investigated for better product yield and purity. Secondly, the inlet flow rates used in this project should be reconsidered, as these materials were mostly used in excess to ensure proper reactivity and near full conversion. Because of this, there is the potential to reduce the capital and manufacturing costs of the process by reducing the amount of excess materials used down to only the amount required.

Finally, the electrical requirements for the thorium-recovery process was an estimated annual value for the pumps, motor, and agitators. The value was taken from a similar chemical process' requirements but was not derived explicitly for this project's process. Reevaluating the electrical requirement for this process should be reconsidered, as it may yield either a higher or lower utility cost and thus change the overall value of the process. Additionally, further research into more sustainable energy sources should be conducted in order to improve the sustainability of the process.

8.0 References

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Appendices

Appendix A: Capital Cost

To analyze the capital costs of the thorium-recovery process, the economic methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics* were used for each piece of equipment in the process.

<u>Neutrl-1</u>

Because a reaction occurs in the neutralizer, Neutrl-1 is modeled as a Continuous Stirred Tank Reactor (CSTR), using a non-jacketed vaporizer vessel to represent the CSTR. From OLI Flowsheet, the volumetric flow rate into Neutrl-1 is,

$$F = 876,085\ L/hr = 867.09\ m^3/hr = 14.6\ m^3/min$$

Because Neutrl-1 contained a single-phase liquid mixture, it was assumed to need a residence time (τ) of 10 minutes. From Ulrich, et al., the required volume of Neutrl-1 is related to the product of the volumetric flow rate and residence time. The required volume of Neutrl-1 is,

$$V = F\tau = (14.6\ m^3/min)(10\ min) = 146\ m^3$$

Using Figure 5.23, $C_p = \$30,000.00$

Because the neutralizer is carbon steel, $F_{BM} = 3.0$

Because the neutralizer is operating at <5 barg, $F_P = 1.0$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = (3.0)(1.0)(\frac{596.2}{400})(\$30,000) = \$134,145.00$

<u>Agitator-1</u>

From Table 4.16, Agitator-1 is modeled as an axial turbine agitator with a mechanical seal. Using this classification, the power consumption is found to be proportional to the volume of the neutralizer through the following relationship:

$$P = 0.3V^{0.8} = 0.3(146\ m^3)^{0.8} = 16\ kW$$

Using Figure 5.42, $C_p = \$27,000.00$

Because the agitator is carbon steel, $F_{BM} = 2.0$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = (2.0)(\frac{596.2}{400})(\$27,000) = \$80,487.00$

Neutrl-2

Because a reaction occurs in the neutralizer, Neutrl-2 is modeled as a Continuous Stirred Tank Reactor (CSTR), using a non-jacketed vaporizer vessel to represent the CSTR. From OLI Flowsheet, the volumetric flow rate into Neutrl-2 is,

$F = 7,081\ L/hr = 7,081\ m^3/hr = 0.118\ m^3/min$

Because Neutrl-2 contained a two-phase liquid mixture, it was assumed to need a residence time (τ) of 30 minutes. From Ulrich, et al., the required volume of Neutrl-2 is related to the product of the volumetric flow rate and residence time. The required volume of Neutrl-2 is,

$V = F\tau = (0.118\ m^3/min)(30\ min) = 3.54\ m^3$

Using Figure 5.23, $C_p = \$5,250.00$

Because the neutralizer is carbon steel, $F_{BM} = 3.0$

Because the neutralizer is operating at <5 barg, $F_P = 1.0$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = (3.0)(1.0)(\frac{596.2}{400})(\$5,250) = \$23,475.38$

Agitator-2

From Table 4.16, Agitator-2 is modeled as an axial turbine agitator with a mechanical seal. Using this classification, the power consumption is found to be proportional to the volume of the neutralizer through the following relationship:

$P = 0.2V(0.8) = 0.2(3.54\ m^3)(0.8) = 5.66\ kW$

Using Figure 5.42, $C_p = \$16,000.00$

Because the agitator is carbon steel, $F_{BM} = 2.0$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = (2.0)(\frac{596.2}{400})(\$16,000) = \$47,696.00$

Extractor

For the component splitter (Csplit-1) in OLI Flowsheet, a liquid-liquid extractor and a stripper are used in the real thorium-recovery process to separate the phosphates from the inlet stream. The extractor is modeled as a vertical process vessel with six stainless steel trays. From the OLI Flowsheet data, the inlet volumetric flow rate into the extractor is $F = 3,769\ L/h$ of aqueous

materials and phosphates. To separate the phosphates, it is assumed that an equal amount of organics is needed, so the total volumetric flow rate into the extractor is $F = 7,600\ L/h$.

Because the extractor is modeled as a vertical process vessel with six stainless steel trays, the resonance time will be 360 seconds. From Ulrich, et al., the required volume of the extractor is related to the product of the volumetric flow rate and residence time. The required volume of the extractor is,

$$V = F\tau = (7,600\ L/hr)(360\ s)(\tfrac{1\ hr}{3600\ s}) = 0.760\ m^3$$

The diameter of the vertical process vessel is assumed to be 1 meter. Because the extractor will need to have six trays, the height of each tray should be 0.25 meters. From Ulrich, et al., the total height of the extractor, allowing space for the reboiler and condenser, will be 10 meters.

For the vertical process vessel,

Using Ulrich, et al., $C_p = \$30,000.00$

Because the vessel is carbon steel, $F_M = 1.0$, $F_P = 1.6$, and $F_{BM} = 4.3$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = (4.3)(\frac{596.2}{400})(\$30,000) = \$192,274.50$

For the six stainless steel trays,

Using Ulrich, et al., $C_p = \$650.00$

Because the trays are stainless steel, $F_{BM} = 2.2$, $N_{act} = 6$, and $F_q = 1.45$

Using a ChemE Index of 596.2 (Jan. 2020), $C_{BM} = \$12,441.00$

The total cost of the extractor is $\$192,274.50 + \$12,441.00 = \$204,715.50$

<u>Stripper</u>

For the component splitter (Csplit-1) in OLI Flowsheet, a liquid-liquid extractor and a stripper is used in the real thorium-recovery process to separate the phosphates from the inlet stream. The strippeer and the extractor run in tandem to have a continuous phosphate and organic separation, and the organic stream is recycled back into the extractor. Due to the process running at steady state, and the recycling of the organics, the flow rates for the stripper will be the same as the extractor. Thus, the stripper will be sized and priced the same as the extractor. Thus, the total cost of the extractor is $\$192,274.50 + \$12,441.00 = \$204,715.50$.

Filter-1

From Table 4.23b, I assumed that the solid phase entering the filter could be classified as either a concentrated ore or a concentrated slime or clay (>5% solids). Using this classification, the dry solids feed rate (m_s) is found to be proportional to the filter area through the following relationship: $0.02A \; to \; 0.1\,A$.

Dry solids feed rate (m_s), $m_s = (86,748.3\ g/hr)(\frac{1\ kg}{1000\ g})(\frac{1h}{3600\ s}) = 0.02410\ kg/s$
Using the lower end of the relationship ($0.02A$), $A = \frac{m_s}{0.02} = \frac{0.02410\ kg/s}{0.02} = 1.2048\ m^2$
From Figure 5.57b, $C_p = \$31,000.00$
Because the filter is continuous and stainless steel, $F_{BM} = 3.6$
Using a ChemE Index of 596.2 (January 2020), $C_{BM} = (3.6)(\frac{596.2}{400})(\$31,000.00) = \$166,339.80$

Conveyor-1

From Table 4.4, the belt width of a belt conveyor can range from 0.3-2.2 m, and the conveying distance of a belt conveyor can range from 10-8,000 m. Because the belt width and conveying distance depends on the specific chemical plant, I assumed a reasonable belt width of 1.0 m, and a minimum conveying distance of 10 m for the purpose of these calculations.

From Figure 5.14, $C_p = \$15,000.00$
Because the conveyor is a belt conveyor, $F_{BM} = 2.4$
Using a ChemE Index of 596.2 (January 2020), $C_{BM} = (2.4)(\frac{596.2}{400})(\$15,000.00) = \$53,658.00$

Filter-2

From Table 4.23b, I assumed that the solid phase entering the filter could be classified as either a concentrated ore or a concentrated slime or clay (>5% solids). Using this classification, the dry solids feed rate (m_s) is found to be proportional to the filter area through the following relationship: $0.02A \; to \; 0.1\,A$.

Dry solids feed rate (m_s), $m_s = (747,600.4833\ g/hr)(\frac{1\ kg}{1000\ g})(\frac{1h}{3600\ s}) = 0.2077\ kg/s$
Using the lower end of the relationship ($0.02A$), $A = \frac{m_s}{0.02} = \frac{0.2077\ kg/s}{0.02} = 10.3833\ m^2$
From Figure 5.57b, $C_p = \$102,000.00$
Because the filter is continuous and stainless steel, $F_{BM} = 3.6$
Using a ChemE Index of 596.2 (January 2020),
$C_{BM} = (3.6)(\frac{596.2}{400})(\$102,000.00) = \$547,311.60$

Conveyor-2

From Table 4.4, the belt width of a belt conveyor can range from 0.3-2.2 m, and the conveying distance of a belt conveyor can range from 10-8,000 m. Because the belt width and conveying distance depends on the specific chemical plant, I assumed a reasonable belt width of 1.0 m, and a minimum conveying distance of 10 m for the purpose of these calculations.

From Figure 5.14, $C_p = \$15,000.00$
Because the conveyor is a belt conveyor, $F_{BM} = 2.4$

Using a ChemE Index of 596.2 (January 2020), $C_{BM} = (2.4)(\frac{596.2}{400})(\$15,000.00) = \$53,658.00$

Filter-3

From Table 4.23b, I assumed that the solid phase entering the filter could be classified as either a concentrated ore or a concentrated slime or clay (>5% solids). Using this classification, the dry solids feed rate (m_s) is found to be proportional to the filter area through the following relationship: $0.02A \; to \; 0.1\,A$.

Dry solids feed rate (m_s), $m_s = (53,999.6\ g/hr)(\frac{1\ kg}{1000\ g})(\frac{1h}{3600\ s}) = 0.0150\ kg/s$
Using the lower end of the relationship ($0.02A$), $A = \frac{m_s}{0.02} = \frac{0.0150\ kg/s}{0.02} = 0.7500\ m^2$
From Figure 5.57b, $C_p = \$26,000.00$
Because the filter is continuous and stainless steel, $F_{BM} = 3.6$

Using a ChemE Index of 596.2 (January 2020), $C_{BM} = (3.6)(\frac{596.2}{400})(\$26,000.00) = \$139,510.80$

Conveyor-3

From Table 4.4, the belt width of a belt conveyor can range from 0.3-2.2 m, and the conveying distance of a belt conveyor can range from 10-8,000 m. Because the belt width and conveying distance depends on the specific chemical plant, I assumed a reasonable belt width of 1.0 m, and a minimum conveying distance of 10 m for the purpose of these calculations.

From Figure 5.14, $C_p = \$15,000.00$
Because the conveyor is a belt conveyor, $F_{BM} = 2.4$

Using a ChemE Index of 596.2 (January 2020), $C_{BM} = (2.4)(\frac{596.2}{400})(\$15,000.00) = \$53,658.00$

R-1

For Rotary Kiln-1, a residence time of one hour was assumed. From Ulrich, et al., the required volume of Rotary Kiln-1 is related to the product of the volumetric flow rate and residence time. The required volume of Rotary Kiln-1 is,

$V = F\tau = (3.7897\ m^3/hr)(1\ hour) = 3.79\ m^3$

From Figure 5.33, $C_p = \$100,000.00$
Because the rotary kiln has brick lining, $F_{BM} = 5.0$

Using a ChemE Index of 596.2 (January 2020), $C_{BM} = \$500,000.00$

R-2

For Rotary Kiln-2, a residence time of one hour was assumed. From Ulrich, et al., the required volume of Rotary Kiln-2 is related to the product of the volumetric flow rate and residence time. The required volume of Rotary Kiln-2 is,

$V = F\tau = (0.1570\ m^3/hr)(1\ hour) = 0.1570\ m^3$

From Figure 5.33, $C_p = \$12,000.00$
Because the rotary kiln has brick lining, $F_{BM} = 5.0$

Using a ChemE Index of 596.2 (January 2020), $C_{BM} = \$60,000.00$

Appendix B: Manufacturing Cost

To analyze the manufacturing costs of the thorium-recovery process, the economic methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics* were used.

Capital

Fixed capital (C_{FC}) was found from the sum of the capital cost of each piece of equipment, as calculated in Appendix A. The fixed capital was estimated to be $3,921,295.47. From Ulrich, et al., the working capital (C_{WC}) was assumed to be 10-20% of the fixed capital. Assuming the working capital of the thorium-recovery process was 15% of the fixed capital,

$$C_{WC} = 0.15C_{FC} = 0.15(\$3,921,295.47) = \$588,194.32$$

Then, the total capital (C_{TC}) was estimated to be the sum of the fixed capital and the working capital. Thus, the total capital was found to be $4,509,489.79.

Direct Manufacturing Expenses

The raw materials expense was estimated as shown in Table 8, giving an expense of $19,756,660.54. From Ulrich, et al., the supervisory and clerical labor expenses were assumed to be 10-20% of the operating labor expense. Assuming the supervisory and clerical labor expenses were 15% of the operating labor costs,

$$Supervisory\ \&\ Clerical\ Labor = 0.15(Operating\ Labor) = 0.15(\$552,240.00) = \$82,836.00$$

The natural gas utility expense was estimated assuming the heat supplied by the natural gas was mostly used to heat the solids and vaporize any water entering the rotary kilns. Monazite ore was assumed to have a specific heat capacity equal to that of cerium phosphate [10]. Based on this, the natural gas energy rate for Rotary Kilns 1 and 2, respectively, were estimated to be,

$$Energy\ Rate = (solid\ flow)(Cp)(Tout - Tin) + (water\ flow)(\Delta Hvap)$$

For R-1, $(4,192.48\ mol/hr)(108\ J/Kmol)(278 - 298) + (112,508.20\ mol/hr)(43,990\ J/mol)$

$$= \$13.88/hr$$

For R-2, $(3,890.00 mol/hr)(108\ J/Kmol)(421.89 - 396.89 = \$0.03/hr$

Assuming the maintenance and repairs expenses were 5% of the fixed capital,

$$Maintenance\ \&\ Repairs = 0.05(C_{FC}) = 0.05(\$3,921,295.47) = \$196,064.77$$

Assuming the operating supplies expenses were 15% of the maintenance and repairs expenses,

$$Operating\ Supplies = 0.15(Maintenance\ \&\ Repairs) = 0.15(\$196,064.77) = \$29,409.72$$

Assuming the laboratory charges were 15% of the operating labor expense,

$$Laboratory\ Charges = 0.15(Operating\ Labor) = 0.15(\$552,240.00) = \$82,836.00$$

Assuming the patents and royalties expenses were 2% of the sum of all direct manufacturing expenses calculated thus far,

$$Patents\ \&\ Royalties = 0.02(Direct\ Manufacturing) = 0.02(\$20,825,543.96) = \$416,510.88$$

<u>Indirect Manufacturing Expenses</u>

Assuming the overhead, packaging, and storage expenses were 50% of the sum of the operating labor, supervisory and clerical labor, and maintenance and repairs expenses,

$$Overhead = 0.50(Operating,\ Supervisory\ \&\ Maintenance) = 0.50(\$831,140.77) = \$415,570.39$$

Assuming the local taxes were 2% of the fixed capital,

$$Local\ Taxes = 0.02(C_{FC}) = 0.02(\$3,921,295.47) = \$78,425.91$$

Assuming the insurance was 1.5% of the fixed capital,

$$Insurance = 0.015(C_{FC}) = 0.015(\$3,921,295.47) = \$58,819.43$$

General Expenses

Assuming the administrative costs were 25% of the overhead, packaging, and storage expenses,

$$Administrative\ Cost = 0.25(Overhead) = 0.25(\$415,570.39) = \$103,892.60$$

Assuming the distribution and selling expenses were 10% of the total manufacturing expenses,

$$Distribution\ \&\ Selling = 0.10(Total\ Manufacturing) = 0.10(\$21,794,870.56) = \$2,179,487.06$$

Assuming the research and development expenses were 5% of the total manufacturing expenses,

$$Research\ \&\ Development = 0.05(Total\ Manufacturing) = 0.05(\$21,794,870.56) = \$1,089,743.53$$

<u>Depreciation</u>

Depreciation was estimated to be approximately 10% of the fixed capital,

$$Depreciation = 0.10(C_{FC}) = 0.10(\$3,921,295.47) = \$392,129.55$$

Net Annual Profit

After subtracting the total expenses from the revenue earned from product sales, the income taxes (A_{IT}) were calculated by estimating the product of the net annual profit (A_{NP}) and the tax rate, assuming the tax rate was 34%,

$$A_{IT} = 0.34(A_{NP}) = 0.34(\$176,894,577.54) = \$60,144,156.36$$

Subtracting the income taxes from the net annual profit, the final net annual profit (A_{NNP}) of the thorium-recovery process plant after taxes was estimated to be,

$$A_{NNP} = A_{NP} - A_{IT} = (\$176,894,577.54) - (\$60,144,156.36) = \$116,750,421.18$$

The after-tax rate of return (i) was calculated as follows,

$$i = 1.5(\frac{A_{NNP}}{C_{TC}}) \times 100\% = 1.5(\frac{\$116,750,421.18}{\$4,509,489.79}) \times 100\% = 3,883.49\%$$

Appendix C: Chancellor's Honors Program Thesis Reflection (Samantha Karklins)

For my Chancellor's Honors Program Thesis Project, I completed a senior design project within the Department of Chemical & Biomolecular Engineering of the Tickle College of Engineering. Sponsored by the Electric Power Research Institute (EPRI), this project required my group and I to conduct a study-level design and economic analysis of a process extracting thorium (Th) and rare earth elements (REEs) from monazite ore. More specifically, the REEs extracted are cerium (Ce), lanthanum (La), and neodymium (Nd). After extraction from monazite ore, thorium can be used in the generation of thorium-based nuclear power, in which the nucleus of a thorium atom accepts a neutron and beta decays into isotope uranium-233. Then, nuclear fission of the isotope uranium-233 may produce fuel. Using OLI Flowsheet software and the economic methods outlined in Gael D. Ulrich's *Chemical Engineering: Process Design and Economics,* the thorium-recovery process we designed successfully extracted thorium and REE oxides from monazite ore with a net annual profit of approximately $116,000,000.00, yielding a return on investment of 3878%.

Because my honors thesis project was completed as a group, the work required to complete this project was divided among us so that each member made a significant contribution to the outcome of the study-level process design and economic analysis. My role in completing this project consisted of four major tasks. First, I actively assisted my group members in designing the thorium-recovery process using the OLI Flowsheet software installed on my laptop, ensuring the results from OLI Flowsheet were made accessible to all members of the group. Second, I assisted my group members in calculating the capital and manufacturing costs of the thorium recovery process. To assist with the capital cost, I calculated the size and cost of the three filters and conveyor belts involved in the process. To assist with the manufacturing cost, I calculated the annual cost of the natural gas used to operate the process. Third, I assisted my group members in providing further details of our thorium-recovery process by creating a series of charts for this report. I created the project assumptions chart, stream composition charts, and stream hazards chart, writing the respective sections of the report discussing those charts.

Lastly, I assisted my group members in writing this report of the results of our study-level design and economic analysis of the thorium-recovery process.

Although the Chancellor's Honors Program Thesis Projects may be completed in a variety of ways, I would not have wanted to complete my thesis project any other way than through a senior design project for the Department of Chemical & Biomolecular Engineering. By completing this study-level design and economic analysis of a process extracting thorium and rare earth elements from monazite ore, I have seen firsthand how the knowledge of chemical engineering principles I have accumulated throughout my undergraduate education may be used in a career as a chemical engineer. Additionally, through this honors thesis project, I have been able to experience how I can use what I have learned throughout my undergraduate education to help solve one of the grand challenges of engineering: providing energy through more clean and sustainable sources.