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Utilization of Monazite Ore for Sources of Thorium Oxide

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Utilization of Monazite Ore for Sources of Thorium Oxide

CBE 488: Honors Senior Design

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1. Introduction

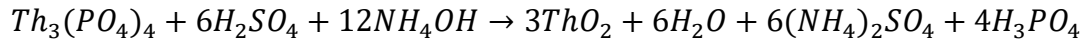
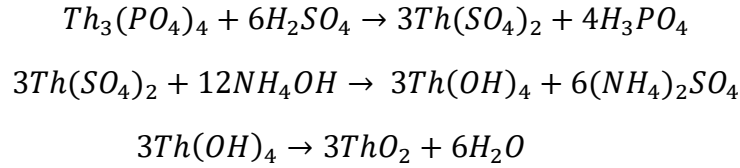
The objective of this paper is to determine the economic and process feasibility of extracting thorium oxide from monazite ore. As nuclear energy advances, fuels other than uranium may be needed as replacements. One potential fuel is thorium. Thorium, along with many other rare earth elements (REEs), are found in a sand-like mineral called monazite. Monazite is a byproduct of the mining industry, so it presents great potential as a source of thorium. In this study, we will evaluate the economic potential and feasibility of an acid extraction of thorium from monazite. The process will separate other REEs and recover phosphoric acid.

Design Objectives
1. Accuracy of economics of + 30% to -20%
2. Process scale: 1000 kg/hr of monazite
3. Feed composition in table 2
4. Process must be safe and environmentally friendly
5. Product streams are to be economically feasible
6. ChE index is 2019

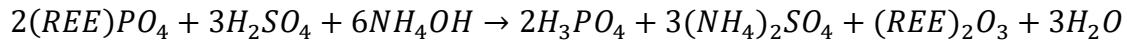
Table 1. List of Design Objectives

2. Synthesis Information for Processes

In the mining industry, monazite is typically a waste product; however, by using sulfuric acid, thorium and uranium can be extracted. The products can then be sold at a profit (2015, Rodliyah et al.). In this process, the desired product is thorium, which means we made all decisions to maximize the conversion to thorium. With this in mind, we used a reactor combined with a filtration process. The series of chemical reactions for thorium is as follows:



The following overall reaction is for the remaining rare earth elements in monazite:



The process relies on the easy separation of thorium from the mixed REE stream since thorium sulfate is solid at 57°C, while the mixed REEs remain in solution. Figure 1 shows the block flow diagram for the separation of thorium and the recycle system for H₂SO₄ and NaOH.

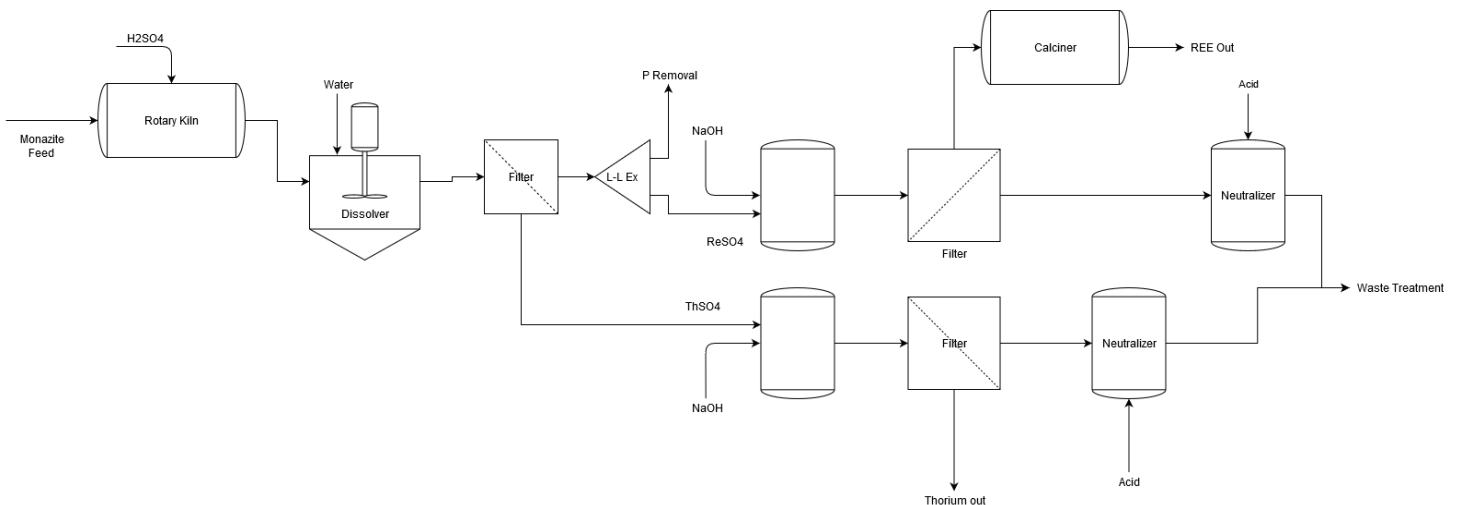


Figure 1: Block Flow Diagram

This process involves several challenges that must be overcome to ensure safety, maintainability, and ultimately, profitability. One such challenge is the abrasiveness of the feed material. Monazite will be damaging to the rotating kiln drum. Further investigation is required to determine design parameters of the drum to ensure it withstands the abrasion for prolonged

periods. This could limit the lifespan of the drum. Variability in the composition of monazite can also cause issues with sizing of the vessels. If the vessel is sized incorrectly it could restrict the maximum inlet flow rate.

2.1 Brief Literature Summary

“Decomposition of Monazite Concentrate in Sulfuric Acid” by Berry, L., V. Agarwai, J. Galvin, and M.S. Safarzadeh (2) looks at the various conditions that optimize the process of separating monazite by using an acid extraction. Based on that research there are three things needed to maximize thorium extraction efficiency. First, is a residence time of 5 hours. Second, a temperature between 180-250 C. Lastly, maintaining a 4:1 sulfuric acid to ore ratio. Thorium extraction is increased when the set temperature is towards the lower end of the range, but has little effect on the other REEs. Particle size during the addition of the acid has little effect on the REEs extraction.

The article “Process development to recover rare earth metals from monazite mineral: A review” by Kumari, Panda, Kumar, Kumar, and Lee (4) evaluates several methods to extract thorium and REEs from monazite. The article also gives an overview of the environmental impacts of extracting REEs from monazite. It cites that using a byproduct like monazite and utilizing reactant/byproduct recovery schemes can reduce the environmental impact of REE mining. This review suggests that NaOH treatments are more effective than H₂SO₄ acid leaching in conversion of sulfates to hydroxides.

The review article by Farzaneh Sadri et al., “A review on the cracking, baking and leaching processes of rare earth element concentrates,” (5) investigates several methods for recovering REE concentrates. The article outlines an industrially acceptable method to extract high purity

REEs. The first step of the process is acid baking. This is followed by neutralization and precipitation by increasing pH. Lastly, the process fluid is re-leached with HCl. It also cites that alkaline cracking is often a more economical process. This is due to avoiding equipment degradation caused by the phosphate-acid interaction.

2.2 Cost Information

The feed to the process includes monazite, H₂SO₄, and NH₄OH. The monazite composition is outlined in Table 4. Of the products, ThO₂ is the most valuable, but Nd₂O₃ is also highly valuable (Table 2.). Since the REEs are to be sold in a mixed stream, we estimated the products are worth half their potential value. Most of the feed cost comes from the monazite. Energy costs for the process are calculated using natural gas as the primary utility (Table 5).

Product	\$/kg
ThO ₂	80
Nd ₂ O ₃	60
La ₂ O ₃	2
Ce ₂ O ₃	2

Table 2. Cost of Products

Feed	\$/kg
Monazite	1
H ₂ SO ₄	0.04
NH ₄ OH	0.10

Table 3. Raw Materials Cost

Component	Mass %
Lanthanum	14.46
Cerium	29.17
Thorium	4.83
Phosphorous	12.89
Neodymium	12.01
Oxygen	26.64

Table 4. Monazite Composition

Utility	\$/1000 ft ³
Natural Gas	10.25

Table 5. Energy Cost

3. Method of Approach

The approach for this project begins with determining the design objectives and chemical reactions needed to achieve the desired result. With these considerations in mind, the block flow diagram was developed and a mass balance was calculated. This was followed by an analysis of process constraints, raw material costs, and economic potential. Next, the process design is modeled and simulated using the OLI Flowsheet software. Using this software, recycle structures and waste streams are examined as well as equipment design. Once the process was simulated and the final design parameters were decided upon, a full cost analysis was completed. This included equipment costs, operating costs, utilities costs, annualized costs, and profitability analysis. From the information gathered throughout this approach, a full report was completed by April 24, 2020 and submitted for review.

4. Results

4.1 Optimization

Certain optimizations can be performed despite the coarse approach taken in a study-level process draft. Much of this comes from determining the minimum materials required to perform the process to an adequate standard. In the first pass of modeling, we used an excess of the reagents and components required to drive the reaction to completion. In subsequent iterations, more care was taken to refine these values. Separate mass balance analyses were performed, in

order to determine more reasonable quantities. This serves to reduce material cost, equipment size, and process waste. Additionally, a scheme to extract the phosphoric acid from the process stream was designed to maximize profit.

4.2 Process Flow Diagram

Figure 2 shows the process flow with a corresponding chart including all flow rates. Not included due to software constraints is a 41.31 m³ rotary kiln located prior to the S-1 stream. The rotary kiln has an inlet flow of pure monazite and sulfuric acid. A second rotary kiln is located after filter 3 with S-16 as the inlet flow. This rotary kiln is 20.5 m³.

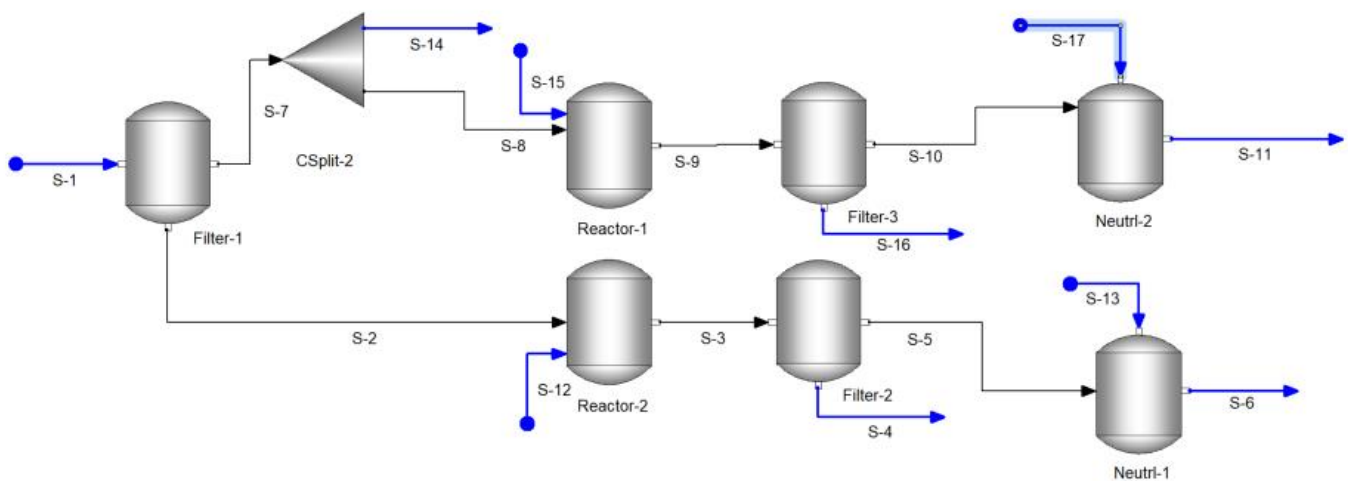


Figure 2. Process Flow Diagram

The first step of the process is feeding 1000 kg/hr of monazite and 1700 kg/hr of sulfuric acid into a rotary kiln at 300°C for 300 minutes. The literature supports that this reaction occurs best at this temperature and residence time to convert the phosphate forms of the REEs and thorium to sulfates. Next, this stream enters the first filter to separate the solid thorium sulfate from the aqueous REE stream. The solid thorium is then dissolved with NaOH in Reactor 2 to produce thorium oxide. In filter 2, the thorium is filtered out and the

waste aqueous stream is neutralized in neutralizer 1. From reactor 1, the aqueous REE stream goes through a liquid-liquid extraction to remove phosphoric acid. After the phosphoric acid is removed, the stream enters reactor 1 and reacts with NaOH to convert sulfates to hydroxides. Filter 3 removes the solid REEs and the liquid waste stream is neutralized in neutralizer 2. The solid REEs are then processed through the final rotary kiln to convert hydroxides into oxides.

Stream	Flow Rate (m³/hr)	Stream Location
S1	3.2	Filter 1 Inlet
S2	0.021	Reactor 2 Inlet
S3	1.13	Filter 2 Inlet
S4	5.57E-03	Thorium Outlet
S5	1.12	Neutralizer 1 Inlet
S6	1.12	Waste Outlet
S7	3.18	Extractor Outlet
S8	3.06	Reactor 1 Inlet
S9	11.7	Filter 3 Inlet
S10	11.5	Neutralizer 2 Inlet
S11	11.8	Waste Outlet
S12	1.1	Reactor 2 Inlet
S13	0.0035	Neutralizer 1 Inlet
S14	0.19	Extractor Outlet
S15	8.1	Reactor 1 Inlet
S16	0.21	Filter 3 Outlet
S17	0.49	Neutralizer 2 Inlet

Table 6. Process Flow Diagram Stream Information

Process Step	Size	Type
Filter 1	15 m ²	Single Vacuum Filter
Calciner 1	41.31 m ³	Stainless Steel Rotary Kiln
Calciner 2	21.91 m ³	Stainless Steel Rotary Kiln
Reactor 1	22.57 m ³	Stainless Steel Continuous Stream Reactor
Reactor 2	0.553 m ³	Stainless Steel Continuous Stream Reactor
Filter 2	8 m ²	Single Vacuum Filter
Filter 3	18 m ²	Single Vacuum Filter
Neutralizer 1	0.094 m ³	Vertical Oriented Stainless Steel Neutralizer
Neutralizer 2	0.839 m ³	Vertical Oriented Stainless Steel Neutralizer
Agitator for Reactor 1	-	Mechanical Seal Agitator with Propellor
Agitator for Reactor 2	-	Mechanical Seal Agitator with Propellor
Liquid-Liquid Extraction	47.12 m ²	Extractor
Liquid-Liquid Extraction	47.12 m ²	Stripper

Table 7. Process Equipment and Costs

4.3 Safety, Health and Environmental Analysis

The safety analysis chart for all chemicals in this process can be found in Appendix A (Table A.1). This process presents several challenges with regards to safety and the environment. In general, the process converts waste into valuable products, therefore it is an inherently environmentally friendly process. However, many hazardous, flammable, and toxic chemicals are used. Corrosion resistant materials must be used throughout the process to reduce the risk of loss of containment. Additionally, the process will be conducted at atmospheric pressure to avoid risks associated with high pressure processes. While more expensive, it is safer to include analyzers at critical sampling points to eliminate the need to take physical samples. Several chemical reactions occur in the process that are exothermic. With any exothermic reaction, there is potential for a runaway reaction. While unlikely in this process, all reactors will be continuous to better control the reaction. Process controls must be present to ensure safe operation. The aqueous waste streams of this process contain acids, bases, and sulfur trioxide. The waste

streams will be neutralized, but further processing and containment is necessary to make the waste streams safe. It is also important to ensure that a high degree of filtration in process streams is achieved to avoid excess thorium in other streams, as it is a safety hazard. An electro dialysis scheme would help to clean the waste stream, but further feasibility analysis must be conducted to design such a scheme in an efficient manner.

4.4 Capital Cost Estimates

To determine the equipment capital cost, it was first necessary to size the equipment based on residence times and flow rates. The cost for all stainless-steel equipment was estimated using charts in Ulrich. The total capital cost was found to be \$9.706 million. A list of the equipment costs is included in Table 8.

Process Step	Type	Bare Module Cost	Annualized Cost
Filter 1	Single Vacuum Filter	\$533,340.00	\$128,001.60
Calciner 1	Stainless Steel Rotary Kiln	\$2,504,040.00	\$600,969.60
Calciner 2	Stainless Steel Rotary Kiln	\$2,012,175.00	\$482,922.00
Reactor 1	Stainless Steel Continuous Stream Reactor	\$988,201.50	\$237,168.36
Reactor 2	Stainless Steel Continuous Stream Reactor	\$116,259.00	\$27,902.16
Filter 2	Single Vacuum Filter	\$426,672.00	\$102,401.28
Filter 3	Single Vacuum Filter	\$800,010.00	\$192,002.40
Neutralizer 1	Vertical Oriented Stainless Steel Neutralizer	\$232,518.00	\$55,804.32
Neutralizer 2	Vertical Oriented Stainless Steel Neutralizer	\$581,295.00	\$139,510.80
Agitator for Reactor 1	Mechanical Seal Agitator with Propellor	\$162,965.00	\$39,111.60
Agitator for Reactor 2	Mechanical Seal Agitator with Propellor	\$37,037.50	\$8,889.00
Liquid-Liquid Extraction	Extractor	\$655,820.00	\$157,396.80
Liquid-Liquid Extraction	Stripper	\$655,820.00	\$157,396.80
		\$9,706,153.00	\$2,329,476.72

Table 8. Capital Costs

4.4 Manufacturing Cost Estimates

The process requires several expenses related to normal operations. This includes direct costs, such as those associated with raw materials and operating labor; indirect costs, associated with overhead, insurance, and taxes; and utilities such as electricity and water. A summary of cost

estimates is shown in Table 9. The most significant of these expenses is raw material costs. The process is designed to use 1000 kg/hr of monazite ore, which is priced at \$1.00 per kg. When dealing with high value materials, it is expected for them to dominate the operating expenses. Necessary reagents for the process such as NaOH also contribute significantly.

Costing Category	Annual Cost
<i>Capital Investment</i>	
Fixed Capital	\$13,163,096.45
Working Capital	\$1,974,464.47
Total	\$15,137,560.92
<i>Direct Costs</i>	
Raw Materials	\$14,998,512.00
Operating Labor	\$1,023,255.06
Supervisory and Clerical Labor	\$204,651.01
Total	\$16,226,418.07
<i>Utility Costs</i>	
Electricity	\$72,558.00
Process Water	\$8,000.00
Waste Disposal	\$86,400.00
Natural Gas	\$112,404.00
Maintenance and Repairs	\$1,316,309.65
Operating Supplies	\$263,261.93
Laboratory Charges	\$204,651.01
Patents and Royalties	\$1,380,218.24
Total	\$3,443,802.83
<i>Indirect Costs</i>	
Overhead	\$859,534.25
Local Taxes	\$394,892.89
Insurance	\$394,892.89
Total	\$1,649,320.03
Total Manufacturing Expense	\$21,319,540.93
<i>General Expenses</i>	
Administrative Costs	\$214,883.56
Distribution and Selling	\$213,195.41
Research and Development	\$1,000,000.00
Total General Expense	\$1,428,078.97
Depreciation	\$1,316,309.65
Total Expenses	\$24,063,929.55
<i>Profit</i>	
Revenue from Sales	\$84,059,032.00
Annual Profit	\$59,995,102.46
Income Taxes	\$20,998,285.86
Annualized Equipment Costs	\$2,329,476.72
Net Annualized Profit	\$36,667,339.88

Table 9. Manufacturing Costs and Net Annualized Profit

5. Discussion of Results

These results summarize the equipment, process, economic commitments and safety concerns of extracting thorium from monazite. In order to optimize this series of reactions, it was determined that 11 pieces of major equipment are needed. Due to the corrosive material, all equipment should be stainless steel and should be sized (Table 7) to function at the appropriate flow rate. The initial capital investment will be \$9.706 million with an annual equipment upkeep cost of \$2.329 million. In addition to these costs, there are other direct and indirect manufacturing expenses, as listed in Table 9 that are \$21,319,540.93 yearly. It is expected that, after all expenses are considered, this process will have an annual profit of \$36,667,339.88 and a return on investment of 401%. While the financial considerations are promising, safety must also be considered. An analysis of each chemical throughout the process was researched and the list of potential concerns is shown in Table A.1. From this research it was determined that the main concerns are flammability, skin irritability, and potential water contamination from the toxicity of the chemicals. Flammability should be mitigated by performing the process at atmospheric pressure and using monitoring equipment to ensure the reactions are occurring properly. Nearly all the chemicals used can cause skin irritation so proper personal protective equipment should be used inside the facility at all times. To keep the water contamination risk low, the waste streams will be neutralized and although it is not required, it is recommended that an electro dialysis process be added in order to further treat the wastewater.

6. Conclusions

The process of extracting thorium from monazite ore to produce product streams of thorium oxide, phosphoric acid, and mixed rare earth oxides is economically incentivized based on the result of this study level design. With a capital investment of just under \$10 million, there is an expected annual net profit of \$40 million. Approximately half of the economic potential comes from the valuable product-thorium, while approximately 30% of the profit comes from phosphoric acid. Each REE was discounted to half worth since the REEs are to be sold in a mixed stream. Since Nd_2O_3 is significantly more valuable than La_2O_3 or Ce_2O_3 , it would likely be incentivized to separate Nd_2O_3 to capture its full value, making that stream more valuable than the thorium oxide stream. Phosphoric acid removal relies on liquid-liquid extraction of the process stream containing thorium and other REEs. The chemistry and costing of this process needs significant further investigation to define the necessary solvent, extractors, and evaporators to provide product-grade acid. Even if phosphoric acid is not refined into a salable product, the process is still profitable based on the profit from thorium and the mixed REEs. Several challenges exist in making this process reliable, safe, and profitable, however, with further study, this process could achieve profitability while maintaining environmental and personal safety.

Material	Flow Rate (\$/kg)	Bulk Price (\$/kg)	\$/yr	% of Cost/Profit
Feeds				
Monazite	1000	\$1.00	\$8,000,000	53%
NaOH	4872	\$0.14	\$5,456,640	36%
H2SO4	4872	\$0.04	\$1,525,872	10%
MIBK	1	\$2.00	\$16,000	0.10%
			\$14,998,512	
Products				
H3PO4	5120	\$0.64	\$26,214,400	31%
ThO2	55	\$80.00	\$35,200,000	42%
La2O3	198	\$1.00	\$1,580,944	2%
Ce2O3	198	\$1.00	\$1,591,408	2%
Nd2O3	81	\$30.00	\$19,472,280	23%
			\$84,059,032	
Economic Potential			\$69,060,520	

Table 10. Economic Potential Analysis

7. Recommendations

This study level design exposed several areas in which further study is necessary. For all processes, we assumed 100% conversion of reactions and 100% filtration. In the next phase of design, lab work is needed to determine more realistic figures for conversion/reaction kinetics and percent filtration. In OLI simulations, thorium sulfate was converted to thorium oxide when exposed to heat and NaOH. This was an unexpected result as thorium sulfate was thought to need calcining to convert to oxide. This result needs to be tested at lab scale.

Next, the liquid-liquid extraction of phosphoric acid by MIBK needs to be refined to better understand the process equipment required, the amount of solvent necessary, and compatibility with the thorium/REEs present in the stream. From an environmental perspective, it is important to explore an electrodialysis scheme to treat the wastewater. While expensive, this could recover

some of the acids and bases used in the process and reduce cost by allowing the water to be internally treated and recycled.

Lastly, the REEs are estimated to sell at a 50% discount when sold in a mixed stream. It would be worth investigating how to efficiently separate all the REEs to capture additional income.

This could result in an additional \$25 million a year.

8. References

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- (3) Kemperman, A.J.B., *Handbook Bipolar Membrane Technology*, 149-219, Twente University Press, Enschede, ISBN 9036515203 (2000)
- (4) Kumari, Archana, Panda, Rekha, Kumar Jha, Manis, Kumar, j. Rajesh, and Lee, Jin Young. "Process development to recover rare earth metals from monazite mineral: A review." *Minerals Engineering*, Vol. 79, p. 102-115. (2015).
- (5) Sadri, F., A.M. Nazari, A. Ghahreman, "A Review of the Cracking, Backing and Leaching Processes of Rare Earth Element Concentrates", *J. of Rare Earths*, 35(8), 739-756 (2017)
- (6) Ulrich, G.D. and P.T. Vasudevan, *Chemical Engineering Process Design and Economics A Practical Guide*, 2nd Edition, Process Publishing, Durham (2004)

Appendix A: Additional Tables & Charts

Compound	Molecular Weight	Melting Point (°C)	Boiling Point (°C)	Liquid Density (g/cm ³)	Flammability Risk	Poisonous	Toxin	Carcinogenic	Mutagenic	Irritating	Corrosive
LaPO4	236.90	N/A	N/A	N/A	N/A	N/A	N/A	N/A	No	Yes	No
CePO4	234.97	N/A	N/A	N/A	Low	N/A	Yes	Yes	N/A	Yes	N/A
Th3(PO4)4	1075.88	3170	4400	9.86	Low	Yes	Yes	Yes	N/A	Yes	N/A
NdPO4	238.97	N/A	N/A	N/A	Yes	N/A	Yes	N/A	N/A	Yes	Yes
U3(PO4)4	1093.88	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
H2SO4	98.60	10.37	279	1.83	No	Yes	No	Yes	N/A	Yes	Yes
NaOH	39.99	318	1388	2.13	No	No	No	No	No	Yes	Yes
La2O3	325.80	2315	4200	6.5	No	No	Yes	No	No	Yes	No
Ce2O3	328.24	2177	3730	6.2	N/A	N/A	N/A	N/A	N/A	N/A	N/A
ThO2	264.00	3390	4400	10	Yes	Yes	Yes	Yes	No	No	No
Nd2O2	320.00	2233	3760	7.24	No	No	No	No	N/A	N/A	No
UO2	270.00	2865	N/A	10.9	Yes	Yes	Yes	Yes	N/A	N/A	Yes
H3PO4	96.97	42.35	158	1.88	No	No	Low	N/A	N/A	Yes	Yes
H2O	18.00	0	99.98	1	No	No	No	No	No	No	No
SO3	80.07	62.42	113	1.92	Yes	Yes	Yes	Yes	N/A	Yes	Yes
MIBK	100.16	-84.7	117	0.802	Yes	No	No	No	No	Yes	N/A

Table A.1 Hazard Analysis

Figure A.1: Graphs used to cost Reactor 1 (Ulrich)

FIGURE 5.44 Purchased equipment costs for (a) horizontally oriented and (b) vertically oriented process vessels. Bases for costs are carbon steel construction and internal pressure less than 4 barg. Installation factors F_{BM}^a for higher pressures and different construction are found in Figure 5.46. For jacketed or internally heated vessels or autoclaves, see Figure 5.23. For packed or tray towers, add bare module costs of packing or trays from Figure 5.47 or 5.48.

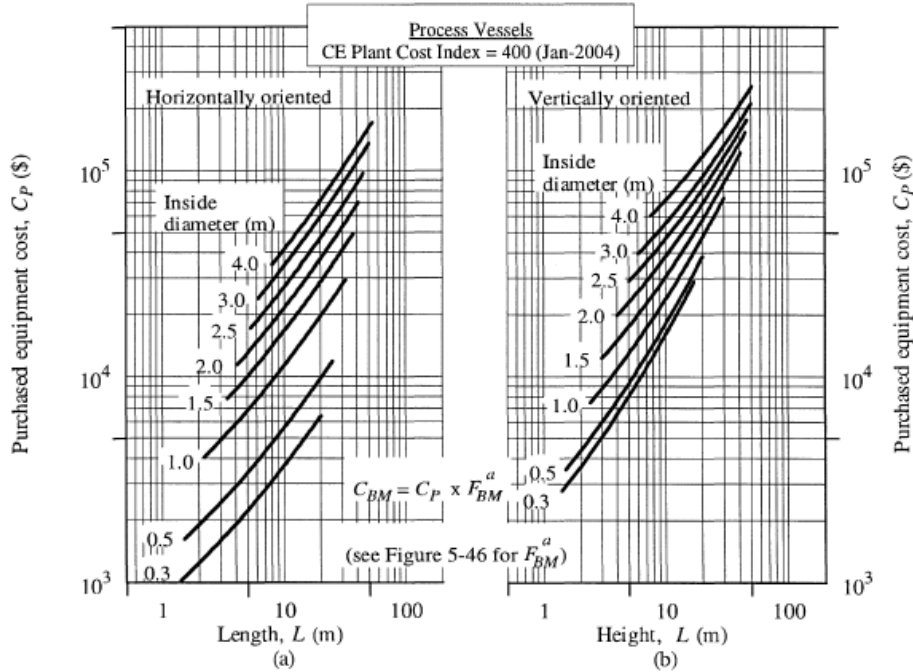


Figure A.2: Graphs used to Cost Reactor 1 Agitator (Ulrich)

FIGURE 5.42 Purchased equipment costs for agitators and inline mixers. Cost of agitators includes motor, speed reducer, and impeller ready for installation in a vessel. Stuffing box seals can contain pressures up to 10 barg. Mechanical seals are suitable for toxic or critical fluids at pressures up to 80 barg. Vessel cost is assessed separately. Inline mixer is assumed to be installed in a pipeline with no vessel (pressures up to 10 barg - for higher pressures use inline mixer installation factors the same as centrifugal pumps as derived from Figures 5.49, 5.50, and 5.51.)

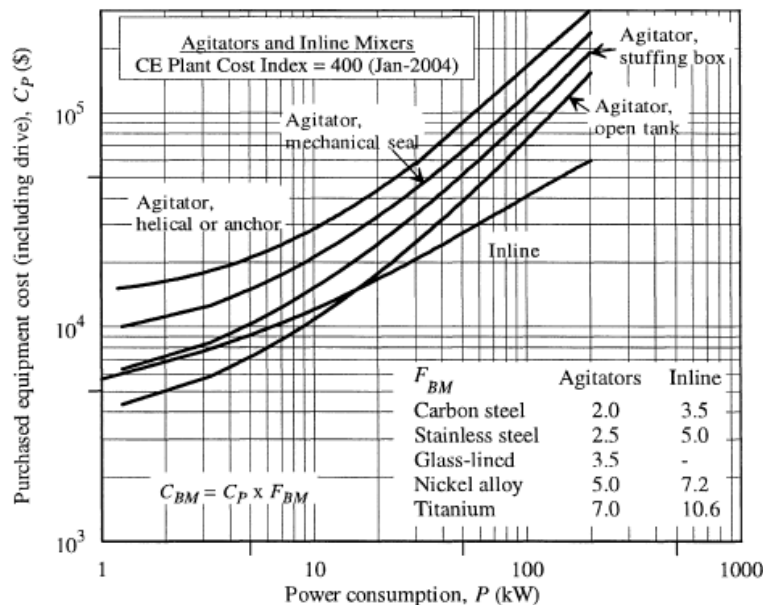


TABLE 4.16 Criteria and Data for the Preliminary Design of Agitators and Mixers

	Type of mixer							
	Fluid-activated				Mechanically-aided			
	Fluid jet	Orific plate (pipeline)	Motionless mixer	Gas sparger	Inline		Agitated	
Pump or agitated line					Rotor-stator	Rotor-stator	Propeller	
<i>Range of common equipment sizes</i>								
Vessel diameter, D_1 (m)	30 D_a	0.005 to 0.5	0.003 to 2.0 ^a	0.01 to 6.0	0.01 to 0.5	0.01 to 0.5	< 1.5	< 50
Vessel length or height, L (m)	100 D_a	0.2 D_1 to 0.5 D_1	0.03 to 80	0.03 to 15	0.05 to 0.5	0.05 to 0.5	< 2.5	< 20
Agitator diameter, D_a (m)	0.001 to 0.1	50 D_1			0.005 to 0.5	0.005 to 0.5	0.005 to 0.5	< 1.5
Vessel volume, V (m^3)				< 360			< 5	< 40,000
<i>Mixed fluid flow rate, \dot{n} (kg/s)</i>								
Gases	0.001 to 100	0.03 to 300	0.001 to 100	^a				
Liquids	0.1 to 10,000	0.16 to 16	0.01 to 1000		0.1 to 400	0.1 to 400		
<i>Typical residence time, θ (s)</i>								
Mixing	0.1 to 200	0.16 to 16	0.02 to 5.0	^a	0.15 to 1 ^b	0.15 to 1 ^b	12,000 ($\mu V/P$) ^{1/2} $V^{0.2}$	60 to 20,000
Liquid-liquid extraction								60
Solids leaching								60
Chemical reaction				^a				From kinetic analysis
<i>Viscosity range, μ (Pa-s)</i>								
	0 to 0.01	0 to 0.1	0 to 2000	0 to 1.0	0 to 1.0	0 to 25	0 to 10	0 to 5
<i>Volume fraction of dispersed medium, ϕ</i>								
	< 0.4	< 0.4		< 0.1	< 0.8	< 0.8	< 0.8	< 0.8
<i>Suitability</i>								
Gas-gas mixing	A	A	A	X	D	X	X	E
Gas-liquid mixing	E	D	B	A	B	B	B	D
Liquid-liquid mixing (miscible)	A	A	A	B	A	D	D	B
Liquid-liquid dispersion (immiscible)	B	B	D	D	B	A	A	B
Liquid-solid suspension	B	B	B	D	E	D	D	B
Paste-past mixing	X	X	A	X	X	X	X	X
Solid-solid mixing	X	X	D	^d	X	X	X	X
Heat-transfer enhancement	A	B	A	B	D	X	X	B
Chemical reaction	A	A	A	A	B	A	A	B
Liquid-solid mixing	D	D	D	B	D	B	B	B
Mixing of sticky materials	E	E	A	X	D	X	X	E
<i>Pressure differential Δp (bar)</i>								
Gases	1 to 3	0.0002 to 0.001	6x10-6	^e				
Liquids	0.3 to 1.0	0.05 to 0.3	0.006 to 0.6	^e				
<i>Power consumption P (kW) ^f</i>								
Gas-gas	1.5 \dot{m} to 5 \dot{m}	0.8 \dot{m}	0.4 \dot{m}		0.15 \dot{m} to 1.5 \dot{m}	-	-	-
Gas-liquid		0.06 \dot{m}	0.03 \dot{m}	^e	0.04 \dot{m}	0.1 \dot{m}	0.2 $V^{0.95}$ to 2 $V^{0.95}$	
Liquid-liquid: Mild	0.007 \dot{m}	0.007 \dot{m}	0.003 \dot{m}	^e	0.007 \dot{m}	-	0.5 V	0.1 $V^{0.8}$ to 0.2 $V^{0.8}$
Vigorous	0.01 \dot{m}	0.02 \dot{m}	0.01 \dot{m}	^e	0.02 \dot{m}	0.06 \dot{m}	1.5 V	0.4 $V^{0.8}$ to 0.4 $V^{0.8}$
Intense	0.02 \dot{m}	0.04 \dot{m}	0.02 \dot{m}	^e	0.04 \dot{m}	0.1 \dot{m}	3 V	0.8 $V^{0.8}$ to 2.0 $V^{0.8}$
Liquid-solid	0.007 \dot{m} to 0.02 \dot{m}	0.007 \dot{m} to 0.04 \dot{m}	0.01 \dot{m} to 0.1 \dot{m}	^e	0.007 \dot{m} to 0.04 \dot{m}	0.01 \dot{m} to 0.1 \dot{m}	3 V	0.1 $V^{0.8}$ to 2.0 $V^{0.8}$
Paste-paste			0.1 \dot{m}					
Solid-solid								
<i>Typical overall heat transfer coefficient, U ($J/m^2 \cdot s \cdot K$)</i>								see Tables 4.15

- A excellent or no limitations
- B modest limitations
- C special units available at higher cost to minimize problems
- D limited in this regard
- E severely limited in this regard
- X unacceptable

^a Gas fluxes typically range from 0.003 to 0.016 m^3 per second per square meter of vessel cross section. Bubble rise velocities are normally between 0.15 to 0.30 m/s. For detailed design procedures, see Tilton and Russell [1982].

^b This is residence time within the pump or line mixer itself. For a pump installed in an external pipe loop, the time required for circulating a volume equal to the tank contents is considered adequate for mixing.

^c See Table 4.5.

^d See fluid beds.

Appendix B: Sample Calculations

Cost of Material per Hour

$$\text{Cost/hour} = \text{Mass flow rate (kg/hr)} * \$/\text{kg} = \$/\text{hr}$$

Level 2 Economic Potential

$$\begin{aligned} \text{Potential} &= \text{Product Value} + \text{Byproduct Value} - \text{Raw Material Costs} \\ &= \frac{\$17,333.23}{\text{hour}} * \frac{24\text{hrs}}{\text{day}} * \frac{365\text{days}}{\text{year}} - \frac{\$1,963.55}{\text{hour}} * \frac{24\text{hours}}{\text{day}} * 365 \frac{\text{days}}{\text{year}} = \frac{\$282,555,945}{\text{year}} \end{aligned}$$

Reactor 1 Sizing

$$\begin{aligned} \text{Volume} &= \text{Volume Stream Feed} / \text{Residence Time} \\ Fv &= 11.285 \frac{\text{m}^3}{\text{hr}} * (60 \text{ mins}) * \left(\frac{1 \text{ hr}}{60 \text{ min}}\right) = 22.57 \text{ m}^3 \\ \text{Volume} &= \frac{3 * \pi * D^3}{4} \\ D &= \left(\frac{4 * 22.57}{\pi}\right)^{\frac{1}{3}} = 3.06 \text{ m} \\ \text{Height} &= 3 * D \\ \text{Height} &= 3 * 2.65 \text{ m} = 9.18 \text{ m} \end{aligned}$$

$$\begin{aligned} &\text{Using Graph A.1} \\ &Fp \times Fm = 4 * 1.5 = 6 \\ &Fbm \rightarrow \text{from graph} = 13 \\ &\text{From graph} \rightarrow Cp = \$38,500 \\ &Cbm = \frac{\text{Index (2019)}}{\text{Index (2004)}} * Fbm * Cp \\ &Cbm = \left(\frac{596.2}{400}\right) * \$52,000 * 13 = \$988,201.50 \\ &\text{Annualized Cost} = Cbm * 0.24 \\ &\text{Annualized Cost} = \$988,201.50 * 0.24 = \$237,168.36 \end{aligned}$$

Agitator Costs for Reactor 1

Agitator Type: Mechanical Seal Propeller agitator

$$\begin{aligned} &\text{Power consumption is determined by Table 4.16 in Ulrich} \\ &\text{Power consumption} = 12.10 \\ &Cb = \$44,000 \end{aligned}$$

$$\begin{aligned}
 F_{bm} &= 2.5 \\
 \text{Chem E Index} &= 596.2 \\
 C_{bm} &= \$162,965 \\
 \text{Annualized Costs} &= \$39,111.60
 \end{aligned}$$

Extractor and Splitter Costs

$$\begin{aligned}
 \text{Volumetric Flow Rate} &= 3.18 \text{ m}^3/\text{hr} \\
 \text{Diameter of Column} &= 2\text{m} \\
 \text{Tray need} &= 4 \text{ stages plus additional height for the top and bottom} \\
 \text{Height} &= 14 \text{ meters}
 \end{aligned}$$

$$\begin{aligned}
 &\text{Using Graph A.1} \\
 F_p \times F_m &= 1.5 \times 4 = 6 \\
 F_{bm} &\rightarrow \text{from graph} = 11 \\
 \text{From graph} &\rightarrow C_p = \$40,000
 \end{aligned}$$

$$C_{bm} = \frac{\$40,000 * 596.2}{400} * 11 = \$655,820.00$$

$$\begin{aligned}
 \text{Annualized Costs} &= \$157,396.80 \\
 \text{Extractor costs} &= \text{Splitter costs} \\
 \text{Total } C_{bm} &= \$1,311,640.00
 \end{aligned}$$

Rotary Kiln / Calciner Sizing

$$T = \frac{0.19 * L}{N * D * S}$$

$$\begin{aligned}
 T &= \text{residence time (min)} \\
 L &= \text{kiln length (ft)} \\
 N &= \text{revolutions / min} \\
 D &= \text{kiln diameter (ft)} \\
 S &= \text{kiln slope (ft/ft)}
 \end{aligned}$$

$$\frac{0.19 * 200}{1 * 0.05 * 3} \approx 4.5 \text{ hrs}$$