

SENIOR DESIGN AND PROFESSIONALISM II SPRING 2020

FINAL PROJECT REPORT

- 1. Apply a systematic design process, incorporate engineering codes, standards, and realistic constraints that include economic; environmental; sustainability; manufacturability; ethical; health and safety; social; and political considerations in solving the design problem.**
- 2. Work in groups**

End of Semester Project Report Format

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1. Letter of Transmittal

The letter of transmittal captures the main points in in the EXECUTIVE SUMMARY.

(i) It clearly spells out why the report has been prepared.

(ii) Essential results, conclusions and recommendations are presented

See your College of Engineering Senior Design Manual for an example.

2. The EXECUTIVE SUMMARY

Your Executive Summary is a carefully distilled form of your complete design report and should have the following components:.

(a) It should clearly state the need for the design and its objectives

(b) It should clearly summarize what was done and main findings (process and production capacity, plant location and plant design and capital cost estimates, the capital investment , return on investment including, discounted cash flow internal rate of return, payback period etc)

(c) How the final results were obtained

(d) Clearly presented conclusions and recommendations.

Design Constraints and Design Basis

External constraints:

1. Physical laws
2. Standards and codes
3. Economic
4. Government and safety regulations
5. Resources

Internal constraints:

1. Choice of process
2. Process conditions
3. Materials and equipment
4. Time (for completion of design)
5. Personnel

Setting the Design Basis

1. Production rate
2. Purity specifications of the main product
3. The system of units to be used
4. The national, local, or company design codes that must be followed
5. Details of raw materials that are available
6. Information on potential sites where the plant might be located, including climate data, seismic conditions, and infrastructure availability
7. Information on the conditions, availability, and price of utility services such as fuel gas, steam, cooling water, process air, process water, and electricity that will be needed to run the process

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 2: An ability to apply engineering design to produce solutions that meet specified needs with consideration of public health, safety, and welfare, as well as global, cultural, social, environmental, and economic factor. Assignments that test competency in this area will include problems testing students' understanding of the design process involving *devising a system, component, or process to meet desired needs and specifications within constraints, through an iterative, creative, decision-making process in which the basic sciences, mathematics, and engineering sciences are applied to convert resources into solutions.*

Name: _____

Title of Assignment: _____ Due Date _____

Competency Area	Points Assigned	Points Received
1. Ability to Define/Understand the Problem and then Plan the Project	60	
(i) Identify the customer and the needs.	8	
(ii) Identify and list the design objectives.	8	
(i) Identify the design constraints.	4	
(ii) Define the design strategy and methodology.	4	
(iii) Identify and break down work into tasks and subtasks and identify the personnel and deliverables for each.	20	
(iv) Develop a Gantt chart and critical path analysis for managing the project.	8	
(v) Establish major milestones for tracking progress and define performance metrics to measure success	8	
2. Ability to Conduct a Review of the Literature	80	
(i) Identify the types of information needed for a complete understanding of all aspects of the project (Based on task described in the project planning).	20	
(ii) Gather information on relevant fundamentals, theory / concept (demonstrate technical competence) and relate them to the design.	40	
(iii) Provide the sources in a list of references properly cited in the literature review section and relevant sections of the report.	20	
3. Ability to perform preliminary design by generation of ideas and applying creativity through generation of ideas and evaluation of candidate solutions	60	
(i) Perform process design and analysis with design constraints; Sorting/screening of process technology alternatives based on availability and cost of raw materials, reaction selectivity and yield, thermodynamic barriers to separating undesired components, and safety and operability	40	
(ii) Evaluate candidate solutions to arrive at feasible designs	20	
4. Ability to Perform Preliminary and Detailed Design	200	
(i) Prepare a process flow diagram (PFD) and present a detailed description of the process indicating the functions of the major equipment components and controls, stream connectivity, safety, and other technical matters of importance	20	
(ii) Carry out a HYSYS/ASPEN simulation of the process showing a converged HYSYS/ASPEN model of the process. State assumptions/parameters used in modeling major equipment	20	
(ii) Prepare a chemical engineering design of (i) reactor(s), (ii) absorption column(s) and (ii) product distillation units, stating the materials of construction and the operating temperatures and pressures	60	
(iii) Prepare mechanical engineering design of the reactor(s) and distillation columns with appropriate dimensioned sketches	20	
(iii) Identify applicable codes and standards for the chemical and mechanical engineering designs. Indicate assumptions made and their justification, and support calculations should be in consistent units; either English Units or SI. Indicate possible sources of error	20	
(iv) Economic Evaluation: Estimate capital costs including engineering, materials, installation, interest, return on investment, taxes and contingency. Total unit conversion cost of product, a sensitivity study on the price of the raw material, and the economic methods used should be included	30	
(v) Prepare a control scheme to ensure safe operation of the plant and carry out Safety Reviews; Hazards, operability and inherent safe (including the use of Safety Data Sheets submitted in the Appendix) or HAZOP analysis	30	
Total	400	

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SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER**

ABET Student Learning Outcome 3: An ability to communicate effectively with a range of audiences (Written Technical Communication)

Assignments that test competency in written communication will include problems that test the student's ability to logically arrange information, use correct grammar, use appropriate format, use discipline specific conventions including citations, and use acceptable graphical conventions.

Name: _____

Title of Assignment: _____ Due Date _____

Competency Area	Points Assigned	Points Received
1. Ability to prepare an executive summary for the report	10	
Students are able to prepare an executive summary of 5 to 8 pages	10	
2. Ability to organize, plan and properly format a written technical report	80	
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	
(ii) Students provide Title Page, and Table of Contents, list of Figures, and List of Tables properly formatted.	12	
(iii) Students provide figure number and title for each figure in the report, reference each figure, and completely discuss each figure in the report in accord with standards in the project manual.	12	
(iv) Students provide table number and title for each table in the report in accord with standards in the project manual, reference each table, and completely discuss each table in the report.	12	
(v) Students properly cite references in the report and provide well formatted reference list at the end.	12	
(vi) Students prepare the written report in accord with standard report formatting provided in the Senior Projects Report Manual.	12	
3. Ability to compose original texts and properly apply the conventions of written language	50	
(i) Properly apply capitalization, punctuation, and penmanship, to communicate clearly.	15	
(ii) Spell proficiently	10	
(iii) Apply standard grammar and usage to communicate clearly and effectively in writing including • using complete sentences, varying the types such as compound and complex to match meanings and purposes • properly employing standard English usage in writing for audiences, including subject-verb agreement, pronoun referents, and parts of speech • properly using adjectives (comparative and superlative forms) and adverbs appropriately to make writing vivid or precise • properly using prepositional phrases to elaborate written ideas • properly using conjunctions to connect ideas meaningfully	20	
(vi) Use available technology to support aspects of creating, revising, editing, spell checking, and publishing the report.	5	
4. Ability to provide appropriate discussion, conclusions and recommendations	60	
(i) Summarize the goals, objectives, and indicate whether they were met.	10	
(ii) Summarize the results.	30	
(iii) Summarize constraints and codes and indicate whether they were met.	10	
(iv) Provide logical conclusions and recommendations (including strengths and weaknesses).	10	
Total	200	

**ABOVE AVERAGE
PROJECT REPORT**

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 2: An ability to apply engineering design to produce solutions that meet specified needs with consideration of public health, safety, and welfare, as well as global, cultural, social, environmental, and economic factor. Assignments that test competency in this area will include problems testing students' understanding of the design process involving devising a system, component, or process to meet desired needs and specifications within constraints, through an iterative, creative, decision-making process in which the basic sciences, mathematics, and engineering sciences are applied to convert resources into solutions.

Name: Group 2 (Bio)

Title of Assignment: Final Report Due Date: 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to Define/Understand the Problem and then Plan the Project	60	55
(i) Identify the customer and the needs.	8	8
(ii) Identify and list the design objectives.	8	8
(i) Identify the design constraints.	4	4
(ii) Define the design strategy and methodology.	4	4
(iii) Identify and break down work into tasks and subtasks and identify the personnel and deliverables for each.	20	20
(iv) Develop a Gantt chart and critical path analysis for managing the project.	8	8
(v) Establish major milestones for tracking progress and define performance metrics to measure success	8	8
2. Ability to Conduct a Review of the Literature	80	80
(i) Identify the types of information needed for a complete understanding of all aspects of the project (Based on task described in the project planning).	20	20
(ii) Gather information on relevant fundamentals, theory / concept (demonstrate technical competence) and relate them to the design.	40	40
(iii) Provide the sources in a list of references properly cited in the literature review section and relevant sections of the report.	20	20
3. Ability to perform preliminary design by generation of ideas and applying creativity through generation of ideas and evaluation of candidate solutions	60	60
(i) Perform process design and analysis with design constraints; Sorting/screening of process technology alternatives based on availability and cost of raw materials, reaction selectivity and yield, thermodynamic barriers to separating undesired components, and safety and operability	40	40 40
(ii) Evaluate candidate solutions to arrive at feasible designs	20	20 20
4. Ability to Perform Preliminary and Detailed Design	200	195
(i) Prepare a process flow diagram (PFD) and present a detailed description of the process indicating the functions of the major equipment components and controls, stream connectivity, safety, and other technical matters of importance	20	20 20
(ii) Carry out a HYSYS/ASPEN simulation of the process showing a converged HYSYS/ASPEN model of the process. State assumptions/parameters used in modeling major equipment	20	20 20
(ii) Prepare a chemical engineering design of (i) reactor(s), (ii) absorption column(s) and (ii) product distillation units, stating the materials of construction and the operating temperatures and pressures	60	60 60
(iii) Prepare mechanical engineering design of the reactor(s) and distillation columns with appropriate dimensioned sketches	20	20 20
(iii) Identify applicable codes and standards for the chemical and mechanical engineering designs. Indicate assumptions made and their justification, and support calculations should be in consistent units, either English Units or SI. Indicate possible sources of error	20	20 20
(iv) Economic Evaluation: Estimate capital costs including engineering, materials, installation, interest, return on investment, taxes and contingency. Total unit conversion cost of product, a sensitivity study on the price of the raw material, and the economic methods used should be included	30	25 25
(v) Prepare a control scheme to ensure safe operation of the plant and carry out Safety Reviews; Hazards, operability and inherent safe (including the use of Safety Data Sheets submitted in the Appendix) or HAZOP analysis	30	30 30
Total	400	395

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Name: Group 2 (Bio)

Title of Assignment: Final Report Due Date: 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to prepare an executive summary for the report	10	10
Students are able to prepare an executive summary of 5 to 8 pages	10	10
2. Ability to organize, plan and properly format a written technical report	80	80
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	20
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	20
(ii) Students provide Title Page, and Table of Contents, list of Figures, and List of Tables properly formatted.	12	12
(iii) Students provide figure number and title for each figure in the report, reference each figure, and completely discuss each figure in the report in accord with standards in the project manual.	12	12
(iv) Students provide table number and title for each table in the report in accord with standards in the project manual, reference each table, and completely discuss each table in the report.	12	12
(v) Students properly cite references in the report and provide well formatted reference list at the end.	12	12
(vi) Students prepare the written report in accord with standard report formatting provided in the Senior Projects Report Manual.	12	12
3. Ability to compose original texts and properly apply the conventions of written language	50	50
(i) Properly apply capitalization, punctuation, and penmanship, to communicate clearly.	15	15
(ii) Spell proficiently	10	10
(iii) Apply standard grammar and usage to communicate clearly and effectively in writing including • using complete sentences, varying the types such as compound and complex to match meanings and purposes • properly employing standard English usage in writing for audiences, including subject-verb agreement, pronoun referents, and parts of speech • properly using adjectives (comparative and superlative forms) and adverbs appropriately to make writing vivid or precise • properly using prepositional phrases to elaborate written ideas • properly using conjunctions to connect ideas meaningfully	20	20
(vi) Use available technology to support aspects of creating, revising, editing, spell checking, and publishing the report.	5	5
4. Ability to provide appropriate discussion, conclusions and recommendations	60	50
(i) Summarize the goals, objectives, and indicate whether they were met.	10	10
(ii) Summarize the results.	30	25
(iii) Summarize constraints and codes and indicate whether they were met.	10	7
(iv) Provide logical conclusions and recommendations (including strengths and weaknesses).	10	8
Total	200	190

“Design of a Grass-root Plant for the Production of Glucose-Fructose Syrup”

Senior Design & Professionalism II Final Report

Submitted by

Charles Kellum

Kameron Flournoy

Sabrina Noe

Temiloluwa Oluyadi

Unyime Effiong

To

Dr. Lealon Martin

Senior Design Instructor

Department of Chemical Engineering

College of Engineering

Prairie View A&M University

May 4, 2020

*Very good
report
Overall.*

*You need
to be
careful
with tens*

LETTER OF TRANSMITTAL

May 4th, 2020

Department of Chemical Engineering
Prairie View A&M University
700 University Drive
Prairie View, TX 77446

Dear Dr. Martin,

Based on our senior design problem statement, we were able to complete the design of a chemical plant that produces high fructose corn syrup (HFCS) after evaluating various design alternatives using a theoretical economic indicator. Enclosed in this report, you will find our plant design to produce approximately 705 million pounds per year of HFCS through enzymatic corn starch hydrolysis. Of this total syrup production, 52% will be syrup containing 55% fructose (364 million annually) and the remaining will be syrup containing 42% fructose (341 million annually). Enzymatic corn starch hydrolysis was the production route selected and information about design specifications of major equipment, a HAZOP analysis of the packed bed reactor for glucose isomerization, and an economic analysis of the entire plant are provided

Sincerely,


Sabrina Noe


Unyime Effiong


Charles Kellum


Kameron Flournoy


Temiloluwa Oluyadi

ACKNOWLEDGEMENTS

We would like to acknowledge the staff of the Roy G. Perry College of Engineering-Department of Chemical Engineering at Prairie View A&M University for providing resources and advice that were essential to the completion of this senior design project. Special thanks to Dr. Lealon Martin and Dr. Michael Gyamerah for advising the bioengineering senior design groups. The assistance of the Industrial Advisory Council in judges the design presentations as well as providing helpful feedback on the project design is also highly appreciated.

EXECUTIVE SUMMARY

The goal of this project is to design a chemical plant that produces glucose-fructose syrup that will cater to the supply and demand for this product within the food and beverage industry. This will be accomplished by making a complete design for all major equipment, conducting a plant economic analysis, creating a control scheme and performing a HAZOP analysis for one major equipment. The production of high fructose corn syrup (HFCS) involves three catalytic reactions (liquefaction, saccharification, and glucose isomerization) which are carried out within our major equipment. In the process of production, purification steps using ion exchangers, absorption columns, and a rotary filter are essential to ensure product quality specifications are met. The total plant capacity is 706 MM lbs/yr (1.5 M metric tons) with a production run time of 330 days annually. Two grades of syrup; HFCS 42 (used for food processes and baked goods) and HFCS 55 (used for carbonated drinks and juices) will be produced at an annual rate of 341 MM lbs/yr and 365 MM lbs/yr respectively. The plant is located in Asbury, Iowa and the plant life is 20 years. HFCS 42 and HFCS 55 are sold at 25.5 ¢/lb and 32.75 ¢/lb respectively. Based on economic analysis, the total module cost of all the equipment is about \$73 million and the cost of manufacturing without depreciation is about \$113 million. The total capital investment is about \$88 million and net annual profit is about \$57 million. The discounted cash flow rate of return and return on investment are about 50% and 78% respectively with a discounted payback period of about 2 years. Overall, this plant design is deemed as safe and profitable based on results provided throughout this report.

Assumptions made during the design of this process and are presented in Section 10.1 of this report.

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1. PROJECT SCOPE

1.1 Problem Statement

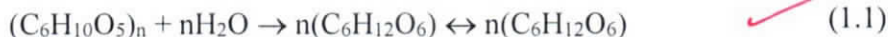
1st tell me what you're doing. What is the problem

Glucose – fructose syrup (GFS) is a liquid sweetener composed of multiple glucose and fructose chains. It can be obtained from various sources of starch such as corn, wheat, potato, tapioca, etc. GFS has become a staple alternative sweetener from graduated cane sugar in the food and beverage industry, owing to various attributes such as a high sweetness content, its non-crystalline nature, its ability to keep products fresh by maintaining content moisture, etc. GFS obtained from corn starch is generally referred to as high fructose corn syrup (HFCS). HFCS 55 with 55% fructose and HFCS 42 with 42% fructose are the grades that will be produced with this design. Corn gluten feed is also obtained from the purification processes and will be sold to local farmers. HFCS 42 is mainly used in processed and baked goods due to its mild sweetness level while HFCS 55 is used in carbonated drinks and juices because of its high level of sweetness.

Granulated cane sugar has become a relatively expensive sweetener for use in the food and beverage industry, therefore the search for an alternative sweetener is the ultimate goal for these industries. HFCS serves as an appropriate replacement for cane sugar in the food and beverage industry due to it being a relatively cheaper alternative compared to granulated cane sugar. From the preliminary economic analysis of different raw materials, corn starch was selected for use as it was the cheapest.

The production begins when a starch slurry is formed which undergoes three enzymatic reactions (liquefaction, saccharification, and glucose isomerization) with additional purification steps. Liquefaction gelatinizes the slurry using the α -amylase enzyme, saccharification creates the glucose syrup using the glucoamylase enzyme, and glucose isomerization creates the fructose-glucose ratio in the syrup using the glucose isomerase enzyme. The overall reaction

showing the conversion of corn starch to glucose followed by subsequent isomerization is represented by Equation 1.1 below:



The plant will produce 706 MM lb of HFCS annually that will be sent to the identified consumers 52% will be HFCS 55 and 48% will be HFCS 42. Some proteins, oils, and fibers which constitute corn gluten feed is also derived during this process and do not pose any health or safety threats. They will be separated out and converted to animal feed that can be used by local farmers. There is a high demand for HFCS in the North American food and beverage markets. For this reason, the plant will be located in Asbury, Iowa United States.

Our design basis can be found in Section 11.1 of this report. Our team will submit a final report that summarizes our project by May 4th, 2020.

1.2 Client Identification and Recognition of Need

The target companies for the final HFCS products are Pepperidge Farm in Pleasant Prairie, Wisconsin and Coca Cola Company in Chattanooga, Tennessee. These companies were selected as potential clients because HFCS is a vital raw material in the production of their respective products. In addition to this, the reputation built by these companies and their longevity as major suppliers of their respective products also guarantee the success of this project in the long run. Asbury, Iowa is an ideal plant location. It is at a reasonable distance from both major clients, and the infrastructures set in place allows for easy transportation so that the quality of the product is maintained before delivery to the aforementioned companies.

HFCS 55 is highly useful in carbonated drinks and juices because of its high sweetness level; it is also favored over sugar because of its cheapness (Hanover 728). Therefore, Coca Cola would use this specific grade for its carbonated drink products. HFCS 42 is mainly used for

canned fruits, sauces, processed foods, and condiments because of its mild sweetness and food preservative properties it offers (Hanover 728). Pepperidge Farm would use this specific grade for preservative and sweetness purposes, for products such as bread, cereal, and granola.

1.3 Project Goals and Objectives

The goal of this project was to design a chemical plant that produces glucose-fructose syrup that would cater to the supply and demand of the product within chemical industries. Our project goals are to develop all major equipment design specifications within our plant, conduct a plant economic analysis, give the final presentation, and create a final report. A Gantt Chart was created to show the project's schedule to achieve these goals as well as the project milestones.

1.4 Contemporary Issues and Impact of Project on Society

1.4.1 Contemporary Issues Relevant to Project

High Fructose Corn Syrup (HFCS) has become a popularly used sweetener for the food and beverage industry due to it being more cost-effective and easier to handle over refined cane sugar (Tou 1). However, it has gained a bad reputation within public health community over the concerns of rising rate in obesity, diabetes, and other cardiovascular diseases have been attributed to the consumption of foods containing HFCS (Tou 1). The negative news on HFCS is used as a proxy for the relation between HFCS consumption and obesity in the United States (Kennedy 4). Furthermore, the negative news on HFCS is found to have a negative and significant effect on the demand for soft drinks, which is the main driver for the demand of HFCS, but soft drink advertising has more than offset the effect of negative news produced (Kennedy 4).

1.4.2 Potential Impact of Project on Society

High Fructose Corn Syrup (HFCS) has become the main driver for alternative sweeteners for many products within the food and beverage industry. Sucrose was commonly used before

modified corn syrup in the food and beverage industry (Intelligence 1). The high and fluctuating sugar prices, along with the abundant use of HFCS in the beverage industry, has almost replaced sugar in the beverage industry (Intelligence 1). Our product HFCS 55 was strategic design to have the same relative sweetness as sucrose, so it can easily substitute sucrose in food and beverage industry (Intelligence 1). Same for the other grade HFCS 42 with a lower relative sweetness can easily replace sucrose (Intelligence 1). The plant is located in Asbury, Iowa which is not a very industrialized town. The establishment of this plant here could provide employment to the town locals while not posing a threat to the safety and well-being of these individuals.

1.5 Project Constraints

One of the constraints for this project involved the process choices in terms of available production routes, but the main constraint was choosing whether or not to produce the raw material on-site or purchase it especially since there are multiple raw materials that could be used to obtain starch (tapioca, potatoes, corn, beets, etc.). Choosing to purchase the raw material would require narrowing our source of starch to the cheapest source that can produce the amount required for our plant. Some important process conditions to consider for our plant were temperature and pH due to the sensitivity of the enzymes involved in the process. The design and operation temperatures for our equipment had to be feasible for both the functioning of the system and the environment. Stainless steel had to be used for a vast majority of the equipment in order to meet food-handling hygiene specifications and this limited the amount of optimization that could have been performed to reduce equipment cost. Also, additional equipment had to be added to the plant design such as ion exchangers, adsorption columns, and rotary filters to remove waste, mineral ions, and decolorization agents that would diminish the quality of our final product. Standard and codes for designing most equipment were based on the

American Institute of Chemical Engineering (AIChE) regulations. For pressure vessels, the standard and codes used were based on the American Society of Mechanical Engineers (ASME) regulations. Safety and Environmental regulations required for proper plant design had to be based on Occupational Safety and Health Administration (OSHA) and the Environmental Protection Agency (EPA) regulations.

2. PROJECT PLANNING AND TASK IDENTIFICATION

2.1 Task Identification and Timeline

The project was organized into four primary major tasks: Design Specifications, Economic Analysis, Final Presentation Preparation, and the Final Report. These major tasks and their respective subtasks are shown in Table 2.1, along with their timeline. Design specifications had a focus on major equipment sizing and preliminary costing, as well as preliminary optimization which was noted by the individual process efficiency task. On completion of those tasks, converging a simulation of our design could then be accomplished. The economic analysis task utilized the preliminary cost data in the design specification task and expanded on that costing data. It was scheduled to follow the design specification major task for that reason. This section of the project involved the capital cost associated with the major equipment, operating cost, and net present value (NPV) and rate of return on investment (ROROI). Other costing tasks are shown in Table 2.1. Final presentation preparation was the shortest task with about 11 days. Subtasks included an agreed upon finalized design through design optimization and presentation practice. The last major task was the Final Report. This section encompassed compiling data over the span of the whole project timeline and writing a rough draft of the final report.

Table 2.1 Project Task Summary

Task Name	Duration	Start	Finish
I. DESIGN SPECIFICATIONS	32 days	Thu 1/23/20	Fri 3/6/20
a. Unit Sizing and Preliminary Cost	27 days	Thu 1/23/20	Fri 2/28/20
i. Packed Bed Reactor			
ii. Liquid Chromatography Column/Ion Exchange			
iii. STR'S			
iv. Pumps			
v. Heat Exchangers			
b. Individual Process Efficiency	32 days	Thu 1/23/20	Fri 3/6/20
i. Energy Balance			
ii. Refine MEB			
iii. Overall Process Efficiency			
iv. Safety Parameters			
v. Control Unit for PBR			
c. Converge Aspen	30 days	Mon 3/2/20	Fri 4/10/20
i. Unit Specifications			
ii. Heat Exchangers			
iii. Final Simulated Design			
II. ECONOMIC ANALYSIS	36 days	Mon 2/10/20	Mon 3/30/20
i. Major Equipment Capital Cost			
ii. Estimated Operating Cost			
iii. Plant Logistics			
iv. Price for Utilities			
v. Price for Transportation and Waste management			
vi. Present Net Value and Investor ROI			
III. FINAL PRESENTATION PREPARATION	11 days	Mon 4/13/20	Sat 4/25/20
i. Design Optimization			
ii. Presentation Practice			
iii. Final Practice			
IV. FINAL REPORT	69 days	Thu 1/23/20	Tue 4/28/20
i. Compiling Information in Report			
ii. Revise Rough Draft			

2.2 Gantt Chart

The data in the Table 2.1 was compiled in the form of a project Gantt chart shown in Figure 2.1. Milestones were noted upon completion of the following tasks: Major Equipment Sizing, Individual Process Efficiency, Converge Aspen, Economic Analysis, Final Presentation Preparation, and the Final Report. Specified task allocation is also noted in Figure 2.1. ✓

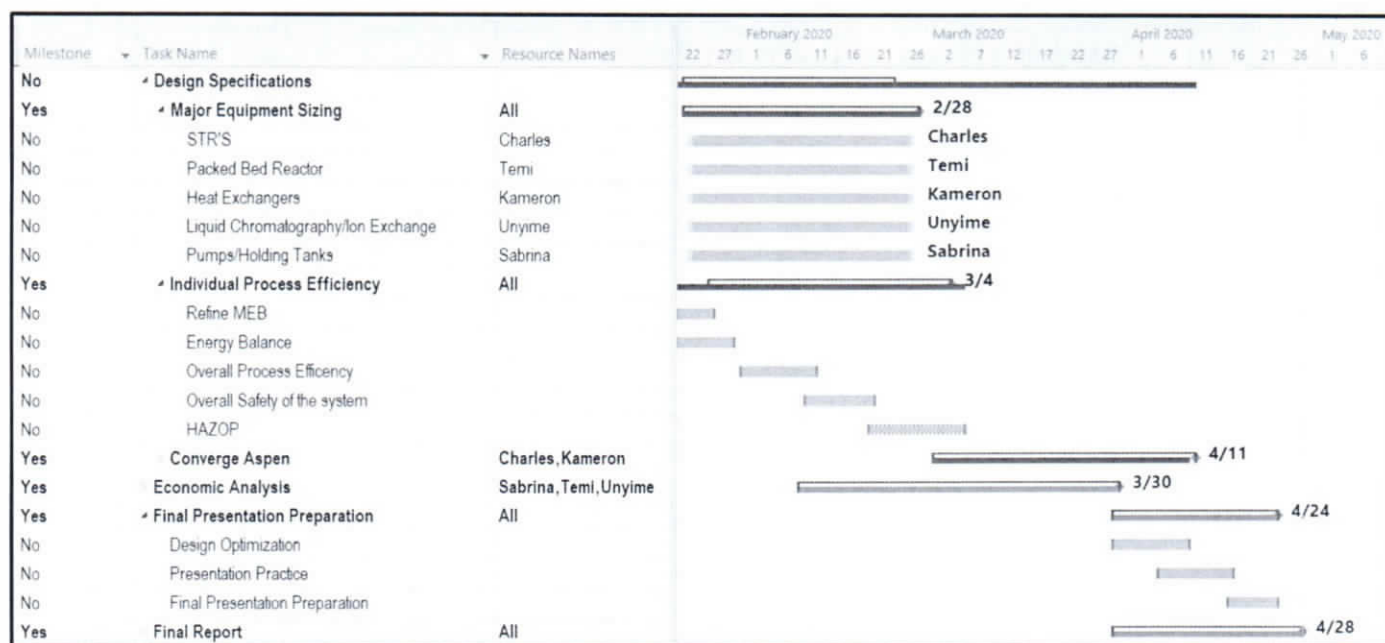


Figure 2.1 Gantt Chart

3. OVERVIEW OF THE MANUFACTURE OF GLUCOSE – FRUCTOSE SYRUP

High Fructose Corn Syrup is commonly produced by hydrolyzing starch to yield glucose (dextrose) syrup, and then converting glucose syrup to fructose syrup through isomerization. Using isolation and purification techniques, the concentrations of 42% and 55% fructose can be achieved in the final product syrup (Yuceer 783). Though there are deviations in sources of raw

material starch, the chemistry of producing high fructose syrups generally follows three major reaction steps: liquefaction, saccharification, and isomerization (Johnson 617).

3.1 Production Routes

Good overview

Glucose-Fructose Syrups can be made by either inversion of sucrose through hydrolysis, or the hydrolysis of starch that could be obtained from a variety of raw material sources such as corn, potatoes, cassava, and wheat (Whitehurst 210).

3.1.1 Starch Hydrolysis

Starch hydrolysis is the breakdown of bonds in the starch molecule using water. This process can be facilitated by the use of acids or enzymes (amylases) (Carr 2443). Acidic hydrolysis gives a lower yield and increased undesired by-products; therefore, enzymatic hydrolysis is more favorable in industrial use (Azmi 265). First, starch which is extracted from the raw material undergoes gelatinization where the starch is broken down with the addition of heat and water, forming a slurry (Johnson 618). Liquefaction describes the process of partial hydrolysis of the starch to yield oligosaccharides like maltose, as well as glucose (Lima 425). This step of hydrolysis is catalyzed using thermostable α -amylase enzyme and the resulting product generally has a dextrose equivalent amount of 8-12% (Whitehurst 206). In saccharification, glycosidic bonds of the starch molecule are broken down into simple sugars using glucoamylase enzyme as the catalyst (Converti 101). In the case of HFCS, a DE amount reached by saccharification is usually about 95-97% (Whitehurst 210). After hydrolysis, isomerization converts the glucose syrup to fructose. The reaction is catalyzed by glucose isomerase as the reaction enzyme (Barker 101).

3.1.2 Sucrose Inversion

Sucrose inversion is the hydrolysis of sucrose to an equimolar mixture of glucose and fructose (Martinez 22). This equimolar mixture is also known as invert sugar which is sent to an enrichment column where the fructose concentration is increased. Any remaining glucose syrup can be subject to isomerization until the desired fructose concentration is achieved. Invertase is the commercial enzyme used for sucrose inversion which is usually carried out in a fixed bed reactor with a strong cation resin (Gehlawat 28). Research showed into the use of immobilized invertase shows successful sucrose hydrolysis (Kilara 166) and suggested operational temperature for the enzyme in the reactor is 55 – 60°C.

3.2 Physical and Chemical Properties

Due to HFCS being a complex mixture, specific most chemical properties are not easy to characterize. Section 3.2.1 highlights some of its physical properties while Section 3.2.2 highlights some physical and chemical properties of glucose syrup (corn syrup).

3.2.1 Physical Properties of HFCS (OSHA 1996)

- For HFS-42 density (lb/ft³): 83.26 (liquid), 59.10 (solid)
- For HFS-55 density (lb/ft³): 85.73 (liquid), 65.98 (solid)
- Colorless, odorless liquid

3.2.2 Physical and Chemical Properties of Glucose Syrup (FDA, Title 21)

- Molecular Formula: C₆H₁₄O₇
- Molecular Weight: 198.17 g/mol
- Viscous, odorless, colorless liquid
- Density (g/mL): 1.4

- Freezing/Melting Point: 217-221 °F
- Decomposition Temperature: 217-221 °F
- Solubility: Freely Soluble
- Non-flammable
- Chemical Stability: Stable under normal conditions

are these really on the same range?

3.3 Handling, Storage and Safety

3.3.1 Handling and Storage

HFCS should be handled with care while properly observing proper industrial hygiene. Appropriate protective equipment should be worn and contact with the eye should be avoided (High Fructose Corn Syrup MSDS). HFCS should be stored at room temperature in a cool and dry place with proper ventilation. The container should be sealed tightly, and the product should be kept away from incompatible materials (NCBE).

Proper handling of enzymes consists of storage at low temperatures in dry environments. Both thermostable α -amylase and glucoamylase should be stored between 37-41°F, to retain their activity. In that temperature range, thermostable α -amylase can retain activity for about a year, and glucoamylase for about 6 months (Novozymes). Some enzymes may irritate the skin, eyes and, mucous membranes upon prolonged contact. These enzymes can create easily inhaled aerosols if vigorously mixed or stirred. If spilled the dried enzyme's residue can create dust. This can be a dust explosion hazard if not addressed properly. Also, any spilled enzyme should be flushed away with water to avoid any splashing (Novozymes).

3.3.2 Safety

3.3.2.1 EPA, OSHA Regulations and Environmental Impact

The Food and Drug Association (FDA) recognizes HFCS as a generally safe substance, however, some product requirements have been stipulated (FDA, Title 21). Such requirements include:

- Arsenic (as As) content no more than 1 mg/kg
 - Heavy metals (as Pb), not more than 5 mg/kg
 - Lead, not more than 0.1 mg/kg
 - Sulfur dioxide, not more than 0.003 percent
 - Color, within the range specified by the vendor which is primarily colorless
- very good!*

The Occupational Safety and Health Administration (OSHA) regulations on starch handling specify a dust accumulation limit of 1/8th of an inch, to lower risk of explosion (OSHA 1996). For this reason, areas nearby any bulk storage of starch must be properly cleaned when approaching this limit. OSHA also recommends limited exposure to the enzyme alpha-amylase used during liquefaction as it has been shown to cause chronic respiratory issues (Brisman 553). For the prevention of respiratory risk, OSHA mandates respirators to be clean and easily accessible following the CFR 1910.134 code (OSHA 1998). ✓

3.4 Uses

HFCS is usually named based on the composition of fructose present in the syrup (Lima 425). The three common forms produced industrially are HFCS 42, HFCS 55, and HFCS 90 (Yuceer 783). HFCS 90 is the major product obtained and is further processed to yield HFCS 42 and HFCS 55 which are used in the food and beverage industries as sweeteners (Hanover 729). HFCS 55 is widely used in the production of carbonated soft drinks while HFCS 42 which is not as sweet is generally used in the production of baked goods, noncarbonated beverages, and many

processed foods (Tou 3). When compared to sucrose, HFCS possesses qualities that are beneficial in the food industry. Firstly, it is relatively cheaper than sucrose (Gehlawat 30). Secondly, it has preservative properties because of its high acidity (Parker 72). Thirdly, it is more soluble and does not crystallize in solutions (Hanover 727). In addition to this, it helps to lower the freezing point to retain moisture in certain processed foods (Yuceer 784).

3.5 Product Specification

HFCS 42, HFCS 55, and HFCS 90 are the three grades that can be produced; however, HFCS 42 and HFCS 55 are the two products produced and sold by this plant. Both of our products must abide by FDA requirements previously stated (Schenck 18).

3.6 Raw Material Preparation

active, not passive voice.

Isolating starch within the corn is achieved by a process called wet milling. Corn being the supply of the starch is primarily delivered using rail and truck. The corn, delivered as shelled corn, is sent to receiving pits to be scaled and sampled and is then cleaned to remove foreign solids and dirt. After the corn is cleaned it is moved to a series of steeping tanks causing corn kernels to swell which loosens the contents of the corn by means of soaking the corn in a solution warm water and sulfur dioxide (Rausch 468). Steeped corn is then passed through degerminating mills where the kernel is torn apart exposing the germ, gluten, fiber, and starch of the corn. Liquid cyclones then are used to separate the germ from the gluten, fiber, and starch mixture. In a series of washers, grinders, and screens the fiber can be removed and eventually sold as animal feed. The remaining mixture of primarily starch, gluten, and water is sent to centrifuges to separate gluten and starch, based on varying densities (Ramírez). Lastly, the mixture of starch and water can be further washed and filtered to remove other organics in the

stream. This product slurry contains close to 35% starch and 65% water. This slurry can then be reacted and used to yield glucose syrup for the HFCS production (Hanover 725).

4. PROCESS SELECTION

4.1 Market Analysis and Trends

Four main companies control about 85% of the \$2.6 billion U.S. HFCS Industry (Parker 72). The four main companies are Archer Daniels Midland, Cargill, Staley Manufacturing Co, and CPC International. An analysis of the supply and demand in the high fructose corn syrup market was obtained based on results gathered by Dr. Lyn Kennedy of Louisiana State University and Dr. Garcia-Fuentes of Midwestern State University (Garcia-Fuentes). The study began ran from 1992 to 2013, and currently, the market supply and demand are decreasing every year (Garcia-Fuentes). For instance, from 2017 to 2018 the total U.S production of all HFCS fell from 8.2 to 8 million short tons a year. In parallel, deliveries of HFCS in the U.S decreased from 6.9 to 6.8 million short tons from 2017 to 2018. However, the average sale price is steadily increasing and the demand for HFCS is expected to be high for at 20 years (USDA). The analysis describes how events such as negative news and the soft drink market are major factors that affect the demand for HFCS. Figure 4.1 shows that the HFCS and soft drink markets almost mirror each other. This is due to the enormous use of high fructose corn syrup in the soft drink market. Around 75% of HFCS production goes into the soft drink market per the USDA and HFCS 55 is used more than HFCS 42. Therefore, it is recommended that more HFCS 55 should be produced from a designed plant.

✓ So then why
percentage Farm?

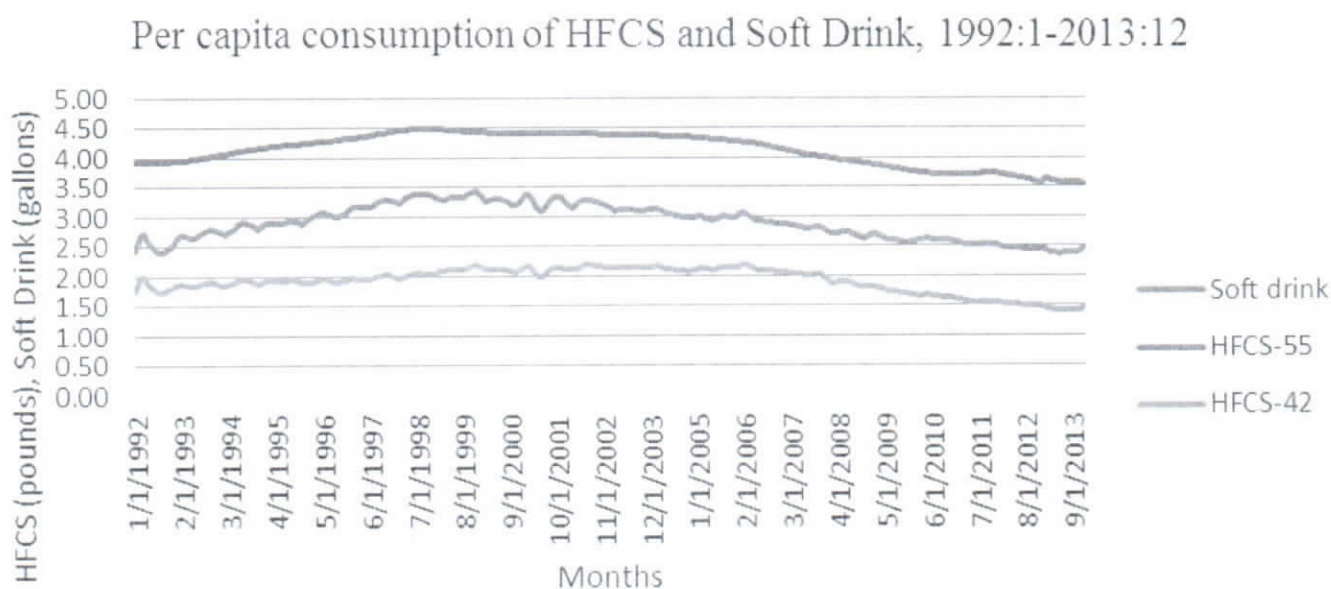
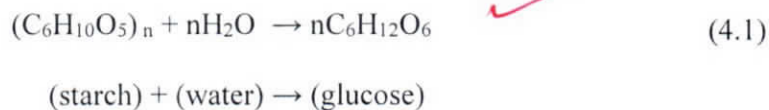


Figure 4.1 HFCS and Soft Drink Consumption Trend in the U.S

4.2 Reaction Routes to Producing Product

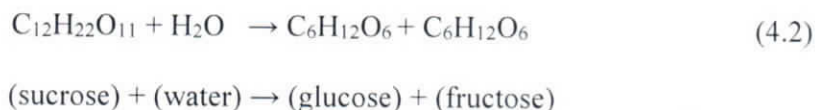
4.2.1 Enzymatic Starch Hydrolysis Followed by Glucose Isomerization

For starch hydrolysis, water molecules break down the polysaccharide into long glucose chains using the α -amylase enzyme. This long chain is further broken down into smaller glucose units using the glucoamylase enzyme. The glucose syrup obtained is purified before being isomerized to obtain some fructose. The immediate product of this reaction is HFCS 42 which when blended with HFCS 90 (obtained through HFCS 42 enrichment) gives HFCS 55. The overall reaction for the conversion of starch to glucose is shown in Equation 4.1 below where n is from 50 to thousands:



4.2.2 Enzymatic Sucrose Inversion Followed by Fructose Enrichment

With sucrose inversion, sucrose is broken down to a 50:50 glucose-fructose mixture using water and in the presence of the invertase enzyme. The fructose obtained in this resulting solution can further be enriched to give the desired concentration. The overall reaction for sucrose inversion to invert sugar is represented by Equation 4.2 below:



4.3 Process Screening using Theoretical Economic Indicator

Good!!

A theoretical economic indicator for the main reaction routes was performed to give a preliminary idea of the gross profit that could be attained depending on the reaction route selected. The results are shown below in Table 4.1 and Table 4.2

Sucrose Theoretical Economic Indicator

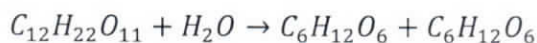


Table 4.1 Analysis for HFCS Using Sucrose

	Sucrose	Water	HFCS	
			HFCS-42	HFCS-55
lb/mole	1	1	1	1
MW	342	18	360	360
lb	342	18	360	360
lb/HFCS	.95	.05	1	1
¢/lb	59.6	1	35.21	41.23

$$\text{Gross Profit} = 35.21(1) - 59.6(.95) - 18(0) = -21.41 \text{ ¢/lb HFCS 42\%}$$

$$\text{Gross Profit} = 41.23(1) - 59.6(.95) - 18(0) = -15.39 \text{ ¢/lb HFCS 55\%}$$

Starch Theoretical Economic Indicator

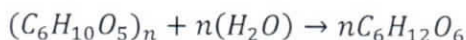


Table 4.2 Analysis for HFCS Using Starch

	Starch	Water	HFCS	
			HFCS-42	HFCS-55
lb/mole	2	2	1	1
MW	162	18	360	360
lb	324	36	360	360
lb/HFCS	.9	.1	1	1
¢/lb	6.26	0	35.21	41.23

$$\text{Gross Profit} = 35.21(1) - 6.26(.9) - 18(0) = 29.58 \text{ ¢/lb HFCS 42\%}$$

$$\text{Gross Profit} = 41.23(1) - 6.26(.9) - 18(0) = 35.60 \text{ ¢/lb HFCS 55\%}$$

4.4 Process Comparison and Selection

The HFCS industry is dominated by the starch hydrolysis route. Based on the results of the theoretical economic indicator, starch hydrolysis was the more viable reaction route for both HFCS 42 and HFCS 55 because the sucrose inversion route gave negative profit values. In selecting the raw material to be used for production, two factors played a major role. The first was accessibility and the second profitability of raw material in the U.S food and beverage market. In the U.S, our target market, the mid-west is host to several major producers of corn starch which is mostly used due to its abundance near the target market as well as its cheapness relative to alternative starch sources (Parker 74). The cost of corn starch is 6.26 cents per pound compared to sucrose at 59.6 cents per pound. Also, from the indicator, starch has the potential to produce profit for both grades of HFCS. Therefore, a hydrolysis process using starch as the raw material was the most suitable option and was selected for this project.

You only use economic considerations for process selection??

5. PROCESS DESCRIPTION

5.1 Block Flow Diagram

The block flow diagram for the production of HFCS 42 and HFCS 55 from corn starch is represented in Figure 5.1 below.

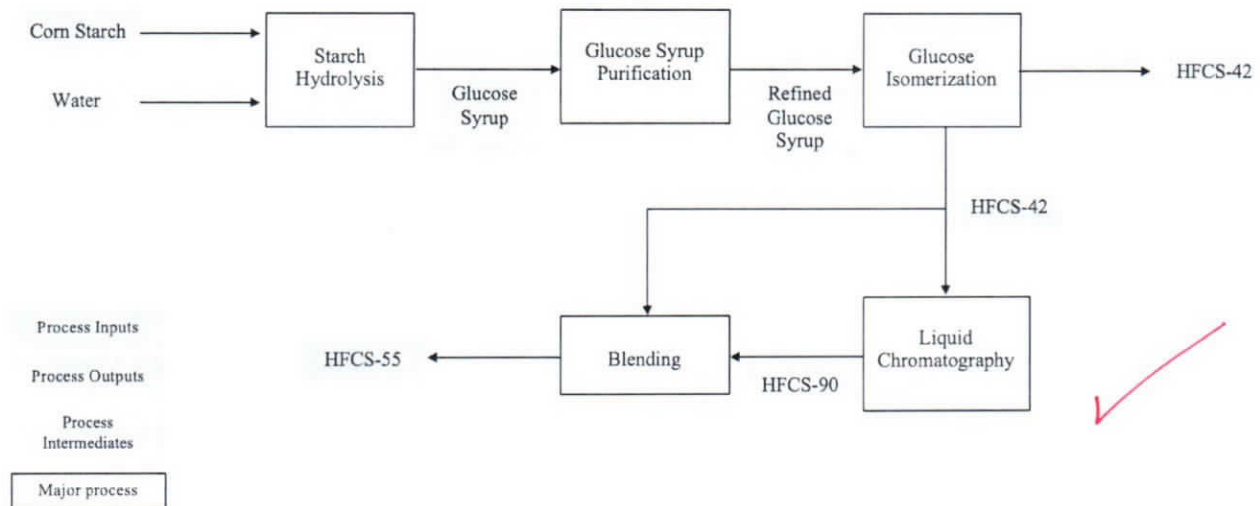


Figure 5.1 Block Flow Diagram

5.2 Process Description

The process involved in the production of high fructose corn syrup (HFCS) can be divided into three main processes: hydrolysis, purification, and isomerization. These three stages encompass the entire process and are comprised of various single units. A detailed description of the entire process is described below:

Hydrolysis

A feed of dry corn starch is mixed with water in a stirred tank to make starch slurry containing between 30% to 35% solid starch to allow the slurry to be easily pumped and prevent clogging in the pipes. A positive displacement pump is used to move this slurry because it has a high viscosity of about 1300 cP (List of Typical Viscosities). The starch slurry obtained from the tank is sent to a jet cooker which uses steam to gelatinize the slurry for 5-10 minutes (Derez).

The steam used in this process is culinary steam and is more hygienic than plant steam because it contains FDA approved boiler chemicals. After this, liquefaction is done by mixing α -amylase with the slurry for 2 hours to convert a high percentage of starch to a dextrin hydrolysate with a dextrose equivalent (DE) of 10% and a 50% w/w solid solution. The glucose slurry is then cooled in preparation for saccharification. Glucoamylase is fed into the next stirred tank reactor to complete the breakdown of starch into dextrose (glucose). Saccharification is the longest phase in the process with a duration of 48-96 hours (Carr 2443). The starch hydrolysis process is shown in Figure 5.2 below.

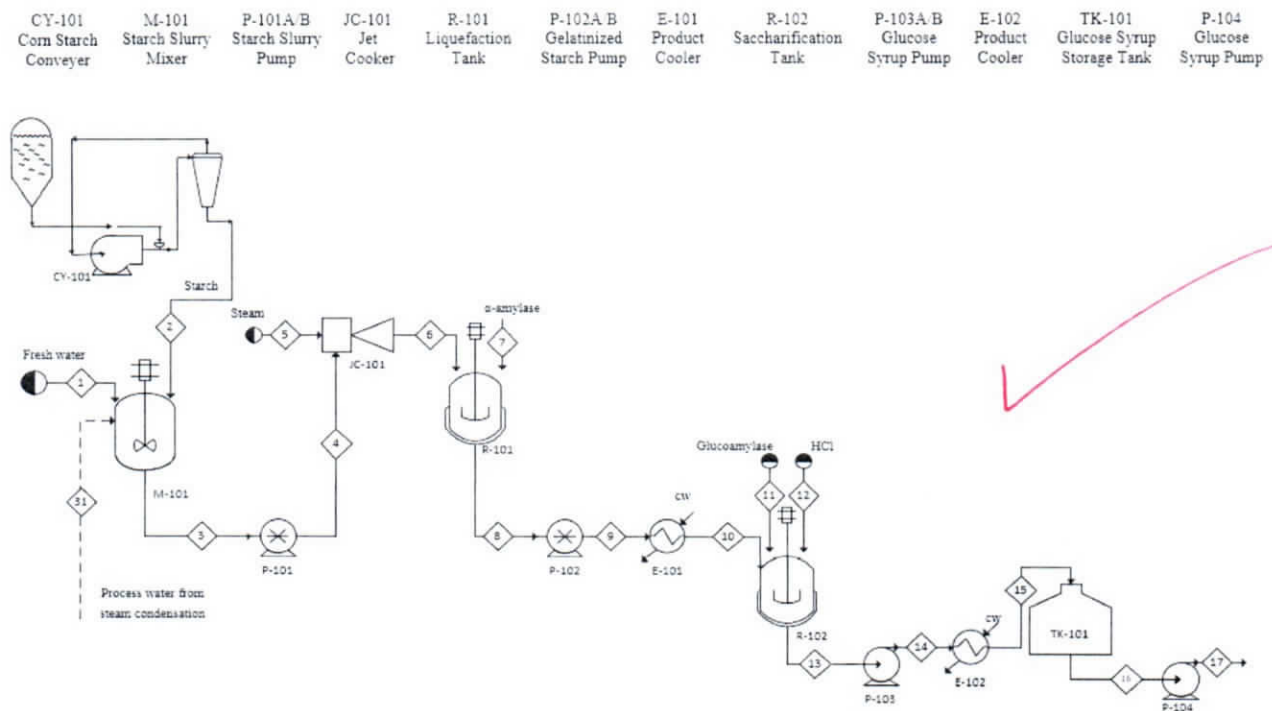


Figure 5.2 Process Flow Diagram for Starch Hydrolysis

Syrup Purification

The glucose syrup obtained is sent through a series of purification stages. First, it goes through a rotary filter to remove proteins, fibers, and lipids. Then it passes through a carbon filter to remove ash, a mixture of salts obtained from processed corn. Lastly, a 50:50 solid to liquid

product goes through an ion-exchange column to remove unwanted ions. After purification, the syrup is sent into a mixer where sodium carbonate (Na_2CO_3) and magnesium sulfate (MgSO_4) are added to increase the pH and ensure enzyme stability respectively in preparation for isomerization. The product from the mixer goes into an evaporator where steam is used to concentrate glucose syrup into an 80% w/w solution. The vapor that leaves the evaporator goes into a condenser and is condensed into liquid water that is recycled into the first stirred tank. The syrup purification process is shown in Figure 5.3 below.

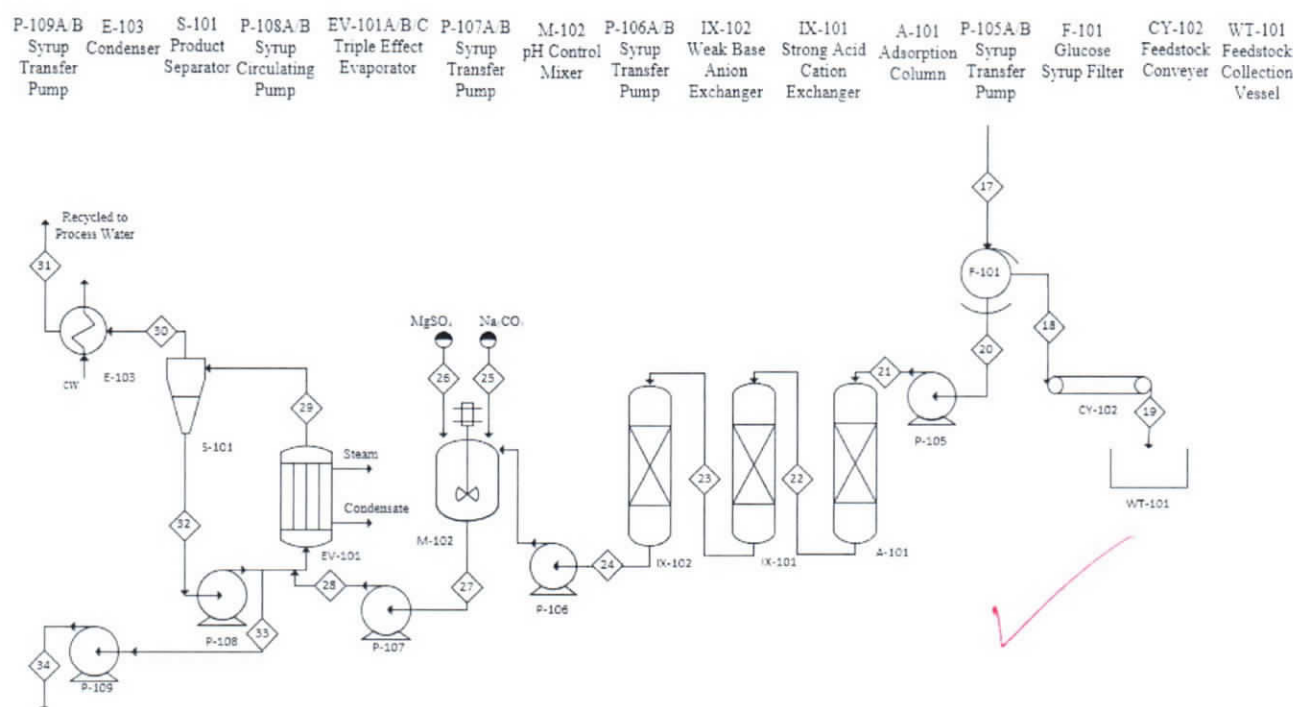


Figure 5.3 Process Flow Diagram for Syrup Purification

Isomerization and Fructose Enrichment

The purified liquid syrup leaves the evaporator at a temperature of 121°F and is pumped into a jacketed packed bed reactor (PBR); the purpose of the jacket is to supply heat and maintain a temperature of 140°F in the reactor for isomerization to take place. Isomerization

occurs for about 4 hours, as glucose isomerase converts the glucose syrup to HFCS 42 (Venkatasubramanian 146). The HFCS 42 product is split into two streams, one stream is sent into storage tanks while the other is passed through a chromatography column to concentrate it to HFCS 90. The two grades of HFCS are then mixed to give a final HFCS 55 product. The glucose isomerization and syrup enrichment process are shown in Figure 5.4 below.

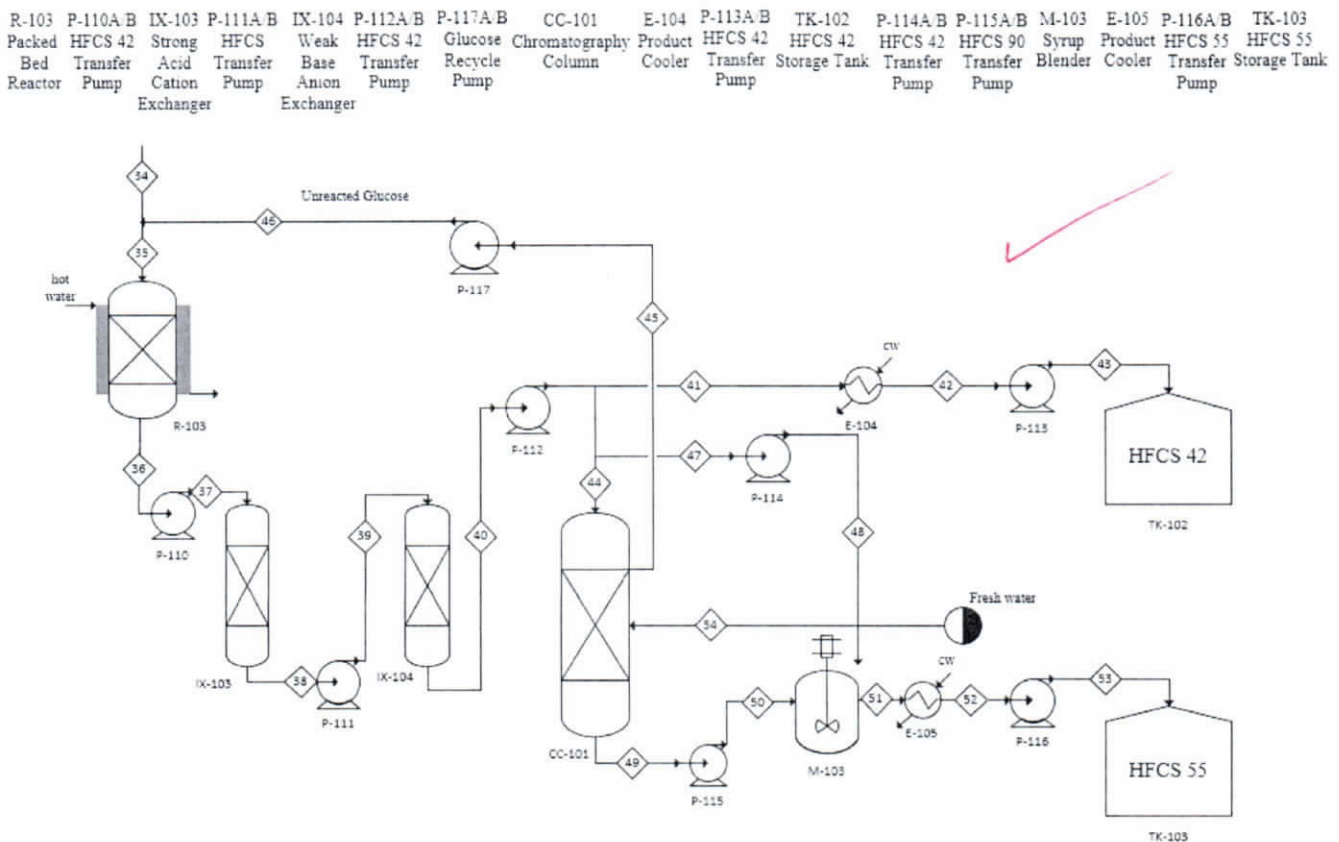
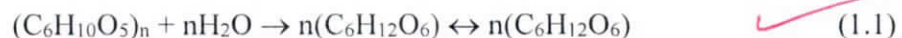


Figure 5.4 Process Flow Diagram for Fructose Enrichment

5.3 Process Chemistry

5.3.1 Reaction Stoichiometry

The overall reaction equation for the conversion of corn starch to HFCS is represented by Equation 1.1.



For the first reaction, the stoichiometric coefficients for H₂O and C₆H₁₂O₆ are equal and dependent on the number of glucose monomers that make up the starch chain to be hydrolyzed. For the second reaction, both C₆H₁₂O₆ isomers have equal stoichiometric coefficients (1:1).

5.3.2 Kinetics and Rate Equations

Hydrolysis Reactions

Hydrolysis is a reaction in which water reacts with a given substrate. For HFCS production the substrate used is corn starch. The reaction rate is highly dependent on the efficiency of the enzyme used. For that reason, two different catalysts are used to maximize conversion and reaction rate, α -amylase and glucoamylase, which splits hydrolysis into two major steps: Liquefaction and Saccharification.

Liquefaction

The process of liquefaction hydrolyzes swelled starch granules by cleaving α -1,4-glycosidic linkages to achieve the 8-12% DE product. The reaction is best characterized as exergonic, meaning that ΔG is negative, making the reaction spontaneous (Bismark 113). This biological reaction can be catalyzed with varying enzymes, classified as either endoenzymes or exoenzymes depending on how they were derived (Novozymes). This distinction is important as enzymes in hydrolysis can greatly affect the overall reaction kinetics of the system. This can be demonstrated by the substrate-enzyme mechanism shown in chemical Equations 5.1, 5.2 and 5.3.



This mechanism gives a reaction rate equation that relates the concentration of substrate (C_S), maximum reaction rate when the substrate is saturated with the enzyme (v_{max}), and the

concentration of substrate needed to half of the v_{max} value (k_m). This relationship is shown below in Equation 5.4 and was used to determine the reaction rate of our system.

$$r_p = \frac{v_{max}(C_S)}{k_m + C_S} \quad (5.4)$$

For HFCS hydrolysis, starch serves as the substrate in the system. The concentration was found based upon converting the calculated mass flow rate to the molar flow rate and dividing the value by the calculated volumetric flow. The concentration of starch was roughly 6.98 mol/L. The enzyme used for our process is α -amylase derived from *Bacillus licheniformis*, but for a more accurate k_m value a *Pyrococcus furiosus* α -amylase was used (Rollings 1481). The degree of deviation from reaction conditions of our system, such as optimum pH and temperature, warranted this change. The k_m value was approximately 0.52 M. Using that value, v_{max} could be solved for using a slope value (k_m/v_{max}) of a Lineweaver Burke plot of starch hydrolysis. The slope value was 2.58 and v_{max} was solved to be 0.2 M. With the values of v_{max} , k_m , and C_S , the reaction rate was solved to be 0.188 M/s.

Saccharification

To achieve higher concentrated amounts of glucose product (greater than 95% DE), the saccharification step of hydrolysis has to occur. This is necessary for HFCS production. Saccharification breaks down amylose and amylopectin chains through cleavage of both α -1,4 and α -1,6 bonds to yield dextrans and maltodextrans. This reaction step can be modeled as a pseudo 1st Order based upon the limited and disproportionate amount of water concentration active in the reaction (Butterworth 1295), compared to the concentration of broken-down starch substrate (amylose and amylopectin). The method for modeling kinetics for saccharification could not be achieved in the same way liquefaction was, due to varying data at our disposal. So,

the 1st Order Batch Reaction equation, shown in Equation 5.5, was used to relate reaction time (t), conversion (X), and k-constant values (k).

$$t = \frac{1}{k} \ln \left(\frac{1}{1-X} \right) \quad (5.5)$$

An optimization curve shown in Figure 5.5 was used to obtain the reaction time and conversion values to be used in Equation 5.5. Using the same figure (Figure 5.5) enzyme dosage amount, chosen for our system, was 0.5/1000 kg of DS to minimize reaction time but also maximize the conversion of glucose.

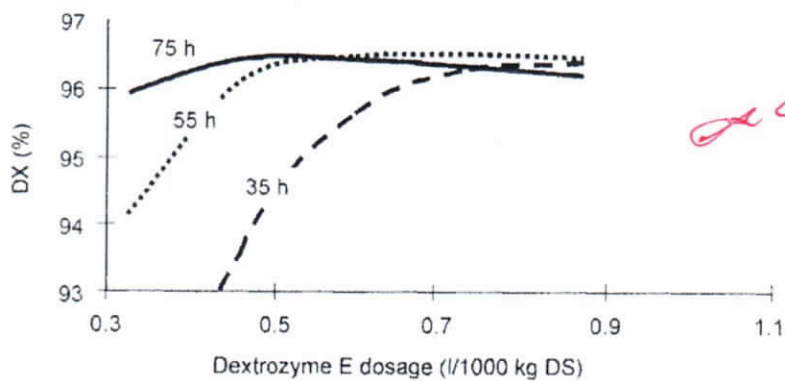


Figure 5.5 Conversion to Dextrose (%) vs Enzyme Dosage (Glucoamylase)

To solve for the k-constant value the reaction times 35, 55, and 75 hours with their respective conversion values of 94.3, 96.2, and 96.3% were used. Those values were chosen due to correspondence with our selected dosage amount. Using Equation 5.5, k-constant values could be calculated, and the results of their change with respect to time can be seen in Figure 5.6. ✓

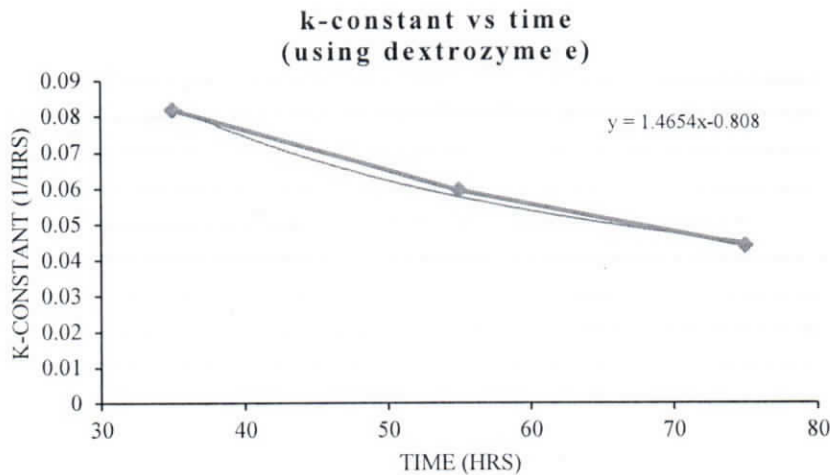


Figure 5.6 k-constant values vs reaction times at (0.5/1000kg of DS)

The most optimum reaction time given our dosage was 55 hours, and the respective k-constant value is 0.06M/hr. Because the behavior of hydrolysis is pseudo 1st order the reaction rate equation can also be depicted as 1st order, as shown in Equation 5.6.

$$-r_A = kC_A \quad (5.6)$$

Solving for reaction rate could now be achieved given the known k-constant value and concentration of substrate (A). Upon calculating with the assumption of 10% loss (Kodoma 988) in mass from starch substrate, the reaction rate was about 0.342M/s.

5.4 Product to Storage

Stainless steel tanks are the recommended storage tanks used to contain the final products of HFCS 42 and HFCS 55 before they are transported to their respective consumers. They are sized to be large enough to store 3-6 days' worth of syrup and the temperature in the tanks should not exceed 118°F (National Sugar Marketing).

5.5 Material Balances

5.5.1 Basis of Calculations

All calculations were made using English units of pound per hour as the flowrate for each stream. A typical HFCS plant capacity of 1000 metric tons per day was used as the basis for determining the output streams of HFCS 42 and HFCS 55. 46,000 lbs/hr of HFCS 55 and 43,000 lbs/hr of HFCS 42 were used for calculation and determined using the assumptions presented in Section 10.1. The sugar compositions of the product streams were to emulate that of commercial-grade HFCS; hence, they were known. Given these values, calculations were made in a reverse manner until the initial process stream values were obtained.

*why did you use this basis?
what about eq. plant sizing??*

5.5.2 Unit Descriptions and Material Balance

Calculations for the material balance for components in each stream were made for all the units and are shown below with a description of each unit. All the major equipment that are in direct contact with the syrup is made of high-grade stainless steel. In the food processing industry, stainless steel is the most important type of steel used in direct contact with food because of its ductility and high resistance to corrosion. (Saravacos 57).

M-101

This is a vertical mixing tank with a volume of 758 ft³ and a residence time of 10 minutes. Starch slurry is made in the tank by mixing corn starch at 77°F with water. The starch is mixed for several hours with internal agitators until a thick slurry is attained. The pH of the slurry made is 6.5. The stirred tank is made of Stainless Steel 316 and has a bare cost module of \$38,225. See page 49 for the unit specification sheet.

P-101

This is a carbon steel progressive cavity pump. It is a type of positive displacement pump that is used for high viscosity fluids. It moves 10,302 gallons of glucose slurry per hour, generating 8 ft

of pump head. Despite its high cost of maintenance, the advantage of using this pump is that the same volume of slurry is pumped at a constant rate and this prevents clogging in pipes. It consumes 1.17 kW electrical power and has a bare module cost of \$29,600. See page 50 for the unit specification sheet.

JC-101

This is a horizontal stainless-steel jet cooker that provides steam at 105°F needed for gelatinization by steam injection. Gelatinization converts the starch to thick gel and prepares it for liquefaction. Culinary steam is introduced to the jet cooker at a pressure between 10-12 bar. Steam is a useful heating utility and is favored above other process heating methods because the temperature at which the heat is released can be controlled by the pressure of the steam. The starch solution leaving the jet cooker contains 20% w/w solids. The unit has a bare module cost of \$1,816. See page 51 for the unit specification sheet.

R-101

This is a stainless steel 316 stirred tank reactor for the liquefaction of starch. Gelatinization of the starch occurs for 5-10 mins at high temperatures. A 20% w/w solid solution enters the STR at a pH of 6.5 and α -amylase containing Ca^{2+} (with a concentration between 20-150 ppm) is fed to the slurry to break down the starch molecules into large glucose chains. Seven reactors are operated in series each with a volume of 1069 ft³. Flat blade agitators are used for the continuous mixing at an operating pressure of 14 psi. The unit has a bare module cost of \$1,575,000. See page 52 for the unit specification sheet.

P-102

This is also a carbon steel progressive cavity pump. The glucose slurry still has a high viscosity of about 600cP. The pump head generated is 8ft and the amount of power consumed is 1.17kW. The unit has a bare module cost of \$29,601. See page 52 for the unit specification sheet.

E-101

This is a shell and tube heat exchanger responsible for cooling the glucose slurry from 221°F to 140°F in preparation for saccharification. It has a heat transfer area of 381 ft² and the material of construction is stainless steel. A heat duty of 6.0e6 is required and it has a heat transfer coefficient of 150 Btu/hr-ft²-°f. The unit has a bare module cost of \$150,093. See page 54 for the unit specification sheet.

R-102

R-102 is a stainless-steel stirred tank reactor. 29 reactors are arranged in series each with a volume of 7,063 ft³. For saccharification to occur the pH must be at 4.5, HCl is added to decrease the pH. This unit involves a feedforward control process that measures the pH of the incoming slurry. A pH sensor is used on the inlet side of the CSTR to signal chemical pumps to feed hydrochloric acid (HCl) into the reactor to maintain a pH level of 4.5. Glucoamylase is added to break the glucose chain into monomers; the glucoamylase used is made from *Aspergillus niger*. The glucose syrup leaving the reactor has a DE of 95% and leaves as a 50% w/w solid solution. Saccharification completes the breakdown of starch into dextrose and it occurs for a period of 24 to 72 hours. The bare cost module for this unit is \$34,800,800. See page 55 for the unit specification sheet.

P-103

This is a carbon steel centrifugal pump used to move glucose syrup into the storage tank. It moves 9,246 gallons of glucose syrup per hour and consumes 2kW of electrical power. It has a

bare module cost of \$13,475. The viscosity of the glucose syrup decreases after saccharification and the pumps used to move the glucose syrup after this stage are regular centrifugal pumps. The unit specification section gives all the details of the other pumps used in this process. See page 56 for the unit specification sheet.

E-102

This is a stainless-steel shell and tube heat exchanger for cooling glucose syrup to 77°F for storage in the tank. The syrup must be cooled to this temperature to avoid microbial growth in the tank. See page 57 for the unit specification sheet.

TK-101

This is a stainless storage tank that stores 3 days' worth of liquid glucose syrup. Two storage tanks will hold 9,245 gallons each of glucose syrup at 14 psi. The purpose of this tank is to hold the syrup that is being made in the second STR during saccharification. Since this process takes about 72 hours the glucose syrup must be stored until the saccharification process is completed. The stored syrup is what is pumped into the next unit to begin continuous production while another batch of glucose syrup is being made in the second STR. The bare cost module for the unit is \$2,310,540. See page 58 for the unit specification sheet.

F-101

This is a 304L stainless steel rotary vacuum filter with a precoat discharge. It is a separation device that removes waste products from the syrup through a vacuum mechanism. The filter has a capacity of 200 lbs/hr.ft² and a filtration surface area of 729 ft². The waste particles are scraped off with a knife attached to the filter. About 14578 lbs of the filter cake are separated per hour. The filter cake obtained is known as corn gluten feed and it would be sold to local farmers as

feedstock. The unit has a bare module cost of \$908,107. See page 60 for the unit specification sheet.

A-101

This is an adsorption column made of stainless steel and is used in the decolorization step of the syrup purification process. The function of this unit is the removal of color-forming components like hydroxymethylfurfural (HMF) and some ions. The unit is packed with DOWEX OPTIPORE SD-2 adsorbent which is an effective alternative to traditional activated carbon systems. For this process, it is assumed that all the ash is removed in this unit. Two adsorption columns are used for this unit each with a volume of 390 ft³, a height of 10.86 ft, and a diameter of 6.76 ft. The bare model cost for this unit is \$187,450. See page 62 for the unit specification sheet.

IX-101

This is a strong acid cation exchanger made of stainless steel and is used in the deashing step of the syrup purification process. The unit is packed with DOWEX Monoshpere 88 resins which have applications in HFCS purification. The function of the cation resins is the removal of mineral ions such as Na⁺, K⁺, Fe²⁺, Mg²⁺, and Ca²⁺ from the glucose syrup. This unit is comprised of two cation exchangers each operated in series with an anion exchanger. The vessel volume is 113 ft³. The bare module cost for both exchangers is \$102,652. See page 63 for the unit specification sheet.

IX-102

This is a weak base anion exchanger made of stainless steel and is also used in the deashing step of the syrup purification process. The unit is packed with DOWEX Monoshpere 88 resins which have applications in HFCS purification. The function of the cation resins is the removal of mineral ions such as NO³⁻, SO₄²⁻, and Cl⁻ from the glucose syrup. Two anion exchangers are also

used for this unit; each anion exchanger is arranged directly after a cation exchanger in series. The vessel volume calculated for the first and second exchangers is 935 ft³. Both anion exchangers have a bare module cost of \$379,368. See page 64 for the unit specification sheet.

M-102

M-102 is a vertical, stainless-steel mixing tank with a volume of 186 ft³ and a residence time of 10 minutes. The tank is continuously stirred by internal agitators and uses 0.04kW of electricity. A pH sensor is also used at the inlet of the mixer to signal pumps to feed Na₂CO₃ into the mixer to maintain the pH at 7.5. MgSO₄ is added to ensure enzyme stability of the glucose isomerase used in the isomerization unit. The unit has a bare module cost of \$144,470. See page 66 for the unit specification sheet.

EV-101

This is a stainless-steel forced circulation evaporator with a residence time of 25 minutes. It is operated as a forward-feed triple effect evaporator with an average area of 1421 ft² for each effect. Steam enters the evaporator at 248°F to vaporize water from the glucose syrup. The glucose syrup leaves the evaporator at a temperature of 121°F. The average heat duty required for concentrating the syrup in each effect is 1.6e7 Btu/hr with a steam economy of 2.25. The bare module cost for the unit is \$6,690,000. See page 68 for the unit specification sheet.

E-103

This is a carbon steel counter-current barometric condenser with a heat load of 2.0e7. The water vapor enters at 111°F to be condensed into a liquid. The water obtained is at room temperature and is recycled into the first stirred tank. 540 pounds of water is consumed for every pound of vapor that is condensed. The bare cost module for the unit is \$120,624. See page 69 for the unit specification sheet.

R-103

This is a jacketed packed bed reactor (PBR) for isomerization. This unit has 14 reactors running in parallel each with a volume of 477 ft³. Glucose isomerase enzyme is used to convert glucose syrup to HFCS 42. The molarity of Mg²⁺ needed for enzyme activation is 0.0004 M. The HFCS 42 product leaving this unit has a 75:25 solid to liquid weight percentage. The orientation of the reactor is vertical, and the material of construction is Stainless Steel 316. The pressure drop across the reactor is 2 psi and the heat duty required for heating is 34639 Btu/hr. The bare cost module for the unit is \$5,149,200. See page 71 for the unit specification sheet.

IX-103

This is a strong acid cation exchanger made of stainless steel and is used in the deashing step of the syrup purification process. The unit is packed with DOWEX Monoshpere 88 resins which have applications in HFCS purification. The function of the cation resins is the removal of Na⁺, Mg²⁺, and Ca²⁺ from the HFCS 42. This unit is comprised of two cation exchangers each operated in series with an anion exchanger. The vessel volume is 56 ft³. The bare module cost for both exchangers is \$81,600. See page 73 for the unit specification sheet.

IX-104

This is a weak base anion exchanger made of stainless steel and is also used in the deashing step of the syrup purification process. The unit is packed with DOWEX Monoshpere 77 resins which have applications in HFCS purification. The function of the cation resins is the removal of mineral ions such as CO₃²⁻ and SO₄²⁻ from the HFCS 42. Two anion exchangers are also used for this unit; each anion exchanger is arranged directly after a cation exchanger in series. The vessel volume calculated for the first and second exchangers is 236 ft³. Both anion exchangers have a bare module cost of \$158,100. See page 74 for the unit specification sheet.

Split point

This point is before the chromatography column. Half of the HFCS 42 produced goes straight to storage and the other half of the HFCS 42 goes through this point and is further split into the chromatography column and a mixer.

CC-101

This is a stainless-steel chromatography column that increases the fructose concentration in the syrup. 18.6% of the HFCS 42 goes into the chromatography column and is enriched to HFCS 90 before it is sent into the mixer. The unit is packed with DOWEX Monoshpere 99 resins in the Ca^{2+} form which has applications in chromatographic separation of HFCS. This unit is designed as a simulated moving bed and has twelve fixed bed chromatography columns in series with two inlet and two outlet streams. The length of the column is 9 ft with an inner diameter of 3 ft. The unit has a bare module cost of \$380,800. See page 77 for the unit specification sheet.

M-103

This is a stainless- steel mixer where HFCS 42 and HFCS 90 are mixed to give HFCS 55. The amount of HFCS 42 coming from the split point into the mixer is 81.4%. The final HFCS 55 product sent into the storage vessels has a 75:25 solid to liquid weight percentage. The volume of the mixer is 36 ft³ and it has a residence time of 10 minutes. Internal agitators are used for continuous stirring and they use 0.32 kW of electricity. The bare cost module for this unit is \$101,907. See page 82 for the unit specification sheet.

E-104 and E-105

These are shell and tube heat exchangers that cool HFCS 42 and HFCS 55 respectively before they are stored in tanks. They are made of stainless steel and cool down the syrup to 77°F at 14.7

psi. A heat duty of 1.7×10^6 is required and they both have a heat transfer coefficient of 150 Btu/hr-ft²-°F. The unit has a bare module cost of \$132,664. See pages 83 and 84 for the unit specification sheet.

TK-102 and TK-103

These are storage tanks that hold 3 days' worth of HFCS products. A total of 29 storage tanks are used to hold HFCS 42 while 30 storage tanks are used to hold HFCS 55. The bare module cost for each tank unit is \$2,320,00 and \$2,400,000 respectively. See pages 85 and 86 for the unit specification sheet.

5.5.3 Degrees of Freedom Analysis

To ascertain that stream calculations could be made, a degree of freedom analysis (DOF) was performed for each unit. These calculations for the major units in each section are shown below. If the DOF was greater than 0, specifications discussed in Section 10.1 were made.

Table 5.1 Degrees of Freedom Analysis for Hydrolysis Step

	Mixer 1		Jet Cooker		STR 1		STR 2	
No. of Variables								
Stream Variables	7		7		7		8	
Chemical Reactions	0		0		1		1	
No. of Equations								
Flows		0		0		0		0
Stream Composition		1		1		1		0
System Performance		0		1		0		0
Material Balances		3		3		5		4
Sum	7	4	7	5	8	6	9	4
DOF	3		2		2		5	

Table 5.2 Degrees of Freedom Analysis for Purification Step

	Rotary Filter		Adsorption Column		Ion Exchangers	
No. of Variables						
Stream Variables	8		6		6	
Chemical Reactions	0		0		0	
No. of Equations						
Flows		0		0		0
Stream Composition		1		0		0
System Performance		1		0		0
Material Balances		4		3		3
Sum	8	6	6	3	6	3
DOF	2		3		3	

Table 5.3 Degrees of Freedom Analysis for Isomerization and Enrichment Step

	PBR		Chromatography Column		Mixer 3	
No. of Variables						
Stream Variables	7		11		12	
Chemical Reactions	1		0		0	
No. of Equations						
Flows		0		0		1
Stream Composition		1		3		3
System Performance		1		1		0
Material Balances		4		4		4
Sum	8	6	11	8	12	8
DOF	2		3		4	

You need a nomenclature sheet to define variables.

5.6 Energy Balances

The heat duty for the coolers used during this process was calculated using Equation 5.7 below because no phase change occurred. In the absence of information on the specific heat capacity of HFCS, the value used for calculation was for corn syrup of 90 DE and 19.8 w/w% solids for E-101 and 79.4 for E-102, E-103, and E-104 as the closest specifications to the glucose syrup produced by this plant design. These values were selected closest to room temperature and were 0.889 Btu/lb°F and 0.583 Btu/lb°F (Critical Data Tables).

$$Q = \dot{m} * C_p * \Delta T \quad (5.7)$$

These values for E-101, E-102, and E-103 came out to be 6.4e6, 2.0e6, and 1.6e6 Btu/hr respectively.

The heat duty for the PBR was obtained using the overall heat transfer equation by combined conduction and convection in Equation 5.8. The heat duty value obtained was 34639 Btu/hr. Lastly, the heat duty for the evaporator and condenser were obtained using Equation 5.9 and 5.10 below, and the values obtained were 1.6e7 Btu/hr and 2.0e7 Btu/hr respectively.

$$Q = \frac{T_1 - T_4}{\frac{1}{h_i A_i} + \frac{r_o - r_i}{K_A A_{ALM}} + \frac{1}{h_o A_o}} \quad (5.8)$$

$$Q = UA\Delta T \quad (5.9)$$

$$Q = \dot{m}\lambda \quad (5.10)$$

5.7 Overall Material Balance with Process Streams

Material balances over each unit in the process were performed and is represented in Table 5.4 below. The component mass flow rates, as well as the stream temperature and pressure, are

represented as well as the total stream amount and this value is in units of pounds per hour (lbs/hr).

Table 5.4 Material Balance by Process Streams

Stream Number	1	2	3	4	5	6	8	10	13
Temperature (°F)	77	77	77	77	221	203	221	140	140
Pressure (psi)	14.7	14.7	14.7	10.7	17.5	14.7	14.7	10.7	10.7
Component Mass Flow (lbs/hr)									
Starch	0	25,608	25,608	25,608	0	38,485	0	0	0
Water	18,700	3,201	71,102	71,102	48,778	106,855	58,311	58,311	58,310
Waste Feed	0	291	950	950	0	438	14,578	14,578	14,578
Oligosaccharides	0	0	0	0	0	0	65,600	65,600	3,645
Glucose	0	0	0	0	0	0	7,289	7,289	69,245
Fructose	0	0	0	0	0	0	0	0	0
TOTAL	18,700	29,100	97,660	97,660	48,778	145,778	145,778	145,778	145,778

Stream Number	14	15	17	18	20	21	22	24	27
Temperature (°F)	140	77	86	77	86	86	86	86	77
Pressure (psi)	14.7	14.7	14.7	14.7	14.7	6	6	6	14.7
Component Mass Flow (lbs/hr)									
Starch	0	0	0	0	0	0	0	0	0
Water	58,310	58,310	58,310	0	65,600	65,600	65,600	65,600	65,600
Waste Feed	14,578	14,578	14,578	14,578	0	0	0	0	0
Oligosaccharides	3,645	3,645	3,645	0	3,280	3,280	3,280	3,280	3,280
Glucose	69,245	69,245	69,245	0	62,320	62,320	62,320	62,320	62,320
Fructose	0	0	0	0	0	0	0	0	0
TOTAL	145,778	145,778	145,778	14,578	131,200	131,200	131,200	131,200	131,200

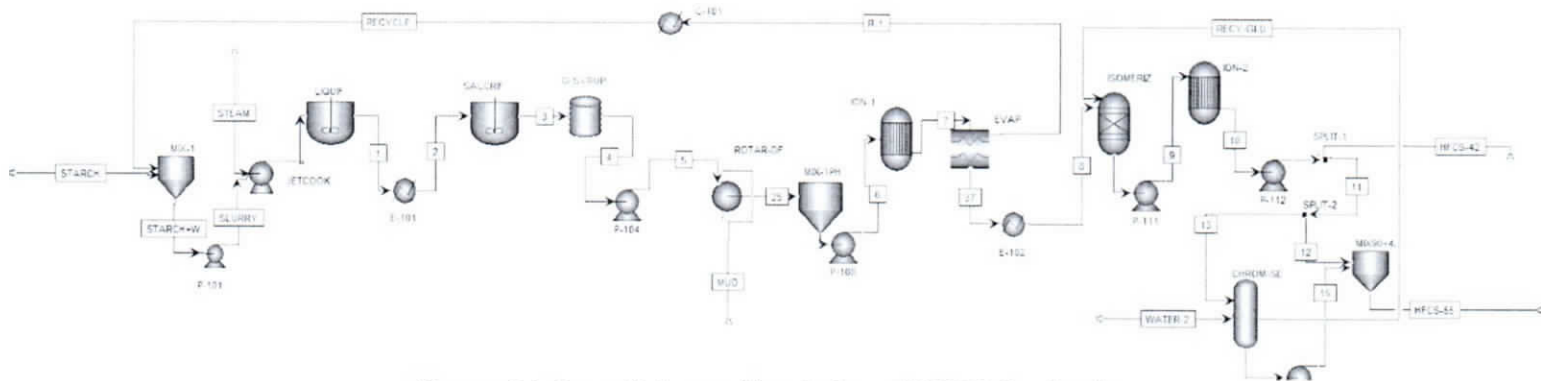
Stream Number	28	30	31	32	34	35	36	38	40
Temperature (°F)	77	111	77	121	121	121	140	140	140
Pressure (psi)	6	0.1	14.7	14.7	59	59	57	24	20
Component Mass Flow (lbs/hr)									
Starch	0	0	0	0	0	0	0	0	0
Water	65,600	49,200	49,200	16400	16400	17,400	21,500	21,500	21,500
Waste Feed	0	0	0	0	0	0	0	0	0
Oligosaccharides	3,280	0	0	3280	3280	3,300	1,900	1,900	1,900
Glucose	62,320	0	0	62320	62320	65,300	35,500	35,500	35,500
Fructose	0	0	0	0	0	0	27,100	27,100	27,100
TOTAL	131,200	49,200	49,200	82,000	82,000	86,000	86,000	86,000	86,000

Stream Number	42	43	44	45	47	49	52	53	54
Temperature (°F)	81	81	140	140	140	140	81	81	77
Pressure (psi)	14.7	14.7	15.1	37	15.1	13	17.7	14.7	14.7
Component Mass Flow (lbs/hr)									
Starch	0	0	0	0	0	0	0	0	0
Water	10,750	10,750	2,000	1,010	8,740	1,890	11,500	11,500	7,030
Waste Feed	0	0	0	0	0	0	0	0	0
Oligosaccharides	967	967	180	0	790	640	1,000	1,000	0
Glucose	17,738	17,738	3,300	3,020	14,430	270	14,500	14,500	0
Fructose	13,545	13,545	2,520	0	11,040	8,200	19,000	19,000	0
TOTAL	43,000	43,000	8,000	4,030	35,000	11,000	46,000	46,000	7,030

6. ASPEN SIMULATION OF PROCESS

This is for a bio process. Good job.

Our plant design of HFCS was simulated using Aspen V9 Plus and our system converged with the truncated model, as shown in Figure 6.2. The property package used for our system was the UNIQUAC model. This determination was set through consideration of the HFCS process being primarily liquid phase and polar. The original design, shown in Figure 6.1, encompassed both major reactions, the hydrolysis of starch slurry, and the isomerization of glucose syrup. The starch feed could be substituted with cellulose as both are composed of glucose subunits (monomers). Recycle from the evaporator (EVAP) would be sent to the recycle stream that would originally be fed to the first mixer. With this configuration, there were notable unspecified values, particularly in the liquefaction and saccharification reactors. This is a result of limited kinetic data for this exergonic reaction, namely the kinetic factor information such as activation energy and k-constant values. For that reason, the truncated model shown in Figure 6.2, was made.



compensate for excess water in the loop. The isomerization reaction which is typically held in a packed bed reactor was to be initially modeled using a stoichiometric reactor with an equilibrium reaction scheme. Though for simplicity a yield reactor was used with an assumption of glucose yielding 42% fructose. The yield amount is described by the equilibrium reaction of glucose to fructose. There should be 50% of glucose converted with 100% reactor efficiency, but given the PBR for this reactor realistically has an efficiency of 85%, only 42.5 % can be converted.

Subsequent splitters, Split-1 and Split-2, have split fractions of 0.5. For liquid chromatography, a basic separator was used with a specification of the primary outlet stream being 90% fructose.

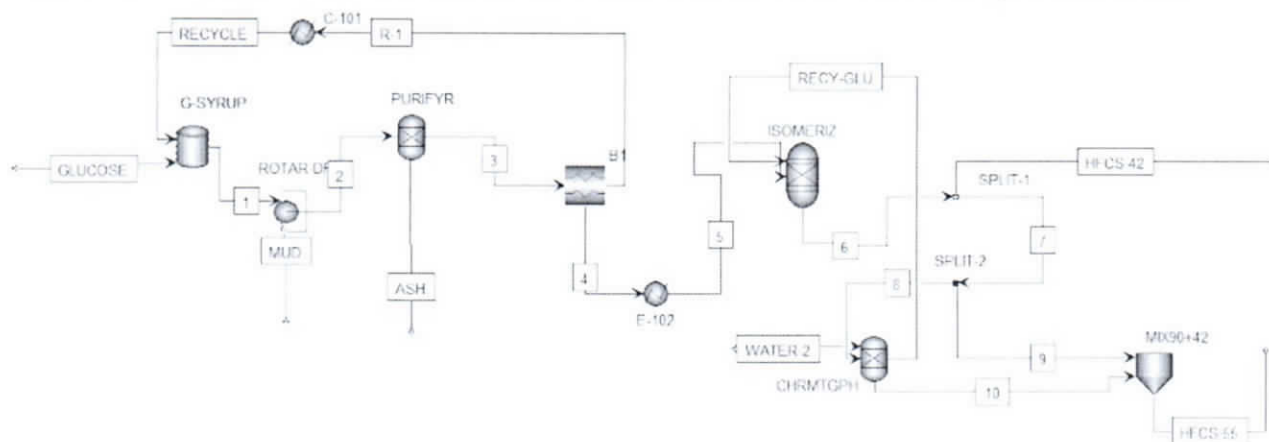


Figure 6.2 Truncated Aspen Simulation of HFCS Production

Our results for both products from the ASPEN simulation proved to be fairly close to our theoretical MEB calculations as shown in by data in Table 6.1 below.

Table 6.1 Simulation Results Comparison

	HFCS 42		HFCS 55	
	Simulated Mass Flow (lb/hr)	Theoretical Mass Flow (lb/hr)	Simulated Mass Flow (lb/hr)	Theoretical Mass Flow (lb/hr)
Fructose	13,738	13,545	19,555	19,000
Glucose	18,263	17,738	16,005	14,500
Water	10,733	10,750	10,741	11,500
TOTAL	42,734*	43,000*	46,302*	46,000*

*Reflects total including oligosaccharide component not shown on table

7. CHEMICAL ENGINEERING DESIGN

7.1 Design of Major Equipment

R-101 and R-102

The first step in sizing the stirred tank reactor (STR) was determining the residence time. For the first hydrolysis step (liquefaction), a residence time of 2 hours was used as the basis of calculation (Buchholz 406). For saccharification, a residence time of 55 hours obtained from Figure 5.5 was used because it is in the range of residence time values given by Buchholz on page 407. *Alpha*-amylase Termamyl and Dextrozyme are the enzymes used for liquefaction and saccharification respectively (Whitehurst 215). After the residence time had been obtained, the total volume needed for liquefaction and saccharification was calculated using the relationship between residence time and volumetric flow rate as shown in Equation 7.1 below.

$$\text{Residence time } (\tau) = \frac{\text{volume } (v)}{\text{volumetric flow rate } (q)} \quad (7.1)$$

*Don't
in correct.*

The volume of the STR for liquefaction was calculated to be 1069 ft³. A much larger volume of 7063 ft³ was obtained for saccharification. Data obtained from research verifies that this volume is appropriate for very large bio-reaction capacities (Humbird 4). Basic dimensions for each reactor were found using heuristics for STR design. These include a length to diameter ratio of 1.5 and an impeller diameter one-third of tank diameter (Brodkey 371). The height and diameter of the reactors were calculated using the formula for finding the volume of a cylinder also while considering heuristics given for vessels with elliptical bottoms. The volume of a cylinder with an elliptical bottom is obtained by the formula (cylinder volume + ($\pi/24 * d^3$)). Headspace allowance was factored by making the diameter equal to that of liquid height. The design of the vessel internals is just as important as sizing the volume of the vessel. The viscosity of the fluid being stirred plays a huge role in determining the agitator type that would be used for the reactor.

The viscosity of the fluid during hydrolysis is high with a value of about 1300cP during liquefaction and 600cP during saccharification. At these viscosities, a flat bade with a disk impeller would be appropriate. In addition to this, baffles were added because notable data shows that their addition aids efficient mixing in the reactor (Geankoplis 178). After selecting the impeller type, the power required (P) for agitation was calculated. Reynold's number was calculated using Equation 7.2 shown below.

$$N'_{Re} = \frac{\rho * N * D_a}{\mu} \quad (7.2)$$

Using Figure 7.1 below, a Reynold's number of 6.0e5 was plotted against curve 1 to get a power number (Np) for the liquefaction and saccharification phase (Geankoplis 181). With that power number, the power needed was calculated using Equation 7.3 below.

$$N_p = \frac{P}{\rho * N^3 * D_a^5} \quad (7.3)$$

Where: Da = agitator blade diameter, ft

N = agitator speed, revs/s

P = power

ρ = density, lb/ft³

} $\Rightarrow$ values?

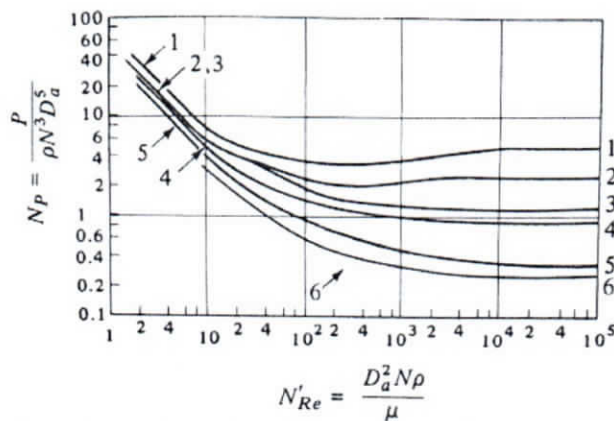


Figure 7.2-4 Power correlations for various impellers and baffles (see Fig. 7.2-3c for dimension D_a , D_i , J , and W). Data from D. W. Dodge and A. B. Metzner. A.I.Ch.E. J., 5, 189, © 1959 American Institute of Chemical Engineers.

- Curve 1. Flat six-blade turbine with disk (like Fig. 7.2-3 but six blades); $D_a/W = 5$; four baffles each $D_i/J = 12$.
- Curve 2. Flat six-blade open turbine (like Fig. 7.2-2c); $D_a/W = 8$; four baffles each $D_i/J = 12$.
- Curve 3. Six-blade open turbine (pitched-blade) but blades at 45° (like Fig. 7.2-2d); $D_a/W = 8$; four baffles each $D_i/J = 12$.
- Curve 4. Propeller (like Fig. 7.2-1); pitch = $2D_a$; four baffles each $D_i/J = 10$; also holds for same propeller in angular off-center position with no baffles.
- Curve 5. Propeller; pitch = D_a ; four baffles each $D_i/J = 10$; also holds for same propeller in angular off-center position with no baffles.
- Curve 6. High-efficiency impeller; four baffles each $D_i/J = 12$.
- Curves 1, 2, and 3 reprinted with permission from R. L. Bates, P. L. Fondy, and R. R. Corpstein, Ind. Eng. Chem. Proc. Des. Dev., 2, 310 (1963). Copyright by the American Chemical Society. Curves 4 and 5 from J. H. Rushton, E. W. Costich, and H. J. Everett, Chem. Eng. Progr., 46, 395, 467 (1950). With permission.

Figure 7.1 Graph for obtaining the power number used to calculate power requirement

R-103

The design of a packed bed reactor is heavily influenced by the catalyst, which in turn depends on certain specifications such as shape, loading selection, dimensions, and life activity. Cylindrical catalysts will be used in this reactor because they usually have low voidage values (Afandizadeh 672). Low voidage facilitates an increase in surface area and the accommodation of more pellets. The loading selection that would be used is snowstorm filling, which involves passing the packing material over staggered wires or wire meshes. This method provides the

how low? range?

densest packing and gives the least voidage (Afandizadeh 676). In the design of packed beds, it is customary to assume that the particles are of equal size although this is far from reality. The catalyst used for this process has a diameter of 0.0098 ft and a length of 0.0082 ft. *No: days??* (Venkatasubramanian 144). Under the optimum operation conditions that have been described earlier, the typical half-life is in a range of 1800 - 1900 hours, and the enzyme can be used for about 3.5 half-lives (Antrim 135). With this type of activity, the enzyme can be used for 235 days before regeneration is needed.

To obtain the bed volume (working volume) for the packed bed reactor, the residence time equation from Equation 7.1 was used as the design equation. Under the given optimum operating conditions, a residence time of 4 hours is suitable (Venkatasubramanian 146). The volumetric flow rate was obtained using the density of glucose syrup and the mass inlet mass flow rate in Equation 7.4 below.

$$q = \frac{\text{mass flow rate (m)}}{\text{density (\rho)}} \quad \text{ok.} \quad (7.4)$$

At a range of 131°F – 140°F the density of glucose syrup is 11.73 lb/gallon. Using this information, the volumetric flow rate obtained was 980 ft³/hr. Using the volumetric flow rate obtained and the residence time, the working volume for the reactor is 3920 ft³. For this design, 14 reactors each with a working volume of 280 ft³ ~~would be operated~~ in parallel (Buchholz 408). A typical industrial scale PBR has a diameter to height ratio of 0.3 and with this information, the height and diameter for one reactor were obtained to be 15.8 ft and 4.8 ft, respectively (Buchholz 408). It is important to note that the dimensions obtained above are for the reactor bed. The reactor vessel dimension was calculated with an additional 3ft added to the bed height. The additional space added would contain ceramic balls to promote uniform velocity across the bed

(Silla 392). The weight of catalyst in the bed for one reactor was calculated using Equation 7.5 below. The weight obtained for one reactor was 8450 lbs. The pressure drop across the reactor is 2 psi. It was found using Ergun's equation for an isothermal operation. The design equation used is Equation 7.6 below. Lastly, a heat duty of 34639 Btu/h was obtained using Equation 5.2 as described in section 5.5 above.

$$W = (1 - \phi)A_c Z * \rho_c \quad (7.5)$$

$$p = \frac{P}{P_o} = (1 - \alpha W)^{\frac{1}{2}} \quad (7.6)$$

$$\alpha = \frac{2\beta_o}{A_c(1-\phi)\rho_c P_o} \quad (lb^{-1}) \quad (7.7)$$

$$\beta_o = \frac{G(1-\phi)}{\rho_o g_c D_p \phi^3} * \left[\frac{150(1-\phi)\mu}{D_p} + 1.75G \right] * \frac{14.7}{12} \quad (psi) \quad (7.8)$$

$$G \text{ (superficial mass velocity)} = \rho_o u \quad \left(\frac{lb}{ft^2 \cdot h} \right) \quad (7.9)$$

Where:

Φ = void fraction

D_p = diameter of the particle in the bed, (ft)

μ = viscosity of liquid passing through the bed, lb/ft²·h

ρ_o = liquid density, lb/ft³

ρ_c = catalyst density, lb/ft³

g_c = 32.174 lb·ft/s²·lb_f (conversion factor)

A_c = cross-sectional area of bed

place in
nomenclature
section

Z = Bed height

P_0 = inlet pressure

u = superficial velocity = volumetric flow rate $\div$ cross-sectional area

CC-101

A simulated moving bed is applied in this design to allow for continuous countercurrent chromatographic separation of the glucose-fructose syrup mixture (Schulte 3). The most common configuration of this method involves arranging about twelve fixed bed chromatography columns in series with two inlet and two outlet streams. The inlet streams being the incoming syrup to be separated and the eluant which is water and the outlet streams being the enriched syrup stream as the extract and the recycled glucose stream as the raffinate. The inlet and outlet ports are shifted one single column ahead in the direction of the fluid flow after the established switching period and this allows for the simulation of continuous flow (Azevedo 171). A separation cycle is completed when the initial location of the input and output ports have been reached.

In the design of the single fixed bed chromatography column, a laboratory experiment was scaled up to obtain the dimensions of the columns used in this design (Wang 545). The laboratory column had a length of 0.98 ft and an internal diameter of 0.049 ft. The retention times of glucose and fructose obtained from the chromatogram were 10.5 and 8.8 minutes respectively and this was used in Equation 7.10 to find the average theoretical number of plates for the analytical column to be 2057.

$$\text{Theoretical Number of Plates (N)} = 16 * \left(\frac{\text{retention time (t}_r\text{)}}{\text{baseline peak width (w)}} \right)^2 \quad (7.10)$$

This value was then used in Equation 7.11 to calculate the height equivalent of a theoretical plate (HETP) to be 0.00048 ft.

$$\text{HETP} = \frac{\text{column length (L)}}{N} \quad (7.11)$$

Peak resolution gives the degree of separation of two components in a mixture (Chromatographic Methods). This was calculated using Equation 7.12 below to give a peak resolution of 1.71.

$$\text{Resolution (R)} = \frac{2(t_{ra} - t_{rb})}{w_a + w_b} \quad (7.12)$$

For this design a peak resolution of 3.5 was used to achieve better component separation (Theory of HPLC) and using Equation 7.10, the theoretical number of plates for this column's design was calculated to be 11255.

$$R = \frac{1}{4} * \sqrt{N} * \frac{(\alpha - 1)}{(\alpha)} * \frac{(k)}{(1+k)} \quad (7.13)$$

Where α and k are component selectivity and retention factors respectively. These values are independent of column length and do not change with it (Chromatographic Methods), therefore, the number of theoretical plates of a different column can be calculated using:

$$\frac{R_1}{R_2} = \frac{\sqrt{N_1}}{\sqrt{N_2}} \quad (7.14)$$

The packed column length for this design was then calculated using Equation 7.11 and the height equivalent of a theoretical plate of 0.00048 ft and this came to be 4.13 ft.

The packing used for this design was the Dowex Monoshpere 99 in the Ca^{2+} form which is a gel-type resin with a particle size of 320 microns and is used in HFCS enrichment. In calculating the total column length, freeboard accounting for 70% backwash expansion of the resin was added with an additional 2 ft for the safety factor (DOW). This brought the total

column length to 9 ft. The column diameter was calculated using a recommended length to diameter ratio of 3 for ion exchange columns (Desal Supplies) and this gave a value of 3 ft.

The column switching time can be calculated using Equation 7.15

$$t^* = \frac{L_c}{u_s} \quad (7.15)$$

With a solid velocity (u_s) of 0.0019 ft/sec and a column length (L_c) of 9 feet, the switching time (t^*) was calculated to be 78 minutes.

M-103

what is the phase in this mixer?

The volume of the mixer was determined using Equation 7.1. With a residence time of 10 minutes, the volume obtained was 36 ft³. To determine the amount of power needed to agitate the mixer, Equation 7.3 was used. The power (kW) for mixers 101, 102, and 103 are 2.7, 0.04, and 0.32 respectively. The density varied between 30% or 50% of the dry substance of the slurry. The agitator speed was set to 0.832 revolutions per second (Lele 1). The agitator diameter was based on the diameter of the mixing tank. The dimensionless power number was set to 0.28 for gate paddle agitators (Lele 1). The volumes of the mixers 101 through 103 are 758, 186, and 36 ft³ respectively. The residence times between 5 – 10 minutes were used for each mixer (Perry 11). To determine the diameter of the mixer, Equation 7.16 was used.

$$V = \frac{\pi}{4} * d^2 * h \quad (7.16)$$

Where,

V = volume of the cylinder (ft³)

d = diameter of the cylinder (ft)

h = height of cylinder (ft)

The diameters for mixers 101 through 103 are 8, 4, and 3 ft, respectively. The height of mixers 101 through 103 are 14, 18, and 6 ft, respectively. The construction material recommended is stainless steel 316 and the type of agitator is gate paddles (Walas 287).

7.2 Equipment Listing and Specifications

The specification sheets for the major units and equipment involved in the overall process are shown subsequently. The cost values shown are for the total number of units required including spares and backups and the design data is per vessel.

M-101				
Identification	Item:	Mixer		
	Item No:	1	Date:	05/05/20
	No. Required:	1		
	Total Purchase Cost:	\$15,290	Bare Module Cost:	\$38,225
Function: Mixing of starch powder and water to produce starch slurry				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	29,100	97,660		
Volumetric Flow (U.S. gallons/hr):	3,092	10,307		
Component Flow (lb/hr):				
Starch	25,608	25,608		
Water	71,101	71,102		
Waste Feed	291	950		
Operating Conditions:				
Normal Operating Temp. (°F):	77	Normal Operating Pressure (psi):	14.7	
Max. Design Temp. (°F):	127	Max. Design Pressure (psi):	150	
Min. Design Temp. (°F):	52	Design Pressure (psi):	38.7	
Residence Time (mins):	5-10			
Design Data:				
Height (ft):	14	Material:	SS 316	
Diameter (ft):	8	Type of agitator:	Gate Paddles	
Volume (ft³):	758	Diameter of agitator (ft):	7	
Power Consumed (kW):	2.7	No. of impellers:	1	
		Orientation:	Vertical	
Comments:	Open mixer vessel			

P-101				
Identification	Item:	Progressive Cavity Pump		
	Item No:	1	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$8,970	Bare Module Cost:	\$29,601
Function: Transportation of starch slurry to STR for liquefaction				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	97,600	97,600		
Volumetric Flow (U.S. gallons/hr):	10,307	10,307		
Component Flow (lb/hr):				
Starch	25,608	25,608		
Water	71,102	71,102		
Waste Feed	950	950		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	9.84	Material:	SS 316	
Hydraulic Power (hp):	0.41	Allowable ΔP (psi):	4.04	
Brake Power (hp):	0.58	Efficiency (%):	70	
Shaft work required (kW):	0.44	Head (ft):	8	

JC-101				
Identification	Item:	Jet Cooker		
	Item No:	1	Date:	05/05/20
	No. Required:	1		
	Total Purchase Cost:	\$454	Bare Module Cost:	\$1,816
Function: Applying steam to heat starch slurry to optimum temperature for liquefaction in STR				
Operation: Batch process				
	Feed	Product	Steam	
Materials Handled:				
Total Flow (lb/hr):	97,660	145,778	48,778	
Volumetric Flow (U.S. gallons/hr):	40,822	61,355	7,017	
Component Flow (lb/hr):				
Starch	25,608	38,485	-	
Water	71,102	106,855	48,778	
Waste feed	950	438	-	
Operating Conditions:				
Phase	Liquid	Liquid	Vapor	
Operating Temperature (°F):	105	203	221	
Operating Pressure (psi)	14.7	14.7	17.5	
Design Data:				
Material:	SS 316			
Orientation:	Horizontal			
Comments:	Steam injection operation			

R-101				
Identification	Item:	Stirred Tank Reactor		
	Item No:	1	Date:	05/05/20
	No Required:	7		
	Total Purchase Cost:	\$525,000	Bare Module Cost:	\$1,575,000
Function: Liquefaction of corn starch				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow (lb/hr):				
Starch	38,485	-		
Glucose	-	7,289		
Oligosaccharides	-	65,600		
Water	106,855	58,311		
Waste Feed	438	14,578		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	221	221		
Pressure (psi)	14.7	14.7		
Residence Time (hours)	2			
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	1,069	Agitator Type	Flat blade	
Height (ft)	14.5	Impeller Type	Disk	
Diameter (ft)	9.68	No. of Impellers	3	
Skirt Height (ft)	2.9	Impeller Diameter (ft)	3.23	
Liquid Height (ft)	9.68	Impeller Disk Diameter (ft)	2.42	
Material	SS 316	Impeller Speed (rpm)	97.08	
		Power Requirement (kW)	21.40	
		No. of Blades	6	
		Blade Length (ft)	0.81	
		Blade Width (ft)	0.65	
		Baffle Width (ft)	0.97	
		Baffle tip distance from vessel bottom (ft)	1.936	
Comments:	7 reactors in series			

P-102				
Identification	Item:	Progressive Cavity Pump		
	Item No:	2	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$214,058	Bare Module Cost:	\$706,391
Function: Transportation of starch slurry to STR for saccharification				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow (lb/hr):				
Glucose	7,289	7,289		
Oligosaccharides	65,600	65,600		
Water	58,311	58,311		
Waste Feed	14,578	14,578		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	9.84	Material:	SS 316	
Hydraulic Power (hp):	5.32	Allowable ΔP (psi):	17.82	
Brake Power (hp):	8.18	Efficiency (%):	65	
Shaft work required (kW):	6.0	Head (ft):	72	

E-101				
Identification	Item:	Shell and Tube Heat Exchanger		
	Item No:	1	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$47,348	Bare Module Cost:	\$150,093
Function: Cooling starch slurry for saccharification				
Operation: Batch process				
	Tube Side	Shell Side		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow (lb/hr):				
Glucose	7,289	7,289		
Oligosaccharides	65,600	65,600		
Water	58,311	58,311		
Waste Feed	14,578	14,578		
Operating Conditions:				
Normal Operating Temp. (°F):	68-221	Normal Operating Pressure (psi):	14.7	
Max. Operating Temp. (°F):	271	Max. Operating Pressure (psi):	40	
Min. Operating Temp. (°F):	43	Design Operating Pressure (psi):	65	
Design Data:				
Heat Transfer Area (ft²):	381	Material:	SS 316	
ΔT_{LM} (°F):	104	Shell Diameter(in):	24	
Shell-Tube Arrangement:	2-4	Tube Outside Diameter(in):	0.75 on 1" Triangles	
Heat required (kW):	1,760	Tube Count:	428	
Overall Heat Transfer Coef. (BTU/hr-ft²-R):	150	TEMA Config.:	AEL	

R-102				
Identification	Item:	Stirred Tank Reactor		
	Item No:	2	Date:	05/05/20
	No Required:	29		
	Total Purchase Cost:	\$11,600,000	Bare Module Cost:	\$34,800,000
Function: Saccharification of corn starch to dextrose				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow (lb/hr):				
Glucose	7,289	69,245		
Oligosaccharides	65,600	3,645		
Water	58,311	58,310		
Waste feed	14,578	14,578		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	140	140		
Pressure (psi)	10.7	10.7		
Residence Time (hours)	55			
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	7,063	Agitator Type	Flat-blade	
Height (ft)	27.3	Impeller Type	Disk	
Diameter (ft)	18.2	No. of Impellers	3	
Skirt Height (ft)	5.46	Impeller Diameter (ft)	6.07	
Liquid Height (ft)	18.2	Impeller Disk Diameter (ft)	4.55	
Material	SS 316	Impeller Speed (rpm)	51.63	
		Power Requirement (kW)	21.40	
		No. of Blades	6	
		Blade Length (ft)	1.52	
		Blade Width (ft)	1.21	
		Baffle Width (ft)	1.82	
		Baffle tip distance from vessel bottom (ft)	3.64	
Comments:	29 reactors in series			

P-103				
Identification	Item:	Centrifugal Pump		
	Item No:	3	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$4,165	Bare Module Cost:	\$13,745
Function: Transportation of glucose syrup to glucose syrup tank				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	145,778	145,778		
Volumetric Flow (U.S. Gallons/hr)	15,485	15,485		
Component Flow (lb/hr):				
Glucose	69,245	69,245		
Oligosaccharides	3,645	3,645		
Water	58,310	58,310		
Waste Feed	14,578	14,578		
Operating Conditions:				
Phase:	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.79	Allowable ΔP (psi):	8.73	
Brake Power (hp):	1.13	Efficiency (%):	70	
Shaft work required (kW):	0.52	Impeller Size (inch):	5	
Impeller specific speed (rpm):	3450	Head (ft):	11	

E-102				
Identification	Item:	Shell and Tube Heat Exchanger		
	Item No:	2	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$47,348	Bare Module Cost:	\$150,093
Function: Cooling glucose syrup for storage in tank				
Operation: Batch process				
	Tube Side	Shell Side		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow(lb/hr):				
Starch	69,245	69,245		
Oligosaccharides	3,645	3,645		
Water	58,310	58,310		
Waste Feed	14,578	14,578		
Operating Conditions:				
Normal Operating Temp. (°F):	68-221	Normal Operating Pressure (psi):	14.7	
Max. Operating Temp. (°F):	271	Max. Operating Pressure (psi):	40	
Min. Operating Temp. (°F):	43	Design Operating Pressure (psi):	65	
Design Data:				
Heat Transfer Area (ft ²):	381	Material:	SS 316	
ΔT_{LM} (°F):	104	Shell Diameter(in):	24	
Shell-Tube Arrangement:	2-4	Tube Outside Diameter(in):	0.75 on 1" Triangles	
Heat required (kW):	1,760	Tube Count:	428	
Overall Heat Transfer Coef. (BTU/hr-ft ² -R):	150	TEMA Config.:	AEL	

TK-101				
Identification	Item:	Tank		
	Item No:	1	Date:	05/05/20
	No. Required:	1		
	Total Purchase Cost:	\$5,300	Bare Module Cost:	\$15,900
Function: Holding tank for glucose syrup before entering rotary filter				
Operation: Batch process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	145,778	145,778		
Volumetric Flow (U.S. gallons/hr):	15,485	15,485		
Component Flow (lb/hr):				
Glucose	69,245	69,245		
Oligosaccharides	3,645	3,645		
Water	58,310	58,310		
Waste Feed	14,578	14,578		
Operating Conditions:				
Phase	Liquid	Liquid		
Normal Operating Temperature (°F):	77	Normal Operating Pressure (psi):	14	
Max. Operating Temperature (°F):	127	Max. Operating Pressure (psi):	150	
Min. Operating Temperature (°F):	52	Design Operating Pressure (psi):	30	
Design Data:				
Orientation:	Vertical	Material:	SS 316	
Height (ft):	16	Type:	Conical Roof	
Diameter (ft):	9	Volume (ft³):	1240	

P -104				
Identification	Item:	Centrifugal Pump		
	Item No:	4	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$5,286	Bare Module Cost:	\$17,444
Function: Transportation of glucose syrup to rotary filter				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	145,778	145,778		
Volumetric Flow (U.S. gallons/hr)	15,485	15,485		
Component Flow (lb/hr):				
Glucose	69,245	69,245		
Oligosaccharides	3,645	3,645		
Water	58,310	58,310		
Waste Feed	14,578	14,578		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.60	Allowable ΔP (psi):	2.01	
Brake Power (hp):	0.86	Efficiency (%):	70	
Shaft work required (kW):	0.18	Impeller Size (inch):	5	
Impeller specific speed (rpm):	3450	Head (ft):	8	

P-101				
Identification	Item:	Rotary Filter		
	Item No:	1	Date:	05/05/20
	No Required:	1		
	Total Purchase Cost:	\$252,252	Bare Module Cost:	\$908,107
Function: Separation of impurities from glucose syrup				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	145,778	131,200		
Volumetric Flow (U.S. gallons/hr):	15,485	11,184		
Component Flow (lb/hr):				
Glucose	69,245	62,320		
Oligosaccharides	3,645	3,280		
Water	58,310	65,600		
Waste Feed	14,578	-		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	86	86		
Pressure (psi)	14.7	14.7		
Design Data:				
Vessel				
Orientation	Horizontal drum			
Drum Diameter (ft)	7			
Drum Length (ft)	10			
Rotating Speed (rpm)	2			
Precoat Discharge	Diatomaceous earth			
Capacity (lb/hr*ft ²)	200			
Surface area of filtration (ft ²)	729			
Material	SS 304L			

P-105				
Identification	Item:	Centrifugal Pump		
	Item No:	5	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$13,882	Bare Module Cost:	\$45,811
Function: Transportation of glucose syrup from rotary filter to adsorption column				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	131,200	131,200		
Volumetric Flow (U.S. gallons/hr)	11,184	11,184		
Component Flow (lb/hr):				
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.68	Allowable ΔP (psi):	8.33	
Brake Power (hp):	0.97	Efficiency (%):	70	
Shaft work required (kW):	0.65	Impeller Size (inch):	5	
Impeller specific speed (rpm):	2880	Head (ft):	10	

A-101				
Identification	Item:	Adsorption Column		
	Item No:	1	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$110,265	Bare Module Cost:	\$187,451
Function: Removal of color forming components in the glucose syrup product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	131,200	131,200		
Volumetric Flow (U.S. gallons/hr):	12,676	12,676		
Component Flow (lb/hr):				
Fructose	0	0		
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	86	86		
Pressure (psi)	6	6		
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	390	Bed volume (ft ³)	271	
Height (ft)	10.86	Bed height (ft)	7.55	
Material	SS 316	Bed diameter (ft)	6.76	
		Packing type	Macroporous Resins	
		Dry weight capacity (meq/g)	0.8	
		Amount of Resin needed (lb)	17,425	
		Resin regeneration time (mins)	45	

IX-101				
Identification	Item:	Strong Acid Cation Exchanger		
	Item No:	1	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$60,384	Bare Module Cost:	\$102,653
Function: Removal of mineral ions in the glucose syrup product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	131,200	131,200		
Volumetric Flow (U.S. gallons/hr):	12,676	12,676		
Component Flow (lb/hr):				
Fructose	0	0		
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	86	86		
Pressure (psi)	6	6		
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	112.6	Bed volume (ft ³)	50.5	
Height (ft)	11.7	Bed height (ft)	5.25	
Material	SS 316	Bed diameter (ft)	3.5	
		Packing type	Macroporous Resins	
		Total exchange capacity (eq/L)	1.8	
		Resin backwash	80%	
		Amount of Resin needed (lb)	3,783	
		Resin regeneration time (mins)	45	
Comments:	2 columns are running in series before the anion exchangers			

IX-102				
Identification	Item:	Weak Base Anion Exchanger		
	Item No:	2	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$223,158	Bare Module Cost:	\$379,369
Function: Removal of mineral ions in the glucose syrup product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	131,200	131,200		
Volumetric Flow (U.S. gallons/hr):	12,676	12,676		
Component Flow (lb/hr):				
Fructose	0	0		
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	86	86		
Pressure (psi)	6	6		
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	934.8	Bed volume (ft ³)	391.6	
Height (ft)	25	Bed height (ft)	10.4	
Material	SS 316	Bed diameter (ft)	6.9	
		Packing type	Macroporous Resins	
		Total exchange capacity (eq/L)	1.5	
		Resin backwash	100%	
		Amount of Resin needed (lb)	25,428	
		Resin regeneration time (mins)	45	
Comments:	2 columns are running in series after the cation exchangers			

P-106				
Identification	Item:	Centrifugal Pump		
	Item No:	6	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$13,882	Bare Module Cost:	\$45,811
Function: Transportation of glucose syrup from ion exchanger to M-102				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	131,200	131,200		
Volumetric Flow (U.S. Gallons/hr)	11,184	11,184		
Component Flow (lb/hr):				
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase:	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.68	Allowable ΔP (psi):	8.33	
Brake Power (hp):	0.97	Efficiency (%):	70	
Shaft work required (kW):	0.65	Impeller Size (inch):	5	
Impeller specific speed (rpm):	2880	Head (ft):	10	

M-102				
Identification	Item:	Mixer		
	Item No:	2	Date:	05/05/20
	No. Required:	1		
	Total Purchase Cost:	\$57,788	Bare Module Cost:	\$144,470
Function: Change in pH and enzyme stabilization				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	131,200	131,200		
Volumetric Flow (U.S. gallons/hr):	11,184	11,184		
Component Flow (lb/hr):				
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Normal Operating Temp. (°F):	77	Normal Operating Pressure (psi):	14.7	
Max. Design Temp. (°F):	127	Max. Design Pressure (psi):	150	
Min. Design Temp. (°F):	52	Design Pressure (psi):	38.7	
Residence Time (mins):	5-10			
Design Data:				
Height (ft):	18	Material:	SS 316	
Diameter (ft):	4	Type of agitator:	Gate Paddles	
Volume (ft³):	186	Diameter of agitator (ft):	3	
Power Consumed (kW):	0.04	No. of impellers:	1	
		Orientation:	Vertical	
Comments:	Open mixer vessel			

P-107				
Identification	Item:	Centrifugal Pump		
	Item No:	7	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$14,096	Bare Module Cost:	\$46,517
Function: Transportation of glucose syrup from M-102 to evaporator				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	131,200	131,200		
Volumetric Flow (U.S. gallons/hr)	11,184	11,184		
Component Flow (lb/hr):				
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	65,600	65,600		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.69	Allowable ΔP (psi):	8.33	
Brake Power (hp):	0.98	Efficiency (%):	70	
Shaft work required (kW):	0.66	Impeller Size (inch):	5	
Impeller specific speed (rpm):	2880	Head (ft):	10	

EY 101				
Identification	Item:	Evaporator		
	Item No:	1	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$1,806,387	Bare Module Cost:	\$6,683,632
Function: Concentration of glucose syrup				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	131,200	131,200		
Volumetric Flow (U.S. gallons/hr):	11,184	11,184		
Component Flow (lb/hr):	62,320	62,320		
Glucose	3,280	3,280		
Oligosaccharides	65,600	65,600		
Water	131,200	131,200		
Operating Conditions:				
	Effect 1	Effect 2	Effect 3	
Overall heat transfer coefficient (Btu/hr* ft^2 *°F)	550	350	200	
Temperature (°F)	227	188	121	
Pressure (psi)	29	14.7	1.89	
Steam temperature (°F)	248	221	180	
Heat duty (Btu/hr)	2.1×10^7	1.3×10^7	1.6×10^7	
Design Data:				
Type	Forced Circulation			
Method of operation	Triple Effect			
Residence time (minutes)	25			
Area (ft^2)	1421	1421	1421	
Steam economy	2.25			
Material	SS 316			

E-103				
Identification	Item:	Condenser		
	Item No:	1	Date:	05/05/20
	No Required:	1		
	Total Purchase Cost:	\$40,208	Bare Module Cost:	\$120,624
Function: Concentration of glucose syrup				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	49,200	49,200		
Volumetric Flow (U.S. gallons/hr):	5,896	5,896		
Component Flow (lb/hr):				
Glucose	-	-		
Oligosaccharides	-	-		
Water	49,200	49,200		
Operating Conditions:				
Phase	Vapor	Liquid		
Pressure (psi)	14.7	14.7		
Heat load (Btu/hr)	2×10^7			
Vapor temperature (°F)	111.43	77		
Cooling water temperature (°F)	85	96.8		
Water consumption (lb water/lb vapor)	540			
Design Data:				
Type	Counter-current barometric			
Material	Carbon Steel			
Diameter (ft)	49			
Height (ft)	34			
Area (ft ²)	2434			

P-109				
Identification	Item:	Centrifugal Pump		
	Item No:	9	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$14,416	Bare Module Cost:	\$47,573
Function: Transportation of glucose syrup from evaporator to PBR				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	82,000	82,000		
Volumetric Flow (U.S. Gallons/hr)	2,906	2,906		
Component Flow (lb/hr):				
Glucose	62,320	62,320		
Oligosaccharides	3,280	3,280		
Water	16,400	16,400		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	1.70	Allowable ΔP (psi):	57.24	
Brake Power (hp):	2.10	Efficiency (%):	78	
Shaft work required (kW):	1.50	Impeller Size (inch):	5	
Impeller specific speed (rpm):	2880	Head (ft):	39	

R-103				
Identification	Item:	Packed Bed Reactor		
	Item No:	3	Date:	05/05/20
	No Required:	21		
	Total Purchase Cost:	\$1,287,300	Bare Module Cost:	\$5,149,200
Function: Isomerization of glucose to fructose				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	86,000	86,000		
Volumetric Flow (U.S. gallons/hr):	7,334	7,334		
Component Flow (lb/hr):				
Fructose	-	27,100		
Glucose	65,300	35,500		
Oligosaccharides	3,300	1,900		
Water	17,400	21,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	121	140		
Pressure (psi)	14.7	14.7		
Residence time (hours)	4			
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	477	Bed volume (ft ³)	283	
Height (ft)	20	Bed height (ft)	17	
Outer diameter (ft)	5	Bed diameter (ft)	4.94	
Wall thickness (ft)	0.03	Catalyst weight (lb)	8,450	
Pressure drop (psi)	2	Pellet shape	Cylindrical beads	
Material	SS 316	Catalyst life (days)	235	
Heating medium	Hot water			
Heat duty (Btu/hr)	34,639			
Hot water flow (lb/hr)	68,200			
Comments:	14 reactors are running in parallel 7 backup reactors			

P-110				
Identification	Item:	Centrifugal Pump		
	Item No:	10	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$16,800	Bare Module Cost:	\$55,440
Function: Transports HFCS 42 syrup from PBR to ion exchanger				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	86,000	86,000		
Volumetric Flow (U.S. Gallons/hr)	7,334	7,334		
Component Flow (lb/hr):				
Fructose	27,100	27,100		
Glucose	35,500	35,500		
Oligosaccharides	1,900	1,900		
Water	21,500	21,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	1.0	Allowable ΔP (psi):	31.29	
Brake Power (hp):	1.3	Efficiency (%):	77	
Shaft work required (kW):	4.5	Impeller Size (inch):	5	
Impeller specific speed (rpm):	1880	Head (ft):	23	

IX-103				
Identification	Item:	Strong Acid Cation Exchanger		
	Item No:	3	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$48,000	Bare Module Cost:	\$81,600
Function: Removal of mineral ions in the HFCS 42 product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	86,000	86,000		
Volumetric Flow (U.S. gallons/hr):	7658	7658		
Component Flow (lb/hr):				
Fructose	27,100	27,100		
Glucose	35,500	35,500		
Oligosaccharides	1,900	1,900		
Water	21,500	21,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	140	140		
Pressure (psi)	24	24		
Design Data:				
Vessel		Vessel Internals		
Volume (ft³)	57.3	Bed volume (ft³)	23	
Height (ft)	10	Bed height (ft)	4	
Material	SS 316	Bed diameter (ft)	2.7	
		Packing type	Macroporous Resins	
		Total exchange capacity (eq/L)	1.8	
		Resin backwash	90%	
		Amount of Resin needed (lb)	1728	
		Resin regeneration time (mins)	45	
Comments:	2 columns are running in series before the anion exchangers			

EX-104				
Identification	Item:	Weak Base Anion Exchanger		
	Item No:	4	Date:	05/05/20
	No Required:	3		
	Total Purchase Cost:	\$93,000	Bare Module Cost:	\$158,100
Function: Removal of mineral ions in the HFCS 42 product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	86,000	86,000		
Volumetric Flow (U.S. gallons/hr):	7658	7658		
Component Flow (lb/hr):				
Fructose	27,100	27,100		
Glucose	35,500	35,500		
Oligosaccharides	1,900	1,900		
Water	21,500	21,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	140	140		
Pressure (psi)	20	20		
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	235.6	Bed volume (ft ³)	93.23	
Height (ft)	16.3	Bed height (ft)	6.46	
Material	SS 316	Bed diameter (ft)	4.29	
		Packing type	Macroporous Resins	
		Total exchange capacity (eq/L)	1.5	
		Resin backwash	100%	
		Amount of Resin needed (lb)	6053	
		Resin regeneration time (mins)	45	
Comments:	2 columns are running in series after the cation exchangers			

P-112				
Identification	Item:	Centrifugal Pump		
	Item No:	12	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$16,800	Bare Module Cost:	\$55,440
Function: Transportation of HFCS 42 syrup from ion exchanger to storage tank				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	86,000	86,000		
Volumetric Flow (U.S. Gallons/hr)	7,334	7,334		
Component Flow (lb/hr):				
Fructose	27,100	27,100		
Glucose	35,500	35,500		
Oligosaccharides	1,900	1,900		
Water	21,500	21,500		
Operation Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.13	Allowable ΔP (psi):	4.88	
Brake Power (hp):	0.17	Efficiency (%):	76	
Shaft work required (kW):	0.13	Impeller Size (inch):	5	
Impeller specific speed (rpm):	1700	Head (ft):	6	

P-113				
Identification	Item:	Centrifugal Pump		
	Item No:	13	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$11,700	Bare Module Cost:	\$38,610
Function: Transportation of HFCS 42 syrup into storage tank				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	86,000	86,000		
Volumetric Flow (U.S. gallons/hr)	7,334	7,334		
Component Flow (lb/hr):				
Fructose	27,100	27,100		
Glucose	35,500	35,500		
Oligosaccharides	1,900	1,900		
Water	21,500	21,500		
Operation Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.13	Allowable ΔP (psi):	4.88	
Brake Power (hp):	0.17	Efficiency (%):	76	
Shaft work required (kW):	0.13	Impeller Size (inch):	5	
Impeller specific speed (rpm):	1700	Head (ft):	6	

GC-101				
Identification	Item:	Chromatography Column		
	Item No:	1	Date:	05/05/20
	No Required:	16		
	Total Purchase Cost:	\$224,000	Bare Module Cost:	\$380,800
Function: Enrichment of fructose concentration in HFCS 42				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	15,030	15,030		
Volumetric Flow (U.S. gallons/hr):	712.4	712.4		
Component Flow (lb/hr):				
Fructose	2,520	8,200		
Glucose	3,300	3,290		
Oligosaccharides	180	640		
Water	9,030	2,900		
Operating Conditions:				
Phase	Liquid	Liquid		
Temperature (°F)	140	140		
Pressure (psi)	15.1	13		
Design Data:				
Vessel		Vessel Internals		
Volume (ft ³)	63	Bed volume (ft ³)	29	
Height (ft)	9	Bed height (ft)	4.1	
Material	SS 316	Bed diameter (ft)	3	
		Packing type	Gel Resins	
		Resin backwash	70%	
		Amount of Resin needed (lb)	4985	
		Resin regeneration time (mins)	45	
		Switching time	78	
		Peak Resolution	3.5	
Comments:	Simulated Moving Bed approach is used with 12 columns running			

P-114				
Identification	Item:	Centrifugal Pump		
	Item No:	14	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$16,800	Bare Module Cost:	\$55,440
Function: Transportation of HFCS 42 from split point to M-103				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	35,000	35,000		
Volumetric Flow (U.S. gallons/hr)	2,985	2,985		
Component Flow (lb/hr):				
Fructose	11,040	11,040		
Glucose	14,430	14,430		
Oligosaccharides	790	790		
Water	8,740	8,740		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.72	Allowable ΔP (psi):	192.13	
Brake Power (hp):	1.01	Efficiency (%):	70	
Shaft work required (kW):	0.73	Impeller Size (inch):	5	
Impeller specific speed (rpm):	4000	Head (ft):	131	

P-115				
Identification	Item:	Centrifugal Pump		
	Item No:	15	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$10,420	Bare Module Cost:	\$34,386
Function: Transportation of HFCS 90 from chromatography column to M-103				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	11,000	11,000		
Volumetric Flow (U.S. gallons/hr)	938	938		
Component Flow (lb/hr):				
Fructose	8,200	8,200		
Glucose	270	270		
Oligosaccharides	640	640		
Water	1,890	1,890		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.72	Allowable ΔP (psi):	192.13	
Brake Power (hp):	1.01	Efficiency (%):	70	
Shaft work required (kW):	0.73	Impeller Size (in):	5	
Impeller specific speed (rpm):	4000	Head (ft):	131	

Identification				
Item:	Centrifugal Pump			
Item No:	16	Date:	05/05/20	
No. Required:	2			
Total Purchase Cost:	\$14,420	Bare Module Cost:	\$47,586	
Function: Transportation of HFCS 55 syrup from M-103 to storage tank				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	46,000	46,000		
Volumetric Flow (U.S. gallons/hr)	3,922	3,922		
Component Flow (lb/hr):				
Fructose	19,000	19,000		
Glucose	14,500	14,500		
Oligosaccharides	1,000	1,000		
Water	11,500	11,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	1.40	Allowable ΔP(psi):	0.2	
Brake Power (hp):	1.88	Efficiency (%):	76	
Shaft work required (kW):	1.50	Impeller Size (in):	5	
Impeller specific speed (rpm):	2000	Head (ft):	62	

P-117				
Identification	Item:	Centrifugal Pump		
	Item No:	17	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$10,420	Bare Module Cost:	\$34,386
Function: Transportation of unconverted glucose from liquid chromatography to PBR				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr)	4,030	4,030		
Volumetric Flow (U.S. gallons/hr)	344	344		
Component Flow (lb/hr):				
Unconverted Glucose	3,020	3,020		
Water	1,010	1,010		
Operating Conditions:				
Phase	Liquid	Liquid		
Design Data:				
NPSH (ft):	5-25	Material:	Carbon Steel 45	
Hydraulic Power (hp):	0.10	Allowable ΔP (psi):	36	
Brake Power (hp):	0.11	Efficiency (%):	55	
Shaft work required (kW):	0.10	Head (ft):	30	
Impeller Speed (rpm):	2000	Impeller Size (in):	5	

M-103				
Identification	Item:	Mixer		
	Item No:	3	Date:	05/05/20
	No. Required:	1		
	Total Purchase Cost:	\$40,763	Bare Module Cost:	\$101,908
Function: Mixing HFCS 90 and HFC 42 to produce HFCS 55				
Operation: Continuous process				
	HFCS 90 Feed	HFCS 42 Feed	HFCS 55 Product	
Materials Handled:				
Total Flow (lb/hr):	11,000	35,000	46,000	
Volumetric Flow (U.S. gallons/hr):	938	2985	3922	
Component Flow (lb/hr):				
Fructose	8,200	11,040	19,000	
Glucose	270	14,430	14,500	
Oligosaccharides	640	790	1,000	
Water	1,890	8,740	11,500	
Operating Conditions:				
Normal Operating Temp. (°F):	77	Normal Operating Pressure (psi):	14.7	
Max. Design Temp. (°F):	127	Max. Design Pressure (psi):	150	
Min. Design Temp. (°F):	52	Design Pressure (psi):	38.7	
Residence Time (mins):	5-10			
Design Data:				
Height (ft):	6	Material:	SS 316	
Diameter (ft):	3	Type of agitator:	Gate Paddles	
Volume (ft³):	36	Diameter of agitator (ft):	1	
Power Consumed (kW):	0.32	No. of impellers:	1	
		Orientation:	Vertical	
Comments:	Open mixer vessel			

E-104				
Identification	Item:	Shell and Tube Heat Exchanger		
	Item No:	4	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$41,850	Bare Module Cost:	\$132,665
Function: Cooling HFCS 42 product stream for storage				
Operation: Continuous process				
	Tube Side	Shell Side		
Materials Handled:				
Total Flow (lb/hr):	43,000	43,000		
Volumetric Flow (U.S. gallons/hr):	3,666	3,666		
Component Flow (lb/hr):				
Fructose	13,545	13,545		
Glucose	17,738	17,738		
Oligosaccharides	967	967		
Water	22,200	1,287		
Operating Conditions:				
Normal Operating Temp. (°F):	68-140	Normal Operating Pressure (psi):	14.7	
Max. Operating Temp. (°F):	190	Max. Operating Pressure (psi):	40	
Min. Operating Temp. (°F):	43	Design Operating Pressure (psi):	65	
Design Data:				
Heat Transfer Area (ft ²):	310	Material:	SS 316	
ΔT_{LM} (°F):	36	Shell Diameter(in):	21.25	
Shell-Tube Arrangement:	2-4	Tube Outside Diameter(in):	0.75 on 1" Triangles	
Heat required (kW):	490	Tube Count:	326	
Overall Heat Transfer Coefficient (Btu/hr-ft ² -R):	150	TEMA Config.:	AEL	

10-105				
Identification	Item:	Shell and Tube Heat Exchanger		
	Item No:	2	Date:	05/05/20
	No. Required:	2		
	Total Purchase Cost:	\$41,850	Bare Module Cost:	\$132,665
Function: Cooling HFCS-55 product stream for storage				
Operation: Continuous process				
	Tube Side	Shell Side		
Materials Handled:				
Total Flow (lb/hr):	46,000	46,000		
Volumetric Flow (U.S. gallons/hr):	3,922	3,922		
Component Flow (lb/hr):				
Fructose	19,000	19,000		
Glucose	14,500	14,500		
Oligosaccharides	1,000	1,000		
Water	11,500	11,500		
Operating Conditions:				
Normal Operating Temp. (°F):	68-140	Normal Operating Pressure (psi):	14.7	
Max. Operating Temp. (°F):	190	Max. Operating Pressure (psi):	40	
Min. Operating Temp. (°F):	43	Design Operating Pressure (psi):	65	
Design Data:				
Heat Transfer Area (ft ²):	310	Material:	SS 316	
ΔT_{LM} (°F):	36	Shell Diameter(in):	21.25	
Shell-Tube Arrangement:	2-4	Tube Outside Diameter(in):	0.75 on 1" Triangles	
Heat required (kW):	470	Tube Count:	326	
Overall Heat Transfer Coefficient (Btu/hr-ft ² -R):	150	TEMA Config:	AEL	

TK-102				
Identification	Item:	Tank		
	Item No:	5	Date:	05/05/20
	No. Required:	29		
	Total Purchase Cost:	\$580,000	Bare Module Cost:	\$2,320,000
Function: Holding tank for HFCS 42 product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	43,000	43,000		
Volumetric Flow (U.S. gallons/hr):	3,665	3,665		
Component Flow (lb/hr):				
Fructose	13,545	13,545		
Glucose	17,738	17,738		
Oligosaccharides	967	967		
Water	10,750	10,750		
Operating Conditions:				
Phase	Liquid	Liquid		
Normal Operating Temperature (°F):	81	Normal Operating Pressure (psi):	14	
Max. Operating Temperature (°F):	127	Max. Operating Pressure (psi):	150	
Min. Operating Temperature (°F):	52	Design Operating Pressure (psi):	30	
Design Data:				
Orientation:	Vertical	Material:	SS 316	
Height (ft):	16	Type:	Conical Roof	
Diameter (ft):	9	Volume (U.S. gallons):	15,000	
Comments:	Holds about 3 days' worth of syrup before transport to final consumer			

TK-103				
Identification	Item:	Tank		
	Item No:	5	Date:	05/05/20
	No. Required:	30		
	Total Purchase Cost:	\$600,000	Bare Module Cost:	\$2,400,000
Function: Holding tank for HFCS 55 product				
Operation: Continuous process				
	Feed	Product		
Materials Handled:				
Total Flow (lb/hr):	46,000	46,000		
Volumetric Flow (U.S. gallons/hr):	3,922	3,922		
Component Flow (lb/hr):				
Fructose	19,000	19,000		
Glucose	14,500	14,500		
Oligosaccharides	1,000	1,000		
Water	11,500	11,500		
Operating Conditions:				
Phase	Liquid	Liquid		
Normal Operating Temperature (°F):	81	Normal Operating Pressure (psi):	14	
Max. Operating Temperature (°F):	127	Max. Operating Pressure (psi):	150	
Min. Operating Temperature (°F):	52	Design Operating Pressure (psi):	30	
Design Data:				
Orientation:	Vertical	Material:	SS 316	
Height (ft):	16	Type:	Conical Roof	
Diameter (ft):	9	Volume (U.S. gallons):	15,000	
Comments:	Holds about 3 days' worth of syrup before transport to final consumer			

8. ECONOMIC EVALUATION

8.1 Capital Cost

8.1.1 Fixed Capital Investment

The bare model cost of the main equipment involved in this design is shown in Table 8.1 below.

Table 8.1 Equipment Summary List

UNIT	Unit Type	Unit Purchase Cost (\$)	No	Total Purchase Cost (\$)	Bare Module Factor	Bare Module Cost (\$)
CY-101	Pneumatic Conveyor	72,809	1	72,809	2.2	160,180
M-101	Mixer	15,290	1	15,290	2.5	38,225
TK-101	Tank	5,300	1	5,300	3	15,900
JC-101	Jet Cooker	454	1	454	4	1,816
P-101	Pump	4,485	2	8,970	3.3	29,601
P-102	Pump	107,029	2	214,058	3.3	706,391
P-103	Pump	2,083	2	4,165	3.3	13,745
P-104	Pump	2,643	2	5,286	3.3	17,444
E-101	Cooler	23,674	2	47,348	3.17	150,093
R-101	Reactor	75,000	7	525,000	3	1,575,000
R-102	Reactor	400,000	29	11,600,000	3	34,800,000
F-101	Vacuum Rotary Filter	252,252	1	252,252	3.6	908,107
CY-102	Conveyor Belt	11,704	1	11,704	2.4	28,090
WT-101	Waste Vessel	385,090	2	770,180	3	2,310,540
P-105	Pump	6,941	2	13,882	3.3	45,811
A-101	Adsorption Column	36,755	3	110,265	1.7	187,451
IX-101	Cation Exchanger	20,128	3	60,384	1.7	102,653
IX-102	Anion Exchanger	74,386	3	223,158	1.7	379,369
P-106	Pump	6,941	2	13,882	3.3	45,811
M-102	Mixer	57,788	1	57,788	2.5	144,470
P-107	Pump	7,048	2	14,096	3.3	46,517
EV-101	Evaporator	602,129	3	1,806,387	3.7	6,683,632
E-103	Condenser	40,208	1	40,208	3	120,624
P-108	Pump	8,009	2	16,018	3.3	52,859
P-109	Pump	7,208	2	14,416	3.3	47,573
R-103	Reactor	61,300	21	1,287,300	4	5,149,200
P-110	Pump	8,400	2	16,800	3.3	55,440
IX-103	Cation Exchanger	16,000	3	48,000	1.7	81,600
P-111	Pump	8,400	2	16,800	3.3	55,440
IX-104	Anion Exchanger	31,000	3	93,000	1.7	158,100
P-112	Pump	8,400	2	16,800	3.3	55,440
CC-101	Chromatography Column	14,000	16	224,000	1.7	380,800
P-113	Pump	5,850	2	11,700	3.3	38,610
P-114	Pump	8,400	2	16,800	3.3	55,440
E-104	Cooler	20,925	2	41,850	3.17	132,665
E-105	Cooler	20,925	2	41,850	3.17	132,665
P-115	Pump	5,210	2	10,420	3.3	34,386
P-116	Pump	7,210	2	14,420	3.3	47,586
M-103	Mixer	40,763	1	40,763	2.5	101,908
TK-102	Storage Tank	20,000	29	580,000	4	2,320,000
TK-103	Storage Tank	20,000	30	600,000	4	2,400,000
	Water treatment		1			500,000
				\$	18,948,513	\$ 60,311,182

Font

The unit purchase costs were obtained from tables and graphs published by *Turton, et al.* 2018 and *Ulrich 1984* and adjusted to the 2019 cost index value of 540. The total purchase cost was obtained for the total number of equipment and the bare model cost was obtained by multiplying this value by the corresponding bare module factors. The bare model factors were also obtained from *Turton, et al.* 2018 and *Ulrich 1984* to give a total bare module cost of about \$60.4 million. This value was used to obtain the total module cost by multiplying the value by 1.18, giving a total module cost of about \$73 million. This value was also used as the fixed capital investment cost (FCI) to determine the cost of manufacturing for the project.

what is this called?

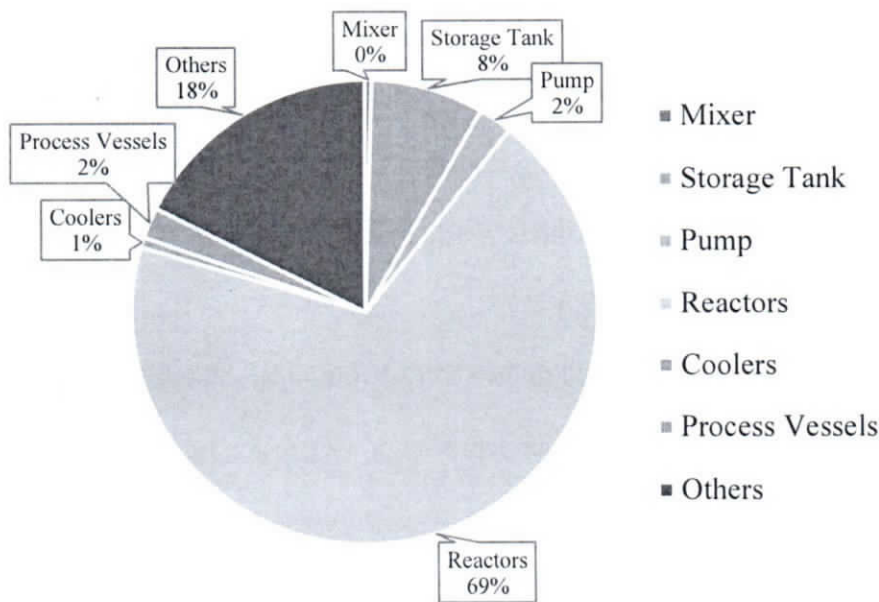


Figure 8.1 Breakdown of Major Equipment Cost

From Figure 8.1 above, the majority of the equipment costs are associated with the reactors in the process. The next major costs are associated with the cost of conveyors, filters, jet cooker, and waste handling with the bulk of this cost deriving from the water treatment plant.

*This section
is weak
on specifics.*

8.1.2 Working Capital

The working capital for this project was obtained from the fixed capital investment. It was assumed that the working capital which involves the costs of storage of raw materials and products, salaries, payable taxes, etc. was 20% of the fixed capital investment value (Turton 248). This value was obtained to be \$14.6 million.

8.1.3 Total Capital Investment

Total Capital Investment (TCI) is the sum of the fixed capital investment and the working capital. For this project, the value of TCI was calculated to be \$88 million.

8.2 Operating Cost

The cost of manufacturing encompasses direct and fixed manufacturing costs as well as general expenses. This was calculated using Equation 8.1 below where C_{OL} , C_{UT} , C_{WT} , C_{RM} represent the cost of operating labor, utilities, waste treatment, and raw materials respectively.

$$COM = 0.280FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) \quad (8.1)$$

The value for this project was calculated to give \$112.5 million and this takes depreciation into account. The cost of manufacturing without depreciation COM_d can be calculated using Equation 8.2 below.

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) \quad (8.2)$$

The value for this project was calculated to be \$105 million and this value was used to generate the cumulative cash flow diagram.

Some main categories of the manufacturing costs are shown in Table 8.2 below. This shows the cost breakdown of the elements that influence the cost of manufacturing.

Table 8.2 Summary of Manufacturing Costs

Item	Cost (\$million/yr)
VARIABLE COST	
Raw materials	50.0
Utilities	15.0
Waste Treatment	1.4
Packaging and Shipping	0.2
Total	66.6
FIXED COST	
Operating labor	3.8
Operating supervision	0.6
Salaries Overhead	2.2
Maintenance and repairs	4.4
Taxes (property) & Insurance	2.3
Land	0.4
Royalties	3.4
Plant overhead, general	5.3
Total	22.4
GENERAL EXPENSES	
Administration	1.7
Distribution & selling	8.7
Research & Development	7.0
Total	17.4

ok.

Cost of Raw Materials

The cost of materials represents the bulk of the variable costs (about 75%) and this value includes the cost of other process consumables such as catalysts, resins, acids, and bases. The main raw material required for this process is corn starch and a unit purchase cost of \$0.46/lb was used (Tate&Lyle). About 157,000 lbs are used annually, therefore, the annual cost of corn starch purchase is \$8 million. The cost of the acids and bases were obtained to be \$16,000. The cost of all catalysts was obtained from the list price of Creative Enzymes Inc. and given to be

\$40.6 million annually. Finally, the resins costs were obtained from the list price of DOWEX resins and given to be \$2 million annually.

Cost of Utility

Table 8.3 gives the factors that contribute to the total annual cost of utilities for the process.

Table 8.3 Summary of Process Utilities

Electricity	\$10,069,488
Steam	\$4,036,890
Water	\$409,591
Waste Treatment	\$1,425,282
TOTAL	\$15,941,251



Electricity represents the greatest of these costs and this is because the power consumption by the evaporator and condenser are high and this is in addition to other electricity requirements through the process. The total annual power usage for the process is 23,982 kW and the power cost is \$0.053/kWh, therefore giving an annual electricity cost to be \$10 million. The steam used in this process is culinary steam because it is required for food-grade production and the unit purchase cost was given to be \$0.004/lb for an annual amount of 940 million lbs. The water cost includes the cost of process water treatment, cooling water for the coolers and vessel jackets, and hot water to give an annual cost of about \$410,000. The cost of waste treatment is derived from the cost of treating waste involved in catalyst and resin regeneration to give an annual cost of \$1.4 million.

Cost of Operating Labor

The cost of operating labor was determined by the number of operators per unit as determined by the Bureau of Labor Statistics. A factor of 4.5 was multiplied by the number of operators needed for each equipment to obtain the total cost of operating labor. For the overall process, a total of

64 operators working 8-hour shifts are needed and this gives an annual cost of operating labor of \$3.8 million.

8.3 Project Analysis

There are three sources of revenue generation for this project, sales of HFCS 42, HFCS 55, and the waste feed from the rotary filter which is sold as wet corn gluten feed. The sales price for both HFCS products were obtained from the United States Department of Agriculture. The sales price for both HFCS 42 and HFCS 55 used to calculate the revenue are the wholesale spot prices for both products as of April 2020 which were 25.5 ¢/lb and 32.75 ¢/lb respectively (USDA). Given an annual output of 333 million pounds for HFCS 42 and 364 million pounds for HFCS 55, the annual revenue generated for each was calculated to be \$85 million for HFCS 42 and \$119 million for HFCS 55. The sales price for the corn gluten feed was taken to be \$0.04/lb and for an annual output of 115 million pounds, the annual revenue generated from this was calculated to be \$4.6 million. This gives a total annual revenue of about \$209 million.

The gross profit obtained was calculated by subtracting the sum of all fixed and variable costs from the main product revenues. The variable costs of production were calculated to be about \$62 million and the fixed costs of production were calculated to be \$ 22.3 million. Since HFCS 42 and HFCS 55 are considered the main products, the total revenue generated from them is \$204 million. The annual gross profit was then calculated to be about \$120 million. The net profit was calculated from this value by subtracting the value of taxes. A tax rate of 48% was used for this project which includes federal, state tax for Iowa, and city tax for Asbury, Iowa. With this value, the annual taxes were calculated to be \$57.5 million and this gave an annual net profit of \$62.3 million.

This is extremely high!!

The non-discounted and discounted cumulative cash flow for the project was generated and is shown in Figure 8.2.

n=20 years	i=0.48	i=0.1	FCI=\$72,972,900	WC=\$14,594,580	Land=\$388,302	R=\$208,756,310				
End of year (k)	Capital Investment	Depreciation Allowance (d)	[FCIL - sum(dk)]	Revenue from sales (R)	Discounted Cost of Manufacturing (COMd)	After-Tax Cash Flow	Nondiscounted Cash Flow	Cumulative Cash Flow	Discounted Cash Flow	Cumulative Discounted Cash Flow
0	(388,302)	-	72,972,900	-	-	-	(388,302)	(388,302)	(388,302)	(388,302)
1	(43,783,740)	-	72,972,900	-	-	-	(43,783,740)	(44,172,042)	(39,803,400)	(40,191,702)
2	(14,594,580)	-	72,972,900	-	-	-	(14,594,580)	(58,766,622)	(12,061,636)	(52,253,338)
3	-	14,594,580	58,378,320	104,378,155	105,234,400	6,560,151	6,560,151	(52,206,471)	4,928,739	(47,324,600)
4	-	23,351,328	35,026,992	187,880,679	105,234,400	54,184,703	54,184,703	1,978,232	37,008,881	(10,315,719)
5	-	14,010,797	21,016,195	208,756,310	105,234,400	60,556,576	60,556,576	62,534,807	37,600,869	27,285,150
6	-	8,391,884	12,624,312	208,756,310	105,234,400	57,859,497	57,859,497	120,394,304	32,660,178	59,945,328
7	-	8,391,884	4,232,428	208,756,310	105,234,400	57,859,497	57,859,497	178,253,802	29,691,071	89,636,399
8	-	4,232,428	-	208,756,310	105,234,400	55,862,959	55,862,959	234,116,760	26,060,483	115,696,881
9	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	287,948,154	22,829,766	138,526,647
10	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	341,779,547	20,754,332	159,280,979
11	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	395,610,940	18,867,575	178,148,554
12	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	449,442,333	17,152,341	195,300,895
13	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	503,273,726	15,593,037	210,893,932
14	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	557,105,120	14,175,488	225,069,420
15	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	610,936,513	12,886,808	237,956,228
16	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	664,767,906	11,715,280	249,671,508
17	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	718,599,299	10,650,254	260,321,762
18	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	772,430,692	9,682,049	270,003,811
19	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	826,262,086	8,801,863	278,805,674
20	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	880,093,479	8,001,694	286,807,368
21	-	-	-	208,756,310	105,234,400	53,831,393	53,831,393	933,924,872	7,274,267	294,081,634
22	14,982,882	-	-	209,144,612	105,234,400	54,033,310	69,016,192	1,002,941,064	8,478,361	302,559,996
TOTALS		\$ 72,972,900		\$ 4,050,260,716		\$ 1,046,724,802			\$ 302,559,998	

Figure 8.2 Cumulative Discounted Cash Flow

A plant life of 20 years was used for this analysis with the first two years for plant construction and start-up. A discount rate of 10% was applied and the Modified Accelerated Cost Recovery System (MACRS) was used to determine the depreciation allowance using a 5-year recovery period. A ramp-up revenue generation was used with the first year of revenue generation running at 50% capacity, the second year of generation running at 90% capacity before 100% capacity is used till the end of the plant life. The cumulative discounted cashflow values gave negative values until year 5 where it became positive. At the end of the plant life, the net present value from the cumulative discounted cash flow was about \$303 million. The cash flow diagram is

represented in Figure 8.3 below. It shows that the break-even point is after year 4 indicating that profit can be made early on in the project.

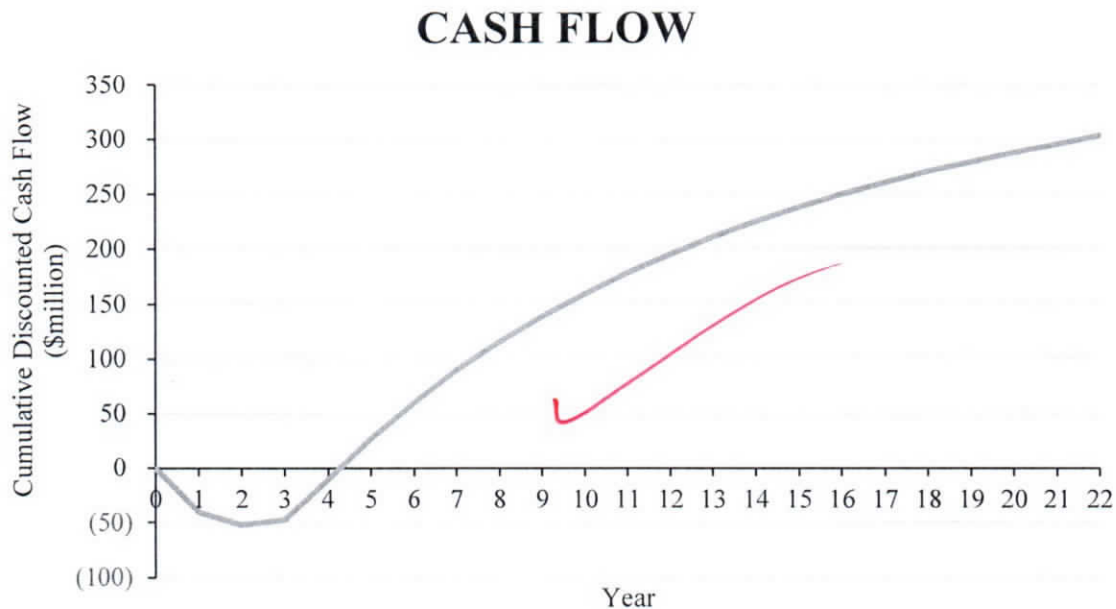


Figure 8.3 Cash Flow Diagram

The present value ratio was calculated to be 6.79 which is much greater than 1 which is an indication that the process is profitable (Turton 271). The return on investment (ROI) was calculated by dividing annual net profit by the fixed capital investment to give a value of 85.3% with a discounted payback period (DPBP) of 2 years. The discounted cash flow rate of return (DCFROR) represents the value discount rate for which the net positive value is 0 and this value was calculated to be 49.6%. A project is considered profitable if the DCFROR is greater than the internal discount rate (Turton 273) and 49.6% is greater than 10%, therefore, this is another indication that this project is profitable. Since the payback period is relatively short for the great return on investment, this project overall is highly profitable.

9. SAFETY AND HAZOP ANALYSIS

9.1 Hazard and Operability (HAZOP) Study

The vessel selected for this HAZOP study is the packed bed reactor (R-103). This reactor is responsible for the conversion of glucose in the incoming glucose syrup to fructose to give High Fructose Corn Syrup containing 42% fructose through an isomerization reaction. Two lines will be considered for this analysis and the details are represented in HAZOP tables. A piping and instrumentation diagram (P&ID) diagram showing the control scheme for the operation of this reactor is represented in Figure 9.1 below.

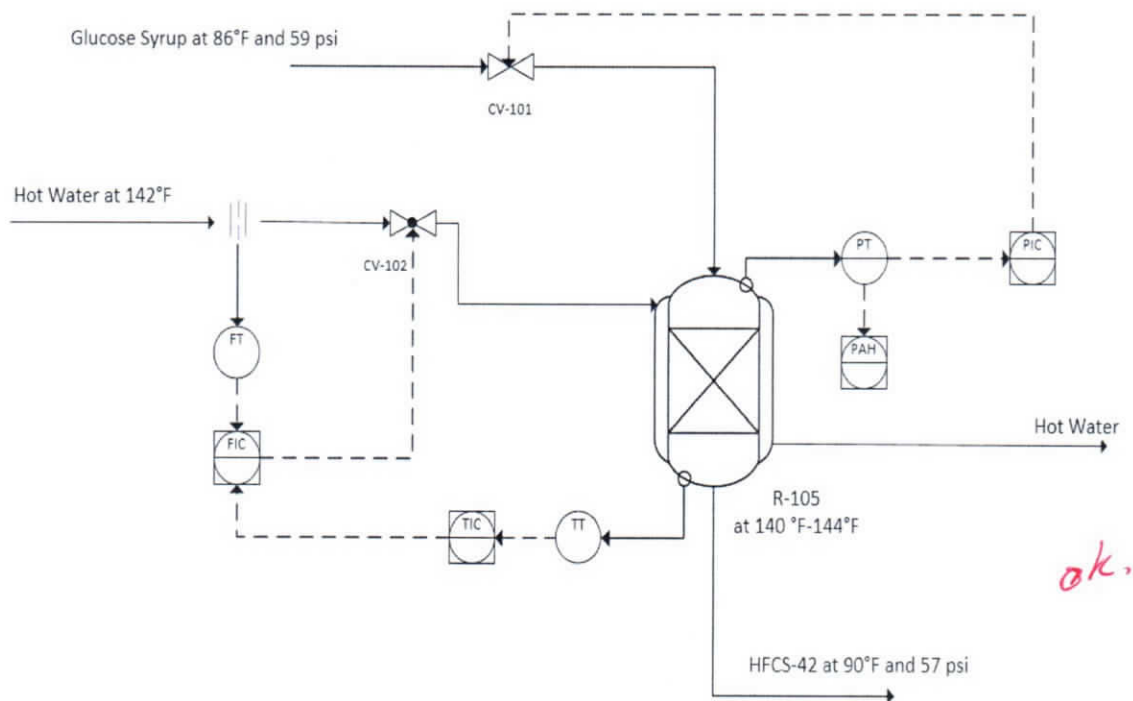


Figure 9.1 Control Scheme for the Packed Bed Reactor (R-103)

why control two parameters??

The control scheme for the packed bed reactor is centered around controlling the internal column pressure and temperature. For the temperature, the desired set point is 140°F which is required for optimum enzyme activity. The temperature is to be controlled by manipulating the flow rate of hot water into the vessel jacket. An orifice measures the flow of the hot water and is connected to the flow transmitter which sends a signal to the flow indicator controller. A temperature transmitter on the vessel sends the temperature reading to the temperature indicator controller which sends the signal to the flow indicator controller as well. The flow indicator controller gives the go-ahead to the globe valve on the hot water feed stream to open or shut depending on if the reactor temperature is too low and more hot water needs to be fed or if the reactor temperature is too high and the flow of hot water is to be stopped or slowed.

For the internal pressure control, the desired set point is 14 psi. The pressure build-up is caused by a blockage in the fluid path through the packed bed. Blockage can be enhanced when the enzyme pellets become crushed due to the higher-pressure condition within the reactor. A pressure transmitter is put in place to record the internal reactor pressure at any given time. The signal is sent to the pressure indicator controller which triggers the gate valve on the glucose syrup feed stream to shut off if the pressure recorded exceeds the setpoint. A high-pressure alarm is also installed as an extra precautionary measure. The reading from the pressure transmitter is also sent to the high-pressure alarm. In the case where the pressure indicator controller or the gate valve is faulty or clogged, the alarm will go off if a pressure around the maximum allowable working pressure for the reactor is sensed. A distributed control system is used for the plant, hence, the gate valve on the glucose feed stream can be automatically manipulated once the high-pressure alarm goes off. The HAZOP analysis is shown below in Table 9.1 and Table 9.2.

Table 9.1 HAZOP Analysis on Line 35 of HFCS Production Plant

Study Node	Keyword	Guide Word	Deviation	Cause	Consequences	Action
Line 35 (incoming Glucose syrup stream)	Flow	No	No flow of glucose syrup through packed bed	Blocked catalyst bed	1. No isomerization reaction 2. Desired product (HFCS 42) not formed 3.	1. Unclog reactor packed bed
	Pressure	More	Pressure in reactor above 14 psi	Blocked catalyst bed	1. Pressure buildup in the reactor 2. Enzymes get crushed leading to more flow blockage 3. Increased chance of an explosion	1. Install pressure transmitter to signal the pressure controller and shut off valve 2. Shut off glucose syrup flow 3. Install high pressure alarm system to alert operator 4. Pressure relief valve discharges 5. Unclog reactor packed bed

Table 9.2 HAZOP analysis on line 55 of HFCS Production Plant

Study Node	Keyword	Guide Word	Deviation	Cause	Consequences	Action
Line 55 (Hot water stream for cooling jacket)	Temperature	High	Temperature in reactor above 158°F	Excessive flow of heating fluid	1. Slower reaction rate due to denaturation of enzyme 2. Lower yield of fructose in product stream 3. Compromised product quality due to enzyme poisoning	1. Install temperature transmitter to record reactor temperature 2. Install flow meter to record hot water flow rate 3. Set up controllers to control valve function
		Low	Temperature in reactor below 140°F	Insufficient flow of heating fluid	1. Compromised product quality due to enzyme poisoning	

9.2 Safety Considerations

One concern in this plant is the accumulation of combustible dust. The causes of these include the release of grain dust during loading, cooling, transferring, and storage. The dust can cause explosions when in contact with an ignition source. Ignition sources range from a hot bearing on the belt conveyor, smoldering materials caught in the dust collection system, to electrostatic discharge from process equipment (Oregon OSHA). Controls to mitigate dust explosions are as follows:

- Use dust collection to collect any dust that may be released in the air.
- Use automatic systems to detect sparks and smoldering materials.
- Equipment prone to electrostatic discharge is properly grounded and bonded.
- A fire detection system in dust collectors and other equipment prone to fire

- Install explosion relief venting system.
- Routine maintenance of ventilation systems
- No smoking in or near this area.
- Keep hot work away from potential dust accumulation.

Furthermore, the handling of hazardous materials such as hydrochloric acid and sodium hydroxide. These materials are corrosive and considered irritants to the skin. The vapors from said materials are toxic to the human body. In order to avoid any potential hazards, ventilation systems are to be put in place. Additionally, proper personal protective equipment must be worn in the process and storage areas where hazardous materials are used or stored (Oregon OSHA).

10. DISCUSSION

10.1 Assumptions for Calculations

To derive the output stream flows for HFCS 42 and HFCS 55, the typical plant capacity was converted to metric tons per hour (41.6 tons/hr) and then to pounds per hour (91,700 lbs/hr). Of this amount, the assumption that 50% of plant production is HFCS 55 and 46% of plant production is HFCS 42 was used. The rest was assumed to be HFCS for miscellaneous use such as in pharmaceuticals or agriculture (Venkatasubramania 142). With the values of the output stream, further calculations were made with including some assumptions to determine the values of the anterior streams. These assumptions, however, will be presented in order of the first unit to the last unit. For the first stirred, it was assumed that a 30% w/w solid solution must be leaving because a higher solid concentration would be harder to move through the pump. It was assumed that a 20% w/w solid solution left the jet cooker as the slurry became more gelatinized due to the contact with culinary steam in the cooker. Calculations for the first stirred tank reactor were

made assuming. The liquefaction product typically has a dextrose equivalent (DE) of 10-15% (Buchholz 494) and 10% was used for calculations. After the liquefaction, the saccharification product was assumed to be a 50% w/w solid solution with a 95% DE (Buchholz 496). It was assumed that all ash was lost in the deashing and decolorization steps although, in reality, a very small amount makes its way to the final HFCS products. The final ash amount is so small it was assumed negligible for calculations in this design. It was also assumed that a 50% w/w solid product exited this unit. After evaporation, it was assumed that the concentrated product had an 80% w/w solid composition. In the packed bed reactor, a conversion of 42% was used for calculations. This value ideally is 50%, however, in practical applications, this value is usually in the range of 90% of the theoretical maximum (Illanes 200). It was also assumed that all streams exiting this unit contained a 75% w/w solid composition. After isomerization, it was assumed that 50% of the HFCS 42 produced went to the storage tank and the other half went to a split point. At this split point, 81.6% of the syrup (HFCS 42) goes straight to the final mixer where it is blended with the HFCS 90 from the chromatography column is to give the HFCS 55. It was assumed that the syrup leaving the final mixer had a solid composition of 75% w/w. These product stream solids composition was based on the specification that commercial-grade HFCS must have a solids concentration of no less than 70% (Schenck 18).

10.2 Discussion

The degrees of freedom analysis for the units give values greater than zero but no more than 5. This is a good range of values to allow for better design flexibility which aids the optimization process after the design has been completed. After the degrees of freedom analysis, an overall material balance for the process was performed. The input streams consist of the water streams (freshwater and process water) and starch stream into the first mixer and the freshwater

did you perform any optimization?

fed to the chromatography column. The output streams consist of the process waste feed from the rotary drum filter and the HFCS 42 and 55 output streams. The input and output totals are 104,030 lb/hr and 103,478 lb/hr and these values are in a very close range with a difference of 0.5%. A possible cause for the slight difference in output and input totals could be the assumptions used in making the calculations. In simulating the process on ASPEN, there were some limitations in component and equipment packages, for example, undefined starch properties as well as the absence of library models to properly simulate processes like the ion exchange and chromatography. To combat this, the flowsheet was truncated using subroutines to obtain a simplified flowsheet that converged. The results of the converged flowsheet in Table 6.1 showed that there was a difference of about 0.62% and 0.65% in the results of the simulated flow rates compared to the hand calculations for HFCS 42 and HFCS 55 respectively. From the economic analysis performed for the plant, a longer plant life, a lower discount rate, and fewer depreciation recovery years altogether give a greater net present value for the plant at the end of its life. The process is generally safe with the most hazardous components being the acid and salt solutions used occasionally for pH control and enzyme activation. Dust explosions from the corn starch could also occur and for these reasons, appropriate precautionary measures are implemented and emphasized such as the installation of an explosion relief system and the use of appropriate personal protective equipment.

11. CONCLUSIONS AND RECOMMENDATIONS

A detailed design and economic analysis of a high fructose corn syrup (HFCS) plant with a capacity of 706 MMlbs/year was accomplished for this project. All equipment was properly designed and sized to accommodate the total yearly capacity. From the detailed design performed on each equipment, the total cost of equipment needed for this project was obtained and that value was used to determine the Fixed Capital Investment (FCI) amount needed to start the project. In addition to this, the yearly capital and operating cost for a plant life of twenty years was calculated. A discounted cumulative cash flow was obtained, and a detailed economic analysis was performed to obtain important information such as the rate of return on investment and the payback period it would take to recover the cost of the initial investment. Lastly, a detailed hazard and operability analysis (HAZOP analysis) was performed over a section of the plant to prevent damage to personnel and equipment, and based on the analysis done, the project poses no threat to human life or damage to equipment. The data obtained from the economic analysis and the current situation caused by the COVID-19 virus both favor this project as a good investment. As the virus declines and the economy begins to boom, the demand for chief products that contain HFCS will begin to rise; ultimately increasing the demand for HFCS and making it easy to break into the market. In conclusion, this project was deemed safe and profitable, therefore, investment in this project is recommended.

*Very good report
overall !!*

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12. APPENDIX

12.1 Plant Design Basis

Plant Capacity

Design Capacity: 706 MM lbs per year (1.5 MM metric tons)

Design Operating Time: 330 days per year

Raw Materials

Corn Starch (white solid powder), may form combustible dust concentrations in air

Product Glucose-Fructose Syrup Specifications

HFCS 42 & 55, clear viscous liquid, sweet odor, pH 5.9, melting/freezing point 146.1°C

Utility Specifications for Enzymatic Reactors

Thermostable α - amylase, optimum pH of 6.5 at 105 °C

Glucoamylase, optimum pH of 4.5 at 60 °C

Immobilized glucose isomerase, optimum of pH 7.5 at 60 °C

Ambient Conditions

Maximum Temperature: 95 °F

Minimum Temperature: -10 °F

Relative Humidity: 25-95%

Maximum Windspeed: 30 mph

12.2 Sample Calculations

12.2.1 Example of Material Balance Calculation

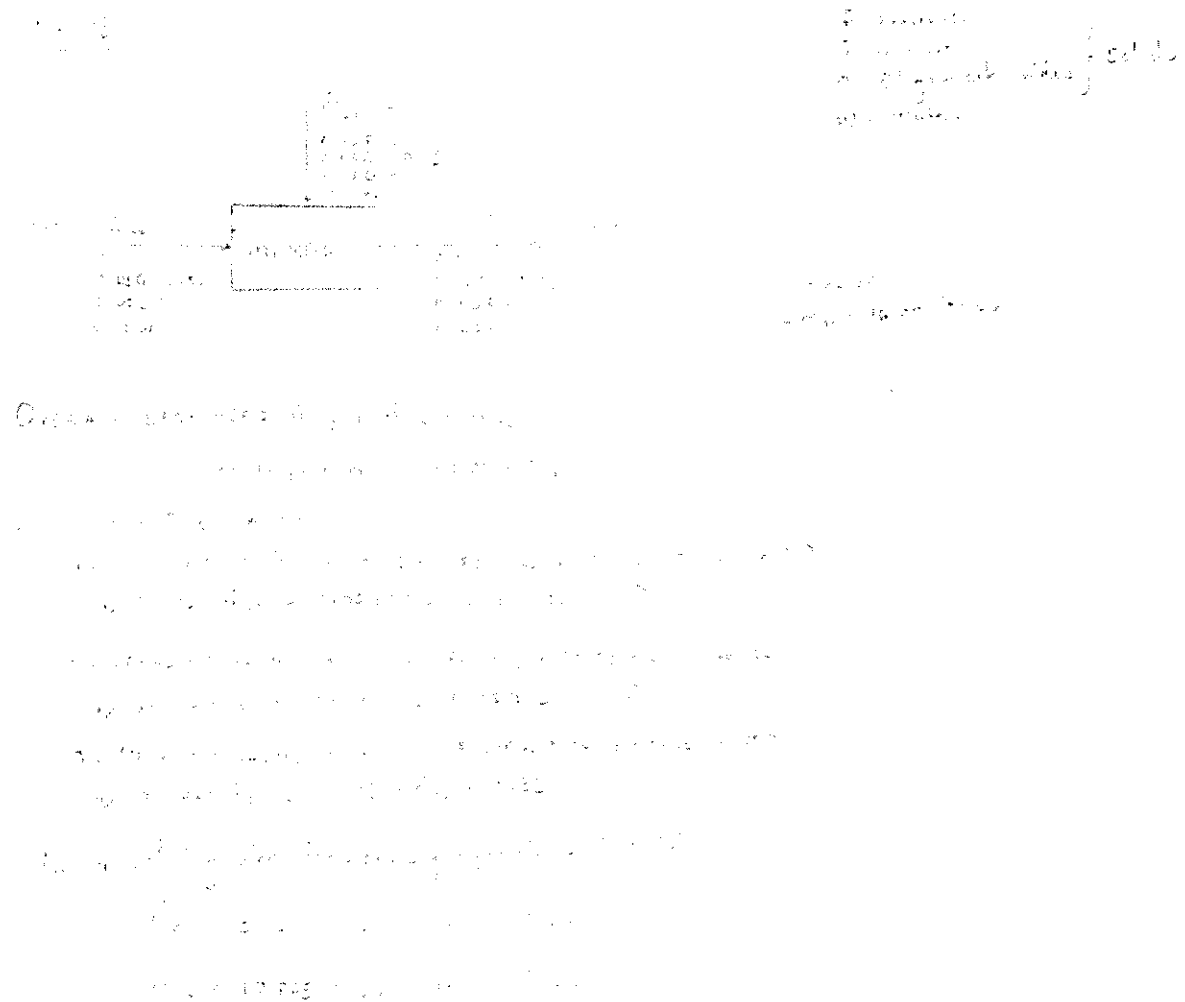


Figure 12.1 Example Material Balance Hand Calculation for Mixer 3

12.2.2 Example of Equipment Design Calculations

* To find the heat duty required to heat the reactor, the overall heat transfer equation by combined conduction & convection is used

$$Q = \frac{T_1 - T_4}{\frac{1}{h_i A_i} + \frac{r_o - r_i}{k_r A_r} + \frac{1}{h_o A_o}} \quad \text{--- (1)}$$

$$Q = U_i A_i (T_1 - T_4) = U_o A_o (T_1 - T_4) \quad \text{--- (2)}$$

* To obtain h_i and h_o the eq below is used

$$h = 1422 (1 + 0.0146 T^{0.2}) \frac{1}{L^{0.2}}$$

The speed (V) of the heating fluid is obtained using Reynolds number formula for a Reynolds number $\approx 9 \times 10^5$

$$h_i = 273 \text{ W/m}^2\cdot\text{K}$$

$$h_o = 353 \text{ W/m}^2\cdot\text{K}$$

* k_r for stainless steel 316 = $15 \text{ W/m}^2\cdot\text{K}$

$$* A_{rw} = \frac{A_o - A_i}{\ln(A_o/A_i)}$$

$$* T_1 = 60^\circ\text{C} = 140^\circ\text{F}$$

$$T_4 = 50^\circ\text{C} = 122^\circ\text{F}$$

$$\text{Finally ; } \boxed{Q = 10152 \text{ W} = 34639 \text{ Btu/h}}$$

$$\boxed{U_i = U_o = 52 \text{ W/m}^2\cdot\text{K} = 9.16 \text{ Btu/h}\cdot\text{ft}^2\cdot^\circ\text{F}}$$

* To obtain the pressure drop across the filter Ergun's equation is used

$$P = \frac{P_2}{P_0} = (1 - \alpha v)^{1/2}$$

$$\text{where: } \alpha = \frac{2 \beta_0}{A (1 - \phi) \rho_0 \phi^3}$$

$$\text{and } \beta_0 = \frac{G (1 - \phi)}{\rho_0 \phi_0 \phi^3} \left[\frac{150 (1 - \phi) \mu}{D_p} + 1.75 G \right]$$

$$G = P \mu = \frac{P \dot{V}_0}{A} = \left(3170.3 \frac{\text{lb}}{\text{m}^3} \right) \left(\frac{22.44 \text{ m}^3/\text{hr}}{1.65 \text{ m}^2} \right) = 52723.05 \frac{\text{lb}}{\text{m}^2 \cdot \text{hr}}$$

$$\phi = 0.3$$

$$\mu = 1.8 \text{ cP} = 4.35 \frac{\text{lbm}}{\text{ft} \cdot \text{s}}$$

$$\rho_0 = 3170.3 \frac{\text{lb}}{\text{m}^3}$$

$$D_p = 3 \text{ mm}$$

$$G_c = 4.17 \times 10^8 \frac{\text{lbm} \cdot \text{ft}}{\text{lb}_f \cdot \text{hr}^2}$$

$$P_0 = 4 \text{ atm}$$

Using all these information and solving;

$$\beta_0 = 0.029 \frac{\text{atm}}{\text{m}}$$

$$\alpha = 7.42 \times 10^{-6} \text{ cm}^{-1}$$

$$P = 0.97$$

$$P_2 = 3.88$$

$$\Delta P = P_0 - P_2 = 0.12 \text{ atm} = 1.8 \text{ psi}$$

* From headwater obtained for a PBR, an additional 3.2 ft is added to the bed height to obtain the vessel height

* adding 1.5 ft to the top and bottom of the bed height

$$\text{Resulting vessel height} = 12.2 \text{ ft} \approx 19 \text{ ft}$$

* Calculating the thickness to get the vessel diameter and volume

$$t = \frac{P_i d_i}{2\sigma(1 - P_i)} + C ; \quad \begin{aligned} P_i &= 0.10325 \text{ MPa} \\ d_i &= 1500 \text{ mm} \\ \sigma &= 136.637 \text{ MPa} \\ P_i &= 0.25 \\ C &= 7 \text{ mm} \end{aligned}$$

$$t = 8 \text{ mm} = 0.008 \text{ m} = 0.03 \text{ ft}$$

$$* \quad D_o = 2(0.03 \text{ ft}) + 12.2 \text{ ft} = 12.26 \text{ ft} \quad \text{--- outer diameter}$$

$$* \quad \text{Total volume} = \pi r_o^2 h = 477 \text{ ft}^3$$

Bed dimensions

$$V = 280 \text{ ft}^3$$

$$h = 15.28 \text{ ft}$$

$$D_i = 4.8 \text{ ft}$$

Vessel dimensions

$$V = 477 \text{ ft}^3$$

$$h = 19 \text{ ft}$$

$$D_o = 12.26 \text{ ft}$$

Starting with the design equation

$$\text{residence time } (\tau) = \frac{\text{Volume } (V)}{\text{volumetric flowrate } (Q_0)}$$

for a residence time of 4 hours and a volumetric flowrate of 7334 U.S. gal/hr ($27.44 \text{ m}^3/\text{hr}$)

$$V = 4 \text{ hours} \left(\frac{7334 \text{ U.S. gal/hr}}{1} \right) = 29,336 \text{ U.S. gal/hr}$$

$$V = 29,336 \text{ U.S. gal/hr} = 3220 \text{ ft}^3$$

According to a reference obtained from research, a plant producing 1000 tons/day of PEGS uses 20 reactors in parallel.

For this design 14 reactors would be used; this would give a volume of 220 m^3 for one reactor

* From literature, packed bed reactors usually have a $\frac{L}{D}$ ratio of 0.5

$$V = \frac{\pi D^2 L}{4} \quad \text{where } L = 0.5D$$

$$\text{Solving } ; \quad L = 15.22 \text{ ft}$$

$$D = 4.8 \text{ ft}$$

* Therefore ; for one reactor the bed dimensions are

$$\left. \begin{array}{l} V = 220 \text{ ft}^3 \\ L = 15.22 \text{ ft} \\ D = 4.8 \text{ ft} \end{array} \right\} \begin{array}{l} \text{This is the volume} \\ \text{occupied by the catalyst} \end{array}$$

12.3 Tables of Data Used

Table 12.1 Relative Sweetness of Sugars

Sugar or sweetener	Relative sweetness
Sucrose	1.0
Invert sugar	0.85 - 1.0
Fructose	1.3
Glucose	0.56
Galactose	0.4-0.6
Maltose	0.3-0.5
Lactose	0.2-0.3
Xylitol	1.01
Cyclamates	30-80
Acesulfame K (Sunnette®)	200
Aspartame (Equal®, Nutrasweet®)	100-200
Saccharin (The Pink Stuff)	200-300
Stevioside	300
Sucralose (Splenda®)	600
Thaumatococin (Talin®)	2000-3000

Source: Parker, Kay, et al. *Biotechnology and Molecular Biology Review*, Dec. 2010, pp. 71–78.

Table 12.2 Nutrition Values per 100g HFCS

Nutritional Items	Value
Energy	1,176 kJ (281 kcal)
Carbohydrates	76 g
Dietary fiber	0 g
Fat	0 g
Protein	0 g
Water	24 g
Riboflavin (Vitamin B 2)	0.019 mg (1%)
Niacin (Vitamin B3)	0 mg (0%)
Pantothenic acid (Vitamin B5)	0.011 mg (0%)
Vitamin B6	0.024 mg (2%)
Folic acid (Vitamin B9)	0 µg (0%)
Vitamin C	0 mg (0%)
Calcium	6 mg (1%)
Iron	0.42 mg (3%)
Magnesium	2 mg (1%)
Phosphorus	4 mg (1%)
Potassium	0 mg (0%)
Sodium	2 mg (0%)
Zinc	0.22 mg (2%)

Source: Parker, Kay, et al. *Biotechnology and Molecular Biology Review*, Dec. 2010, pp. 71–78.

<i>Calendar Year</i>	<i>Sugar^a</i>	<i>HFCS 42^b</i>	<i>HFCS 55^b</i>
2000	42.41	12.76	15.72
2001	43.42	13.66	17.05
2002	43.10	15.76	18.50
2003	42.68	16.38	19.07
2004	42.64	17.98	20.54
2005	43.54	18.62	21.12
2006	49.58	19.35	21.77
2007	51.48	23.41	25.49
2008	52.91	27.57	29.32
2009	57.03	31.51	32.95

^a USDA 2010, Table 23.6—US retail refined sugar price, cents per pound

^b USDA 2010, Table 23.9—US prices for high fructose corn syrup, wholesale list price, cents per pound dry weight

Figure 12.2 Relative Sugar Prices from 2010 United States Department of Agriculture (USDA) Report

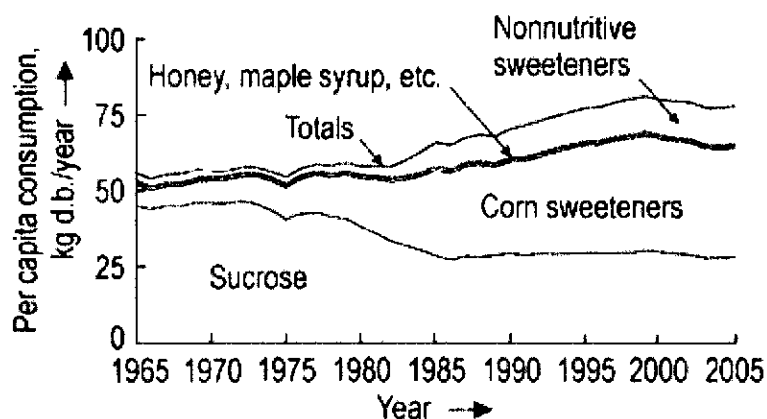


Figure 12.3 Consumption Trend of Various Syrups from Schenck, Fred W. *Ullmann's Encyclopedia of Industrial Chemistry*, 2006

90 D.E. Corn
(Sugar) Syrup

Commercial Baume	Percent dry Substance by Wt.	Temperature °F				
		40	80	120	160	200
		<u>BTU/lb °F</u>	<u>BTU/lb °F</u>	<u>BTU/lb °F</u>	<u>BTU/lb °F</u>	<u>BTU/lb °F</u>
-1.0	0.0	1.004	.998	.998	1.001	1.005
0.0	1.8	0.994	.989	.990	.993	.998
5.0	10.8	.938	.939	.944	.950	.958
10.0	19.8	.881	.889	.898	.908	.919
15.0	29.0	.825	.840	.853	.866	.879
20.0	38.3	.770	.790	.807	.822	.837
25.0	47.7	.716	.739	.759	.777	.794
30.0	57.3	.662	.689	.712	.732	.750
32.0	61.2	.641	.669	.693	.713	.731
34.0	65.2	.620	.649	.674	.694	.713
36.0	69.2	.600	.630	.655	.675	.694
38.0	73.2	.582	.611	.635	.657	.675
40.0	77.3	.564	.593	.617	.638	.655
41.0	79.4	.556	.583	.607	.628	.646
42.0	81.5	.547	.574	.598	.618	.636
43.0	83.6	.539	.565	.588	.607	.625
44.0	85.7	.532	.556	.578	.597	.615
45.0	87.8	.525	.548	.568	.587	.604
46.0	89.9	.517	.539	.558	.577	.594

Figure 12.4 90 DE Corn Syrup Heat Capacity from Corn Refiners Association, Inc., 2018

12.4 Product Information Sheet

Product Information



DOWEX™ MONOSPHERE™ 77

Ion Exchange Resin for Sweetener Applications

Product	Type	Matrix	Functional group
DOWEX™ MONOSPHERE™ 77	Weak base anion	Styrene-DVB, macroporous	Tertiary amine

Typical Physical and Chemical Properties

Ionic form as produced		FB (free base)
Total exchange capacity, min.	eq/L	1.7
Weak base capacity, min.	eq/L	1.5
Water content	%	40 - 50
Bead size distribution		
Volume median diameter	µm	475 - 600
Total swelling (FB → HCl)	%	22
Whole uncracked beads, min.	%	95
Particle density	g/mL	1.04
Shipping weight**	g/L lbs/ft ³	640 40

60°C (140°F)

Recommended

0 - 7

Operating • Conditions

- Maximum operating temperature 91 cm (3 ft)
- (H⁺ form) pH range Bed depth, min.
- Flow rates:
 - Service 2 - 4 bed volumes/hour See Figure 1 30 - 45 min.
 - Backwash 30 - 45 min.
 - Regeneration time 2 - 10 bed volumes /hour
 - Displacement rinse
 - Fast rinse (if applicable)
- Total rinse requirement 3 - 5 bed volumes

Regenerants	NaOH [†]	Na ₂ CO ₃	NH ₄ OH
Concentration (%)	4	5	5
Level, 100% basis ^{††}			
lbs/ft ³	4 - 5	6 - 7	4 - 5
kg/m ³	64 - 80	96 - 112	64 - 80
Temperature, max.	60°C (140°F)	60°C (140°F)	60°C (140°F)

[†] Recommended

^{††} Regeneration level may be lower for counter-current regeneration systems.

^{**} As per the backwashed and settled density of the resin, determined by ASTM D-2187.

Page 1 of 2
01710-0209

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Typical Properties DOWEX[™] MONOSPHERE[™] 77 resin is the newest in the DOWEX line of weak base anion **and Applications** resins made using a Dow-patented process which produces beads with remarkable size

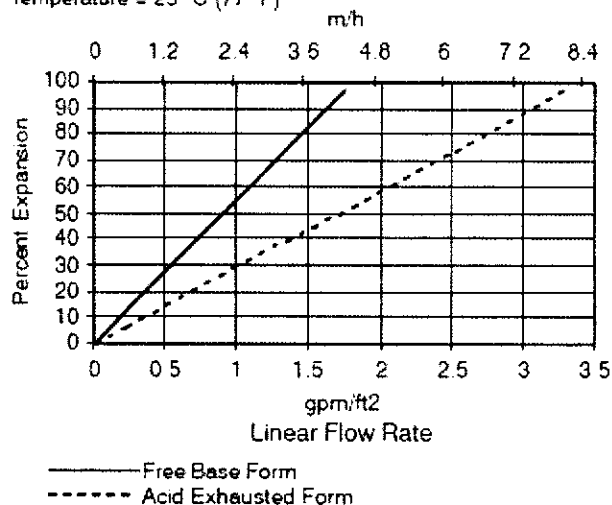
uniformity. Chemically & physically optimized for syrup processing, DOWEX

MONOSPHERE 77 provides an improved balance of high operating capacity, excellent physical strength, economical regeneration, long resin life and low operating costs.

Packaging 25 liter bags, 5 cubic feet fiber drums or 1 cubic meter super sacks.

Figure 1. Backwash Expansion Data

Temperature = 25° C (77° F)



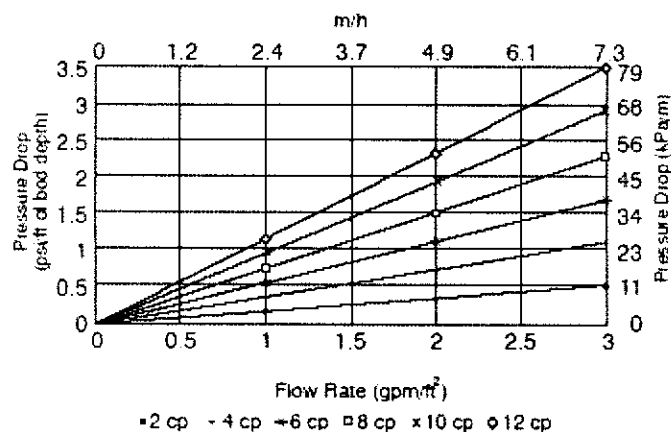
For other temperatures use:

$$F_T = F_{77^\circ F} [1 + 0.008 (T_F - 77)], \text{ where } F \equiv \text{gpm/ft}^2$$

$$F_T = F_{25^\circ C} [1 + 0.008 (1.8T_C - 45)], \text{ where } F \equiv \text{m/h}$$

Figure 2. Pressure Drop Data

Temperature = 20° C (68° F)



For other temperatures use:

$$P_T = P_{20^\circ C} / (0.026 T_C + 0.48), \text{ where } P \equiv \text{bar/m}$$

$$P_T = P_{68^\circ F} / (0.014 T_F + 0.05), \text{ where } P \equiv \text{psi/ft}$$

DOWEX™ Ion Exchange Resins

For more information about DOWEX resins, call the Dow Water Solutions business:

North America: 1-800-447-4369
Latin America: (+55) 11-5188-9222
Europe: (+32) 3-450-2240
Pacific: +60 3 7958 3392
Japan: +813 5460 2100 China: +86 21 2301 9000

<http://www.dowwatersolutions.com>

Warning: Oxidizing agents such as nitric acid attack organic ion exchange resins under certain conditions. This could lead to anything from slight resin degradation to a violent exothermic reaction (explosion). Before using strong oxidizing agents, consult sources knowledgeable in handling such materials.

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Lenntech

DOWEX™ MONOSPHERE™ 88

Ion Exchange Resin for Sweetener Applications

Product	Type	Matrix	Functional group
DOWEX™ MONOSPHERE™ 88	Strong acid cation	Styrene-DVB, macroporous	Sulfonate

Typical Physical and Chemical Properties

Ionic form as produced		Na ⁺
Total exchange capacity, min.	eq/L	1.8
Water content	%	42 - 50
Bead size distribution		
Volume median		500 - 600
diameter 400 - 720 µm, min.	µm %	95
Total swelling (Na ⁺ → H ⁺)	%	5
Whole uncracked beads, min.	%	95
Particle density	g/mL	1.2
Shipping weight**	g/L lbs/ft ³	800 50

Recommended	• Maximum operating temperature (H ⁺ form)	93°C (200°F)
Operating Conditions	• pH range	0 - 14
	• Bed depth, min.	91 cm (3 ft)
	• Flow rates:	
	Service	2 - 4 bed volumes/hour
	Backwash	See Figure 1
	Regeneration time	30 - 45 min.
	Displacement rinse	30 - 45 min.
	Fast rinse (if applicable)	2 - 10 bed volumes /hour

- Total rinse requirement 2 - 5
bed volumes
- Regenerant:

Concentration	7% HCl
Level, 100% basis [†]	5 - 6 lbs/ft ³ 80 - 96 kg/m ³
Temperature, max.	93°C (200°F)

[†] Regeneration level may be lower for counter-current regeneration systems.

** As per the backwashed and settled density of the resin, determined by ASTM D-2187.

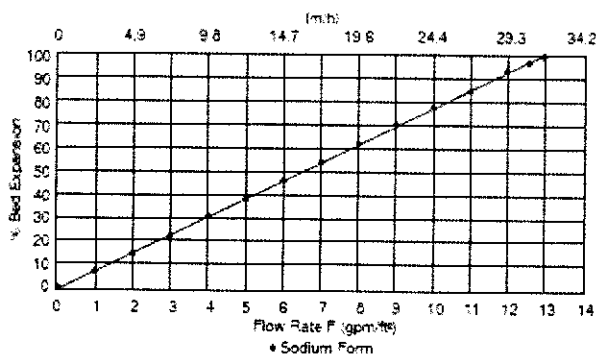
Page 1 of 2
01685-0209

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Typical Properties DOWEX™ MONOSPHERE™ 88 resin is a strong acid cation resin made using a Dow and **Applications** patented process which produces beads with remarkable size uniformity. Chemically optimized for syrup processing, they provide an ideal balance of high operating capacity, excellent physical strength, economical regeneration, long resin life and low operating costs.

Packaging 25 liter bags, 5 cubic feet fiber drums or 1 cubic meter super sacks.

Figure 1. Backwash Expansion Data

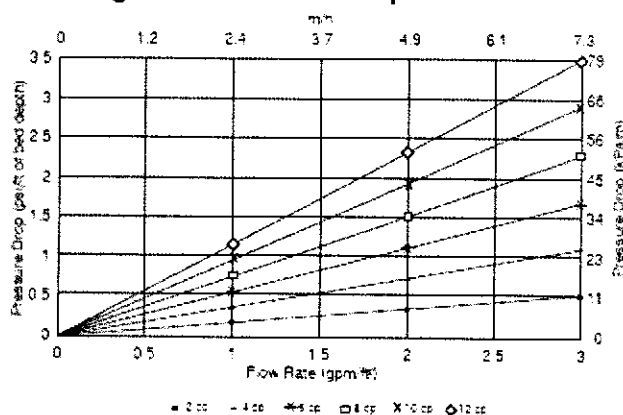


For other temperatures use:

$$F_T = F_{77^\circ\text{F}} [1 + 0.008 (T_F - 77)], \text{ where } F \equiv \text{gpm/ft}^2$$

$$F_T = F_{25^\circ\text{C}} [1 + 0.008 (1.8T_C - 45)], \text{ where } F \equiv \text{m/h}$$

Figure 2. Pressure Drop Data



For other temperatures use:

$$P_T = P_{20^\circ\text{C}} / (0.026 T_C + 0.48), \text{ where } P \equiv \text{bar/m}$$

$$P_T = P_{68^\circ\text{F}} / (0.014 T_F + 0.05), \text{ where } P \equiv \text{psi/ft}$$

Warning: Oxidizing agents such as nitric acid attack organic ion exchange resins under certain conditions. This could lead to anything from slight resin degradation to a violent exothermic reaction (explosion). Before using strong oxidizing agents, consult sources knowledgeable in handling such materials.

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DOWEX™ MONOSPHERE™ 99 Ca/320

Separation Resin for Chromatographic Separations (Primarily Sugars and Sugar Alcohols)

Product	Type	Matrix	Functional group
DOWEX™ MONOSPHERE™ 99 Ca/320	Strong acid cation	Styrene-DVB, gel	Sulfonate

Typical Physical and Chemical Properties

Ionic form as delivered			Ca ⁺⁺
Total capacity, min.		eq/L	1.5 (H ⁺ form)
Water content		%	57 - 61 (H ⁺ form)
Whole uncracked beads, min.		%	98
Particle density		g/mL	1.27
Apparent bulk density:**			
Backwashed, settled and drained	g/L	lbs/ft ³	770 - 820
Tapped and settled		g/L	48 - 51
		lbs/ft ³	785 - 850
			49 - 53

Bead Size Distribution (Light Obscuration Instrument Particle Size)

		Microns
Volume median diameter		300 - 330
Broad range	min. 90%	285 - 350
Narrow range	min. 75%	300 - 335
Fine fines	max. 1%	< 260
Coarse fines	max. 4%	< 280
Fine coarse	max. 4%	> 380
Coarse coarse	max. 1%	> 450

** As per the backwashed and settled density of the resin, determined by ASTM D-2187.

Figure 1. Backwash Expansion Data

Temperature = 25° C (77° F)

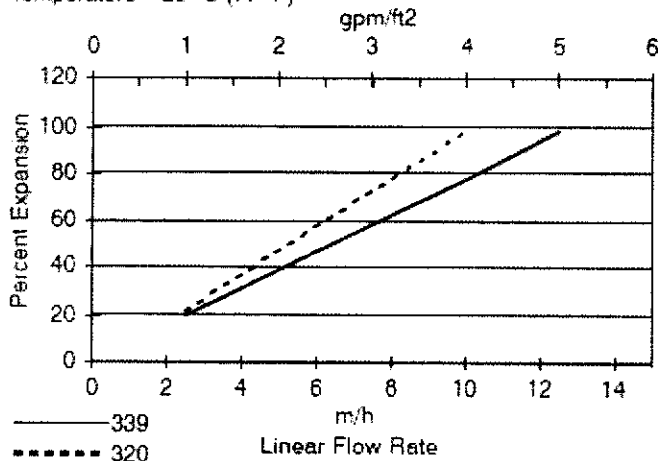
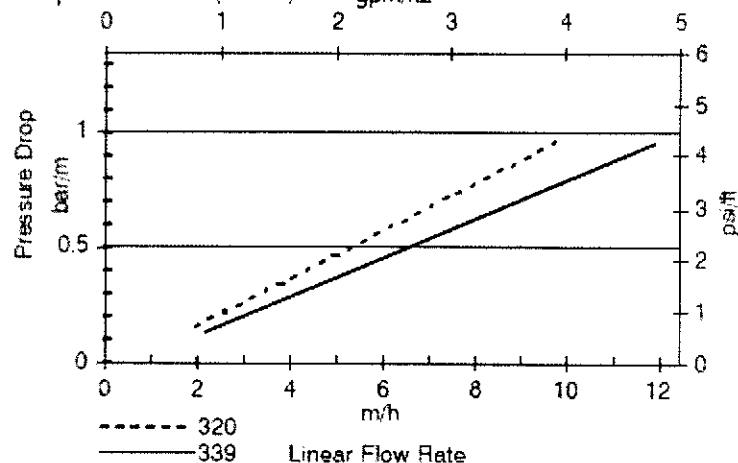


Figure 2. Pressure Drop Data

PRESSURE DROP - 42% Fructose @ 50% DS

Temperature 60° C (140° F)



For other temperatures use:

$$F_T = F_{77°F} [1 + 0.008 (T_F - 77)], \text{ where } F \equiv \text{gpm/ft}^2$$

$$F_T = F_{25°C} [1 + 0.008 (1.8T_C - 45)], \text{ where } F \equiv \text{m/h}$$

DOWEX™ Ion Exchange Resins

For more information about

DOWEX resins, call the Dow Water Solutions business:

North America: 1-800-447-4369

Latin America: (+55) 11-5188-9222

Europe: (+32) 3-450-2240

Pacific: +60 3 7958 3392

Japan: +813 5460 2100 China:

+86 21 2301 9000

<http://www.dowwatersolutions.com>

Warning: Oxidizing agents such as nitric acid attack organic ion exchange resins under certain conditions. This could lead to anything from slight resin degradation to a violent exothermic reaction (explosion). Before using strong oxidizing agents, consult sources knowledgeable in handling such materials.

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DOWEX™ OPTIPORE™ SD-2 Adsorbent

Polymeric Adsorbent for Sucrose and Other Sweetener Applications

Applications surface area

DOWEX™ OPTIPORE™ SD-2 adsorbent is an adsorbent with a high specific

and high porosity which shows exceptional mechanical, thermal and chemical stability. It is specially designed for decolorization as well as taste and odor removal in sweetener applications and complies with the U.S. Food, Drug and Cosmetic Act as amended under Food Additive Regulation 21 CFR 173.25.

This adsorbent has similar pore size distribution and adsorption properties to activated carbon so that it can be used as a direct replacement for carbon in many sweeteners applications. DOWEX OPTIPORE SD-2 has the additional advantage of containing macropores which improve the bulk movement of solutions in and out of the bead. The resulting improvement in kinetics leads to faster production flow rates and smaller beds.

The adsorbent is lightly functionalized with weak base groups to provide a hydrophilic character giving the adsorbent good wettability and compatibility with acid and base regenerants. Regeneration requires only dilute caustic, dilute acid, and hot water.

More detailed information on the use of DOWEX OPTIPORE SD-2 adsorbent in your particular application can be obtained from your Dow technical contact or from your Dow sales representative.

Product Specifications

Product	Type	Matrix	Functional group
DOWEX OPTIPORE SD-2 Adsorbent	Adsorbent	Macroporous styrene-divinylbenzene copolymer	Tertiary amine

General Properties

Dry weight capacity	0.8 meq/g
Typical surface area	800 m ² /g
Average pore diameter	50 Angstrom

Water content (%)	50 - 62
Particle size:	
> 1,190 μm	$\leq 2\%$
< 297 μm (total)	$\leq 3\%$
< 210 μm	0.3% max.
< 105 μm	0.0% max.
	5%
	90%
Density	1.04 g/mL
Shipping weight**	670 g/L (42 lb/ft ³)

** As per the backwashed and settled density of the resin, determined by ASTM D-2187.

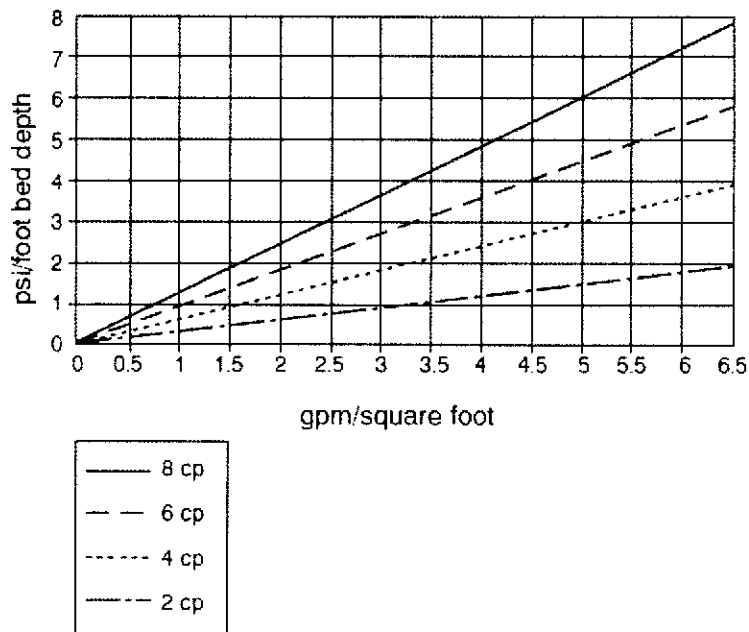
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01666-0209

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Typical Operating Condition Ranges (dependent on application)

Operating temperature	50 - 85°C
Operating pH	3 - 8.5
Service/regeneration flow rates	2 - 6 BV/hr
Regenerant type	NaOH, HCl or H ₂ SO ₄

Figure 1. Pressure Drop



[11] Patent number: 4,950,332
[45] Date of patent: August 21, 1990

[57] Abstract

A process for decolorizing an aqueous sugar solution by contacting the solution with an absorbent resin, wherein the adsorbent resin is a macroporous copolymer being post-crosslinked in a swollen state in the presence of a Friedel-Crafts catalyst and functionalized with hydrophilic groups prior to contact with aqueous sugar solution.

[75] Inventors

Richard T. Springfield, H. Robert Goltz, Seth I. Norman, Upen J. Bharwada, Robert L LaBrie, all of Midland, Mich.

[73] Assignee

The Dow Chemical Company – Midland, Mich.

DOWEX™ Ion Exchange Resins

For more information about

DOWEX resins, call the Dow Water Solutions business:

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+86 21 2301 9000

<http://www.dowwatersolutions.com>

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AVERAGE
PROJECT REPORT

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 2: An ability to apply engineering design to produce solutions that meet specified needs with consideration of public health, safety, and welfare, as well as global, cultural, social, environmental, and economic factor. Assignments that test competency in this area will include problems testing students' understanding of the design process involving devising a system, component, or process to meet desired needs and specifications within constraints, through an iterative, creative, decision-making process in which the basic sciences, mathematics, and engineering sciences are applied to convert resources into solutions.

Name: Group 4

Title of Assignment: Final Report Due Date 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to Define/Understand the Problem and then Plan the Project	60	45 45
(i) Identify the customer and the needs.	8	6 6
(ii) Identify and list the design objectives.	8	6 6
(i) Identify the design constraints.	4	2 1
(ii) Define the design strategy and methodology.	4	2 1
(iii) Identify and break down work into tasks and subtasks and identify the personnel and deliverables for each.	20	20 20
(iv) Develop a Gantt chart and critical path analysis for managing the project.	8	6 6
(v) Establish major milestones for tracking progress and define performance metrics to measure success	8	5 5
2. Ability to Conduct a Review of the Literature	80	60 60
(i) Identify the types of information needed for a complete understanding of all aspects of the project (Based on task described in the project planning).	20	20 20
(ii) Gather information on relevant fundamentals, theory / concept (demonstrate technical competence) and relate them to the design.	40	30 30
(iii) Provide the sources in a list of references properly cited in the literature review section and relevant sections of the report.	20	10 10
3. Ability to perform preliminary design by generation of ideas and applying creativity through generation of ideas and evaluation of candidate solutions	60	45 45
(i) Perform process design and analysis with design constraints; Sorting/screening of process technology alternatives based on availability and cost of raw materials, reaction selectivity and yield, thermodynamic barriers to separating undesired components, and safety and operability	40	30 30
(ii) Evaluate candidate solutions to arrive at feasible designs	20	15 15
4. Ability to Perform Preliminary and Detailed Design	200	175 175
(i) Prepare a process flow diagram (PFD) and present a detailed description of the process indicating the functions of the major equipment components and controls, stream connectivity, safety, and other technical matters of importance	20	15 15
(ii) Carry out a HYSYS/ASPEN simulation of the process showing a converged HYSYS/ASPEN model of the process. State assumptions/parameters used in modeling major equipment	20	20 20
(ii) Prepare a chemical engineering design of (i) reactor(s), (ii) absorption column(s) and (ii) product distillation units, stating the materials of construction and the operating temperatures and pressures	60	55 55
(iii) Prepare mechanical engineering design of the reactor(s) and distillation columns with appropriate dimensioned sketches	20	15 15
(iii) Identify applicable codes and standards for the chemical and mechanical engineering designs. Indicate assumptions made and their justification, and support calculations should be in consistent units, either English Units or SI. Indicate possible sources of error	20	15 15
(iv) Economic Evaluation: Estimate capital costs including engineering, materials, installation, interest, return on investment, taxes and contingency. Total unit conversion cost of product, a sensitivity study on the price of the raw material, and the economic methods used should be included	30	25 25
(v) Prepare a control scheme to ensure safe operation of the plant and carry out Safety Reviews; Hazards, operability and inherent safe (including the use of Safety Data Sheets submitted in the Appendix) or HAZOP analysis	30	30 30
Total	400	325 325

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 3: An ability to communicate effectively with a range of audiences (Written Technical Communication)

Assignments that test competency in written communication will include problems that test the student's ability to logically arrange information, use correct grammar, use appropriate format, use discipline specific conventions including citations, and use acceptable graphical conventions.

Name: Group 4

Title of Assignment: Final Report Due Date: 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to prepare an executive summary for the report	10	10
Students are able to prepare an executive summary of 5 to 8 pages	10	10
2. Ability to organize, plan and properly format a written technical report	80	80
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	20
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	20
(ii) Students provide Title Page, and Table of Contents, list of Figures, and List of Tables properly formatted.	12	20
(iii) Students provide figure number and title for each figure in the report, reference each figure, and completely discuss each figure in the report in accord with standards in the project manual.	12	12
(iv) Students provide table number and title for each table in the report in accord with standards in the project manual, reference each table, and completely discuss each table in the report.	12	12
(v) Students properly cite references in the report and provide well formatted reference list at the end.	12	12
(vi) Students prepare the written report in accord with standard report formatting provided in the Senior Projects Report Manual.	12	12
3. Ability to compose original texts and properly apply the conventions of written language	50	50
(i) Properly apply capitalization, punctuation, and penmanship, to communicate clearly.	15	15
(ii) Spell proficiently	10	10
(iii) Apply standard grammar and usage to communicate clearly and effectively in writing including • using complete sentences, varying the types such as compound and complex to match meanings and purposes • properly employing standard English usage in writing for audiences, including subject-verb agreement, pronoun referents, and parts of speech • properly using adjectives (comparative and superlative forms) and adverbs appropriately to make writing vivid or precise • properly using prepositional phrases to elaborate written ideas • properly using conjunctions to connect ideas meaningfully	20	20
(vi) Use available technology to support aspects of creating, revising, editing, spell checking, and publishing the report.	5	5
4. Ability to provide appropriate discussion, conclusions and recommendations	60	40
(i) Summarize the goals, objectives, and indicate whether they were met.	10	8
(ii) Summarize the results.	30	20
(iii) Summarize constraints and codes and indicate whether they were met.	10	6
(iv) Provide logical conclusions and recommendations (including strengths and weaknesses).	10	6
Total	200	180

“Plant Design for the Production of Ethylene Oxide”

Senior Design and Professionalism II
Final Report

Submitted By

Leah McClellan ✓

Natalie Russell ✓

Chloe Sanders ✓

Francis Saneii ✓

Ariel Taylor ✓

To

Dr Lealon Martin
Senior Design Instructor

Roy G Perry College of Engineering
Department of Chemical Engineering
Prairie View A&M University
Prairie View, Texas 77446

May 6th, 2020

LETTER OF TRANSMITTAL

May 6th, 2020

Department of Chemical Engineering
Prairie View A&M University
100 University Dr.
Prairie View, TX 77446

Greetings Dr Martin,

This group has completed the spring requirements for designing a grass roots ethylene oxide plant. After completion of the fall semester requirements, this group was able to successfully complete the chemical engineering detailed design of the major pieces of equipment including the reactor and the refining columns. This group was also able to successfully complete a safety and HAZOP analysis around the most critical piece of equipment, in our case the reactor. Finally, an economic analysis was completed around the designed equipment and a 15-year cash flow diagram was obtained and an ROROI for the proposed project was calculated.

Regards,

Leah McClellan
Natalie Russell
Chloe Sanders
Francis Saneii
Ariel Taylor

ACKNOWLEDGEMENT

We would like to express our gratitude to everyone in the chemical engineering department at Prairie View A&M University. As senior design is a capstone course, this group would not be able to achieve what we have so far without the guidance and aid of all of the professors we have had throughout this engineering endeavor. Specifically, we would like to thank Dr Lealon Martin, our professor, for being an exceptional mentor and teaching us how to be good engineers. We would like to express appreciation for Dr Michael Gyamerah, the chemical engineering department head, for his immense support and guidance throughout this project. We would also like to thank Dr Kazeem Olanrewaju for his insights on material balances. Last but not least, we would like to thank the Prairie View A&M Alumni Industry Council for challenging us during our presentation and for their feedback on this project.

PROJECT EXECUTIVE SUMMARY

The purpose of this project proposes a design for a grass roots plant to produce ethylene oxide. Supply and demand for ethylene oxide were not readily available to us so the supply and demand of ethylene glycol, an ethylene oxide derivative, influenced the basis of production of our plant. Using market analysis and comparing the average of top ethylene oxide producers, we calculated a yearly production rate of 1.1 billion pounds of ethylene oxide. Additionally, we chose the United States Gulf Coast for the location of this plant due to the availability of ethylene feedstock and the access to multiple routes of transporting this product to consumers. We chose direct catalytic oxidation of ethylene using oxygen and a silver alumina catalyst as the production route for this project. Moreover, we refined the material balance in order to move on to the detailed design of the major pieces of equipment such as the reactor and refining columns. Simultaneously, we made a control scheme around the reactor as well as a detailed HAZOP analysis. After completing those deliverables, we established an economic analysis for a 15-year lifespan of this plant, calculated a ROROI of 38%, and we recommend the project move forward as designed.

What is the
total annualized
cost of the plant?
feed stocks?

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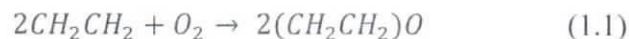
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1. PROJECT SCOPE

1.1 Problem Statement

The demand for ethylene oxide expects to rise 32,610 thousand tons, or roughly 6.5 billion pounds, by the end of 2023 (Markets, 2018). Our team has been asked to design a grass-roots ethylene oxide plant to ~~produce 1.1 billion pounds per year~~. The primary end-use of ethylene oxide distributes out to produce ethylene glycol. In 2018 ethylene glycol production accounted for 73% of the ethylene oxide market (IHS Markit, 2019). With that being said, our plant will produce ethylene oxide at a purity of 99.5%.

To produce ethylene oxide, our team chose an industry standard route: Catalytic oxidation of ethylene in the presence of a silver-supported catalyst. The primary result of this reaction is the partial oxidation of ethylene to ethylene oxide. The secondary result is complete oxidation of ethylene to carbon dioxide and water. Shell supplies varying ethylene oxide catalysts in order to meet process selectivity and catalyst life specifications. For our process, we need a catalyst with high selectivity and a long catalyst life. The main reaction for the process is as follows:



1.2 Client Identification and Recognition of Need

Ethylene oxide is an intermediate primarily used to produce ethylene glycol- an ethylene oxide derivative. The demand for ethylene glycol comes from the automotive industry such as antifreeze, brake fluids, polyester, fibers, paint, and plastic. Outside of the automotive industry, industrial coolants in air conditioning systems, gas compressors, and ice-skating rinks derives from ethylene glycol. Also, within the medical industry, ethylene glycol sanitizes equipment that is sensitive to high temperature steam. Therefore, our purified ethylene oxide will sell to ethylene glycol factories as starting material to create different products.

1.3 Project Goals and Objectives

2 *by main cycle*

Our goals for the design of our ethylene oxide plant include producing a 99.5% purity of ethylene oxide with a catalyst, while considering reflux. While completing these production objectives, we are making safety a main priority throughout the process and making the most efficient decisions to save money and increase selectivity to optimize production and appeal to investors. This poses a challenge because creating the right conditions for the desired reaction creates a ~~hassle~~ especially with such a highly flammable and potentially explosive process.

Throughout the project, we focused on reaching our goals in a timely manner while making the most efficient process. Efficiency within a chemical engineering production process identifies as one of the most important keys to success. Increasing the efficiency means less time, resources and money used to increase profit made. ~~We must be able to sell this project to investors so we can make this plan for success become a reality.~~

1.4 Contemporary Issues and the Impact of the Project on Society

yes. not well-described.

Because of ethylene oxide's daily use in the automotive and medical industry, our contribution to the creation of more resources from its' production, will increase the industries that utilize it and will affect the economy positively. Therefore, creating more resources for people who use their car for work every day as well as keeping medical equipment sanitary is important to ~~keeping the public happy and healthy.~~

1.4.1 Contemporary Issues Relevant to the Project

Alternatively, ethylene oxide produces a sterilization for medical devices that are sensitive to high temperatures. Almost half of medical devices today use EO for sterilization. The demand for ethylene oxide continues to grow with the economy as well as technological

advances. Recently, due to the COVID-19 outbreak the usage of ethylene oxide has decreased. The increased demand for N-95 masks has on the contrary negatively affected the demand for ethylene oxide. Large companies such as Medline who extensively uses ethylene oxide to sanitize masks before sending them out have decreased usage. Ethylene oxide, a known carcinogen, may leave residual amounts in the mask and has the potential to cause harm. This consequently has shut down many ethylene oxide plants. ✓

1.4.2 Potential Impact of the Project on Society

Because Ethylene oxide identifies as a derivative for ethylene glycol, the automotive industry (and other industries) takes advantage of its contributing factors. For instance, the medicinal industry uses instruments such as endoscopes that sterilize in storage cubes which contain concentrated amounts of ethylene oxide gas. This means that the growing demand for the auto industry (and others) directly impacts the demand for ethylene oxide.

On the other hand, it is important to note that many ethylene oxide plants are being shut down due to the COVID-19 pandemic. This pandemic has caused a lot of health issues.

Formosa, a major ethylene oxide producer and petrochemical company, was in the works of starting a new plant in Louisiana. The plant was ultimately put on hold due to the Environmental Protection Agency and environmental health issues. Although Formosa stated that they would not be emitting more than the permittable emissions, it was ultimately decided that it would be best to put the 14-part plastics plant (that a large included ethylene oxide unit) on pause due to the health crisis at hand.

1.5 Project Assumptions/Constraints

In order to complete the assigned deliverables, the following assumptions list below:

1. Feed impurities are negligible *justification*
2. Polymer grade ethylene feed (99.999% Purity) *ok*
3. Oxygen feed 99.9% pure *ok*
4. Water to EO ratio 22:1 (molar) *why?*
5. Methane used as an inert gas to keep reactor feed and reactor effluent outside of the flammability limits *is this an assumption?*
6. Negligible formation of aldehydes
7. Negligible losses of ethylene oxide in the ethylene oxide stripping column

2. PROJECT PLANNING AND TASK IDENTIFICATION

2.1 Task Identification

Task identification for this project was straightforward. Major tasks were given for the project and were broken down into subtasks. The subtasks were broken down into a timeline. All of these elements are listed in the following subsections.

2.1.1 Major Tasks

1. Finalize MEB; Duration: 9 days
2. Equipment Design; Duration: 30 days
3. Cost Analysis; Duration: 45 days
4. Optimization; Duration: 10 days
5. HAZOP Analysis; Duration: 6 days
6. Mechanical Engineering Design; Duration: 11 days
7. Presentation; Duration: 64 days
8. Report; Duration: 67 days
9. Aspen Simulation; Duration: 54 days

2.1.2 Subtasks and Timeline

The following list is the breakdown of our major tasks into phases, subtasks, and milestones. The list contains information on which major tasks are in the specific phases and the link between major tasks and subtasks.

(a) Phases/module

1. Phase 1: Tasks 1 & 2
2. Phase 2: Tasks 3, 4, 5 & 6
3. Phase 3: Tasks 7, 8 & 10

(b) Subtasks

1. Reactor MEB; Duration: 9 days
2. EO Absorber MEB; Duration: 9 days
3. CO₂ Absorber MEB; Duration: 9 days
4. CO₂ Stripper MEB; Duration: 9 days
5. EO Stripper MEB; Duration: 9 days
6. EO Light-ends Column MEB; Duration: 9 days
7. Preliminary Equipment Design; Duration: 16 days
8. Final Equipment Design; Duration: 16 days
9. Pump Design; Duration: 6 days
10. Heat Exchanger Design; Duration: 6 days
11. Preliminary Cost Analysis; Duration: 16 days
12. Final Capital Cost Analysis; Duration: 25 days
13. Final Operating Cost Analysis; Duration: 6 days
14. Plant Layout Design; Duration: 11 days

where do these subtasks fall within phases?

15. Mechanical Design of Major Equipment; Duration: 11 days

16. Preliminary ASPEN Simulation; Duration: 13 days

17. Final ASPEN Simulation; Duration: 41 days

(c) Link of Major Tasks with Subtasks

1. Major Task 1 Linked with Subtasks 1 – 6
2. Major Task 2 Linked with Subtasks 7 – 10
3. Major Task 3 Linked with Subtasks 11- 13
4. Major Task 6 Linked with Subtasks 14 & 15
5. Major Task 9 Linked with Subtasks 16 & 17

ok. ✓

(d) Milestones & Deadlines

1. Milestone 1: Finish Preliminary Equipment Design and Cost Analysis; Deadline:
3/2/20
2. Milestone 2: Finish Final Equipment Design and Cost Analysis; Deadline:
4/10/20
3. Milestone 3: HAZOP Analysis of Critical Equipment; Deadline: 3/27/20
4. Milestone 4: Finish Final ASPEN Simulation; Deadline: 4/13/20
5. Milestone 5: Finish Presentation Draft for Dry-Runs; Deadline: 4/13/20

✓

2.2 Gantt Chart

Figure 1 below represents the simplified Gantt chart constructed for this semester made using Microsoft excel. We used the major tasks to represent the overall timeline for this project. The time has been given in weeks rather than days in order to be displayed properly.

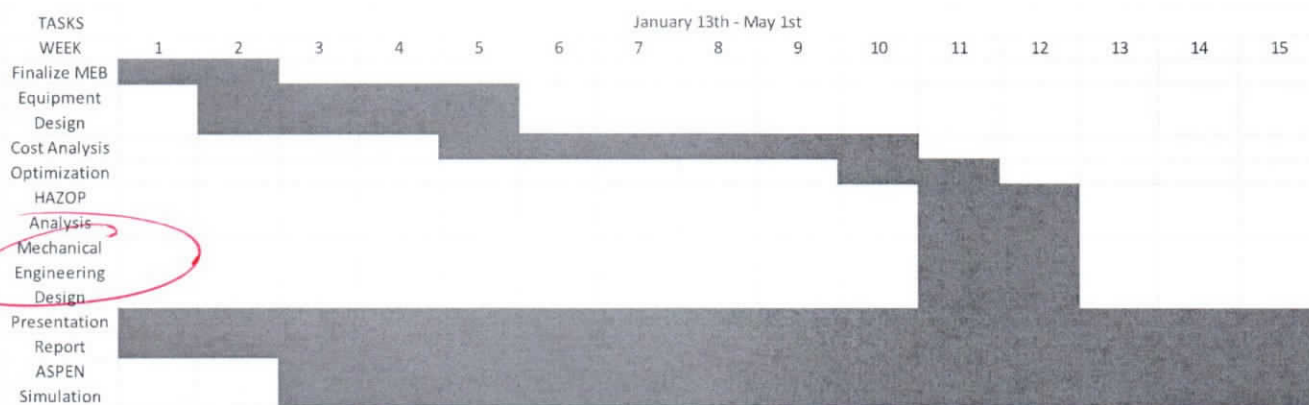


Figure 1. Simplified Gantt Chart

3. OVERVIEW OF THE MANUFACTURE OF ETHYLENE OXIDE

Historically, two manufacturing processes exist in the commercial production of ethylene oxide. The first process is the chlorohydrin process, performed as a lab synthesis in 1914 by Badische Anilin-und Soda-Fabrik (Speight, 2019). During World War I, Union Carbide Corporation became the first company to produce ethylene oxide commercially. ✓

In 1931, Lefort discovered a more efficient process to produce ethylene oxide that involved the direct oxidation ethylene by air. Shell Oil Company improved Lefort's discovery in 1958 by replacing air with oxygen (Speight, 2019). Currently, the direct oxidation process is the primary method of use in the commercial production of ethylene oxide.

3.1 Production Routes

3.1.1 Chlorohydrin Process

Although the chlorohydrin process is no longer used for production of ethylene oxide, the process was considered for production routes. The chlorohydrin process is a multi-step process that can be divided into two phases. During the first phase of production, ethylene is reacted with hypochlorous acid to form chlorohydrin:

Same?



The second phase of production is the dehydrochlorination step, which involves the reaction of the chlorohydrin with a base to give ethylene oxide and by product calcium chloride:



Eqn not balanced!

where is this Cl₂ coming from??

The problem with the chlorohydrin process is that it was not very economic. A large portion of the chlorine, which was introduced in the first step as hypochlorous ended up as calcium chloride (Burdick, Leffler, & Dodd, 2000). This not only causes disposal issues; it can cause build-up in the equipment and reduce the mechanical integrity of any equipment that contains the build-up.

3.1.2 Direct Oxidation

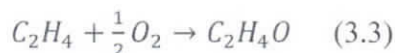
how is this done with?

The direct oxidation process has replaced the chlorohydrin process and involves the direct oxidation of ethylene in the presence of a silver catalyst. The process is highly exothermic and heat control is considered a major design constraint for the process. If the process temperature exceeds the appropriate conditions, it can lead to catalyst deterioration and reduce the yield of ethylene oxide (Sami Hatch & Lewis, 2001).

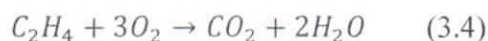
The process can be performed using two routes. The routes are classified by considering the source of the oxidizing agent, which are air and oxygen. Although the routes of production are similar, there are a few differences. One of the primary differences between them is that air-based direct oxidation produces a large amount of nitrogen, which ends up concentrated in the recycle gas stream. The recycle stream must be vented to prevent the nitrogen concentration from getting too high (Rebsdatt & Mayer, 2017).

OK.

Currently, the oxygen-based direct oxidation is the main production route used for the commercial production of ethylene oxide (Rebsdats & Mayer, 2017). The main reaction involved in the direct oxidation process is the reaction of ethylene with oxygen over a bed of silver catalyst.



A side reaction occurs and results in the formation of byproducts, carbon dioxide and water:



The side reaction is limited based on catalyst selectivity. An optimized design of the direct oxidation process will produce high amounts of ethylene oxide over a catalyst cycle and reduces the amount of ethylene consumed during production (Rebsdats & Mayer, 2017).

3.2 Physical and Chemical Properties

3.2.1 Physical Properties

Name: Ethylene oxide (also known as Oxirane, Epoxyethane, Ethene Oxide and more)

Molecular formula: $C_2H_4O_2$

Molecular weight: 44.05 g/mol (lb/lbmol)

Ethylene Oxide is a colorless and toxic gas that is used as a chemical intermediate, a sterilant in the medical fields, and more. It has a sweet odor that resembles the soft smell of overly ripe apples. It is possible to smell in water if the volume is at, or above, 140 mg per liter and can also be detected in air, if it is present at, or above, 430ppm.

Ethylene oxide at room temperature is extremely reactive and flammable. If heated, it has the potential to explode and will also explode in the presence of ammonia, amines, alkali metal

hydroxides. However, ethylene oxide is completely soluble in water and breaks down in water and/or destroyed by bacteria in a span of a few days. Ethylene oxide levels decrease as it is broken down into other substances.

3.2.2 Chemical Properties

Table 1 below represents the chemical properties of ethylene oxide. These properties have been and will continue to be referenced throughout the duration of this project. Properties, such as boiling point, will be used to help determine the best conditions for separation processes such as distillation.

Table 1: Chemical Properties of Ethylene Oxide

Property	SI (AES) Units
Boiling Point	11°C (51°F)
Freezing point	-112°C(-169°F)
Flash Point	-29°C to -30°C (-20° F through -22°F)
Specific Gravity	0.88 <i>units</i>
Density	.822 at 50°F (.882 at 10°C/10°C)
Vapor Density	1.49 relative to air
Vapor Pressure	1068 mmHg at 68°F
Water Solubility	22.7 M
Melting point	-170.5 ° F (-111.7°C)
Exposure limits (highest)	800ppm
Exposure limits (lowest)	< .1ppm ✓

Table 2 below represents important properties of the other chemicals used in this process.

The properties shown are the components boiling points, lower explosive limits (LEL) and upper explosive limits (UEL) if available. The range between the LEL and UEL is the amount of chemical that can be present in oxygen, by volume percent, that can potentially cause a fire or explosion. The amount of chemical present that is below the LEL and above the UEL will not cause a potential fire or explosion. The basis of the explosive limits are mixtures within a two-inch tube at atmospheric pressure and room temperature.

Table 2. Chemical Components: Boiling Points and Flammability Limits

Component	Boiling Point	LEL – UEL
Ethylene Oxide	51.3 F	3.6 - 100 % by volume*
Ethylene	-154.7 F	2.7 - 36 % by volume*
Oxygen	-297.3 F	N/A
Water	212 F	N/A
Carbon Dioxide	-109.2 F	N/A
Acetaldehyde	68.36 F	N/A
Formaldehyde	-2.2 F	N/A

3.3 Handling, Storage and Safety

3.3.1 Handling

Safety is of the utmost importance when it comes to the handling and storage of ethylene oxide. When it comes to storage, the vessels considered here are above ground, mounded and buried vessels. Types, advantages, and disadvantages will be discussed as well as the best materials of construction in order to mitigate ethylene oxide hazards and handle the product safely.

3.3.2 Storage

Above ground storage vessels offer three different types for storing ethylene oxide: Vertical or horizontal cylindrical tanks and spherical tanks. Vertical cylindrical tanks offer minimal contact between the product and flame sources, otherwise known as flame impingement. This type of vessel also allows for the inventory to be monitored easily as well as minimizing the potential for the product to accumulate polymer. Horizontal cylindrical tanks minimize the potential of impact from any flying debris. (American Chemistry Council's Ethylene Oxide/Ethylene Glycols Panel, 2007)

Handling ethylene oxide should be done in such a way that mitigates the hazards the product presents. In order to do so, it is best to keep in mind the materials that make up the process

equipment as well as storage tanks. The equipment used to handle, and store ethylene oxide is usually made from austenitic stainless steel or mild carbon steel. Stainless steel minimizes the potential of the formation of rust which facilitates the polymerization of ethylene oxide. However, depending on the conditions of the environment, stainless steel has the potential to corrode on the outside, but this can be prevented by using a protective outer coating. Similarly, carbon steel can corrode on the outside which can also be prevented. However, carbon steel can accumulate rust internally and cause ethylene oxide to polymerize. Therefore, when it comes to choosing what materials to use for storage and handling, it is imperative to understand the process and ambient conditions in order to protect the equipment's integrity (American Chemistry Council's Ethylene Oxide/Ethylene Glycols Panel, 2007). For this process, we will need to do some further investigating into storage alternatives given that ethylene oxide has several storage constraints that need to be mitigated.

project is over. This should be dealt with now.

3.3.3 Safety

It is important to keep safety in mind when handling or around ethylene oxide. Protective gear should always be used. Goggles and a hard hat when around any chemical production site is important. As well as earplugs for ear protection in most production sites because there is a lot of noise. Being around loud machines for long periods of time without ear protection will cause long term damage. Lastly, fire resistant coveralls are important in any process and must be used to protect personnel against the impact of flash fires.

The basics when around chemical plants are important but when around ethylene oxide there are many short-term effects of the chemicals such as, according to OSHA, respiratory irritation, lung injury, headaches, nausea, vomiting, diarrhea, shortness of breath and skin turning blue. Chronic long-term damage includes reproductive effects, changes or damage to DNA, impact on

learning, central and peripheral nervous systems. Ethylene Oxide May also cause lymphoma and leukemia as well as stomach cancer and breast cancer in both men and women.

Production site dangers include the potential of a runaway reaction given that the reaction is highly exothermic. Any spark or high temperature change may create a potential for an explosion. Ethylene Oxide is a highly flammable and corrosive chemical. It is important for the production site to have on campus additional cooling water as well as low temperature cooling water in order to combat a runaway reaction during a worst-case scenario such as loss of power during the reaction process. Adding gas washes to clean out the reactor before catalyst coking is important to make sure there are no reactions in the reactor during downtime. An extra extraction chamber just in case to be able to hold the reaction will also be important to decrease the size of the potential explosion. Keeping a diluent ready just in case to absorb the heat in order to avoid heat sink.

Keeping a gas monitoring system to sense how much ethylene oxide is left after gas washes as well as in other pieces of equipment connected to the reactor to make sure there are no trace amounts of ethylene oxide moving throughout the cleaning process. Shatter resistant glass for all control windows and when possible use cameras for safe distance monitoring. Oxidizers must have gas monitors, safety interlocks when moving from one piece of equipment to another as well as alarms if there is a leak. National fire protection agency requires safeguards at all ethylene oxide facilities including explosion damage control devices for worst case scenarios. Pressure relief valves are used when the pressure is too high during a runaway reaction that there will be less damage done in surrounding areas in the case of an explosion.

Mandatory training for ethylene oxide regularly as well as process hazard training whenever there is a new hire. As well as the process hazard analysis program for oxidizers, oxygen, and their potential for explosion, for all employees. ✓

3.3.3.1 EPA, OSHA Regulations and Environmental Impact

Table 3 below represents the impact on an individual that is exposed to ethylene oxide on an acute and chronic basis. These effects are presented by OSHA and were explained in the safety section.

Table 3. OSHA Short-term and Long-term Exposure

Acute (Short term) effects of the chemicals	Chronic (long-term) damage	Carcinogenic Effects
•Respiratory irritation	•Reproductive effects	•Lymphoma
•Lung injury	•Changes or damage to DNA	•Breast
•Headaches	•Impact on learning	•Stomach Cancer

Table 4 below represents the environmental impact regulations from the EPA. These regulations are in place in order to mitigate pollution and adverse effects on people living in the surrounding areas of the site.

Table 4. Environmental Impact Regulations

Environmental Impact
5,000 ppm or .5% of CO ₂ can be admitted into the air (Permissible Exposure Limit PEL)
Methane NIOSH permissible amounts released in the air 1,000ppm (.1%)
Ethylene is not harmful or toxic to humans however at or above 27,000ppm is combustible

→ do you incorporate these limits into your design.

3.4 Uses

Ethylene oxide is used as an intermediate in the production of ethylene glycol and makes up the majority of ethylene oxide consumption. It can also be used in the production of ethoxylates which can be made into detergents, paint solvents and other textile industry applications. ✓

Moreover, another major application is the use of ethylene oxide to produce ethanolamine.

(Burdick, Leffler, & Dodd, 2000)

3.5 Production Specification

Ethylene oxide is produced with a purity >99.5%. Because ethylene oxide primary end-use is to make ethylene glycol, the purity needs to be high in order to appeal to ethylene glycol producers. High purity technical grade ethylene oxide is produced with a purity of 99.7%.

this is not a reason.

(Burdick, Leffler, & Dodd, 2000) ✓

4. PROCESS SELECTION

where is this comparison? A comparison of the two processes shows that the direct oxidation is favored from an economic standpoint. One of the disadvantages with the chlorohydrin process is the cost of operation, primarily the feedstock. The chlorohydrin process requires chlorine and is costly. Another disadvantage of the chlorohydrin process is accumulation of calcium chloride which causes issues with disposal and does not have any economic benefit. (Burdick, Leffler, & Dodd, 2000).

Do you have a Theoretical Economic Analysis

4.1 Market Analysis and Trends

Little information is available to us for the ethylene oxide market. Since ethylene glycol production accounts for 73% of the ethylene oxide market, this will be the basis of the market analysis we have developed (IHS Markit, 2019). We assumed the supply and demand of ethylene oxide was dependent on the ethylene glycol market. Figure 2 below ~~represents~~ ^{shows} the supply and demand for ethylene glycol developed for analysis.

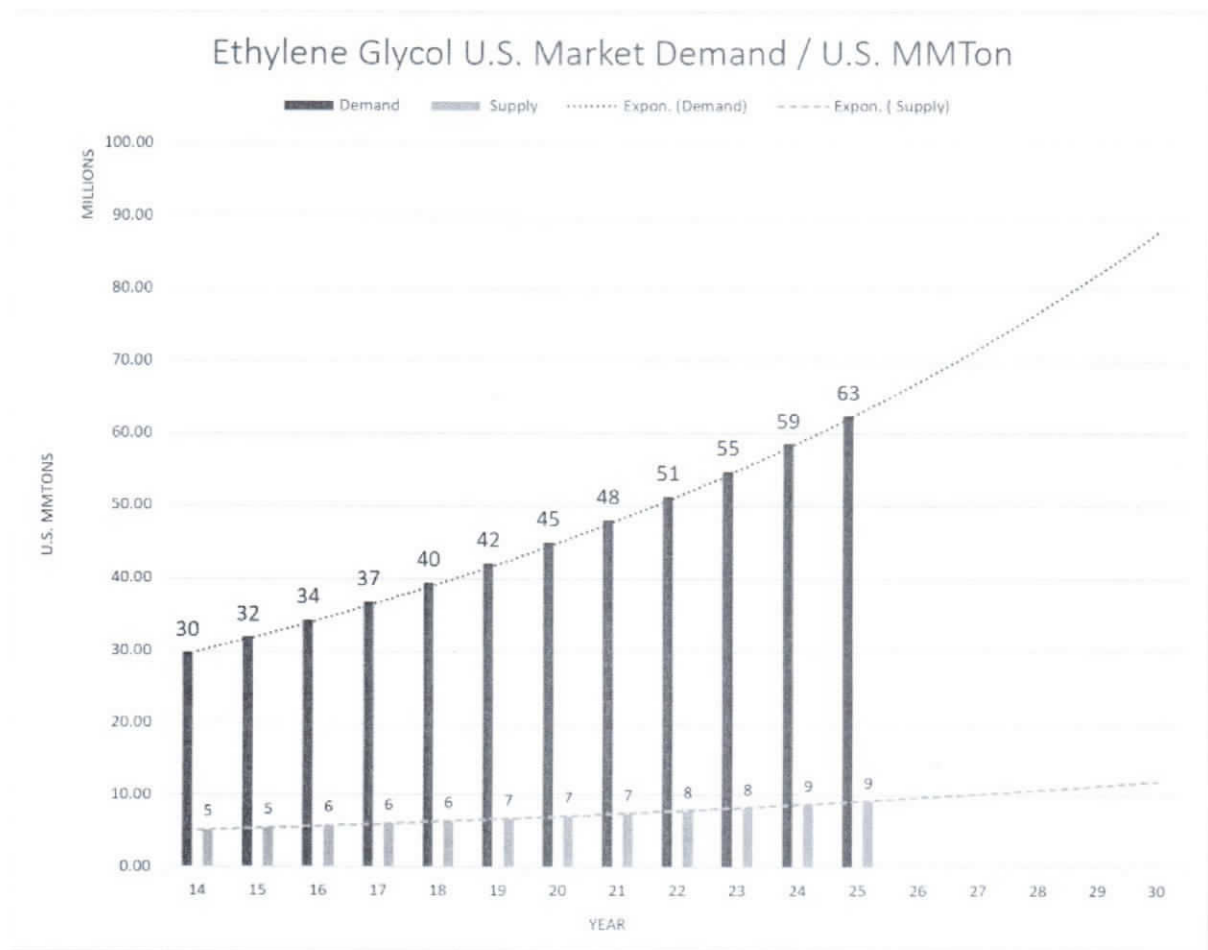


Figure 2. Supply vs Demand of Ethylene Glycol

The 42.19 MM pounds for the year of 2019 makes up 15% of the gap. This is a practical amount compared to other top companies. Our Gross Profit Analysis of 28 cents of Ethylene Oxide per pound translated to a Total Profit of 308 million dollars per year based on our yearly production rate of 1.1 billion pounds.

Since ethylene oxide is an ethylene glycol derivative, ethylene oxide will be in demand if ethylene glycol is. According to the U.S. glycol market demand study, ethylene glycol demand will continue to increase going into 2025. This is important to note so that we will be able to plan for a growth in production to meet the growing demand.

Ethylene Glycol global market is 30.41 billion in 2018. It is projected to expand at a compound annual growth rate (CAGR) of 7%. Currently within production regions the middle east holds 36% of global production in the world. North America following second with 21% of global production. Europe, China, and Taiwan follow, respectively.

not sure why this is here.

The industry for the global Gross Domestic Product (GDP) includes the automotive (the fastest growing application), polyester (fiber and resins), pharmaceuticals, food & beverage processes, maintenance for pipelines, Heating Ventilation and Air Conditioning (HVAC), textiles, aviation, and medical. The GDP expanded by 3.7% within developed nations. The glycol global economy has a promising future.

4.2 Reaction Routes to Producing Product

4.2.1 Industrial Processes and Lab/Patent Routes

The main industrial process used to produce ethylene oxide is the direct oxidation of ethylene. This is well known in industry although there are optimizations for the process. For example: the use of methane as an inert gas instead of nitrogen. This is an optimization used to eliminate any amines that might form in the reactor which can pose a threat to the ethylene oxide process. Ethylene oxide in the presence of amines has the potential to polymerize explosively (Ethylene Oxide).

4.2.2 Process Comparison and Selection

There are three different routes to produce ethylene oxide as mentioned in sections 3.1.1 and 3.1.2. Each route has its pros and cons and are presented in table 5 below.

should include gross economic analysis here.

Table 5. Process Comparison: Advantages and Disadvantages

	Chlorohydrin	Air-Based Oxidation	Direct Oxidation
Advantages	• No Catalyst required • Low Operating Temperatures • 80% Selectivity	• Lower operating cost • Low purity ethylene	• High selectivity • No purge reactor required • Higher capacities are more economical
Disadvantages	• Requires a large amount of chlorine (corrosive & expensive) • Calcium chloride byproduct	• Air purification required • Requires a purge reactor • Lowest selectivity	• Expensive instruments • Extra feedstock required for CO₂ removal

The chlorohydrin process is no longer used in industry due to the use of chlorine, which is expensive and results in a build-up of calcium chloride that is difficult to dispose of. Oxidation of ethylene is the common route in industry today. For this project, the direct oxidation of ethylene using pure oxygen as a primary feedstock has been selected.

5. PROCESS DESCRIPTION

5.1 Process Chemistry

5.1.1 Reaction Stoichiometry

The reactor used for commercial production of ethylene oxide is a multi-tubular plug flow reactor. The reactor tubes filled with the catalyst are surrounded by a coolant, which can be either water or a high-boiling hydrocarbon that removes the reaction heat and allows for temperature control. The operational pressure inside the reactor varies depending on process characteristics, but it is generally in the range of 15-25 atm pressure (Aryana, 2009). The raw materials consist of ethylene (purity > 99.5 % and free of sulfur), oxygen (purity > 99%), and

what coolant?

good!

recycle gas is heated and fed to the reactor. In general, three parallel reactions take place during the oxidation of ethylene inside the reactor, which shows as follows:



The reaction is highly exothermic and produces a large amount of heat. The hot outlet gas is cooled by using steam generation or will be used directly to heat the reactor inlet gases before going to the next stage of the process.

5.1.2 Kinetics and Rate Equations

Table 6 below displays the kinetic constants that will be used when designing and modeling the reactor in ASPEN.

Table 6. Reaction Kinetics

	Activation Energy	Pre-Exponential Factor
Ethylene Oxide Reaction	41,946 Btu/lbmol	$5.83 \times 10^9 \text{ (ft}^3/\text{lbmol)}^5/\text{hr}$
Combustion Reaction	7,527 Btu/lbmol	$7.15 \times 10^{14} \text{ (ft}^3/\text{lbmol)}^3/\text{hr}$

This data was found in a reactor design report from Rice University. The pre-exponential factors were found using an iterative process (Kim & Palombo 1997). The following equations are used to determine the kinetic factor, k , and the rate of reaction for both reactions present in our system. We assumed elementary reactions both first order in terms of ethylene oxide.

$$k = Ae^{\frac{-E_a}{RT}} \quad (5.4)$$

$$r_{\text{Ethylene Oxide}} = kC_{C_2H_4}C_{O_2}^{0.5} \quad (5.5)$$

$$r_{Combustion} = kC_{C_2H_4}C_{O_2}^3 \quad (5.6)$$

5.2 Process Flow Diagrams and Process Description

5.2.1 Block Flow Diagram

Figure 3 below represents the block flow diagram developed for this project.

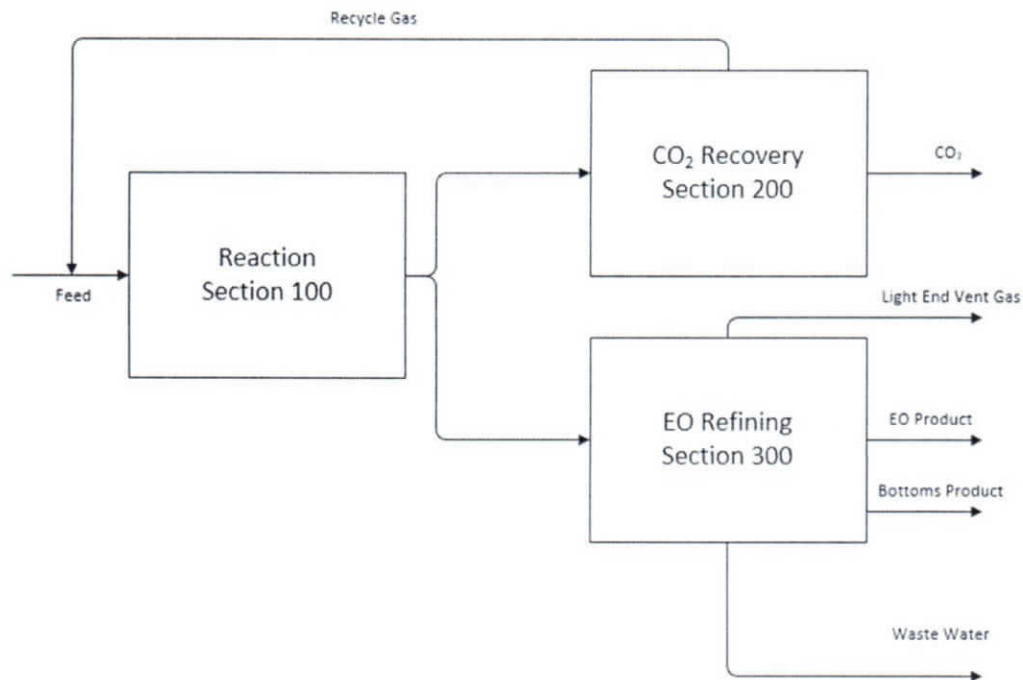


Figure 3. Block Flow Diagram for Ethylene Oxide Plant

This diagram was used as an overview of the plant and is broken down into sections which are referenced in the process flow diagram. Section 100 includes the reactor and the first stage of ethylene oxide separation, section 200 includes the units separating carbon dioxide from our reactor effluent and section 300 includes the remaining units used to separate ethylene oxide to obtain our final product.

5.2.2 Process Flow Diagram

Figure 4 below represents the process flow diagram developed for our project.

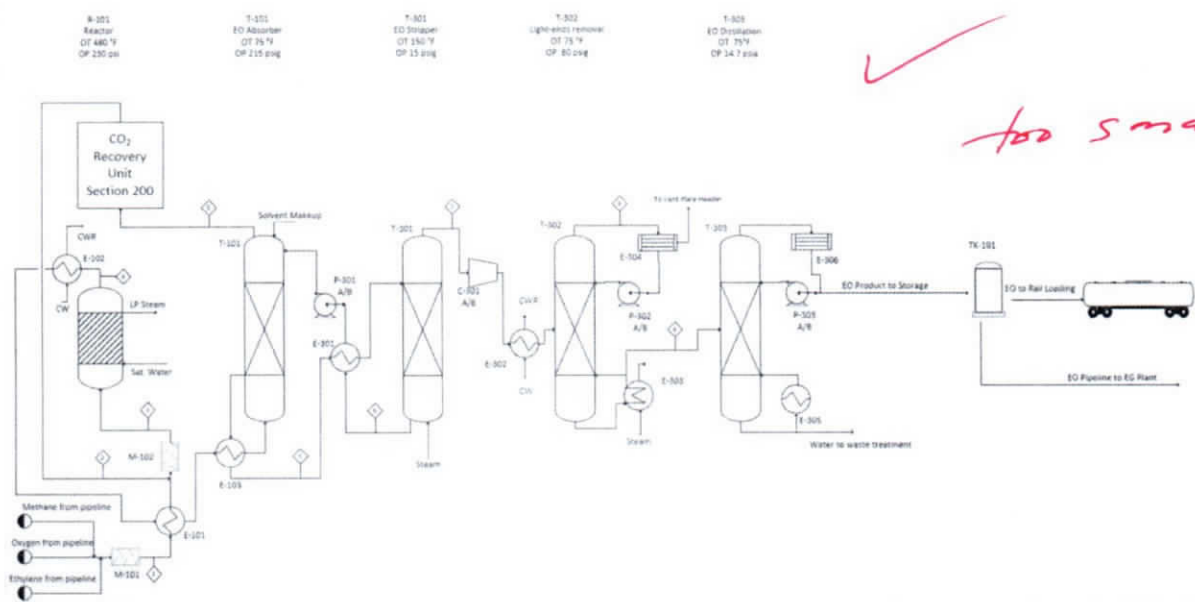


Figure 4. Process Flow Diagram for Ethylene Oxide Production

This diagram represents the plant layout in more detail than the block flow diagram. This diagram displays the units used in each part of the process. The remaining parts of this section describe the process and each unit shown above.

5.2.2.1 Raw Material Preparation

The raw materials used in this process are ethylene and oxygen with methane used as an inert gas to dilute the process enough for lower flammability limits. It will be heated in an exchanger that utilizes the heat of the reactor effluent to prepare the feed to the appropriate conditions to enter the reactor. This will also remove heat from the reactor effluent to prepare for entrance into the ethylene oxide scrubber.

5.2.2.2 R-101 Ethylene Oxide Reactor

The reactor used for commercial production of ethylene oxide is a multi-tubular plug flow reactor, which consists of a large bundle of several thousand tubes. The reactors and the sections of the plant that contain ethylene oxide are usually made of mild steel or stainless steel to reduce the formation of acetaldehyde by isomerization of ethylene oxide on rust (Rebsdats & Mayer, 2017).

The reactor tubes filled with the catalyst are surrounded by a coolant, which can be either water or a high-boiling point hydrocarbon that removes the reaction heat and allows for temperature control. The temperature inside the reactor depends on the type of catalyst and varies from 480-520 °F. (Aryana, 2009). The operational pressure inside the reactor depending on process characteristics, but it is generally in the range of 150-440 psi (Aryana, 2009). The raw materials consist of ethylene (purity > 99.5 % and free of sulfur), oxygen (purity > 99%), and recycle gas is heated and fed to the reactor. In general, three parallel reactions take place during the oxidation of ethylene in the presence of catalyst. The partial oxidation of ethylene which results in ethylene oxide is our desirable reaction. The selectivity of this reaction is 80% and conversion with respect to ethylene is 20%. *Source ??*

The reaction is highly exothermic and produces a large amount of heat. Therefore, an efficient cooling system is used to remove the heat of reaction by pumping the coolant to the reactor to avoid hot spots. The hot outlet gas is cooled by using steam generation or will be used directly to heat the reactor inlet gases before going to the next stage of the process. (Rebsdats & Mayer, 2017)

5.2.3 Separations

5.2.3.1 Ethylene Oxide Refining

5.2.3.1.1 T-101 Ethylene Oxide Absorber

After the gas from the reactor has been cooled, ethylene oxide (1 – 2%) must be removed by scrubbing with water. In the ethylene oxide scrubber, practically all the ethylene oxide and small amounts of constituents such as CO₂, N₂, CH₄, CH₂CH₂, and aldehydes within natural gas dissolve in water and separate from other gaseous products. The operation temperature and pressure of the ethylene oxide scrubber is 90-100 °F and slightly less than the reactor pressure about 180-190 psi respectively (Rebsdatt & Mayer, 2017).

5.2.3.1.2 T-301 Ethylene Oxide Stripper

The aqueous stream leaving tower T-101 composed of ethylene oxide, water, and a small amount of impurities is sent to the ethylene oxide stripper. Ethylene oxide is recovered using steam and comes out as overhead product. The bottoms are sent back to the ethylene oxide absorber as recycled solvent.

5.2.3.1.3 T-302 Light Ends Column

The overhead from T-301, containing recovered ethylene oxide and some impurities, is sent to the light ends column (T-302). The ethylene oxide is stripped from the lighter boiling point components such as some carbon dioxide, unreacted ethylene, and oxygen as well as some of the methane used as inert gas. Further research into disposal or re-use of these impurities will be done in order to limit release of these components into the atmosphere.

*non-condensable
vapors. Why
not use a chaper
flash?*

5.2.3.1.4 T-303 Ethylene Oxide Distillation

The solution of ethylene oxide and water leaves the light ends column as bottom product and goes to the distillation tower (T-303) to separate the ethylene oxide along as the final product that will be transferred to storage. Water and a small amount of ethylene oxide leaves the distillation column as bottom product which can be used as feed for ethylene glycol production.

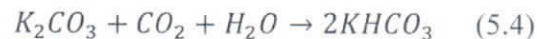
More research will be done on the possibility of using this bottom product in the ethylene oxide process. 

5.2.3.2 Carbon Dioxide Separation

5.2.3.2.1 T-201 Carbon Dioxide Absorber

The recycle gas from the ethylene oxide scrubber (5-10% CO₂) is compressed, and a side stream loses carbon dioxide by scrubbing with hot aqueous potassium carbonate solution.

Removing the CO₂ is not only by physical dissolution but also by a reaction with potassium carbonate to create potassium bicarbonate. 



5.2.3.2.2 T-202 Carbon Dioxide Stripper

The potassium carbonate solution is sent to the carbon dioxide stripper. The carbon dioxide is stripped out of the solution at atmospheric pressure and leaves the column as overhead product. Further investigation into uses and potential buyers of this carbon dioxide product will be done in order to reduce emissions and increase revenue. The recycle gas, which is free from ethylene oxide and has a lower carbon dioxide content, is mixed with the feed before being sent back into the reactor (Rebsdatt & Mayer, 2017). 

5.2.4 Product to Storage

Moving the product to storage will need to be monitored. Any backflow of the liquid could cause a rupture within the pipeline. It is important to install multiple check valves of different types to prevent similar modes of failure between the same types of device. 

6. Material and Energy Balances

6.1 Material Balance

6.1.1 Basis of Calculations

The desired production rate for this project is 1.1 billion pounds per year. With an operating time of about 92% of the year, roughly 8,064 hours, the hourly production rate comes out to be about 136 thousand pounds. The figure below represents the preliminary overall material balance done using Microsoft Excel. The overall balance shows all the components entering and leaving the plant as well as recycle.

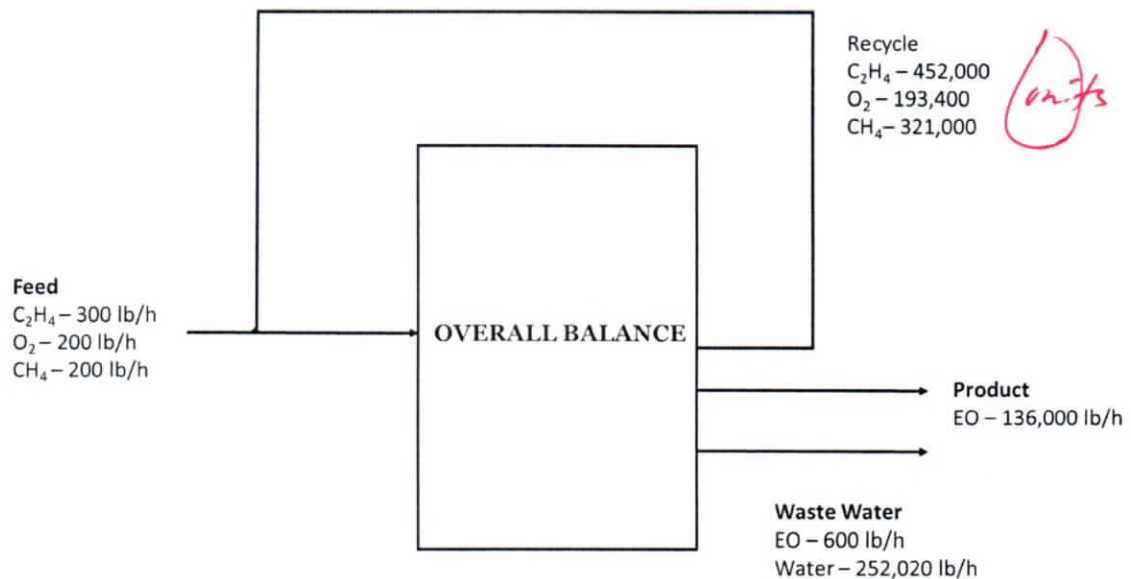


Figure 5. Overall Material Balance

6.1.2 Unit Balance

The purpose of this section is to display the individual unit balances for the main pieces of equipment. Specifically, the reactor and all the ethylene oxide separation units. The carbon dioxide separating section was not considered for the unit balances at this time.

6.1.2.1 R-101 Reactor Balance

Figure 6 below represents the unit balance for the ethylene oxide reactor. Negligible formation of aldehydes was considered for this calculation. This was assumed due to the formation of acetaldehyde being reliant on the material of the reactor that causes ethylene oxide to isomerize into acetaldehyde. Also, the extent of the formation of formaldehyde is unknown at this point but will be investigated further when refining the material balance.

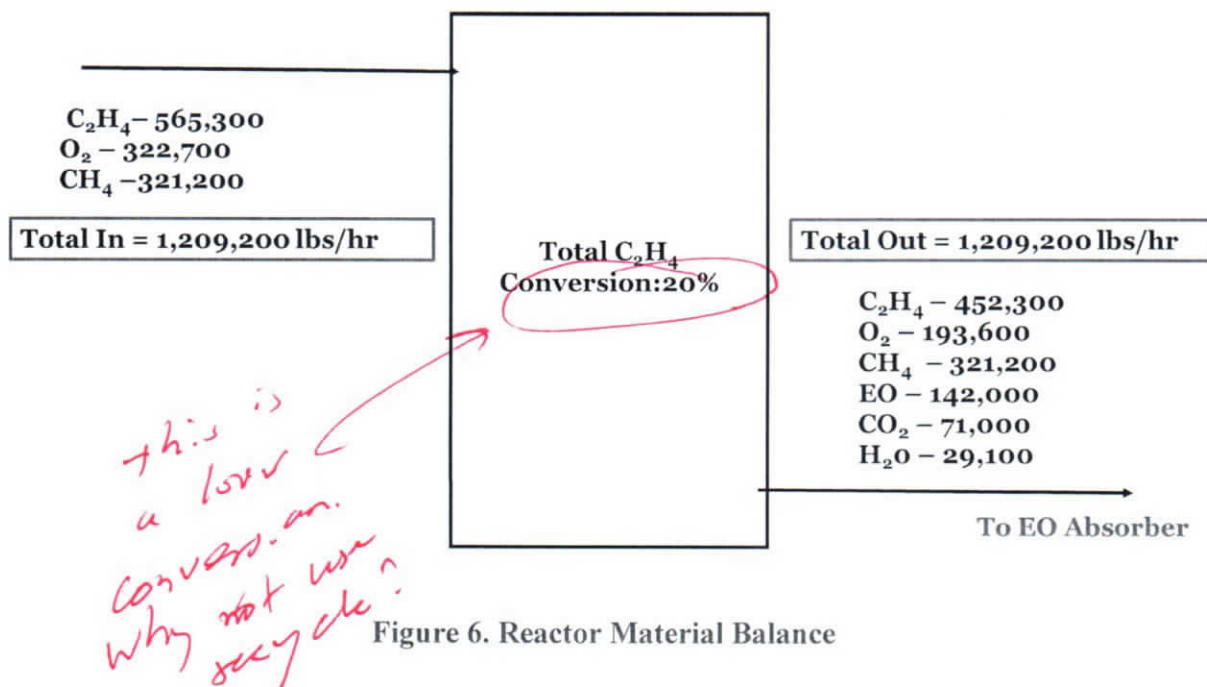


Figure 6. Reactor Material Balance

6.1.2.2 T-101 Ethylene Oxide Absorber

Figure 7 below represents the material balance around the first stage separation process. We assumed 99% of the ethylene oxide was absorbed in the solvent, water.

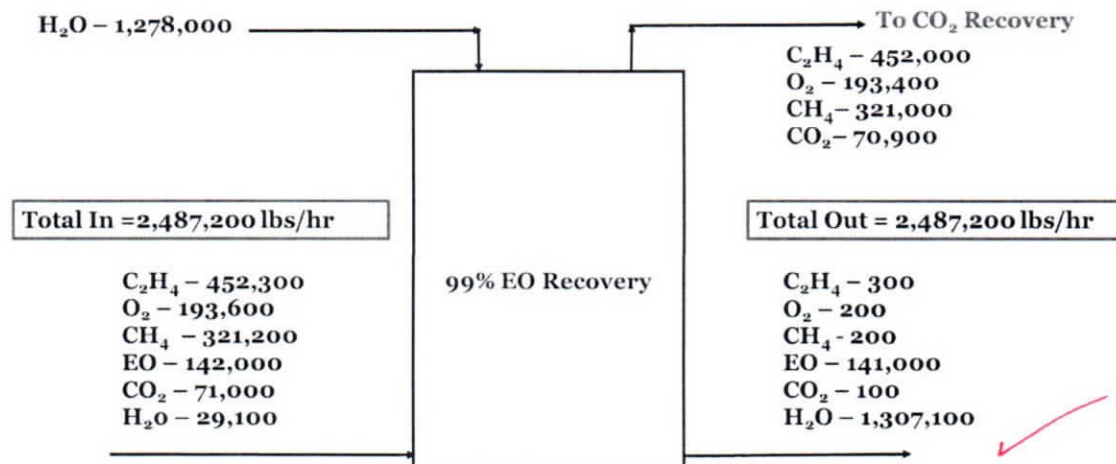


Figure 7. Ethylene Oxide Absorber Material Balance

6.1.2.3 T-301 Ethylene Oxide Stripper

Figure 7 below represents the material balance for the ethylene oxide steam-stripping column. We assumed 99% of the ethylene oxide and 100% of the light-ends were recovered in the overhead product.

↓
how can you justify

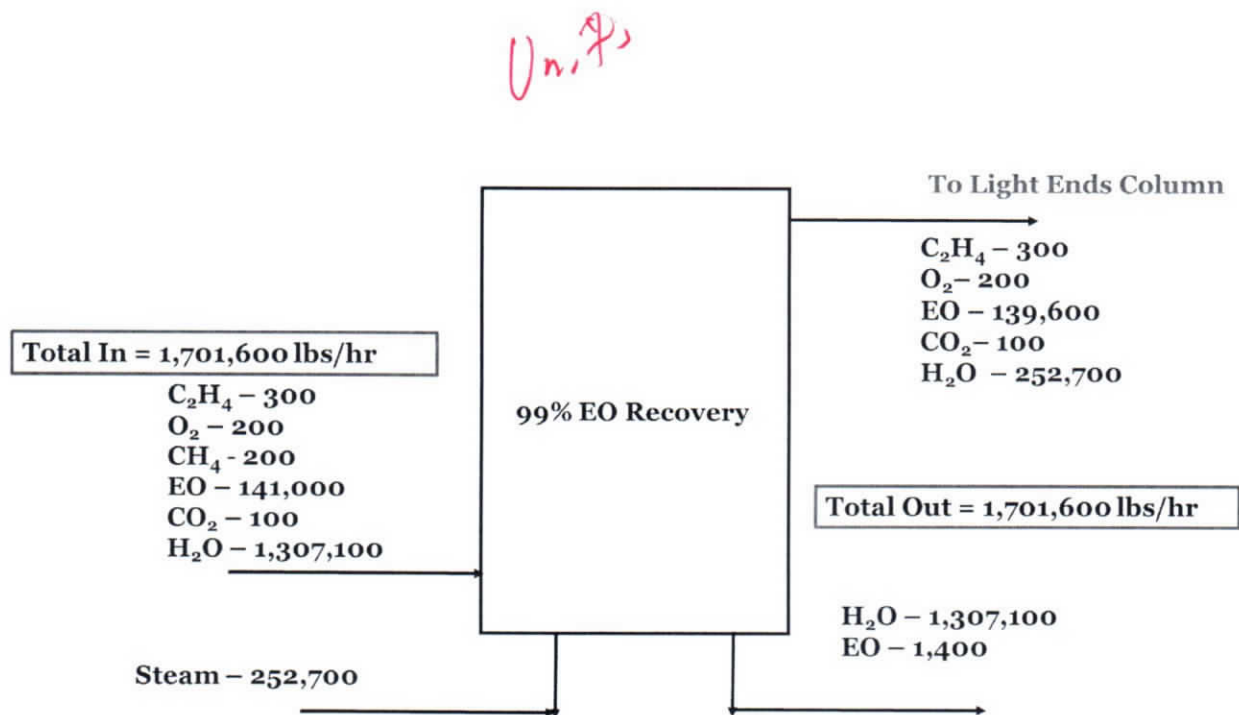


Figure 8. Ethylene Oxide Stripper Material Balance

6.1.2.4 T-302 Light Ends Column

Figure 9 below represents the material balance around the light ends column. This column is used to remove all the low-boiling components from the main product. We assumed all of the overhead product would be emitted into the atmosphere.

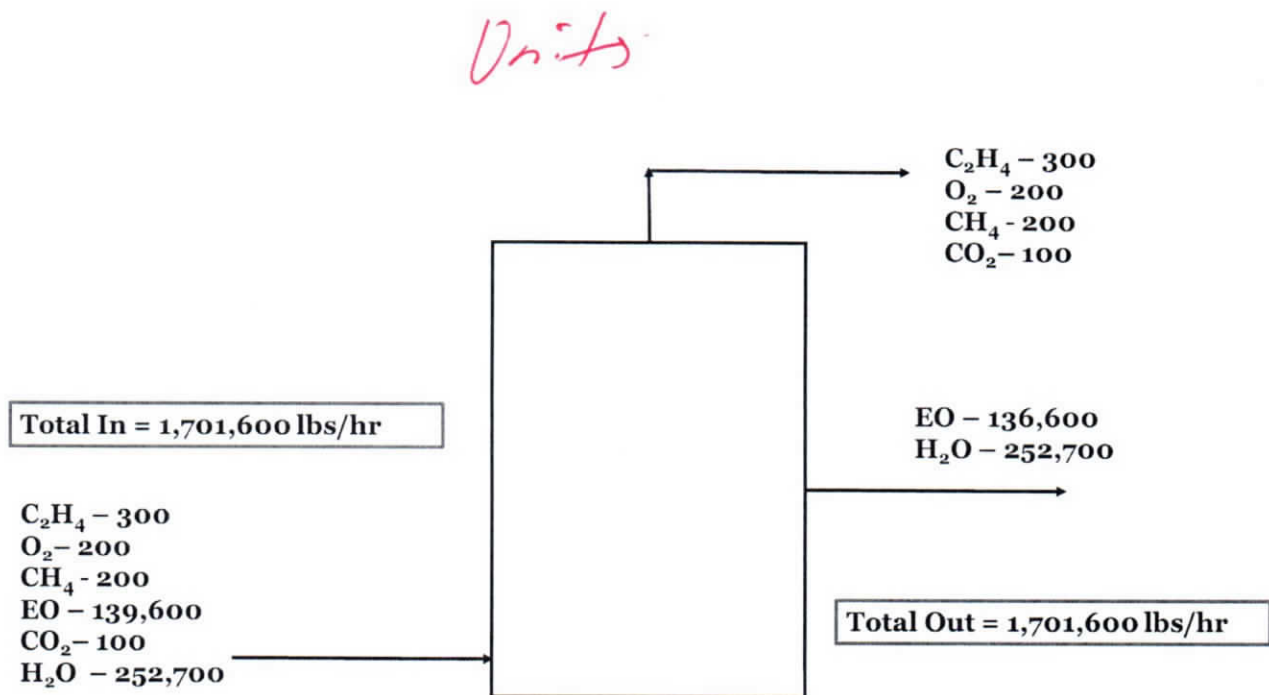


Figure 9. Light Ends Column Material Balance

6.1.2.5 T-303 Ethylene Oxide Distillation

Figure 10 below represents the material balance around the final unit in this process, the ethylene oxide distillation column. We based these calculations on the required amount of ethylene oxide we need as well as the purity specification required, 99.7%.

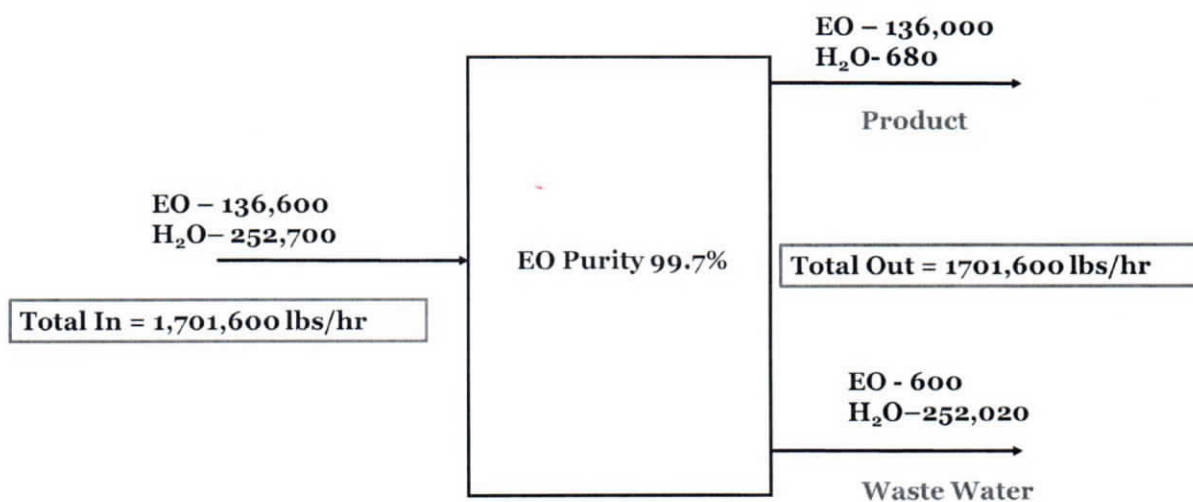


Figure 10. Distillation Column Material Balance

6.2 Energy Balances

We did an energy balance, by hand, around our reactor while the other energy balances were done using ASPEN Plus. The energy balance for the reactor was done in order to determine the amount of cooling water required to absorb the energy released during our reaction. The total amount of energy released is given by the equation below.

$$Q = (mC_p\Delta T)_{Feed} + (n\Delta H)_{Reaction\ 1} + (n\Delta H)_{Reaction\ 2} \quad (6.1)$$

We calculated the total amount of heat being released to be about 117.2 MMBTU released per hour. With eight reactors, the energy released per reactor is 14.65 MMBTU. This was used to calculate how much of the cooling water, in our case saturated water at 35.4 psig, needed in order to absorb the energy being released. The energy is absorbed by the saturated water which turns to saturated steam at 35.4 psig. We calculated the amount of saturated water required using the following equation:

$$Q = n\Delta H_{vap} \quad (6.2)$$

Where the energy absorbed by the water is the energy released during the reaction and ΔH_{vap} is the enthalpy of vaporization required to convert the saturated water into steam. We calculated the amount of water needed to be about 125 Milbmol per hour.

why not in gallons??

The table below represents the energy balance data we obtained for the remaining equipment, T-101, T-301, T-302 and T-303, using ASPEN Plus.

Table 7. ASPEN Energy Balance Data

Unit	T-101				T-301			
Stream	Inlet (T)	Inlet (B)	Outlet (T)	Outlet (B)	Inlet (T)	Inlet (B)	Outlet (T)	Outlet (B)
Enthalpy (BTU/h)	-8.73E+09	-8.34E+08	-5.71E+08	-8.99E+09	-8.89E+14	-1.44E+09	-1.02E+09	-9.30E+09
Unit	T-302			T-303				
Stream	Feed	Outlet (T)	Outlet (B)	Feed	Product	Bottoms		
Enthalpy (BTU/h)	-1.24E+09	-2.06E+06	1.23E+09	-1.16E+09	-7.37E+07	-1.06E+09		

7. HYSYS/ASPEN Simulation of Process

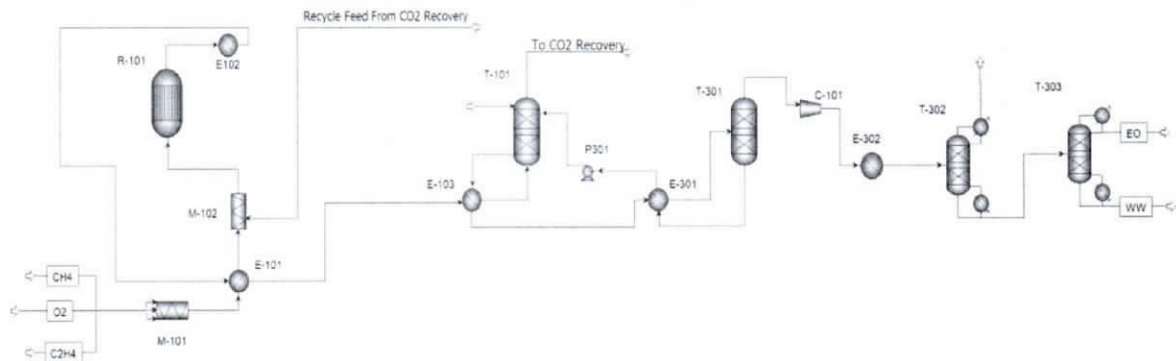


Figure 11. ASPEN Simulation of Entire Process

The figure above represents the overall ASPEN simulation diagram for our process. Our design group used Aspen Plus V9 to perform the process modeling and simulation. During this time, the design group was not able to successfully converge all the equipment on one flowsheet. After further investigation, we found that this was a common issue in Aspen Plus due to the property package requirements for our process (Al-Malah, 2017). The property package requirements for the units are different for several reasons, including operating pressure and

binary interactions (Al-Malah, 2017). As a result, our design group performed the simulations separately for each column.

Initially, we modeled our reactor as a stoichiometric reactor to obtain the appropriate flowrates for our desired production rate. As we remodeled the reactor as a packed bed tubular reactor using kinetics, we found out that the initial assumptions we made regarding the reaction, mentioned in section 5, were incorrect. Unfortunately, by the time we came to this realization, it was too late to fix.

To model the high-pressure ethylene oxide absorption column (T-101), we used the Predictive Redlich-Kwong-Soave (PSRK)un property package, which is particularly useful in high-pressure applications (Al-Malah, 2017). Our simulation converged successfully without any errors or warnings. We validated the model by comparing our results with our hand calculations from the material and energy balance and found that they aligned closely. After we confirmed the results, we used the Aspen hydraulics feature to perform a rigorous model of the column, which also successfully converged.

When we modeled the ethylene oxide stripping column (T-301), we simulated the tower with multiple recommended property packages, including the Non-Random Two-Liquid (NRTL), PSRK, and Wilson property methods (Adams,2018). When we compared the simulation results with our hand calculations, the PSRK method showed the most accurate results. We performed a sensitivity analysis to determine the steam flowrate into the column. After we validated the results from the property package, we did rigorous modeling of the column internals hydraulic data in Aspen. We found that we were operating on the operating line with no flooding or weeping.



The design team performed multiple simulations to determine the most accurate property method for the light ends column (T-302). The recommended property packages for this column were the NRTL and Wilson property packages (Aspen Plus User Guide, 2000). When we compared the property packages, we found that the NRTL property package showed the most accurate results. Therefore, the design team used the NRTL property package to model this column. Once we validated our model, we optimized the model by varying the feed stage location, reflux ratio, and the number of stages. We noticed we noticed a small error in the light ends overhead vapor stream. We believe the error was due to our specified distillate rate in Aspen, which was extremely sensitive to error. *why?*

We modeled the ethylene oxide distillation column (T-303) using the NRTL property package, which was the recommended property package for this column (Al-Malah, 2017). Although the results from our simulation were validated, we ran into issues when we tried to obtain the column diameter using the hydraulic analysis feature in Aspen. We had no warnings or errors in our simulation, but Aspen reported the diameter to be 12 ft, which is too large for a column that is only 31 ft tall. We believe that the errors in our hydraulic analysis were due to the way we input the packing information in our simulation.

8. Chemical Engineering Design

8.1 Design of Major Equipment

Ethylene Oxide Reactor (R-101)

Our group conducted a detailed design and analysis on several major pieces of equipment, including the ethylene oxide reactor (R-101), ethylene oxide absorption column (T-101), ethylene oxide stripping column (T-301), the light-ends column (T-302), and the ethylene oxide

distillation column (T-303). The following section will provide a brief overview of our detailed equipment design.

The following design equations were used for the design of the ethylene oxide reactor, R-101. We used the following equations to design our ethylene oxide reactor (Fogler, 20018).

$$\text{Total Bed Volume (ft}^3\text{)} = \text{Bed diameter (ft)} \times \text{Cross Sectional Area (ft}^2\text{)} \quad (8.1)$$

$$\text{Total Mass of Catalyst (lbm)} = \text{Catalyst Density } \left(\frac{\text{lb}}{\text{ft}^3}\right) \times \text{Total Bed Volume (ft}^3\text{)} \quad (8.2)$$

$$\text{Bed Volume Per Tube (ft}^3\text{)} = \text{Cross-Sectional area of tube} \times \text{Length} = \pi \left(\frac{D^2(\text{ft})}{4}\right) \times h(\text{ft}) \quad (8.3)$$

$$\text{Catalyst Volume Req (ft}^3\text{)} = \frac{\text{Total Amount of Catalyst (lb)}}{\text{Density of Catalyst } \left(\frac{\text{lb}}{\text{ft}^3}\right)} \quad (8.4)$$

$$\text{Total Number of Tubes} = \frac{\text{Catalyst Volume Req (ft}^3\text{)}}{\text{Bed Volume per Tube (ft}^3\text{)}} \quad (8.5)$$

$$\text{Number of Tubes per Reactor} = \frac{\text{Total Number of Tubes}}{8 \text{ Reactors}} \quad (8.6)$$

$$\text{Catalyst Mass per Tube (lbm)} = \frac{\text{Total Mass of Req Catalyst (lb)}}{\text{Total Number of Tubes}} \quad (8.7)$$

The design of this reactor was initially done as a singular, packed bed reactor to obtain the required volume of catalyst for the reaction to take place. Our reactor is a multi-tubular packed bed reactor so the amount of catalyst found for the singular packed bed was used to back-calculate the amount of catalyst that can be present in each individual tube by first determining the amount of bed volume each tube can hold. Reactor dimensions were obtained from a member in industry that previously worked on an ethylene oxide expansion project for one of the top 5 chemical companies in the world. The values for these dimensions are presented in the table below.

these should be calculated ~~and~~ for comparison.

Table 8. Ethylene Oxide Reactor Dimensions

Reactor Diameter	30 ft
Reactor Length	70 ft
Tube Diameter	0.083 ft
Tube Length	70 ft

The total number of tubes required for this reaction is 55,543. The total diameter of 30 feet for each reactor determined that we need 8 reactors with 6,943 tubes in each. The tubes will be arranged in a triangular formation to reduce the overall diameter that is taken up by the tubes. ✓

While the heuristics that we were required to use stated we should refrain from a diameter greater than 13.5 feet, we would require more than twice as many reactors which is not typical in this industry according to our ethylene oxide contact.

High Pressure Ethylene Oxide Absorber (T-101)

The purpose of our ethylene oxide absorption column is to recover 99% of the ethylene oxide in the gaseous reactor effluent. The solvent used to recover the ethylene oxide is water. Our design group assumed the ratio of water to ethylene oxide to be 22:1 (molar), which we mentioned previously in this report

During the preliminary design phases of the ethylene oxide absorption column, the design group evaluated whether we would use trays or packing in the column. Literature indicates that structured packings produce minimal pressure drop when compared to trays, which is important in units treating a large volume of gas (Weston, 2012). The design group used the method of transfer units to calculate the height of the column and the number of transfer units.

Determination of the packed height can be either gas or liquid phase (Seader & Henley, 1998)

awkward.

For the design of this column, the gas phase transfer units were considered. We used the following equations in our detailed design of the absorption column (Seader & Henley, 1998).

$$N_{OG} = \frac{A}{A-1} \ln \left\{ \frac{(A-1)(y_{in}-K*x_{in})}{A(y_{out}-K*x_{in})} + \frac{1}{A} \right\} \quad (8.8) \quad \checkmark$$

$$H_{OG} = \frac{V}{K_y a * S} \quad (8.9)$$

Using the formulas above, we calculated the number of gas-phase transfer units to be 9 transfer units with about 13 feet of packing per section of packed height. We calculated the total height of the 127 ft (117 ft of packing plus 5 ft added to the top and bottom of the column) and assumed a diameter of 7 feet initially but after simulating in ASPEN Plus, we found an optimal diameter of 12.5 feet for this column. 

Ethylene Oxide Stripper (T-301)

During the design of our ethylene oxide stripper (T-301), we also used the method of transfer units, except we focused on the number of transfer units required for the liquid phase because the liquid phase is being stripped off (Seader & Henley, 1998). Our design group also decided to model the stripper with structured packing because structured packings occupy less height than trays, have a smaller pressure drop, and optimize the energy consumption of the column (Weston, 2012). We calculated the liquid phase transfer units to be about 5 with about 12 feet of packing per section of packed height, we also assumed a diameter of 13.5 feet. We used the following equations to design T-301 (Seader & Henley, 1998).

$$N_{OL} = A * \ln \left[\left(\frac{(A-1)(y_{in}-K*x_{in})}{A(y_{out}-K*x_{in})} \right) + \frac{1}{A} \right] \quad (8.10)$$

$$H_{OL} = \frac{L}{K_x a * S} \quad (8.11)$$

We calculated 11 transfer units and a height of packing per stage of 5.5 feet, the total height of the column is 70.5 feet (60.5 feet of packing with 5 feet added to the top and bottom of the column) and the diameter is 13.5 feet.

We designed the light-ends column, T-302, as a multi-component distillation column and used the following equations:

$$N = \frac{\log \left(\frac{LK_d}{HK_d} \cdot \frac{HK_b}{LK_b} \right)}{\log(\alpha_{avg})} + 1 \quad (8.12) \quad \checkmark$$

$$\alpha_{avg} = \sqrt{\alpha_{top} * \alpha_{bot}} \quad (8.13) \quad \checkmark$$

$$\alpha_{top} = \frac{y_{LK}/x_{LK}}{y_{HK}/x_{HK}} \quad (8.14) \quad \checkmark$$

$$\alpha_{bot} = \frac{y_{LK}/x_{LK}}{y_{HK}/x_{HK}} \quad (8.15) \quad \checkmark$$

We used this method since this is a multicomponent distillation with two heavy key and three light key components. We determined that the light key, highest boiling of the other light components, is carbon dioxide and the heavy key, lowest boiling of the heavier components, is ethylene oxide. We calculated the relative volatilities using those two components and calculated 5 theoretical stages, 10 real stages and a height of packing of about 7 feet. We realized this sizing was small with respect to the flowrates we have entering the column so we finalized the design using ASPEN Plus where we obtained a total height of 22.5 feet (12.5 feet of packing with 5 feet added to the top and bottom of the column) and a diameter of 12.5 feet.

why not use a flash drum for this??

We designed the ethylene oxide distillation column, T-303, using the HETP method. We used this method since the column is a packed distillation column and used the following equations:

$$\alpha = \frac{y(1-x)}{x(1-y)} \quad (8.15)$$

$$R_{op} = 1.2R_{min} \quad (8.16)$$

$$y = \frac{q}{q-1} x_i - \frac{1}{q-1} x_F \quad (8.17)$$

$$y = \frac{R}{1+R} + \frac{1}{1+R} x_D \quad (8.18)$$

$$y = \frac{\bar{L}}{\bar{V}} x - \frac{B}{\bar{V}} x_B \quad (8.19)$$

$$H_{packed} = N_{ideal} * HETP \quad (8.20)$$

We used these equations to determine the number of ideal stages. We did this by first using equations 8.17 through 8.19 to plot on a graph for the McCabe Thiele method to determine the number of theoretical stages. We then multiplied the number of stages by the HETP factor for structured packing, which was 0.66 feet (Seader & Henley, 1998). The number of ideal stages we found for T-303 is 6 and the total height of packing is 3 feet given. Since this design is incredibly small, we had to finalize the design using ASPEN Plus. We found the total height to be 31 feet (21 feet of packing with 5 feet added to the top and bottom of the column).

8.2 Design Equipment Specification Sheets

Table 9. R-101 Specification Sheet

Identification:		Item:	Reactor
		Item No: R-101	Date: 4/24/2020
		No Required: 8	
Function:		Convert ethylene and oxygen to produce to ethylene oxide <i>phase?</i>	
Operation:		Continuous	
Materials Handled:		Feed	Outlet
Total flow (MMSCF/hr):		19.1	18.5
Component Flows (MMSCF/hr)			
Ethylene		7.7	6.3
Oxygen		3.8	2.3
Methane		7.6	7.6
Ethylene Oxide			1.2
Water			0.6
Carbon Dioxide			0.6
		Operating	Design
Temperature (F):		480	720
Pressure (psia):		230	480
Design Data: PER REACTOR			
Height	70 ft	Material	
Total Diameter	30 ft	Shell	316L SS clad carbon steel
# Tubes	6943	Tubes	316 L SS
Tube I.D	0.083 ft	Tubesheet	316L SS clad carbon steel
Tube Length	70 ft		
Tube Formation	Triangular	Q(MMBTU/hr)	14.65
Amount of Catalyst (lbs)	108,000		

Table 10. T-101 Specification Sheet

Identification:		Item:		Ethylene Oxide Absorber	
		Item No:		T-101	
		No Required:		1	
				Date: 4/24/2020	
Function:		Recover 99% of Ethylene Oxide by scrubbing with water			
Operation:		Continuous			
		Inlet		Outlet	
		Feed		Solvent	
				Overhead	
				Bottoms	
Materials Handled:					
Total flow (MMSCF/hr):		18.4		0.02	
				17.16	
				1.248	
Component Flows (MMSCF/hr)					
Ethylene		6.1		6.09	
Oxygen		2.3		2.30	
Methane		7.6		7.59	
Ethylene Oxide		1.2		0	
Water		0.6		0.02	
Carbon Dioxide		0.6		5.99E-01	
				7.59E-04	
		Operating		Design	
Temperature (F):		75		112.5	
Pressure (psig):		215		480	
Design Data:					
Tower Height (ft)		127			
Tower Diameter (ft)		12.5			
Total Packing Height (ft)		117			
Packing Diameter (ft)		12.5			
		Internal		External	
Material		316L SS clad to carbon steel		Carbon steel	
Packing		Mellpak 125X			

Table 11. T-301 Specification Sheet

Identification:	Item:	Ethylene Oxide Stripper			
	Item No:	T-301		Date:	4/24/2020
	No Required:	1			
Function: Recover 99% Ethylene Oxide and regenerate T-101 Solvent by steam stripping					
Operation: Continuous					
		Inlet		Outlet	
		Feed	Steam	Overhead	Bottoms
Materials Handled:					
Total flow (MMSCF/hr):		1.248	5.3	6.53	0.021
Component Flows (MMSCF/hr)					
Ethylene		0.006		0.006	
Oxygen		0.002		0.002	
Methane		0.008		0.008	
Ethylene Oxide		1.21		1.2	0.012
Water		0.021	5.3	5.3	0.02
Carbon Dioxide		7.59E-04		7.59E-04	
		Operating	Design		
Temperature (F):		150	225		
Pressure (psig):		15	400		
Design Data:					
Tower Height (ft)		70.5			
Tower Diameter (ft)		13.4			
Total Packing Height (ft)		60.5			
Packing Diameter (ft)		13.4			
		Internal	External		
Material		316L SS clad to carbon steel	Carbon steel		
Packing		Mellapak 125X			

Table 12. T-302 Specification Sheet

Identification:	Item:	Light-ends Column	
	Item No:	T-302	Date: 4/24/2020
	No Required:	1	
Function:	Remove VOC from Ethylene Oxide and Water Stream		
Operation:	Continuous		
	Inlet	Outlet	
	Feed	Overhead	Bottoms
Materials Handled:			
Total flow (MMSCF/hr):	6.5	0.017	6.5
Component Flows (MMSCF/hr)			
Ethylene	0.006	0.006	
Oxygen	0.002	0.002	
Methane	0.008	0.008	
Ethylene Oxide	1.2	-	1.2
Water	5.3	-	5.3
Carbon Dioxide	7.49E-04	0.00075	
	Operating	Design	
Temperature (F):	75	112.5	
Pressure (psig):	80	400	
Design Data:			
Tower Height (ft)	22.5		
Tower Diameter (ft)	5.68		
Total Packing Height (ft)	12.5		
Packing Diameter (ft)	5.68		
	Internal	External	
Material	316L SS clad to carbon steel	Carbon steel	
Packing	Mellapak 125Y		

Table 13. T-303 Specification Sheet

Identification:	Item:	Ethylene Oxide Distillation	
	Item No:	T-303	Date: 4/24/2020
	No Required:	1	
Function:	Recover Ethylene Oxide Product with 99.7% purity by distillation		
Operation:	Continuous		
	Inlet	Outlet	
	Feed	Overhead	Bottoms
Materials Handled:			
Total flow (MMSCF/hr):	6.5	1.18	5.32
Component Flows (MMSCF/hr)			
Ethylene			
Oxygen			
Methane			
Ethylene Oxide	1.2	1.18	0.02
Water	5.3	2.70E-05	5.3
Carbon Dioxide			
	Operating	Design	
Temperature (F):	75	112.5	
Pressure (psig):	0	400	
Design Data:			
Tower Height (ft)	31		
Tower Diameter (ft)	12.5		
Total Packing Height (ft)	21		
Packing Diameter (ft)	12.5		
	Internal		External
Material	316L SS clad to carbon steel		Carbon steel
Packing	Mellapak 452Y		

9. Plant Economic Evaluation

9.1 Capital Cost

The capital cost of our ethylene oxide plant includes the total equipment cost, FCI, fixed capital investment, and the cost of the land required to build this plant. The total cost of equipment, which includes what we calculated using CAPCOST software as well as the extra cost of adding clad steel and the cost of packing required for our packed columns is about \$1.4 million. The FCI that we used is a total of \$1.5 billion. This FCI was given to us by our ethylene

oxide contact as a reference for a complete ethylene oxide plant. The FCI that we calculated using the CAPCOST software was about \$1.5 million which is off from the reference by three orders of magnitude. Therefore, we used the FCI given to us to observe a more realistic cash flow analysis. The FCI includes all of the elements that were outside of the scope of this analysis. These elements include pipelines within the plant, cost of other pieces of equipment such as pumps, compressors and heat exchangers, and the cost of the buildings to house employees as well as the structures around specific units such as scaffolding. The amount of land that we require for our plant is about 40 acres, this information was obtained from our ethylene oxide contact. The cost of land for industrial use is about \$40 thousand an acre. The total cost of land required is about \$1.6 million.

why??

units

9.2 Operating Cost

The total cost of operating the plant includes the cost of utilities, such as cooling water, the cost of personnel to run the process units, otherwise known as the working capital, WC, and the cost of raw materials. The total cost of the raw materials that we require is about \$1.9 billion. While the cost of ethylene has dropped dramatically since the beginning of the COVID-19 crisis, about \$0.03 per pound, we used the cost of ethylene prior to this pandemic, about \$0.56 per pound. We received raw materials cost information from a utilities engineer that works intimately within the markets of the raw materials that we require, ethylene, methane, and oxygen. We calculated the total cost of the WC using an excel spreadsheet that was used by a prior student in the PVAMU chemical engineering department which helped us determine the optimum number of operators and administrative positions required for our process. The total cost of the WC was about \$2.3 million. The total cost of utilities, which we based primarily on the water needed for the ethylene oxide reactor and absorption column since that is the bulk of

the utility usage, was about \$64.8 million for about 1.3 billion gallons per year at \$0.05 per gallon. We assumed that the water we will be using is received from a local water source in the Texas Gulf Coast area, such as Chocolate Bayou, where the water will need extensive treatment in order to be at the purity specification required for our plant.

9.3 Cash Flow Diagram and ROROI

The table below represents a cost breakdown that was used to develop the cash flow diagram that will be discussed in this section. The profit that we expect to receive from selling our ethylene oxide was calculated using the CAPCOST software in excel. *ok*

Table 14. Grass-roots Ethylene Oxide Cost Breakdown

Land:	\$1,600,000
FCI:	\$1,500,000,000
WC:	\$2,261,950.34
Salvage:	\$0.00
Equipment & Raw Materials	\$1,941,601,419.36
Profit:	\$567,560,247

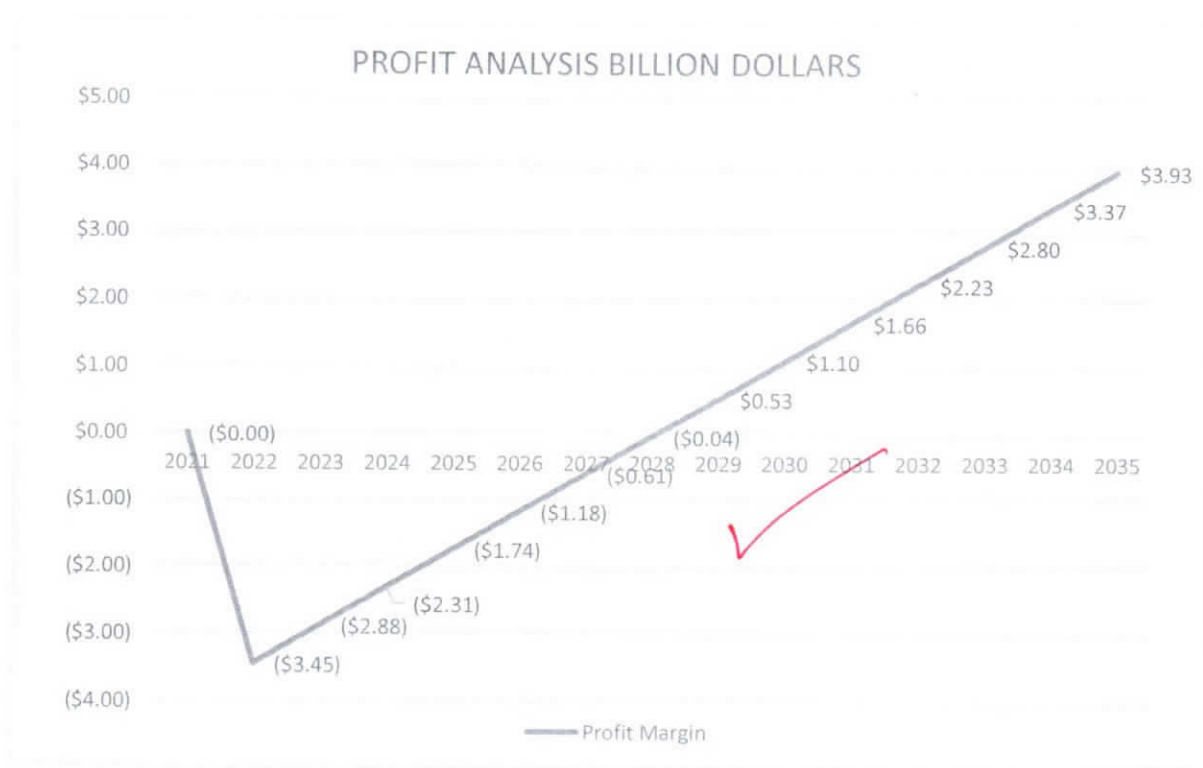


Figure 12. Grass-roots Ethylene Oxide 15-year Cash Flow Diagram

The graph above is the 15-year cash flow diagram of our project. This cash flow diagram illustrates a break-even point about 7 years after the initial startup. The ROROI, rate of return on investment, of our project was calculated using the following equation.

$$\text{ROROI} = \frac{\text{Profit per year}}{\text{FCI}} \quad (9.1)$$

We calculated an ROROI of 38% which exceeds the hurdle rate of 15%. Therefore, we recommend this project proceeds as designed.

10. Safety and HAZOP Analysis

The figure below represents the control scheme that we established for our most critical piece of equipment, the reactor, R-101.

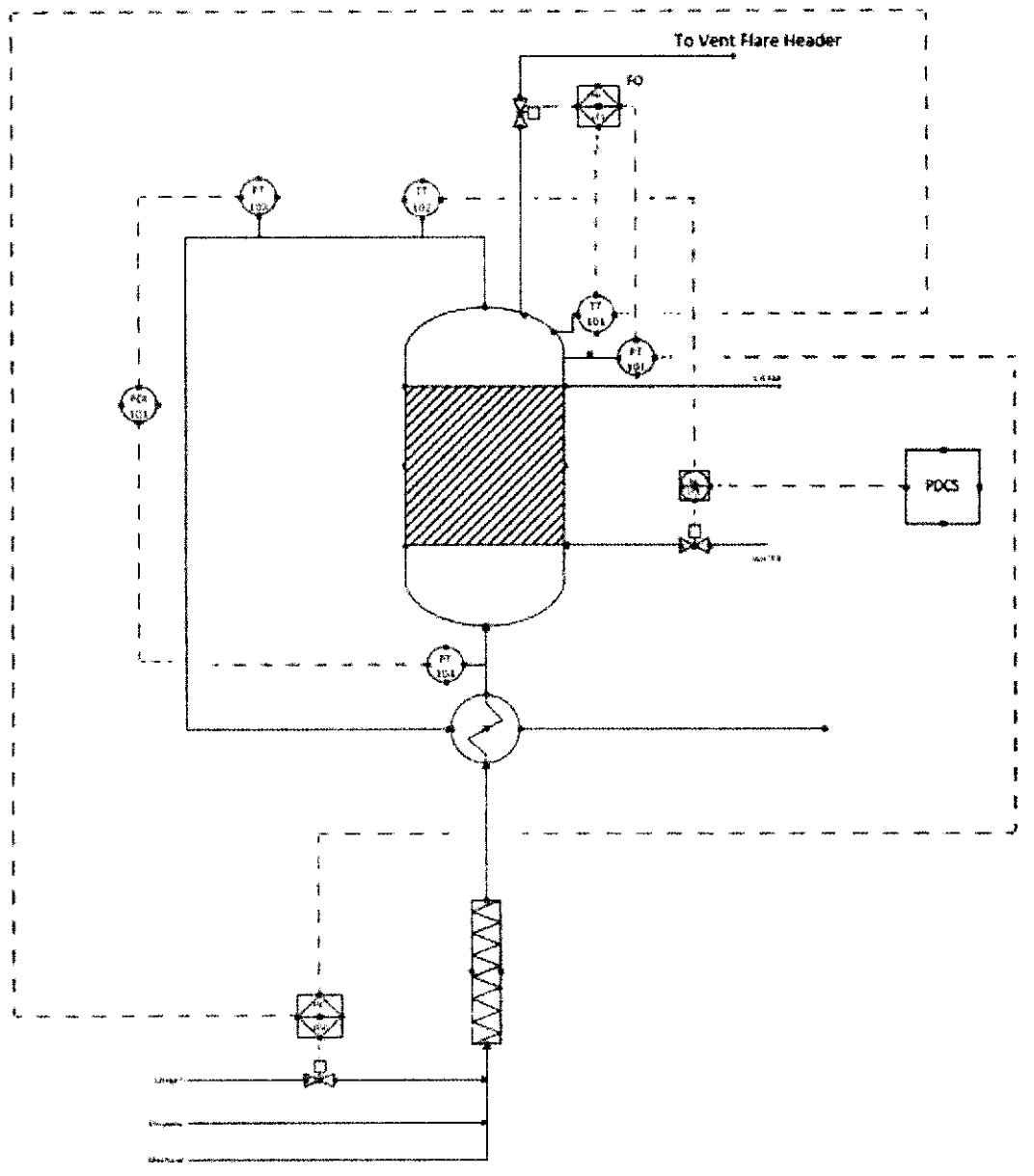


Figure 13. Reactor, R-101 Control Scheme

Our reactor is the most critical piece of equipment due to the high pressure, 230 psia, high temperature, 480 °F, and the exothermic reaction occurring therein. Also, since our reactor is several thousand packed tubes, there is a special interest in the pressure drop that occurs from the

inlet to the outlet. This is mitigated by installing two pressure transmitters, one at the inlet, PT-104, and one at the outlet, PT-103, that relays information to the pressure differential indicator, PDI 101. This will tell operators at the control board if something is wrong with the tube internals such a blockage from disturbed catalyst. *language.*

The energy released during the reaction is mitigated by the flow controller located at the inlet of the cooling water stream, FC-101. A temperature transmitter, TT-102, is connected to the flow controller and will relay information back to the control board through the process digital control system, PDCS, which will prompt the operators to increase the flowrate of cooling water into the reactor. *this makes little sense.*

In the event of an uncontrollable high pressure and high temperature incident, such as a runaway reaction, the event indicting controllers, YIC-101 and -102, which are connected to pressure transmitter, PT-101, and temperature transmitter, TT-101, will activate. YIC-102 is located on the oxygen feed line and will close which will stop the reaction from occurring. YIC-101 is located on an exit line from the reactor which will relieve the pressure inside the reactor by removing the runaway reaction products, mainly water and carbon dioxide, by sending it to the vent flare header. This control valve will also fail open in the event the valve fails.

The table below represents the detailed HAZOP analysis that was conducted for our reactor.

Very Good!

Table 15. HAZOP Analysis around Reactor, R-101

Deviation		Possible Causes	Consequences	Recommendation
Parameter	Guideword			
Flow to the reactor	No	Feed inlet control valve failed closed	No reaction takes place	Install low flow alarm to monitor and regularly measure the flow entering the reactor
	More	Valve failed open	Too much flow into the reactor causes a runaway heat reaction and possible explosion	Install high flow alarm and backup control valve
	Less	Leaking of pipe	Desired amount not produced possible accumulation in pipeline can lead to potential explosion	Install Pressure indicator and bypass valve
Pressure in the reactor	Less	High pressure downstream due to pump/compressor failure	EO line could back flow into reactor and uncontrolled amounts of oxygen and ethylene oxide could cause Runaway reaction	Monitor reactor temperature and pressure on DCS system and have an emergency shutdown procedure in place. Add thermal jacket to reactor
	More	Reactor material/valve failure	Temperature will increase runaway reaction takes place leading to potential explosion	Monitor reactor temperature and pressure on DCS and have emergency shutdown procedures. Add thermal jacket to reactor. Include pressure release valve to reactor.
Temperature in the reactor	Low	Cooling water valve failed open	Reaction kinetics would be affected, and pressure could decrease	Install temperature alarm for jacketed cooling. Add thermal control jacket to reactor. Monitor reactor temperature and pressure on DCS and have an emergency shutdown procedure in place
	High	Cooling water valve failed closed	Increase Temperature in reactor could cause Explosion	Add thermal control jacket to reactor and monitor reactor temperature and pressure on DCS system and have automatic shutdown procedures

11. Discussion

We were able to design our plant preliminarily using the design equations for each respective unit given in section 8. We determined that some of our calculations, mainly the refining columns T-302 – T-303 were not correct and were able to refine the designs using ASPEN Plus. The economic analysis we did for this plant only considers the equipment we designed, it does

sensitivity analysis, optimization, other analysis missing!!

not account for any of the heat exchangers, pumps, compressors or the pipes needed to transfer materials throughout the plant and provide the necessary unit operations to bring the materials to the appropriate conditions. We also manually calculated the cost of clad steel required for each of our process units as well as the cost of packing for the packed columns. The FCI was given to us by our ethylene oxide contact which we assumed accounts for all the components we did not reference for this report.

12. Conclusions and Recommendations

What are key hurdles and obstacles??

Overall, we recommend this project move forward given the economic analysis. Our plant will have a rate of return of 38% and will achieve a break-even point about 7 years after the initial startup. However, when we combined our Aspen flowsheets, we noticed that some of the temperatures and pressures of the columns in the ethylene oxide refining section need to be redesigned, due to large pressure differences and high utility cost. First and foremost, we recommend taking another look into the kinetics of this reaction to appropriately model the reactor. We also recommend increasing the operating pressure of the ethylene oxide stripping column and redesigning the column as a distillation unit. Due to unrealistic column dimensions, we also recommend that the ethylene oxide light ends column (T-302) and the ethylene oxide distillation column (T-303) be redesigned. Our column dimensions for these two columns does not appear to be realistic. The column heights seem too small and the diameters too large with respect to the height. To improve the accuracy of the economic evaluation, we recommend obtaining a more accurate value for the cost of the clad steel and packing used in our design. Finally, we recommend that another economic evaluation be completed after the process modifications have been completed.

13. References

you didn't do this why?

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14. Appendix

14.1 Sample Calculations

Reactor (R-101)

Good.

$$\text{Total Bed Volume} = 30 \text{ (ft)} \times 707 \text{ (ft}^2\text{)} = 21,206 \text{ (ft}^3\text{)}$$

$$\text{Total Mass of Catalyst} = 40.36 \left(\frac{\text{lb}}{\text{ft}^3} \right) \times 21,206 \text{ (ft}^3\text{)} = 860,105 \text{ (lbs)}$$

$$\text{Bed Volume per Tube} = \pi \left(\frac{0.083333 \text{ (ft)}}{2} \right)^2 \times 70 \text{ (ft)} = 0.3818 \text{ (ft}^3\text{)}$$

$$\text{Total Number of Tubes} = \frac{21,206 \text{ (ft}^3\text{)}}{0.3818 \text{ (ft}^3\text{)}} = 55,543 \text{ Tubes}$$

$$\text{Number of Tubes per Reactor} = \frac{55,543}{8} = 6,943$$

$$\text{Catalyst Mass per Tube} = \frac{860,105}{55,543} = 15.5 \text{ lb}$$

Ethylene Oxide Absorber (T-101)

$$A = \frac{L}{K \times V} = \frac{71,000 \left(\frac{\text{lbmol}}{\text{hr}} \right)}{0.85 \times 48,735 \left(\frac{\text{lbmol}}{\text{hr}} \right)} = 1.714$$

$$N_{OG} = \frac{1.714}{(1.714 - 1)} \ln \left(\frac{(1.714 - 1)(0.07 - (0.85 \times 0))}{1.714(0.0007 - (0.85 \times 0))} + \frac{1}{1.714} \right) = 2.4 \ln(42.24) = 8.98$$

$\cong 9$ Transfer Units

$$H_{OG} \text{ (ft)} = \frac{V}{K_{ya} \times S} = \frac{48,735 \left(\frac{\text{lbmol}}{\text{hr}} \right)}{98.51 \times 38.5 \text{ ft}^2} = 12.849 \cong 13 \text{ ft Height}$$

$$K_{ya} = \left(\frac{1}{k_{ya}} + \frac{K}{k_{xa}} \right)^{-1} = \left(\frac{1}{200} + \frac{0.85}{165} \right)^{-1} = 98.51 \left(\frac{\text{lbmol}}{\text{ft}^3 \text{ hr}} \right)$$

$$\text{Packing Height} = 9 \times 13 \text{ ft} = 117 \text{ ft}$$

Ethylene Oxide Stripping Column (T-301)

$$A = \frac{L}{K * V} = \frac{75,853 \frac{\text{lbmol}}{\text{hr}}}{0.85 * 11,800 \frac{\text{lbmol}}{\text{hr}}} = 7.567$$

$$N_{OL} = A * \ln \left[\left(\frac{(A-1)(y_{in} - K * x_{in})}{A(y_{out} - K * x_{in})} \right) + \frac{1}{A} \right]$$

$$= (7.567) \ln \left[\left(\frac{(6.567 - (0.85 * 0.042))}{7.567(0.28 - (0.85 * 0.42))} \right) + \left(\frac{1}{7.567} \right) \right] = 10.8 = 11$$

$$H_{OL} = \frac{L}{K_x a * S} = \frac{75,895 \frac{\text{lbmol}}{\text{hr}}}{96.99 \frac{\text{lbmol}}{\text{hr} * \text{ft}^3} * 143.1 \text{ft}^2} = 5.5 \text{ ft}$$

$$S = \frac{\pi * d^2}{4} = \frac{\pi * (13.5 \text{ft})^2}{4} = 143.1 \text{ft}^2$$

$$K_x a = \left(\frac{1}{k_x a} + \frac{K}{k_y a} \right)^{-1} = \left(\frac{1}{165 \frac{\text{lbmol}}{\text{hr} * \text{ft}^3}} + \frac{0.85}{200 \frac{\text{lbmol}}{\text{hr} * \text{ft}^3}} \right)^{-1} = 96.99 \frac{\text{lbmol}}{\text{hr} * \text{ft}^3}$$

$$\text{Total packing height} = N_{OL} * H_{OL} = 11 * 5.5 \text{ft} = 60.5 \text{ft}$$

Light-Ends Column (T-302)

$$\alpha_{bot} = \frac{y_{LK}/x_{LK}}{y_{HK}/x_{HK}} = \frac{0.00049548/0.000036562}{0.96013/0.38507} = 5.435078609$$

$$\alpha_{top} = \frac{y_{LK}/x_{LK}}{y_{HK}/x_{HK}} = \frac{0.12065/0.071129}{0.0092291/0.60693} = 111.547509$$

$$\alpha_{avg} = \sqrt{\alpha_{top} * \alpha_{bot}} = \sqrt{111.547509 * 5.435078609} = 24.62254008$$

$$N = \frac{\log \left(\frac{LK_d}{HK_d} * \frac{HK_b}{LK_b} \right)}{\log(\alpha_{avg})} = \frac{\log \left(\frac{0.7592}{0.00922909} * \frac{0.265919}{0.0000909418} \right)}{\log(24.62254008)} + 1 = 5$$

Ethylene Oxide Distillation Column (T-303)

Equilibrium Data for Ethylene Oxide-Water Data was obtained from Aspen Plus V9

Property Package: NRTL

The y-x diagram was generated at 0 psig

equilibrium curve	
x	y
0	0
0.1	0.702
0.2	0.879
0.3	0.94
0.4	0.967
0.5	0.981
0.6	0.988
0.7	0.993
0.8	0.996
0.9	0.999
1	1

Relative Volatility with Respect to More Volatile Component)

$$\alpha = \frac{y(1-x)}{x(1-y)}$$

$$\alpha = \frac{0.702(1-0.1)}{0.1(1-0.702)}$$

$$\alpha = 21.2$$

Operating Reflux

$$R_{op} = 1.2R_{min}$$

$$R_{op} = 1.2$$

McCabe Thiele Method for EO Distillation Column (T-303)

Rectifying Section Operating Line

$$y = \frac{R}{1+R} + \frac{1}{1+R} x_D$$

$$y = 0.545 + .453 x_D$$

$$y = 0.545 + 0.453(.997)$$

$$y = 0.545 + 0.453 (1)$$

Feed Line Section Operating Line

$$y = \frac{q}{q-1} x_i - \frac{1}{q-1} x_F$$

$$y = \frac{0.82}{0.82-1} x_i - \frac{1}{0.82-1} x_F$$

$$y = -4.56 x_i + 5.56 x_F$$

$$y = -4.56 x_i + .56(0.266)$$

$$y = -4.56 x_i + 1.47 (2)$$

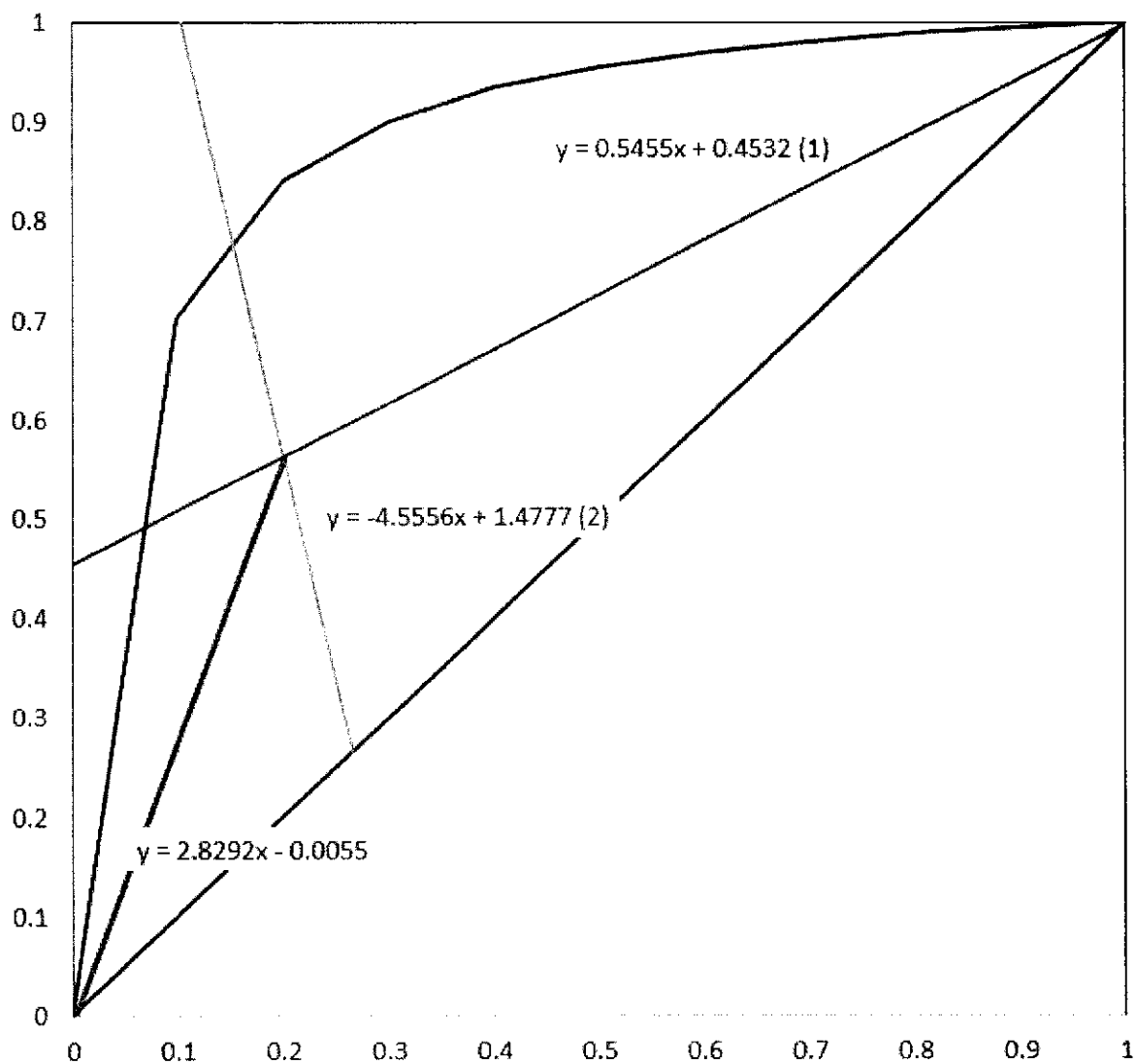
Stripping Section Operating Line

$$y = \frac{\bar{L}}{\bar{V}} x - + \frac{B}{V} x_B$$

$$y = 2.83x - 0.55 (3)$$

McCabe Thiele Plot for EO Distillation Column (T-303)

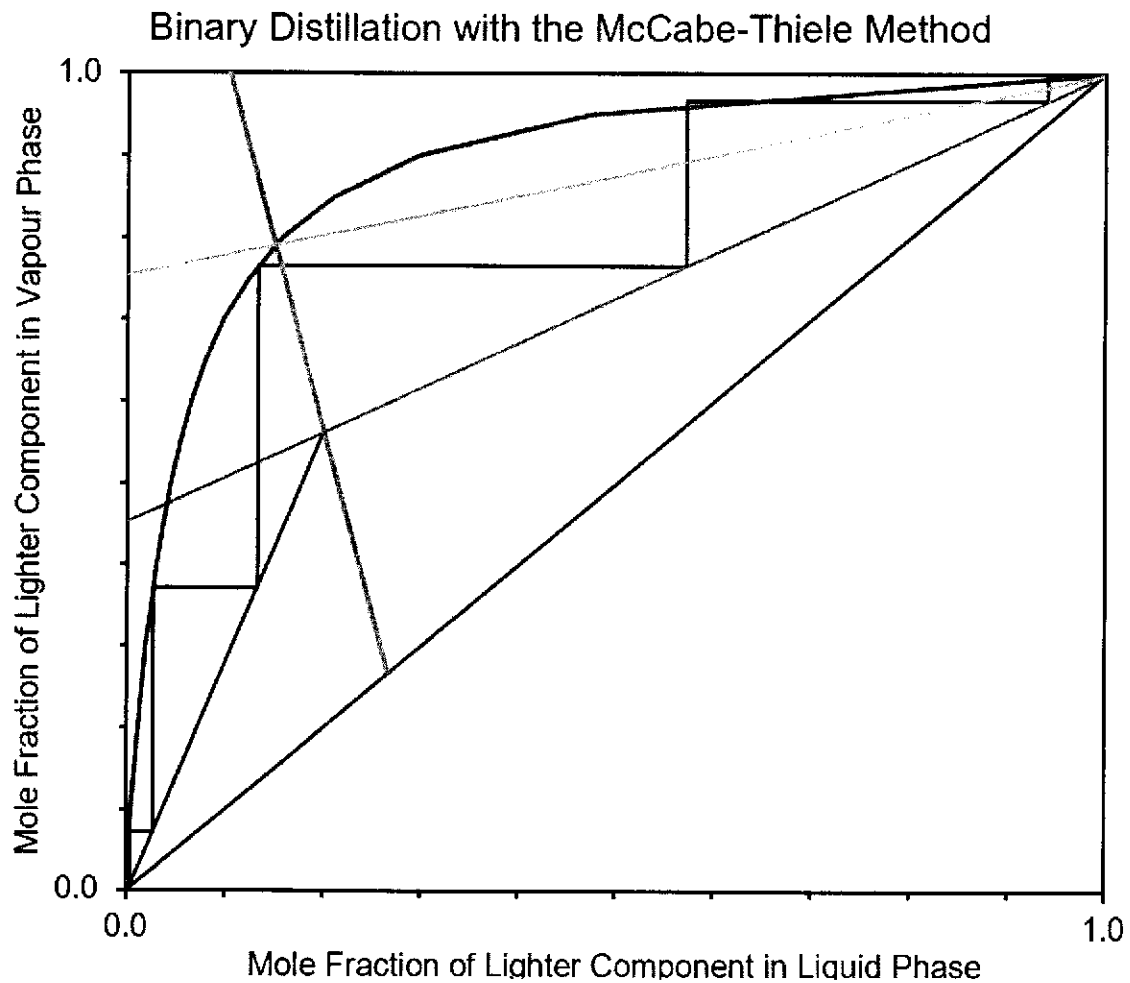
YX- For EO-WATER Mixture



Rectifying Section Operating Line $y = 0.545 + 0.453$ (1)

Feed Line Section Operating Line $y = -4.56 x_t + 1.47$ (2)

Stripping Section Operating Line $y = 2.83x - 0.55$ (3)



Height Equivalent to Theoretical Plate (HETP)

$$H_{packed} = N_{ideal} * HETP$$

Structured Packing HETP=0.66 ft (Seader & Henley, 1998)

$$H_{packed} = 5 * 0.66 \text{ ft}$$

$$H_{packed} = 3.3 \text{ ft}$$

14.1.1 Explanation of High-Pressure Ethylene Oxide Absorption Column T-101

Calculations

Number of transfer units, N_{OG} :

$$N_{OG} = \frac{A}{A-1} \ln \left\{ \frac{(A-1)(y_{in} - K * x_{in})}{A(y_{out} - K * x_{in})} + \frac{1}{A} \right\}$$

Where:

y_{in} = mole fraction of EO entering in the gas stream; 0.07

y_{out} = mole fraction of EO exiting in the gas stream; 0.0007

x_{in} = mole fraction of EO entering in the liquid stream; 0

K = dimensionless equilibrium distribution coefficient; 0.85

A = Absorption factor; $A = \frac{L}{KV}$

Where:

L = solvent flowrate entering (lbmol/hr)

V = vapor flowrate entering (lbmol/hr)

Height of transfer units, H_{OG} :

$$H_{OG} = \frac{V}{K_y a * S}$$

Where:

S = cross-sectional area of the column in ft^2 ; 38.5

$K_y a$ = overall mass transfer coefficient; $K_y a = \left(\frac{1}{k_y a} + \frac{K}{k_x a} \right)^{-1}$

Where:

$k_y a$ = volumetric mass transfer coefficient in the gas phase in $\text{lbmol}/(\text{hr} * \text{ft}^3)$; 200

$k_x a$ = volumetric mass transfer coefficient in the liquid phase in $\text{lbmol}/(\text{hr} * \text{ft}^3)$; 165

Total height of packing

$$\text{Packing height} = N_{OG} * H_{OG}$$

14.1.2 Computer Simulations Used

R-101 Stream Summary

Material			
Stream Name	Units	FEED	EFFLUENT
Description			
From			R-101
To		R-101	
Stream Class		CONVEN	CONVEN
Maximum Relative Error			
Cost Flow	\$/hr		
MIXED Substream			
Phase		Vapor	Vapor
Temperature	F	480	480
Pressure	psig		215
Molar Vapor Fraction		1	1
Molar Liquid Fraction		0	0
Molar Solid Fraction		0	0
Mass Vapor Fraction		1	1
Mass Liquid Fraction		0	0
Mass Solid Fraction		0	0
Molar Enthalpy	Btu/lbmol	491.903	-11968.2
Mass Enthalpy	Btu/lb	20.4445	-481.455
Molar Entropy	Btu/lbmol-R	-10.379	-11.3447
Mass Entropy	Btu/lb-R	-0.431374	-0.45637
Molar Density	lbmol/cuft	0.0228141	0.0228915
Mass Density	lb/cuft	0.548917	0.569047
Enthalpy Flow	Btu/hr	2.47E+07	-5.82E+08
Average MW		24.0604	24.8585
Mole Flows	lbmol/hr	50256.8	48643.3
C2H4	lbmol/hr	20150.6	16116.6
O2	lbmol/hr	1.01E+04	6050.25
CH4	lbmol/hr	2.00E+04	20021.5
C2H4O	lbmol/hr	0	3227
CO2	lbmol/hr	0	1614
H2O	lbmol/hr	0	1614
Mole Fractions			
C2H4		0.400952	0.331322
O2		0.20064	0.12438
CH4		0.398383	0.411598
C2H4O		0	0.06634
CO2		0	0.0331803
H2O		0	0.0331803

T-101 Stream Summary

Stream Name	Units	Material			
		H2O	LIOUT	RGASES	VOUT
Description					
From			T-101		T-101
To		T-101		T-101	
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error					
Cost Flow	\$/hr				
MIXED Substream					
Phase		Liquid	Liquid	Mixed	Vapor
Temperature	F	75	97.0891	75	76.4168
Pressure	psig	215	215	215	215
Molar Vapor Fraction		0	0	0.965063	1
Molar Liquid Fraction		1	1	0.0349373	0
Molar Solid Fraction		0	0	0	0
Mass Vapor Fraction		0	0	0.97058	1
Mass Liquid Fraction		1	1	0.02942	0
Mass Solid Fraction		0	0	0	0
Enthalpy Flow	Btu/hr	-8.73E+09	-8.99E+09	-8.34E+08	-5.71E+08
Average MW		18.0153	19.1402	24.8516	23.6676
Mole Flows	lbmol/hr	71000	75895.1	48735	43839.9
C2H4	lbmol/hr	0	153.517	16155	16001.5
O2	lbmol/hr	0	5.97382	6051	6045.03
CH4	lbmol/hr	0	16.6668	20074	20057.3
C2H4O	lbmol/hr	0	3193.32	3227	33.681
CO2	lbmol/hr	0	24.3893	1614	1589.61
H2O	lbmol/hr	71000	72501.2	1614	112.767
Mole Fractions					
C2H4		0	0.00202275	0.331487	0.364998
O2		0	7.87E-05	0.124161	0.137889
CH4		0	0.0002196	0.411901	0.457513
C2H4O		0	0.0420754	0.0662152	0.00076827
CO2		0	0.00032136	0.0331179	0.0362595
H2O		1	0.955282	0.0331179	0.00257224

T-301 Stream Summary

Stream Name	Units	Material			
		FEED	L	STEAM	V
Description					
From			T-301		T-301
To		T-301		T-301	
Stream Class		CONVEN	CONVEN	CONVEN	CONVEN
Maximum Relative Error					
Cost Flow	\$/hr				
MIXED Substream					
Phase		Mixed	Liquid	Vapor	Vapor
Temperature	F	150	211.903	212	196.62
Pressure	psig	0	0	0	0
Molar Vapor Fraction		0.0360871	0	1	1
Molar Liquid Fraction		0.963913	1	0	0
Molar Solid Fraction		0	0	0	0
Mass Vapor Fraction		0.0686446	0	1	1
Mass Liquid Fraction		0.931355	1	0	0
Mass Solid Fraction		0	0	0	0
Enthalpy Flow	Btu/hr	-8.89E+09	-9.30E+09	-1.44E+09	-1.02E+09
Average MW		19.1402	18.0153	18.0153	24.7503
Mole Flows	lbmol/hr	75895.1	77219.1	14000	12676
C2H4	lbmol/hr	153.517	4.16E-30	0	153.517
O2	lbmol/hr	5.97382	2.36E-45	0	5.97382
CH4	lbmol/hr	16.6668	1.07E-44	0	16.6668
C2H4O	lbmol/hr	3193.32	4.42E-10	0	3193.32
CO2	lbmol/hr	24.3893	1.08E-32	0	24.3893
H2O	lbmol/hr	72501.2	77219.1	14000	9282.11
Mole Fractions					
C2H4		0.00202275	5.38E-35	0	0.0121108
O2		7.87E-05	3.05E-50	0	0.00047127
CH4		0.0002196	1.39E-49	0	0.00131483
C2H4O		0.0420754	5.72E-15	0	0.251919
CO2		0.00032136	1.40E-37	0	0.00192405
H2O		0.955282	1	1	0.73226

T-302 Stream Summary

Stream Name	Units	Material		
		BTMS	FEED	LIGHTS
Description				
From		B2		B2
To			B2	
Stream Class		CONVEN	CONVEN	CONVEN
Maximum Relative Error				
Cost Flow	\$/hr			
MIXED Substream				
Phase		Liquid	Mixed	Vapor
Temperature	F	166.899	75	5.39974
Pressure	psig	80	80	80
Molar Vapor Fraction		0	0.0101417	1
Molar Liquid Fraction		1	0.989858	0
Molar Solid Fraction		0	0	0
Mass Vapor Fraction		0	0.0126573	1
Mass Liquid Fraction		1	0.987343	0
Mass Solid Fraction		0	0	0
Average MW		24.6655	24.7503	29.7842
Mole Flows	lbmol/hr	12466	12676	210
C2H4	lbmol/hr	0.00021838	153.517	153.517
O2	lbmol/hr	8.32E-06	5.97382	5.97381
CH4	lbmol/hr	1.43E-08	16.6668	16.6668
C2H4O	lbmol/hr	3183.86	3193.32	9.46488
CO2	lbmol/hr	0.026934	24.3893	24.3624
H2O	lbmol/hr	9282.09	9282.11	0.0153599
Mole Fractions				
C2H4		1.75E-08	0.0121109	0.731032
O2		6.67E-10	0.00047127	0.0284467
CH4		1.15E-12	0.00131483	0.0793657
C2H4O		0.255404	0.251919	0.0450709
CO2		2.16E-06	0.00192406	0.116011
H2O		0.744594	0.73226	7.31E-05

T-303 Stream Summary

Stream Name	Units	Material		
		FEED	VAPOR	WW
Description				
From			T-303	T-303
To		T-303		
Stream Class		CONVEN	CONVEN	CONVEN
Maximum Relative Error				
Cost Flow	\$/hr			
MIXED Substream				
Phase		Mixed	Vapor	Liquid
Temperature	F	75	50.7563	209.9
Pressure	psig	-1.78E-15	-1.78E-15	-1.78E-15
Molar Vapor Fraction		0.153753	1	0
Molar Liquid Fraction		0.846247	0	1
Molar Solid Fraction		0	0	0
Mass Vapor Fraction		0.270362	1	0
Mass Liquid Fraction		0.729638	0	1
Mass Solid Fraction		0	0	0
Enthalpy Flow	Btu/hr	-1.23E+09	-7.37E+07	-1.12E+09
Average MW		24.6655	43.9905	18.0309
Mole Flows	lbmol/hr	12466	3186	9280
C2H4	lbmol/hr	0.00021838	0.00021838	2.83E-10
O2	lbmol/hr	8.32E-06	8.32E-06	1.51E-15
CH4	lbmol/hr	1.43E-08	1.43E-08	4.44E-17
C2H4O	lbmol/hr	3183.87	3178.31	5.56045
CO2	lbmol/hr	0.026934	0.026934	3.41E-08
H2O	lbmol/hr	9282.11	7.66745	9274.44
Mole Fractions				
C2H4		1.75E-08	6.85E-08	3.05E-14
O2		6.67E-10	2.61E-09	1.63E-19
CH4		1.15E-12	4.49E-12	4.78E-21
C2H4O		0.255404	0.997585	0.00059919
CO2		2.16E-06	8.45E-06	3.67E-12
H2O		0.744594	0.00240661	0.999401

BELOW AVERAGE
PROJECT REPORT

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 2: An ability to apply engineering design to produce solutions that meet specified needs with consideration of public health, safety, and welfare, as well as global, cultural, social, environmental, and economic factor. Assignments that test competency in this area will include problems testing students' understanding of the design process involving devising a system, component, or process to meet desired needs and specifications within constraints, through an iterative, creative, decision-making process in which the basic sciences, mathematics, and engineering sciences are applied to convert resources into solutions.

Name: Gray 3

Title of Assignment: Final Report

Due Date: 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to Define/Understand the Problem and then Plan the Project	60	40
(i) Identify the customer and the needs.	8	5
(ii) Identify and list the design objectives.	8	5
(i) Identify the design constraints.	4	2
(ii) Define the design strategy and methodology.	4	2
(iii) Identify and break down work into tasks and subtasks and identify the personnel and deliverables for each.	20	20
(iv) Develop a Gantt chart and critical path analysis for managing the project.	8	3
(v) Establish major milestones for tracking progress and define performance metrics to measure success.	8	3
2. Ability to Conduct a Review of the Literature	80	60
(i) Identify the types of information needed for a complete understanding of all aspects of the project (Based on task described in the project planning).	20	15
(ii) Gather information on relevant fundamentals, theory / concept (demonstrate technical competence) and relate them to the design.	40	25
(iii) Provide the sources in a list of references properly cited in the literature review section and relevant sections of the report.	20	20
3. Ability to perform preliminary design by generation of ideas and applying creativity through generation of ideas and evaluation of candidate solutions	60	40
(i) Perform process design and analysis with design constraints; Sorting/screening of process technology alternatives based on availability and cost of raw materials, reaction selectivity and yield, thermodynamic barriers to separating undesired components, and safety and operability.	40	30
(ii) Evaluate candidate solutions to arrive at feasible designs.	20	10
4. Ability to Perform Preliminary and Detailed Design	200	117
(i) Prepare a process flow diagram (PFD) and present a detailed description of the process indicating the functions of the major equipment components and controls, stream connectivity, safety, and other technical matters of importance.	20	12
(ii) Carry out a HYSYS/ASPEN simulation of the process showing a converged HYSYS/ASPEN model of the process. State assumptions/parameters used in modeling major equipment.	20	10
(ii) Prepare a chemical engineering design of (i) reactor(s), (ii) absorption column(s) and (ii) product distillation units, stating the materials of construction and the operating temperatures and pressures.	60	40
(iii) Prepare mechanical engineering design of the reactor(s) and distillation columns with appropriate dimensioned sketches.	20	10
(iii) Identify applicable codes and standards for the chemical and mechanical engineering designs. Indicate assumptions made and their justification, and support calculations should be in consistent units, either English Units or SI. Indicate possible sources of error.	20	15
(iv) Economic Evaluation: Estimate capital costs including engineering, materials, installation, interest, return on investment, taxes and contingency. Total unit conversion cost of product, a sensitivity study on the price of the raw material, and the economic methods used should be included.	30	15
(v) Prepare a control scheme to ensure safe operation of the plant and carry out Safety Reviews, Hazards, operability and inherent safe (including the use of Safety Data Sheets submitted in the Appendix or HAZOP analysis).	30	15
Total	400	257

DEPARTMENT OF CHEMICAL ENGINEERING
SENIOR DESIGN & PROFESSIONALISM II CHEG 4482
COURSE ASSESSMENT GRADING & OUTCOME
ASSESSMENT SHEET FOR SPRING 2019 SEMESTER

ABET Student Learning Outcome 3: An ability to communicate effectively with a range of audiences (Written Technical Communication)

Assignments that test competency in written communication will include problems that test the student's ability to logically arrange information, use correct grammar, use appropriate format, use discipline specific conventions including citations, and use acceptable graphical conventions.

Name: Gray 3

Title of Assignment: Final Report

Due Date: 5/6/2020

Competency Area	Points Assigned	Points Received
1. Ability to prepare an executive summary for the report	10	<u>10</u>
Students are able to prepare an executive summary of 5 to 8 pages	10	<u>10</u>
2. Ability to organize, plan and properly format a written technical report	80	<u>50</u>
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	<u>13</u>
(i) Students are able to organize report by categorizing ideas for the report into well and logically organized chapters, major sections, subsections and paragraphs blended within the larger units.	20	<u>12</u>
(ii) Students provide Title Page, and Table of Contents, list of Figures, and List of Tables properly formatted.	12	<u>5</u>
(iii) Students provide figure number and title for each figure in the report, reference each figure, and completely discuss each figure in the report in accord with standards in the project manual.	12	<u>5</u>
(iv) Students provide table number and title for each table in the report in accord with standards in the project manual, reference each table, and completely discuss each table in the report.	12	<u>5</u>
(v) Students properly cite references in the report and provide well formatted reference list at the end.	12	<u>5</u>
(vi) Students prepare the written report in accord with standard report formatting provided in the Senior Projects Report Manual.	12	<u>5</u>
3. Ability to compose original texts and properly apply the conventions of written language	50	<u>35</u>
(i) Properly apply capitalization, punctuation, and penmanship, to communicate clearly.	15	<u>10</u>
(ii) Spell proficiently	10	<u>8</u>
(iii) Apply standard grammar and usage to communicate clearly and effectively in writing including	20	<u>15</u>
• using complete sentences, varying the types such as compound and complex to match meanings and purposes • properly employing standard English usage in writing for audiences, including subject-verb agreement, pronoun referents, and parts of speech • properly using adjectives (comparative and superlative forms) and adverbs appropriately to make writing vivid or precise • properly using prepositional phrases to elaborate written ideas • properly using conjunctions to connect ideas meaningfully		
(vi) Use available technology to support aspects of creating, revising, editing, spell checking, and publishing the report.	5	<u>2</u>
4. Ability to provide appropriate discussion, conclusions and recommendations	60	<u>35</u>
(i) Summarize the goals, objectives, and indicate whether they were met.	10	<u>10</u>
(ii) Summarize the results.	30	<u>15</u>
(iii) Summarize constraints and codes and indicate whether they were met.	10	<u>5</u>
(iv) Provide logical conclusions and recommendations (including strengths and weaknesses).	10	<u>5</u>
Total	200	<u>130</u>

“Plant Design for the Production of Nitric Acid”

Senior Design & Professionalism II
Final Report

Submitted By:
James Fearon
Guty Francisco
Ryan Hooks
Maya Hunter
Robert Jackson II

To:
Dr. Lealon Martin
Senior Design Instructor

Roy G. Perry College of Engineering
Chemical Engineering Department
Prairie View A&M University

May 6th, 2020

1

LETTER OF TRANSMITTAL

May 6th, 2020

Department of Chemical Engineering
Prairie View A&M University
100 University Dr.
Prairie View, TX 77446

Dear Dr. Martin,

The Nitry Inc., Plant has completed the requirements that have been set for the Spring 2020 semester for the design of a chemical plant to produce nitric acid. After extensive research and going through a design phase, this group has completed the market and economic analysis, selection of optimum production routes and design of major process units. This report holds our design to produce 400,000 tons of nitric acid per year via the Ostwald process.

Sincerely,

No Signatures!

Maya Hunter, Team Lead

James Fearon

Guty Francisco

Ryan Hooks

Robert Jackson II

il

ACKNOWLEDGMENTS

Nitry, Inc., would like to thank all those who dedicated time and energy into helping us with the completion of this project and report. We would like to thank our technical advisor, Dr. Lealon Martin for all of his feedback and words of encouragement throughout this strenuous process. The guidance and wise words we have received from you have been continuous motivation and will follow us beyond our collegiate years.

We would also like to thank the El Dorado Nitrogen, LLC. for hosting us at their nitric acid plant in Baytown, TX. Dr. Michael Griffin and James Cartwright opened their doors to us and gave us an amazing walk thru and breakdown of their plant

Lastly, we would like to acknowledge all of our group members and fellow classmates for their hard work and determination this semester. This was not an easy semester, but we pushed thru and did not let COVID-19 be an excuse for poor work. Wishing the Chemical Engineering Class of 2020, the best and good luck in their future endeavors.

Pici

PROJECT EXECUTIVE SUMMARY

The purpose of the project was to design an aqueous ~~solution~~ of nitric acid. This project was completed by doing an in-depth literature review of the raw materials, product, reaction mechanisms and plant necessities to build a durable and safe plant. In order to know the amount of product that will be produced yearly by the plant, a market analysis was completed, along with a gross profit analysis and proper material and energy balances. The plant capacity will be determined based on the market analysis findings which put the plant at a production capacity of 400,000 tons per year in Baytown, Texas. This location was chosen due to the proximity to a water source, the easy access to raw materials and product transportation, the current presence of piping to import raw materials and the industrial surroundings. This product will be done via the Ostwald Process which is a 3-step process using a platinum rhodium catalyst. A HAZOP analysis will be done for the main process unit, the reactor, and safety parameters will be outlined for major portions of the plant

*This does not summarize
your activities and findings in
a complete way*

- 1. raw materials*
- 2. process*
- 3. cost estimate*
- 4. payback period*
- 5. safety / control*

All missing

4/18

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1. Project Scope

1.1 Problem Statement

what is the problem statement??

In response to growing demands for nitric acid (HNO_3), The Nitry Inc., Plant will be building a chemical plant in Texas. This plant will be estimated to produce and sell 400,000 tons of nitric acid per year with the main customer base being fertilizer companies in and near the area. Due to the need of being near a water base, the plant will be located in Baytown, Texas near the San Jacinto River on Galveston Bay. The price breakdown of the plant will be based on producing nitric acid at \$290 per ton [El Dorado Nitrogen, LLC].

The nitric acid which will be produced at this plant will be a 60 wt% aqueous solution of nitric acid (HNO_3). This percentage has been chosen due to nitric acid of less than 70 wt% being best used for our fertilizer plant customers and the best fit for the production process that has been chosen for the plant, the dual pressure process. The major units that were focused on this semester for the design basis were the evaporator, reactor, knockout drum, absorption column and stripping column.

1.2 Client Identification and Recognition of Need

Due to the location of the Nitry Inc. Plant being in Baytown, Texas, this plant is in what would be classified as the gulf coast region of the United States. This means that our nitric acid will be a top seller amongst fertilizer plants. Most of the nitric acid which is produced in the United States is used in the agriculture industry, especially the product that is produced in the mid-Western and Gulf Coast regions. However, if the plant were to be located in the far Western region of the United States, the market would lean more towards the explosives or polymer industry.

who are you selling to?

The plant to customer transportation process would utilize stainless steel tanks to transport the nitric acid to fertilizer plants in the states of Texas, Louisiana and Mississippi.

Table 1 shows the top customers of our plant.

Companies	Locations
Ouachita Fertilizer	Monroe, LA
Gavilan Fertilizer, LLC	Shreveport, LA
Helene Chemical Company	Guntown, MS
Mosaic Faustina	Hockley, TX
Nelson Plant Food	Bellville, TX
Southwest Fertilizer	Houston, TX
Lavaca Feed & Fertilizer	Hallettsville, TX
Helena Chemical Company	Garwood, TX
CF Industries	Donaldsonville, LA
Abilene AG Services & Supply Inc.	Abilene, TX

Table 1: Nitry Inc. Chemical Plant Customers and Locations

1.3 Project Goals and Objectives

The goal of this project is to design a chemical plant that is profitable and built to hold a production rate of 400,000 tons per year of nitric acid while running at 8,000 hours per year.

The production time will be based on both scheduled downtime for maintenance and cleaning of the plant and units but also considering unplanned downtime that may occur.

This plant will be able to produce enough product to competitively enter the nitric acid market. The production process that will be used will operate within the EPA guidelines for nitric acid production while operating for an environmentally friendly environment.

1.4 Contemporary Issues and Impact of Project on Society

1.4.1 Contemporary Issues Relevant to Project

In the production of nitric acid, there will be a stream of tail gas that may be released into the atmosphere from our absorption column. This tail gas will be a product of our nitrogen dioxide partaking in the third step of our chosen process with water and oxygen. The release of NO_x gas to the air can cause potential environmental and health concerns in humans. The NO_x gas can potentially enter the respiratory system of humans and cause lung irritation, laryngitis and bronchitis. The release of these gases can cause small harm to the environment but major issues in humans due to the small level nitrogen oxides that are released with the tail gas. Within the environment, the tail gas being released into the atmosphere can have a long-time effect on plants as well as agriculture near the plant site.

1.4.2 Potential Impact of Project on Society

With the location of the Nitry, Inc. Plant being within the gulf coastal region, the plant is in the perfect location to aid the fertilizer industry. A potential constraint for the plant could be access to raw materials and transport of both raw materials and the product however the industrial area is perfect for these needs. The coastal location works due to a large agriculture point being along the coast for farming and crops. The high demand of fertilizer to ensure proper cropping and food growth will bring the perfect opportunity for our plant to gain a wide customer base and aid those fertilizer companies in and around the states of Texas. This company would also be able to aid in the production of explosives if the opportunity is given for any explosive companies. This plant will open doors for various job positions for plant operators, a finance team, plant managers, control teams and other plant staff or personnel.

1.5 Project Constraints

No constraint listed.

For the project to be successful, the group must first look over the economic feasibility of the project. If a project is not economically feasible then it will not be of benefit of the company and will result in the company losing money. The goal is never to prepare a project and a process before analyzing the economic value of it in the long run. Secondly, the safety and health concerns that the plant will bring to the surroundings and the employees needs to be considered. The emissions that will be released from the plant must be within the EPA and OSHA guidelines and regulations to avoid any fines or plant closures that could happen. Also, if the economic analysis of the plant does not show a profit and proper rate of return for the plant then the profit that will be made will be a constraint. Operating a plant but not being able to make a profit to support both the plant and the maintenance of the plant can result in possible closure.

2 Project Planning and Task Identification

2.1 Task Identification

In order to properly plan the manufacturing process for this plant, there is a layout and breakdown of tasks and deadlines that need to be implemented. Project managements requires comparing timelines of a typical project planning time and giving each task of event an estimated amount of time in either days, weeks or in some case months. However, due to only having a few months for this project, this will be broken down into days. This second stretch of the project will take place from mid-January to the end of April. This will include the design of major units, a HAZOP and safety analysis and an economic cost analysis. Below is a list of the major tasks that were necessary for the completion of the project. The tasks were as follows:

1. Define Project and Tasks Timeline
2. Literature Review

3.4 Handling, Storage and Safety

3.5 Handling

When handling HNO_3 , the proper personal protective equipment (PPE) should be worn such as safety goggles, rubber gloves and steeled toe boots that are designed and suitable for possible exposure to acidic products. However, when in the presence of HNO_3 at the concentration and volume that will be produced in this plant, a positive pressure self-contained breathing apparatus will need to be in close proximity. *why?*

3.6 Storage

The product will be stored in stainless steel cylindrical, insulated storage tanks which will be above ground and have a capacity of $30,000 \frac{\text{lb}}{\text{hr}}$. The volume of these storage tanks will be 66.8 m^3 and store HNO_3 at a “slightly negative pressure which assures no NO_x vapors seep out of the tanks” as stated by El Dorado, LLC. General Manager, Michael Griffin. The storage tanks also have small vents at the top that allow for any thermal air that is put off by the chemical to be released. The diameter of the storage tanks will be 6.5 meters and the height will be 19.6 meters. *gallons!!*

Equipment Type	Nitric Acid storage tank
Shape	Cylindrical
Nature	Insulated
Material of Construction	Stainless Steel
Capacity	50000000kg/hr
Volume(m3)	23.8
Diameter(m)	4.6
Height(m)	13.9

Equipment Type	Ammonia Storage tank
Shape	Cylindrical
Nature	Insulated
Material of Construction	Stainless Steel
Capacity	13565kg/hr
Volume(m3)	66.8
Diameter(m)	6.5
Height (m)	19.6

Figure 3 : Storage Tank Specifications

3.7 Safety

The safety for the product nitric acid will focus on the exposure to large concentrations of the chemical, if released in large concentrations or volumes, water spray or resistant foam will be located around the plant and can be accessed by hose. The staff will be trained on how to properly spray nitric acid concentrations due to having to not spray the chemical with a straight stream. In case of exposure, the first step after containment will be to eliminate the source or locate the pipe causing the leak. Throughout the plant there will be safety signs near locations where nitric acid plants may hover or be in close proximity of. During rounds of the plant, possible tours or when supervisors or operators are directly in the plant with units, safety gear such as personal protective equipment (PPE) and a positive self-contained breathing apparatus will have to be carried. This breathing apparatus will be a small oxygen tank that personnel will utilize if possible exposure happens within the plant. Personnel and visitors will also wear "Safe Air" chemical detection badges on their shirts or near their badges that will change colors when a chemical such as ammonia, nitric acid or other NO_x gas emissions or exposure is above the normal limit. The color change of the badge will notify personnel that something is wrong within the plant or the area, telling them that it is time to take cover or shelter within the control room or a main building. Nitric acid reacts explosively with metallic powders, hydrogen sulfide and ethanol, however within this plant there will not be moments when the product will come in contact with any of these chemicals.

In regard to other safety measure that will be taken within the plant, the platinum rhodium catalyst that is used within the reactor will have to be replaced or changed periodically. The catalyst degrades over time and the full decomposition of the catalysts can lead to an issue within the chemical reaction or oxidation taking place within the reactor. Just as though the platinum rhodium catalyst will need to be changed regularly, the air filters within the air filter at the

beginning of the plant will need to be changed as well. Overtime, a filter will become dirty or blocked by filtration and this will allow the wrong chemical to enter into the wrong units of the plant. The changing of filters will typically happen during the turnaround period or planned downtime of the plant, but this will depend on production and the upkeep needed for the individual units.

Lastly, the tail gas for the process must be within the Environmental and Protection Agency's guidelines before being released into the atmosphere. The limits for the tail gas for the Baytown, Texas location of the plant will be less than 100 part per million (ppm).

3.7.1 EPA, OSHA Regulations and Environmental Impact

The EPA, The Environmental Protection Agency, focuses on the emission made from the tail gas of the absorption tower. The tail gas coming from the absorption tower consists of NO_x gases such as nitrogen monoxide, nitrogen dioxide, small traces of nitric acid that may pass through and ammonia. Majority of the tail gas outlet is NO_x gases because in the absorber is where the main reaction takes place to produce the nitric acid. Section 8.8.2 of the EPA's report on Nitric Acid production states that there are "two common techniques used to control absorption tower tail gas emission" [EPA], which are extended absorption and catalytic reduction. The process of extended absorption will utilize an additional absorption tower which will continue to breakdown the NO_x gases as a continuation of reaction on the sieve trays. The process of catalytic reduction would additional NO_x gases would be run through another catalytic bed which is where it would also be able to continue reacting.

Through our absorption tower which utilizes sieve trays, there is a sieve absorption process that takes place and oxidizes and absorbs the NO and NO_2 . The process of sieve tray absorption reduces the amount of NO_x emission in the plants [EPA Source]. It is stated that although this is

the best alternative, it is an expensive process and has high energy costs, but this is the best way and the way which has been chosen for this plant. Another alternative which is proposed by the EPA is to have a wet scrubber that scrubs that NO_x gas that exits the absorption tower and “scrubs” away any excess NO_x gas.

The exposure limit that has been highlighted in an Occupational Health Guideline for Nitric Acid, states that the permissible exposure limit for the exposure of nitric acid is 2 part per million (ppm) over the average work shift. *Cred?*

3.8 Uses

(Sources)

HNO_3 is used in the production of fertilizer, dyes, explosives and rocket fuels. The location of these production plants makes an impact on the industry that the product is used for. Most HNO_3 production plants that are located in the mid-West and the Gulf Coast region of the United States, supply fertilizer plants which nitric acid. However, if a HNO_3 plant is located in the west coast region of the United States, the plant will normally supply companies that produce explosives, polymers, dyes or rocket fuel. Due to this plant being located in Baytown, Texas near Galveston and the San Jacinto River, we will supply fertilizer plants.

3.9 Product Specifications

Table 3 shows chemical specifications for the 60 wt% nitric acid:

Nitric Acid Specifications	
Chemical Analysis	
Nitric Acid (HNO_3), wt%	56-64
Water	Balance
Oxides of Nitrogen, ppm	1500-5000
Physical Analysis	
Specific Gravity at 60°F	1.352
Freezing Point	< 0°F

Table 3 : Production Specification for HNO_3 (CF Industries)

60 wt% ??

4 Process Selection

4.1 Market Analysis and Trends

Due to our plant being located in the United States, the supply and demand of the nitric acid we will be producing will be based on the demand of fertilizer and the industry. Approximately 80% of the supply of nitric acid is used in the fertilizer industry today and is projected to continue growing. The Grand View Research Center reports that for 2019, the global market size for nitric acid was valued at \$24.0 billion. The source also reports that the market size is projected to grow at 3.3% from 2020 to 2027. [Grand View Research Center]. Figure 4 shows the United States Nitric Acid Market Demand for 2016-2027 with the fertilizer demand being the bottom portion of the bar graph. The chart has been utilized due to having to pay for access to the sample report with exact numbers. However, the 80% made be viewed on each bar.

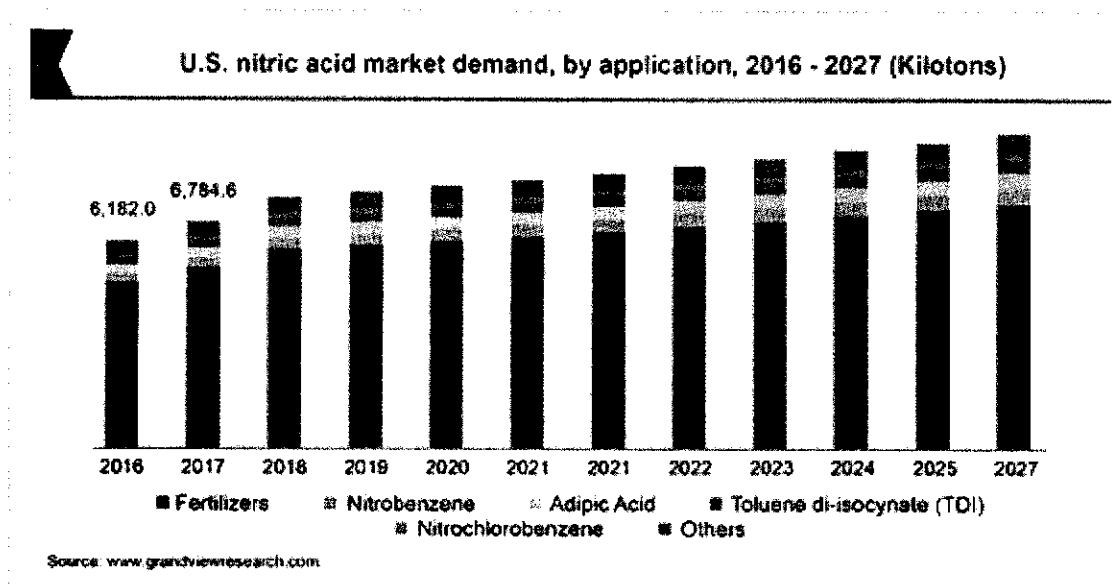


Figure 4: US Nitric Acid Market Demand for Grand View Research Center for 2016-2027

In contrast to the graph showing the United States Nitric Acid Market Demand until 2027, the figure below which was also obtained from the Grand View Research Center shows the Regional percentage of the market share for nitric acid production from 2019. The dark purple shows the North American portion of this market and is expected to continue growing significantly in the coming year. The demand for nitric acid for fertilizer production is set to continue to increase as the population of not only the United States but the world continues to increase.

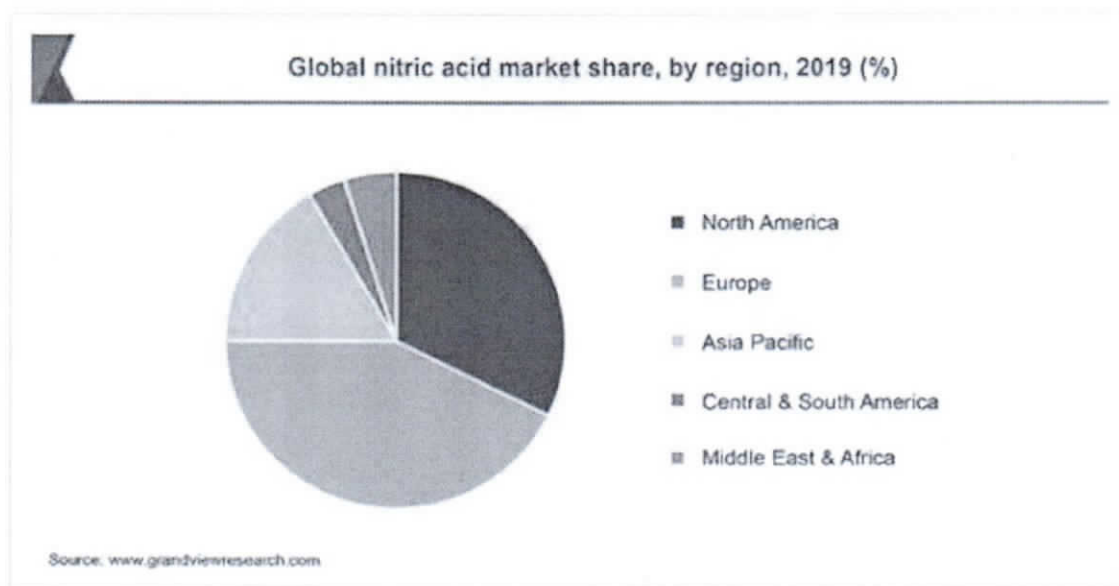


Figure 5 : Nitric Acid Market Share for 2019 from Grand View Research Center

4.2 Reaction Routes to Producing Product

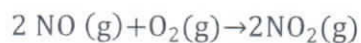
4.2.1 Ostwald Process

The Ostwald Process is a simple three step process that involves a two-step oxidation and one absorption step to form HNO_3 . In this process NH_3 is first mixed with air before being introduced into the reactor. This will cause the NH_3 to oxidize with the oxygen molecules found in the air to form nitric oxide. The nitric oxide is then oxidized to form nitrogen dioxide and dinitrogen tetroxide which later reacts with water within the absorption tower to produce HNO_3 .

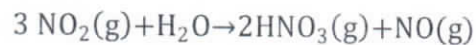
First Oxidation of NH_3



Second Oxidation of NO



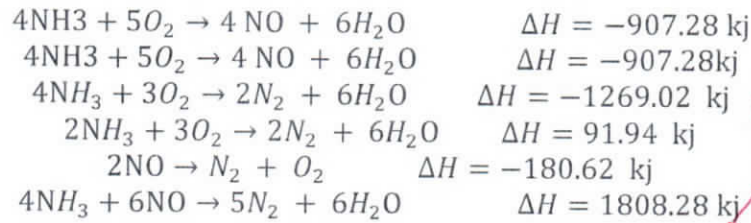
Absorption of NO_x to HNO_3



5 Process Description

5.1 Process Chemistry

5.1.1 Reaction Stoichiometry



What's the difference?

5.1.2 Kinetics and Rate Equations

In regard to the rate of reaction of ammonia in our reactor, the rate of reaction can be expressed in terms of a mass coefficient, surface area per unit volume of the catalyst and partial pressure of ammonia

$$-r_A = k_{gA}^s a_{wr} P_A$$

nomenclature

5.2 Process (Block) Flow Diagram and Process Description

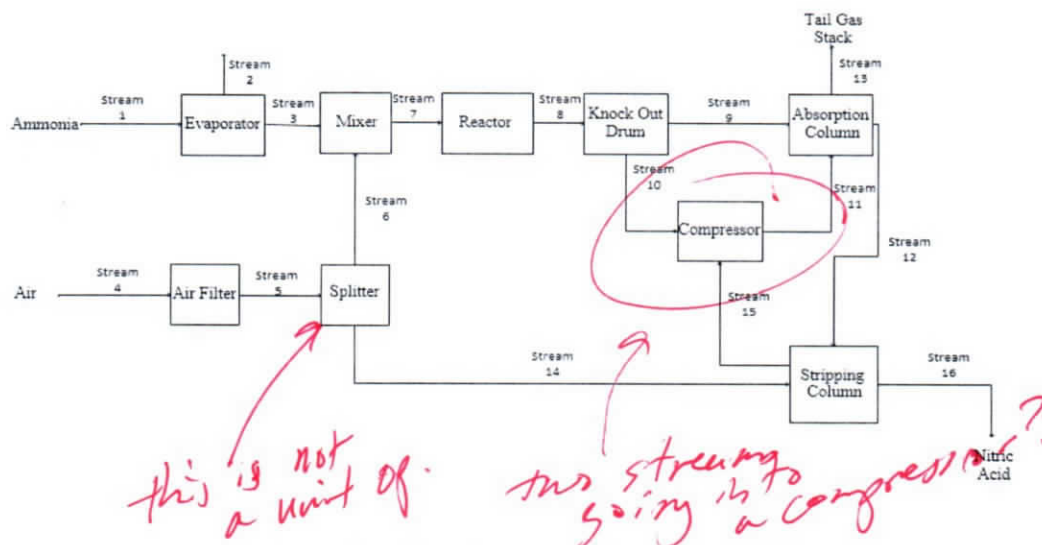


Figure 6 : Block Flow Diagram for Dual Pressure Process

The block flow diagram for the production of nitric acid through Ostwald's dual pressure process is presented in Figure 6.

The block flow diagram is a simple diagram that showcases the overall process that is used to make nitric acid in this chemical plant. A block diagram shows the major components or units and flow of product throughout the plant. The block flow diagram is a simplified version of the complex process flow diagram in Figure 7.

6h The block diagram shows liquid ammonia entering the process to an evaporator where the pure ammonia exits and any liquid or impurities (mainly water) will exit out of the top stream. The second stream that enters the process will be air that will pass through an air filter and into a splitter which will split the oxygen and nitrogen into two separate streams. Oxygen will be sent to the mixer to be mixed with ammonia and the nitrogen will move towards the end of the process and into the stripping column. Once the ammonia and oxygen are mixed in the mixing column, the mixture will be sent to the reactor which is where the first oxidation step will take place. This first step is where the ammonia and oxygen will be oxidized to produce nitrogen monoxide. After exiting the reactor, the nitrogen monoxide the nitric monoxide exits the knockout drum and is split between two streams, one stream enters the compressor and the second stream enters the absorption column. Nitrogen dioxide enters the absorption column which is where the nitrogen dioxide participates in the absorption process then exiting to the stripping column. The stripping column is where the nitric acid is stripped of all NO_x gases from the second stream that comes from the air splitter at the beginning of the process. The stripping column is the last unit in the process before the

A large, hand-drawn pink oval shape, resembling a stylized letter 'O' or a simple face. It is drawn with a thick pink line on a white background. There is a small, dark pink mark on the right side of the oval, possibly indicating a starting point or a specific feature.

~~H-101 Evaporator M-101 Mixer F-101 Filter C-101 Compressor V-101 Splitter R-101 Reactor E-101 Waste Heat Boiler E-102 Economizer E-103 Tail Gas Preheater D-101 Knockout Drum P-101 Pump A-101 Absorption Column A-102 Stripping Column C-103 Steam Turbine C-104 Expander C-102 Compressor E-106 Condenser E-110 Condenser~~

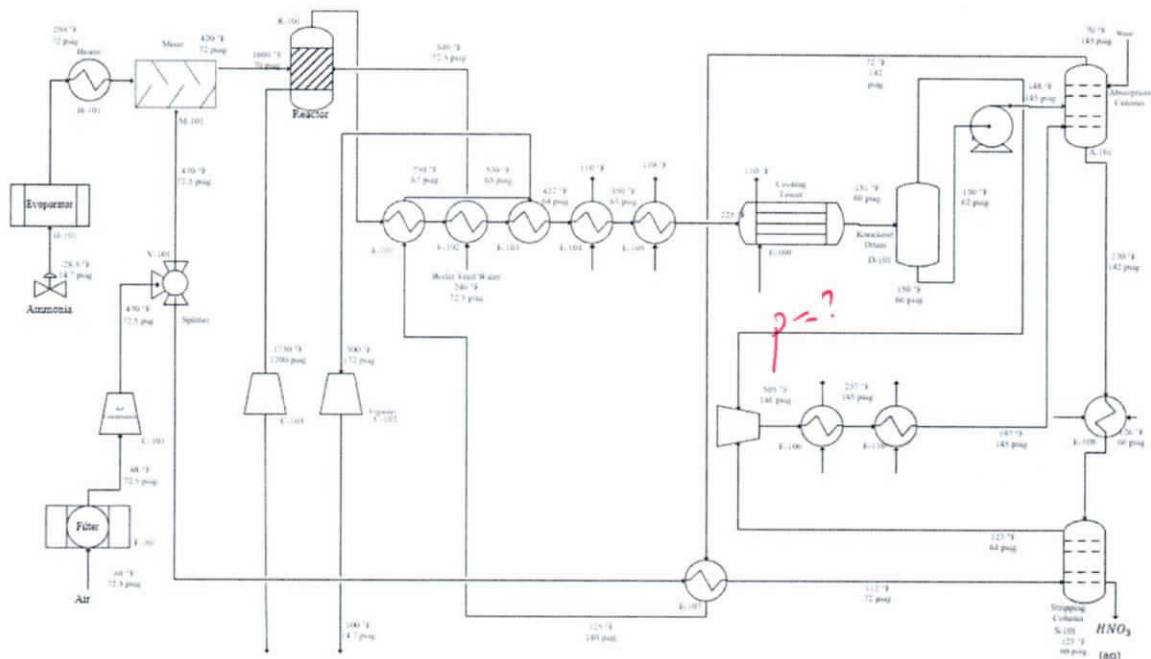


Figure 7 : Process Flow Diagram for Dual Pressure Process

5.2.1 Catalyst Preparation

The catalyst which will be used within the reactor will be a platinum rhodium catalyst that is constructed on 90% platinum and 10% rhodium. This is a fairly expensive catalyst that will be layered within the reactor walls or jacket. Overtime there are a variety of catalyst that have been used during this process however the platinum rhodium catalyst has proved to have the highest conversion although it is more expensive. The catalyst is in the form of a finely put together wire gauze. Gauze, by definition, is “a very fine wire mesh”. The construction of the catalyst typically uses 5-10% rhodium to increase the durability of the catalyst. The gauze, by definition is “a very fine wire mesh”, takes the 90% platinum and the 10% rhodium and they are placed along the inside of the reactor. As the ammonia and oxygen mixture passes over the gauze at the higher temperatures of the reactor, the first oxidation step of the process takes place.

[Thiemann] *yes?*

5.2.2 Raw Material Preparation

For this process, the main raw materials are NH_3 and O_2 . The NH_3 will be brought into the plant from an outside supplier and the O_2 will be obtained from the air. The raw material preparation process begins with the NH_3 passing thru an evaporator where any oils or liquids that may also be in the NH_3 are evaporated. This evaporator prepares the ammonia for the mixer where the chemical is mixed with the other raw material, O_2 . After the NH_3 and O_2 are mixed in the mixer, the mixture enters the NH_3 Oxidizing Plug Flow Reactor via stream 7. The first oxidation step of the Ostwald Process occurs in this reactor while the mixture is heated to up to 1600°F and pressurized at 70 psig.

5.2.3 Reaction and Reactor Operation

The ammonia and air mixture is sent from the mixing unit to the packed bed catalytic reactor which may operate at temperatures to about 1700°F and a pressure of 70 psig. After entering into the reactor, the mixture passes over a catalyst bed made of 90% platinum and 10% rhodium catalyst. As the reaction takes place with the passing mixture, the nitric oxide then exits the reactor at a temperature of 1600°F and passes through a series of heat exchangers. The heat exchangers include a super heater, a waste heat boiler and a tail gas preheater that takes the mixture from 750°F to 225°F. After leaving the heat exchangers and passing through a cooling tower to cool the liquid-vapor mixture down to 151°F. There is a knockout drum or flash drum that is used to split the liquid and vapor stream as the liquid goes to the absorption column and the vapor stream to the air compressor. Another reaction takes place in the absorption tower as the liquid passes through the tower and interacts with the sieve trays. The acid that leaves the bottom of the absorption tower contains some NO_x , but is mostly a colored nitric acid. The colored nitric acid is sent to the stripping column where it is stripped of any colored gases and comes out of the exit stream as a pure 60 wt% aqueous solution of nitric acid that is now prepared to be sent out to Nitry Inc.'s customers.

5.2.4 Product to Storage

The finished product of 60 wt% HNO_3 will exit the stripping column via stream 16 and go straight into the cylindrical storage tanks. These tanks will be made of stainless steel with a vented roof to allow the release of any thermal gases that will be produced while the product is settling in the storage tanks. This plant will have four storage tanks and each tank will have level indicators that will be connected directly to the control room and a shut off control for the plant. Level indicators will be in each tank therefore when tank one reaches its full capacity, the

indicator will notify the control room and the inlet stream for tank one will close and the inlet stream for tank two will open. This will go on from tank one to tank four and in the event that tank four reaches the full capacity, the control room will be notified, and the plant will automatically shut off. The ideal shut off procedure will allow the plant enough time to finish the production process of the HNO_3 until all raw materials that are moving through the system have been converted and utilized. After plant shut off, the shut off will carry out until the plant is able to push out some of the product and allow room for more HNO_3 . ok.

6 Material and Energy Balances

6.1 Material Balances

Material balances are an important aspect in the design of chemical plants. They are used to ensure that the amount of product and flows that are entering a plant are also being released or recycled at the end of the process. These calculations are used to not only help with the sizing of units but can also help with the economic analysis.

6.1.1 Basis Calculations

The design of this plant was done using the basis of the plant producing 400,000 tons/yr of the product, 60 wt% aqueous solution of nitric acid.

$$\text{Operational Hours} = 8,000 \text{ hrs/yr}$$

$$\text{Mass Flow Rate} = 400,000 \text{ tons/yr}$$

$$= 400,000 \text{ tons/yr} \times 2,000 \text{ lb/1 ton}$$

$$= 800,000,000 \text{ lb/yr} \times 1 \text{ yr/8000 hrs}$$

$$= 100,000 \text{ lb/hr}$$

For 60 wt% aqueous solution of Nitric Acid,

Compound	Stream 1 (lb/hr)	Stream 2 (lb/hr)	Stream 3 (lb/hr)	Stream 4 (lb/hr)	Stream 5 (lb/hr)	Stream 6 (lb/hr)
NH ₃	29762	0	29762	0	0	0
H ₂ O	0	1093	143	0	0	0
O ₂	0	0	0	109614	109614	94269
N ₂	0	0	0	412440	412440	354635
NO	0	0	0	0	0	0
HNO ₃	0	0	0	0	0	0
NO ₂	0	0	0	0	0	0
Total	29762	1093	29906	522054	522054	448904.15

Table 4: Material Balance for Stream 1 to 6

Compound	Stream 7 (lb/hr)	Stream 8 (lb/hr)	Stream 9 (lb/hr)	Stream 10 (lb/hr)	Stream 12 (lb/hr)	Stream 13 (lb/hr)
NH ₃	29762	615	615	0	595	1210
H ₂ O	143	46429	46429	0	61994	55801
O ₂	94270	25706	0	25706	0	41050
N ₂	354635	354635	0	635	635	1286212
NO	0	0	0	51425	18960	707385
HNO ₃	0	0	0	0	66139	0
NO ₂	0	0	0	0	1482	1482
Total	68401	427385	6721	77766	149804	2093140

Table 5: Material Balance for Stream 7 to 13

Compound	Stream 14 (lb/hr)	Stream 15 (lb/hr)	Stream 16 (lb/hr)	Stream 17 (lb/hr)
NH ₃	0	0	0	0
H ₂ O	0	0	44092	9392
O ₂	15344	8885	0	0
N ₂	576942	354635	0	0
NO	0	1279	0	0
HNO ₃	0	0	66139	0
NO ₂	0	16006	0	0
Total	592286	380804	110231	9392

Table 6: Material Balance for Stream 14 to Stream 17.

6.2 Energy Balances

Overall energy balance equation is :

$$U_1 + P_1 V_1 + U \frac{1}{2} g + Z_1 g + Q = U_2 + P_2 V_2 + U \frac{2}{2} g + Z_2 g + W$$

Q=Heat Transfer

W=Work Done

P=Pressure

$$\text{HNO}_3 = 100,000 \text{ lb/hr} \times 0.6 \text{ HNO}_3$$

$$= 60,000 \text{ lb/hr HNO}_3$$

$$\text{H}_2\text{O} = 100,000 \text{ lb/hr} - 60,000 \text{ lb/hr}$$

$$= 40,000 \text{ lb/hr}$$

6.1.2 Material Balance Around Each Unit

The material balance for the plant from stream 1 to stream 17 which can be compared with the Block Flow Diagram in Figure 6.

6.1.2.1 Assumptions and Use of Heuristics

These are assumptions which were made around the overall design process of this nitric acid production plant.

- Steady State
- Assumed downtime, 91% full production and 9% projected downtime, planned and unplanned
- Nitric Acid will be produced at 60 wt%
- 400,000 tons/yr of nitric acid production
- Projected flow rate of product: 100,000 lb/hr

6.1.3 Overall Material Balance with Process Streams

Total Input = Total Output

Stream 1 + Stream 4 = Stream 16 + Stream 17

V=Volume

U=Velocity

Z=Height

G=Acceleration due to gravity

We know that:

$$H=U+PV$$

$$H_2-H_1 = Q-W \rightarrow W=0$$

$$H_2-H_1=Q-W$$

Assumptions:

- Steady state
- No heat loss
- Effect of pressure is ignored
- Potential and kinetic energy are negligible

Energy Balance on Reactor

Composition	Mass Flowrate In (lb/hr)	H_{in} (btu/lb)	Mass Flowrate Out (lb/hr)	H_{out} (btu/lb)
NH ₃	929762	0	595	816
H ₂ O	143	563	46429	563
O ₂	94269	263	25705	263
N ₂	354238	298	354635	298
NO ₂	0	0	51412	263

Table 7 : Energy Balance on NH₃Oxidizing Reactor

Energy Balance on Absorption Tower

Composition	Mass Flowrate In	H_{in}	Mass Flowrate Out	H_{out}
-------------	------------------	----------	-------------------	-----------

	(lb/hr)	(B tu/lb)	(lb/hr)	(B tu/lb)
NH_3	595	0	615	248
H_2O	61994	50620	52536	42905
N_2	354635	72150	354635	42905
NO	3219	619	18960	3696
NO_2	73921	18907	1478	379
HNO_3	0	0	66139	22211

Table 8 : Energy Balance on Absorber

7 HYSYS/ASPEN Simulation Process

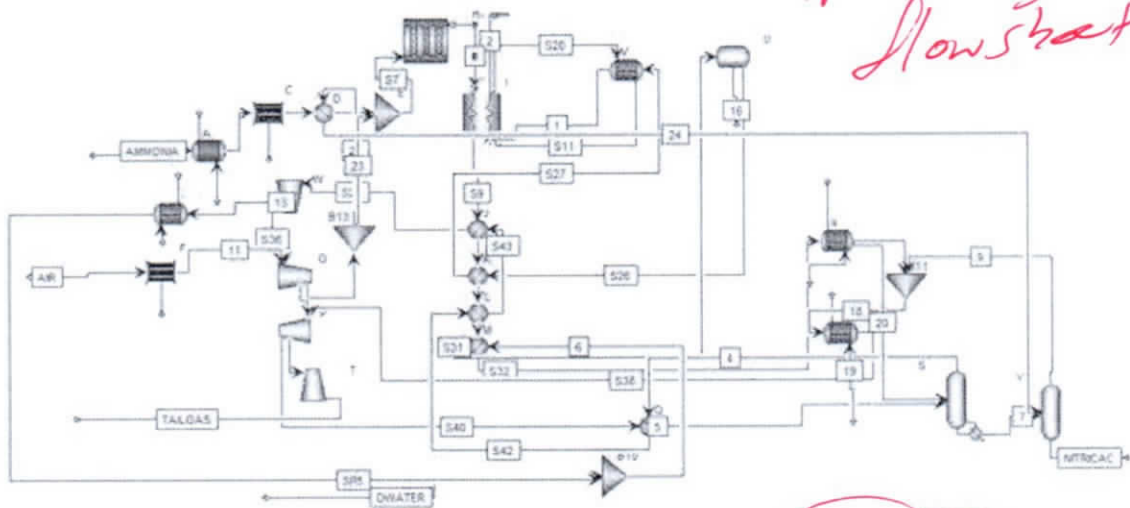


Figure 8 : Aspen Simulation for Nitry Inc. Chemical Plant – Unconverged

8 Chemical Engineering Design

This section will include descriptions and calculations for the detailed design of five major units for the plant: Evaporator, NH_3 Oxidizing Plug Flow Reactor, Knockout Drum, Stripping Column and Absorption Tower.

8.1 Design of Major Equipment

8.1.1 Evaporator Design

The evaporator for this process is implemented at the beginning and the inlet stream for the raw material, ammonia (NH_3). An evaporator is, by definition, used to remove a solvent in its vapor form from a solution. The evaporator is used to remove compounds such as organic compounds, acids, bases, oils, debris, etc. This plant will utilize a Calandria Evaporator (Robert Evaporator) which is a tubular shaped evaporator that holds a long diameter and typically has a down comer area that utilizes about 40 to 70% of the total cross-sectional area of the evaporator tube. The most common solvent in most of the evaporation systems is water. Evaporator consists of a heat exchanger for boiling the solution with special provisions for separation of liquid and vapor phases. Below will show the steps and equations used to calculate the evaporator design, as well as being followed by the specification sheet in Figure 8 for the evaporator.

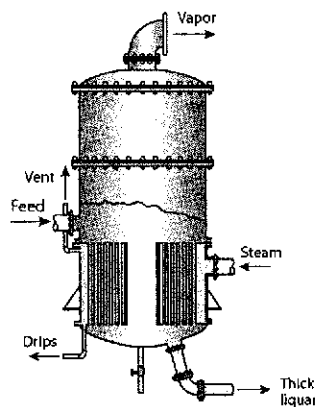


Figure 9: Calandria Type Evaporator.

Above we have a Short-tube vertical evaporator which consist of a short tube bundle (about 4 to 10 ft in length) enclosed in a cylindrical shell. This is called calandria. A evaporator of this type is shown in Figure 9 . The feed is introduced above the upper tube sheet and steam is

introduced to the shell or steam chest of the calandria. The solution is heated and partly vaporized in the tubes.

The central tube in a calandria is of longer diameter. Its down comer area is taken as 40 to 70% of the total cross-sectional area of tubes. The circulation rate through the down comer/down take is many times the feed rate. The flow area of the down take is normally approximately equal to the total tubular flow area.

Steps in calculation of Evaporator Design:

Number of tubes (N):

$$\text{Mean Dia of Tube } (D_m) = \text{Tube OD} - \text{Tube Thickness}$$

- *Effective Length of the tube (L) = Tube length – 2(Tube plate thickness)-2(Tube expansion allowance)*
- *Number of tubes = Heating $\frac{\text{Surface}}{\pi \cdot D_m \cdot L}$*
- The thickness of the tube generally take for ammonia heaters and for evaporator 18G and for pans take 16G. (18g = 1.22mm , 16g = 1.625mm, 14g = 1.8mm).

*Need heating Surface Area!
But you need to know the heat duty!!*

Tube plate & Downtaker diameter :

- **Tube Pitch (P)** = OD of the tube + Legment of the tube + tube clearance +hole clearance
- Proportional factor(β) = Generally β value taken for multiple pass (i.e Ammonia heaters) 0.6 to 0.8 and for single pass(i.e evaporators) 8 to 1.0.
- Its value lies in the range 10 to 20% on area occupied for tubes.

$$\text{Tube plate area required for tubes only } (AT) = (0.866 \times P^2 \times N / \beta) \times \% \text{extra}$$

$$\text{Tube plate diameter required for tubes only} = \text{SQRT } (AT \times 4/\pi)$$

- Diameter of the central downtaker in multiple down takes design:

Fonts!!

$$\sqrt{[(\text{Area of singledowntake} - \text{total area of peripheraldowntakes}) \times \frac{4}{\pi}]}$$

- The final required tube plate diameter.
- Final Diameter of the tube plate:

$$\sqrt{[(\text{area of tubeplate for tubes} + \text{downtake area}) \times \frac{4}{\pi}]}$$

- **Diameter required for vapor inlet & Diameter of the Calendria in radial steam/ vapor entry**
- Area required for the each vapor entry (ft²) = Volume of the vapor in each (ft³/sec) / Velocity of vapor.
- Diameter of the each steam entry

$$\sqrt{[\text{area required for } \frac{\text{entry}}{\text{no. of vapour entries}} \times (4/\pi)]}$$

- Calendria diameter at the entry of the steam/vapor jacket
- Area for the inlet vapor (If more than one connection to calandria than take each inlet vapor area):

$$\pi/4 \times (\text{dia of vapour inlet})^2$$

- Height of the steam entry = Take Effective Length of the tube.
- Width of the steam entry = Area for the inlet vapor / Height of the steam entry.

Vapor outlet pipe diameter:

- Vapor volume outlet vapor in ft³/sec

$$\text{heating surface} \times \text{evap. rate} \times \text{sp. volume of outlet } \frac{\text{vapour}}{3600}$$

- Vapor outlet pipe diameter in meters:

$$\sqrt{\left[\frac{\text{vapour volume}}{0.785 \times \text{velocity of vapour}}\right]}$$

Diameter of the condensate line :

- Number of condensate withdrawal points = Consider minimum two numbers of withdrawal points and it can be increase according to the diameter of the body.
- Volume of the condensate in ft³/sec:

$$\frac{\text{heating surface} \times \text{evap. rate}}{\text{density of water} \times 3600}$$

- Diameter of the each condensate line : $\sqrt{\left(\frac{\text{vol. of condensate}}{0.785 \times \text{velocity of condensate}}\right)}$

Noxious gases connections :

Generally, 10 m² heating surface area required 1 cm² area for removal of non-condensable gases.

Cross section area of non-condensable gases in cm² = Heating surface in m² /10

- Diameter of the each non condensable gases line : $\sqrt{\left(\frac{\text{total area of NCG}}{0.785 \times \text{No. of points}}\right)}$

Vapor space height :

It will take for lost effect 2.5 times on calandria tube height.

Velocity in vapor space of body (Cross checking of the system):

- It is helpful to avoid the entertainment of the system. The velocity of vapor leaving the liquid surface would then be approximately 10 in/sec.
- In present designs, the vapor velocities in the vessels to be maintained below 3.6ft/sec and for lost body it can go up to 4.6 ft/sec .

Calendria shell thickness :

- P = Maximum allowable pressure in lb/ft²
- Di = ID of the Calendria in mm
- F = Allowable stress in lb/in²
- J = Welding Joint efficiency in inch
- C= corrosion allowance in inch
- Calendria shell thickness in inch:

$$\left(\frac{PD_i}{2FJ - p} \right) + C$$

Vapor shell Thickness :

- P = Maximum allowable pressure in lb/cm²
- F = Allowable stress in lb/cm²
- J = Welding Joint efficiency in in
- C= corrosion allowance in in
- Di = ID of the Calendria in in
- Vapor shell thickness

$$\left(\frac{PD_i}{2FJ - p} \right) + C$$

Tube plate thickness:

- C= corrosion allowance in in
- F = Allowable stress in lb/cm²

- P = Maximum allowable pressure in lb/cm²
- E_s = Modulus factor for MS sheet in lb/cm²
- E_t = Modulus factor for SS sheet in lb/cm²
- G = ID of the shell in inch
- t_s = Thickness of the shell in inch
- t_t = Thickness of tube in inch
- d_o = OD of the tube in inch
- d_o = OD of the calendria sheet in inch
- N_t = Number of tubes

$$K = (E_s \times t_s \times (D_o - t_s)) / N_t \times E_t \times t_t \times (D_o - t_t)$$

$$f = \sqrt{\left(\frac{K}{2 + 3K}\right)}$$

- Tube plate thickness in inch = $f \times G \sqrt{((0.25 \times P)/F)}$

Vapor doom diameter :

- Generally for Robert type bodies vapor doom diameter taken 2 to 2.5 times for vapor outlet pipe area.
- Cross sectional area of the vapor doom = $2 \times \text{vapour outlet pipe}$
- Vapor doom dia = $\sqrt{\text{area of vapour doom} \times \frac{4}{\pi}}$

Top cone Height :

- Top cone angle (ϕ) = 35 deg.
- Top cone height =

$$\tan\phi(ID \text{ of body} - ID \text{ of doom})$$

Center Umbrella area :

- Area of Umbrella = consider 60 to 65% on body diameter
- Gap between Umbrella to top cone at the place of vapor inlet (H_i) =

$$\frac{\text{area of vapour doom}}{\pi \times \text{dia of umbrella}}$$

- Gap between Umbrella to top cone at the place of vapor outlet (H_o) =

$$\frac{\text{area of vapour doom}}{\pi \times ID \text{ of vapour doom}}$$

THERMAL CALCULATIONS:

Description	Sign	UOM	Values
Collection of the required data.			
Ammonia Flow rate	Q_j	Lb/hr	30998.71
Hot water (heating media) Inlet temperature	t_i	°F	194
Hot water (heating media) Inlet temperature	t_o	°F	143.6
Velocity of ammonia in tubes	V_j	ft/sec	5.85
Collect the physical properties for Ammonia			
Specific heat of the ammonia	C_j	Btu/lb °F	0.52
Density of ammonia	ρ_j	lb/ft ³	38.55

Ammonia inlet temperature	Ti	°F	113
Ammonia outlet temperature	To	°F	149
Viscosity of ammonia	μ_j	Poise	270×10^{-5}
Thermal conductivity of ammonia	Kj	[Btu(IT)/(h ft °F)]	0.5714
Collect the physical properties for condensate water at 85 °C			
Density of water	ρ_w	kg/M ³	62.4
Specific heat of hot water	Cw	Btu/lb-°F	1.003
Thermal conductivity of water	Kw	[Btu(IT)/(h ft °F)]	0.3343
Viscosity of water	μ_w	Poise	0.0091
Assume OHTC and heater design parameters			
Heat transfer coefficient	Uo ass	Btu/ft ² /hr/°F	225
Tube OD	OD	In	1.77
Tube Thk	Tk	In	0.047
Tube plate thk	Tpk	In	0.98
Tube ID	D	In	1.68
Legment	Lg	In	0.47
Tube length	L	In	157.4803
Fundamental formula for heat balance			
Heat received by Ammonia = Heat rejected by condensate water ✓			
$Q_j \times C_j \times \Delta T (t_o - t_i) = W \times C_w \times \Delta T (T_i - T_o)$			
Quantity of Ammonia	Qj	lb/hr	507063.203

Weight of hot water	W	lb/hr	419480.16
Calculating number of tubes per pass			
Volume of ammonia	Qj	ft ³ /sec	2.152
Cross sectional area of one tube	At	ft ²	0.0152
Number of tubes per pass			23.77
i.e		tubes/pass	24
Heat transfer coefficient	U _{oas}	Kcal/ft ² °C/hr	102.049345
$\Delta T = T_o - T_i$		°C	20.00
$\Delta T_i = t_i - T_o$		°C	19.00
$\Delta T_e = t_o - T_i$		°C	17.00
$\ln(\Delta T_i / \Delta T_e)$		°C	0.11
$\Delta T_m = \Delta T_i - \Delta T_e / (\ln(\Delta T_i / \Delta T_e))$			17.98
Heating Surface	S	ft ²	2281.19553
Mean Dia of the tube	D _m	In	1.72
Effective tube length	E _f	In	155.11
Number of tubes	N _T		391.23
i.e		no.s	392
Tube plate Area = $0.866 \times P^2 \times N_T / \beta$			
Pitch = OD+ legment +tube tolerance +hole tolerance	P		57.6
Proportional factor	β		0.7

Tube plate area	A		1.61
Tube plate dia / heater shell dia	Ds	Ft	4.691
Tube side Ammonia coefficient (Hf)			
Sieder and Tate formula (1974)			
$H_f = 0.023(K_j/D) \times (DG/\mu_j)^{0.8} \times (C_j \mu_j/K_j)^{0.333} \times (\mu_j/\mu_w)^{0.14}$			
Tube side mass flux	$G = Q_j / A_j$		
Area of the all tubes in pass	Aj	ft ²	0.366
	Qj	lb/hr	507063.2
G		lb/hr/ft ²	1 377 818.27
Re = Reynolds number = $G D / \mu_j$			
Re			95334.12
Pr = Prandtl number = $C_j \mu_j / K_j$			
Pr			5.395
Tube side Ammonia coefficient	Hf	Kcal/hr/ft ² /°C	480.35
Consider baffles in shell side for hot water passing	B	no.s	3
$Re = d_e V_w \rho_w / \mu_w$			
Shell side mass velocity $V_o = m / A_s$			
Cross flow $A_s = (P - D) D_s L_b / P$			
Pitch	P	Ft	0.189
OD of tube	D	Ft	0.147
Shell inside dia	Ds	Ft	4.697

Baffle spacing	Lb	Ft	1.56
Cross flow	As	ft ²	1.609
Fluid flow rate on the shell	m	lb/hr	116.52
	Qw	M ³ /sec	1.942
Velocity of water in shell side	Vw	m/sec	1.197
		m/hr	4314.96
de (here consider triangular pitch)		M	0.117
$Re = de \cdot Vw \cdot \rho_w / \mu_w$			37923.04
$Pr = Cw \times \mu_w / Kw$			2.072
Hs		Kcal/hr/ft ² /°C	227.3
Material coefficient			
Thickness of the tube	Tk	M	0.0012
Thermal conductivity SS tubes	Kt	Btu/Ft/hr/°F	26
		Kcal/hr/m/°C	13.96
Material heat transfer coefficient	Kt / Tk	Kcal/hr/ft ² /°C	1080.7
Fouling factor for light hydrocarbons		Kcal/hr/ft ² /°C	399.39
$1/U_{sc}$			0.000233
Uo cal = Overall heat transfer coefficient			
1/Hf		Tube side	0.000193
1/Hs		Shell side	0.000409

Kt/Tk		Wall	8.6E-05
1/Usc		Fouling factor	0.000233
1/Uo cal			0.0009206
	Uo cal	Kcal/hr/ft ² /°C	100.911

Table 9: Evaporator Design Calculations and Equations

These calculations should be in the Appendix. This is regarding.

In this calculation the percentage of variations between: U_{oas} & U_{ocal} is less-than 3%. So U_{ocal} value is correct.

Calandria Evaporator Specification Sheet

Identification	Item:	Vertical Vessel		
	Item No.		Date:	04/25/2020
	No. Required	1		
	C(\$):			

Function: Evaporator is to removal of solvent (Water) as vapor from a solution (Ammonia) shurry

or suspension of solid in a liquid.

Operation: Adiabatic

*?
what is the solid?*

	Feed Stream (lb/hr)	Vapor Stream (lb/hr)	Liquid Stream (lb/hr)
Material Handled: Ammonia, Water			
Oxygen:	0	0	0
Nitrogen:	0	0	0
Ammonia:	29762.3	0	29762.3
Water:	1256.4	1113.1	143.3
Nitrogen Oxide:	0		
Nitric Acid:	0		
Nitrogen Dioxide	0		
Total:	31018.7	1113.1	29905.6
Temperature (°F):		143.6	
Pressure (psig):			
Design Data:			
Area (ft. ²):			
Height (ft.)			
Diameter (ft):			
Volume (ft ³):	2.152		
Material	Stainless Steel		
Construction:			
Vapor	5.85	1.197	1.8
Velocity (ft/s):			

only 2 ft³ to handle 29 tons/hr of NH₃?

Table 11: Evaporator Specification Sheet

8.1.2 Reactor Design

Typically, in reactor designed to produce nitric oxides for nitric acid production there are usually standard operating temperatures ranging from 1490°F to 1724°F. In order to generate sufficient amounts of nitric oxide for viable commercial use we designed the catalytic packed bed to operate at a temperature of about 1700°F and a pressure of 70 psig. The purpose of designing the reactor with a high operating temperature is to increase the rate of reaction of ammonia to nitric oxide. After entering into the reactor, the mixture passes over a catalyst bed made of 90% platinum and 10% rhodium. As the mixture of gases passes through the platinum rhodium catalyst, the nitric oxide then exits the reactor at a temperature of 1600°F where the remaining materials exit through the reactor passes through a series of heat exchangers. In order to determine the number of gauzes needed to help stimulate reaction of ammonia to nitrogen oxide the equation presented below is:

$$n_s = -\ln(1 - X_A) \varepsilon_w^{.352} d_w^{.648} G^{.648} \mu_f^{.019} / (5.81761 * 10^{-5}) f_w T_i^{.333} (28.85 + 11.82 y_{AO})^{.667}$$

poorly presented equation

where as: ε_w represent the porosity factor, d_w represents diameter of the gauze wire, G represents the mass velocity, μ_f is equivalent to viscosity, f_w represents the wire area per cross sectional area of one gauze, while T_i represents mean film temperature. In order to determine the height of the catalyst bed the equation below is utilized which also encompasses using the number of gauze formula.

$$h_c = 2d_w n_s$$

Also, in determining the volume of the catalyst bed,

$$V_c = \left(\frac{1}{4}\right) \pi D^2 (h_c)$$

bad sig figs!

Reactor Specification Sheet		
Reactor Temperature	1700	degree Fahrenheit or Celsius
Reactor Pressure	70	PSI or kPa <i>which one?</i>
Conversion of Ammonia	0.787788829	
NO Yield	0.414610701	
Rate	0.6467	(lbmol of NH ₃ /ft ³ of catalyst*hr)
Catalyst Name	Platinum- Rhodium	
Bulk Density	1248.6	lb/ft ³
Volume of Reactor	2707.17	ft ³
Weight of Catalyst	3380174.02	lb
Catalyst Diameter	0.0003	in
Catalyst Length	1	in
Catalyst specific surface area		ft ² /lb
Feed Mass Flowrate	478810.4	lb/hr
Feed Molar Flowrate	17370.2	lbmol/hr
Feed Volumetric Flowrate	192687.7	ft ³ /hr
Product Mass Flowrate	478810.4	lb/hr
Product Molar Flowrate	17798.6	lbmol/hr
Product Volumetric Flowrate	77635.0	ft ³ /hr
Residence Time	0.014049528	hr
Space Velocity	71.17676894	hr ⁻¹
Diameter	10.44916121	ft
Height	31.34748363	ft
Cross sectional Area	85.75367488	ft ²
Material of Construction	Stainless Steel	
MAWT	2100	Degrees Fahrenheit or Celsius
MAWP	150	PSI or kPa

Table 11: NH₃ Oxidizing Plug Flow Reactor

*what does this mean?
which one?*

8.1.3 Knockout Drum Design

A knockout drum, sometimes called a flash drum, is a vapor liquid separator that is used to split the feed into two phases. The vapor stream of this knockout drum will be sent to an air compressor and the liquid stream will be sent to the absorption column. The vapor stream

contains oxygen, nitrogen and nitrogen monoxide and the liquid stream contains water and a small composition of ammonia. [Wankant]. *yes*

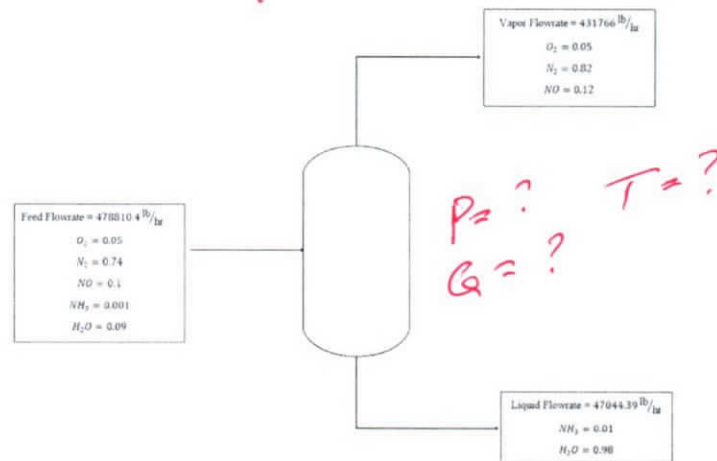


Figure 10 : Knockout Drum Schematic Drawing

Knockout Drum Calculations

1. Liquid Density

The average molecular weight of the liquid is:

$$MW_L = x_{H_2O} MW_{H_2O} + x_{NH_3} MW_{NH_3} = \left(\frac{.09}{.091}\right) \left(18 \frac{g}{mol}\right) + \left(\frac{.001}{.091}\right) \left(17 \frac{g}{mol}\right) = 18 \frac{g}{mol}$$

MW_L = Molecular weight of the liquid

x_{H_2O} = Mole fraction of Water

MW_{H_2O} = Molecular weight of Water

x_{HNO_3} = Mole fraction of Ammonia

MW_{HNO_3} = Molecular weight of Ammonia

The volume of the liquid is:

$$V_L = x_{H_2O} \frac{MW_{H_2O}}{\rho_{H_2O}} + x_{NH_3} \frac{MW_{NH_3}}{\rho_{NH_3}} = (.09) \left(\frac{18 \frac{g}{mol}}{0.9 \frac{g}{cm^3}} \right) + (0.001) \left(\frac{17 \frac{g}{mol}}{0.00073 \frac{g}{cm^3}} \right) = 25$$

V_L = Volumetric flow rate of the liquid

ρ_{H_2O} = Density of Water

ρ_{NH_3} = Density of Ammonia

Therefore the density of the liquid is:

$$\rho_L = \frac{MW_L}{V_L} = \frac{18 \frac{\text{g}}{\text{mol}}}{25} = 0.72$$

ρ_L = Density of the liquid

2. Vapor Density

(? deal gas ?? really ??)

The average molecular weight of the vapor is:

$$MW_V = x_{\text{O}_2} MW_{\text{O}_2} + x_{\text{N}_2} MW_{\text{N}_2} + x_{\text{NO}} MW_{\text{NO}}$$

$$MW_V = (.06)(32 \frac{\text{g}}{\text{mol}}) + (.82)(28 \frac{\text{g}}{\text{mol}}) + (.11)(30 \frac{\text{g}}{\text{mol}})$$

Therefore, the density of the vapor is:

$$\rho_V = \frac{P_{\text{atm}} MW_V}{RT} = \frac{(10 \text{ atm})(28.18 \frac{\text{g}}{\text{mol}})}{(82.0575)(333.3 \text{ K})}$$

P_{atm} = Pressure of the vessel in atmospheric pressure

R = Gas Constant

T = Temperature in Kelvin

3. Permissibility

$$u_{\text{perm}} = K_{\text{drum}} \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} = (0.38) \sqrt{\frac{0.72 \frac{\text{g}}{\text{mol}} - 0.01 \frac{\text{g}}{\text{mol}}}{0.01 \frac{\text{g}}{\text{mol}}}} = 1.8 \frac{\text{ft}}{\text{s}}$$

u_{perm} = permissible vapor velocity

4. Cross-Sectional Area

$$A_C = \frac{(V)(MW_V)}{(u_{\text{perm}})(3600)\rho_V} = \frac{(431766)(28.18 \frac{\text{g}}{\text{mol}})(454 \frac{\text{g}}{\text{lb}})}{(3.2 \frac{\text{ft}}{\text{s}})(3600)(0.01)(28316.85 \frac{\text{mL}}{\text{ft}^3})} = 423 \text{ ft}^2$$

A_C = Cross sectional area of the vessel

5. Diameter

$$D = \sqrt{\frac{4A_C}{\pi}} = \frac{4(423)}{\pi} = 23.2 \text{ ft.}$$

6. Total Height

$$H_{\text{total}} = 3D = (3)(23.2 \text{ ft}) = 69 \text{ ft.}$$

H_{total} = Total height of the vessel

7. Maximum Volume

$$V_{\text{max}} = \frac{\pi D^2 H}{4} = \frac{\pi (23.2 \text{ ft.})(69 \text{ ft.})}{4} = 29168.5 \text{ ft}^3$$

V_{max} = Maximum volume of the vessel

8. Knockout Drum Specification Sheet

Knock Out Drum				
Identification	Item:	Vertical Vessel		
	Item No.		Date:	04/25/2020
	No. Required	1		
	C(\$):			
Function: To separate vapor and liquid mixture of Nitrogen, Ammonia, Water, Nitrogen Monoxide and Nitrogen Dioxide				
Operation: Adiabatic				
	Feed Stream (lb/hr)	Vapor Stream (lb/hr)	Liquid Stream (lb/hr)	
Material Handled: Nitrogen, Ammonia, Water, Nitrogen Monoxide, Nitrogen Dioxide				
Oxygen:	25,705.8	25,705.9	0	
Nitrogen:	354,635.2	354,635.2	0	
Ammonia:	615.09	0	615.1	
Water:	46,429.3	0	46,429.3	
Nitrogen Oxide:	51,424	51,425		
Nitric Acid:	0	0	0	
Nitrogen Dioxide	0	0	0	
Total:	478,810.4	431,766	47,044.4	
Temperature (°F):				
Pressure (psig):				
Design Data:				
Area (ft. ²):				423
Height (ft.):				69
Diameter (ft.):				23.2
Volume (ft. ³):				29168.5
Material	Stainless Steel			
Construction:				
Vapor				1.8
Velocity (ft/s):				

70 ft
high
knockout
drum?
why not
just use
distillation?

Table 12: Specification Sheet for Knockout Drum

8.1.4 Stripping Column

A stripping column is a separation column which allows a solute (nitrogen dioxide and other nitrogen and oxygen containing compounds known as NO_x in nitric acid production) to be stripped from a liquid (aqueous nitric acid from absorption column) in other words, used to purify a liquid. In this section, there will be a detailed design of a stripping column. A stripping column is sometimes a cylindrical column that is lined with trays from the top to the bottom and is mounted vertically in the plant.

In this plant, the stripping column, sometimes referred to as a bleaching column, is the final unit where 60 wt aqueous nitric acid is at the end of the process, after the splitter and the absorption column.. After air is separated in the separator, the oxygen will be mixed with the ammonia in the mixer and the air will be sent to the stripping column. For this process, 85% of the air will pass to the reactor and the remaining 15% will be sent to the stripping column to aid in the stripping process.

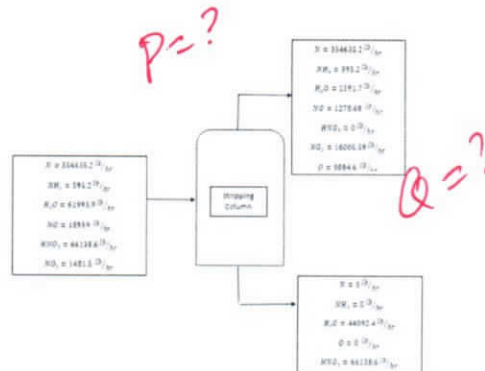


Figure --. Stripping Tower Schematic

8.1.5 Absorption Tower Design

The Absorption tower is where the nitric acid is made. We have water, ammonia, oxygen, nitrogen, nitric oxide, and nitrogen dioxide coming in. The stream going to the stripping column includes nitric acid and water as well as other compounds such as nitrogen, nitrogen dioxide, nitric oxide, and a little ammonia. In order for this reaction to take place, our absorption column will require a total of 36 trays. The tray spacing required will vary between each individual tray. Sieve trays will be used to have the best movement inside the column and the material of construction will be stainless steel. The walls will be .875 feet thick. Because our tower will have a height of 100 feet, we will have to meet local wind loads to ensure that the tower is steady. At the bottom of the tower we need a wind load speed of $25.9 \frac{\text{lb}}{\text{sq}\cdot\text{ft}}$ and the strength increases as the tower gets taller.

how
do
you
get
this?

The calculations required to do the McCabe-Thiele method are to know the vapor and liquid flow rates at each part of the tower as well as the component material balances for each species.

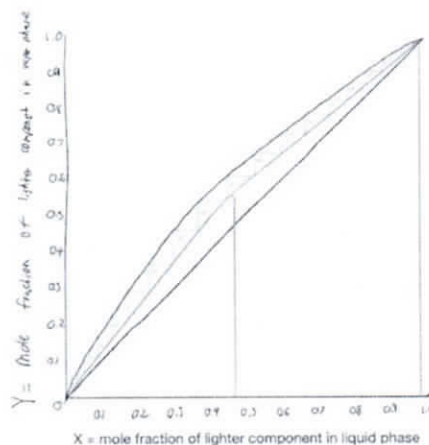


Figure 10: McCabe Thiele for Absorption Tower

Absorption Tower				
Identification	Item:	Vertical Vessel	Date:	04/25/2020
	Item No.			
	No. Required	1		
Function: To create Nitric Acid vapor and liquid mixture of Nitrogen, Ammonia, Water, Nitrogen Monoxide and Nitrogen Dioxide				
Operation: Adiabatic				
	Feed Stream 1 (lb/hr)	Feed Stream 2 (lb/hr)	Exit Stream 1 (lb/hr)	Exit Stream 2 (lb/hr)
Material Handled: Nitrogen, Ammonia, Water, Nitrogen Oxide, Nitric Acid, Nitrogen Dioxide				
Oxygen:	0	0	0	0
Nitrogen:	0	354635.2	0	354636.2
Ammonia:	595.25	0	595.25	0
Water:	61993.91	0	61993.91	0
Nitrogen Oxide:	0	3218.745	0	18959.73
Nitric Acid:	0	0	66138.6	0
Nitrogen Dioxide	0	73920.91	0	1481.505
Total:	62589.16	431774.8	128727.76	375077.435
Temperature (°F):	148	197	170	70
Pressure (psig):	145	145	142	145
Wall Thickness (ft):	0.8125			
Height (ft.)	100.7			
Diameter (ft):	12			
Number of Stages	36			
Material	Stainless Steel			
Construction:				
Tray Type	Sieve			

Table 13: Absorption Tower Specification Sheet

9 Economic Evaluation

The economic aspect of the Nitry Inc. Plant was determined using the given Capcost economic analysis document and the Cost and Evaluation Workbook accompanying the Plant Design and Economics for Chemical Engineers [Peters]. The Capcost sheet was used to gather the total price of the major and minor units of the plant. The total cost of purchased equipment for the plant was estimated to be \$40.8 million [CapCost]. However, after applying the cost of installation of the equipment, delivery fees, piping installation, electrical systems, buildings, yard improvement and service facility cost, the total direct cost was estimated at \$162 million.

Another adding factor to the total equipment cost and direct cost being high is the material of all units, piping and product transportation vessels will be made of stainless steel.

For the materials and labor of the plant, the cost of the raw material, ammonia, in 2019 was priced at \$0.37 per kilogram. With the amount of ammonia that will be needed for this plant's yearly capacity, the plant will spend \$40 million on raw materials. Following the price of ammonia, the selling price of nitric acid in 2019 and also the price that Nitry Inc. would project to sell their nitric acid for would be \$0.31 per kilogram, bringing the plant \$74 million per year from the selling of their product. Lastly, the average salary for nitric acid plant operators in Texas (based on the average from comparably database) for 2019 was about \$53,000 per year which falls at least 10% above the national average salary of these operators. Due to the low supervision and controlling that is needed for a nitric acid plant, the plant will have a team of 2 plant operators per shift with 2 shifts per day, a day shift and a night shift. The plant will have a front-end team that works at the beginning of the week, typically Sunday to Wednesday and a back-end team that will work from Wednesday to Saturday.

An economic analysis is typically done during the beginning and planning stages of chemical plant construction but for the purposes of this project, we calculated economic cost at the end of the project. Unfortunately, the return on investment for the Nitry Inc. plant would be -17.0% with a payback period of -9.4% [Peters]. This means that through the life of the plant it would take almost 10 years of production for the plant to begin making money. The cost analysis which was done shows that the plant would not be feasible, and it is not recommended to move forward with the building of this plant. All plant cost and economic information may be found in the tables below.

Where is your
profitability graph?
that shows
payback period

Fluid Processing Plant	Fraction	Calculated Values, \$Million
Direct Costs		
Purchased Equipment, E'		40.8
Delivery, fraction of E'	0.1	4
Subtotal: Delivered Equipment		45
Purchased Equipment Installation	0.47	21
Instrumentation & Control (Installed)	0.36	16
Piping (Installed)	0.68	30.5
Electrical Systems (Installed)	0.11	5
Building (Including Services)	0.18	8
Yard Improvements	0.1	4.4
Services Facilities (Installed)	0.7	31.4
Total Direct Cost	2.6	162

Table 14 : Total Direct Cost of Nitry Inc. Plant – Cost & Evaluation Workbook

Indirect Cost		
Engineering and Supervision	0.33	15
Construction Expenses	0.41	18.4
Legal Expenses	0.04	1.8
Contractor's Fee	0.22	10
Contingency	0.44	19.7
Total Indirect Cost	1.44	64.6
Fixed Capital Investment (FCI)		226

Working Capital (WC)	0.89	40
----------------------	------	----

Total Capital Investment (TCI)	266
--------------------------------	-----

Table 15 : Total Indirect Cost and Total Capital Investment of Nitry Inc. Plant – Cost & Evaluation Workbook

Annual Expenses

Items	Calculated Values, \$Millions
Raw Materials	39.96
Operating Labor	0.3
Operating Supervision	0.3
Utilities	2.0
Maintenance and Repairs	232
Operating Supplies	14
Laboratory Charges	0.3
Royalties (if not, on lump-sum basis)	85.3
Catalyst and Solvents	
Variable Cost	59
Taxes (Property)	4.6
Insurance	2.3
Fixed Charges	7
Plant Overhead	8.5
Manufacturing Cost	75
Administration	2.8
Distribution & Selling	4.2
Research & Development	3.4

General Expense	10.5
-----------------	------

Total Product Cost without Depreciation	85.3
---	------

Table 16: Annual Total Product Cost for Nitry Inc. Plant- Cost and Evaluation Workbook, Operational Cost

The Capcost sheet was used to calculate the total cost of each piece of equipment that will be needed to complete the Nitry Inc. chemical plant.

Equipment	Cost
Evaporator	\$78,800
Mixer	\$133,000
Air Filter	\$755,000
NH ₃ Oxidizing Plug Flow Reactor	\$88,800
Cooling Tower	\$599,000
Knockout Drum	\$88,700
Absorption Tower	\$1,920,00
Stripping Column	\$117,000
Heat Exchangers (x10)	\$297,000
Air Compressors (x4)	\$2,293,000
Pumps (x1)	\$31,400
Total	\$40,800,000

Table 17 : Total Equipment Cost for Nitry Inc. Chemical Plant

10 Safety and HAZOP Analysis

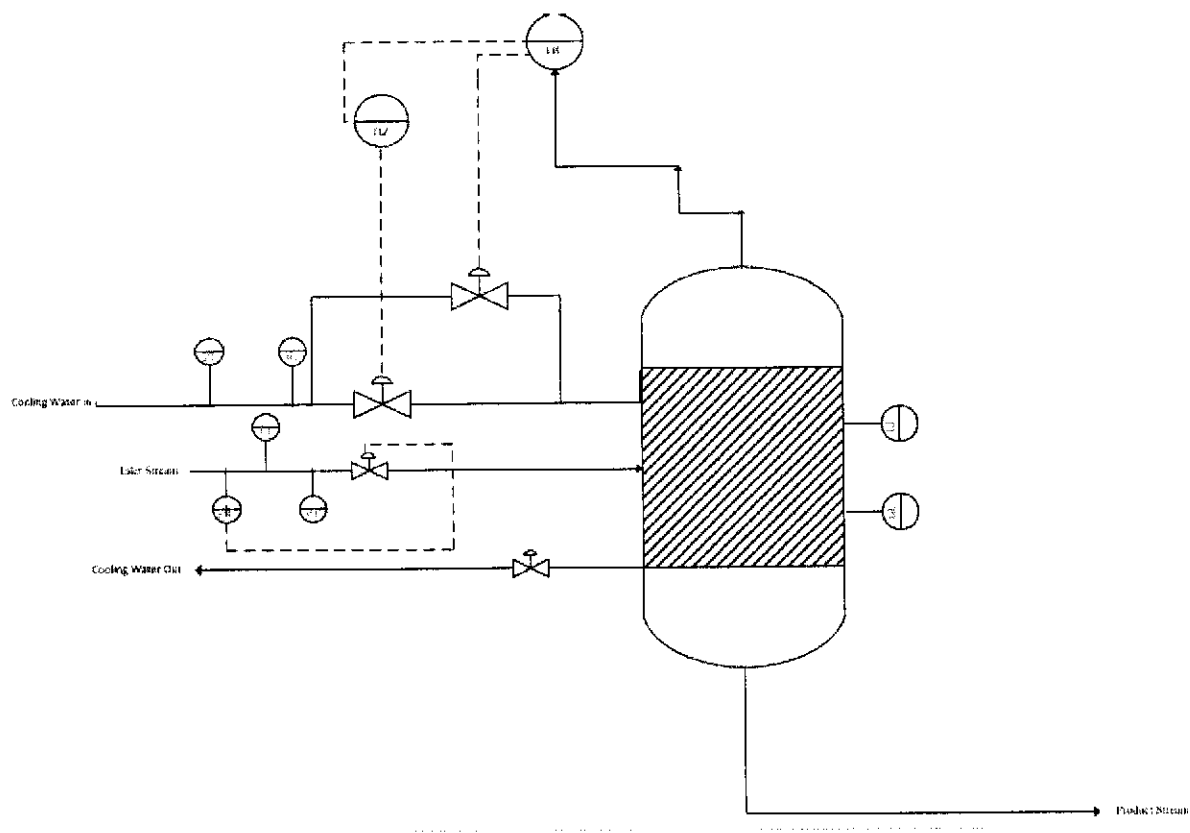


Figure 13: HAZOP Analysis for Cooling Water of NH₃ Oxidizing Plug Flow Reactor

No.	Item	Function/ Purpose	Parameter	Guideword	Deviation	Consequence	Cause	Hazard
1	2	3	4	5	6	7	8	9

1	Cooling Water	Reactor System	Flow	No	No cooling water flow to reactor	Gives reactor potential to overheat leading to possible explosion	Blockage of cooling water flow pipe, malfunctioned control valve for cooling water entry. Break in reactor coolant pipe	Overheating of reactor, runaway reaction and potential explosion
2	Cooling Water	Reactor System	Flow	More	Too much cooling water to reactor	Gives reactor potential to overcool and slow down reaction	Malfunctioned control valve for cooling water entry cannot	Overcooling and decrease reactor temperature

							close. Broken valve	
3	Cooling Water	Reactor System	Flow	Less	Not enough cooling water to reactor	Slow down heating of reactor	Cooling water control valve blockage, partially open valve	Overheating, increased temperature and runaway reaction
4	Feed Inlet Stream	Reactor System	Flow	More	Too much cooling water to reactor	Possible overflow of reactor	Inlet flow rate being above flow rate that can be held by inlet flow, inlet valve malfunction	Overflow of reactor

5	Feed Inlet Stream	Reactor System	Concen tration	High	Ammonia, Water or Air concentration is higher than what is needed by the reactor to react	Ammonia is highly flammable at high volumes in air		Improper reaction taking place
---	-------------------------	-------------------	-------------------	------	---	--	--	--------------------------------------

Table 18: Safety Analysis for Reactor

These are not really "consequences", "cause", and "hazards"

11 Discussion

The main goal of this project is to design a plant that produces nitric acid from the raw materials of ammonia and oxygen. After reaching out to El Dorado Nitrogen, it is concluded that the selling price of nitric acid in 2019 was about \$450 per ton, we are set to have a return on investment that will not make up for the cost of our raw materials which makes this plant unfeasible. Although this plant would cause no harm to the environment while also operating safely, the economic value will not be beneficial.

is this competitive?

12 Conclusion and Recommendations

This plant was designed to produce 400,000 tons of nitric acid per year in a chemical plant in Baytown, Texas. This location was based on close proximity to a water source and also being in an industrial area with an existing pipeline for the transport of the raw material. This aids the process selection of the dual pressure Ostwald process by having an accessible body of cooling water and a proper area for the release of tail gas into the atmosphere.

A process flow diagram was created for the plants operation process and was based off of comparison of multiple nitric acid plants both in literature and from a site visit. After a detailed

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look into the operating procedure and process, the pressure and temperature are more aligned with operating guidelines for a plant this size. For this project, the Aspen Plus simulation was not able to converge but converged for every portion leading up to the absorption tower. The absorption tower was the most complex portion of the plant due to the change from nitrogen dioxide to nitric acid amongst the trays. The complexity of this process shows that temperature and pressures throughout the process will always vary. The recommendation is that more attention be added to the units that will be needed for the process so the plant design will be economically feasible for production, return on investment and a positive payback period.

Very good report. Missing detailed reactor design, trade-off analysis, sensitivity analysis, cost flow diagram, etc.

- Also, this is not well-organized.

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13 Appendix

Stripping Column Design

- The Kremser equation for theoretical stages:
- $$N = \frac{\ln \left[4 + (1-A) \left(\frac{x_2 - \frac{y_1}{m}}{x_1 - \frac{y_1}{m}} \right) \right]}{\ln \left(\frac{1}{A} \right)}$$
- Where, $\frac{1}{A} = \frac{mV}{L}$ where $\frac{mV}{L}$ - Stripping factor
- Calculation of Efficiency using Drickamer - Bradford Equation
- $E_o = 13.3 - 66.8 \log(\mu)$
- Since, $N_{\text{theo}} = 5$ stages,
- $E_o = 0.4$
- $N_{\text{act}} = 13$ stages
- Calculation of vapor velocity
- $$V_{nf} = K_1 \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$
- $V_{nf} = 2.3 \frac{\text{ft}}{\text{s}}$
- Actual Vapor Velocity 50-90 % of net vapor velocity
- $V_n = 0.80 \times V_{nf}$
- $V_n = 1.84 \frac{\text{ft}}{\text{s}}$
- Diameter of the stripping column
- Net area for separation process
- $A_n = \frac{m \cdot v}{V_n}$
- with $m_v = 122.4 \frac{\text{ft}^3}{\text{s}}$
- $A_n = 68 \text{ ft}^2$
- Assuming down comer occupies 15%
- $A_c = \frac{A_n}{0.85} = 80 \text{ ft}^2$
- $$D_c = \sqrt{\frac{A_c \times 4}{\pi}}$$
- $D_c = 10.1 \text{ ft}$
- Height of the Stripping Column