

Production of Light Olefins from Methanol

Final Design Report

04/25/2025



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LIST OF NOMENCLATURE AND NOTATIONS

Abbreviation or Symbol	Description
Process Nomenclature	
DCS	Distributed Control System
LEL	Lower Explosive Limit
MtO	Methanol to Olefins
SAPO	Silicoaluminophosphate
ZSM	Zeolite Socony Mobil
Chemical and Utility Abbreviations	
C2	Hydrocarbons containing two carbon atoms
C3	Hydrocarbons containing three carbon atoms
C4	Hydrocarbons containing four carbon atoms
C5+	Hydrocarbons containing more than four carbon atoms
CR	Condensate return
CHWR	Chilled water return
CHWS	Chilled water supply
CWR	Cooling water return
CWS	Cooling water supply
DME	Dimethyl ether
HP	High pressure
LP	Low pressure
SS	Saturated steam
Equipment	
BL	Blowers, fans
C	Compressors
E	Heat exchangers
H	Direct fired heaters
P	Pumps
PSV	Pressure Safety Valve
R	Reactors
S	Separators
T	Towers
TK	Storage tanks
V	Valves
Units	
Mass Flow Rate	kg/h
Volumetric Flow Rate	m ³ /s
Velocity	m/s
Mass Flux	kg/m ² s
Length	m
Area	m ²

Volume	m^3
Thickness	mm
Density	kg/m^3
Surface Tension	dyne/cm
Viscosity	Pa^*s
Power	kW, MW
Head Loss	m
Overall Heat Transfer Coefficient	$\text{W}/\text{m}^2*^\circ\text{C}$
Heat of Vaporization	kJ/kg
Specific Heat	$\text{kJ}/\text{kg}^\circ\text{C}$
Temperature	$^\circ\text{C}$
Pressure	bar
Stress	N/mm^2
Olefin Production Rate	MMTPA
Economic Terminology	
DPBP	Discounted Payback Period
IRR	Internal Rate of Return
NPV	Net Present Value
ISBL	Inside Boundary Limits
OSBL	Outside Boundary Limits
FCI	Fixed Capital Investment
TCI	Total Capital Investment

EXECUTIVE SUMMARY

MichiChem has identified the methanol to olefins (MtO) process as a potentially greener alternative than steam cracking to produce ethylene, propylene, and other olefins with large market shares and values, especially as green methanol could be entering the market in the near future. Thus, we were tasked with designing a methanol to olefins production facility, proposing an optimal production capacity and plant location, and financially evaluating the proposed facility to provide MichiChem with recommendations on whether to proceed with the plant. The MtO process relies on the SAPO-34 catalyst to break apart methanol first into a dimethyl ether intermediate and then into olefins (ethylene, propylene, butylene, C5+ compounds), paraffins (methane, ethane, and propane) and large amounts of water.

Based on cost of materials and existing market size, our design considers a plant location in China producing 0.5 million tonnes per year of polymer-grade ethylene and propylene. Although a wide number of designs including D-MTO, S-MTO, and the Lurgi Process have industrially produced olefins, we proceeded with the design by UOP/Norsk Hydro due to the availability of data on product distributions out of the reactor. Although alternative designs for catalysts have been proposed, we proceed with a traditional SAPO-34 catalyst and emphasize putting further research into nanosized and Ca-modified silicoaluminophosphate catalysts. The overall design of the project relies on the reaction of the methanol to olefins, purification via absorption, scrubbing, and drying, and product separation with distillation columns.

Our team designed a methanol unloading station with a tank farm, integrated heat exchange with methanol feed and reactor effluent, a reactor/regenerator system including catalyst fluxes and cooling duties required, cooling via quench towers and heat exchangers, removal of methanol by quenching and absorption, removal of carbon dioxide and water by scrubbing, removal of additional water by drying, a system of seven distillation columns to separate the products/byproducts (methane, ethane, ethylene, propane, propylene, butylene, and C5+ compounds), a storage and unloading system for C5+ compounds, and three refrigeration loops to provide cooling for the cryogenic de-methanizer, de-ethanizer, and C2 splitter. The design contains equipment sizes and materials of construction, utilities required, and rough estimates of pressure drops through lines and pieces of equipment. It also includes a sketch of a proposed plant layout.

We also provide recommendations for environmental, health, and safety considerations, including the installation of pressure relief equipment and containment devices along with personal protective equipment and routine inspections and maintenance. We suggest adding a wastewater treatment plant to decontaminate quencher water.

Based on our Class 5 economic analysis, we do not recommend pursuing this project as the net present value (NPV) of the plant is estimated at \$-2.86 billion after 20 years of operation. However, if utility and waste costs are cut by 15% and methanol costs decrease to \$212 per tonne, the process could become profitable, with an NPV of \$114 million after 20 years.

INTRODUCTION

The following sections provide a discussion of the problem statement for our overall process as well as the specific objectives we aim to complete.

Problem Statement

Olefins are an important class of unsaturated hydrocarbons widely used for the manufacturing of plastics, rubber, detergent, and adhesives [1]. Currently produced by steam cracking of light hydrocarbons derived from non-renewable sources such as oil, alternatives using compounds sourced from renewable, less carbon-intensive processes are increasingly sought after [2].

MichiChem has identified green methanol as a potential carbon source entering the market within the next ten years. Establishing a light olefin production facility using methanol as an input provides MichiChem an opportunity to invest in its future, synthesizing chemicals like propylene and ethylene from renewable carbon sources.

The core objectives of this project are to design a methanol to olefins (MtO) production facility, propose an optimal production capacity and plant location, and financially evaluate the proposed facility to provide MichiChem with recommendations on whether to proceed with the plant. The project requires deliverables including a conceptual plant design for a plant producing 0.5 million tonnes per year (MMTPA) of polymer-grade olefins (ethylene and propylene) as well as a variety of hydrocarbon byproducts, a detailed plant design with operating conditions, and an economic feasibility report of the process.

Having successfully delivered a conceptual design report for our process in accordance with the project timeline, we now present a detailed plant design. This report aims to present the project team's systematic approach to the MtO project, providing MichiChem with the research, analysis, and design process behind the proposed MtO plant. The report covers topics including the basics of MtO processes, possible alternatives to our design, a summary of our conceptual design, a block flow diagram describing our process broadly, individual process stage diagrams and descriptions, analysis of potential safety concerns within our process, and management of waste within our process.

Project Objectives

The overall design of our process has been set to satisfy the core objectives of the project. We have designed our MtO production facility to produce 0.5 MMTPA of light olefins via the dehydration of methanol, producing approximately equal amounts of polymer-grade ethylene (>99.9 wt%) and propylene (>99.5 wt%). In addition to our light olefins products, we also produce a number of byproducts including butylene, methane, ethane, propane, and C5+ hydrocarbons. The light olefins will be sent to adjacent polymer production facilities via pipelines, and the byproducts will be sold via pipelines (methane, ethane, propane, and butylene) or will be stored on site and sold via tanker trucks (C5+ hydrocarbons). Our plant has been designed for 24/7 operation for a total of 340 days annually with a methanol and C5+ hydrocarbon storage capacity of 3 days of operation to account for supply chain disruptions.

BACKGROUND

The following sections provide necessary background information for our proposed process. Specifically, discussions of the basics of MtO processes, possible alternatives processes to our design, possible alternative decisions within our design, and a conceptual design of our process are included.

Basic Process Information

The following sections provide an overview of the MtO process, with specific emphasis on the UOP/Norsk Hydro process, the process on which we based our design.

Reaction. The general goal for all MtO processes is to produce ethylene and propylene via the dehydration of methanol. This reaction is accomplished with the use of a catalyst, typically SAPO-34, but the reaction is not highly selective, producing a variety of olefins, aliphatic hydrocarbons, and aromatic compounds [3] [4]. Figure 1 below depicts the general reaction scheme.

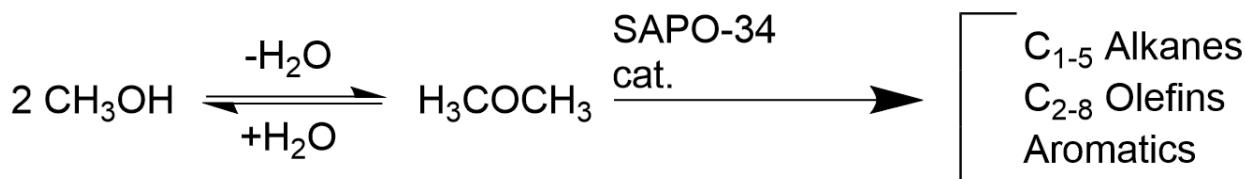


Figure 1. Reaction scheme for MtO using SAPO-34 as the catalyst [3]

In the UOP/Norsk Hydro process, this reaction is typically carried out in a reactor analogous to the FCC reactors used in petroleum refineries. Specifically, the reactor is a fluidized bed containing two distinct regions of fluidized catalyst, the dense region and the transition region. Methanol is vaporized and pressurized before being fed into the dense region of the reactor where it reacts essentially to completion before passing through the transition region and then into a series of cyclones, removing entrained catalyst particles before the effluent leaves the reactor. The reaction produces significant amounts of coke that covers the active surface of the catalyst particles. Accordingly, a regenerator unit is also attached to the reactor to continuously regenerate the catalyst by burning off the coke [5] [6] [7] [8].

Product Purification. The effluent from the reactor contains the desired olefin products and a mixture of additional byproducts and waste. Accordingly, the effluent is passed through a variety of separation units to purify and isolate the products. It is sent through a quencher to remove water and residual methanol, a caustic scrubber to remove carbon dioxide, a dryer to remove residual water accumulated in the scrubber, and then a series of distillation columns to separate the products and byproducts. Given the very low boiling points of ethylene, propylene, and the aliphatic byproducts, some of the distillation columns operate under cryogenic conditions, requiring additional design in the form of refrigeration systems [5] [6] [7] [8]. A simplified diagram of the process as a whole is depicted below in **Figure 2**.

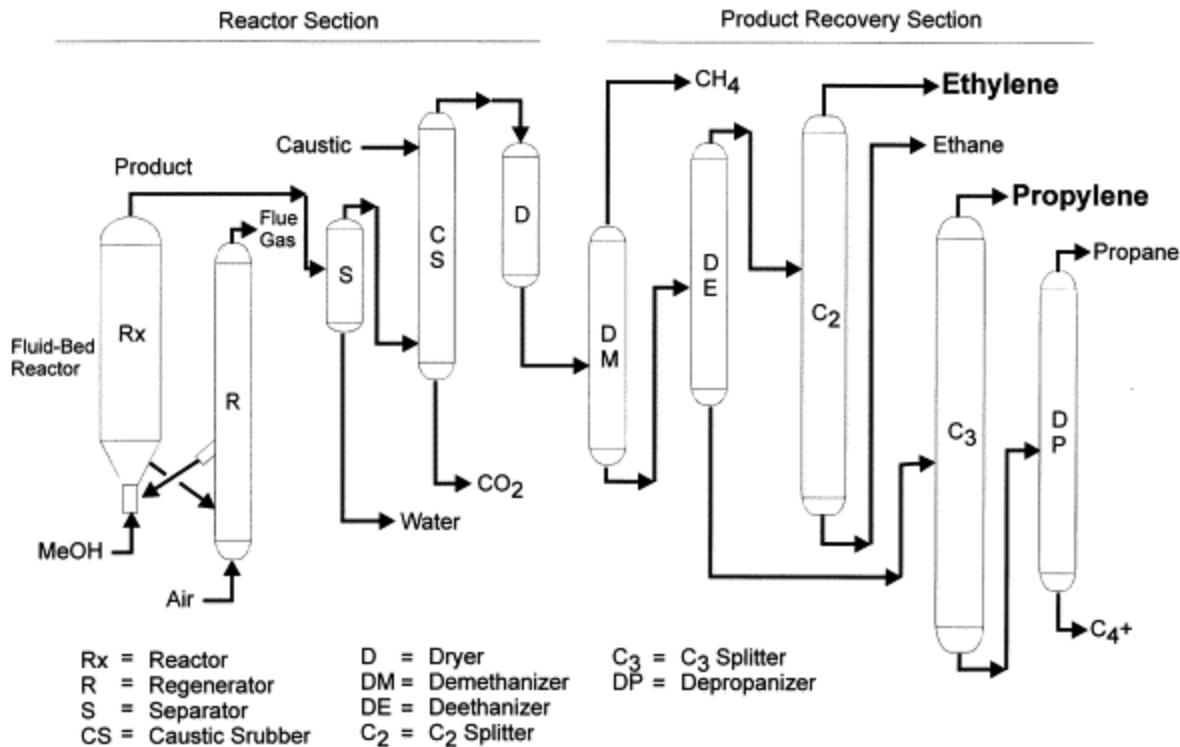


Figure 2. A flow diagram depicting the process units used in the UOP/Norsk Hydro MtO process [8]

Alternative Processes to our Design

The following section details alternative process designs that we did not pursue for this project. Specifically, discussions of the D-MtO/S-MtO process as well as the Lurgi process are included.

D-MTO/S-MTO Process. The D-MTO and S-MTO processes are relatively newer MtO processes developed and utilized primarily by Chinese petrochemical companies. The processes are nearly identical apart from the fact that S-MTO uses a modified version of the SAPO-34 catalyst that yields a higher ethylene to propylene product ratio [8]. Given this rather small difference and the similarity in all other aspects of the processes, the two processes will be hereafter treated as a single process.

The D-MTO/S-MTO process shares many characteristics with the UOP/Norsk Hydro MtO process, but there are a couple of key differences. The biggest difference is that the two processes work with streams in different phases. In contrast to UOP/Norsk Hydro, which works with a vaporized feed and keeps the reactor effluent in the vapor phase, D-MTO/S-MTO uses a liquid feed and recondenses the reactor effluent for later purification steps. Similar to the UOP/Norsk Hydro process, the D-MTO/S-MTO process begins by passing a stream of methanol through a fluidized bed reactor; however, the process differs from UOP/Norsk Hydro in that the reactor is typically operated in a slightly different range of conditions: it is operated in a temperature range of 400-500°C and a pressure range of 1-3 bar. Naturally, given the difference in stream phases, the effluent from the D-MTO/S-MTO reactor is processed in a different but analogous way to the effluent from the UOP/Norsk Hydro reactor [8] [9]. A simplified flow chart of the process can be seen below in **Figure 3**.

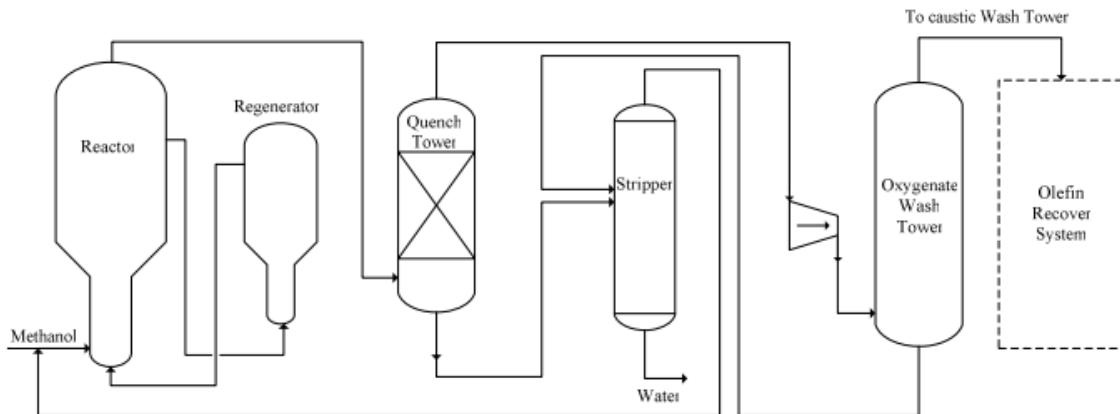


Figure 3. A flow diagram depicting the general process units of the D-MTO/S-MTO process [10]

As can be ascertained from **Figure 3**, the general flow of D-MTO/S-MTO is very similar to UOP/Norsk Hydro. Of course, **Figure 3** does not include the train of separation units for isolating the desired products, but they will typically be very similar to those shown in **Figure 2**.

Ultimately, we did not choose to base our design on the D-MTO/S-MTO process due to the lack of readily accessible information on product distributions within D-MTO/S-MTO reactors. Literature on the UOP/Norsk Hydro process provides tabulated product distributions for given reactor operating conditions, but we were unable to find similar information for the D-MTO/S-MTO process, so we elected to base our design around the process with more publicly available data.

Lurgi Process. The Lurgi process differs from other MtO processes in that it specifically targets propylene as the sole desired product, naturally leading to a different flow of process units as well. In general, the Lurgi process begins by feeding a vaporized methanol feed to a dimethyl ether (DME) reactor operating at a temperature of 260°C. In this reactor, an acidic catalyst is used to dehydrate the methanol and form DME and water, achieving a total conversion of 75%. The DME, water, and remaining methanol are then heated to 400°C and passed through a series of three fluidized bed reactors with an ultimate DME/methanol conversion over 99% and propylene being the predominant product. After the reactor train, the effluent is processed similarly to the UOP/Norsk Hydro process with the exception of the target final product being specifically polymer-grade propylene [8] [10]. A simplified flow diagram of the Lurgi process is depicted below in **Figure 4**.

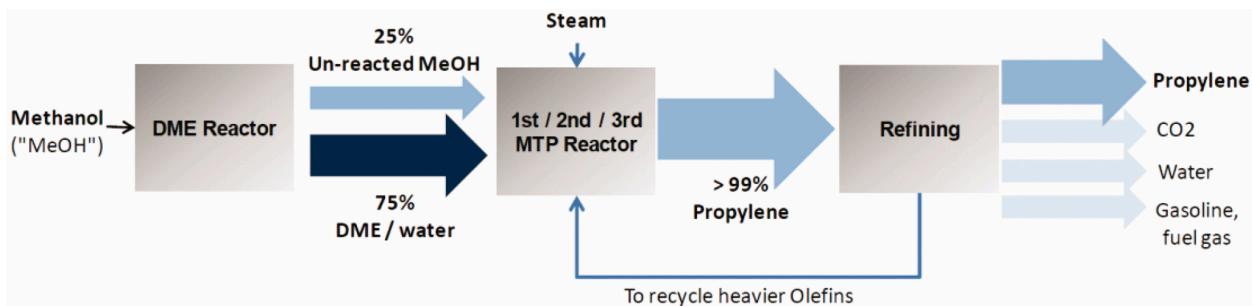


Figure 4. A flow diagram depicting the Lurgi process [8].

Similar to the D-MTO/S-MTO process, we were unable to find data on product distributions in the reactor under given operating conditions for the Lurgi process. Accordingly, we decided that the UOP/Norsk Hydro would be the better process around which to base our design.

Alternative Decisions within our Design

This section discusses possible alternate approaches to the MtO process that may not be fully commercialized. Due to the likely shortage of crude oil in the foreseeable future, the MtO process has undergone tremendous amounts of research and development to further improve the process and search for more efficient catalysts.

Ca-modified HZSM-5 Catalyst. The MtO process uses a well-known acid-catalyst reaction to produce light olefins; however, the reaction of methanol over acid catalysts leads to the formation of undesired byproducts such as heavy olefins [11]. Research suggests that the addition of alkaline earth metals, such as calcium, may be able to effectively lower the catalyst acidity and improve the catalyst life and selectivity for light olefins. By loading different percentages of calcium into the HZSM-5 catalysts, a study was able to conclude that the calcium modification of the HZSM-5 catalyst increased both the catalytic stability and selectivity for light olefins in the MtO process; however, they found that there is an optimal amount of calcium loading, as adding too much to the catalyst would cause the basic sites to be too strong and stop the reaction from effectively proceeding, thus decreasing selectivity for light olefins [11].

Nanosized SAPO-34. SAPO-34 is one of the main catalysts used in current MtO processes and is mainly favored for its exceptionally high selectivity of light olefins, typically above 80%. However, SAPO-34 suffers from rapid deactivation due to coke formations from retained organic species in its cavity and pores. To alleviate this diffusion limitation, the intracrystalline diffusion path length can be shortened by reducing the crystal size, which is done by creating nanosized SAPO-34 through various methods such as microwave synthesis, dry gel conversion, and high-temperature synthesis to list a few. This reduction in diffusion path length greatly enhances mass flow to and away from catalytic centers, overcomes the diffusion limitation, prohibits undesired secondary reactions, reduces coke formation, and increases catalyst lifetime [12].

Though both alternative catalyst choices could help improve the overall selectivity for light olefins and reduce regeneration costs by lowering coke formation, they are not currently available at an affordable commercial scale. Accordingly, though we recommend that Michichem consider exploring these options in its research division, we decided to design our process using normal SAPO-34.

Conceptual Design

Our proposed MtO process for 0.5 MMTPA ethylene and propylene is based on the UOP/Norsk Hydro process rather than the D-MTO/S-MTO and Lurgi processes due to the more readily available data and widespread use. In addition to ethylene and propylene, other hydrocarbons such as methane, ethane, propane, C4, and C5+ will be produced and sold. At a high level, the process progresses through four main stages: methanol unloading and feed preparation, conversion of methanol to light olefins,

purification of the reaction effluent to remove non-hydrocarbon impurities, and separation of the hydrocarbons into pure products.

The methanol unloading process concerns the unloading of methanol from trains into tanks and the subsequent pumping of methanol into the pipes and heat exchangers. The tanks must be large enough to hold three days worth of methanol feedstock while being short enough to minimize hydrostatic pressure.

The conversion of methanol to light olefins process begins with heated and vaporized methanol entering the reactor containing SAPO-34 catalyst in a fast fluidized bed, where the methanol reacts to form a variety of olefins, paraffins, and aromatics, but primarily ethylene and butylene. The reaction also produces coke, which decreases the selectivity and activity of the catalyst and is burned off in the regenerator. The regenerator is also a fluidized bed reactor feeding air to burn off coked catalyst, returning more pure catalyst to the reactor. The reactor-regenerator system requires significant cooling, as both reactions are exothermic, and temperatures exceeding 500°C in the reactor can alter product purity.

The reaction effluent, which contains the aforementioned hydrocarbons in addition to significant amounts of water, some unreacted methanol, and small amounts of carbon dioxide, is cooled directly with water in quench towers. The cooled mixture goes through an adsorption tower to remove nearly all the methanol with significant amounts of water. The quenched vapor stream then goes through a caustic scrubber containing aqueous NaOH to remove carbon dioxide, which can disrupt cryogenic distillation by freezing into dry ice. The resulting vapor stream is sent to the dryer to remove The quenched reaction mixture will be sent to the caustic scrubber, which will combine with a flow of aqueous NaOH. This basic solution will remove carbon dioxide by converting it to sodium bicarbonate and carbonate. The resulting hydrocarbon stream containing some water will be sent to the dryer, where temperature-swing adsorption beds selectively adsorb water from the mixture and then regenerate by having hot air blown over them, creating humid air. Similarly to carbon dioxide, water must be removed prior to cryogenic distillation to prevent solids from accumulating within the column.

The more pure hydrocarbon stream is sent to a distillation train after undergoing some cooling and compression. Distillation was chosen as the preferred method of separation over other liquid-liquid methods such as liquid-liquid extraction and pervaporation because it is well-established for hydrocarbon separations and because the products have a wide range of boiling points (-161.5°C for methane compared to 36.1°C for pentane, both at atmospheric pressure) that allow for relatively easy separation. The purified hydrocarbons are then separated from one another in a distillation train. The distillation train includes a de-methanizer, which separates methane from the rest of the mixture; a de-ethanizer, which splits C2 hydrocarbons from heavier C3+ hydrocarbons; C2 splitter, which splits ethane and ethylene; de-propanizer, which separates C3 hydrocarbons from heavier C4+ hydrocarbons; C3 splitter, which splits propane and propylene; and the C4/C5 splitter, which separates the vapor-phase C4 compounds from the liquid-phase C5+ compounds. All but the C5+ hydrocarbons are sent directly to a pipeline, while the C5+ products are pumped to storage tanks, which are loaded into trucks.

The plant location for the proposed facility has been recommended to be in China, due to the relatively cheap cost of raw materials, as compared to locations with similar infrastructure such as the United States Gulf Coast or Saudi Arabia. Additionally, the majority of operating MtO facilities are located in China, so the expertise of the available workforce in the Northeast Asia region is of interest to MichiChem.

PROCESS DESIGN

The following section details the process design of a proposed MtO process for producing 0.5 MMTPA polymer-grade ethylene and propylene that is adapted from the UOP/Norsk Hydro process. To explain the design at a high level, this section provides a block flow diagram depicting the major operational units in the process and their interconnectivity. Additionally, this section provides stage diagrams supplemented with descriptions of each stage to further detail each step of the process.

Overall Process

The overall MtO process was broken down into 12 stages, nine for the main sequence of unloading methanol, reaction of methanol to olefins, purification of hydrocarbons, separation of hydrocarbons, and shipping of C5+ compounds, and three for maintaining refrigeration loops, necessary for the cryogenic distillation columns. The nine main stages are, in order from Stage 1 to 9, Methanol Storage and Unloading, Reaction of Methanol to Hydrocarbons, Hydrocarbon Purification, Distillation Train Preparation, Separation of C1 and C2 Hydrocarbons, Ethylene Recovery, Propylene Recovery, Butylene Recovery, and C5+ Hydrocarbons Storage and Shipping. The refrigeration stages, Stages 10 through 12, are named after the stage using the refrigeration created in that loop: Separation of C1 and C2 Hydrocarbons Refrigeration, Ethylene Recovery Refrigeration, and Distillation Train Preparation Refrigeration. The block flow diagram for the process is shown in **Figure 5**.

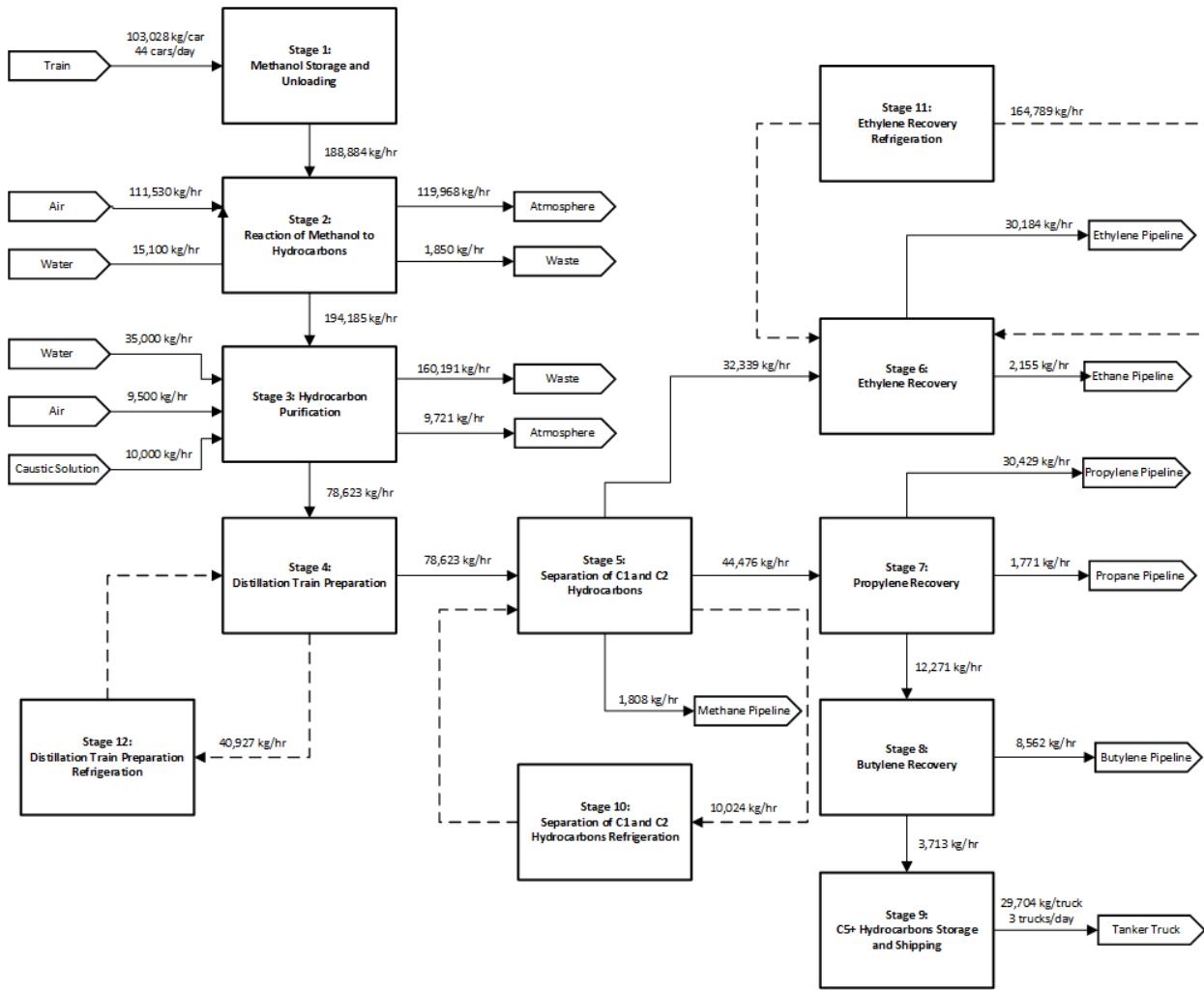


Figure 5: Block flow diagram for the methanol to olefins process

Overall, 44 train cars each holding 103,028 kg methanol are unloaded daily into storage in Stage 1 and then sent to Stage 2 at a rate of 188,884 kg/h. In Stage 2, the methanol is converted into products and byproducts including methane, ethane, ethylene, propane, propylene, butylene, and C5+ hydrocarbons, as well as water, carbon dioxide, and coke. The coke is burned in the regenerator using 111,530 kg/h air, and the resulting flue gas is released to the atmosphere at 119,968 kg/h. After the reaction, the mixture is quenched using 15,100 kg/h water and 1,850 kg/h leaves the second quench tower as waste. The remaining reaction effluent is sent to Stage 3 for product purification at 194,185 kg/h. This product vapor is sent through a series of quenching, scrubbing, and drying steps, where 35,000 kg/h water is used for quenching, 10,000 kg/h caustic solution is used for scrubbing, and 9,500 kg/h air is used for drying. Waste streams containing mostly water leave at 160,191 kg/h, while 9,721 kg/h air containing carbon dioxide enters the atmosphere. The remaining 78,623 kg/h hydrocarbons exit to Stage 4. In Stage 4, the hydrocarbon mixture is compressed and cooled to cryogenic conditions and sent to Stage 5 at the same flow rate. In Stage 5, 1,808 kg/h methanol is removed in the first distillation column and exits via a pipeline, while the second distillation column separates the C2 and C3+ products. 32,339 kg/h ethane and ethylene goes to Stage 6, and 44,476 kg/h C3+ goes to Stage 7. In Stage 6, the distillation column separates the ethylene, leaving at 30,184 kg/h to exit in a pipeline, with the ethane, which leaves at

2,155 kg/h, also leaves in a pipeline. In Stage 7, two distillation columns separate the C3+ components into propylene at 30,429 kg/h, propane at 1,771 kg/h, and C4+ at 12,271 kg/h. The propylene and propane exit to pipelines, while the C4+ enter Stage 8. Stage 8 contains the final distillation column in the process, which separates the butylene, leaving for a pipeline at 8,562 kg/h from the heavier C5+ hydrocarbons, which are sent to Stage 9 at 3,713 kg/h. Stream 9 contains tanks and unloads the C5+ hydrocarbons with 3 trucks per day, each carrying 29,704 kg.

Stage 10 cools 10,024 kg/h ethylene to be used for cooling in Stage 5; Stage 11 cools 164,789 kg/h propylene for cooling in Stage 6; and Stage 12 cools 40,927 kg/h additional propylene for cooling in Stage 4.

Stage Descriptions and Diagrams

The following sections provide detailed descriptions and diagrams of each stage within our process.

Stage 1 (Unloading and storage of methanol feedstock)

The purpose of Stage 1 is to transfer methanol feedstock from railcars to tanks and ultimately feeding the methanol to the process. This stage includes the delivery schedule and tank farm design for the sourcing of methanol for the proposed MtO process and is shown in **Figure 6**.

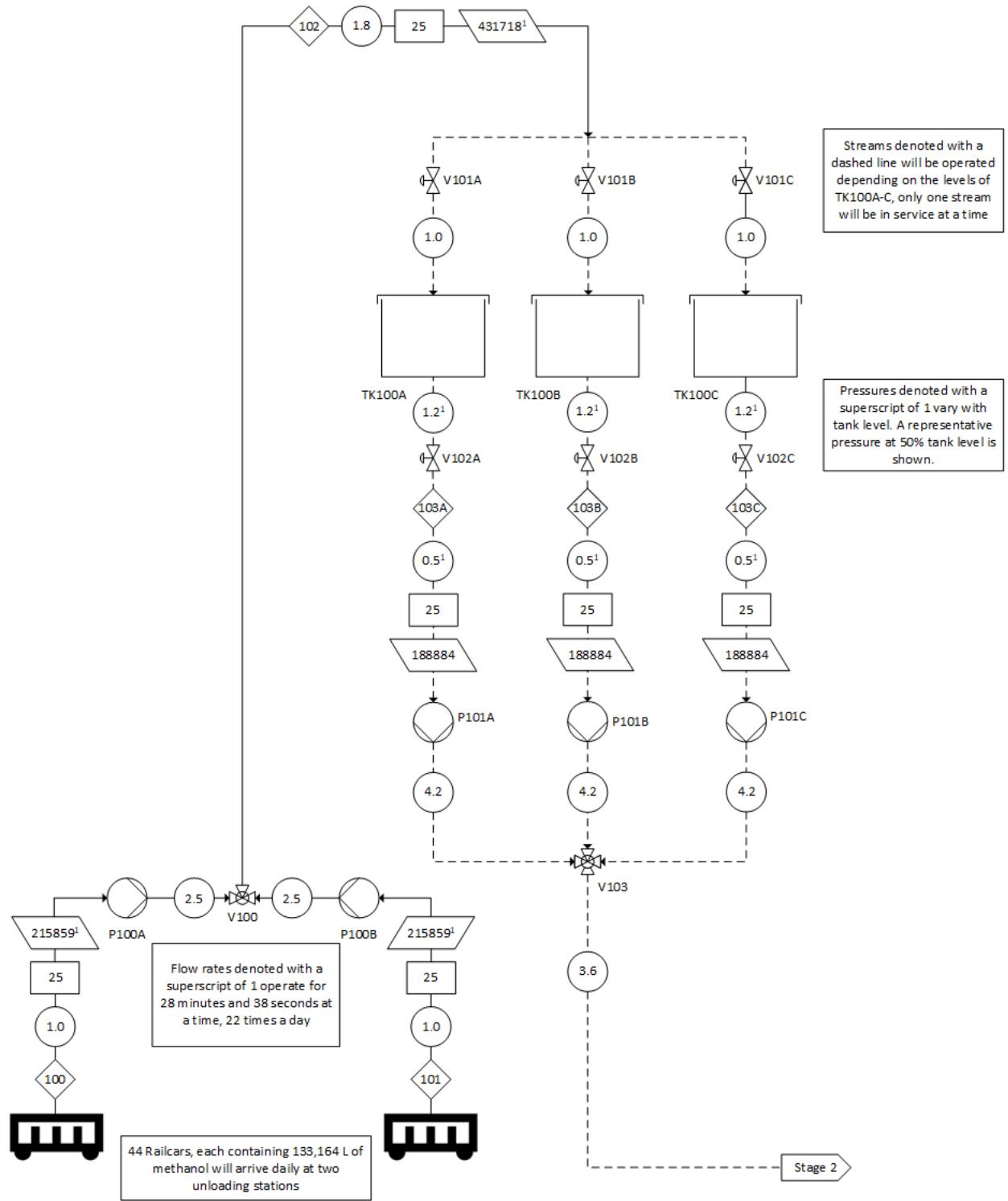


Figure 6: Stage 1 process diagram for the unloading schedule for and storage of methanol

Methanol Unloading. To meet the proposed MtO facility's specified capacity of 0.5 MMTPA olefins, methanol must be transported to the site by railcar. Forty-four railcars will arrive at the site daily, unloaded at two separate unloading stations. Each railcar will contain 133,164 L of methanol at 25°C and 1 bar, and will be unloaded as Streams 100 and 101 at flow rates of 215,859 kg/h, equivalent to 317 LPM for 28 minutes and 38 seconds per railcar. This unloading will be facilitated by a set of two identical stainless steel centrifugal pumps, P100A-B, each with 1.1 kW of power input, which pressurize the methanol to 2.5 bar to accommodate for subsequent pressure drops arising from control valves. Streams 100 and 101 will be added to each other by pumping them through combination valve V100, reducing the pressure of the Stream to 1.8 bar, creating Stream 102, which will then be sent to the methanol tank farm.

Methanol Tank Farm. The methanol tank farm consists of three identically configured tanks, TK100A-C. Constructed from carbon steel, each tank measures 46 m in diameter, 4.60 m in height, and 14 mm in wall thickness, is equipped with an external floating roof, and contributes to a total available storage capacity of 22.9 ML. Tanks will be filled by Stream 102 with the tank being filled varying based on available tank level, and can be changed by opening/closing V101A-C, which will decrease the pressure of the incoming methanol by 0.7 bar. TK100A-C will store methanol at 1 bar and 25°C. Tanks will also be continuously emptied at a rate of 188,884 kg/h to feed the process, but the outlet pressure of a given tank will depend on its liquid level, so a representative outlet pressure at 50% tank level is shown in **Figure 6**. The specific tank being emptied can be controlled by opening/closing V102A-C, which leads to a pressure drop of 0.7 bar, necessitating the installation of P101A-C to increase the pressure of Stream 103A-C to 4.2 bar to overcome the 0.7 bar pressure drop associated with combination valve V103 to supply the stream to Stage 2 at 3.6 bar. P101A-C are three identically configured carbon steel centrifugal pumps, each with 37.3 kW of input power.

Stage 2A (Integrated heat exchange of methanol reactant and hydrocarbon effluent)

The purpose of Stage 2A in our process is to increase the inlet methanol temperature and reduce the exit temperature of the reactor effluent. The inlet temperature of the methanol must increase enough so that in the reactor, the methanol will react to form the desired hydrocarbon products rather than DME, which is favored at lower reaction temperatures [6]. Additionally, the reaction effluent must be immediately cooled to about 420°C to prevent degradation of the products [13], and since the later separations must occur at cryogenic temperatures, the products have further usable heat. **Figure 7** shows the integrated heat exchange process.

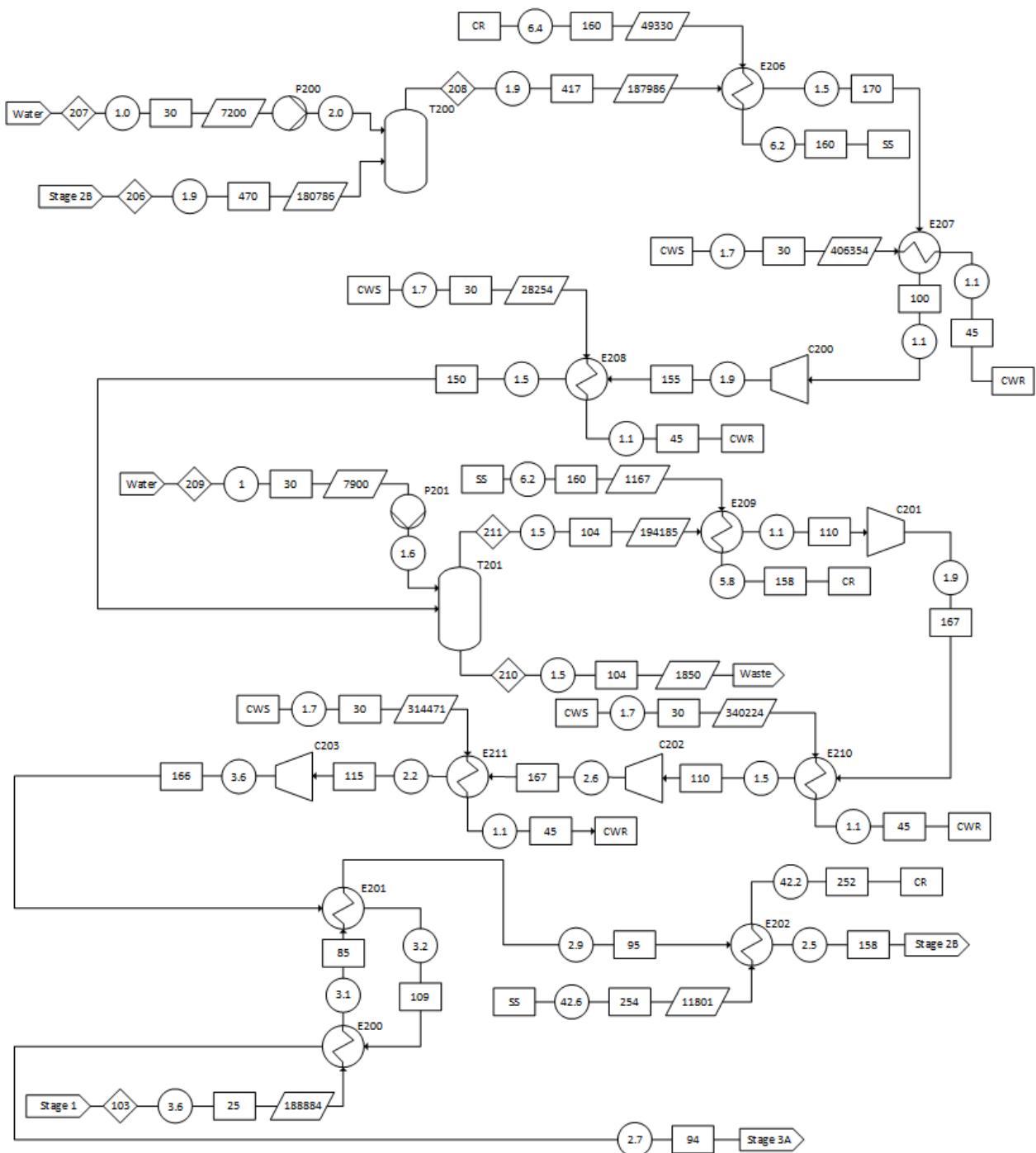


Figure 7: Stage 2A process diagram for the integrated heat exchange of methanol reactant and hydrocarbon effluent

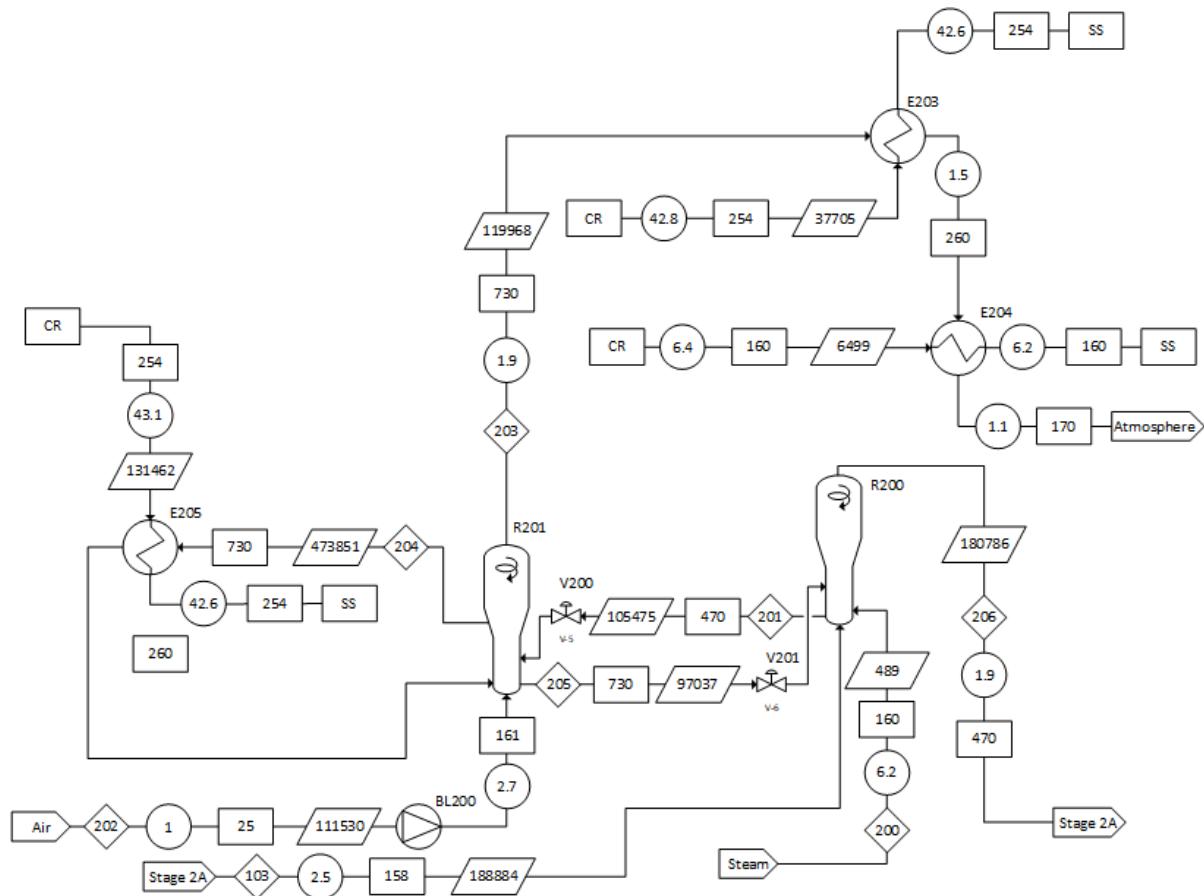
Quenching the Reactor Effluent. To prevent product degradation, the reaction effluent from Stage 2B is sent to a T200, a 21.2m tall, 10.6 m diameter carbon steel quencher. Stream 206 contains 180,786 kg/h effluent at 470°C, which is combined with 7,200 kg/h water at 30°C and pumped to 2 bar using a 0.543 kW carbon steel centrifugal pump in Stream 207. The vapor in Stream 208 leaves at 417°C and enters E206 to create valuable high pressure steam using excess heat. E206 is a 1,596 m² carbon steel shell and tube heat exchanger with a tube outer diameter of 1.59 cm, a shell outer diameter of 1.63 m, and total length of 9.3 m. The vapor leaves at 170°C and enters another carbon steel shell and tube heat exchanger, E207, which has a length of 6.1 m, a shell outer diameter of 0.86 m, a tube outer diameter of 1.59 cm, and an area of 304 m². Due to the pressure drops through the two heat exchangers, the vapor is compressed to 1.9 bar in C200, a 5.49 MW carbon steel centrifugal compressor, thus increasing the vapor temperature to 155°C. To enter the second quencher at the desired temperature of 150°C, the vapor is cooled slightly to 150°C in E208, a carbon steel shell and tube heat exchanger with 0.41 m shell outer diameter, 1.59 cm tube outer diameter, 18 m² area, and 2.2 m length. This vapor is combined with water in Stream 209, which has a flow rate of 7,900 kg/h and is pumped to 1.6 bar with P201, a 0.344 kW carbon steel centrifugal pump. The second quencher T201 is carbon steel with a diameter of 9.2 m and height of 18.4 m, and it is operated at a low enough temperature to have a liquid waste stream, Stream 210, which contains mostly water and flows at 1850 kg/h. This quenching step is assumed to remove most of the remaining solids that were not removed by the cyclones [13]. The remaining product vapor leaves as Stream 211 at 194,185 kg/h and 104°C to undergo additional heat exchange.

Product Side of Integrated Heat Exchange. The resulting vapor must be compressed over 3 bar to undergo additional heat exchange and enter a third quencher in Stage 3A at 2.2 bar, however, compressors cannot operate at temperatures exceeding about 167°C, so the vapor must be compressed in stages with heat exchange in between. Stream 211, a saturated vapor stream, is first heated to 110°C in E209 using saturated steam to ensure the vapor entering the following condenser does not contain any portion of liquid, which can damage the compressor. E209, a carbon steel, 56 m² shell and tube heat exchanger has dimensions of 0.46 m shell outer diameter, 1.59 cm tube outer diameter, and 4.0 m length. The superheated vapor stream exits C201, a 5.87 MW carbon steel centrifugal compressor, at 1.9 bar. The vapor enters E210, a 401 m², carbon steel shell and tube heat exchanger with length 5.8 m and outer diameters of 1.07 m and 1.59 cm for the shell and tube, respectively. Using cooling water, the cooled vapor leaves at 110°C and 1.5 bar and is compressed in C202, a 5.28 MW carbon steel centrifugal compressor, to 2.6 bar and 167°C. Stream 210 continues through E211, a 365 m², carbon steel shell and tube heat exchanger 5.7 m long and with 1.02 m shell outer diameter and 1.59 cm tube outer diameter. The vapor at 115°C is compressed in C203 (a 5.89 MW carbon steel compressor) to 3.6 bar and exits at a temperature of 166°C. The high-pressure vapor then enters E201 to heat the methanol going into the reactor. E201 is a 5,721 m², 11.9 m length, 3.05 m shell diameter, 1.59 cm tube diameter, carbon steel shell and tube heat exchanger, and the resulting vapor enters E200 at 109°C and 3.2 bar. E200 is a carbon steel shell and tube heat exchanger using 1,128 m² area. It is 8.8 m long with 1.37 m shell outer diameter and 1.59 cm tube outer diameter. The cooled vapor leaves for Stage 3A at 2.7 bar and 94°C.

Methanol Side of Integrated Heat Exchange. The inlet methanol, coming from Stage 1 in Stream 103 at 3.6 bar, 25°C, and 188,884 kg/h, first enters E200 to increase the temperature near to that of the boiling point. The methanol, at 85°C then enters E201 to vaporize, leaving at a temperature of 95°C. Finally, the methanol is heated by saturated steam in E202 to further increase the methanol temperature entering the reactor. E202 is a 1.59 cm tube diameter, 0.81 m shell diameter, 4.7 m long, 191 m², carbon steel shell and tube heat exchanger. The methanol vapor leaves E202 at 158°C and 2.5 bar and enters Stage 2B.

Stage 2B (Reaction of methanol to hydrocarbons)

The purpose of Stage 2B in our process is to convert our preheated methanol from Stage 2A into the products, ethylene and propylene, using the SAPO-34 catalyst. The fluidized bed method is required for efficient methanol to olefins technology because fluidized beds allow for superior heat transfer and continuous regeneration of the catalyst while also maintaining good catalyst-reactant contact. This stage also regenerates the catalyst in the catalyst regenerator, provides cooling for the highly exothermic regenerator, and quenches our initial products to prevent degradation, eventually returning the hot products to Stage 2A to perform heat exchange with the colder methanol, as shown in **Figure 8**.



Reactor. Stream 103, containing 188,884 kg/h methanol vapor pressurized to 2.5 bar during Stage 1A, flows into R200, the refractory-lined carbon steel reactor, at 151°C. The vapor enters the 9.5 m diameter reactor at a superficial velocity of 1 m/s to fluidize 97,037 kg SAPO-34 in the fast fluidization regime, which allows for good mass transfer while also using a smaller reactor than other fluidization regimes such as the bubbling regime [14]. The dense phase of this zone, where the majority of the fluidized catalyst is located, is approximately 7 m long. During the dense phase, the methanol reacts with the aid of the catalyst to form products including hydrocarbons, carbon dioxide, and water in an exothermic reaction raising the temperature to 470°C. The reactor narrows in the transition zone, increasing the superficial velocity of the fluid to 3 m/s. In the disengaging zone, 14.5 m in diameter, catalyst particles begin to settle. Two-stage cyclones in the disengaging zone separate the resulting products from the catalyst such that only trace catalyst exists in Stream 206, which contains 180,935 kg/h products. The major products include 30,637 kg/h propylene, 30,637 kg/h ethylene, and 99,609 kg/h water, and Stream 206 exits the reactor after a pressure drop of 0.6 bar and is sent to Stage 2A for quenching and additional heat exchange.

Regenerator. During the reaction, the catalyst also produces coke, a solid, mostly-carbon material, that hinders the reaction by covering active sites on the catalyst. Thus, the spent catalyst containing coke, controlled via valve V200, is circulated to R201, the refractory-lined carbon steel regenerator, via Stream 201. The stream contains 97,037 kg/h catalyst and 8,438 kg/h coke at 470°C. In the regenerator, this catalyst is fluidized by the combustion air of Stream 202. The air, flowing at 111,530 kg/h, is pressurized by blower BL200, a carbon steel axial flow blower with 4263 kW, to 2.7 bar and contains 2% excess oxygen. The air fluidizes the catalyst in the fast fluidization regime with a velocity of 0.84 m/s in the initially 4.66 m diameter regenerator. Similarly to the reactor, the disengaging zone widens to a diameter of 5.96 m, where the catalyst settles and is further separated from the flue gas by two stages of cyclones. Stream 203, the flue gas, exits the reactor at a temperature of 730°C with a mass flow rate of 119,968 kg/h and is sent to Stage 2A for heat recovery through steam generation. Stream 205, containing pure catalyst at 730°C, is sent to R200.

Catalyst Cooler. During the combustion process, a significant amount of heat is generated, necessitating the use of a catalyst cooler, which provides 63 MW of cooling. In this process, Stream 204, containing 473,851 kg/h catalyst fluidized by additional air, is circulated through E205, cooling from 730°C to 260°C. E205 is a carbon steel, vertically oriented shell and tube heat exchanger (1,044 m² area with height 8.7 m, shell diameter 1.32 m, and tube diameter 1.59 cm), cooled by 134,185 kg/h high pressure condensate returned as high pressure saturated steam at 254°C.

Stage 3A (Removal of Water and Methanol)

The purpose of Stage 3A is to continue cooling the reactor effluent and remove water and any remaining unreacted methanol from the effluent stream. A quench tower removes water by flashing the stream, followed by an absorption tower that removes the unreacted methanol. **Figure 9** shows the process diagram for the removal of methanol and water from the hydrocarbons.

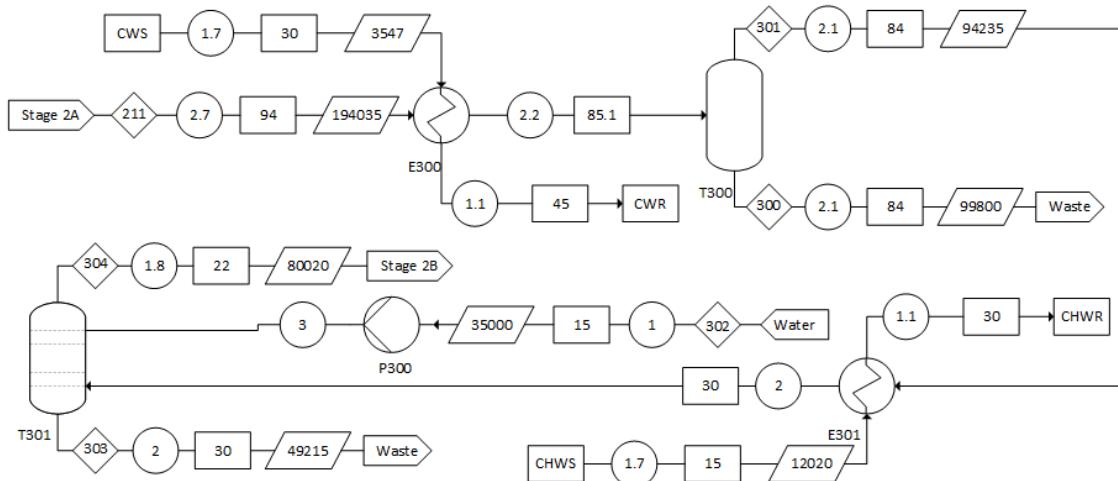


Figure 9. Stage 3A process diagram for the removal of water and methanol

Flash Quench Tower. Stream 211 exits Stage 2A and enters Stage 3A at a pressure of 2.7 bar, a temperature of 94°C, and a total flow rate of 194,035 kg/h. To cool Stream 211 it enters E300, a carbon steel shell and tube heat exchanger with an area of 134 m², and containing 1,096 tubes, each with an outer diameter of 1.59 cm. E300 has a shell outer diameter of 0.76 m and a length 4.1 m. Stream 211 is cooled to 85.1°C and is sent to T300, a carbon steel quench tower with a length of 12.14 m, a diameter of 4.05 m, and a wall thickness of 9 mm, where it is flashed to 2.1 bar to separate the liquid and vapor phases of Stream 211. The liquid waste stream of mostly water leaves as Stream 300 at 2.1 bar, 84°C, and a total flow rate of 99,800 kg/h, while the vapor stream leaves as Stream 301 at 2.1 bar, 84°C, and a total flow rate of 94,235 kg/h.

Methanol Removal Absorption Tower. Stream 301 is sent to be cooled further by E301, another carbon steel shell and tube heat exchanger with an area of 2,644 m², a shell outer diameter of 2.08 m, and a length of 10.1 m. E301 is equipped with 8,678 tubes, each with an outer diameter of 1.59 cm. Stream 301 is cooled to 30°C and is sent to T301 to remove the unreacted methanol in the stream via absorption. T301 is an absorption tower made of carbon steel that is 24 m tall, with an inner diameter of 0.77 m, and a wall thickness of 24 mm. Inside T301, there is 1.83 m of liquid holdup space, 1.22 m of vapor disengagement space, and a total of 29 trays, spaced 0.61 m apart. The trays are made of carbon steel, have a diameter of 0.77 m, and an effective area of 0.38 m². A stream of 35,000 kg/h water at 15°C, Stream 302, is pumped to 3 bar by P300, a 3.73 kW 304 stainless steel centrifugal pump, and sent to T301 to absorb the unreacted methanol. The liquid waste stream comes out as Stream 303 at 2 bar, 30°C, and a total flow rate of 49,215 kg/h, while the vapor stream comes out as Stream 304 at 1.8 bar, 22°C, and a total flow rate of 80,020 kg/h and enters Stage 3B.

Stage 3B (Removal of Carbon Dioxide and Water Vapor)

The purpose of Stage 3B is to remove carbon dioxide and water vapor from the vapor stream before the stream is prepared for the distillation column train. A caustic scrubber is utilized to scrub the carbon dioxide out of the stream, followed by a temperature swing adsorption bed (TSA bed) to remove any

remaining water vapor left in the stream before being prepped for distillation in Stage 4, as shown in **Figure 10**.

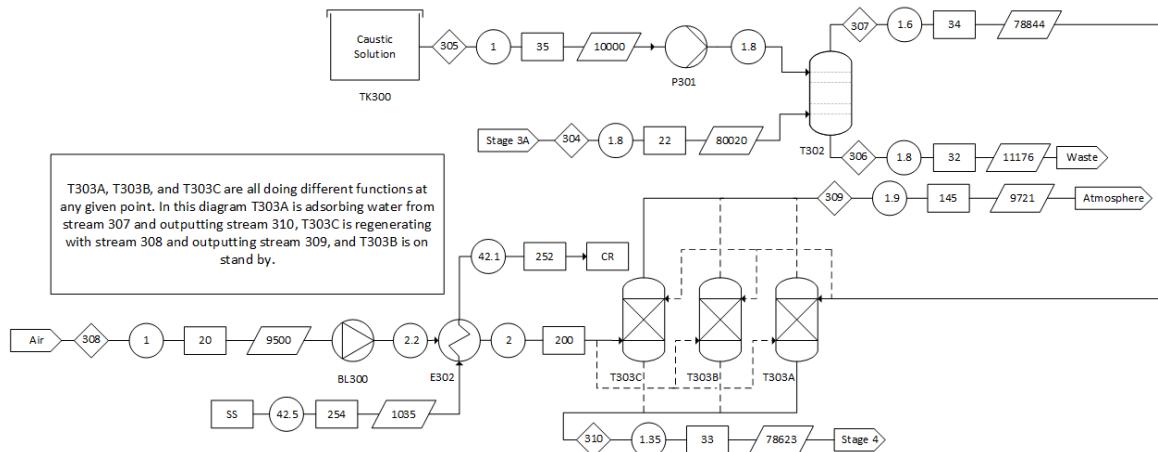


Figure 10. Stage 3B process diagram for the removal of carbon dioxide and water vapor

Caustic Scrubber. Stream 304 exits Stage 3A and enters Stage 3B at a pressure of 1.8 bar, a temperature of 22°C, and a total flow rate of 80,020 kg/h. Stream 304 enters T302, a caustic scrubber made of 304 stainless steel and has a height of 24 m, with an inner diameter of 0.73 m, and a wall thickness of 15 mm. Inside T302, there is 1.83 m of liquid holdup space, 1.22 m of vapor disengagement space, and a total of 29 trays, spaced 0.61 m apart. The trays have a diameter of 0.73 m and an effective area of 0.33 m². Carbon dioxide is scrubbed from the stream by a 50% sodium hydroxide aqueous solution that is stored in TK300 at 1 bar and 35°C. TK300 is a storage tank made of 304 stainless steel, has a height of 2 m, a diameter of 20 m, and a wall thickness of 4 mm. Stream 305 is pumped into T302 at 1.8 bar by a 1.12 kW 304 stainless steel centrifugal pump, P301. The liquid waste stream comes out as stream 306 at 1.8 bar, 32°C, and a total flow rate of 11,176 kg/h, while the vapor stream leaves as Stream 307 at 1.6 bar, 34°C, and a total flow rate of 78,844 kg/h and is sent to T303A-C for drying.

Temperature Swing Adsorption Beds. Stream 307 still has water vapor present in the stream that can cause issues when cryogenic temperatures are reached in distillation. To avoid this issue a TSA bed is used to remove the remaining water vapor through adsorption using silica gel as sorbent. T303A-C are all carbon steel TSA beds with a height of 34 m, a diameter of 1.8 m, a wall thickness of 12 mm, and 18.67 m of silica gel sorbent. Silica gel has a max capacity of water it can adsorb which means that it will need to be regenerated once this capacity is reached. To accommodate this, three TSA beds will be used, each performing a different function at any given time. In **Figure 10**, T303C is being regenerated for 4 hours using 9,500 kg/h air in Stream 308 entering at 1 bar and 20°C, pressurized to 2.2 bar by BL300, a carbon steel axial flow blower with 287 kW. It is then sent to E302, a carbon steel shell and tube heat exchanger with an area of 26 m², 212 tubes each with an outer diameter of 1.59 cm, a shell diameter of 0.36 m, and a length of 3.3 m, which heats the air to 200°C, the regeneration temperature of silica gel. The stream enters T303C, vaporizes the water in the sorbent, and leaves as Stream 309 at 1.9 bar, 145°C, and a total flow rate of 9,721 kg/h. T303B is cooling or on stand-by, ready to begin adsorbing once T303A has reached capacity. Until it reaches capacity after 12 hours, T303A will perform adsorption on Stream 307 and have a vapor stream leave as Stream 310 at 1.35 bar, 33°C, and a total flow rate of 78,623 kg/h and enter Stage 4.

Stage 4 (Distillation Train Preparation)

The purpose of Stage 4 is to prepare the vapor stream for the distillation train that begins in Stage 5A. Multiple compressors and heat exchangers will be used to pressurize and cool the vapor olefins to condense them into a liquid phase for distillation, as can be seen in **Figure 11**.

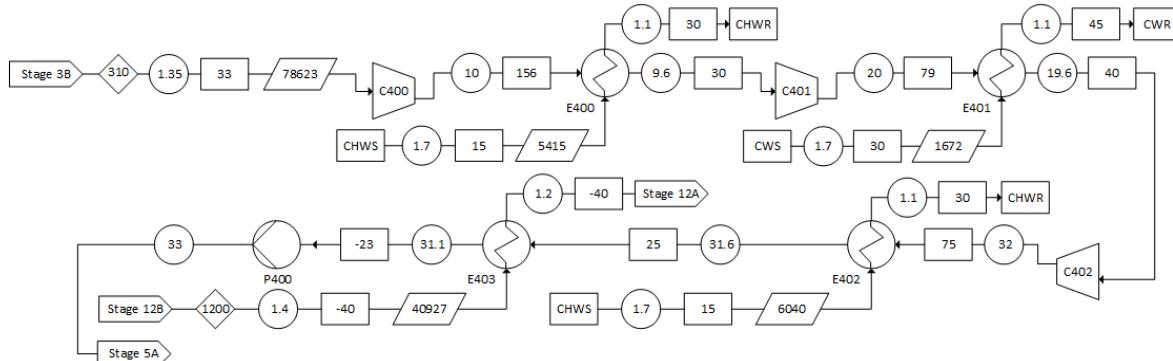


Figure 11. Stage 4 process diagram for distillation train preparation

Stream 310 exits Stage 3B and enters Stage 4 at a pressure of 1.35 bar, a temperature of 33°C, and a total flow rate of 78,623 kg/h. Due to the similarities in structure and properties of all of the hydrocarbons entering the separation train and the scale at which the process operates, distillation was chosen as the best separation technique for these compounds. Because the boiling point for the light olefins, like ethylene and propylene, is normally extremely low at atmospheric pressure, by increasing the pressure and decreasing the temperature of the vapor stream, the boiling point of all the olefins will increase to more physically realizable temperatures. Stream 310 is first pressurized to 10 bar by C400, a 4.75 MW carbon steel centrifugal compressor, and is accompanied by an increase in temperature to 156°C. The stream is then cooled to 30°C by E400, a carbon steel shell and tube heat exchanger with an area of 669 m², 2,195 tubes each with an outer diameter of 1.59 cm, a shell diameter of 1.07 m, and a length of 8.2 m. The stream is pressurized again to 20 bar by C401, a 1.52 MW carbon steel centrifugal compressor, which also increases the temperature to 79°C. The stream is cooled again to 40°C by E401, a carbon steel shell and tube heat exchanger with an area of 304 m², 1,994 tubes each with an outer diameter of 1.59 cm, a shell diameter of 1.02 m, and a length of 5.1 m. The stream is pressurized a third time to 32 bar by C402, a 0.91 MW carbon steel centrifugal compressor, which again increases the temperature to 75°C. The pressure is now sufficiently high enough for the stream to begin condensing so it is sent to E402, a carbon steel shell and tube heat exchanger with an area of 2,146 m², 7,042 tubes each with an outer diameter of 1.59 cm, a shell diameter of 1.88 m, and a length of 9.7 m, where it is cooled to 25°C and partially condenses. To fully condense the vapor, the stream is sent to E403, a 304 stainless steel shell and tube heat exchanger with an area of 275 m², 1,802 tubes each with an outer diameter of 1.59 cm, a shell diameter of 0.97 m, and a length of 5.1 m, where it is cooled to -23°C. To achieve the cryogenic temperature, Stream 1200, which comes from Stage 12B, has a flow rate of 40,927 kg/h of pure propylene and is supplied to the tubes of E403 at 1.4 bar and -40°C. Stream 1200 partially vaporizes while moving through E403 and returns to Stage 12A at 1.2 bar and -40°C. Stream 310 is pumped to 33 bar by P400, a 29.9 kW 304 stainless steel centrifugal pump, and sent to Stage 5A to begin distillation. D

Stage 5A (De-Methanizer)

The purpose of Stage 5A is to remove methane from the rest of the reactor products and byproducts. Due to the very low boiling point of methane, we needed to design the de-methanizer to operate under cryogenic conditions. To achieve these cryogenic temperatures, we designed a refrigeration system using pure ethylene as the refrigerant. This system is described in Stages 10A and 10B. The de-methanizer removes methane from Stream 312 as distillate, and the bottoms are sent to the de-ethanizer, Stage 5B. A diagram of the de-methanizer is shown in **Figure 12**.

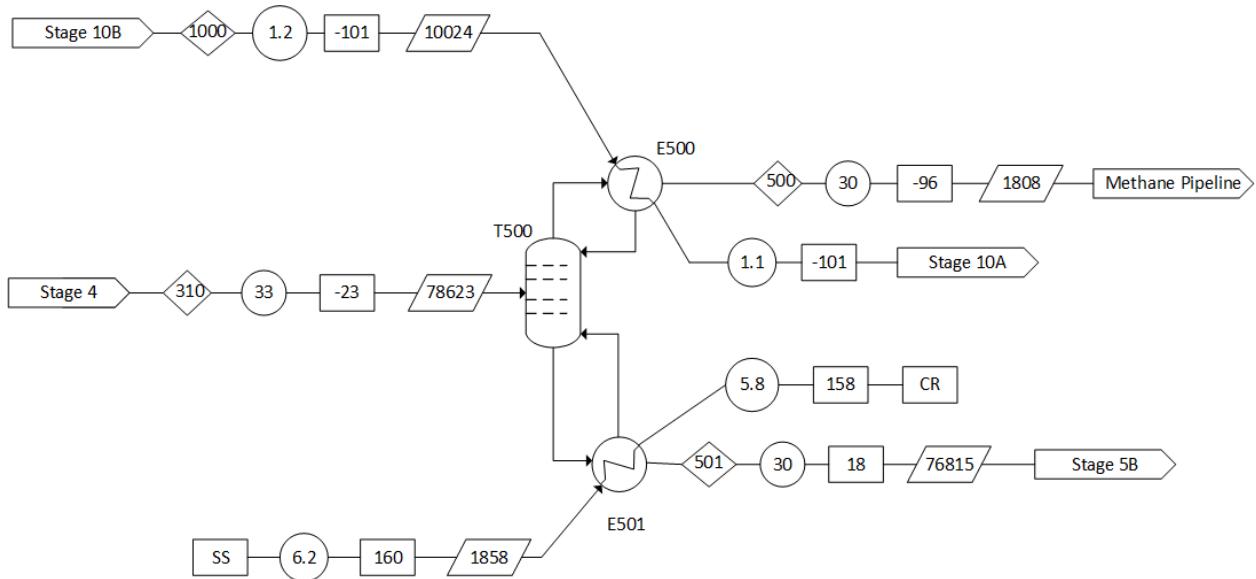


Figure 12. Stage 5A process diagram for the separation of methane from the other hydrocarbon products

Stream 310 exits Stage 4 and enters Stage 5A at a pressure of 33 bar, a temperature of -23°C, and a total flow rate of 78,623 kg/h. T500, the de-methanizer, separates methane from the olefin products (ethylene, propylene, and butylene) and the other byproducts (ethane, propane, and C5+ hydrocarbons). It is constructed from 304 stainless steel, is 41.05 m tall, has an inner diameter of 1.90 m, and a wall thickness of 62 mm. T500 operates with a reflux ratio of 8 and a distillate-to-feed ratio of 0.023. Inside T500, there is 1.83 m of liquid holdup space, 1.22 m of vapor disengagement space, and a total of 68 trays, spaced 0.56 m apart. The trays have a diameter of 1.90 m and an effective area of 2.26 m².

The vapor stream leaving T500's tray 1 has a flow rate of 16275 kg/h and is sent to the shell of E500, a 1-1 pass shell-and-tube 304 stainless steel partial condenser with 377 m² of available heat exchange area. E500 has a shell OD of 1.02 m, and it contains 1475 tubes, each with an OD of 1.91 cm. E500 is 6.07 m long.

In the shell-side of E500, 14467 kg/h of material is condensed and returned to the column at tray 1. The remaining un-condensed vapor leaves E500 as stream 500 with a flow rate of 1808 kg/h of 98% pure methane and is sent to a methane pipeline at a pressure of 30 bar and a temperature of -96°C. The balance of Stream 500 consists of 1.8% ethylene and trace amounts of ethane. Stream 1000, which

comes from Stage 10B, has a flow rate of 10024 kg/h of pure ethylene and is supplied to the tubes of E500 at 1.2 bar and -101°C. Stream 1000 partially vaporizes while moving through E500 and returns to Stage 10A at a pressure of 1.1 bar and a temperature of -101°C.

The liquid stream leaving T500's tray 68 has a flow rate of 117132 kg/h and is sent to the shell of E501, a 1-1 pass shell-and-tube 304 stainless steel kettle reboiler with 56 m² of available heat exchange area. E501 has a shell OD of 0.46 m, and it contains 367 tubes, each with an OD of 1.56 cm. E501 is 4.04 m long.

In the shell-side of E501, 40317 kg/h of material is vaporized and returned to the column at tray 70. The remaining un-vaporized liquid leaves E501 as Stream 501 with a flow rate of 76815 kg/h and consists of the olefin products and remaining byproducts. It is sent to Stage 5B at a pressure of 30 bar and a temperature of 18°C for additional purification. Saturated steam at a flow rate of 1858 kg/h, a pressure of 6.2 bar, and a temperature of 160°C is supplied to the tubes of E501 where it condenses inside the tubes and returns as steam condensate at a pressure of 5.8 bar and a temperature of 158°C.

Stage 5B (De-Ethanizer)

The purpose of Stage 5B is to remove ethylene and ethane from the rest of the hydrocarbon products. Similar to the de-methanizer, we needed to design the de-ethanizer to operate under cryogenic conditions to effectively isolate ethylene and ethane, the two lightest components at this point in the process. To achieve these cryogenic temperatures, we designed a refrigeration system using pure propylene as the refrigerant. This system is described in Stages 11A and 11B. The de-ethanizer removes ethylene and ethane from Stream 501 as distillate, and the bottoms are sent to the de-propanizer, Stage 7A. **Figure 13** displays the de-ethanizer process.

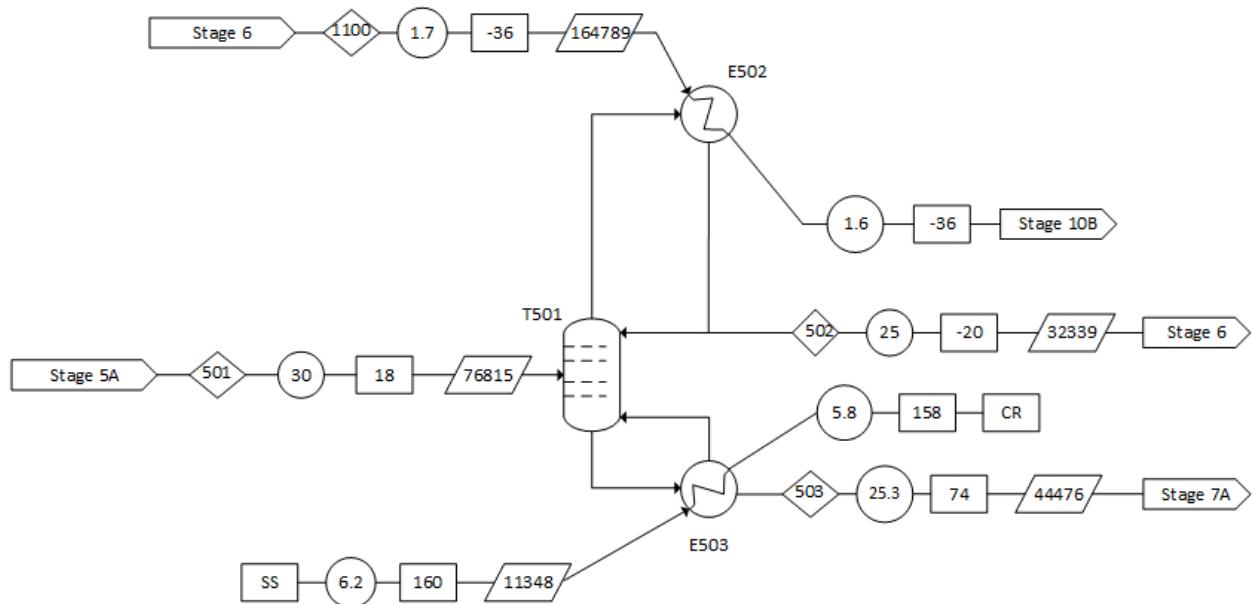


Figure 13. Stage 5B process diagram for the separation of ethylene and ethane from other hydrocarbon products

Stream 501 exits Stage 5A and enters Stage 5B at a pressure of 30 bar, a temperature of 18°C, and a total flow rate of 76,815 kg/h. T501, the de-ethanizer, separates ethane and ethylene from the other olefin products (propylene, and butylene) and the remaining byproducts (propane and C5+ hydrocarbons). It is constructed from 304 stainless steel, is 24.28 m tall, has an inner diameter of 3.01 m, and a wall thickness of 81 mm. T501 operates with a reflux ratio of 1.5 and a distillate-to-feed ratio of 0.421. Inside T501, there is 1.83 m of liquid holdup space, 1.22 m of vapor disengagement space, and a total of 38 trays, spaced 0.56 m apart. The trays have a diameter of 3.01 m and an effective area of 5.69 m².

The vapor stream leaving T501's tray 1 has a flow rate of 80848 kg/h and is sent to the shell of E502, a 1-1 pass shell-and-tube 304 stainless steel total condenser with 811 m² of available heat exchange area. E502 has a shell OD of 1.17 m, and it contains 6107 tubes, each with an OD of 1.59 cm. E502 is 8.44 m long.

In the shell-side of E502, all entering material is condensed, and 48509 kg/h is returned to the column at tray 1. The remaining liquid forms Stream 502, which has a flow rate of 32339 kg/h of 94.5% pure ethylene and 5.5% pure ethane and is sent to Stage 6 at a pressure of 25 bar and a temperature of -20°C. Stream 1100, which comes from Stage 6, has a flow rate of 164789 kg/h of pure propylene and is supplied to the tubes of E502 at 1.7 bar and -36°C. Stream 1100 partially vaporizes while moving through E502 and moves to Stage 10B at a pressure of 1.6 bar and a temperature of -36°C.

The liquid stream leaving T501's tray 38 has a flow rate of 149465 kg/h and is sent to the shell of E503, a 1-1 pass shell-and-tube carbon steel kettle reboiler with 187 m² of available heat exchange area. E503 has a shell OD of 0.71 m, and it contains 384 tubes, each with an OD of 2.54 cm. E503 is 7.90 m long.

In the shell-side of E503, 104990 kg/h of material is vaporized and returned to the column at tray 38. The remaining un-vaporized liquid leaves E503 as Stream 503 with a flow rate of 44476 kg/h and consists of the remaining olefin products (propylene and butylene) and remaining byproducts (propane and C5+ hydrocarbons). It is sent to Stage 7A at a pressure of 25.3 bar and a temperature of 74°C for additional purification. Saturated steam at a flow rate of 11348 kg/h, a pressure of 6.2 bar, and a temperature of 160°C is supplied to the tubes of E503 where it condenses inside the tubes and returns as steam condensate at a pressure of 5.8 bar and a temperature of 158°C.

Stage 6 (C2 Splitter)

Stage 6's purpose is to purify ethylene to polymer-grade purity of 99.9% by removing ethane and trace amounts of heavier hydrocarbons from the distillate of the de-ethanizer outlined in Stage 5B. A diagram of the process is shown in **Figure 14**.

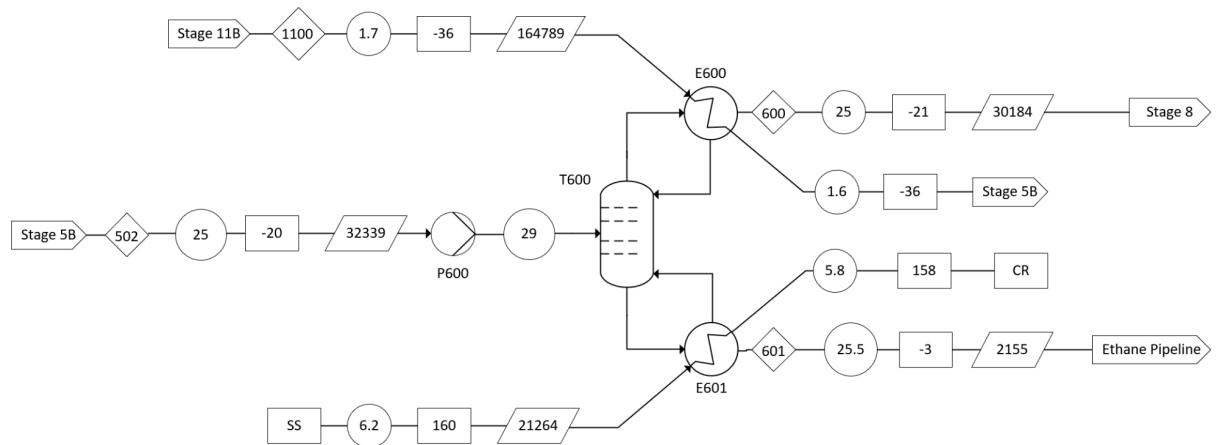


Figure 14. Stage 6 process diagram for the separation of ethylene from ethane

Stream 502 is sent to Stage 6 as the distillate product of the de-ethanizer T501 and contains 30,573 kg/h of ethylene and 1,766 kg/h of ethane, pressurized to 25 bar, at a temperature of -20°C. Stream 502 is fed through P600, a 304 stainless steel centrifugal pump, operating with a power input of 37 kW. P600 pressurizes Stream 502 to 29 bar to overcome the pressure drop arising from pumping Stream 502 10 m vertically to reach the feed tray of T600. T600 is referred to as the C2 splitter as it separates the C2 hydrocarbons produced in the MtO process, ethylene and ethane, from each other, taking ethylene as the light key and ethane as the heavy key. It is constructed from 304 stainless steel, and is 41.05 m tall, has an inner diameter of 4.50 m, and a wall thickness of 121 mm. T600 operates with a condenser pressure of 25 bar, a reflux ratio of 4.05, and a distillate-to-feed ratio of 0.945. The internals of the column are comprised of 1.83 m of liquid holdup, 1.22 m of vapor disengagement, and 68 trays, spaced 0.56 m apart. The trays have a diameter of 4.50 m, and an effective area of 12.72 m².

The vapor leaves tray 1 of T600 at a flow rate of 30184 kg/h and is sent to the shell of partial condenser E600, a 1-1 pass, shell-and-tube 304 stainless steel heat exchanger, with an effective area of 1,306 m². E600 has a shell OD of 1.47 m, and contains 4284 tubes , each with an OD of 1.59 cm. E600 is 9 m long.

Heat is removed from this vapor stream by running 164,789 kg/h of 100% pure ethylene at -36°C from Stage 11 through the tube side of the exchanger. This refrigerant exits E600 at 1.6 bar and -36°C, and is sent to Stage 5 for further heat removal. 30,184 kg/h of 99.9% pure uncondensed ethylene vapor exists E600 at -21°C and 25 bar, and is sent to Stage 8 for removal of excess heat from the C5+ product stream, while the remainder of the shell fluid is condensed and returned to tray 1 of T600 at 122,247 kg/h.

The liquid stream leaving tray 68 of T600 at a flow rate of 154,027 kg/h is sent to the shell of E601, a 1-1 pass shell-and-tube 304 stainless steel heat exchanger, with an effective area of 174 m². E601 has a shell OD of 0.86 m, and contains 1424 tubes, each of which has an OD of 1.59 cm. E601 is 4.24 m long.

The tube fluid entering E601 is 21,264 kg/h of low pressure steam, supplied at 6.2 bar and 160°C. This saturated steam condenses in the tubes of E601, releasing heat and exiting the exchanger at 5.8 bar and

158°C as steam condensate. On the shell side, 2,155 kg/h of 82.8% pure ethane exits E601 as a liquid stream at 25.5 bar and -3°C, and is sent to the ethane pipeline, while the remainder of the fluid fed to E601 is vaporized and sent to tray 68 of T600 at a rate of 151,873 kg/h.

Stage 7A (De-Propanizer)

The role of Stage 7A in the proposed process is to separate C3 hydrocarbons propylene and propane from the residual heavier hydrocarbons present in the bottoms product of the de-ethanizer outlined in Stage 5B. **Figure 15** describes the de-propanizer process discussed below.

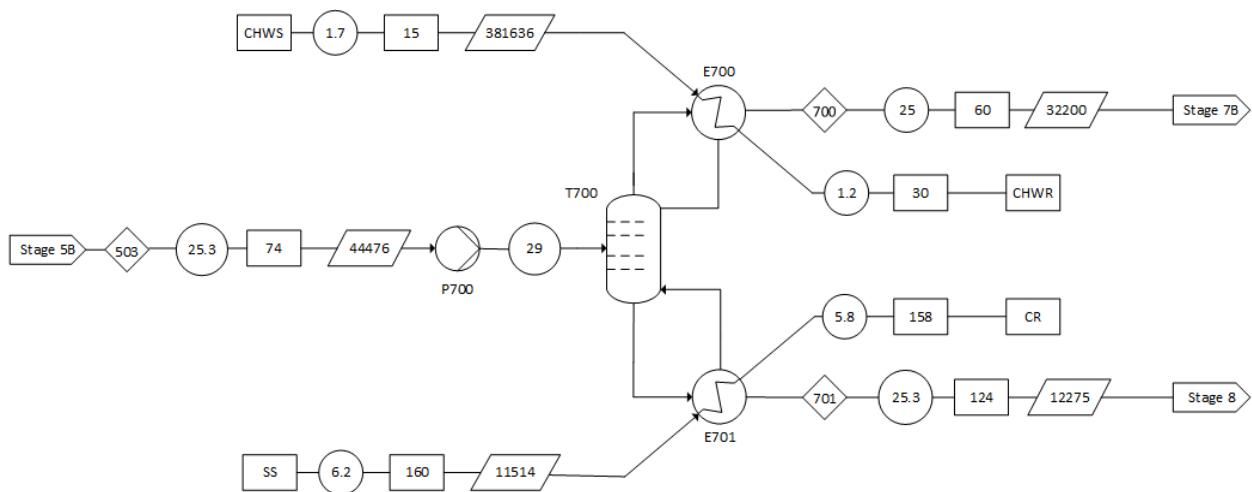


Figure 15. Stage 7A process diagram for the separation of C3 hydrocarbons from the remaining C3+ hydrocarbons

Stream 503 is sent to Stage 7A as the bottoms product of the de-ethanizer T501 from Stage 5B at 25.3 bar, 74°C, and at a flow rate of 44,476 kg/h. This process stream is pressurized to 29 bar by P700, a carbon steel centrifugal pump operating with 18.66 kW of power input to overcome the pressure loss associated with pumping Stream 503 13 m vertically to reach the feed tray of distillation column T700. T700, referred to as the de-propanizer, separates the olefin product propylene and C3 byproduct propane from the heavier butylene and C5+ hydrocarbons comprising Stream 503. It is constructed from carbon steel, is 29.87 m tall, has an inner diameter of 3.61 m, and a wall thickness of 98 mm. T700 operates with a condenser pressure of 25 bar, a reflux ratio of 2, and a distillate-to-feed ratio of 0.724. The internals of the column are comprised of 1.83 m of liquid holdup, 1.22 m of vapor disengagement, and 48 trays, spaced 0.56 m apart. The trays have a diameter of 3.61 m and an effective area of 8.19 m².

The vapor stream leaving T700's tray 1 has a flow rate of 96,600 kg/h and is sent to E700, a 1-1 pass shell-and-tube carbon steel partial condenser with 296 m² of available heat exchange area. E700 has a shell OD of 0.91 m, and it contains 1617 tubes, each with an OD of 1.59 cm. E700 is 5.56 m long.

The shell side of E700 condenses 64,400 kg/h of material and returns it to tray 1 of T700. The remainder of the vapor stream leaves E700 as Stream 700, which sends 32,200 kg/h of 94.43% pure propylene

vapor with the balance propane and trace amounts of ethane, butylene, and ethylene at 25 bar and 60°C to Stage 7B for further purification. The tube fluid of E700 is 381,636 kg/h of chilled water, supplied to the process at 1.7 bar and 15°C and returned at 1.2 bar and 30°C.

The liquid stream leaving tray 49 of T700 has a flow rate of 146,898 kg/h, and is passed through the shell of E701, a 1-1 pass shell-and-tube carbon steel kettle reboiler, which has 469 m² of heat transfer area. E701 has a shell OD of 1.07 m, and it contains 2195 tubes, each with an OD of 1.59 cm. E701 is 6.37 m long.

The tube fluid of E701 is 11,514 kg/h of low-pressure steam, supplied to the process at 6.2 bar and 160°C. Within the tubes of E701, this steam condenses, releasing heat to the process fluid, and returning steam condensate at 5.8 bar and 158°C. This heat exchange vaporizes 134,623 kg/h of material on the shell side, which is returned to tray 49 of T700, while the remaining 12,275 kg/h of the liquid process stream exits the exchanger at 25.3 bar and 124°C, and is sent to Stage 8 for subsequent purification.

Stage 7B (C3 Splitter)

Stage 7B is fed the distillate product of the de-propanizer detailed in Stage 7A, and purifies propylene to polymer-grade purity of 99.5% while simultaneously generating propane as a saleable byproduct, as shown in **Figure 16**.

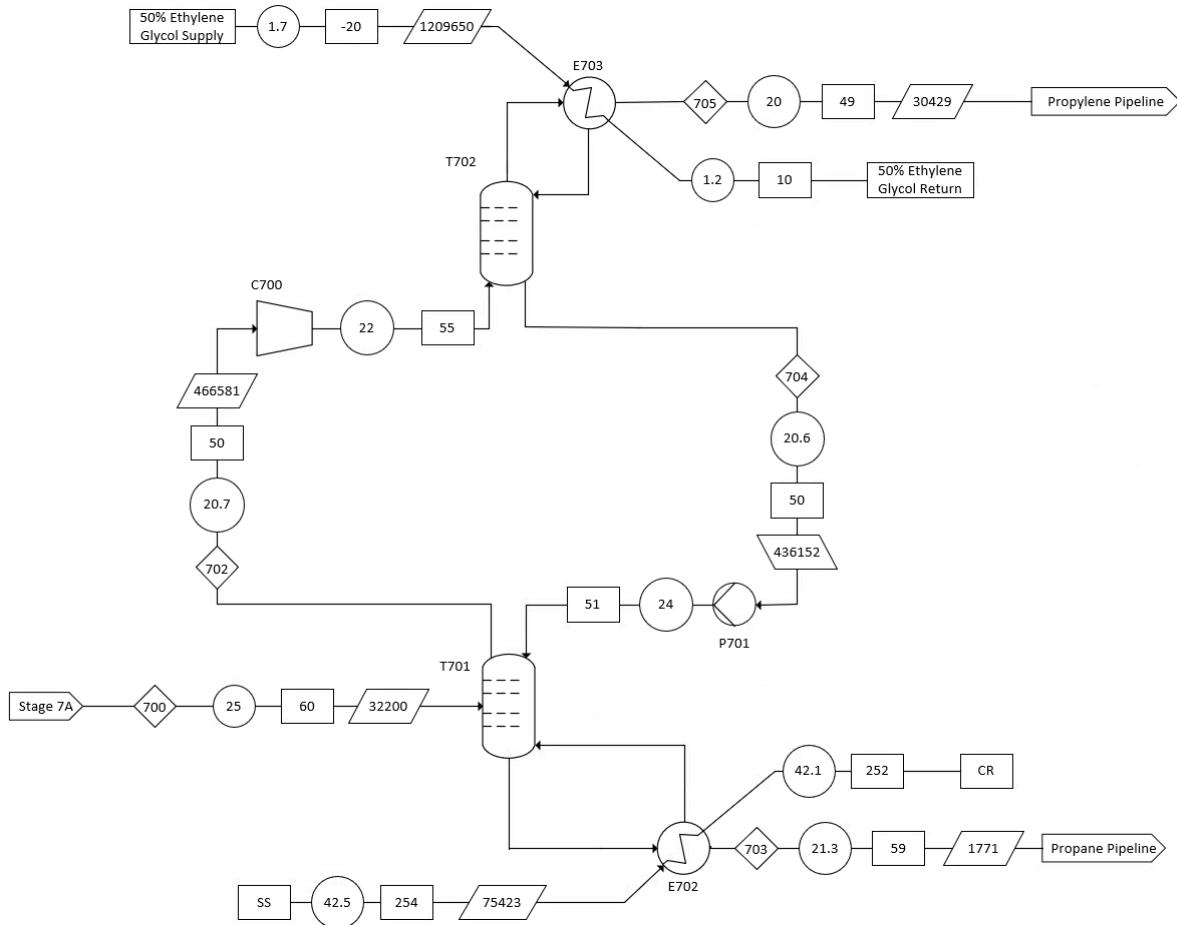


Figure 16. Stage 7B process diagram for the separation of propylene from propane

C3 Splitter Stripping Section. Stream 700, the distillate of the de-propanizer described in Stage 7A, is fed at 25 bar and 60°C to tray 20 of the T701, which is referred to as the C3 splitter stripping column. Necessitated by maximum column height, the split between propylene and propane must be split into two equally sized sections, allowing for more equilibrium stages than would be possible with a single column. The C3 splitter stripping column is constructed from carbon steel, and is 52.22 m tall, has an inner diameter of 8.84 m, and has a wall thickness of 198 mm. T701 is equipped with a kettle reboiler and no condenser, and operates with a constant bottoms rate of 1771 kg/h. The internals of the T701 are comprised of 1.83 m of liquid holdup, 1.22 m of vapor disengagement, and 89 trays, spaced 0.56 m apart. The trays have a diameter of 8.84 m, and an effective area of 49.1 m². The purpose and design of T702, known as the C3 rectifying column, will be detailed in the subsequent section.

The liquid stream leaving tray 89 of the C3 stripping column has a flow rate of 497,259 kg/h, and is sent to E702, a carbon steel kettle reboiler, designed with 426 m² of available heat transfer area. E702 has a shell OD of 1.02 m, and it contains 1994 tubes, each with an OD of 1.59 cm. E702 is 6.34 m long. The tube fluid of E702 is 75,423 kg/h of high pressure steam, supplied to the process at 42.5 bar and 254°C. This steam condenses, and is returned as steam condensate at 42.1 bar and 252°C. On the shell side, 495,488 kg/h of the material fed is vaporized and returned to tray 89 of T701, while the remaining

1,771 kg/h of material exits as Stream 703. Stream 703 leaves the process as 93.1% pure propane at 21.3 bar and 59°C, and is sent to the propane pipeline to be sold for its energy content.

Because the separation of propylene from propane is necessarily split into two disjointed distillation columns operating as one, vapor must be sent from tray 1 of the C3 stripping column to tray 89 of the C3 rectifying column, and liquid must be returned from tray 89 of the C3 rectifying column to tray 1 of the C3 stripping column.

Internal vapor Stream 702, containing 466,581 kg/h 95.45% pure propylene vapor pressurized to 20.7 bar and at a temperature of 50°C leaves tray 1 of T701 and is compressed to 22 bar by C700. This ensures a sufficient differential pressure between tray 1 of T701 and tray 89 of T702, and the pressurized stream is fed to T702. C700 is a carbon steel centrifugal compressor, with a power requirement of 502 kW.

Internal liquid Stream 704, containing 436,152 kg/h of 94.39% pure propylene with the balance propane exists tray 89 of T702 at 20.6 bar and 50°C. Stream 704 is then pressurized to 24 bar by P701 in order to overcome pressure losses associated with pumping the stream to tray 1 of T701, which is where the stream is fed to. P701 is a carbon steel centrifugal pump, operating with 224 kW of power input.

C3 Splitter Rectifying Section. As previously described, the split between propylene and propane must be split into two sections to avoid an unfeasibly tall column design. T702's purpose is to take the vapor stream coming from tray 1 of T701, and purify it further to the specification of 99.5% pure propylene. T702 is equipped with a partial condenser and no reboiler, and operates with a distillate rate of 30,429 kg/h. T702 is constructed from carbon steel, is 52.22 m tall, has an inner diameter of 7.7 m, and a wall thickness of 167 mm. Inside T702 is 1.83 m of liquid holdup, 1.22 m of vapor disengagement, and 89 trays, which are spaced 0.56 m apart, have a diameter of 7.7 m, and an effective area of 37.29 m².

The 99.5% pure propylene vapor exiting tray 1 of T702 is fed to the shell side of E703 at a flow rate of 57,672 kg/h. E703 is a carbon steel partial condenser, equipped with 11,680 m² of heat exchange area. E703 has a shell OD of 1.35 m, and it contains 3561 tubes, each with an OD of 1.59 cm. E703 is 8.77 m long.

E703 condenses 27,251 kg/h of material on the shell side, giving rise to product Stream 705, which exits E703 at 20 bar and 49°C at a flow rate of 30,429 kg/h and is sent to the ethylene pipeline. The tube fluid servicing E703 is 1,209,650 kg/h of 50% ethylene glycol, supplied to the process at a pressure of 1.7 bar and a temperature of -20°C. Heat is transferred to the process fluid in the shell, and the ethylene glycol is returned at 1.2 bar and 10°C.

Stage 8 (C4/C5 Splitter)

The purpose of Stage 8 is to act as the final separation step of the proposed MtO process by separating butylene at a saleable purity of >95% from the heavier C5+ hydrocarbons remaining in the process stream, and the process diagram is shown in **Figure 17**.

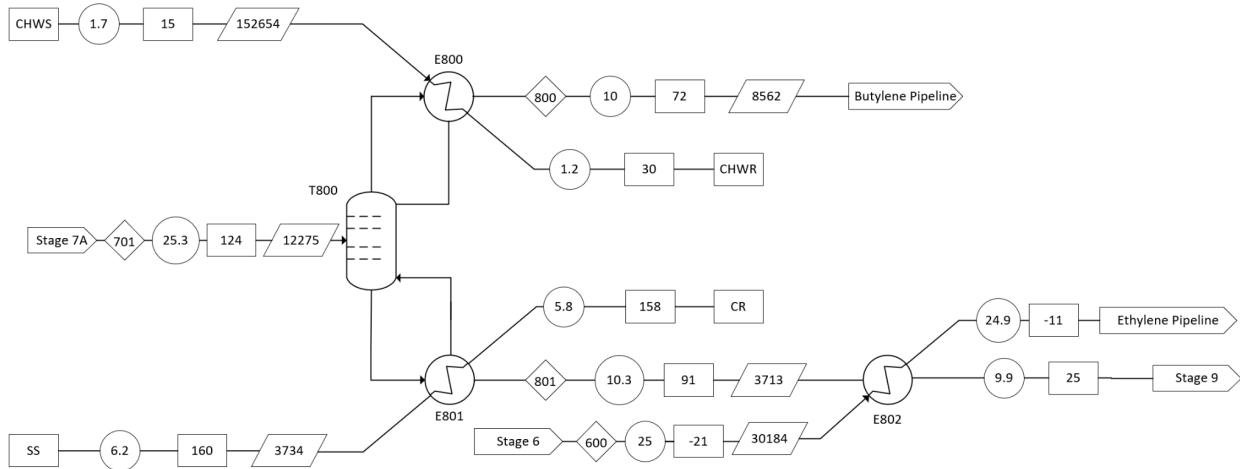


Figure 17. Stage 8 process diagram for the separation of butylene from C5+ hydrocarbons

Stream 701 is sent to Stage 8 as the bottoms product of the de-propanizer T700 outlined in Stage 7A and is composed of 71.64% butylene and 28.21% C5+ hydrocarbons, with the balance being smaller hydrocarbons remaining from the upstream process. Stream 701 enters at 25.3 bar and 124°C at a flow rate of 12275 kg/h, and is fed on tray 22 of T800. T800, also referred to as the C4/C5 splitter is designed and operated to purify the remaining butylene to a purity of 95.72%, where it meets saleable purity specification. T800 is constructed from caron steel, is 29.87 m tall, has an inner diameter of 1.71 m, and a wall thickness of 19 mm. T800 is operated with a reflux ratio of 2.7, and a distillate-to-feed ratio of 0.737. It is equipped with a partial condenser with an outlet pressure of 10 bar, and internally is designed with 1.83 m of liquid holdup, 1.22 m of vapor disengagement, and 48 trays, spaced 0.56 m apart. The trays have a diameter of 1.71 m, and an effective area of 1.83 m².

The vapor stream leaving T800's tray 1 exits at a flow rate of 31,679 kg/h, and is fed to E800, a partial 1-1 pass shell-and-tube condenser designed with 87 m² of heat exchange area and constructed from carbon steel. E800 has a shell OD of 0.56 m, and it contains 575 tubes, each with an OD of 1.59 cm. E800 is 4.23 m long.

E800 condenses 23,117 kg/h of this vapor stream, leaving 8,562 kg/h of this 95.77% pure butylene as vapor, which emerges from E800 at 10 bar and 72°C, and is sent to the butylene pipeline to be sold. E800 is cooled by 152,654 kg/h of chilled water, supplied to the process at 1.7 bar and 15°C. This chilled water flows through the tubes of E600, where it is heated to 30°C and its pressure is reduced to 1.2 bar before being returned.

The liquid stream leaving T800's tray 48 has a flow rate of 34,722 kg/h is sent to E801, which is carbon steel 1-1 pass shell-and-tube kettle reboiler with 74 m² of heat exchange area. E801 has a shell OD of 0.66 m, and it contains 811 tubes, each with an OD of 1.59 cm. E801 is 3.27 m long.

E801 vaporizes 31,009 kg/h of this liquid, allowing for 3,713 kg/h of 83.78% pure C5+ hydrocarbons with the balance butylene to leave the the shell at 10.3 bar and 91°C. E801's tube side is serviced with 3,374 kg/h of low pressure steam, provided at 6.2 bar and 160°C. This steam condenses in the tubes of the exchanger, and is returned at a pressure of 5.8 bar and a temperature of 158°C as steam condensate.

To prepare the C5+ byproduct stream for storage, it is fed to the shell side of E802 and cooled to 25°C. E802 is a 1-1 pass shell-and-tube 304 stainless steel heat exchanger with 11 m² of available heat exchange area. E802 has a shell OD of 0.27 m, and it contains 121 tubes, each with an OD of 1.59 cm. E802 is 2.54 m long.

The ethylene product produced in Stage 6 of the process enters the tube side of E802 as Stream 600 at 25 bar and -21°C. Stream 600 exits E802 at 24.9 bar and -11°C, and is sent to the ethylene pipeline at a flow rate of 30,184 kg/h. Stream 801 exits E802 at 9.9 bar and 25°C, and is sent to Stage 9 at a flow rate of 3,713 kg/h for product storage.

Stage 9 (C5+ Storage and Shipping)

Stage 9 functions to take the heavier C5+ hydrocarbon byproducts remaining in the bottoms product of the C4/C5 splitter outlined in Stage 8 and store them in a tank farm, as well as detail the operational schedule of emptying these tanks and selling the material. **Figure 18** shows the storage and shipping process for C5+ hydrocarbons.

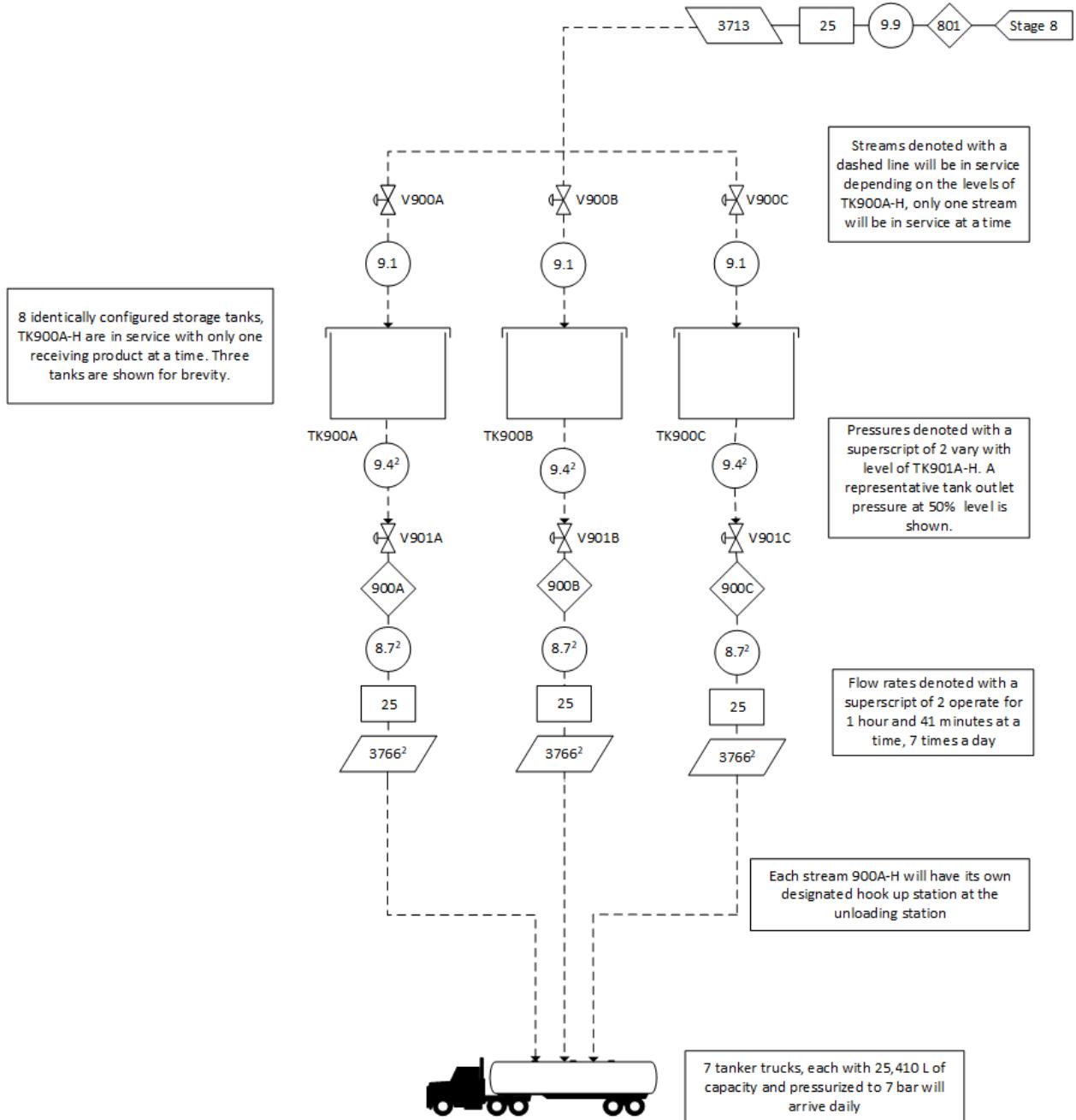


Figure 18: Stage 9 process diagram for the storage of and selling of C5+ hydrocarbons

C5+ Tank Farm. Stream 801, containing 83.78% pure C5+ hydrocarbons is fed to Stage 9 at a rate of 3,713 kg/h at 9.9 bar and 25°C, ultimately to fill TK900A-H, a set of 8 identically configured carbon steel storage tanks. Each tank is pressurized to 9.1 bar, is 9.15 m tall, has a diameter of 3.81 m, and a wall thickness of 29 mm. To fill these tanks, Stream 801 is passed through V900A-H, a set of 8 identically configured control valves that open and close depending on the level of storage tanks TK900A-H, contributing to a 0.7 bar pressure reduction of the product stream.

C5+ Tank Unloading. 7 tanker trucks, each pressurized to 7 bar, will arrive daily with an available capacity of 25,410 L to receive the C5+ product. To fill these trucks, one of the valves in the identical set V901A-H will be opened depending on the level of TK900A-H, and Stream 900 will flow at 3,766 kg/h (251 LPM) for 1 hour and 41 minutes. Outlet pressure of each of these tanks is dependent on tank level, so an outlet pressure at 50% tank level is shown in **Figure 18**. This difference in pressure between the tank outlet and the tanker truck will drive the product stream to the unloading station, where each Stream 900A-H will have its own hookup station. After filling, each tanker truck will leave the site and be sent to various customers.

Stage 10A (Compressing the Ethylene Refrigerant)

The purpose of Stage 10A is to prepare the ethylene refrigerant for flashing that occurs in Stage 10B. Serving as the “start” of the refrigeration loop, the stage begins with a liquid-vapor mixture of ethylene that comes from Stage 5A. The stream is vaporized, combined with the vapor stream produced from flashing, and then compressed before being sent to Stage 10B. This process is shown in **Figure 19**.

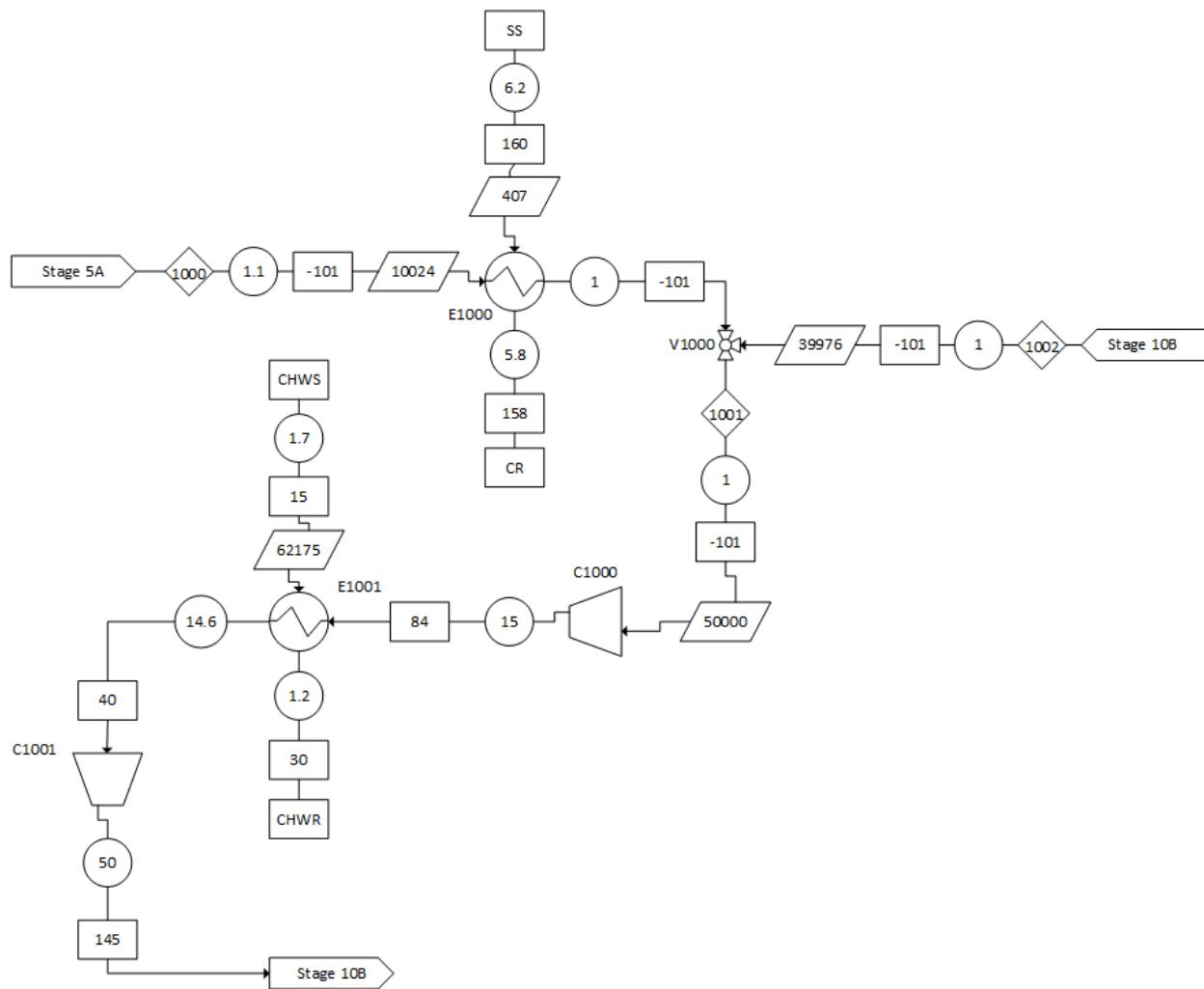


Figure 19. Stage 10A process diagram for the compression of ethylene refrigerant in preparation for flashing

Stream 1000 exits Stage 5A and enters Stage 10A as a vapor-liquid mixture at a pressure of 1.1 bar, a temperature of -101°C, and a total flow rate of 10,024 kg/h and enters the tubes of E1000, a 1-1 pass shell-and-tube 304 stainless steel heat exchanger with 2 m² of available heat exchange area. E1000 has a shell OD of 0.17 m, and it contains 31 tubes, each with an OD of 1.91 cm. E1000 is 1.73 m long. All liquid present in Stream 1000 is vaporized in E1000's tube side and exits E1000 at a pressure of 1 bar and a temperature of -101°C. Saturated steam at a flow rate of 407 kg/h, a pressure of 6.2 bar, and a temperature of 160°C is supplied to the shell of E1000 where it condenses inside the shell and returns as steam condensate at a pressure of 5.8 bar and a temperature of 158°C.

After exiting E1000, Stream 1000 enters V1000, a three-way valve, where it mixes with Stream 1002 from Stage 10B; Stream 1002 has a total flow rate of 39,976 kg/h, a pressure of 1 bar, and a temperature of -101°C. The two streams mix to form Stream 1001, which leaves V1000 at a flow rate of 50,000 kg/h, a pressure of 1 bar, and a temperature of -101°C. Stream 1001 enters C1000, a 3.51 MW 304 stainless steel centrifugal compressor, and exits at a pressure of 15 bar and a temperature of 84°C. It then enters the tubes of E1001, a 1-1 pass shell-and-tube carbon steel heat exchanger with 58 m² of available heat exchange area. E1001 has a shell OD of 0.71 m, and it contains 949 tubes, each with an OD of 1.59 cm. E1001 is 2.68 m long. Stream 1001 is cooled in the tubes of E1001 and exits at a pressure of 14.6 bar and a temperature of 40°C. Chilled water at a flow rate of 62,175 kg/h, a pressure of 1.7 bar, and a temperature of 15°C is supplied to the shell of E1001 where it heats up, leaving at a pressure of 1.2 bar and a temperature of 30°C.

After exiting E1001, Stream 1001 enters C1001, a 2.23 MW carbon steel centrifugal compressor, and exits at a pressure of 50 bar and a temperature of 145°C. It then moves to Stage 10B.

Stage 10B (Flashing the Ethylene Refrigerant)

The purpose of Stage 10B is to finish preparing the ethylene refrigerant for flashing and to finally flash it, dropping its temperature to a cold enough level for the liquid to be used in the condenser of the de-methanizer. The vapor resulting from the flash is then fed back to Stage 10A to be recombined with the liquid stream, closing the loop. **Figure 20** describes the flashing of the ethylene refrigerant.

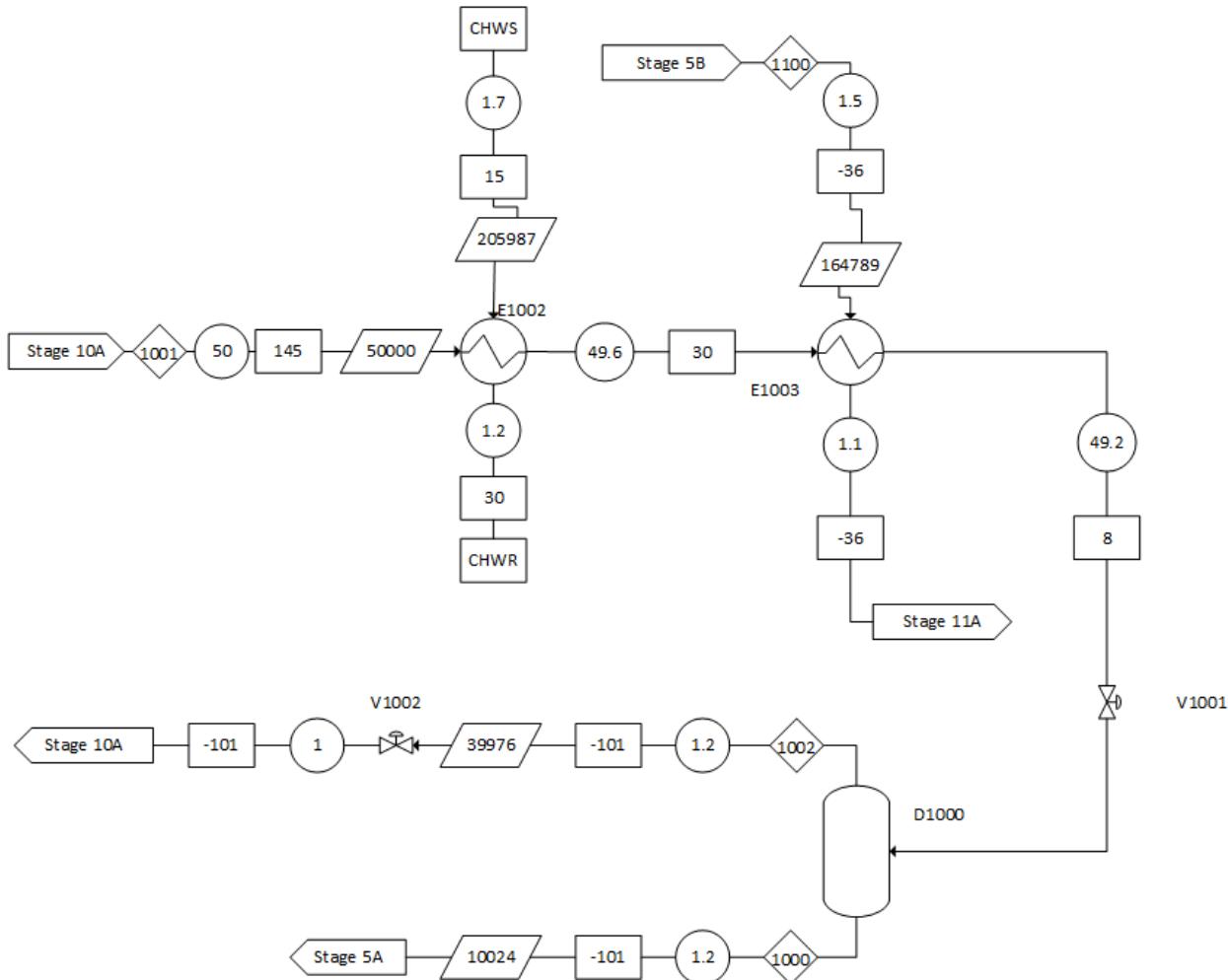


Figure 20. Stage 10B process diagram for the flashing of ethylene refrigerant

Stream 1001 enters Stage 10B at a flow rate of 50,000 kg/h, a pressure of 50 bar, and a temperature of 145°C and enters the tubes of E1002, a 1-1 pass shell-and-tube carbon steel heat exchanger with 146 m² of available heat exchange area. E1002 has a shell OD of 0.97 m, and it contains 1337 tubes, each with an OD of 1.91 cm. E1002 is 3.88 m long. Stream 1001 is cooled in the tubes of E1002 and exits at a pressure of 49.6 bar and a temperature of 30°C. Chilled water at a flow rate of 205,987 kg/h, a pressure of 1.7 bar, and a temperature of 15°C is supplied to the shell of E1002 where it heats up, leaving at a pressure of 1.2 bar, and a temperature of 30°C.

After exiting E1002, Stream 1001 enters the tubes of E1003, a 1-1 pass shell-and-tube 304 stainless steel heat exchanger with 202 m² of available heat exchange area. E1003 has a shell OD of 0.81 m, and it contains 518 tubes, each with an OD of 2.54 cm. E1003 is 6.54 m long. In the tube side of E1003, Stream 1001 is condensed and exits at a pressure of 49.2 bar and a temperature of 8°C. Stream 1100, which contains propylene refrigerant at a pressure of 1.5 bar, a temperature of -36°C, and a flow rate of 164,789 kg/h, is supplied from Stage 5B to the shell side of E1003. Inside the shell, Stream 1100 partially evaporates before leaving and returning to Stage 11A at a pressure of 1.1 bar and a temperature of -36°C.

After exiting E1003, Stream 1001 passes through V1001, a throttling valve that drops the pressure to 1.2 bar, before entering D1000, a 304 stainless steel flash drum with a diameter of 2.88 m, a wall thickness of 10 mm, and a length of 8.65 m. Inside D1000, Stream 1001 splits into a vapor stream and a liquid stream, exiting the drum as Streams 1002 and 1000, respectively. Stream 1002 exits the drum at a pressure of 1.2 bar, a temperature of -101°C, and a flow rate of 39,976 kg/h and passes through V1002, a throttling valve that drops the pressure to 1 bar. After passing through the valve, Stream 1002 returns to Stage 10A where it recombines with Stream 1000 via V1000. Stream 1000 exits D1000 at a pressure of 1.2 bar, a temperature of -101°C, and a flow rate of 10,024 kg/h and moves to Stage 5A, where it passes through the tubes of E500 before returning to Stage 10A to restart the cycle.

Stage 11A (Compressing the Propylene Refrigerant)

The purpose of Stage 11A is to prepare the propylene refrigerant for flashing that occurs in Stage 11B. Serving as the “start” of the refrigeration loop, the stage begins with a liquid-vapor mixture of propylene that comes from Stage 10B. The stream is vaporized, combined with the vapor stream produced from flashing, and then compressed before being sent to Stage 11B. **Figure 21** shows the preparation of the propylene refrigerant through compression and heat exchange.

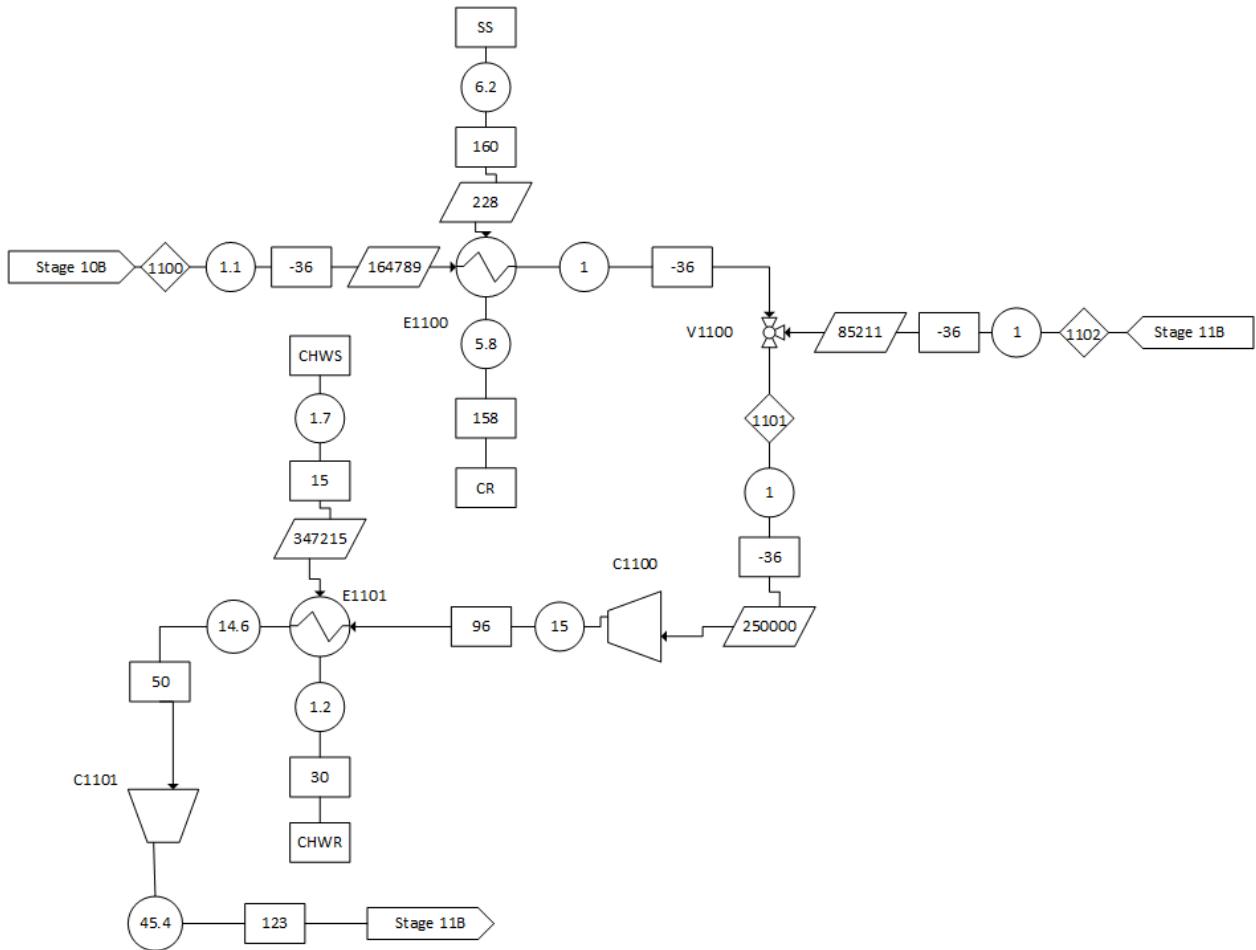


Figure 21. Stage 11A process diagram for the compression of propylene refrigerant in preparation for flashing

Stream 1100 exits Stage 10B and enters Stage 11A as a vapor-liquid mixture at a pressure of 1.1 bar, a temperature of -36°C , and a total flow rate of 164,789 kg/h and enters the tubes of E1100, a 1-1 pass shell-and-tube 304 stainless steel heat exchanger with 2 m^2 of available heat exchange area. E1100 has a shell OD of 0.17 m, and it contains 13 tubes, each with an OD of 2.54 cm. E1100 is 1.73 m long. All liquid present in stream 1100 is vaporized in E1100's tubes and exits E1100 at a pressure of 1 bar and a temperature of -36°C . Saturated steam at a flow rate of 228 kg/h, a pressure of 6.2 bar, and a temperature of 160°C is supplied to the shell of E1100 where it condenses inside the shell and returns as steam condensate at a pressure of 5.8 bar and a temperature of 158°C .

After exiting E1100, Stream 1100 enters V1100, a three-way valve, where it mixes with Stream 1102 from Stage 11B; Stream 1102 has a total flow rate of 85,211 kg/h, a pressure of 1 bar, and a temperature of -36°C . The two streams mix to form Stream 1101, which leaves V1100 at a flow rate of 250,000 kg/h, a pressure of 1 bar, and a temperature of -36°C . Stream 1101 enters C1100, a 13.32 MW 304 stainless steel centrifugal compressor, and exits at a pressure of 15 bar and a temperature of 96°C. It then enters the tubes of E1101, a 1-1 pass shell-and-tube carbon steel heat exchanger with 252 m^2 of available heat exchange area. E1101 has a shell OD of 0.81 m, and it contains 518 tubes, each with an OD of 2.54 cm.

E1101 is 7.76 m long. Stream 1101 is cooled in the tubes of E1101 and exits at a pressure of 14.6 bar and a temperature of 50°C. Chilled water at a flow rate of 347,215 kg/h, a pressure of 1.7 bar, and a temperature of 15°C is supplied to the shell of E1101 where it heats up, leaving at a pressure of 1.2 bar and a temperature of 30°C.

After exiting E1101, Stream 1101 enters C1101, a 5.34 MW carbon steel centrifugal compressor, and exits at a pressure of 45.4 bar and a temperature of 123°C. It then moves to Stage 11B.

Stage 11B (Flashing the Propylene Refrigerant)

The purpose of Stage 11B is to finish preparing the propylene refrigerant for flashing and to finally flash it, dropping its temperature to a cold enough level for the liquid to be used in the condensers of the de-ethanizer, C2 splitter, and Stage 10B. The vapor resulting from the flash is then fed back to Stage 11A to be recombined with the liquid stream, closing the loop, shown in **Figure 22**.

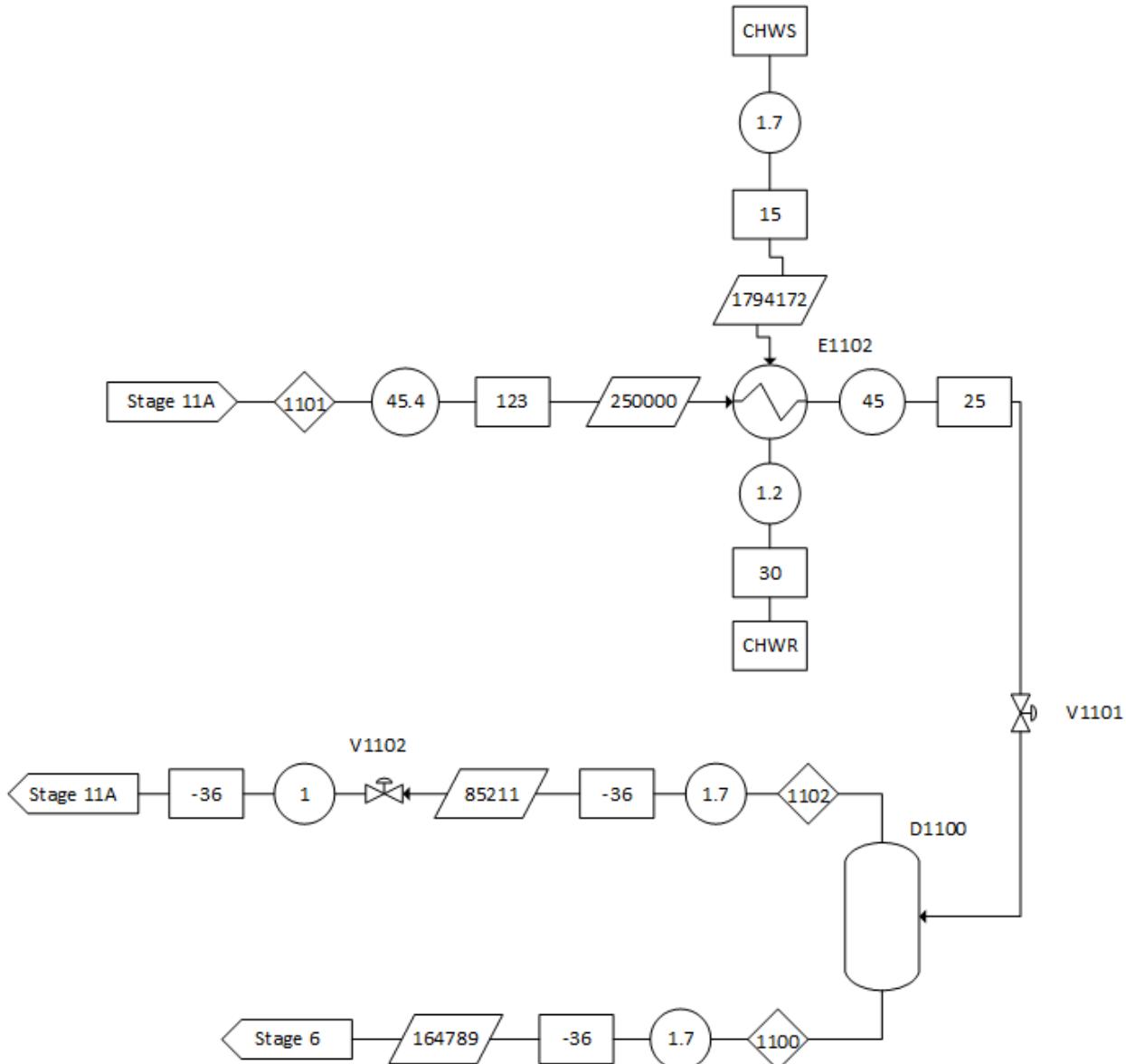


Figure 22. Stage 11B process diagram for the flashing of propylene refrigerant

Stream 1101 enters Stage 11B at a flow rate of 250,000 kg/h, a pressure of 45.4 bar, and a temperature of 123°C and enters the tubes of E1102, a 1-1 pass shell-and-tube carbon steel heat exchanger with 1699 m² of available heat exchange area. E1102 has a shell OD of 1.68 m, and it contains 5577 tubes, each with an OD of 1.59 cm. E1102 is 9.38 m long. In the tube side of E1102, Stream 1101 is cooled and exits at a pressure of 45 bar and a temperature of 25°C. Chilled water at a flow rate of 1,794,172 kg/h, a pressure of 1.7 bar, and a temperature of 15°C is supplied to the shell of E1002 where it heats up, leaving at a pressure of 1.2 bar, and a temperature of 30°C.

After exiting E1102, Stream 1101 passes through V1101, a throttling valve that drops the pressure to 1.7 bar, before entering D1100, a 304 stainless steel flash drum with a diameter of 3.94 m, a wall thickness of 15 mm, and a length of 11.81 m. Inside D1100, Stream 1101 splits into a vapor stream and a liquid

stream, exiting the drum as Streams 1102 and 1100, respectively. Stream 1102 exits the D1100 at a pressure of 1.7 bar, a temperature of -36°C, and a flow rate of 85,211 kg/h and passes through V1102, a throttling valve that drops the pressure to 1 bar. After passing through the valve, Stream 1102 returns to Stage 11A where it recombines with Stream 1100 via V1100. Stream 1100 exits the drum at a pressure of 1.7 bar, a temperature of -36°C, and a flow rate of 164,789 kg/h and moves to Stage 6, where it passes through the tubes of E600. It then moves to Stage 5B, where it passes through E502 and moves to Stage 10B, where it passes through E1003 before finally returning to Stage 11A to restart the cycle.

Stage 12A (Compressing the Propylene Refrigerant for Stage 4)

The purpose of Stage 12A is to prepare the propylene refrigerant for flashing that occurs in Stage 12B. Similar to Stage 11A, the stage begins with a liquid-vapor mixture of propylene that comes from Stage 4. The stream is vaporized, combined with the vapor stream produced from flashing, and then compressed before being sent to Stage 12B, shown in **Figure 23**.

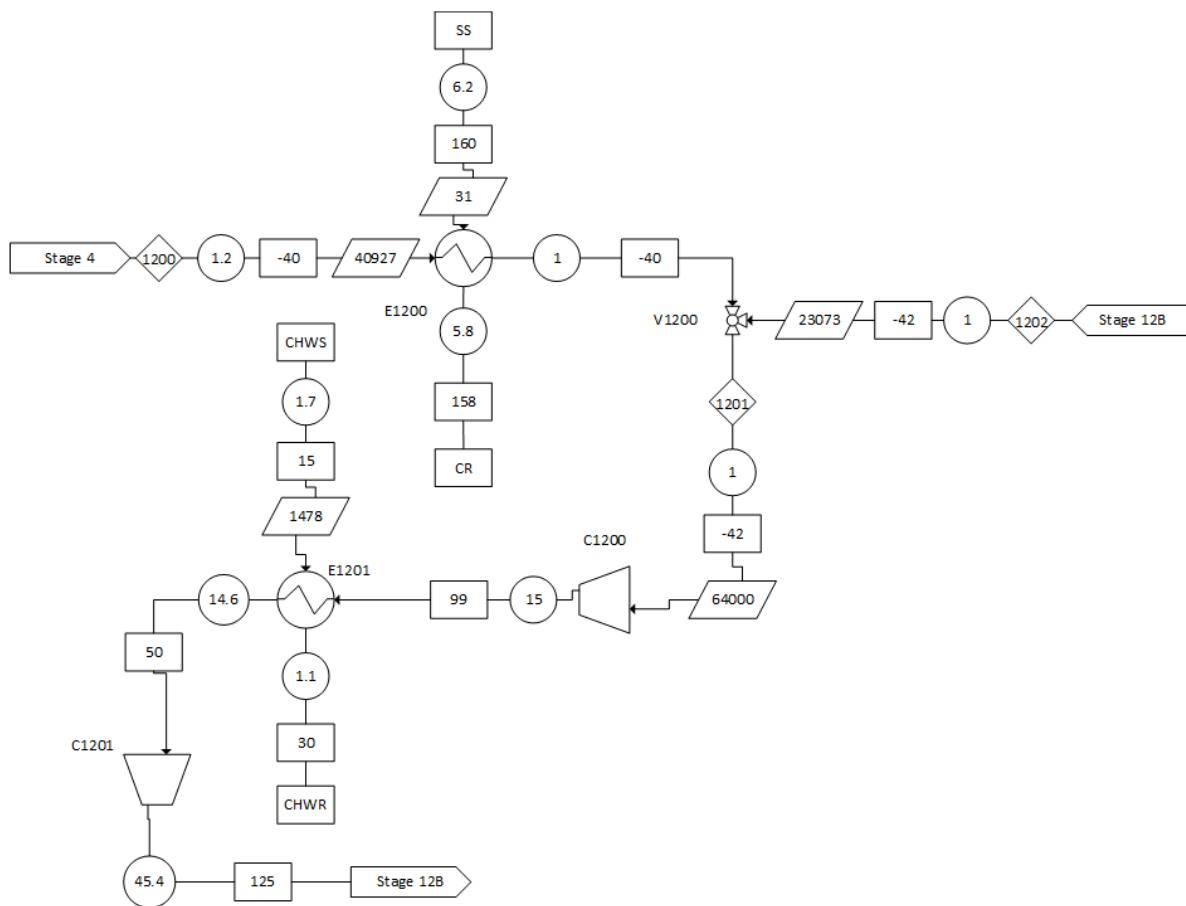


Figure 23. Stage 12A process diagram for the compression of propylene refrigerant in preparation for flashing

Stream 1200 exits Stage 4 and enters Stage 12A as a vapor-liquid mixture at a pressure of 1.2 bar, a temperature of -40°C, and a total flow rate of 40,927 kg/h and enters the tubes of E1200, a 304 stainless

steel shell and tube heat exchanger with an area of 1.3 m^2 , 13 tubes each with an outer diameter of 2.54 cm, a shell diameter of 0.17 m, and a length of 1.7 m. All liquid present in Stream 1200 is vaporized in E1200's tubes and exits at -40°C.

After exiting E1200, Stream 1200 enters V1200, a three-way valve, where it mixes with Stream 1202 from Stage 12B; Stream 1202 has a total flow rate of 23,073 kg/h, a pressure of 1 bar, and a temperature of -42°C. The two streams mix to form Stream 1201, which leaves V1200 at a flow rate of 64,000 kg/h, 1 bar, and -42°C. Stream 1201 enters C1200, a 3.46 MW 304 stainless steel centrifugal compressor, and exits at 15 bar and 99°C. It then enters the tubes of E1201, a carbon steel shell and tube heat exchanger with an area of 67 m^2 , 1,096 tubes each with an outer diameter of 1.59 cm, a shell diameter of 0.76 m, and a length of 2.9 m. Stream 1201 is cooled in the tubes of E1201 and exits at 50°C.

After exiting E1201, Stream 1201 enters C1201, a 1.41 MW carbon steel centrifugal compressor, and exits at a 45.4 bar and 125°C to go to Stage 12B.

Stage 12B (Flashing the Propylene Refrigerant for Stage 4)

The purpose of Stage 12B is to finish preparing the propylene refrigerant for flashing and to finally flash it, dropping its temperature to a cold enough level for the liquid to be used in the condenser in Stage 4. The vapor resulting from the flash is then fed back to Stage 12A to be recombined with the liquid stream, closing the loop. The flashing process for the propylene refrigerant is shown in **Figure 24**.

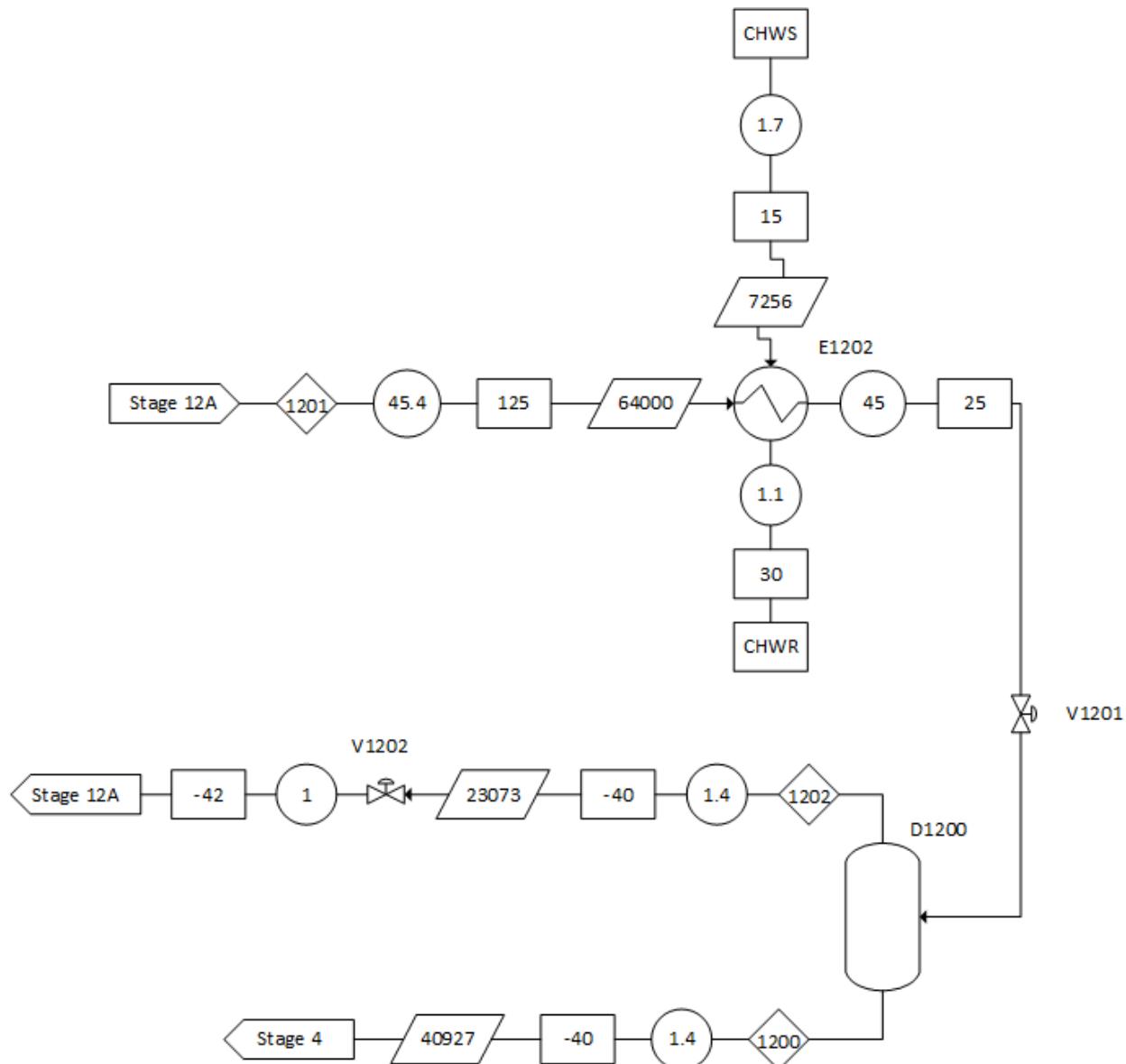


Figure 24. Stage 12B process diagram for the flashing of propylene refrigerant

Stream 1201 enters Stage 12B at a flow rate of 64,000 kg/h, a pressure of 45.4 bar, and a temperature of 125°C and enters the tubes of E1202, a carbon steel shell and tube heat exchanger with an area of 430 m², 1,475 tubes each with an outer diameter of 1.91 cm, a shell diameter of 1.02 m, and a length of 7 m. Stream 1202 is cooled and exits at 25°C.

After exiting E1202, Stream 1201 passes through V1201, a throttling valve that drops the pressure to 1.4 bar, before entering D1200, a 304 stainless steel flash drum with a diameter of 2.12 m, a wall thickness of 9 mm, and a length of 6.35 m. Inside D1200, Stream 1201 splits into a vapor stream and a liquid stream, exiting the drum as Streams 1202 and 1200, respectively. Stream 1202 exits D1200 at 1.4 bar, -40°C, and a flow rate of 23,073 kg/h and passes through V1202, a throttling valve that drops the pressure to 1 bar. After passing through the valve, Stream 1202 returns to Stage 12A where it

recombines with Stream 1200 via V1200. Stream 1200 exits the D1200 at 1.4 bar, -40°C, and a flow rate of 40,927 kg/h and moves to Stage 4, where it passes through the tubes of E403 and then returns to Stage 12A to restart the cycle.

ENVIRONMENTAL, HEALTH, AND SAFETY ANALYSIS

The following section details various process hazards as well as environmental impact of our proposed plant design. It will consider process hazards throughout each stage of our process design as well as discussion of waste streams and their handling.

Stage 1: Methanol Storage and Unloading

Stage 1's main process hazard is a loss of containment of methanol. Methanol liquid and vapor is flammable, is toxic if swallowed, in contact with skin, or inhaled, and can cause serious damage to eyes and other organs [19]. Therefore, the plant considers the safe unloading of methanol to be of the utmost importance. Prior to unloading, the railcar must be completely at rest with the hand brakes set and the wheels locked and be visually inspected for leaks or deformations [20]. Additionally, access to the rail must be blocked by a derail or some other function to prevent collisions. As a company policy, blue caution signs will be set up around the car to identify it as a hazard. Class 1 Div 1 classification will be given to the entire production facility, so cigarettes, lighters, and other sources of ignition will be strictly prohibited in the plant, and the train will be electrically grounded to prevent a buildup of static electricity, as illustrated in **Figure 25** [21]. Methanol storage tanks will be inspected every five years by a certified inspector, as per API 653 guidelines [22].



Figure 25: Electrical grounding of railcar to prevent ignition [23]

All gaskets, fittings, valves, and fasteners will be visually inspected prior to connection to the pumps, which will move the methanol to the large storage tanks. If any equipment appears to be unfit (i.e. is corroded, torn, worn, stripped, or otherwise deformed), unloading at the station will pause until proper equipment is provided. Within the storage tanks, the methanol continues to function as a hazard. To ensure the liquid level of the tanks does not exceed tank capacity, control valves will stop the flow of methanol into the storage tanks at 95% level. Stage 1 does not produce any waste streams. Each tank will be surrounded by a dike large enough to hold the entire volume of methanol within the tank to prevent leaks from spreading to the surrounding areas of the plant in the event of a tank rupture. Emergency alarm switches that will notify the local fire department will be located around the storage and unloading area in the event of ignition, and employees will be trained on where to take shelter in the case of such an event.

Stage 2: Reaction of Methanol to Hydrocarbons

Stage 2 has several hazards relating to high temperature and loss of containment. The reactor-regenerator system has a delicate pressure and temperature balance, and if the flux of catalyst through the cooler is not maintained, temperatures inside the regenerator may damage the catalyst and would likely cause degradation of reaction products. Due to the highly exothermic nature of the reaction, if the cooling rate was lower than necessary, a runaway reaction could occur, which could result in pressure buildup in the vessels and if oxygen came into contact with the reactor, combustion as well. To ensure the catalyst cooling rate is maintained, proper amounts of fluidization and lift air will be provided, and condensate will be pumped at the necessary rate for 131,462 kg/h of high pressure steam. A diagram of the catalyst cooler is shown in **Figure 26**.

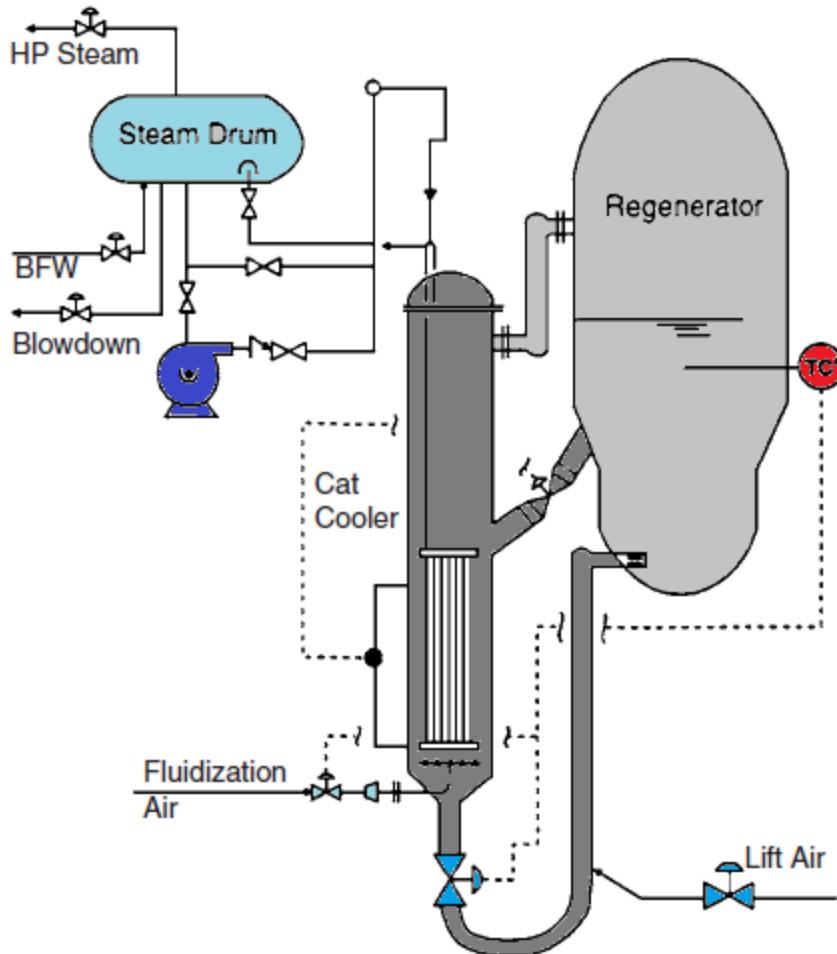


Figure 26: Catalyst cooler to manage heat produced in the regenerator

If the temperature coming off the regenerator exceeds a threshold temperature of 800°C, the spent catalyst valve connecting the reactor to the regenerator will be tightened to stop the flow of coke, which fuels the reaction, to the regenerator. If the temperature of the reactor exceeds 500°C, the inlet methanol flow will be stopped, as products will be off specification if the reactor is not near 470°C. In addition to the health hazards posed by methanol, many of the products are also highly hazardous. Methane, ethane, ethylene, propane, propylene, propane, butylene, and C5+ hydrocarbons are all extremely flammable and can displace oxygen and suffocate humans [24] [25] [26] [27] [28] [29]. Additionally, inhalation of butylene can cause genetic defects and cancer. Regular inspection of pipes and process units, including heat exchangers, quench towers, and compressors, is required to ensure containment. In the event of a major process upset, Stage 2 will be sent into its fail-safe mode by the DCS.

Stage 2 produces waste in Stream 210, which contains nearly 99.9% water by mass. This will be sent to our in-plant wastewater treatment plant. Though a detailed wastewater treatment plant design is outside the scope of our project, the wastewater will be treated similarly to typical wastewater, albeit at

a significantly smaller scale. It will first be sent through a screen as a precaution to remove any large solids, which should not be present in the waste [30]. Then the water will be sent through a grit chamber and sedimentation tank to remove any small particles of corroded pipe and/or catalyst particles that escaped collection by the reactor cyclones. This primary treated wastewater will be sent to an aeration chamber, where the water will be mixed with sludge containing high concentrations of bacteria as well as air. A schematic of the secondary treatment is shown in **Figure 27**.

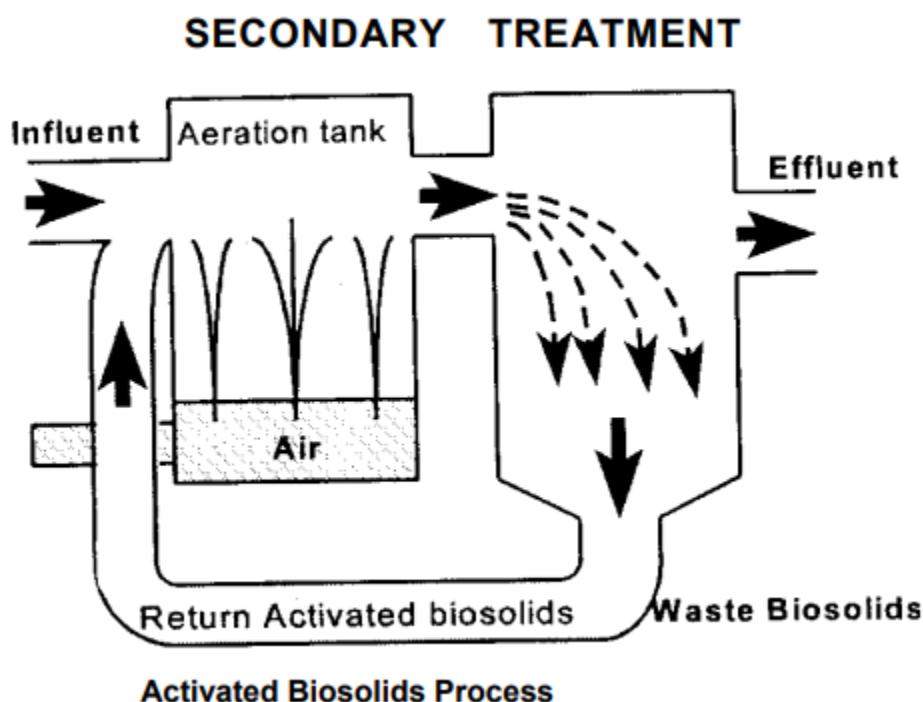


Figure 27: Secondary treatment of wastewater using bacteria [30]

The bacteria will convert approximately 85% of the hydrocarbons and methanol into harmless byproducts such as carbon dioxide. The water will be sent to a large chamber to allow bacteria to settle and then will be disinfected with chlorine before the water is released into the environment. In situations where chlorinated water is harmful to the fish and aquatic environment, alternatives such as ozonation and UV irradiation will be explored instead.

Stages 3-4: Hydrocarbon Purification and Distillation Train Preparation

Stages 3-4 have multiple hazards relating to high temperatures in the TSA bed and pipelines as temperatures can reach upwards of 200°C when regenerating the silica gel. Hydrocarbons may also be adsorbed within the silica gel so desorption may release flammable vapors. During the distillation train preparation the vapor stream can get very hot when being compressed by some compressors and the temperature also reaches cryogenic temperatures that can still be harmful if come into contact with. To account for these hazards, insulation will be used on piping and vessels to mitigate direct contact with

hot and cold equipment. A nitrogen purge stream may also be used before regenerating the TSA beds to remove oxygen and temperature interlocks and alarms will be put in place to prevent overheating.

Pressures within pipelines during the distillation train preparation reach pressures above 20 bar which means over-pressurization can be a significant hazard. To mitigate this, pressure relief valves will be installed along with rupture discs for very severe cases of over-pressurization. Because the hydrocarbons are well above atmospheric pressure at certain points, loss of containment would mean the rapid vaporization of flammable material. To mitigate this possibility, each pressure relief valve and rupture disc will be connected to the DCS and an emergency-stop command can be executed to stop more effluent from moving through the ruptured pipes.

Caustic solution used in the caustic scrubber can cause damage to both personnel and equipment if precautions aren't taken. Sodium hydroxide is corrosive to most metals so poor material selection could lead to failure or leaks. If leaks do occur, a slippery surface could form around the caustic scrubber that can lead to slipping hazards on top of possible chemical burns if direct contact occurs. To mitigate this, the caustic scrubber is made with 304 stainless steel and a proper wall thickness to account for corrosion and limit the possibility of leaks. Proper PPE will be provided when working around the caustic scrubber for further protection against possible leaks.

Multiple towers in stage 3 have liquid waste streams that contain different contaminants. For water waste streams, they will be sent to a wastewater treatment plant, similarly to the water waste streams in stage 2 and treated with the same methods. For the spent caustic solution stream, this will be sent to a hazardous waste treatment plant where the stream will undergo neutralization by adding acids to neutralize the caustic. Once the pH reaches 7 it can go through oxidation using wet air oxidation, which uses high temperature, high pressure air or oxygen to convert organics into carbon dioxide and water.

Stages 5-8: Separation of C1 and C2 Hydrocarbons and Ethylene, Propylene, and Butylene Recovery

Stages 5-8 share similar hazards, all relating to the distillation processes that they run. Within the distillation train, temperatures of process streams range from -96 to 124°C, which can cause severe damage if contacted. For this reason, all lines within Stages 5-8 that are not sent immediately to an underground pipeline will be insulated.

Inherent to the distillation process are high operating pressures. Pressures above 20 bar are maintained in this section of the process, so over-pressurization is a significant hazard. To mitigate this, each column was designed with a wall thickness above the minimum thickness for a pressure vessel, adding on a safety factor to the design. Pressure safety valves will be installed on each of the process lines and equipment within these stages in order to avoid over-pressurization. Moreover, rupture discs will be installed on each column for the same reason. Because the hydrocarbons within this section of the process are well above atmospheric pressure, a loss of containment would mean the rapid vaporization of flammable material. To keep this possibility to a minimum, every PSV and rupture disc will be connected to the DCS, and the emergency-stop command will be executed for the stage that has been overpressurized, putting the stage into its fail-safe condition. Emergency alarm switches will be located around the distillation area of the process, which will notify the local fire department if activated, and

employees will be trained on where to take shelter in the case of such an event. Each column will be inspected once every five years by a certified inspector, per API 510 guidelines [31]. Stages 5-8 do not produce any waste streams.

Stage 9: C5+ Hydrocarbons Storage and Shipping

Similarly to Stage 1, the largest process hazard in Stage 9 is loss of containment of the storage material, which is the C5+ byproduct stream. Because the C5+ stream is composed of flammable hydrocarbons, the entire storage area will be electrically classified as Class 1, Div 1 [32]. Dip tubes will be installed in each C5+ storage tank to prevent the buildup of a static spark when charging each tank. The inlet to each tank will be controlled by valves communicating with the DCS to stop the filling of each tank at 90% level, as 10% of the tank volume is designated for freeboard. LEL alarms will be installed around the storage and unloading areas and will be connected to the DCS to alarm if activated. Each tank will be surrounded by a dike large enough to hold the entire volume of each C5+ within the tank to prevent leaks from spreading to the surrounding areas of the plant in the event of a tank rupture. Emergency alarm switches that will notify the local fire department will be located around the storage and unloading area in the event of ignition, and employees will be trained on where to take shelter in the case of such an event. Each tank will be inspected every five years by a certified inspector, per API 510 guidelines [33]. Stage 9 does not produce any waste streams.

Like in Stage 1, all gaskets, fittings, valves, and fasteners will be visually inspected prior to connection to the pumps, which will move the C5+ byproduct to the tanker trucks. If any equipment appears to be unfit (i.e. is corroded, torn, worn, stripped, or otherwise deformed), unloading at the station will pause until proper equipment is provided.

Stages 10-12: Refrigeration Loops

Stages 10-12 share the same hazards, mainly centered around cryogenic temperatures and loss of containment of refrigerant. To satisfy the temperature demands of our cryogenic distillation units, stages 10-12 have process streams well below 0°C. Furthermore, the compressor outlet streams are at temperatures well over 50°C. Accordingly, all equipment and process lines containing fluids less than 25°C or greater than 30°C will be properly insulated to prevent potential personnel injuries from touching the equipment. Additionally, all personnel working in close proximity to the refrigeration systems will be equipped with PPE including insulated clothing and gloves, splash-proof safety goggles, and respirators to mitigate the dangers of contact with the equipment or inhalation of refrigerant in the event of a refrigerant leak [34].

Ethylene and propylene, used as refrigerants in stage 10 and stages 11 and 12, respectively, are both highly flammable vapors at ambient pressures [27] [28]. All electrical wiring in the building will be approved for Class 1, Division 1, Hazardous Locations, in accordance with OSHA regulations. To prevent loss of containment of the refrigerants, the equipment and process lines will be inspected daily for leaks. When repairs to the equipment are needed, we will obtain hot work permits as necessary and use inert

gas to flush the system of refrigerant before any repairs take place. Additionally, the process lines will be equipped with pressure relief valves and temperature gauges integrated with the DCS to allow close monitoring and quick diagnosis of potential equipment failure or leaks. To mitigate the dangers of containment loss incidents, PSV and temperature alarms can alert operators to put the stage into its fail-safe state. Additionally, emergency alarm switches will be located around this area of the process, which will notify the local fire department if activated, and employees will be trained on where to take shelter in the case of such an event. Stages 10-12 do not produce any waste streams.

ECONOMIC FEASIBILITY ANALYSIS

The following section presents an economic feasibility analysis of our proposed MtO plant. The total capital investment (TCI) for the proposed plant is \$756.88 MM. The annual operating cost is \$947.01 MM, and the annual revenue is \$710.41 MM. As operating costs are larger than revenue, the plant is not expected to provide a return on investment and has an overall net present value (NPV) of \$-2.86 BN. The following analysis includes discussions on capital investment, operating costs, revenue, overall profitability of the process, and a sensitivity analysis of our process that considers possible methods of reaching profitability.

Capital Investment

The following sections detail our results and rationale for estimating the capital investment of our process. Specifically, discussions on inside boundary limits (ISBL) capital costs, outside boundary limits (OSBL) capital costs, fixed capital investment (FCI), and total capital investment (TCI) are included.

Inside Boundary Limits Capital Costs. ISBL capital costs account for costs directly related to the project such as the cost of purchasing and installing the equipment needed for our process. We determined the cost of most of the process units using costing curves, and we used the factorial method to account for materials of construction as well as a variety of costs associated with the implementation of the equipment into the plant. The reactor and several trayed column costs were estimated using reported data on MtO processes [35]. **Table 1** below summarizes the calculated costs for our process, and detailed calculations for each individual piece of equipment are provided in **Appendix D**.

Process Unit	Equipment (\$MM)	Installation (\$MM)	Piping (\$MM)	Electrical (\$MM)	Control Systems (\$MM)	Structures and Paint (\$MM)	Total (\$MM)
Heat Exchangers	11.44	3.31	9.16	3.31	2.20	6.61	36.03
Compressors /Blowers	78.24	21.71	62.59	14.47	21.71	43.42	242.14
Pumps	1.01	0.29	0.81	0.20	0.29	0.59	3.19
Reactor & Regenerator	36.39	10.92	29.11	7.28	10.92	21.83	116.45
Storage Tanks	8.07	2.40	6.46	1.60	2.40	4.80	25.73
Trayed Columns	14.02	3.51	11.22	2.34	3.51	7.02	46.89
Quench Towers	1.27	0.38	1.02	0.25	0.38	0.76	4.06
Dryer	0.75	0.21	0.55	0.14	0.21	0.41	2.27
Flash Drums	0.31	0.08	0.25	0.05	0.08	0.16	0.91

Table 1. Equipment and implementation costs for our process

As can be seen in the table, the equipment costs for our process are dominated by compressors and the reactor regeneration system. The Pareto diagram below (**Figure 28**) summarizes the total costs associated with each process unit type.

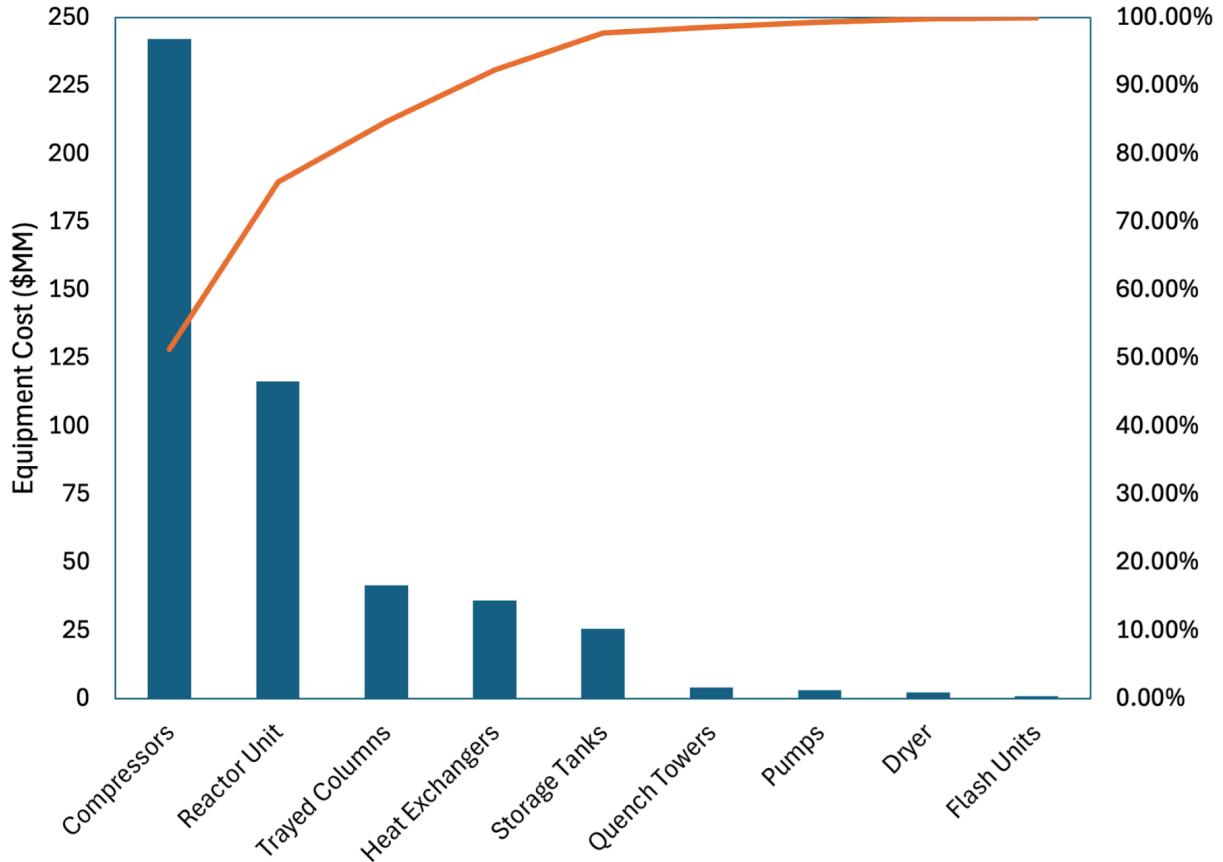


Figure 28. Pareto diagram for the purchase and installation costs of each process unit type

Summing the total cost associated with each process unit type gives us the ISBL for our process: \$472.40 MM.

Outside Boundary Limits Capital Costs. OSBL capital costs account for costs indirectly related to the project such as expansions to existing utility streams or the addition of access roads for the plant. The OSBL is typically estimated as 25-50% of ISBL costs. Since our process utilizes large amounts of utilities such as cooling water and chilled water, we estimated the OSBL costs to be 50% of the ISBL, resulting in a value of \$236.20 MM.

Fixed Capital Investment. The FCI is the sum of ISBL costs, OSBL costs, engineering costs, and money set aside for contingency. In general, engineering costs and contingency money are both estimated as 10% of the ISBL and OSBL costs summed together. Accordingly, we estimated both engineering costs and contingency money to have a value of \$70.86 MM. Taking the sum of all these costs together, we achieve a FCI of \$850.33 MM. However, the previous calculations were all performed under the assumed location of the US gulf coast. Since our plant will be located in China, we need to multiply our FCI by a location factor based on the exchange rate of the US dollar and the yuen [36]. Assuming that we source our equipment internally from China, we can multiply our previously determined FCI by a factor of 0.69, resulting in a value of \$586.73 MM. **Table 2** below summarizes the various contributions to the FCI.

	Cost	\$MM
<i>ISBL</i>		
Equipment	151.51	
Installation	42.80	
Piping	121.16	
Electrical	29.64	
Control Systems	41.70	
Structures and Paint	85.60	
OSBL (50% ISBL)	236.20	
Engineering (10% ISBL+OSBL)	70.86	
Contingency (10% ISBL+OSBL)	70.86	
FCI (ISBL+OSBL+Engineering+Contingency)	850.33	
Adjusted FCI (FCI*location factor)	586.73	

Table 2. A summary of the components that contribute to the FCI

Total Capital Investment. The total capital investment accounts for all capital in the project, including the FCI, working capital, and startup costs. In general, the working capital of a process is the money that will be required for the business to operate during the construction of the plant. It is typically estimated as 15-25% of the FCI; we do not have specific expectations for the amount of operating capital we will need, so we estimated our working capital as 20% of the FCI, resulting in a value of \$117.35 MM. Startup costs account for costs incurred during the initial production phase that are not included in operating costs. They are typically estimated as 8-10% of the FCI. Similar to the working capital, we do not have specific expectations for the startup costs and estimated them as 9% of the FCI, resulting in a value of \$52.81 MM. Taking the sum of these costs and the FCI, we achieve a TCI of \$756.88 MM. **Table 3** below summarizes the various contributions to the TCI.

	Cost	\$MM
FCI		586.73

Working Capital	117.35
Startup Costs	52.81
TCI (FCI+Working+Startup)	756.88

Table 3. A summary of the components that contribute to the TCI

Operating Costs

The following sections detail the various costs associated with operation of our plant. Specifically, discussions on direct production cost, fixed production cost, plant overhead, and general expenses are included.

Direct Production Cost. The direct production cost of a process consists of the costs incurred through daily operation. In general, it includes the costs of raw materials, utilities, waste processing, labor, lab costs, supplies, and maintenance. In terms of costing raw materials, we need to account for methanol as our process feed, SAPO-34 to counter attrition in the reactor, 50% sodium hydroxide solution for our caustic scrubber, and water for our quench towers. Utilities include electricity for compressors and pumps and coolants and heat sources for heat exchangers. Waste processing includes costs for both hazardous and nonhazardous waste, and operating labor consists of operators, maintenance, lab techs, and engineers. More detail on how these costs were calculated can be found in **Appendix D. Table 4** summarizes the contributions to the direct production cost, which totals \$834.64 MM.

Cost	\$MM/yr
Materials	654.33
Utilities	95.52
Waste	71.99
Operating Labor	9.02
Supervisors (10% Operating Labor)	0.90
Lab Costs (10% Operating Labor)	0.90
Maintenance (6% FCI)	0.26
Supplies (15% Maintenance)	0.17
Direct Production Cost	834.66

Table 4: Annual direct production cost for our process

Fixed Production Cost. The fixed production cost accounts for expenditures that the plant must make even while not operating for things including property tax and insurance. This tends to scale with the scale of the plant and is typically estimated by taking 4% of the FCI; thus, we assume a fixed production cost of \$1.16 MM/yr.

Plant Overhead. Plant overhead costs provide funding for supporting personnel such as human resources, security, property maintenance, and janitors. This is typically estimated based on costs for workers more directly associated with the process, including operating labor, maintenance, and supervisors. We estimate plant overhead costs as 60% of the costs of operating labor, maintenance, and supervisors, or \$7.00 MM/yr.

General Expenses. General expenses are company-level expenses for items such as research and development, marketing, and administrative expenses. These are typically estimated as a fraction of the total production cost (TPC), where research and development makes up about 5-50% of the TPC, marketing makes up about 2-20% of the TPC, and administrative expenses make up about 2-6% of the TPC. Assuming MichiChem operates and invests in well-understood processes like MtO and that marketing expenses will be directed towards maintaining contracts with other companies, both the marketing and research expenses will be on the lower end of their range. Thus, we estimate that research and development makes up about 5% of the TPC, marketing expenses are 2% of the TPC, and administrative costs make up 4% of the TPC.

Total Production Cost. The total production cost is the sum of the direct production cost, fixed production cost, plant overhead, and general expenses, accounting for all the costs related to operating the plant, owning the plant, paying for a fraction of the plant's supporting staff, and paying for a fraction of the plant's research, administration, and marketing. **Table 5** summarizes the various contributions to the TPC.

Cost	\$MM/yr
Direct Production Cost	834.66
Fixed Production Cost	1.17
Plant Overhead	7.01
General Expenses	104.17
Research and Development	47.35
Marketing	18.94
Administrative	37.88
Total Production Cost	947.01

Table 5. Annual production costs for our process

The TPC totals \$947.01 MM/yr, and the direct production cost makes up over 85% of the TPC. Thus, to significantly reduce the TPC, materials (which make up almost 80% of the direct production cost) must be purchased at a cheaper price.

Revenue

The following section details the various sources of revenue for our plant. Specifically, revenues from materials and utilities are included. About 76.6% of our yearly revenue comes from selling the desired products of ethylene and propylene, as shown in **Table 6**, and more detailed calculations on how these revenues were calculated can be found in **Appendix D**.

Revenue Source	\$MM/yr
Materials	667.89
Methane	2.40
Ethane	2.86
Propane	1.93
Ethylene	292.85
Propylene	251.28
Butylene	83.84
C5+	32.72
Utilities	42.52
Low Pressure Steam	8.30
High Pressure Steam	34.23
Total Production Revenue	710.41

Table 6. Annual revenues from materials and utilities for our process

The total yearly revenue is \$710.41 MM, which is \$236.6 MM less than our total production cost. Thus, unless revenue increases or the total production cost decreases, the process will not be profitable.

Profitability of the Process

With the capital costs, annual operating costs, and annual revenue determined, we can analyze the net profitability of the process. In general, MtO plants operate for approximately 20 years [37]. As such, we

assume our plant will operate for 20 years, with 1 year of construction (21 years total). We assume that the FCI will be spent during the first year and that the startup costs and working capital will be used during the first year of operation. To account for depreciation of our equipment, we used a 20 year MACRS schedule. Though our plant is located in China, we have been unable to find posted depreciation schedules for China, so we used the MACRS schedule as an estimate. Lamentably, our operating costs are significantly higher than our revenue, so the depreciation rates and tax rates in China do not ultimately matter because our plant will have no profit and, therefore, no taxable income. Assuming a hurdle rate of 11%, as dictated by analysts at Michichem, we can discount the cash flows for each year and construct a cumulative cash flow diagram. The cumulative cash flow diagram for our plant is depicted below in **Figure 29**. See **Appendix D** for detailed profitability calculations and a cash flow table.

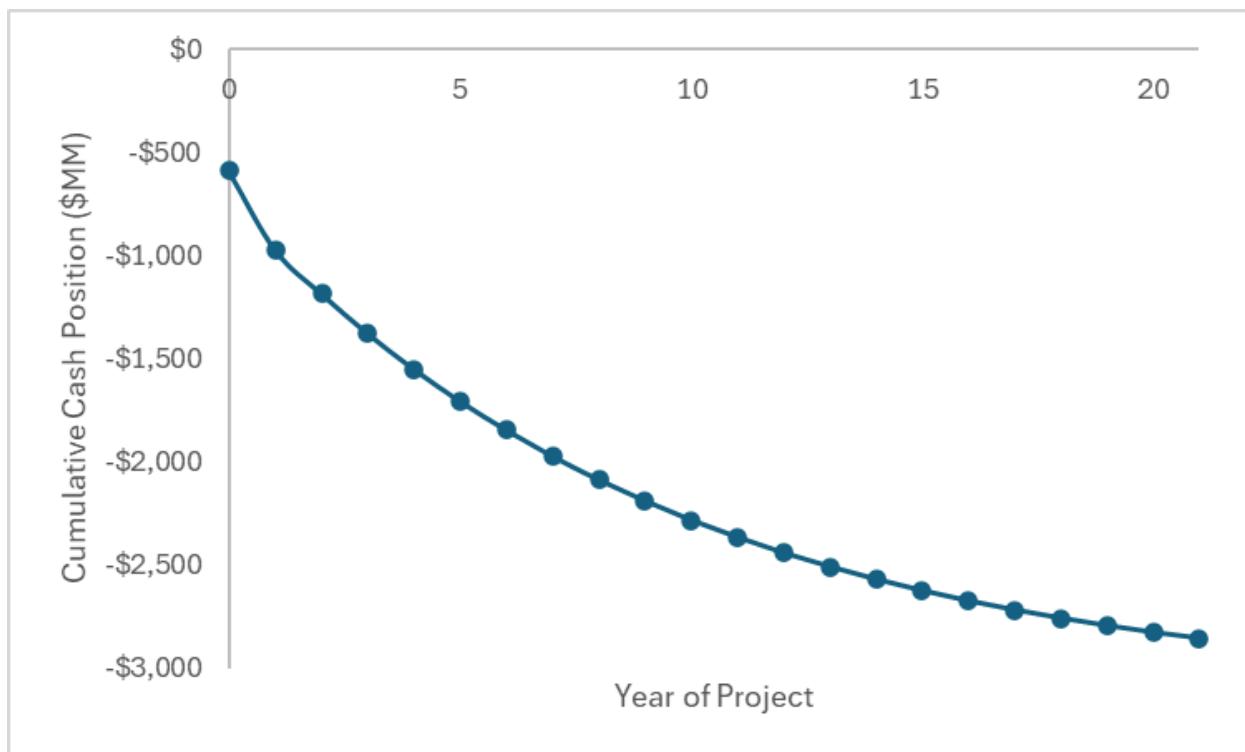


Figure 29. Cumulative cash flow for our process throughout the project lifetime

As can be clearly seen in **Figure 29**, the plant will continually incur debt as it operates and will never provide a return on investment, ultimately leading to a NPV of \$-2.86 BN. Naturally, this means that we cannot determine an internal rate of return (IRR) or a discounted payback period (DPBP).

Sensitivity Analysis

Although the process is not currently economically viable, a sensitivity analysis helps identify the key factors driving this unprofitability and highlights which cash flows have the greatest impact on the economic standing of the process. The following section presents this analysis in detail, outlines the set of conditions that must be met for the process to become profitable, and provides a basic economic

analysis of this profitable scenario. It also adds a brief description of the potential of green methanol and how that would affect taxes.

Conditions for Profitability. To assess which cash flows have the largest impact on the profitability of the process, a spider diagram was created. This visual, provided below in **Figure 30**, illustrates how key financial metrics (revenue, operating cost, capital investment) influence the NPV at an 11% hurdle rate.



Figure 30. Spider diagram of the process varying revenue/price, operational cost, and capital investment

Because higher slopes correspond to higher sensitivity of the NPV to fluctuations in the financial input, we identified from this spider diagram that minimizing operational cost would be the most feasible approach to increasing the profitability of the process.

Currently, our annual operating costs outweigh our annual revenue, so our process loses money each year it operates. To address this, we can utilize a Pareto diagram to visualize our direct production costs to see where in the process we are spending the most money. This Pareto diagram is provided below in **Figure 31**.

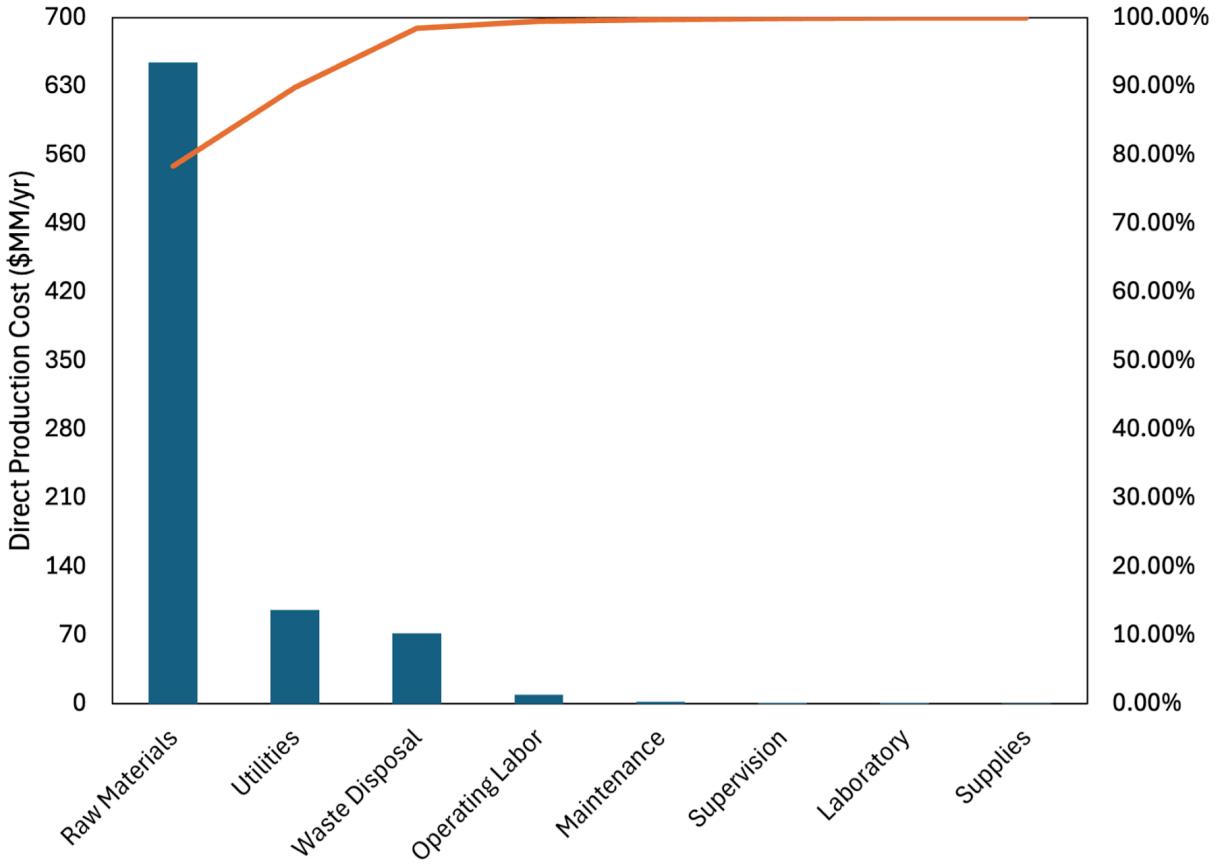


Figure 31. Pareto diagram of direct production costs

Drawing from **Figure 31**, it can be seen that raw materials account for nearly 80% of direct production costs, and the second largest cost is utility usage. Because the cost of raw materials is significantly higher than any of the other costs, another Pareto diagram is provided in **Figure 32** to further visualize what is contributing to this category and to identify which specific materials are driving the majority of the expense.

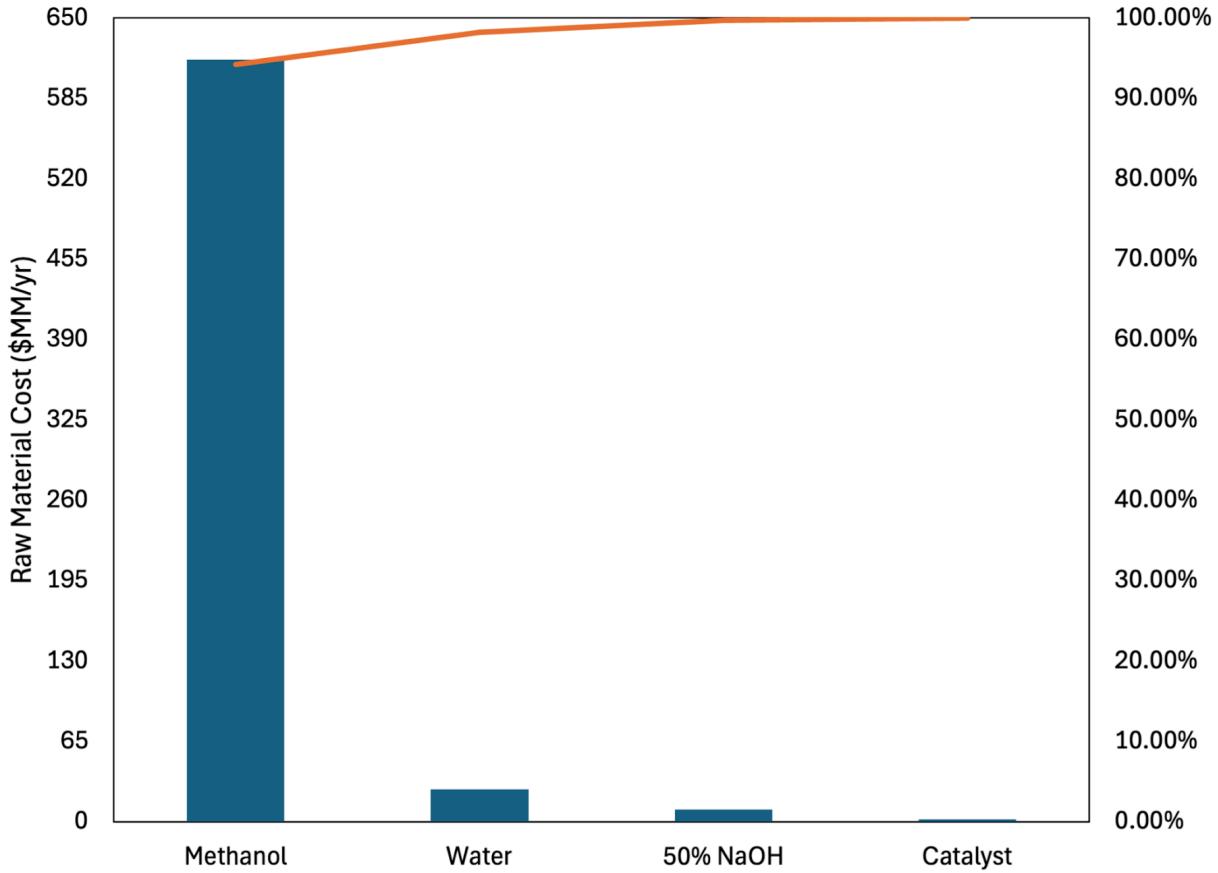


Figure 32. Pareto diagram of raw material costs

The sourcing of methanol feedstock is the largest driver of the raw material costs, corresponding to over 90% of the total expense. Because of the massive influence the methanol feedstock price has over the profitability of the process, this was the most important cash flow to minimize in order to create a scenario where the process is profitable. Utilities and waste treatment also contribute between 10 and 20% of the direct production costs, so these were also variables that were considered in the remainder of the sensitivity analysis.

Although the price of methanol used so far in this economic analysis has been \$400/MT [38], this market value is always fluctuating, as shown below in **Figure 33**.

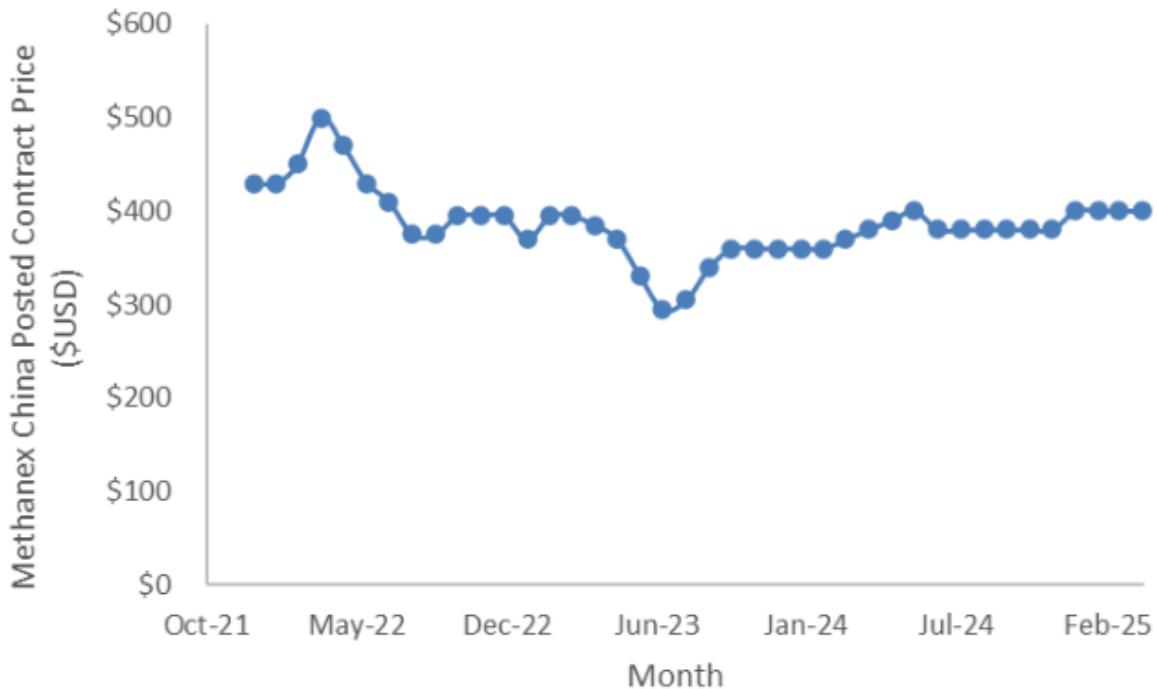


Figure 33. Fluctuations in Chinese methanol prices from January 2022 to present [38]

Although Methanex is widely considered one of the most reliable sources for pricing methanol, other sources [39] have methanol in the Northeast Asia market priced as low as \$265/MT. Using this value significantly increases the NPV of the process to -\$772 MM. Additionally, we can propose a 15% reduction in operating costs related to utilities and waste treatment. In an iterative design, the process could be further optimized in terms of energy usage and waste generation, making a 15% reduction a feasible target. Applying all of these conditions together (\$265/MT methanol, 85% current utility and waste treatment costs), an NPV of -\$534 MM is achieved and annual revenue exceeds annual costs. Because this process is now profitable on an annual basis, the standard 25% corporate income tax for China must be considered as well [40]. A cash flow table for this case, referred to as Scenario One, is included in **Appendix D**. The corresponding cumulative cash flow diagram is shown below in **Figure 34**.

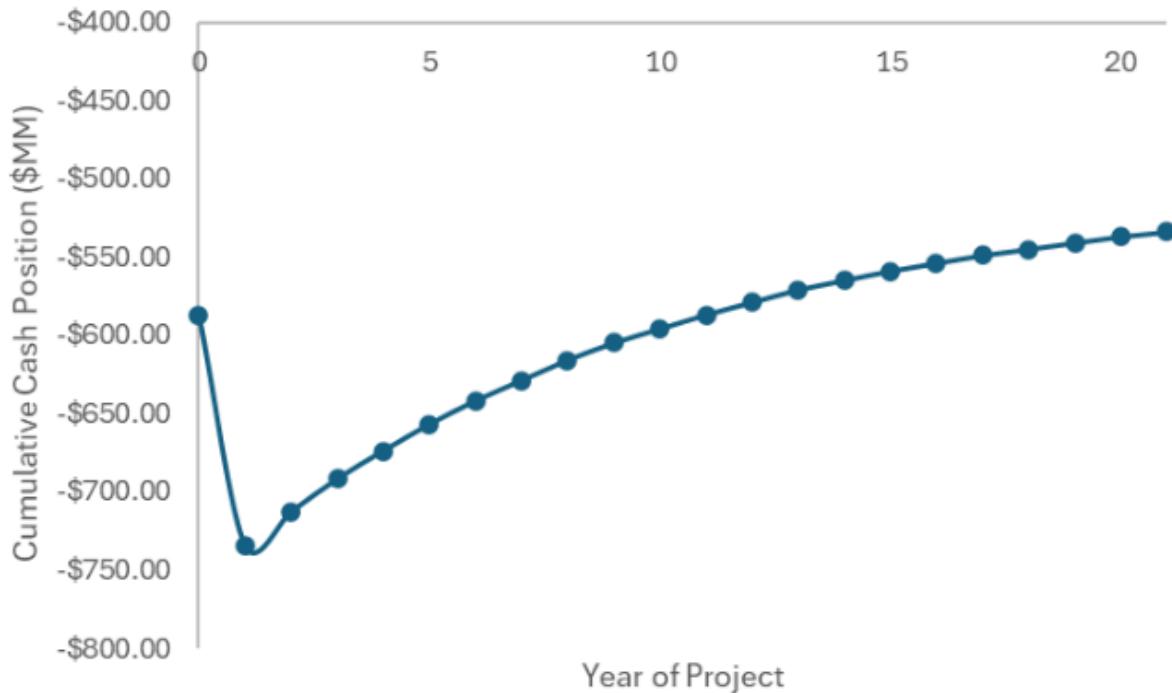


Figure 34. Cumulative cash flow for Scenario One throughout the project lifetime

Because this process still yields a negative NPV, we recognize that the best way to break even would be to further decrease the price of methanol. Because we are operating next to production facilities and pipelines that we send our products to, we could propose a contract with a nearby methanol facility to install a pipeline to supply our methanol feedstock. This is common in industry, and especially feasible because our process would buy this methanol in large quantities. This decision would also decrease the capital investment of the process, as there would be no costs associated with Stage 1 because the methanol is no longer being unloaded into storage tanks. Assuming a 20% decrease in methanol from the \$265/MT methanol price, as well as effectively eliminating Stage 1 by receiving methanol from a pipeline, the NPV of the process would come out to \$114 MM. A cash flow table for this profitable scenario, referred to as Scenario Two, is included in **Appendix D**. The corresponding cumulative cash flow diagram is shown below in **Figure 35**.

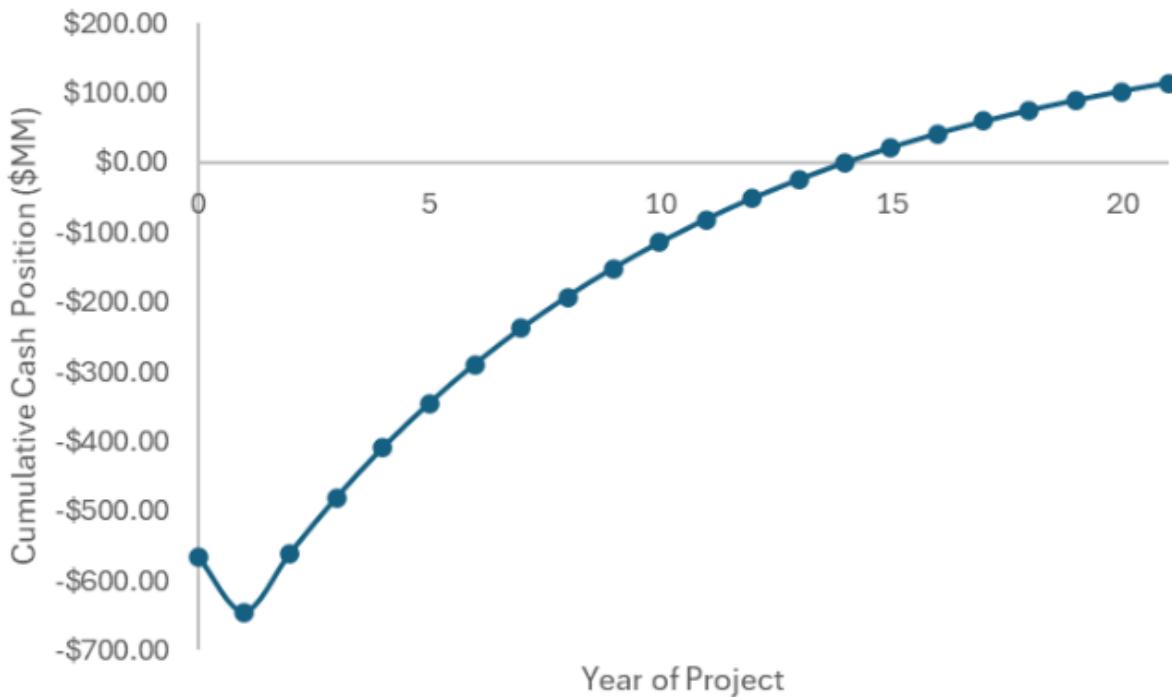


Figure 35. Cumulative cash flow for Scenario Two throughout the project lifetime

Because Scenario Two is profitable over its lifetime, an IRR and DPBP can be obtained. The detailed calculation is shown in **Appendix D**, and the IRR and DPBP come out to 13.6% and 14 years, respectively. This means that the effective return of Scenario Two is 13.6%, and the investment will take 14 years to pay itself off. **Table 7** summarizes all of the economic scenarios discussed in this section, and provides the financial metrics of each.

Scenario	Methanol Cost per MT	Annual Utility Cost (\$MM)	Annual Waste Treatment Cost (\$MM)	NPV (\$MM)	IRR	DPBP (years)
Base Case	\$400	95.52	71.99	-2855	NA	NA
Scenario One	\$265	81.19	61.19	-534	NA	NA
Scenario Two	\$212	81.19	61.19	114	13.6%	14

Table 7. Summary of each economic scenario including assumptions, NPV, IRR, and DPBP

Reflecting on each of these scenarios, our team does not recommend that MichiChem continue the development of this process in its current state. The high costs of raw materials, utilities, and waste treatment lead to the process losing money each year it operates. As demonstrated in Scenario Two, the

process becomes financially attractive if costs are sufficiently reduced. However, as currently designed, the process remains unprofitable.

Green Methanol. In a future where green methanol becomes profitable, tax incentives for the production of green fuels and chemicals could further reduce taxes and increase the yearly profit after taxes, thus increasing the NPV further. Although China currently does not have any tax incentives for green methanol production and use, the United States, under the Clean Fuel Production Credit, offers a tax credit of \$0.20 per gallon of clean fuel produced [41]. Current research indicates that MTO using methanol produced with blue hydrogen costs a minimum of \$3.19 per kg of ethylene produced, while methanol produced with green hydrogen costs a minimum of \$3.25 per kg of ethylene produced, and methanol produced via gray hydrogen costs just \$1.48 per kg of ethylene produced [42]. Thus, while green methanol could reduce global warming potential from 2.15 kg CO₂-equivalents per kg ethylene to -2.3 kg CO₂-equivalents per kg ethylene, it is not currently profitable. It should, however, be monitored and further researched in the future as technologies such as proton exchange membrane water electrolysis continue to produce hydrogen at lower costs, thus reducing the price of hydrogen for methanol production.

CONCLUSION

As current industrial practice for the production of ethylene and propylene involves carbon-intensive petrochemical processes, interest in alternative synthetic routes to these compounds has increased in recent history. The production of light olefins from methanol provides an industrially viable alternative to typical non-renewable processes to produce petrochemical feedstocks.

The primary goals of this project are to design a methanol-to-olefins (Mto) production facility, determine the optimal production capacity and plant location, and conduct a financial analysis to advise MichiChem on whether to move forward with process development. Key deliverables include a conceptual design for a facility capable of producing our specified capacity of olefins, along with various hydrocarbon byproducts; a detailed process design outlining process operating conditions; and an economic feasibility report evaluating the overall financial viability of the plant.

The proposed process feeds methanol into a fast-fluidized bed reactor, where it contacts a SAPO-34 catalyst, and is converted to target compounds ethylene and propylene, as well as many other straight chain hydrocarbon byproducts. This reaction also generates coke on the catalyst, heavy hydrocarbons that do not enter the vapor phase. The catalyst is recycled to a regeneration unit, where it is mixed with air at a high temperature to burn the coke off of the catalyst, and the fresh catalyst is recycled back to the reactor. The vapor effluent leaving the reactor goes through a series of quench towers, heat exchangers, absorption columns, and a caustic scrubber to remove water, carbon dioxide, and unreacted methanol from the process stream. The process stream is then fed through a series of compressors and heat exchangers to condense the process stream, which is then fed through a series of distillation columns to separate methane, ethane, propane, ethylene, propylene, butylene, and heavier hydrocarbons into individual streams.

General EHS considerations revolve around the properties of the materials handled and the conditions at which the process equipment is operated. Because methanol and all of the hydrocarbons produced in the process are flammable, the entire production area is electrically classified as Class 1, Div. 1. There are emergency shutoff switches both in the field around major pieces of production equipment, as well as in the control room through the DCS. Should one of these switches be flipped, the entire process will enter its fail-safe state, and the local fire department will be notified. All on-site personnel will be required to complete training detailing emergency procedures, and will not be able to enter the production area unless completed. Hot work permits must be obtained in order to perform hot work anywhere in the production facility, and dip tubes will be installed on all storage tanks without a floating roof to minimize the possibility of static charge buildup.

A wide range of operating conditions are employed at the proposed facility. Equipment operates at over 20 times ambient pressure, and temperatures range from -101 to 730°C. To avoid overpressuring any part of the process, PSVs will be installed on all pressurized vessels and lines. Moreover, rupture discs will be installed on all pressurized vessels to relieve any pressure buildup that occurs. All PSVs and rupture discs will be routed to the DCS, where operators will be alerted if either piece of equipment is activated, and the area of the process where the failure occurred will be switched into its fail-safe state. To avoid injuries relating to contact of hot or cold equipment, all relevant pipes and vessels will be insulated.

This report has also provided an in-depth financial analysis of the process, which concludes that the process is not currently economically viable, with a NPV of -\$2.86 BN. The high costs of raw materials, utilities, and waste treatment drive annual operating expenses above annual revenue, which results in annual financial losses. We have presented scenarios involving alternative raw material sourcing and process optimization that could yield higher NPVs, potentially becoming positive. However, until these cost reductions are achieved, we do not recommend that MichiChem continues with the development of this process.

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LIST OF APPENDICES

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APPENDIX A – PROCESS FLOW DIAGRAMS.

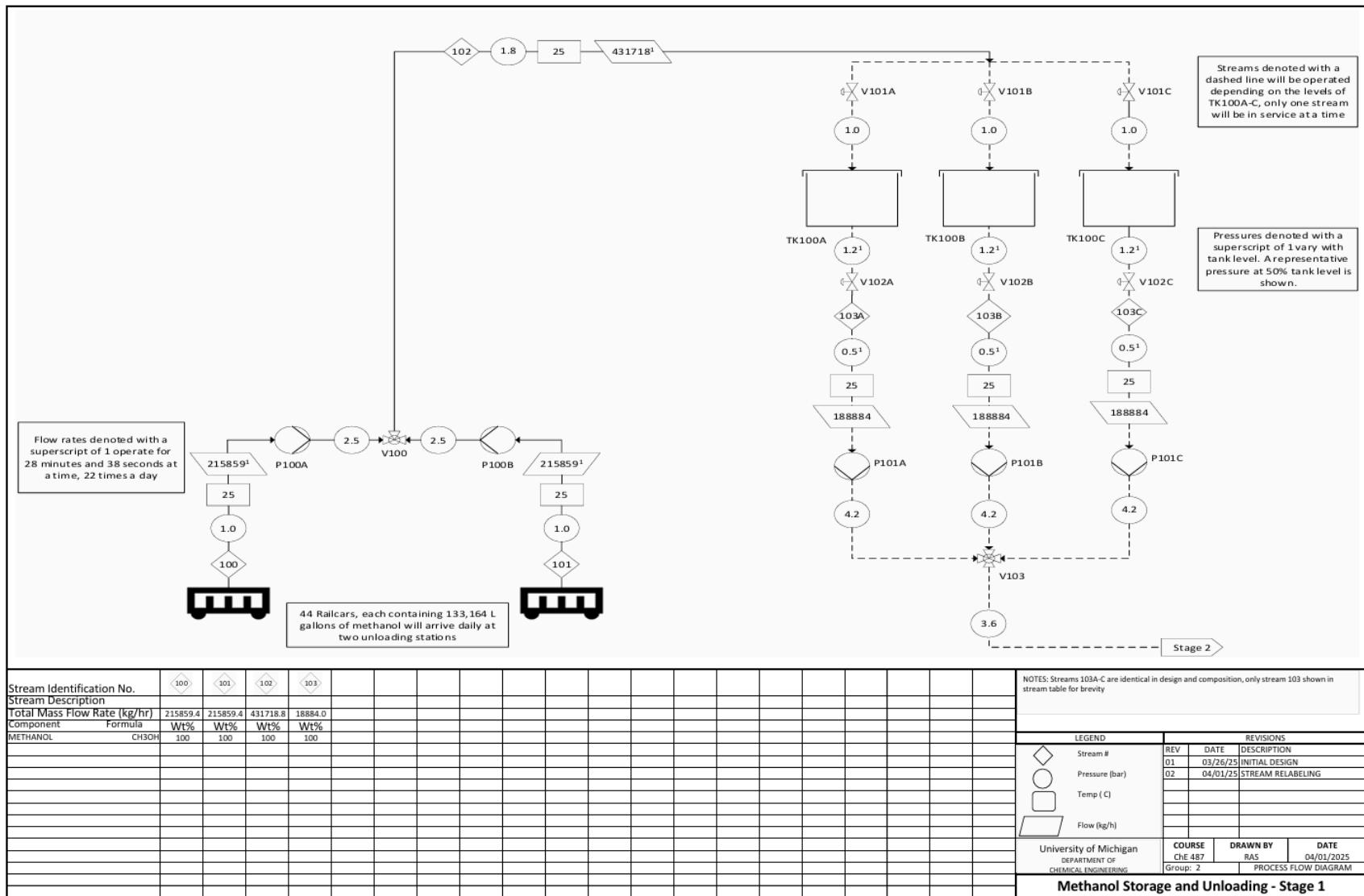
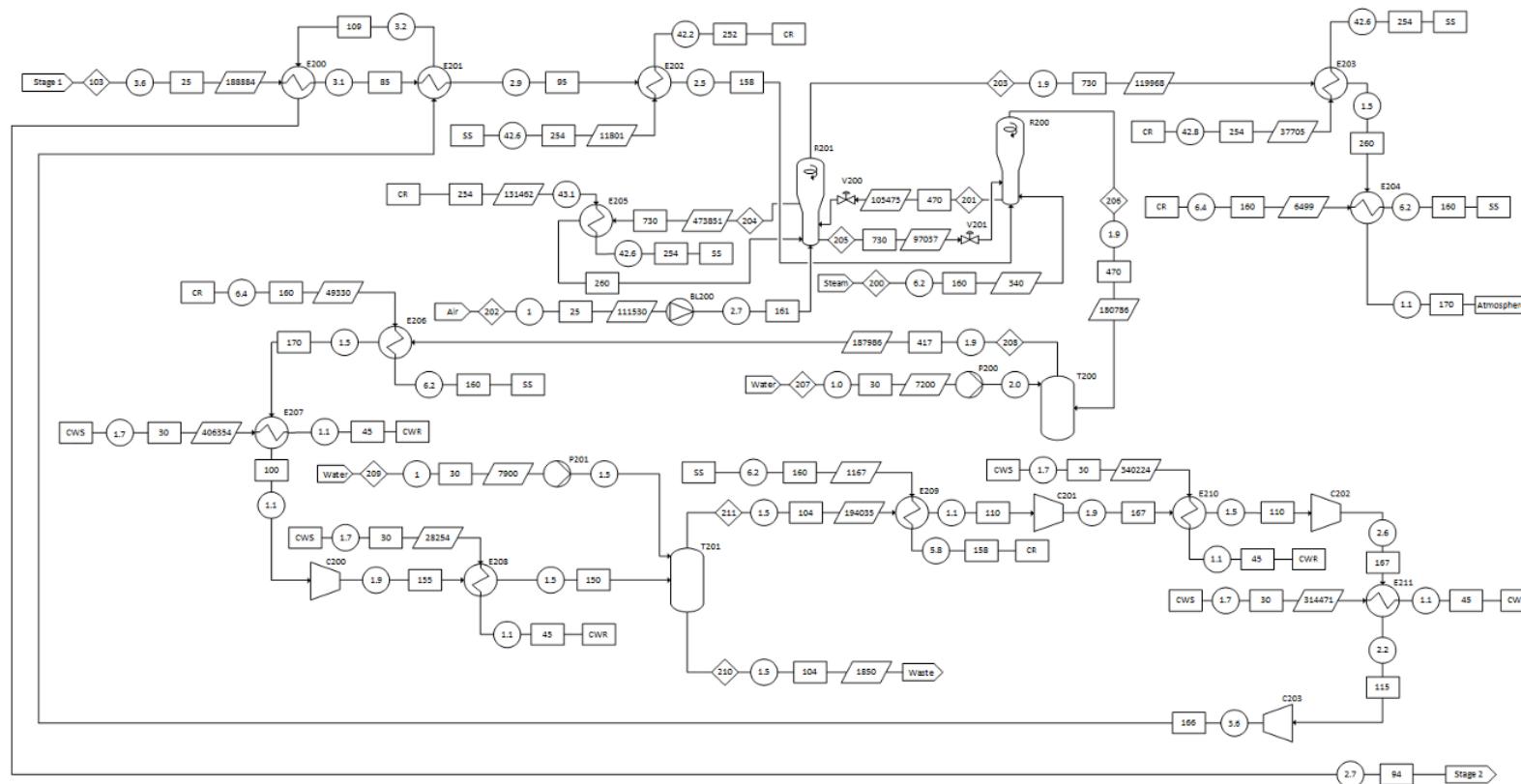


Figure A-1. Stage 1 PFD



Stream Identification No.	103	200	201	202	203	204	205	206	207	208	209	210	211						
Stream Description																			
Total Mass Flow Rate (kg/hr)																			
Component	Formula	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%	Wt%
METHANOL CH ₃ OH		100.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0	1.0	0.0	0.1	0.9					
METHANE CH ₄		0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0	1.0	0.0	trace	0.9					
ETHANE C ₂ H ₆		0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0	1.0	0.0	trace	0.9					
PROPANE C ₃ H ₈		0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	0.0	1.0	0.0	trace	0.9					
C ₅₊ N/A		0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.0	0.0	1.9	0.0	trace	1.9					
ETHYLENE C ₂ H ₄		0.0	0.0	0.0	0.0	0.0	0.0	0.0	16.9	0.0	16.3	0.0	4 ppm	15.8					
PROPYLENE C ₃ H ₆		0.0	0.0	0.0	0.0	0.0	0.0	0.0	16.9	0.0	16.3	0.0	2 ppm	15.8					
WATER H ₂ O		0.0	100.0	0.0	0.0	4.1	0.0	0.0	55.0	100.0	56.7	100.0	99.9	58.1					
CARBON DIOXIDE CO ₂		0.0	0.0	0.0	0.0	24.0	0.0	0.0	0.1	0.0	0.1	0.0	trace	0.1					
BUTYLENE C ₄ H ₈		0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.0	0.0	4.8	0.0	trace	4.6					
NITROGEN N ₂		0.0	0.0	0.0	75.4	70.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0					
OXYGEN O ₂		0.0	0.0	0.0	23.2	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0					
ARGON Ar		0.0	0.0	0.0	1.4	1.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0					
SAPG-34		0.0	0.0	92.0	0.0	0.0	100.0	100.0	trace	0.0	trace	0.0	trace	0.0					
Coke C		0.0	0.0	8.0	0.0	0.0	trace	trace	0.0	0.0	0.0	0.0	0.0	0.0					

NOTES:

Stream #	LEGEND		
	Pressure (bar)	REVISIONS	
		REV	DATE
		3/9/2025	INITIAL DRAFT
		3/26/2025	DESIGN FREEZE
		4/4/2025	FINAL ADJUSTMENTS

University of Michigan
DEPARTMENT OF
CHEMICAL ENGINEERING

COURSE: CHE 487 DRAWN BY: ZRM DATE: 04/04/2025
Group: N/A PROCESS FLOW DIAGRAM

Reaction of Methanol to Hydrocarbons - Stage 2

Figure A-2. Stage 2 PFD

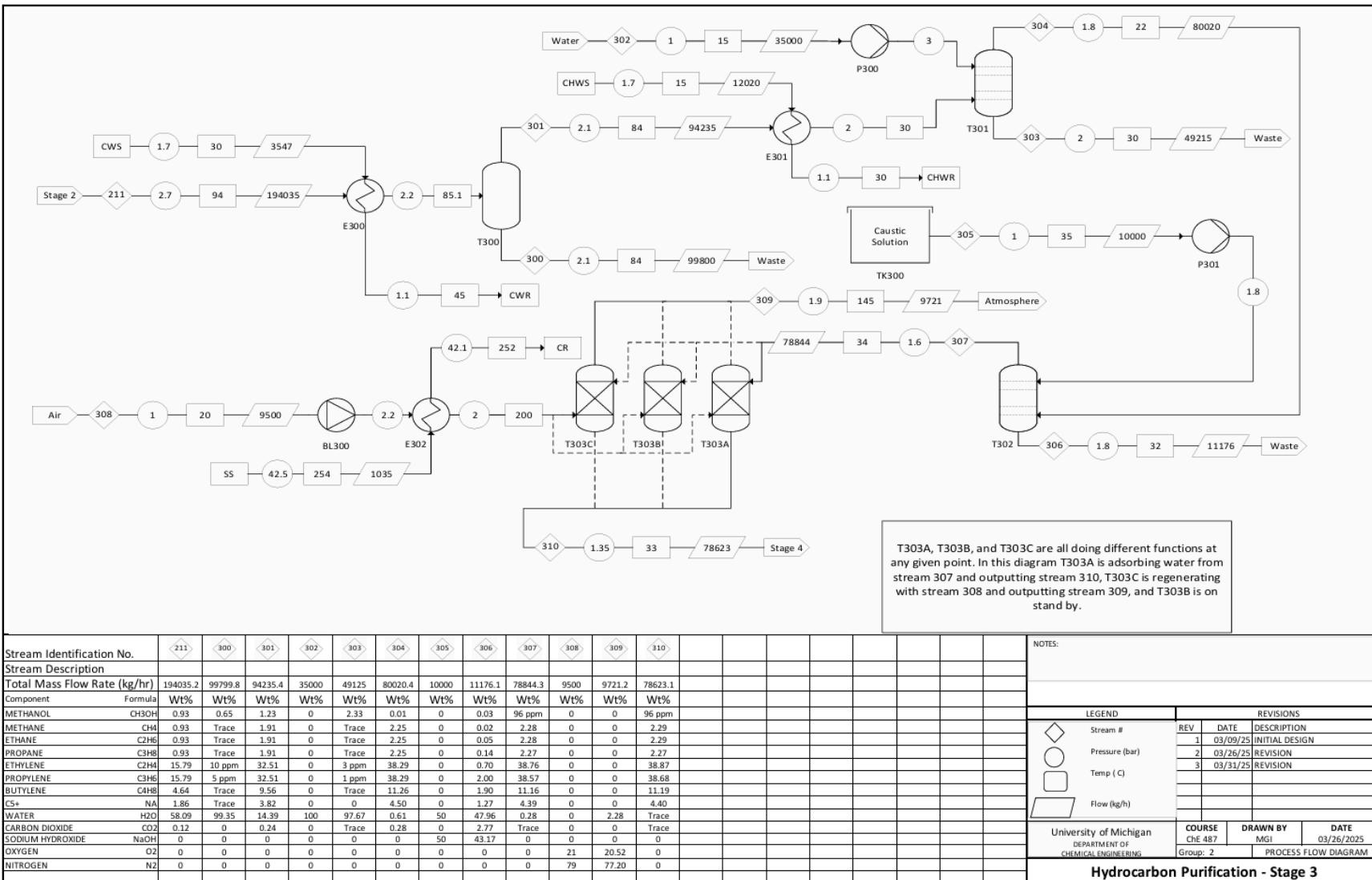


Figure A-3. Stage 3 PFD

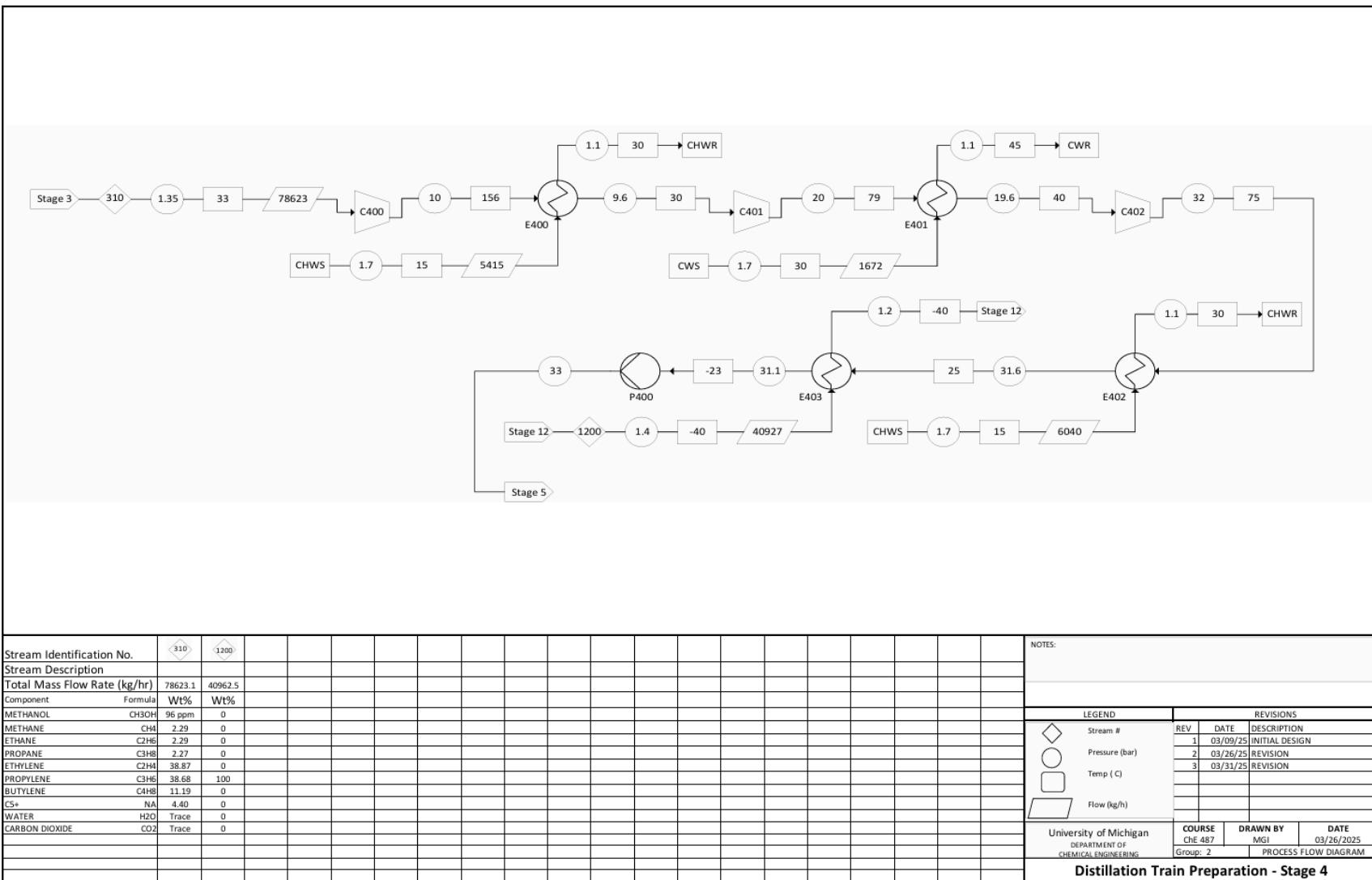


Figure A-4. Stage 4 PFD

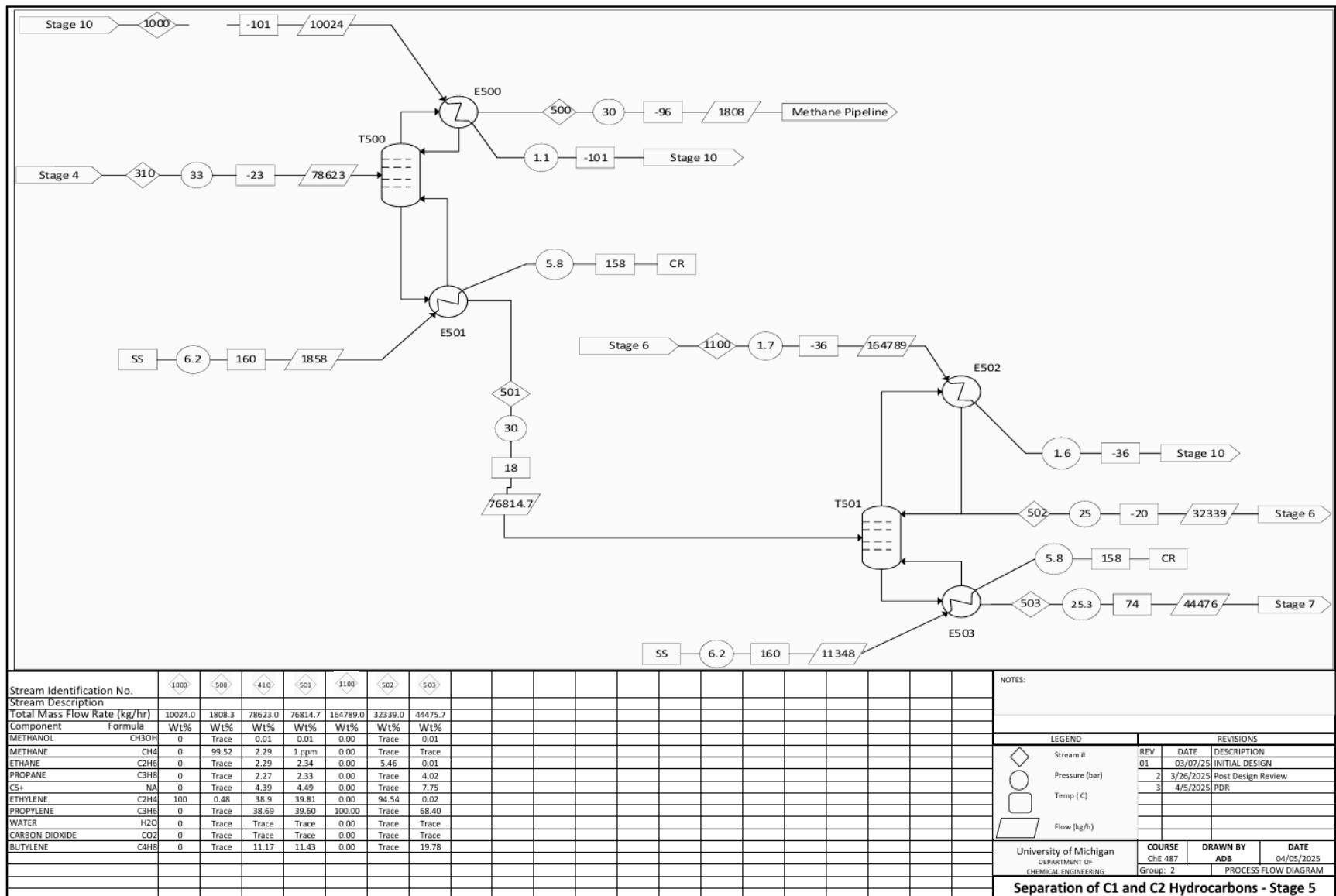


Figure A-5. Stage 5 PFD

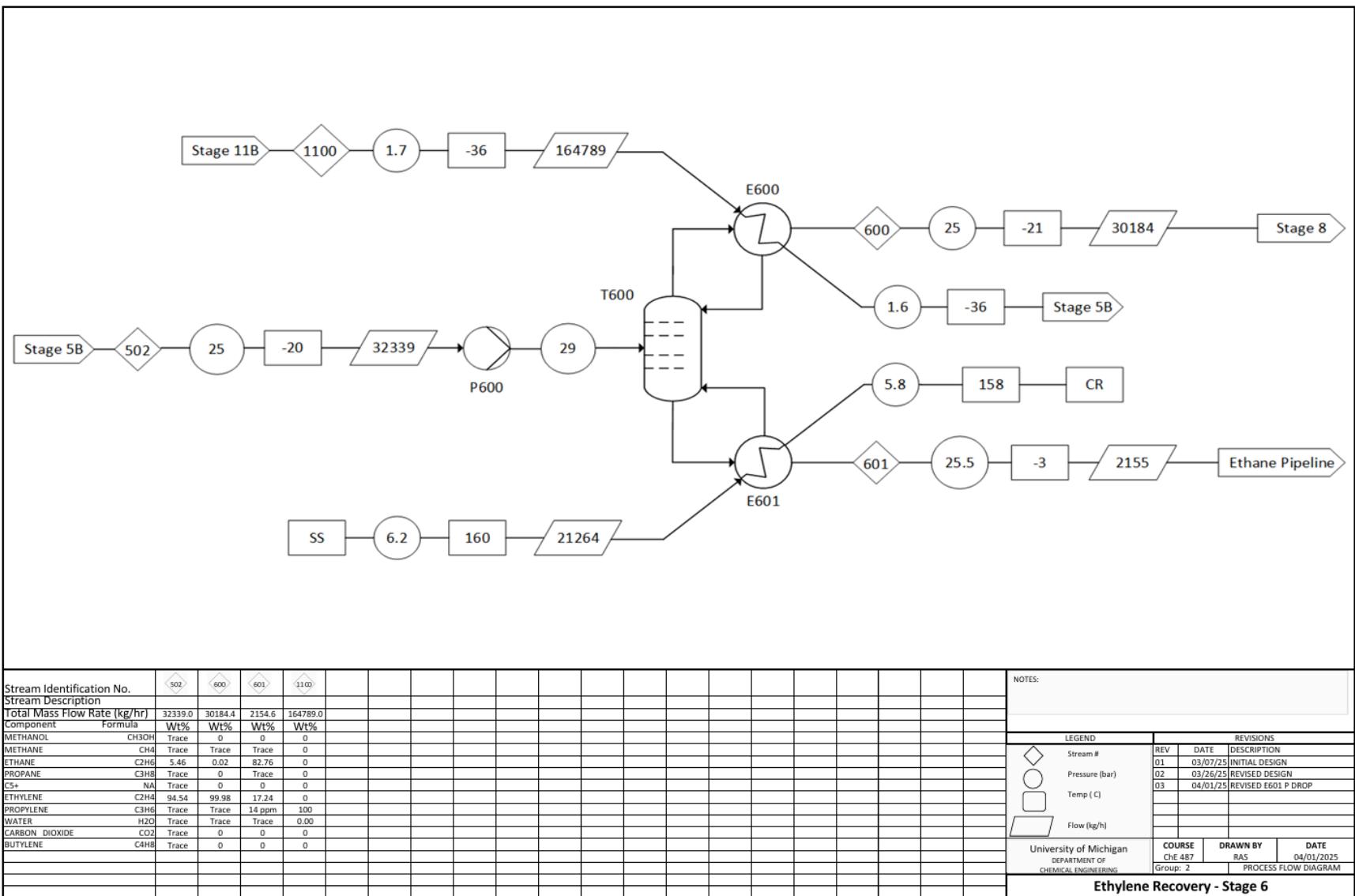


Figure A-6. Stage 6 PFD

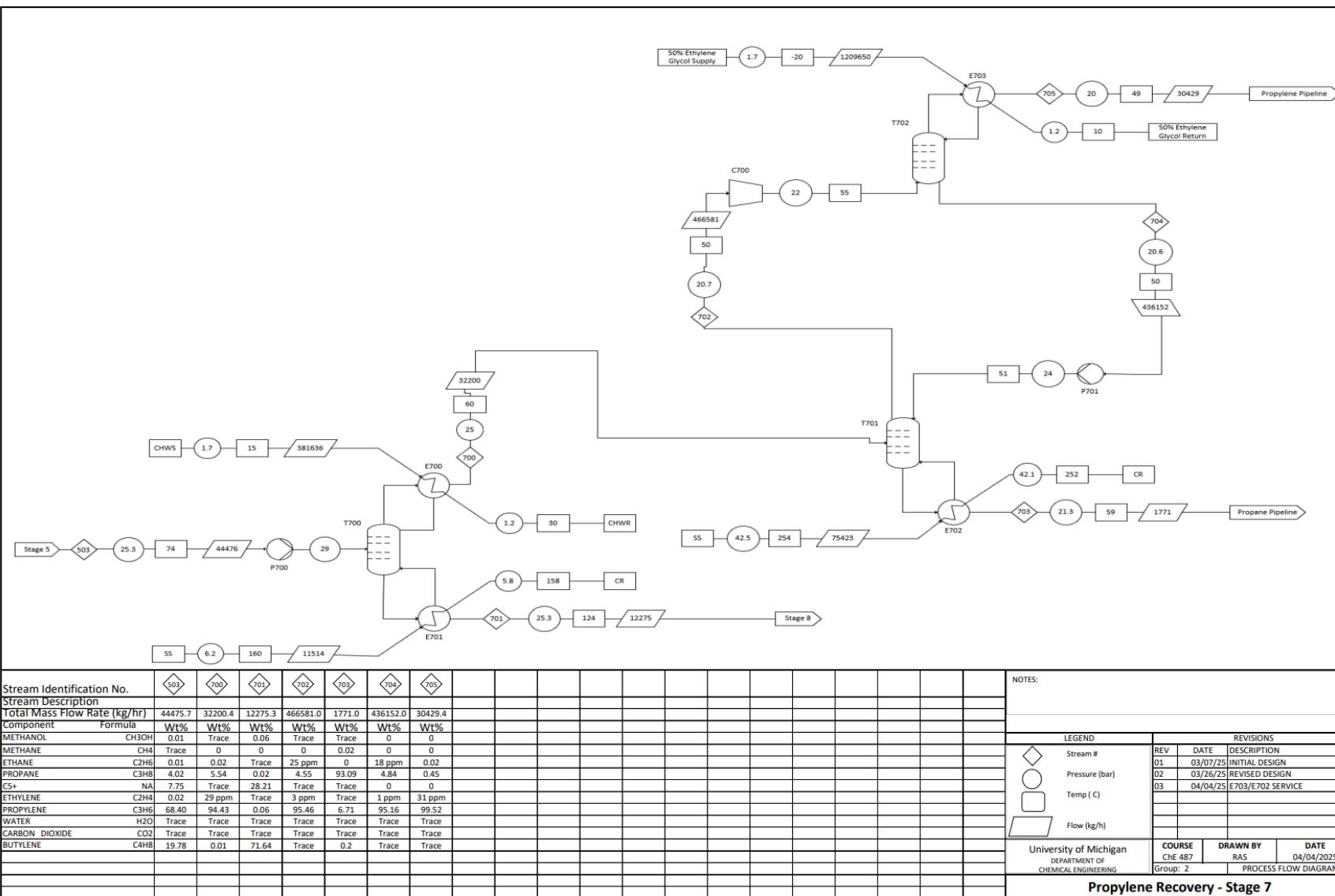


Figure A-7. Stage 7 PFD

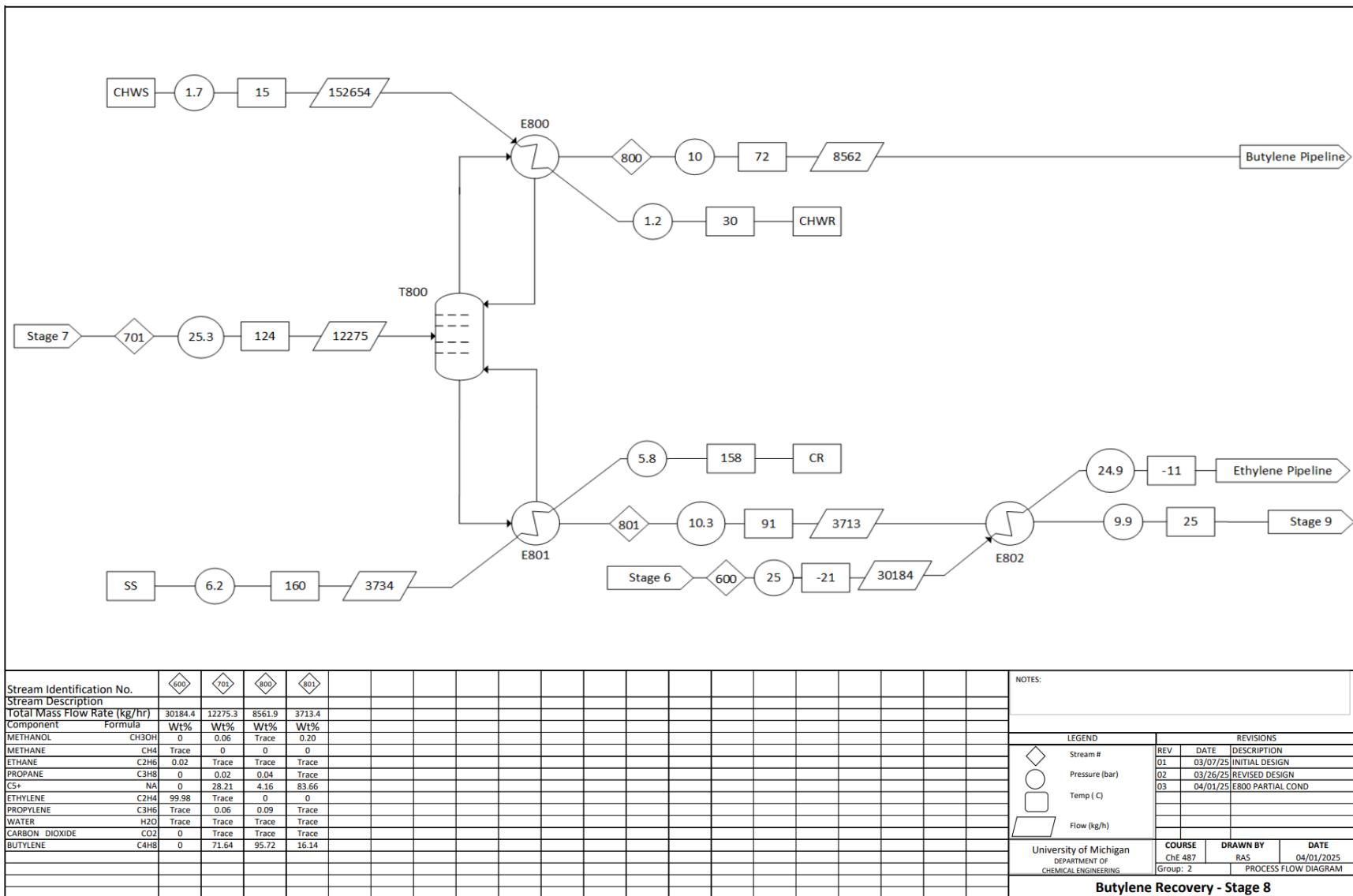


Figure A-8. Stage 8 PFD

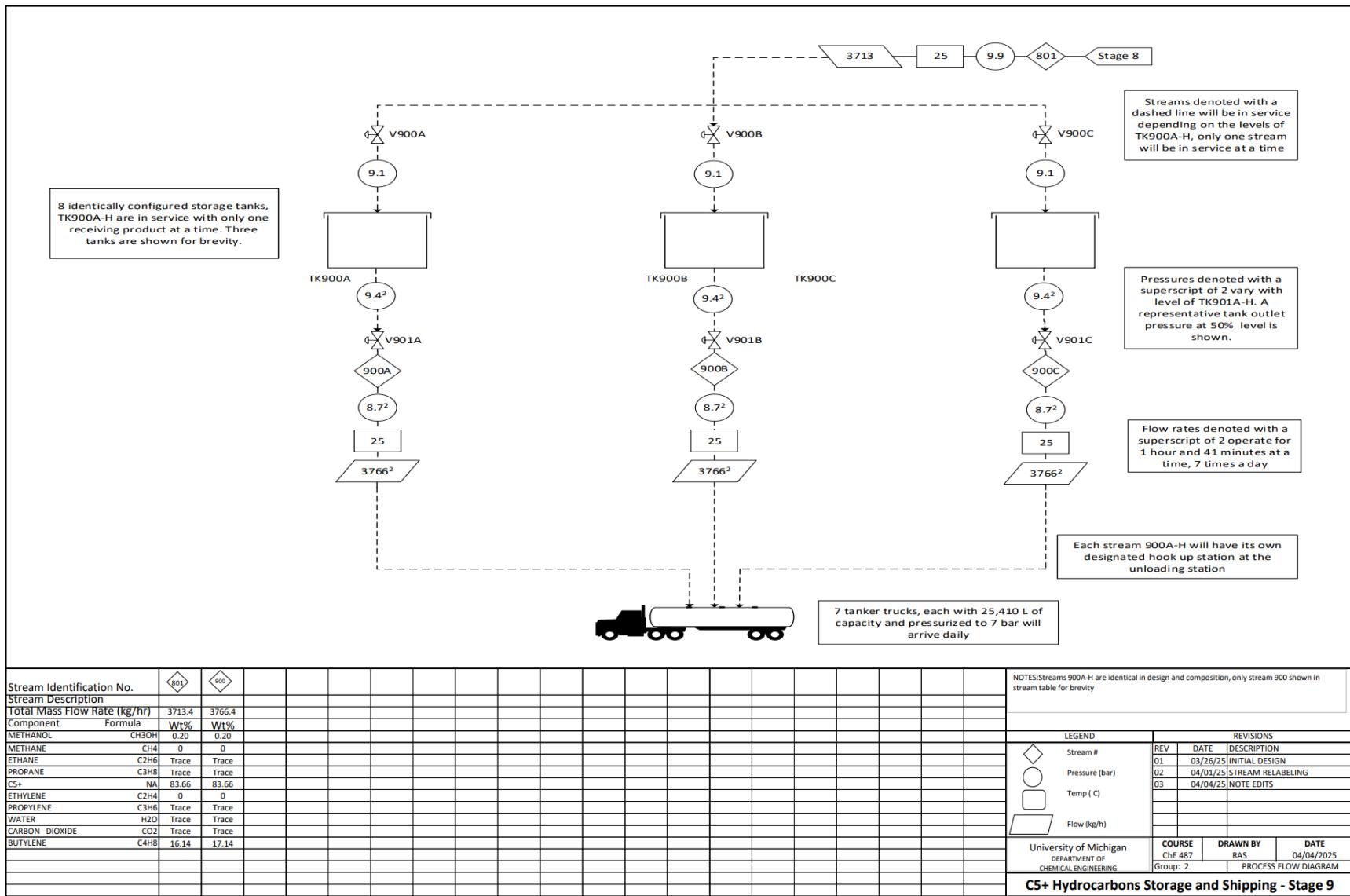


Figure A-9. Stage 9 PFD

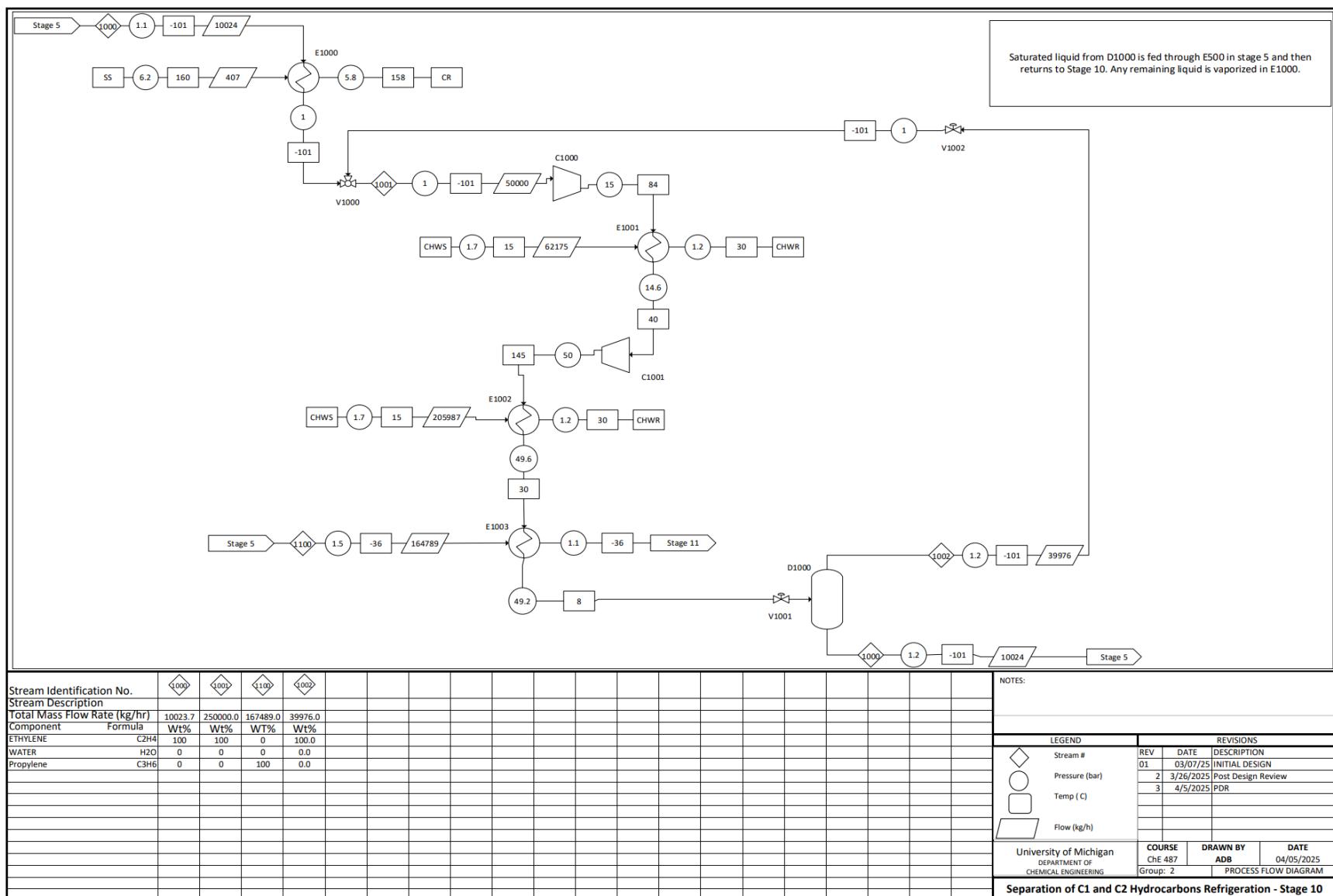


Figure A-10. Stage 10 PFD

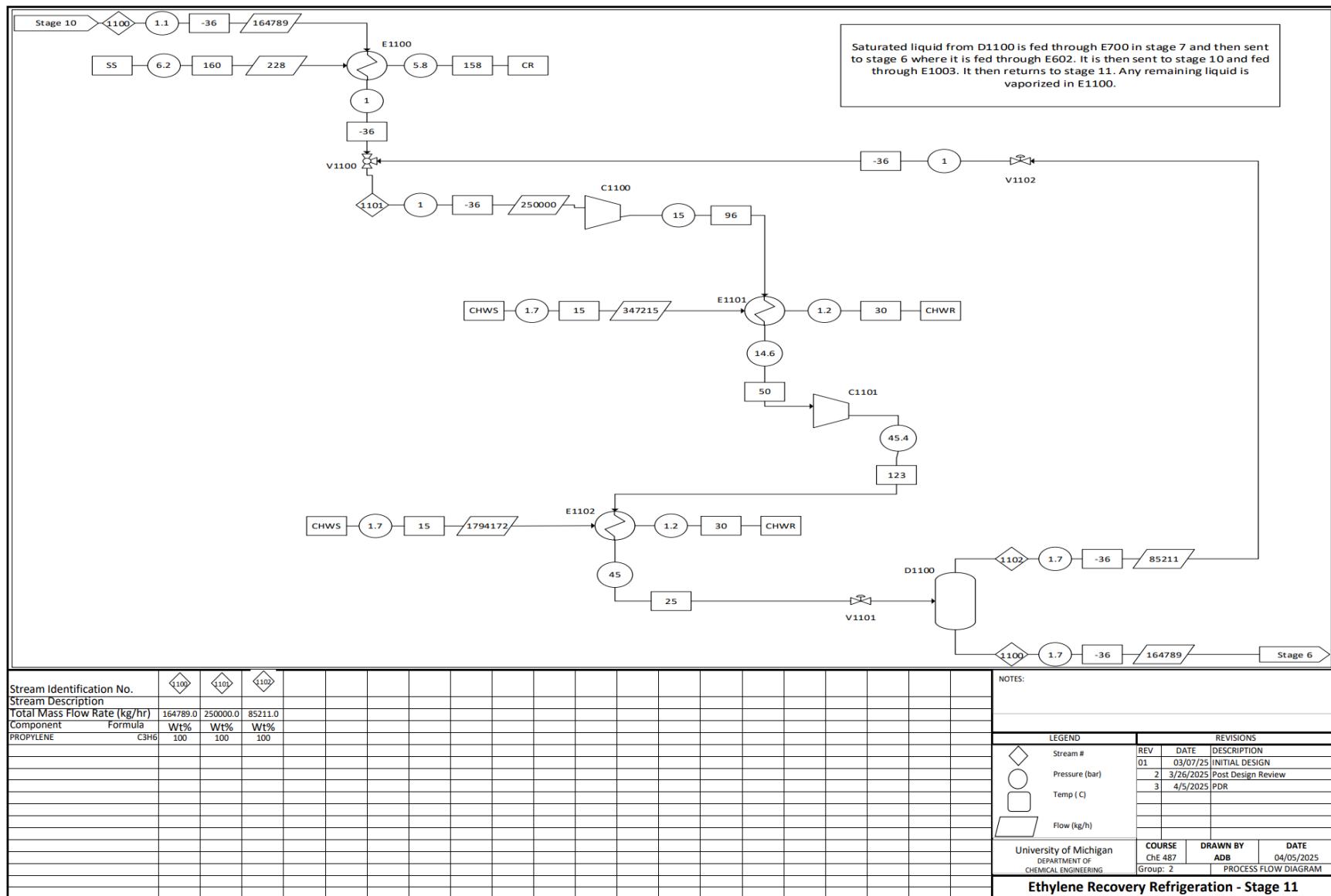


Figure A-11. Stage 11 PFD

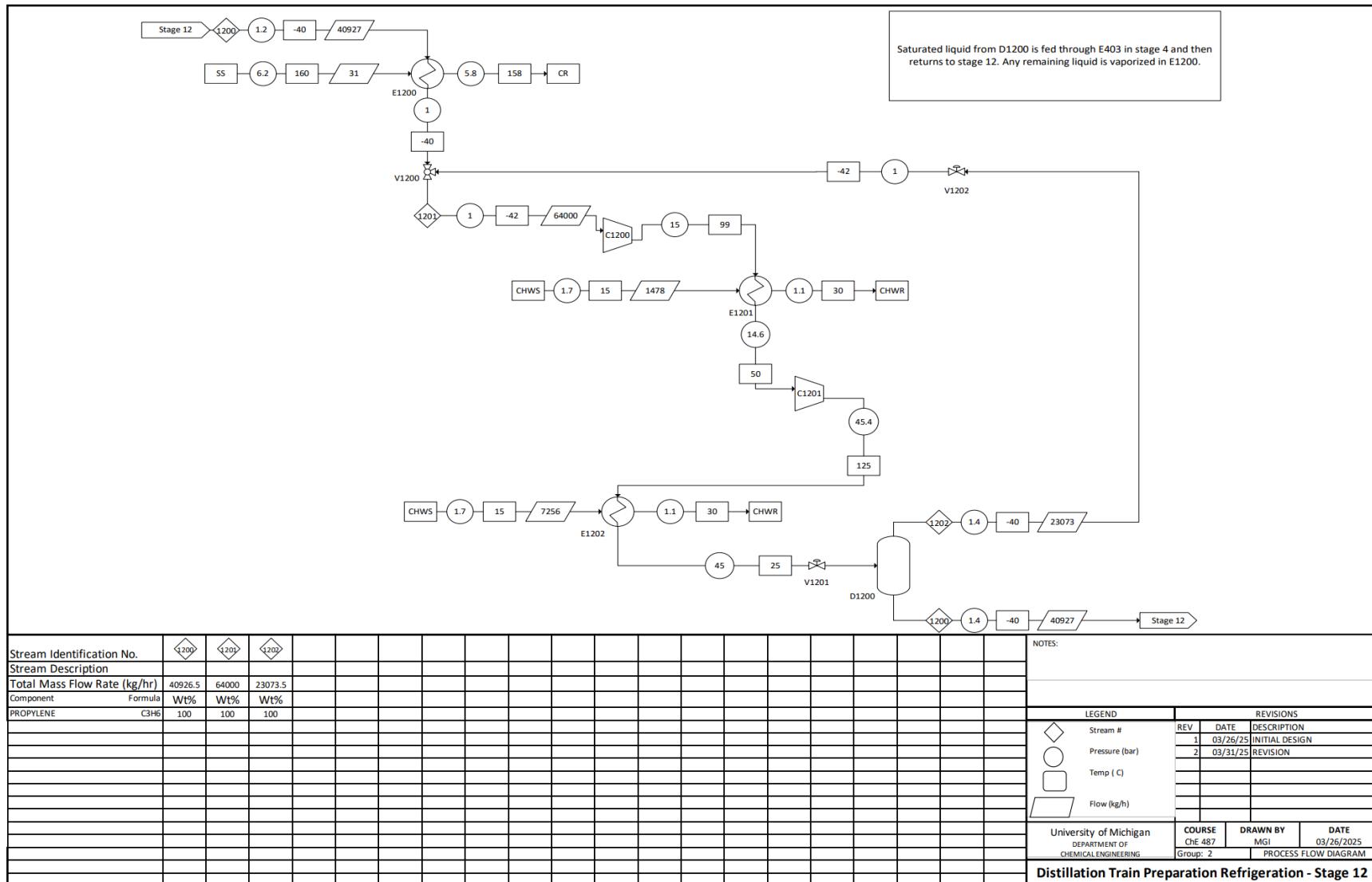


Figure A-12. Stage 12 PFD

APPENDIX B – PLANT LAYOUT

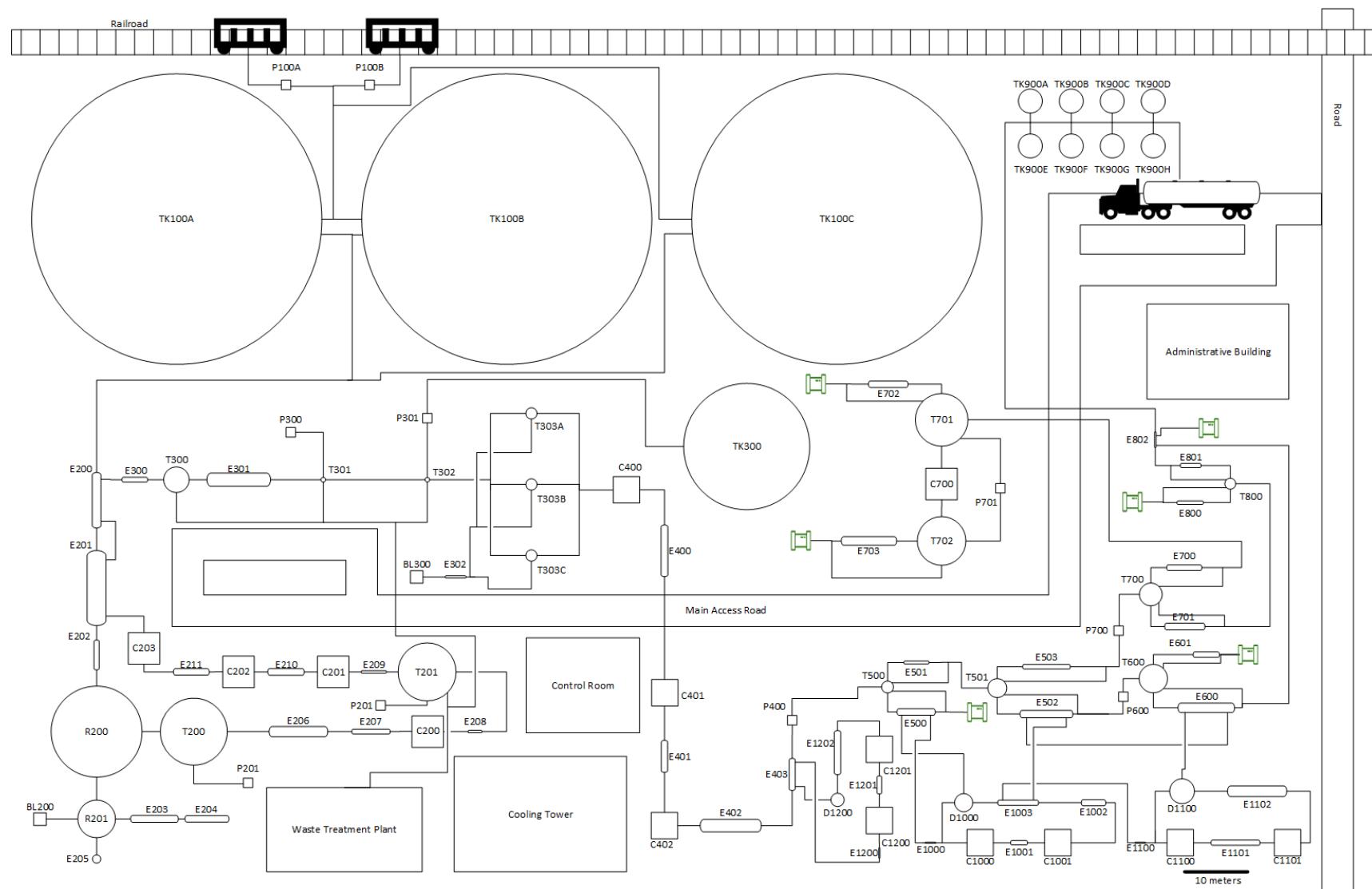


Figure B-1. Plant layout

APPENDIX C – SAMPLE CALCULATIONS

The following sections provide sample calculations for our sizing process for various process units, our material and energy balances around the reactor and regenerator system, and utility requirements. Additionally, each section provides a table summarizing the key sizing parameters for each type of process unit.

Reactor and Regenerator

This section provides calculations for the sizing, material balances, and energy balances around the reactor/regenerator system.

Sizing. To estimate the dimensions of the reactor, we found a superficial velocity v_{sp} from the literature and calculated the cross-sectional area needed using the volumetric flow rate V of our fluid, as given by the following equation:

$$A = \frac{V}{v_{sp}}$$

The volumetric flow rate was derived from ASPEN using the Peng Robinson base method. The remaining dimensions, including lengths and diameters of additional zones, were calculated based on dimensions of existing reactors.

The superficial velocity of the fluid in the dense phase zone of the reactor is typically about 1 m/s [43]. Because the methanol quickly increases in temperature due to heat transfer from the catalyst and the reaction is fast and exothermic, the volumetric flow rate of the fluid was chosen as the effluent leaving the reactor at 254,377 m³/h. The effluent's composition is based on a product distribution given by [13]. Therefore, the reactor cross-sectional area was determined by:

$$A = \frac{V}{v_{sp}} = \frac{254,377 \text{ m}^3/\text{h}}{1 \text{ m/s}} = 70.7 \text{ m}^2$$

The dense phase diameter D_{dense} could thus be calculated from the area:

$$D_{dense} = 2\sqrt{\frac{A}{\pi}} = 2\sqrt{\frac{70.7 \text{ m}^2}{\pi}} = 9.49 \text{ m}$$

To keep the area of the disengaging zone proportional to that of the dense phase, the ratio of the squares of the diameters of an example reactor should be kept constant. For UOP's 5.4 m dense phase diameter, it employed a 7.9 m disengaging zone diameter. Therefore:

$$D_{disengaging} = \frac{D_{dis}^2}{D_{den}^2} D_{dense} = \frac{(7.9 \text{ m})^2}{(5.4 \text{ m})^2} 9.49 \text{ m} = 14.4 \text{ m}$$

Although UOP lacked data for other dimensions, a similar process was employed using an example reactor from Wang et. al [44], pictured in **Figure C-1**.

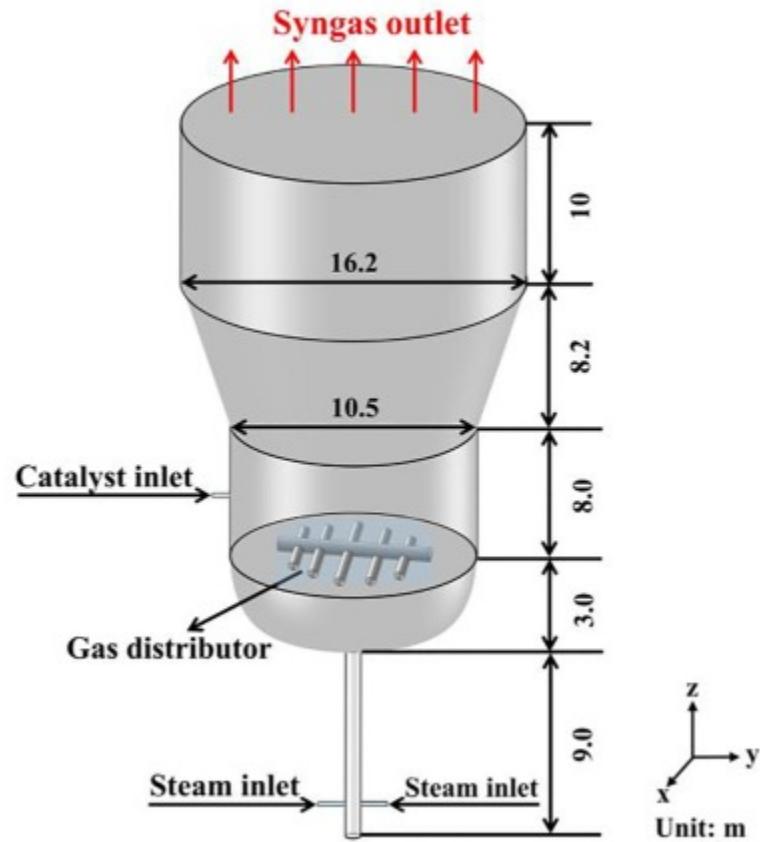


Figure C-1: Model reactor used for dimensioning

For the direction of flow, lengths were assumed to be proportional to the dense phase diameter and to one another. The reactor dimensions are summarized in **Table C-1**.

Item	Height (m)	Diameter (m)
Dense Phase (Reaction) Zone	7.23	9.49
Transition Zone	7.41	9.49-14.4
Disengaging Zone	9.04	14.4

Table C-1: Dimensions of reactor based on superficial velocity and reactors in literature

For the regenerator, an existing regenerator using similar conditions to ours already existed, so we used the dimensions of Chen's model, shown in **Figure C-2** [45].

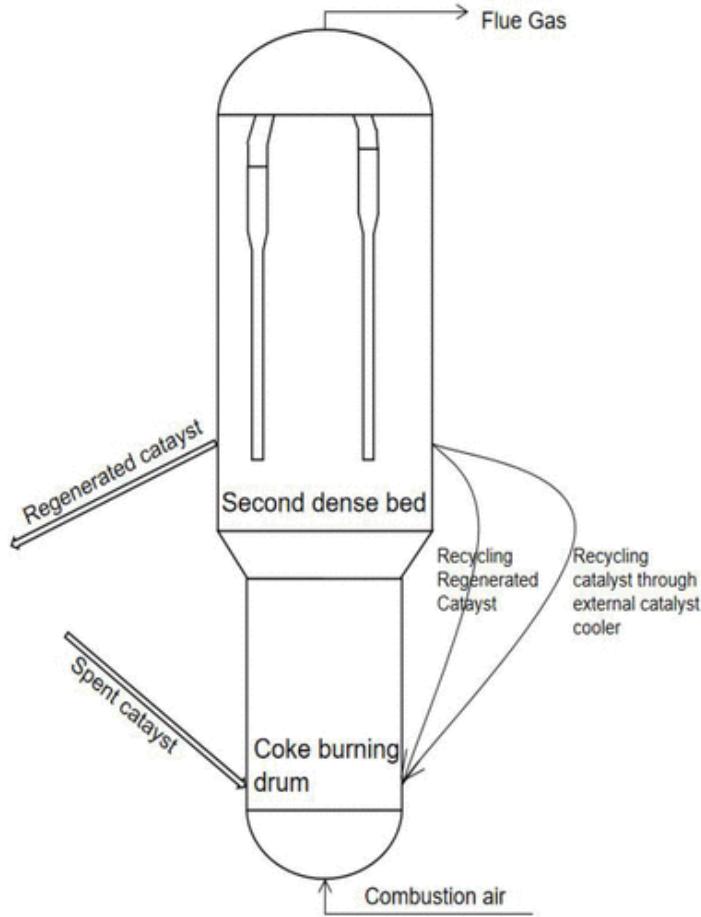


Figure C-2: Model regenerator used for dimensioning

The model was a two-stage regenerator with 50,000-60,000 m³ air, catalyst flow rate from the reactor of 105-150 kg/h, and catalyst flow rate from the catalyst cooler of 400,000-600,000 kg/h, whereas our flow rates of air, catalyst from reactor, and catalyst from cooler were 51,532 m³/h, 97,037 kg/h, and 473,851 kg/h, respectively. The regenerator dimensions are summarized in **Table C-2**.

Item	Height (m)	Diameter (m)
Coke Burning Drum	14.4	4.66
Second Dense Phase Zone	4.00	5.96

Table C-2: Dimensions of regenerator based on superficial velocity and regenerator in literature

Material Balances. To determine the heat requirement on the catalyst cooler, an energy balance was performed on the reactor/regenerator system. However, before the energy balance was conducted, a material balance was made on both the regenerator and reactor so that appropriate thermodynamic data from ASPEN could be gathered. For the reactor, we took a product distribution given by Kiss et al. for reaction occurring at 470°C [13]. We assumed a carbon dioxide mass fraction of 0.0012 in the outlet stream and added additional methanol to balance [46]. We also added water brought in to the system by the stripping steam. For the regenerator, we added enough air to combust all the coke with 2% excess oxygen to prevent afterburn in the cyclones [47]. We assumed the coke had a carbon to hydrogen (C:H) molar ratio of 1.2:1, as the major components of coke at a temperature of 450°C are pyrene (1.4:1) and phenanthrene (1.6:1), but other smaller hydrocarbons with smaller C:H ratios also are formed. To determine the air flow rate, we used stoichiometric properties: a C:H ratio of 1.2:1 corresponds to a mass fraction of 0.935 kg C/kg coke, and we assumed 21 mol% O₂ in the air and a molar mass of 28.96 kg/kmol for air.

$$8438 \text{ kg/h coke} * \frac{0.935 \text{ kg C}}{\text{kg coke}} * \frac{\text{kmol C}}{12.011 \text{ kg C}} * \frac{\text{kmol O}_2}{\text{kmol C}} * \frac{\text{kmol air}}{0.21 \text{ kmol O}_2} * \frac{28.96 \text{ kg air}}{\text{kmol air}} = 90584 \text{ kg/h}$$

$$8438 \text{ kg/h coke} * \frac{0.065 \text{ kg H}}{\text{kg coke}} * \frac{\text{kmol H}}{1.008 \text{ kg H}} * \frac{\text{kmol O}_2}{4 \text{ kmol H}} * \frac{31.998 \text{ kg O}_2}{\text{kmol O}_2} * \frac{\text{kg air}}{0.21 \text{ kg O}_2} = 18759 \text{ kg/h}$$

$$m_{\text{air}} = 1.02 * m_{\text{stoich}} = 1.02 * (90583.9 \text{ kg/h} + 18759.1 \text{ kg/h}) = 111530 \text{ kg/h}$$

Assuming the balance of air was made up of nitrogen (78 mol %) and argon (1 mol %), and assuming complete combustion of C in coke to carbon dioxide and H to water, the outlet flow rates are:

$$N_2: 111530 \text{ kg/h air} * \frac{\text{kmol air}}{28.96 \text{ kg air}} * \frac{0.78 \text{ kmol N}_2}{\text{kmol air}} * \frac{28.02 \text{ kg N}_2}{\text{kmol N}_2} = 84170 \text{ kg/h}$$

$$Ar: 111530 \text{ kg/h air} * \frac{\text{kmol air}}{28.96 \text{ kg air}} * \frac{0.01 \text{ kmol Ar}}{\text{kmol air}} * \frac{39.95 \text{ kg Ar}}{\text{kmol Ar}} = 1539 \text{ kg/h}$$

$$O_2: (111530 - 109343) \text{ kg/h air} * \frac{\text{kmol air}}{28.96 \text{ kg air}} * \frac{0.21 \text{ kmol O}_2}{\text{kmol air}} * \frac{31.998 \text{ kg O}_2}{\text{kmol O}_2} = 507 \text{ kg/h}$$

$$CO_2: 8438 \text{ kg/h coke} * \frac{0.935 \text{ kg C}}{\text{kg coke}} * \frac{\text{kmol C}}{12.011 \text{ kg C}} * \frac{1 \text{ kmol CO}_2}{\text{kmol C}} * \frac{44.009 \text{ kg CO}_2}{\text{kmol CO}_2} = 28908 \text{ kg/h}$$

$$H_2O: 8438 \text{ kg/hh coke} * \frac{0.065 \text{ kg C}}{\text{kg coke}} * \frac{\text{kmol H}}{1.008 \text{ kg H}} * \frac{1 \text{ kmol H}_2O}{2 \text{ kmol H}} * \frac{18.015 \text{ kg H}_2O}{\text{kmol H}_2O} = 4901 \text{ kg/h}$$

The masses add up to 120,026 kg/h, a 0.048% increase from the expected 119,968 kg/h obtained from summing the catalyst and air flow rates and likely due to rounding in the coke composition. The air flue gas flow rate was set to 119,968 kg/h assuming the same composition of gases as calculated through stoichiometry.

To determine catalyst flow rates and inventories, we assumed the catalyst contained 8% coke by mass entering the reactor (assumed to be optimal for kinetics) and the catalyst leaving the reactor was completely pure [6]. Thus,

$$\frac{m_{coke}}{m_{coke} + m_{cat}} = 0.08$$

$$m_{cat} = \frac{0.92}{0.08} * m_{coke} = 11.5 * 8438 \text{ kg/h} = 97037 \text{ kg/h}$$

Furthermore, the catalyst residence time is typically around 1 hour in the reactor [6]. Thus, the inventory in the reactor is 97,037 kg. Finally, the ratio of catalyst in the regenerator to the reactor is typically between 0.1 and 0.9 and preferably between 0.25 and 0.7 [48]. Since the regenerator requires a significant amount of cooling, 0.7 was chosen as the desired ratio. Therefore, the regenerator contains 67,926 kg catalyst.

Stripping steam, necessary for removing products from the catalyst, is typically provided in 2-5 lb per 1000 lb circulating catalyst in fluid catalytic cracking (FCC) units, which operate similarly to MtO units [49].

$$m_{steam} = 0.0035 * m_{cat} = 0.0035 * 97037 \text{ kg/h} = 340 \text{ kg/h}$$

Finally, due to catalyst attrition, the catalyst will need to be replaced in the reactor system. In FCC systems, for 30,000 tons/day catalyst cycled through the system, about 2.2 tons/day need to be replaced [50].

$$m_{replacement} = \frac{2.2 \text{ tons/day}}{30,000 \text{ tons/day}} * 97037 \text{ kg/h} = 7.1 \text{ kg/h}$$

A summary of the flow rates and catalyst inventory is listed in **Table C-3**.

Parameter	Value
Methanol In	188884 kg/h
Water In	340 kg/h
Air In	111530 kg/h

Effluent Out	119968 kg/h
Flue Gas Out	180786 kg/h
Reactor Coke to Regenerator	8438 kg/h
Reactor Catalyst to Regenerator	97037 kg/h
Reactor Catalyst Inventory	97037 kg
Regenerator Catalyst Inventory	67926 kg
Catalyst Replacement Rate	7.1 kg/h

Table C-3: Mass flow rates for reactor/regenerator system

Energy Balances. Energy balances around the reactor and regenerator can be made using the following equations, where Q is the net heat, m_i represents the mass flow rate of species i , and H_i represents the specific enthalpy of species i :

$$Q = (\sum m_{prod} H_{prod} + m_{coke} H_{coke} + m_{cat} H_{cat,out}) - (m_{methanol} H_{methanol} + m_{steam} H_{steam} + m_{cat} H_{cat,in})$$

$$Q = (\sum m_{flue} H_{flue} + m_{cat} H_{cat,out}) - (\sum m_{air} H_{air} + m_{coke} H_{coke} + mH_{cat,in})$$

Assuming catalyst, coke, and products leaving the reactor exit at 470°C, steam enters the reactor at 160°C, catalyst enters the reactor at 730°C, air entering the regenerator enters at 161°C, and flue gas leaves the regenerator at 730°C, we can calculate the enthalpy flows of the product vapor, steam feed, air feed, and flue gas using ASPEN. Assuming the catalyst has a specific heat capacity Cp_{cat} of 1 kJ/kgK, we can rewrite

$m_{cat} H_{cat,out} - m_{cat} H_{cat,in}$ as $m_{cat} Cp_{cat} \Delta T_{cat}$ [51]. Assuming coke is similar to graphitic carbon, the enthalpy flow of coke is

$H^0 + m_{coke} Cp_{coke}(T - T^0)$, where H^0 is 0, Cp_{coke} is 0.85 kJ/kgK, and T^0 is 298 K [52]. Assuming the reactor does not contain any cooling elements, we can solve for $m_{methanol} H_{methanol}$ to determine the inlet methanol temperature where Q is 0. This temperature is 158°C.

$$Q = 0 = (\sum m_{prod} H_{prod} + m_{coke} H_{coke} + m_{cat} Cp_{cat} \Delta T_{cat}) - (m_{methanol} H_{methanol} + m_{steam} H_{steam})$$

$$m_{methanol} H_{methanol} = (\sum m_{prod} H_{prod} + m_{coke} H_{coke} + m_{cat} Cp_{cat} \Delta T_{cat}) - m_{steam} H_{steam} =$$

$$- 315.005 MW + 8438 \text{ kg/h} * 0.85 \text{ kJ/kgK} * (470^\circ\text{C} - 25^\circ\text{C}) +$$

$$97037 \text{ kg/h} * 1 \text{ kJ/kgK} * (470^\circ\text{C} - 730^\circ\text{C}) + 1.790 MW = - 319.336 MW$$

Similarly, we can solve another energy balance in the regenerator to determine Q , the net cooling rate needed in the catalyst cooler. **Table C-4** indicates the enthalpy flows calculated both via ASPEN and by hand. Once given the cooling needed in the regenerator, we can calculate the catalyst flux in the catalyst cooler by the following equation, where ΔT_{cat} is the temperature difference between the catalyst entering at 730°C and the catalyst leaving the cooler at 260°C (being cooled by high pressure steam at 254°C):

$$Q = m_{cat,cool} C_p_{cat} \Delta T_{cat}$$

$$m_{cat,cool} = \frac{Q}{C_p_{cat} \Delta T_{cat}}$$

Parameter	Value
$\Sigma m_{prod} H_{prod}$	-315.10 MW
$m_{coke} H_{coke}$	0.89 MW
$m_{cat} C_p_{cat} \Delta T_{cat}$ (reactor)	-7.01 MW
$m_{cat} C_p_{cat} \Delta T_{cat}$ (regenerator)	7.01 MW
$m_{methanol} H_{methanol}$	-319.34 MW
$m_{steam} H_{steam}$	-1.79 MW
$\Sigma m_{flue} H_{flue}$	-63.73 MW
$\Sigma m_{air} H_{air}$	4.71 MW
$Q_{reactor}$	0.00 MW
$Q_{regenerator}$	-61.86 MW

$m_{cat,cool}$	473851 kg/h
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Table C-4: Enthalpy flows of streams entering and exiting the regenerator and reactor

Heat Exchangers

To estimate the size and dimensions of the heat exchangers, the required duty, Q , was calculated using ASPEN calculations in most cases. An appropriate overall heat transfer coefficient, U , as well as two fouling factors, F_{tube} and F_{shell} , were assumed based on the fluids in the exchanger. The corrected heat transfer coefficient, U_c , was calculated using the equation:

$$U_c = (U^{-1} + F_{tube} + F_{shell})^{-1}$$

The log-mean temperature ΔT_{lm} was given by the temperature difference at the inlet and outlet of the hot and cold ends by the following equation, where T_1 is the temperature of the hot side fluid at the inlet, T_2 is the temperature of the hot side fluid at the outlet, t_1 is the temperature of the cold side fluid at the inlet, and t_2 is the temperature of the cold side fluid at the outlet:

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln(\frac{T_1 - t_2}{T_2 - t_1})}$$

The design equation, given below, was then solved for the required heat exchange area:

$$Q = U_c A \Delta T_{lm}$$

$$A = \frac{Q}{U_c \Delta T_{lm}}$$

The heat exchanger length L , outer diameter D , and tube diameter d were based on the available Doyle and Roth catalog [53]. For any heat exchangers that were too large to be found on the catalog, the tube length and tube diameter were assumed to be fixed at 20 ft (6.10 m) and 0.625 in (1.59 cm), and the shell outer diameter, D , was calculated for our desired surface area. Because the cross-sectional area of the heat exchanger, and therefore the number of tubes and the exchange surface area, increases with shell outer diameter squared, a regression using the available Doyle and Roth diameter sizes and surface areas was used to determine the exchange area of a larger outer diameter shell and tube heat exchanger. The equation for the regression, with D in ft and A in ft^2 is:

$$A = 3.28 * (1.33224D^2 - 3.45498D + 1.15538)$$

Microsoft's GoalSeek function was used to determine the required diameter for the area needed, and the length, L of the heat exchangers in ft was calculated based on the shell outer diameter:

$$L = 0.1533 * D + 20.621$$

As an example, we show the calculations for E200; all other heat exchangers are summarized in **Table C-5**.

$$U_c = (U^{-1} + F_{tube} + F_{shell})^{-1} = ((280 \frac{W}{m^2 K})^{-1} + 0.002 \frac{m^2 K}{W} + 0.002 \frac{m^2 K}{W}) = 251.8 \frac{W}{m^2 K}$$

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln(\frac{(T_1 - t_2)}{(T_2 - t_1)})} = \frac{(109.4 K - 85 K) - (93.9 K - 25 K)}{\ln(\frac{(109.4 K - 85 K)}{(93.9 K - 25 K)})} = 42.9 K$$

$$A = \frac{Q}{U_c \Delta T_{lm}} = \frac{11.3856 * 10^6 W}{251.8 \frac{W}{m^2 K} * 42.9 K} = 1054 m^2$$

Using GoalSeek, the desired shell diameter D is about 52.26 in. Assuming Doyle and Roth can manufacture shell diameters in increments of two inches, the desired diameter is 54 in, or 1.37 m.

$$L = 0.1533 * D + 20.621 = 0.1533 * 54 + 20.621 = 28.299 ft = 8.626 m$$

Heat Exchanger	Q (kW)	U^1 (W/m ² K)	U_c^2 (W/m ² K)	ΔT_{lm} (K)	A (m ²)	D (m)	L (m)	d (cm)
E200	11387	280	252	42.9	1128	1.37	8.8	1.59
E201	59395	280	240	43.8	5721	3.05	11.9	1.59
E202	5553	280	238	124.1	191	0.81	4.7	1.59
E203	17743	280	230	135.6	602	1.07	7.6	1.59
E204	3758	280	230	34.0	486	1.02	7.0	1.59
E205	61864 ³	N/A ⁴	580	107.5	1044	1.32	8.7	1.59
E206	28528	280	238	76.0	1596	1.63	9.3	1.59
E207	7084	280	250	94.9	304	0.86	6.1	1.59
E208	493	280	250	115.0	18	0.41	2.2	1.59
E209	675	280	238	52.0	56	0.46	4.0	1.59
E210	5931	160	150	99.4	401	1.07	5.8	1.59

E211	5482	160	150	102.4	365	1.02	5.7	1.59
E300	3432	625	500	51.9	134	0.76	4.1	1.59
E301	11628	160	150	30.5	2644	2.08	10.1	1.59
E302	487	165	156	122.1	26	0.36	3.3	1.59
E400	5238	160	150	52.2	669	1.07	8.2	1.59
E401	1618	160	150	35.5	304	1.02	5.1	1.59
E402	5843	160	118	23.1	2146	1.88	9.7	1.59
E403	4842	625	500	35.8	275	0.97	5.1	1.59
E500	1079	700	493	5.9	377	1.02	6.1	1.91
E501	3535	600	441	144.1	56	0.46	4.0	1.59
E502	6564	700	493	16.5	811	1.17	8.4	1.59
E503	7383	600	441	90.0	187	0.71	7.6	2.54
E600	9903	700	493	15.4	1305	1.47	9.0	1.59
E601	12342	600	441	162.3	174	0.86	4.2	1.59
E700	6658	850	621	36.6	296	0.91	5.6	1.59
E701	6907	600	441	35.3	468	1.07	6.4	1.59
E702	35531	600	441	193.8	425	1.02	6.3	1.59
E703	34070	850	621	52.3	1085	1.35	8.8	1.59
E800	2664	850	621	49.5	87	0.56	4.2	1.59

E801	2161	600	441	69.0	74	0.66	3.3	1.59
E802	172	250	227	70.2	11	0.27	2.5	1.59
E1000	235	600	441	260.0	2.2	0.17	1.7	1.91
E1001	1084	625	500	37.8	58	0.71	2.7	1.59
E1002	3591	625	500	49.1	146	0.97	3.9	1.91
E1003	2377	250	217	54.3	202	0.81	6.5	2.54
E1100	132	600	441	195.0	1.9	0.17	1.7	2.54
E1101	6053	625	500	48.7	252	0.81	7.8	2.54
E1102	31278	625	500	37.3	1699	1.68	9.4	1.59
E1200	18	600	441	199.0	1.3	0.17	1.7	2.54
E1201	1645	625	500	50.1	67	0.76	2.9	1.59
E1202	8076	625	500	37.7	430	1.02	7.0	1.91

¹Heat transfer coefficients were determined from Towler et al [54] or Luyben [55].

²Fouling factors were determined from Towler et al [54].

³The duty for the catalyst cooler was determined by an energy balance (see “Material and Energy Balances” section). All other duties were calculated in ASPEN.

⁴The heat transfer coefficient for E205 was calculated manually according to Bai [56] (see below) and included the fouling factors in its calculation.

Table C-5: Heat exchanger sizing using heat, log-mean temperature difference, and fouling factor

The following equation describes the calculation for U of E205 where h_i describes the convective heat transfer coefficient of the water to the tube, F_i is the fouling factor inside the tubes, A_i is the inner area of the tubes, A_o is the outer area, h_o is the convective heat transfer coefficient of the fluidized catalyst to the tube, F_o is the fouling factor in the shell, δ_t is the thickness of the tube wall, and λ is the thermal conductivity of the carbon steel tube.

$$U = ((h_i^{-1} + F_i)(A_o/A_i) + (h_o^{-1} + F_o) + (\delta_t/\lambda)(A_o/A_i))^{-1}$$

The values were calculated assuming a 1.9 cm tube diameter with thickness 0.254 cm, $F_i=0.0001 \text{ m}^2\text{K/W}$, $F_o=0$, $h_i=15,000 \text{ W/m}^2\text{K}$, $h_o=700 \text{ W/m}^2\text{K}$, and carbon steel conductivity λ of 43.5 W/mK.

$$U = (((15000 \frac{W}{m^2K})^{-1} + 0.0001 \frac{m^2K}{W})(2.8353 * 10^{-4} \text{ m}^2 / 2.1279 * 10^{-4} \text{ m}^2) + (700 \frac{W}{m^2K})^{-1} + (0.00254 \text{ m} / 43.5 \frac{W}{mK})(2.8353 * 10^{-4} \text{ m}^2 / 2.1279 * 10^{-4} \text{ m}^2))^{-1} = 580 \frac{W}{m^2K}$$

Pumps

To estimate the power requirements of each pump, the volumetric flow rate in GPM and density of the fluid being pumped in $\frac{\text{kg}}{\text{m}^3}$ were obtained from the stream results read-out in ASPEN, utilizing the Peng-Robinson physical property method. Because Stage 1 was not modeled in ASPEN, the density of methanol was approximated using known values [57] and multiplied by the mass flow rate to calculate the volumetric flow rate. The total head loss opposing the flow of the pump in m of head was calculated using the following equation:

$$\text{Head (m)} = \frac{\Delta P}{\rho g}$$

Where ΔP is the difference in pressure between the discharge and suction sides of the pump in Pa,

$$\Delta P = P_{\text{discharge}} - P_{\text{suction}}$$

ρ is the density of the fluid in kg/m^3 , and g is the gravitational constant in m/s^2 .

Using head and GPM, the Gould's Model 3196 Chemical Process Pumps catalog [58] was used to identify the best model for each use case, as well as the power requirement for that given pump.

As an example, we show the calculations for P700; all other pumps are summarized in **Table C-6**.

$$V_{\text{GPM}} = 440.5 \text{ GPM}$$

$$\text{Head (m)} = \frac{P_{\text{discharge}} - P_{\text{suction}}}{\rho g} = \frac{(29 \text{ bar} - 25.3 \text{ bar}) * \frac{100,000 \text{ Pa}}{\text{bar}}}{444.5 \frac{\text{kg}}{\text{m}^3} * 9.81 \frac{\text{m}}{\text{s}^2}} = 84.84 \text{ m}$$

Using the calculated head loss and volumetric flow rate obtained from ASPEN, the hydraulic coverage charts in the Gould catalog can be used. Assuming a motor RPM of 3500 along with a volumetric flow rate of 440.5 GPM and 84.84 m of head, the 2x3-10 model was selected and is shown in **Figure C-3**.

Hydraulic Coverage Model 3196

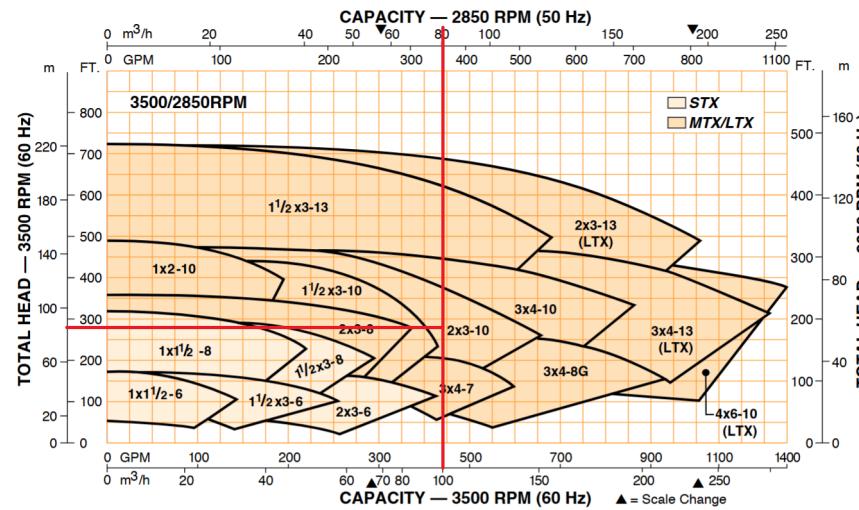


Figure C-3. Selecting pump model using hydraulic coverage chart

After deciding on the model of the pump, the 2x3-10 pump curve can be used. Using the same 440.5 GPM capacity and 84.84 m of head, the pump power is obtained in **Figure C-4**.

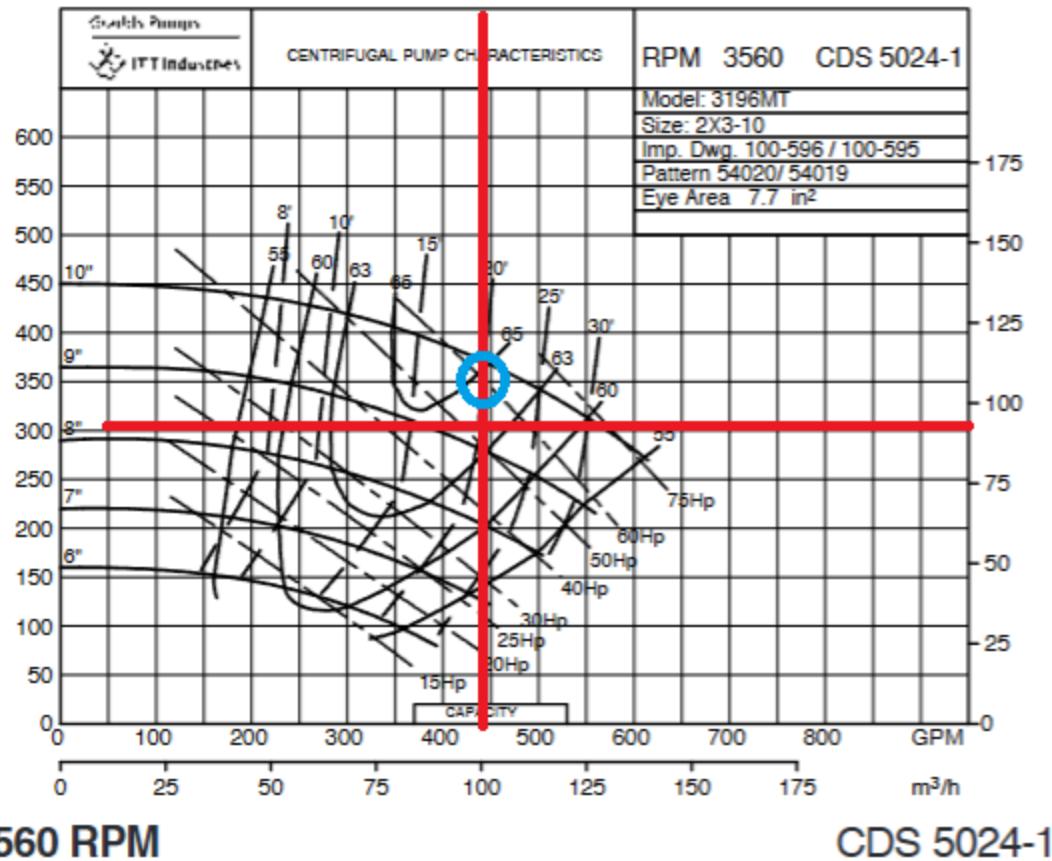


Figure C-4. Determining pump power using pump curve

Because the operating conditions of P700 do not place it directly on a pump curve, the closest pump curve above the operating point must be selected. For P700, this occurs at 60 hp, or 44.78 kW.

Pump	Flow Rate (GPM)	Pressure Drop (Pa)	Density (kg/m^3)	Head Loss (m)	Power (kW)
------	-----------------	--------------------	------------------------------------	---------------	------------

P100A-B	81.8	80,000	792.0	10.30	3.70
P101A-C	1050.0	370,000	792.0	48.62	37.29
P200	32.1	100,000	989.1	10.31	0.54
P201	35.2	60,000	989.1	6.18	0.34
P300	153.7	200,000	1002.6	20.33	3.73
P301	162.1	80,000	1521.7	5.36	1.12
P400	671.1	190,000	515.8	37.55	29.85
P600	341.5	400,000	417.0	97.79	37.31
P700	440.5	370,000	444.5	84.84	44.78
P701	4190.1	340,000	457.5	83.12	224.0 ¹

¹P701 power input exceeded pump curves from Gould catalog, so the pump curve from "Pentair Berkeley B-Series Close Coupled Motor Drive Centrifugal Pumps" was used [59].

Table C-6: Pump sizing using volumetric flow rate, pressure drop, and fluid density

Compressors

The sizing process for compressors is relatively simple, as the key sizing parameter is the power requirement of the compressor. To determine these values, we modeled our stages in ASPEN using the Peng-Robinson physical property method. Within each stage, we modeled our compressors as polytropic centrifugal compressors using ASME method with a polytropic efficiency of 0.75. We used **Figure C-5** below and the volumetric flow rates within our process to estimate the polytropic efficiency and determine the appropriate type of compressor for our process.

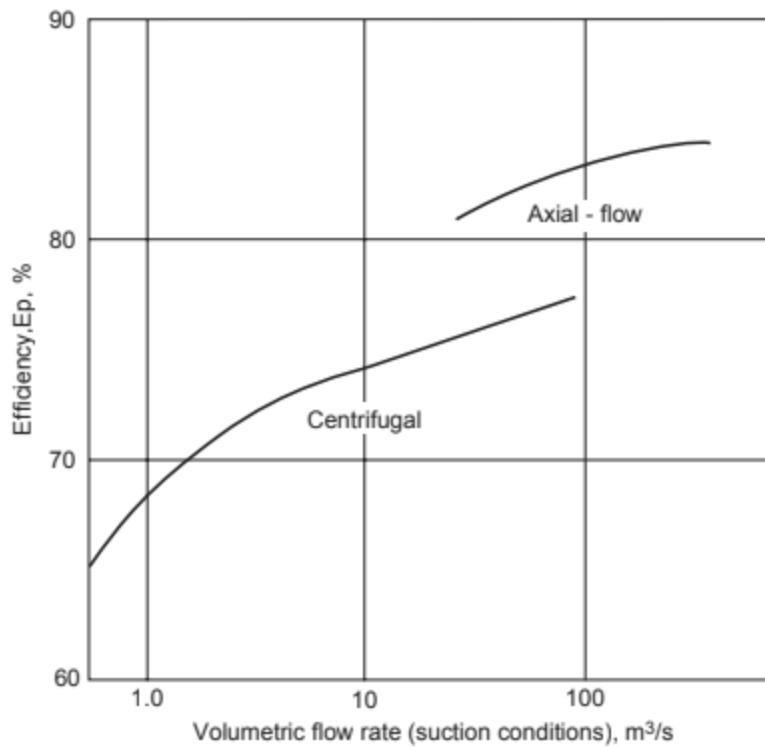


Figure C-5. Polytropic efficiency vs volumetric flow rate at compressor outlet [54]

All of our compressors have inlet volumetric flow rates between 10 and 60 m³/s, so we decided to assume a polytropic efficiency of 0.75 for all of our compressors to simplify our analysis. With the polytropic efficiency and compressor analysis method determined, we could then simply run our ASPEN simulations and record the reported compressor duties. **Figure C-6** below depicts the tab from which we obtained our compressor duties.

Compressor model	ASME polytropic
Phase calculations	Vapor phase calculation
Indicated horsepower	5.49023 MW
Brake horsepower	5.49023 MW
Net work required	5.49023 MW
Power loss	0 MW
Efficiency	0.75
Mechanical efficiency	1
Outlet pressure	1.9 bar
Outlet temperature	155.225 C
Isentropic outlet temperature	141.273 C
Vapor fraction	1
Displacement	
Volumetric efficiency	

Figure C-6. An example of the tabulated data provided by ASPEN for each compressor in our process

For each compressor, we simply recorded the “net work required” value reported by ASPEN. The power requirements of all compressors in our plant are depicted below in **Table C-7**.

Compressor	Power (MW)
C200	5.49
C201	5.87

C202	5.28
C203	5.89
BL200	4.26
BL300	0.287
C400	4.75
C401	1.52
C402	0.915
C700	0.062
C1000	3.51
C1001	2.23
C1100	13.32
C1101	5.34
C1200	3.46
C1201	1.41

Table C-7. Compressor duties for our MtO process

Distillation Columns

Achieving Specified Product Output and Purity. Our general approach to designing our distillation columns was to target the specific product flow rates and purities we needed to achieve our target 0.5 MMTPA of ethylene and propylene. Using ASPEN with the Peng-Robinson property model, we modeled all of our distillation columns with “RadFrac” blocks. We then changed the distillate-to-feed ratio until the distillate flow rate matched the values we needed based on overall mass balances assuming 24/7 operation for a total of 340 days annually. With the distillate-to-feed ratio set, we then would change the reflux ratio and number of trays until we achieved our desired purities. **Figure C-7** below shows the tab in which we changed these properties for each column, and **Table C-8** summarizes our findings for each column. The exception to this process was our design for the C3 splitter, composed of columns T701 and T702. Due to the difficulty of the separation, our C3 splitter would require an unfeasibly tall column to house the necessary number of stages. Accordingly, we designed the C3 splitter as separate stripping and rectifying columns and set the distillate flow rate for the rectifying section and bottoms flow rate for the stripping section. We set these flow rates using the known mass and purity of the stream entering the C3 splitter and knowledge of our desired ethylene product.

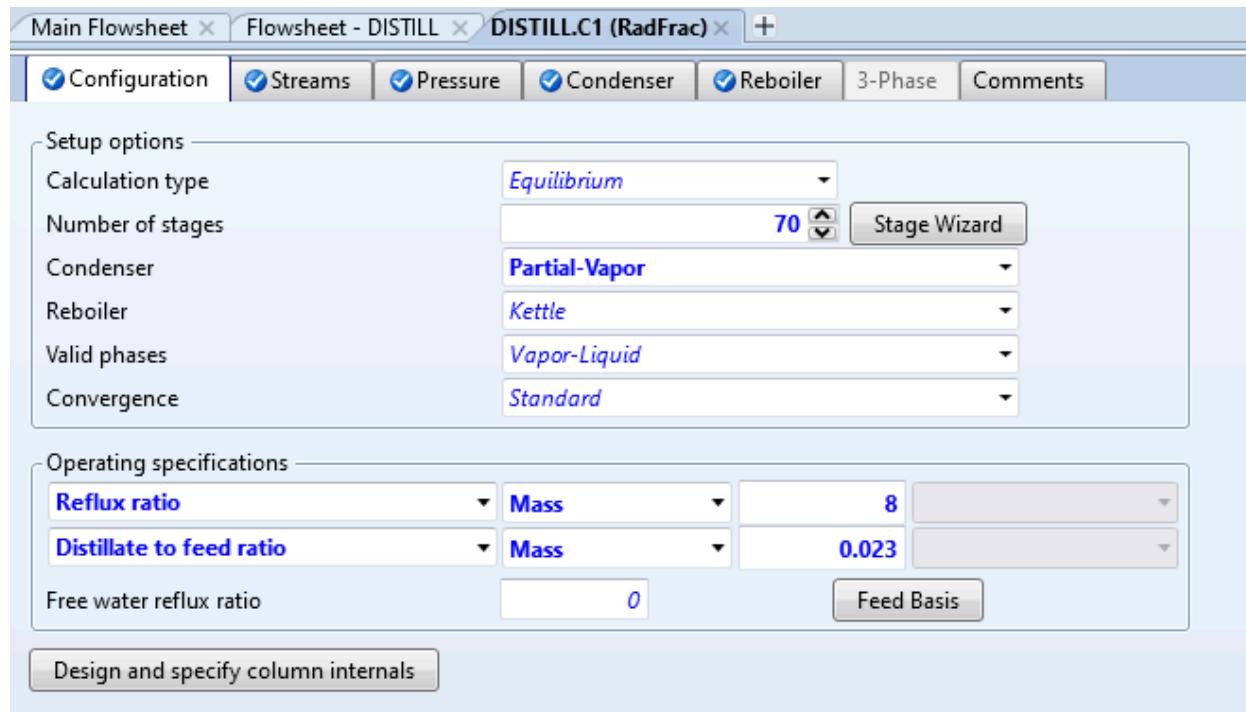


Figure C-7. An example of the tab we used to modify the properties of our columns in ASPEN

Distillation Tower	Distillate to Feed Ratio	Reflux Ratio	Number of Trays
T500	0.023	8.0	68
T501	0.421	1.5	38
T600	0.945	4.5	68
T700	0.724	2.0	48
T701	N/A	N/A	89
T702	N/A	N/A	89
T800	0.737	2.7	48

Table C-8. The column specifications that we used to achieve our desired product flow rates and purities

Sizing. With the internal components of the columns specified, we could then determine the necessary physical dimensions of our distillation columns. To determine the length of the columns, we used the following equation.

$$L = ns + VD + LH$$

Where L is the length of the distillation column in m, n is the number of trays in the columns, s is the spacing between trays (0.5588 m / 22 in for all columns), VD is the vapor disengagement space (1.2192 m / 4 ft for all columns), and LH is the liquid holdup space (1.8288 m / 6 ft for all columns). An example of this process using T500 is shown below.

$$L = (68 * 0.5588 \text{ m}) + 1.2192 \text{ m} + 1.8288 \text{ m} = 41.05 \text{ m}$$

To determine the necessary internal diameter for our columns, we used the following equation.

$$D = \sqrt{\frac{4VRT}{\pi n PaU}}$$

Where D is the internal diameter in m, V is the molar flow rate of vapor within the column in mol/s, R is the molar gas constant in $\text{m}^3\text{barK}^{-1}\text{mol}^{-1}$, T is the temperature within the column in K, P is the pressure within the column in bar, n is the tray efficiency, a is a safety factor, and U is the flood velocity in m/s. To determine the molar flow rate of vapor within the column, we decided to use the molar flow rate at the bottom of the column as that would provide the most conservative sizing estimate. We estimated the temperature and pressure as the average values within

each column. Based on class notes provided in our separations class, we estimated the safety factor as a value of 0.75 and the tray efficiency as a value of 0.8 [60]. To determine the flooding velocity for each column, we used the following equation

$$U = C \left(\frac{\sigma}{20} \right)^{0.2} \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

Where U is the flood velocity in ft/s, C is an empirically determined constant in ft/s, σ is the surface tension of liquid at the top of the column in dyne/cm, ρ_L is the density of liquid at the top of the column in kg/m³, and ρ_V is the density of vapor in the bottom of the column in kg/m³. To determine the surface tension and density of liquid within the columns, we used the properties reported by ASPEN for the liquid leaving the condenser and reentering the column at tray 1. To determine the density of vapor in the column, we used the vapor density reported by ASPEN for the vapor exiting the reboiler and reentering the column. We used the following graph in **Figure C-8** to determine C .

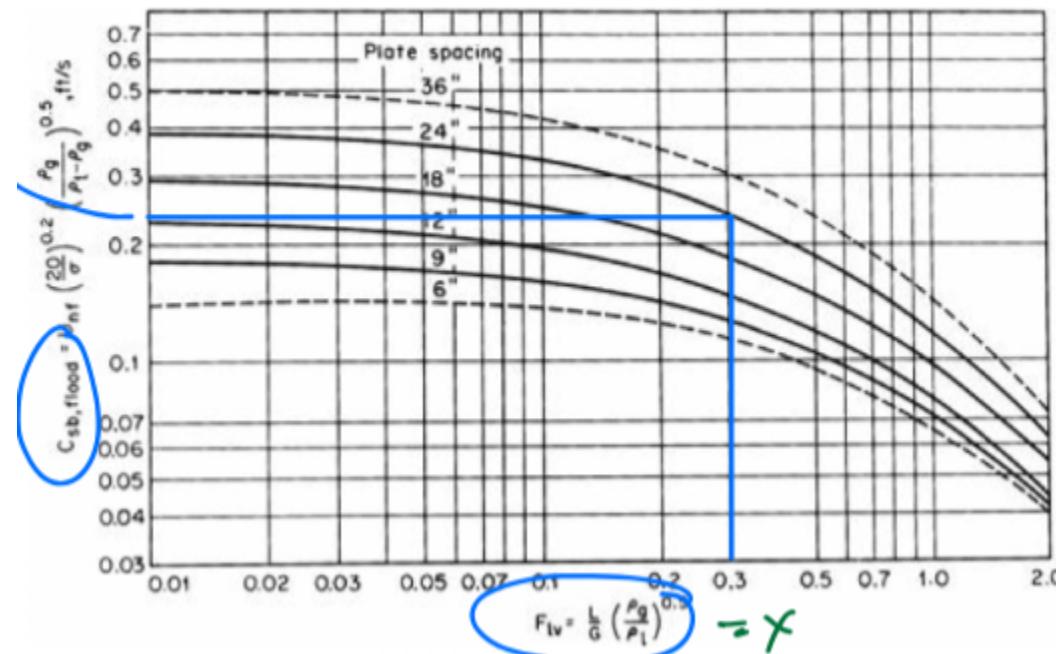


Figure C-8. The graph for determining the value of C as a function of Flv [60]

To use the graph, we needed to calculate the capacity factor using the following equation.

$$Flv = \frac{W_L}{W_V} \sqrt{\frac{\rho_V}{\rho_L}}$$

Where Flv is the capacity factor, W_L is the mass flow rate of liquid in the top of the column in kg/h, W_V is the mass flow rate of vapor in the bottom of the column in kg/h, and the ρ terms are the same densities mentioned previously. We determined the mass flow rates using the liquid stream that exits the condenser and reenters the column and the vapor stream that exits the reboiler and reenters the column, respectively. An example of this process using T500 is shown below.

$$Flv = \frac{14466.6 \frac{kg}{hr}}{40319.3 \frac{kg}{hr}} \sqrt{\frac{59.469 \frac{kg}{m^3}}{285 \frac{kg}{m^3}}} = 0.1639$$

With Flv determined, we could then estimate C from the graph, calculate U , and then perform a unit change to determine the flood velocity in m/s. An example of this process using T500 is shown below. Using the calculated value of Flv , we estimated C to be 0.3.

$$U = 0.3 * \left(\frac{5.32741 \frac{dyne}{cm}}{20} \right)^{0.2} \sqrt{\frac{285 \frac{kg}{m^3} - 59.469 \frac{kg}{m^3}}{59.469 \frac{kg}{m^3}}} = 0.4484 \frac{ft}{s}$$

$$U = 0.4484 \frac{ft}{s} * \frac{0.3048 m}{1 ft} = 0.1367 \frac{m}{s}$$

With all the variables in our diameter equation finally determined, we could calculate the diameters of our columns. An example of this process using T500 is shown below.

$$D = \sqrt{\frac{4 * 357.94 \frac{mol}{s} * 8.314 * 10^{-5} \frac{m^3 bar}{K mol} * 235 K}{3.14 * 0.8 * 30.2 bar * 0.75 * 0.1367 \frac{m}{s}}} = 1.90 m$$

We repeated this process for all of our distillation columns, and the results are summarized in **Table C-9** below.

Distillation Tower	Length (m)	Flv	C (ft/s)	U (m/s)	D (m)
T500	41.05	0.1639	0.30	0.1367	1.90
T501	24.28	0.1838	0.27	0.1527	3.01
T600	41.05	0.3463	0.25	0.1288	4.50

T700	29.87	0.1990	0.26	0.1235	3.61
T701	52.22	0.2908	0.22	0.1429	8.84
T702	52.22	0.3054	0.22	0.1429	7.70
T800	29.87	0.1774	0.27	0.2688	1.71

Table C-9. The physical dimensions of our distillation columns

To determine the thickness of the walls of our columns, we used two different equations to calculate the required thickness and then chose the larger of the two values [61].

$$t_1 = \frac{PD}{2SE - 1.2P}$$

$$t_2 = \frac{PD}{4SE + 0.8P}$$

Where t_1 and t_2 are the wall thickness of the column in inches, P is the internal column pressure in psi, D is the internal diameter in inches, S is the allowable stress of the material in psi, and E is the weld efficiency. Based on our source, we assumed an allowable stress of 10,000 psi and a weld efficiency of 0.7 for all of our columns. An example of this process using T500 is shown below.

$$t_1 = \frac{438.01\text{psi} * 74.65\text{in}}{2 * 10000\text{psi} * 0.7 - 1.2 * 438.01\text{psi}} = 2.43 \text{ in}$$

$$t_2 = \frac{438.01\text{psi} * 74.65\text{in}}{4 * 10000\text{psi} * 0.7 + 0.8 * 438.01\text{psi}} = 1.15 \text{ in}$$

We chose the larger of the two values and then converted it to meters. **Table C-10** below summarizes the wall thicknesses for our columns.

Distillation Tower	t1 (in)	t2 (in)	Selected Thickness (mm)
T500	2.43	1.15	62
T501	3.18	1.53	81
T600	4.78	2.29	121
T700	3.83	1.83	97

T701	7.77	3.75	197
T702	6.54	3.16	80
T800	0.72	0.35	9

Table C-10. Distillation column wall thickness

Flash Drums

To determine the diameter of our flash drums, we used the following equation.

$$D = \sqrt{\frac{4A}{\pi}}$$

Where D is the internal diameter of the drum in m and A is the cross sectional area of the drum in m^2 . To determine the cross sectional area of the drum, we used the following equation.

$$A = \frac{VM}{U\rho_v}$$

Where V is the molar flow rate of vapor in the drum in mol/s , M is the molecular weight of the vapor in kg/mol , U is the permissible velocity in m/s , and ρ_v is the vapor density in kg/m^3 . We determined the molar flow rate of vapor and vapor density using the ASPEN output tables for our streams, and we knew the molar mass of our vapor was either that of ethylene or propylene (depending on the refrigeration loop). To determine the permissible velocity, we used the following equation.

$$U = K \sqrt{\frac{\rho_L - \rho_v}{\rho_v}} * \frac{0.3048m}{1ft}$$

Where K is the drum coefficient in ft/s and ρ_L is the liquid density in kg/m^3 . We determined the liquid density using the ASPEN output tables for our streams. We determined K using the following equation.

$$K = \exp(A + Bln(Flv) + Cln(Flv)^2 + Dln(Flv)^3 + Eln(Flv)^4)$$

Where A, B, C, D , and E are empirical constants (-1.877478097, -0.8145804597, -0.1870744085, -0.0145228667, and -0.0010148518, respectively) and Flv is the capacity factor. We calculated Flv using the following equation.

$$Flv = \frac{w_L}{w_v} \sqrt{\frac{\rho_v}{\rho_L}}$$

Where W_L is the mass flow rate of liquid in the drum in kg/h and W_V is the mass flow rate of vapor in the drum in kg/h. An example of this calculation process for D1000 is included below.

$$Flv = \frac{10024 \frac{\text{kg}}{\text{hr}}}{164789 \frac{\text{kg}}{\text{hr}}} \sqrt{\frac{2.43 \frac{\text{kg}}{\text{m}^3}}{564.816 \frac{\text{kg}}{\text{m}^3}}} = 0.0164$$

$$K = \exp(A + Bln(0.0164) + Cln(0.0164)^2 + Dln(0.0164)^3 + Eln(0.0164)^4) = 0.1509 \frac{\text{ft}}{\text{s}}$$

$$U = 0.1509 \frac{\text{ft}}{\text{s}} \sqrt{\frac{564.816 \frac{\text{kg}}{\text{m}^3} - 2.43 \frac{\text{kg}}{\text{m}^3}}{2.43 \frac{\text{kg}}{\text{m}^3}}} * \frac{0.3048 \text{m}}{1 \text{ft}} = 0.700 \frac{\text{m}}{\text{s}}$$

$$A = \frac{395.88 \frac{\text{mol}}{\text{s}} * 0.02805 \frac{\text{kg}}{\text{mol}}}{0.7 \frac{\text{m}}{\text{s}}} = 6.53 \text{ m}^2$$

$$D = \sqrt{\frac{4 * 6.53 \text{ m}^2}{3.14}} = 2.88 \text{ m}$$

To determine the length of our flash drums, we assumed a length-to-diameter ratio of 3 [62].

Repeating this process for all of our flash drums, we obtain **Table C-11**.

Flash Drum	Flv	K (ft/s)	U (m/s)	A (m ²)	D (m)	L (m)
T300	0.052	0.1466	0.912	12.87	4.05	12.14
D1000	0.016	0.1509	0.700	6.53	2.88	8.65
D1100	0.155	0.1343	0.511	12.17	3.94	11.81
D1200	0.129	0.1373	0.574	3.52	2.12	96.35

Table C-11. The parameters used to calculate the internal diameter for our flash drums

To determine the wall thickness of our flash drums, we used the same equations discussed in the distillation column section. However, we found that the calculated values were all lower than the minimum wall thickness required for the flash drums to stand upright. Accordingly, wall thicknesses were selected based on recommended minimum wall thicknesses for given diameters [63]. **Table C-12** below summarizes the wall thicknesses for our flash drums.

Flash Drum	Wall Thickness (mm)
T300	9
D1000	10
D1100	15
D1200	9

Table C-12. The wall thicknesses of our flash drums

Storage Tanks

The sizing process for storage tanks was largely influenced by the overall capacity needed, as well as the operating pressure.

Floating Roof Atmospheric Storage Tanks. Floating roof atmospheric storage tanks were chosen for the storage of methanol at this facility because of the high volume needed for operation. To obtain desired capacity of the storage tanks, the following equation was used to calculate the volume needed to continue steady-state operation, and to add three days of capacity to that to mitigate supply chain disturbances.

$$V_{capacity} = \frac{4m}{\rho}$$

Where m is the mass flow rate of the supplied chemical into the process in kg/day, and ρ is the density of the fluid in kg/m³. This volume was then split over several tanks if needed, as the maximum capacity found for floating roof atmospheric storage tanks was found to be 2 MGal (7,751 m³) [64]. The equation for this determination is provided below:

$$N_{tanks} = \frac{V_{capacity}}{V_{tank,available}}$$

The L/D of the tank was constrained to 0.1, as this is in the middle of the range found for commercially available floating roof bulk chemical storage [64]. Because L/D was fixed at 0.1, the following relationships could be used:

$$\frac{L}{D} = 0.1$$

$$10L = D$$

$$V = L\left(\frac{D}{2}\right)^2\pi$$

$$V = L \left(\frac{10L}{2}\right)^2 \pi$$

$$L = \sqrt[3]{\frac{V}{25\pi}}$$

The thickness of the tank was calculated using the following equation:

$$t \text{ (mm)} = \frac{\rho H_t g D_t}{2S_t E * 10^3}$$

Where H_t is the maximum liquid depth in meters, g is the gravitational constant in m/s^2 , D_t is the tank diameter in m, S_t is the maximum allowable stress for the tank material in N/mm^2 , and E is the joint efficiency of the tank. As an example, we show the calculations for TK100A-C, and the dimensions of these tanks is provided in **Table C-13**.

$$V_{capacity} \text{ (m}^3\text{)} = \frac{4}{792 \frac{\text{kg}}{\text{m}^3}} \left(\frac{188884 \text{ kg}}{\text{hr}} \right) * \left(\frac{24 \text{ hr}}{\text{day}} \right) = 22,895 \text{ m}^3$$

$$N_{tanks} = \frac{22,895 \text{ m}^3}{7,571 \text{ m}^3} = 3.02$$

Because the number of tanks must be a whole number, the number of tanks was rounded down to 3, and the capacity was reduced by 182,000 L.

$$L = \sqrt[3]{\frac{7571 \text{ m}^3}{25\pi}} = 4.59 \text{ m}$$

$$10L = D = 45.9 \text{ m}$$

$$t \text{ (mm)} = \frac{\frac{792 \text{ kg}}{\text{m}^3} (4.59 \text{ m})(9.81 \frac{\text{m}}{\text{s}^2})(45.9 \text{ m})}{2(88.97 \frac{\text{N}}{\text{mm}^2})(0.7)*10^3} = 13.2 \text{ mm}$$

Because the tank dimensions are not whole numbers and unlikely to be easily manufactured as a result, the tank dimensions were rounded to a length of 4.6 m, a diameter of 46 m, and a thickness of 14 mm. **Table C-12** below summarizes our results.

Tank	Length (m)	Diameter (m)	Max Depth (m)	Joint Efficiency	Material	Maximum Allowable Stress (N/mm^2)	Thickness (mm)
TK100A-C	4.60	46.0	4.60	0.7 ¹	Carbon Steel	88.97 ²	14
TK300	2.00	20.0	1.80	0.7 ¹	304 Stainless Steel	137.93 ²	4

¹Joint efficiency estimated from "Equipment Design Heuristics" [62].

²Maximum allowable stress of carbon steel was determined from Towler et al [54].

Table C-13. The physical dimensions and construction materials of our floating roof atmosphere storage tanks

Pressurized Storage Tanks. The pressurized storage tank procedure necessarily follows a procedure different from that of floating roof atmospheric storage tanks due to design limitations when operating above ambient pressure. Firstly, the following equation was employed to calculate the required capacity needed for the tank:

$$V_{capacity} (m^3) = \frac{4m}{\rho}$$

Where m is the mass flow rate of the supplied chemical into/from the process in kg/day, and ρ is the density of the fluid in kg/m³, both obtained through the stream results read-out in ASPEN.

Because storage tanks under pressure have a fixed design range of lengths and diameters [62], lengths on the higher end of the range, as well as L/D ratios on the lower end, were selected to maximize the capacity of each tank safely. Out of the range provided by [62], a length of 9.15 m and a diameter of 3.81 m were selected, yielding a L/D of 2.4. For each tank, 10% of the volume is designated for freeboard, and this relation is shown in the following equation:

$$V_{tank,available} (m^3) = 0.9V_{tank}$$

If the total volume of material required exceeded the capacity of a tank with these dimensions, multiple tanks were specified utilizing the following equation:

$$N_{tanks} = \frac{V_{capacity}}{V_{tank,available}}$$

To calculate the required wall thickness of the tank, the following equations were used:

$$t_1 (in) = \frac{P_i D_i}{2SE - 1.2P_i}$$

$$t_2 (in) = \frac{P_i D_i}{4SE + 0.8P_i}$$

Where t is the thickness of the tank shell in in, P_i is the internal pressure of the tank in psi, S is the maximum allowable stress of the material in psi, and E is the joint efficiency. The larger of these two thicknesses was selected as the minimum thickness of the tank. As an example, we show the calculations for TK900A-C, and the dimensions of these tanks are provided in **Table C-14**.

$$V_{capacity} = \frac{4}{501 \frac{kg}{m^3}} \left(\frac{3713 \frac{kg}{hr}}{hr} \right) * \left(\frac{24 \frac{hr}{day}}{day} \right) = 711.5 \frac{m^3}{}$$

$$V_{tank} = \left(\frac{3.81m}{2} \right)^2 \pi (9.15m) = 104.3 \frac{m^3}{}$$

$$V_{tank,available} = (0.9)(104.3 \frac{m^3}{}) = 93.9 \frac{m^3}{}$$

$$N_{tanks} = \frac{711.5 \text{ m}^3}{93.9 \text{ m}^3} = 7.57 \text{ tanks}$$

Because we cannot have a fractional tank, the number of storage tanks was rounded to 8.

$$t_1 = \frac{(131.95 \text{ psi})(150 \text{ in})}{2(12900 \text{ psi})(0.7) - 1.2(131.95 \text{ psi})} \left(\frac{25.4 \text{ mm}}{1 \text{ in}}\right) = 28.1 \text{ mm}$$

$$t_2 = \frac{(131.95 \text{ psi})(150 \text{ in})}{4(12900 \text{ psi})(0.7) + 0.8(131.95 \text{ psi})} \left(\frac{25.4 \text{ mm}}{1 \text{ in}}\right) = 13.9 \text{ mm}$$

The larger of these two values is 28.1 mm, which was rounded up to 29 mm.

Tank	Length (m)	Diameter (m)	Operating Pressure (bar)	Joint Efficiency	Material	Maximum Allowable Stress (N/mm ²)	Thickness (mm)
TK900A-H	9.15	3.81	9.1	0.7 ¹	Carbon Steel	88.97 ²	29

¹Joint efficiency estimated from [62].

²Maximum allowable stress of carbon steel was determined from [62].

Table C-14. The physical dimensions and construction materials of our pressurized storage tanks

Quencher Towers

To determine the required area of the quench towers the following equation was used.

$$A = \frac{V}{v}$$

Where A is the area of the quench tower in m², V is the volumetric flow rate in m³/s, and v is the superficial gas velocity in m/s. V was determined using the ASPEN output tables for our streams and v was assumed to be 0.8 m/s [65]. Using the area we were able to calculate the diameter using the following equation.

$$D = 2\sqrt{\frac{A}{\pi}}$$

Where D is the diameter of the quench tower in m. An L/D ratio can now be used to determine the quench tower length, L , in m. An L/D ratio of 2 was used based on Alinejad et al [66]. An example calculation for T200 is included below.

$$A = \frac{50.59 \frac{\text{m}^3}{\text{s}}}{0.8 \frac{\text{m}}{\text{s}}} = 88.23 \text{ m}^2$$

$$D = 2\sqrt{\frac{88.23 \text{ m}^2}{\pi}} = 10.6 \text{ m}$$

$$L = 2 \cdot 10.6 \text{ m} = 21.2 \text{ m}$$

Repeating this process for T201, we obtained the results depicted below in **Table C-15**.

Quench Tower	Diameter (m)	Length (m)
T200	10.6	21.2
T201	9.2	18.4

Table C-15. The physical dimensions of the quench towers

Absorption Towers

The absorption tower and caustic scrubber were designed using the same equations and methods outlined in the **Distillation Columns** sample calculation section. Refer to that section for example calculations. Using these equations, we obtained the results depicted below in **Table C-16**.

Tower	Diameter (m)	Length (m)	Thickness (mm)
T301	0.8	20.7	24
T302	0.7	20.7	15

Table C-16. The physical dimensions of the absorption tower and caustic scrubber

Gas Dryers

To determine the dimensions of the gas dryers, we first need to determine the amount of sorbent we need to adsorb the water from the incoming stream. To do this we use an isotherm for silica gel to find the capacity of water from the percent relative humidity of the stream. To get the percent relative humidity the following equations were used:

$$P_{H_2O} = y_{H_2O} \cdot P_T$$

Where P_{H_2O} is partial pressure of water in bar, y_{H_2O} is the vapor mole fraction of water in the stream which is obtained from our ASPEN stream tables, and P_T is the total pressure of the stream in bar.

$$\log_{10} P_{sat} = A - \frac{B}{C+T}$$

Where P_{sat} is the saturation pressure in mmHg, A, B, and C are Antoine constants all obtained from Felder, and T is the temperature of the stream [67].

$$RH = \frac{P_{H_2O}}{P_{sat}} \cdot 100\%$$

Where RH is the percent relative humidity. An example of the calculation is included below.

$$P_{H_2O} = 0.0055 \cdot 1.6 \text{ bar} = 0.0088 \text{ bar}$$

$$P_{sat} = 10^{8.10765 - \frac{1750.286}{235+34}} = 38.81 \text{ mmHg} \cdot \frac{1.01325 \text{ bar}}{760 \text{ mmHg}} = 0.0517 \text{ bar}$$

$$RH = \frac{0.0088 \text{ bar}}{0.0517 \text{ bar}} \cdot 100\% = 17\%$$

Using the percent relative humidity, we can use a silica gel isotherm to get the capacity of water the sorbent can hold, as shown in **Figure C-9**.

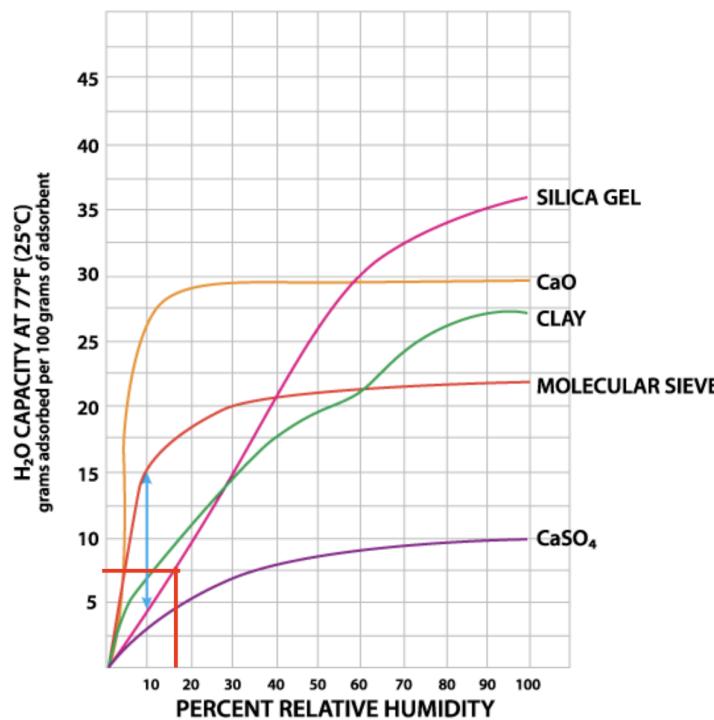


Figure C-9. Silica gel isotherm [68]

From the isotherm, the silica gel will have a capacity of 7.5 g of water per 100 g of adsorbent. Using this ratio we can calculate the amount of silica gel needed from the following equation.

$$M = m \cdot \frac{100 \text{ g sorbent}}{7.5 \text{ g water}} \cdot t$$

Where M is the total mass of silica gel in kg, m is the flow of water into the dryer in kg/h which is obtained from our ASPEN stream tables, and t is the hours of operation for the gas dryer before it needs to be regenerated. An example calculation is included below.

$$M = 221.2 \frac{\text{kg water}}{\text{hr}} \cdot \frac{100 \text{ g sorbent}}{7.5 \text{ g water}} \cdot 12 \text{ hr} = 35401.8 \text{ kg of silica gel}$$

Now that we have the total mass of silica gel we can get the volume of the silica gel and the tower using the following equations.

$$V = \frac{M}{\rho_b}$$

Where V is the volume of the silica gel in m^3 , and ρ_b is the bulk density of silica gel in kg/m^3 [69].

$$V_T = \frac{V}{1-\varepsilon}$$

Where V_T is the volume of the tower in m^3 and ε is the void fraction, obtained from Nastaj and Ambrozek [69]. An example calculation is included below.

$$V = \frac{35401.8 \text{ kg of silica gel}}{721 \frac{\text{kg}}{\text{m}^3}} = 49.1 \text{ m}^3$$

$$V_T = \frac{49.1 \text{ m}^3}{1-0.4} = 81.8 \text{ m}^3$$

Now that we have the volumes we can find the heights from the following equations.

$$A = \pi r^2$$

Where A is the area of the dryer in m^2 and r is the radius of the dryer in m, which is obtained from an assumed diameter of 1.8 m [50].

$$h = \frac{V}{A}$$

Where h is the height of the silica gel in m.

$$H = \frac{V_T}{A}$$

Where H is the height of the tower in m. An example of the calculations is included below.

$$A = \pi \cdot (0.9 \text{ m})^2 = 2.63 \text{ m}^2$$

$$h = \frac{49.1 \text{ m}^3}{2.63 \text{ m}^2} = 18.67 \text{ m of silica gel}$$

$$H = \frac{81.8 \text{ m}^3}{2.63 \text{ m}^2} = 31.11 \text{ m}$$

To account for the freeboard in the tower, 10% of the calculated height is added, so the true height of the dryers is 34 m. Now that the height of packing is obtained, we can also calculate the pressure drop in the dryer using the Ergun equation.

$$\Delta P = \frac{150(1-\varepsilon)^2 \cdot \mu \cdot v_o^2}{\varepsilon^3 \cdot d_p^2 \cdot h} + \frac{1.75(1-\varepsilon) \cdot \rho \cdot v_o^2}{\varepsilon^3 \cdot d_p \cdot h}$$

Where ΔP is the pressure drop in Pa, μ is the viscosity of the fluid in $\text{Pa}\cdot\text{s}$ which is obtained from our ASPEN stream tables, v_o is the superficial velocity in m/s , d_p is the particle diameter of silica gel in m, and ρ is the density of fluid in kg/m^3 which is obtained from our ASPEN tables. An example calculation is included below.

$$\Delta P = \frac{150(1-0.4)^2 (9.64 \times 10^{-6} \text{ Pa}\cdot\text{s})(3.72 \frac{\text{m}}{\text{s}})}{0.4^3 (1.15 \times 10^{-3} \text{ m})^2 (18.67 \text{ m})} + \frac{1.75(1-0.4)(2.237 \frac{\text{kg}}{\text{m}^3})(3.72 \frac{\text{m}}{\text{s}})^2}{0.4^3 (1.15 \times 10^{-3} \text{ m})(18.67 \text{ m})} = 24590 \text{ Pa} \cdot \frac{1 \text{ bar}}{100000 \text{ Pa}} = 0.25 \text{ bar}$$

To calculate the thickness of the dryer, the following equation was used.

$$t = \frac{PD}{2SE - 1.2P}$$

Where t is the wall thickness of the dryer in inches, P is the pressure in psi, D is the diameter in inches, S is the allowable stress of the material in psi, and E is the weld efficiency. Based on our source, we assumed an allowable stress of 12,900 psi and a weld efficiency of 0.7. An example calculation is included below.

$$t = \frac{21.4 \text{ psi}(72 \text{ in})}{2(12900 \text{ psi})(0.7) - 1.2(21.4 \text{ psi})} = 0.085 \text{ in} \cdot \frac{25.4 \text{ mm}}{1 \text{ in}} = 2.2 \text{ mm}$$

This wall thickness is the minimum needed to support the pressure inside the dryer, however it is too small to support the dryer's diameter so we will use the minimum wall thickness required of 12 mm, estimated from [62]. The dryer dimensions are summarized in **Table C-17** below.

Dryer	Diameter (m)	Height (m)	Height of Packing (m)	Wall Thickness (mm)	Pressure Drop (bar)
T303A-C	1.8	34	18.67	12	0.25

Table C-17. Dryer dimensions

Line Sizes, Pressure Drop Estimations, and Line List

Line Sizing. Using volumetric flow rate and heuristics for typical fluid velocities in pipes, the diameter of each stream in the process was calculated. Because a new stream number was not denoted for pressure and temperature changes that affect the volumetric flow rate of the material, a new line diameter was calculated each time that a stream was sent through a piece of process equipment. The volumetric flow rate was obtained from ASPEN in the stream read-out, an example of which is provided below in **Figure C-10**.

Material	Vol. % Curves	Wt. % Curves	Petroleum	Polymers	Solids	Status
					S1	
Description		Units				
From		COMP2				
To		E8				
Stream Class		CONVEN				
Maximum Relative Error						
Cost Flow	\$/hr					
- MIXED Substream						
Phase		Vapor Phase				
Temperature	C	78.6613				
Pressure	bar	20				
Molar Vapor Fraction		1				
Molar Liquid Fraction		0				
Molar Solid Fraction		0				
Mass Vapor Fraction		1				
Mass Liquid Fraction		0				
Mass Solid Fraction		0				
Molar Enthalpy	cal/mol	5406.7				
Mass Enthalpy	cal/gm	153.317				
Molar Entropy	cal/mol-K	-28.6231				
Mass Entropy	cal/gm-K	-0.811662				
Molar Density	mol/cc	0.00078753				
Mass Density	gm/cc	0.0277721				
Enthalpy Flow	cal/sec	3.34841e+06				
Average MW		35.2648				
+ Mole Flows	kmol/hr	2229.51				
+ Mole Fractions						
+ Mass Flows	kg/hr	78623.1				
+ Mass Fractions						
Volume Flow	cum/sec	0.786392				
+ Vapor Phase						
<add properties>						

Figure C-10. An example of the ASPEN readouts we used to determine the volumetric flow rate of our streams

For Stages 1 and 9, which were not simulated, hand calculations using mass flow rates and known densities were employed to determine volumetric flow rates. The following equation was used to calculate the cross sectional area of each pipe:

$$A = \frac{V}{u}$$

Where A is the cross sectional area in m^2 , V is the volumetric flow rate in m^3/s , and u is the fluid velocity in m/s . **Table C-18** below consolidates the fluid velocities used and the sources from which they were taken.

Fluid	Velocity (m/s)
Air	20.0 ¹
Hydrocarbon Vapors	30.0 ¹
Steam	20.0 ¹
Water	1.65 ¹
Sodium Hydroxide	1.83 ¹
Hydrocarbon Liquids	1.83 ²
Methanol	1.83 ²

¹Velocities estimated from “Lecture 14 - Fluids Slides” [70].

²Velocity estimated by taking average of liquid velocities [70].

Table C-18. Fluid velocities used for pipe sizing

In the case of a vapor-liquid mixture, the volumetric flow rates of each phase were divided by their respective fluid velocities and added together to approximate pipe diameter. This approach is summarized by the following equation:

$$A_{mix} = \frac{V_v}{u_v} + \frac{V_L}{u_L}$$

Where A_{mix} is the cross-sectional area of the pipe in m^2 , V_v is the volumetric flow rate of the vapor phase in m^3/s , u_v is the vapor velocity in m/s , V_L is the volumetric flow rate of the liquid phase in m^3/s , and u_L is the liquid velocity in m/s . Pipe diameter was then extracted from the cross sectional area using the geometric relationship provided below:

$$d_{pipe} = \sqrt{\frac{4A}{\pi}}$$

Where d_{pipe} is the diameter of the pipe in m and A is the cross-sectional area of the pipe in m^2 .

Pressure Drop Estimations. In an iterative design, it would be possible to estimate a friction factor associated with each line, use it to calculate a pressure drop, which would be incorporated into the process simulation, which would then alter the performance of the process, and ultimately lead to the re-design of much of the process. However, as this report does not encompass an iterative design, the following pressure drop heuristics were used as a first approximation for this process, and are provided in **Table C-19**.

Class of Equipment	Phase	Phase Change	Pressure Drop (bar)
Pipes	Liquid	N/A	0.1 ^{1,2}
Pipes	Vapor	N/A	~0 ^{1,2}
Heat Exchangers	Liquid	N/A	0.4 ²
Heat Exchangers	Liquid	Vaporizing	0.2 ²
Heat Exchangers	Vapor	N/A	0.4 ²
Heat Exchangers	Vapor	Condensing	0.2 ²
Reactor	Vapor	N/A	0.6 ^{3,4}
Regenerator	Vapor	N/A	0.8 ^{3,4}
Absorption Columns	Mixed	N/A	0.2 ⁴
Adsorption Columns	Mixed	N/A	0.25
Distillation Columns	Mixed	N/A	0.007 ^{2,5}
Control Valves	Liquid	N/A	0.7 ²

¹Pressure drop per 30.5 m of pipe

²Pressure drops calculated from “Equipment Design Heuristics” [62].

³Pressure drop calculated using information from “Fluidization Engineering” [71].

⁴Pressure drop calculated using information from Gupta and Rao [72].

⁵Pressure drop per stage, estimated from “Equipment Design Heuristics” [62].

Table C-19. Pressure drop estimations for various pieces of equipment

For an example of a one-phase stream, we show the calculation for Stream 600 entering E802:

$$A = \frac{0.169 \text{ m}^3/\text{s}}{30 \text{ m/s}} = 0.0056 \text{ m}^2$$

$$d_{\text{pipe}} = \sqrt{\frac{4(0.0056m^2)}{\pi}} = 0.085 \text{ m}$$

For an example of two-phase mixture, we show the calculation for Stream 1001 entering D1000:

$$A_{\text{mix}} = \frac{4.559 \text{ m}^3/\text{s}}{30 \text{ m/s}} + \frac{0.0049 \text{ m}^3/\text{s}}{1.83 \text{ m/s}} = 0.155 \text{ m}^2$$

$$d_{\text{pipe}} = \sqrt{\frac{4(0.155m^2)}{\pi}} = 0.444 \text{ m}$$

Because pipes are typically manufactured at set diameters, all pipe diameters were rounded up to the nearest readily available diameter [73] [74]. As a result of this, the diameter for Stream 600 entering E802 and Stream 1001 entering D1000 were rounded to 1 m and 0.45 m, respectively.

Line List. The diameter of every process line is given below in **Table C-20**, as well as service, material of construction, and whether the line is insulated or not.

Stream No.	Destination	Service	Diameter (m)	Material of Construction	Insulated?
100	V100	Methanol	0.25	Carbon Steel	No
101	V100	Methanol	0.25	Carbon Steel	No
102	V101A-C	Methanol	0.35	Carbon Steel	No
102	TK100A-C	Methanol	0.35	Carbon Steel	No
102	V102A-C	Methanol	0.25	Carbon Steel	No
102	P100A-C	Methanol	0.25	Carbon Steel	No
102	V103	Methanol	0.25	Carbon Steel	No
103	E200	Methanol	0.25	Carbon Steel	No
103	E201	Methanol	0.25	Carbon Steel	Yes
103	E202	Methanol	0.9	Carbon Steel	Yes

103	R200	Methanol	1.0	Carbon Steel	Yes
200	R200	Steam	0.05	Carbon Steel	Yes
201	V200	Spent Catalyst	0.35	Refractory Lined Carbon Steel	Yes
201	R200	Spent Catalyst	0.35	Refractory Lined Carbon Steel	Yes
202	BL200	Air	1.3	Carbon Steel	Yes
202	R201	Air	1.0	Carbon Steel	Yes
203	E203	Flue Gas	1.8	Refractory Lined Carbon Steel	Yes
203	E204	Flue Gas	1.5	Carbon Steel	Yes
203	Atmosphere	Flue Gas	1.6	Carbon Steel	Yes
204	E205	Catalyst	0.75	Refractory Lined Carbon Steel	Yes
204	R201	Catalyst	0.75	Refractory Lined Carbon Steel	Yes
205	V201	Catalyst	0.35	Refractory Lined Carbon Steel	Yes
205	R200	Catalyst	0.35	Refractory Lined Carbon Steel	Yes
206	T200	Process Fluid	1.8	Carbon Steel	Yes
207	P200	Cooling Water	0.04	Carbon Steel	No
207	T200	Cooling Water	0.04	Carbon Steel	No
208	E206	Process Fluid	1.8	Carbon Steel	Yes
208	E207	Process Fluid	1.6	Carbon Steel	Yes
208	C200	Process Fluid	1.7	Carbon Steel	Yes
208	E208	Process Fluid	1.4	Carbon Steel	Yes
208	T201	Process Fluid	1.6	Carbon Steel	Yes
209	P201	Cooling Water	0.05	Carbon Steel	No
209	T201	Cooling Water	0.05	Carbon Steel	No

210	Waste	Wastewater	0.025	Carbon Steel	Yes
211	E209	Process Fluid	1.5	Carbon Steel	Yes
211	C201	Process Fluid	1.8	Carbon Steel	Yes
211	E210	Process Fluid	1.4	Carbon Steel	Yes
211	C202	Process Fluid	1.5	Carbon Steel	Yes
211	E211	Process Fluid	1.2	Carbon Steel	Yes
211	C203	Process Fluid	1.3	Carbon Steel	Yes
211	E201	Process Fluid	1.1	Carbon Steel	Yes
211	E200	Process Fluid	0.7	Carbon Steel	Yes
211	E300	Process Fluid	0.7	Carbon Steel	Yes
300	Waste	Wastewater	0.04	Carbon Steel	Yes
301	T300	Process Fluid	0.75	Carbon Steel	Yes
301	E301	Process Fluid	0.75	Carbon Steel	Yes
301	T301	Process Fluid	0.6	Carbon Steel	Yes
302	P300	Chilled Water	0.1	Carbon Steel	Yes
302	T301	Chilled Water	0.1	Carbon Steel	Yes
303	Waste	Wastewater	0.125	Carbon Steel	Yes
304	T302	Process Fluid	0.6	Carbon Steel	Yes
305	P301	50% NaOH	0.04	304 Stainless Steel	Yes
305	T302	50% NaOH	0.04	304 Stainless Steel	Yes
306	Waste	Wastewater	0.05	304 Stainless Steel	Yes
307	T303A-C	Process Fluid	0.7	Carbon Steel	Yes
308	BL300	Air	0.4	Carbon Steel	Yes

308	E302	Air	0.4	Carbon Steel	Yes
309	Atmosphere	Air	0.4	Carbon Steel	Yes
310	C400	Process Fluid	0.7	Carbon Steel	Yes
310	E400	Process Fluid	0.35	Carbon Steel	Yes
310	C401	Process Fluid	0.25	Carbon Steel	Yes
310	E401	Process Fluid	0.125	Carbon Steel	Yes
310	C402	Process Fluid	0.2	Carbon Steel	Yes
310	E402	Process Fluid	0.15	Carbon Steel	Yes
310	E403	Process Fluid	0.2	Carbon Steel	Yes
310	P400	Process Fluid	0.2	304 Stainless Steel	Yes
310	T500	Process Fluid	0.2	304 Stainless Steel	Yes
500	Pipeline	Methane Byproduct	0.02	304 Stainless Steel	No
501	T501	Process Fluid	0.2	Carbon Steel	Yes
502	P600	Process Fluid	0.125	304 Stainless Steel	Yes
503	P700	Process Fluid	0.15	Carbon Steel	Yes
503	T700	Process Fluid	0.15	Carbon Steel	Yes
600	E802	Ethylene Product	0.1	304 Stainless Steel	Yes
600	Pipeline	Ethylene Product	0.1	304 Stainless Steel	No
601	Pipeline	Ethane Byproduct	0.06	304 Stainless Steel	No
700	T701	Process Fluid	0.06	Carbon Steel	Yes
701	T800	Process Fluid	0.06	Carbon Steel	Yes
702	C700	Process Fluid	0.15	Carbon Steel	Yes
702	T702	Process Fluid	0.15	Carbon Steel	Yes

703	Pipeline	Propane Byproduct	0.06	Carbon Steel	No
704	P701	Process Fluid	0.125	Carbon Steel	Yes
704	T701	Process Fluid	0.125	Carbon Steel	Yes
705	Pipeline	Propylene	0.1	Carbon Steel	No
800	Pipeline	Butylene Product	0.1	Carbon Steel	No
801	E802	C5+ Byproduct	0.06	Carbon Steel	Yes
801	V900A-H	C5+ Byproduct	0.06	Carbon Steel	No
801	TK900A-H	C5+ Byproduct	0.06	Carbon Steel	No
900A-H	V901A-H	C5+ Byproduct	0.08	Carbon Steel	No
900A-H	Trucks	C5+ Byproduct	0.08	Carbon Steel	No
1000	E1000	Ethylene Refrigerant	0.9	304 Stainless Steel	Yes
1000	V1000	Ethylene Refrigerant	0.25	304 Stainless Steel	Yes
1001	C1000	Ethylene Refrigerant	0.6	304 Stainless Steel	Yes
1001	E1001	Ethylene Refrigerant	0.2	Carbon Steel	Yes
1001	C1001	Ethylene Refrigerant	0.2	Carbon Steel	Yes
1001	E1002	Ethylene Refrigerant	0.125	Carbon Steel	Yes
1001	E1003	Ethylene Refrigerant	0.1	Carbon Steel	Yes
1001	V1001	Ethylene Refrigerant	0.2	Carbon Steel	Yes
1001	D1000	Ethylene Refrigerant	0.45	Carbon Steel	Yes
1002	V1002	Ethylene Refrigerant	0.45	304 Stainless Steel	Yes
1002	V1000	Ethylene Refrigerant	0.5	304 Stainless Steel	Yes
1000	E500	Ethylene Refrigerant	0.06	304 Stainless Steel	Yes
1100	E1100	Propylene Refrigerant	0.9	304 Stainless Steel	Yes

1100	V1100	Propylene Refrigerant	1	304 Stainless Steel	Yes
1101	C1100	Propylene Refrigerant	1.2	304 Stainless Steel	Yes
1101	E1101	Propylene Refrigerant	0.4	Carbon Steel	Yes
1101	C1101	Propylene Refrigerant	0.35	Carbon Steel	Yes
1101	E1102	Propylene Refrigerant	0.2	Carbon Steel	Yes
1101	V1101	Propylene Refrigerant	0.08	Carbon Steel	Yes
1101	D1100	Propylene Refrigerant	0.6	Carbon Steel	Yes
1102	V1102	Propylene Refrigerant	0.6	304 Stainless Steel	Yes
1102	V1100	Propylene Refrigerant	0.7	304 Stainless Steel	Yes
1100	E600	Propylene Refrigerant	0.25	304 Stainless Steel	Yes
1100	E502	Propylene Refrigerant	0.6	304 Stainless Steel	Yes
1100	E1003	Propylene Refrigerant	0.75	304 Stainless Steel	Yes
1200	E1200	Propylene Refrigerant	0.4	304 Stainless Steel	Yes
1200	V1200	Propylene Refrigerant	0.4	304 Stainless Steel	Yes
1201	C1200	Propylene Refrigerant	0.6	304 Stainless Steel	Yes
1201	E1201	Propylene Refrigerant	0.2	Carbon Steel	Yes
1201	C1201	Propylene Refrigerant	0.2	Carbon Steel	Yes
1201	E1202	Propylene Refrigerant	0.1	Carbon Steel	Yes
1201	V1201	Propylene Refrigerant	0.06	304 Stainless Steel	Yes
1201	D1200	Propylene Refrigerant	0.25	304 Stainless Steel	Yes
1202	V1202	Propylene Refrigerant	0.3	304 Stainless Steel	Yes
1202	V1200	Propylene Refrigerant	0.35	304 Stainless Steel	Yes
1200	E403	Propylene Refrigerant	0.125	304 Stainless Steel	Yes

Table C-20. Process lines with diameters and materials of construction

The decision whether or not to insulate a given line was made on the basis of if the process would deviate from regular operation if heat was added or removed from a given stream. All pipes are fabricated from carbon steel, except for cryogenic lines, for which 304 stainless steel was selected, or lines interconnecting with the reactor-regenerator system in Stage 2, for which refractory-lined carbon steel was chosen.

Utilities

To calculate the flow rates of utilities needed for our process, we used the duties from ASPEN or calculated the heat requirement manually using energy balances, as in the case of the regenerator. For heat exchangers using the energy of vaporization, the mass flow rate of the utility was calculated by the following equation, where Q is the heat duty, m is the mass flow rate of the utility, and ΔH_{vap} is the heat of vaporization of the utility:

$$Q = m\Delta H_{vap}$$

$$m = \frac{Q}{\Delta H_{vap}}$$

As an example, we show the calculation for E202:

$$m = \frac{Q}{\Delta H_{vap}} = \frac{-5.55\text{MW}}{-1694.1\text{ kJ/kg}} = 11801\text{ kg/h}$$

For heat exchangers not employing vaporization or condensation of a utility, the mass flow rate was calculated using the utility's specific heat capacity C_p and its change in temperature ΔT :

$$Q = mC_p\Delta T$$

$$m = \frac{Q}{C_p\Delta T}$$

For example, we show the calculation for E208:

$$m = \frac{Q}{C_p\Delta T} = \frac{0.493\text{MW}}{4.184\text{ kJ/kgK}(45^\circ\text{C}-30^\circ\text{C})} = 28254\text{ kg/h}$$

Table C-21 indicates the physical properties of the utilities being used.

Utility	$\Delta H_{vaporization}$ (kJ/kg)	C_p (kJ/kgK)	T_{in} (°C)	T_{out} (°C)
High Pressure Steam (HPS)	-1694.1	N/A	254	252
High Pressure Condensate (HPC)	1694.1	N/A	254	254
Low Pressure Steam (LPS)	-2081.9	N/A	160	158
Low Pressure Condensate (LPC)	2081.9	N/A	160	160
Cooling Water (CW)	Not used	4.184	30	45
Chilled Water (CHW)	Not used	4.184	15	30
50% Glycol/Water (GL/W)	Not used	3.381 ¹	-20	10

¹Specific heat of glycol/water taken from Engineering Toolbox [75].

Table C-21. Physical properties of the utilities being used

Table C-22 indicates the utility being used and the mass flow rate for each heat exchanger using utilities in the process.

Heat Exchanger	Q (kW)	Utility	m (kg/h)
E202	5553	HPS	11801
E203	17743	HPC	37705
E204	3758	LPC	6499

E205	61864	HPC	131462
E206	28528	LPC	49330
E207	7084	CW	406354
E208	493	CW	28254
E209	675	LPS	1167
E210	5931	CW	340224
E211	5482	CW	314471
E300	3432	CW	3547
E301	11628	CHW	12020
E302	487	HPS	1035
E400	5238	CHW	5415
E401	1618	CHW	1672
E402	5843	CHW	6040
E501	3535	LPS	1858
E503	7383	LPS	11348
E601	12342	LPS	21264
E700	6658	CHW	381616
E701	6907	LPS	11514
E702	35531	HPS	75423

E703	34070	GL/W	1209650
E800	2664	CHW	152654
E801	2161	LPS	3734
E1000	235	LPS	407
E1001	1084	CHW	62175
E1002	3591	CHW	205987
E1100	132	LPS	228
E1101	6053	CHW	347215
E1102	31278	CHW	1794172
E1200	18	LPS	31
E1201	1645	CHW	1478
E1202	8076	CHW	7256

Table C-22. The exchanger duties and mass flow rates for our utility streams

APPENDIX D – ECONOMIC SAMPLE CALCULATIONS

The following sections provide sample cost calculations for different process units, along with tables that summarize the equipment and installation costs for each unit.

Capital Costs - Equipment, Installation, Piping

Heat Exchangers. Cost curves were employed to estimate the cost of heat exchangers in our process. Using heat exchange area as the key input parameter, curves from reference [76] were implemented to estimate the cost of each exchanger. The equation used is provided below:

$$C_{2010} = a + bS^n$$

Where C is the cost of the equipment, a, b, and n are costing parameters provided from the cost curve, and S is the key process parameter. As mentioned previously, the key process parameter for heat exchanger cost estimations is the total heat exchange area. The calculated cost is for 2010, as this is the year that these cost curves were created. Because the cost curves used in this process are from 2010, a CEPCI index must be used to update the cost from 2010 to 2024. The equation employed to integrate this into the analysis is shown below:

$$C_{2024} = C_{2010} \frac{Index_{2024}}{Index_{2010}}$$

Where the two indices, provided by [76] and [77], scale the 2010 cost to what the 2024 cost would be. Finally, to determine the installation cost of the exchanger, the factorial method was used, and the associated equation is provided below:

$$C_{installed} = C_{2024} [(1 + f_p)f_m + (f_{er} + f_{el} + f_i + f_c + f_s + f_l)]$$

Where $C_{installed}$ is the cost of installing the piece of equipment, and each f factor is an estimate for the cost of different steps of installation. Listing them out, f_p corresponds to piping needed for the equipment, f_m corresponds to the material of construction, f_{er} corresponds to equipment erection, f_{el} corresponds to electrical work that must be performed, f_i corresponds to instrumentation that is tied to the equipment, f_c corresponds to civil work that must be performed, f_s corresponds to structures needed to support the equipment, and f_l corresponds to lagging/paint needed for the equipment. Each f factor was obtained from [78] and the fluids only value was used. An example calculation is shown below for E701.

$$C_{2010} = 28,000 + (54)(468m^2)^{1.2} = \$114,486$$

$$C_{2024} = (\$114,486) \frac{800}{532.9} = \$171,869$$

$$C_{installed} = (\$171,689)[(1 + 0.1)(1) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$549,980$$

This process was repeated for each heat exchanger, and the results are consolidated in **Table D-1**.

Heat Exchanger	S (m²)	C₂₀₂₄³ (\$MM)	C_{installed}² (\$MM)	Total Cost (\$MM)
E200¹	11387	0.41	1.33	1.74
E201¹	59395	2.66	8.51	11.17
E202	5553	0.09	0.28	0.36
E203	17743	0.22	0.70	0.91
E204	3758	0.18	0.57	0.75
E205¹	61864	0.38	1.22	1.60
E206¹	28528	0.61	1.94	2.55
E207	7084	0.12	0.38	0.50
E208	493	0.04	0.14	0.19
E209	675	0.05	0.17	0.22
E210	5931	0.15	0.48	0.63
E211	5482	0.14	0.44	0.58
E300	3432	0.07	0.23	0.30
E301¹	11628	1.08	3.45	4.53
E302	487	0.05	0.15	0.19
E400	5238	0.24	0.77	1.01
E401	1618	0.12	0.38	0.50

E402¹	5843	0.85	2.72	3.57
E403	4842	0.11	0.35	0.46
E500	1079	0.14	0.53	0.67
E501	3535	0.05	0.19	0.25
E502	6564	0.29	1.10 ²	1.39
E503	7383	0.09	0.27	0.36
E600¹	9903	0.49	1.82	2.31
E601	12342	0.08	0.30 ²	0.39
E700	6658	0.12	0.37	0.49
E701	6907	0.17	0.55	0.72
E702	35531	0.16	0.50	0.66
E703¹	34070	0.40	1.27	1.67
E800	2664	0.06	0.19	0.25
E801	2161	0.06	0.18	0.24
E802	172	0.04	0.14	0.18
E1000	235	0.04	0.16	0.20
E1001	1084	0.05	0.20	0.25
E1002	3591	0.07	0.24	0.31
E1003	2377	0.09	0.33	0.42

E1100	132	0.04	0.16	0.20
E1101	6053	0.10	0.39	0.49
E1102¹	31278	0.65	2.09	2.74
E1200	18	0.04	0.16	0.20
E1201	1645	0.05	0.17	0.23
E1202	8076	0.16	0.51	0.67

¹Exchanger area exceeds maximum value from cost curve, cost curve was still used to provide rough estimate

²f_m = 1.3 for 304 stainless steel, 1 for carbon steel from [78]

³a = 28,000 b = 54, n = 1.2 for heat exchangers using [78]

Table D-1: Heat exchanger costing using heat transfer area, cost curves, CEPCI indices, and the factorial method

Pumps. The same approach was taken to cost pumps as outlined in the Heat Exchangers subsection of **Appendix D**. The only variations were that instead of heat exchange area being the key process parameter S, volumetric flow rate in L/s was used, and the cost curve parameters a, b, and n were changed to properly implement the pump cost curve. An example calculation is shown below for P100A, and the remainder of the pumps costed are summarized in **Table D-2**.

$$C_{2010} = \$8,000 + (240)(5.16 \frac{L}{s})^{0.9} = \$9,051$$

$$C_{2024} = (\$9,051) \frac{800}{532.9} = \$13,588$$

$$C_{installed} = (\$13,588)[(1 + 0.1)(1) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$50,819$$

Pump	V (L/s)	C ₂₀₂₄ ² (\$MM)	C _{installed} ¹ (\$MM)	Total Cost (\$MM)
P100A-B⁴	5	0.03	0.10	0.13
P101A-C⁴	66	0.08	0.27	0.35
P200	2	0.01	0.04	0.05
P201	2	0.01	0.04	0.05

P300	10	0.01	0.06	0.07
P301	10	0.01	0.06	0.07
P400	42	0.02	0.08	0.11
P600	22	0.02	0.07	0.08
P700	28	0.02	0.06	0.08
P701	9333	0.76 ³	2.43	3.18

¹f_m = 1.3 for 304 stainless steel, 1 for carbon steel from [78]

²a = 8,000 b = 240 n = 0.9 for pumps using [78]

³Costed as explosion proof motor pump using [78]

⁴Per unit mass shown, cumulative costs for all units shown

Table D-2: Pump costing using volumetric flow rate, cost curves, CEPCL indices, and the factorial method

Compressors and Blowers. The same approach was taken to cost compressors as outlined in the Heat Exchanger subsection in **Appendix D**. The only variations were that the key process parameter S was changed to input power in kW for compressors and volumetric flow rate in m³/h for blowers, and that the cost curve parameters a, b, and n were changed to properly implement the compressor/blower cost curves. An example calculation is shown below for C200, and the remainder of the compressor and blower costs are summarized in **Table D-3**.

$$C_{2010} = 580,000 + (20,000)(5490 \text{ kW})^{0.6} = \$4,085,777$$

$$C_{2024} = (\$4,085,777) \frac{800}{532.9} = \$6,133,649$$

$$C_{\text{installed}} = (\$6,133,649)[(1 + 0.1)(1) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$19,627,678$$

Compressor/Blower	Power (kW)	C ₂₀₂₄ ^{2, 3} (\$MM)	C _{installed} ¹ (\$MM)	Total Cost (\$MM)
C200	5490	6.13	19.63	25.76
C201	5871	6.35	20.32	26.67
C202	5279	6.01	19.24	25.25
C203	5893	6.36	20.36	26.72

BL200	4263	0.83	2.66	3.49
BL300	287	0.12	0.39	0.51
C400	4747	5.69	18.22	23.91
C401	1524	3.31	10.59	13.90
C402	915	2.67	8.53	11.20
C700	502	2.12	6.80	8.92
C1000	3510	4.89	18.31	23.20
C1001	2232	3.94	12.60	16.54
C1100	13319	9.83	36.76	46.58
C1101	5341	6.05	19.35	25.40
C1200	3457	4.86	18.17	23.03
C1201	1414	3.20	10.25	13.45

¹f_m = 1.3 for 304 stainless steel, 1 for carbon steel from [78]

²a = 580,000 b = 20,000 n = 0.6 for compressors using [78]

³a = 4,450 b = 57 n = 0.8 for blowers using [78]

Table D-3: Compressor/blower costing using power/volumetric flow rate, cost curves, CEPCI indices, and the factorial method

Trayed Columns. Trayed columns were costed in a variety of different ways in this economic analysis, largely due to the fact that the columns often exceeded the maximum weight that could be inputted into a cost curve or other online costing resources. This appendix section details three different methods employed to cost the nine total trayed columns in the proposed process. For T301, T302, and T800 the same approach was taken as previous subsections Heat Exchangers, Pumps, and Compressors and Blowers. S was determined to be the mass of the column, and to cost the internal trays of the column, S was the diameter of the tray. Shown below is the formula used to calculate the mass of the column.

$$m_{column} = \frac{\pi L(D^2 - d^2)}{4} \rho$$

Where m_{column} is the mass of the column shell in kg, L is the length of the column in m, D is the outer diameter of the column in m, d is the inner diameter of the column in m, and ρ is the density of the shell material in kg/m³. The two materials utilized in this process are 304 stainless steel and carbon steel, which have densities of 7,930 and 7,800 kg/m³, respectively [79],[80]. The material of the trayed column will also impact the parameters a, b, and n, and these values are reported in footnotes in **Table D-4**, which can be found at the end of this section.

An example calculation is shown below for the shell of T301.

$$m_{column} = \frac{\pi(20.72m)(0.821^2 m^2 - 0.773^2 m^2)}{4} (7,800 \frac{kg}{m^3}) = 9,717 \text{ kg}$$

$$C_{2010} = 11,600 + (34)(9,717 \text{ kg})^{0.85} = \$94,948$$

$$C_{2024} = (\$94,948) \frac{800}{532.9} = \$142,537$$

$$C_{installed} = (\$142,537)[(1 + 0.1)(1) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$456,119$$

T301's internals are comprised of 29 trays, which as previously mentioned, are costed using the tray diameter. In the proposed design, the tray diameter is identical to the inner diameter of the column, so this value was used. The cost of a tray was multiplied by the number of trays in the column. The sample calculation to cost T301 continues below.

$$C_{2010,tray} = 130 + (440)(0.773 \text{ m})^{1.8} = \frac{\$407}{tray}$$

$$C_{2010,total} = (\frac{\$407}{tray})(29 \text{ trays}) = \$11,801$$

$$C_{2024} = (\$11,801) \frac{800}{532.9} = \$17,716$$

$$C_{installed} = (\$17,716)[(1 + 0.1)(1) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$56,691$$

T500, T501, and T700 exceeded the maximum shell weight S that was recommended by [78], so Matches Pressure Vessel Costing[81] was used instead to estimate the cost of the shell by inputting the weight in pounds into the appropriate input field. A screenshot is provided below in **Figure D-1** to cost the shell of T501.

The screenshot shows a web page for 'Matches' with a blue header containing the company logo and navigation links for Home, About, News, Products, Services, Calendar, and Contact. In the top right corner, there is contact information: office (405) 340-2673, fax (866) 583-3035, and toll-free (800) 525-6277. The main content area features a sidebar titled 'Equipment Type' listing various industrial equipment categories. The main form area contains fields for Vessel Type (set to 'Column, No Internals, Large'), Weight (set to 333133 pounds), and Material (set to 'Stainless Steel 304'). A button labeled 'Select For Cost' is present. Below these fields, the cost is displayed as \$1,999,600. The text 'Cost 2014 US \$: 1999600' is highlighted in red. The form also includes a note about F.O.B. Gulf Coast U.S.A. and a section for 'Cost Information Exchange'.

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Figure D-1. Costing of T501 shell using Matches Pressure Vessel Costing

Because this source reports the cost in 2014 dollars, a CEPCI index collected from [77] must be used to convert this amount into 2024 dollars, and this conversion is shown below.

$$C_{2024} = (\$1,999,600) \frac{800}{576.8} = \$2,773,232$$

The trays of T500, T501, and T700 were all costed using the same method that was used for T301, T302, and T800.

The maximum weight that could be inputted into Matches Pressure Vessel Costing was 1 MM lbs, which is lower than the shell weights of T600, T701, and T702. The costs of these columns were approximated using a literature source [35] that cost each of the columns inherent to the MtO process, including the C3 splitter and butene distillation column. As the empirical correlations found in this paper were derived from columns performing similar functions to the ones that exceeded the maximum weight permissible for the Matches resource, these empirical correlations were used to cost these three columns. The assumption was made from the paper that this cost included installation and trays, so only one calculation was performed to cost each column. The equation provided by the paper and employed by our team is provided below.

$$C_{2016} = C_0 \left(\frac{S^f}{S_0} \right)$$

Where C_0 is the base unit cost in \$MM, S_0 is defined as the base capacity, which is the flow rate in kg/s through the unit in the reference state, s_f is the scaling factor used to calculate the cost at different flow rates from the reference state, and S is the key process parameter, defined in units of kg/s feed in the paper. Because the paper is from 2016, a CEPCI cost index from [77] must be employed to convert the cost into 2024 dollars. An example calculation is shown below for T600.

$$C_{2016} = 3.08\text{MM} \left(\frac{(200 \text{ kg/s})^{0.8}}{2.83} \right) = \$6,302,365$$

$$C_{2024} = (\$6,302,365) \frac{800}{541.7} = \$9,307,535$$

Again, this cost derived from the paper includes installation costs and trays, so no further calculations were performed to cost T600, T701, and T702. **Table D-4** below provides the cost of each column, along with key parameters and assumptions used to estimate these costs.

Trayed Vessel	Shell Mass (kg)	C_{2024} (Shell) (\$MM)	$C_{\text{installed}}$ (Shell) (\$MM)	C_{2024} ³ (Trays) (\$MM)	$C_{\text{installed}}$ (Trays) (\$MM)	Total Cost (\$MM)
T301 ^{1,6}	9717	0.14	0.46	0.02	0.06	0.67
T302 ^{2,6}	5733	0.21	0.79	0.02	0.06	1.08
T500 ⁶	124151	2.33 ⁴	8.73	0.16	0.58	11.80
T501 ⁶	151415	2.77 ⁴	10.37	0.19	0.71	14.04
T600 ⁵	571717	9.31	9.06	NA	NA	18.36
T700 ⁶	265981	1.62 ⁴	5.20	0.33	1.05	8.20

T701 ⁵	2290013	3.05	3.05	NA	NA	3.05
T702 ⁵	1681858	3.05	3.05	NA	NA	3.05
T800 ^{1,6}	23991	0.29	0.92	0.09	0.30	1.59

¹a = 11,600 b = 34 n = 0.85 for carbon steel columns using [78]

²a = 17,400 b = 79 n = 0.85 for 304 stainless steel columns using [78]

³a = 130 b = 440 n = 1.8 for column trays using [78]

⁴Cost estimated using [81]

⁵Cost estimated using [35], $C_0 = 3.08 S_0 = 2.83$, sf = 0.67 for C2 distillation, $C_0 = 2.20 S_0 = 4.33$, sf = 0.64 for C3 distillation column

⁶For installation costs, $f_m = 1.3$ for 304 stainless steel, 1 for carbon steel from [78]

Table D-4: Trayed vessel costing using a variety of methods[82]

Reactor and Regenerator. Due to the complexity of the reactor-regenerator system, costing of this unit followed the same approach used to cost T600, T701, and T702 in the Trayed Columns subsection of **Appendix D**. The same source [35] and the same equations used in the aforementioned section were utilized to calculate the equipment and installation cost of the reactor regeneration system. Again, this estimation included piping and installation costs, so the factorial method was not used to calculate this value. This calculation is provided below and is accompanied by a CEPCI index conversion to convert the cost into 2024 dollars.

$$C_{2015} = 65.66 \text{MM} \left(\frac{(52.46 \text{kg/s})^{0.65}}{10.65} \right) = \$81,039,705$$

$$C_{2024} = (\$81,039,705) \frac{800}{556.8} = \$116,436,357$$

For completeness, **Table D-5** provides the cost of the reactor regenerator unit.

Reaction Vessel	C_{2024}^1 (\$MM)
R100, R101	116.44

¹Cost estimated using [35] $C_0 = 6.66 S_0 = 10.63$, sf = 0.65 for methanol-to-olefins unit

Table D-5: Reactor and regenerator costing

Flash Drums, Dryers, and Pressurized Storage Tanks. The same approach was taken to cost the remainder of the pressure vessels in the process as outlined in the *Heat Exchanger* subsection in **Appendix D**. The variations in this process are that the mass of the shell was used as the key process parameter S, and a, b, and n were changed to reflect the cost curve utilized for pressure vessels. Mass of the shell was calculated using

the same formula used for T301, T302, and T800 in the *Trayed Columns* section above. An example calculation is shown below for the 304 stainless steel flash drum D1000, and the pressurized vessel equipment costing is summarized in **Table D-6**.

$$m_{\text{column}} = \frac{\pi(8.65m)(2.88^2 m^2 - 2.87^2 m^2)}{4} (7,930 \frac{\text{kg}}{m^3}) = 557 \text{ kg}$$

$$C_{2010} = 17,400 + (79)(557 \text{ kg})^{0.85} = \$34,448$$

$$C_{2024} = (\$34,448) \frac{800}{532.9} = \$51,637$$

$$C_{\text{installed}} = (51,637)[(1 + 0.1)(1.3) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$193,122$$

Pressure Vessel	Shell Mass (kg)	$C_{2024}^{1,2}$ (\$MM)	$C_{\text{installed}}^5$ (\$MM)	Total Cost (\$MM)
T300	2654	0.06	0.19	0.25
D1000	557	0.05	0.19	0.24
D1100	2009	0.10	0.38	0.48
D1200	257	0.04	0.15	0.19
T303A-C ³	18116	0.69	2.21	2.95 ⁴
TK900A-H ³	24958	2.37	7.59	9.96

¹a = 11,600 b = 34 n = 0.85 for carbon steel pressure vessels using [78]

²a = 17,400 b = 79 n = 0.85 for 304 stainless steel pressure vessels using [78]

³Per unit mass shown, cumulative costs for all units shown

⁴Packing cost estimated using [83], added to total cost

⁵For installation costs, f_m = 1.3 for 304 stainless steel, 1 for carbon steel from [78]

Table D-6: Pressure vessel costing using cost curves, CEPCI indices, and the factorial method

Atmospheric Tanks. The same approach was taken to cost atmospheric storage tanks as outlined in the Heat Exchangers subsection of **Appendix D**. The only variations were that instead of heat exchange area being the key process parameter S , storage capacity in m^3 was used, and the cost curve parameters a , b , and n were changed to properly implement the storage tank cost curves. An example calculation is shown below for the 304 stainless steel storage tank TK300, and the remainder of the tanks cost are summarized in **Table D-7**.

$$C_{2010} = 5,800 + (1,600)(628 m^3)^{0.7} = \$151,295$$

$$C_{2024} = (\$151,295) \frac{800}{532.9} = \$227,126$$

$$C_{\text{installed}} = (\$227,126)[(1 + 0.1)(1.3) + (0.3 + 0.2 + 0.3 + 0.3 + 0.2 + 0.1)] = \$849,451$$

Storage Tank	Storage Capacity (m^3)	C_{2024} (\$MM)	$C_{\text{installed}}^4$ (\$MM)	Total Cost (\$MM)
TK100A-C ^{1,3}	7644	5.40	17.29	22.71
TK300 ²	628	0.23	0.85	1.077

¹a = 113,000 b = 3,250 n = 0.65 for floating roof tanks using [78]

²a = 5,800 b = 1,600 n = 0.7 for cone roof tanks using [78]

³Per unit mass shown, cumulative costs for all units shown

⁴For installation costs, $f_m = 1.3$ for 304 stainless steel, 1 for carbon steel from [78]

Table D-7: Atmospheric storage tank costing using cost curves, CEPCI indices, and the factorial method

Total Production Costs - Materials, Utilities, Waste, and Operating Labor

The following sections provide detailed tables for different aspects of the direct production cost, including materials, utilities, waste, and operating labor.

Materials. The relevant materials that must be costed for this process include methanol feedstock, catalyst to accommodate attrition in the reactor regenerator units, 50% NaOH, and process water. **Table D-8** below summarizes the amounts, per unit costs, and total costs of each of these materials.

Material	Amount Required (kg/h)	Per Unit Cost	Total Hourly Cost (\$)	Total Yearly Cost ⁵ (\$MM)
Methanol ¹	188884	\$400.00/MT	\$75,554	\$617.52
SAPO-34 ²	7	\$34.00/kg	\$241	\$1.97
		\$1,619.81/27		
50% NaOH ³	10000	5 gal	\$1,183	\$9.66
		\$0.06/1000		
Process water ⁴	50100	kg	\$3,206	\$26.16

¹Per unit cost estimated from [38]

²Per unit cost estimated from [84]

³Per unit cost estimated from [85]

⁴Per unit cost estimated from [39]

⁵Assuming 340 days/yr operation

Table D-8: Production costs of process materials

Utilities. A variety of utilities are required to operate this process. They include electricity, low-pressure and high-pressure steam, cooling and chilled water, and ethylene glycol. Ethylene glycol exclusively services E703, which has a duty of 34070 kW, as shown in the **Heat Exchangers** subsection of **Appendix C**. Over one hour, this equates to 123 GJ, which is a necessary calculation to perform to cost the utility. **Table D-9** below details the required capacity of each of these utilities, as well as their per unit and total costs.

Utility	Amount Required	Per Unit Cost	Total Hourly Cost (\$)	Total Yearly Cost ³ (\$MM)
Electricity ¹	64426 kW	\$0.03/kWh	\$1,933	\$15.77
		\$8.28/1000		
LP Steam ²	52200 kg/h	lbs	\$951	\$7.76
		\$11.27/1000		
HP Steam ²	125964 kg/h	lbs	\$3,123	\$25.48

		\$0.08/1000		
Cooling Water ²	1094522 kg/h	gal	\$23	\$0.19
		\$1.73/1000		
Chilled Water ²	2976048 kg/h	gal	\$1,358	\$11.09
		123 GJ heat removed	\$35.20/GJ removed	
Ethylene Glycol ²			\$4,317	\$35.23

¹Per unit cost estimated from [86]

²Per unit cost estimated from [15]

³Assuming 340 days/yr operation

Table D-9: Production costs of utilities

Waste. The proposed process generates both nonhazardous and hazardous wastewater that must be treated. This is accounted for as a production cost and the amount of waste generated as well as the costs associated with its treatment on an hourly and annual basis are provided in **Table D-10**.

Wastewater Type	Amount Generated (kg/h)	Per Unit Treatment Cost	Total Hourly Cost (\$)	Total Yearly Cost ² (\$MM)
Nonhazardous ¹	150865	\$27.50/MT	\$4,149	\$33.85
Hazardous ¹	11176	\$0.92/lb	\$4,674	\$38.14

¹Per unit cost estimated from [15]

³Assuming 340 days/yr operation

Table D-10: Production costs of waste treatment

Labor. The last major production cost associated with the proposed process is labor. The first step in calculating our labor costs is to determine the amount of operators we would need for any given shift. To accomplish this task, we estimated our operator labor requirements based on industry heuristics and input from an industry professional [87]. However, we do not necessarily need to maintain constant labor levels over weekends. Accordingly, we used industry heuristics to determine that we will not have lab techs or engineers work on weekends but that we will keep the number of maintenance workers and operators constant [87]. Ultimately, we decided that we need 12 operators working at any given time. In addition to operators, we also need to consider maintenance workers, lab techs, and engineers. Heuristics typically recommend having one maintenance worker for every 50,000 sq ft of plant; our plant is approximately 150,000 sq ft, so we decided to have 3 maintenance workers per shift. We previously assumed that our lab costs would be relatively small; with this in mind, we decided to only have 2 lab techs per shift. Lastly, we decided to have 2 engineers per shift so that one can engage in managerial duties while the other performs more traditional engineering work. With the number of workers at any given time determined, we then need to determine the cost associated with each position. **Figure D-2** below depicts our assumptions on the labor costs associated with each position.

Role	\$/hr
Operator	54.00
Maint	63.00
Lab	60.00
Super	93.00
Engineer	127.50

Figure D-2. The labor costs per hour associated with each position

Using these values and number of workers in each position, we can determine the total labor costs per hour on weekends and weekdays.

$$C_{\text{weekday}} = (12 * \$54.00) + (3 * \$63.00) + (2 * 60.00) + (2 * \$127.50) = \frac{\$1212}{hr}$$

$$C_{\text{weekend}} = (12 * \$54.00) + (3 * \$63.00) = \frac{\$837}{hr}$$

Since the plant operates for 340 days a year, we estimate the total number of weekdays as being 5/7 of 340, or 243 days. Similarly, we estimate the total number of weekend days as being 2/7 of 340, or 97 days. With these values, we can determine the total yearly labor costs for weekdays and weekends and then sum these values to find total yearly labor costs.

$$C_{weekday,yr} = \frac{\$1212}{hr} * \frac{24hr}{1 day} * 243 days * 10^{-6} = \$7.06 MM$$

$$C_{weekend,yr} = \frac{\$837}{hr} * \frac{24hr}{day} * 97 * 10^{-6} = \$1.95 MM$$

$$C_{total} = \$7.06 MM + \$1.95 MM = \$9.01 MM$$

Table D-11 below summarizes the labor costs for our plant.

Position	Number of Employees (# on weekday/# on weekend)	Hourly Cost (\$/h)	Annual Cost for Weekdays (\$MM/yr)	Annual Cost for Weekends (\$MM/yr)	Total Annual Cost (\$MM/yr)
Operator	12/12	54.00	3.78	1.51	5.29
Maintenance	3/3	63.00	1.10	0.44	1.54
Lab Tech	2/0	60.00	0.70	0	0.70
Engineer	2/0	127.50	1.48	0	1.48
Total					9.02

Table D-11: Labor costs of the proposed process

Total Production Revenue - Materials and Utilities

The following sections provide detailed tables for different aspects of the total production revenue, namely material production and utility generation.

Materials. The majority of the revenue generated by this process stems from the production of purified hydrocarbons. Because the saturated hydrocarbons methane, ethane, and propane are sold as natural gas per MMBTU, an equation is provided below to show how each revenue stream was calculated.

$$E = mED\left(\frac{.000948 \text{ MMBTU}}{1 \text{ MJ}}\right)$$

Where E is the total energy that can be extracted from the material in MMBTU/h, m is the mass flow rate of the product stream in kg/h, and ED is the energy density of the material in MJ/kg. An example calculation is shown for methane, which has an ED value of 55.5 MJ/kg [88].

$$E = (1,808 \frac{\text{kg}}{\text{h}})(55.5 \frac{\text{MJ}}{\text{kg}})(\frac{.000947 \text{ MMBTU}}{1 \text{ MJ}}) = 95 \frac{\text{MMBTU}}{\text{h}}$$

Multiplying this result by the per unit price of methane (\$3.10/MMBTU)[15], we get that the total hourly revenue of methane is \$295.

Table D-12 below summarizes the amounts of each product produced, the per unit prices of these products, as well as the hourly and annual revenue accumulated from each stream.

Material	Sold As	Amount Produced (kg/h)	Per Unit Price ¹	Total Hourly Revenue (\$)	Total Yearly Revenue ⁵ (\$MM)
Methane ²	Natural gas	1808	\$3.10/MMBTU	\$295	\$2,403,771.83
Ethane ³	Natural gas	2155	\$3.10/MMBTU	\$332	\$2,710,244.11
Propane ⁴	Natural gas	1771	\$3.10/MMBTU	\$237	\$1,934,573.50
Ethylene	Polymer grade	30184	\$1,189.00/MT	\$35,889	\$292,852,412.16
Propylene	Polymer grade	30429	\$1,012.00/MT	\$30,794	\$251,280,247.68
Butylene	Polymer grade	8562	\$1,200.00/MT	\$10,274	\$83,839,104.00
C5+	Light naphtha	3713	\$1,080.00/MT	\$4,010	\$32,721,926.40

¹Per unit prices estimated from [39]

²Value of 55.5 MJ/kg used for ED, adapted from[15]

³Value of 52.5 MJ/kg used for ED, adapted from [89]

⁴Value of 45.6 MJ/kg used for ED, adapted from [90]

⁵Assuming 340 days/yr operation

Table D-12: Production revenue from materials

Utilities. Some additional revenue is brought in by the process by using excess heat to create low-pressure and high-pressure steam. The amounts generated, as well as the hourly and annual revenue that result from this utility generation is provided below in **Table D-13**.

Utility Generated	Amount Generated (kg/h)	Per Unit Price ¹	Total Hourly Revenue (\$)	Total Yearly Revenue ² (\$MM)

Low-pressure				
steam	55829	\$8.28/MT	\$1,017	\$8.30
High-pressure		\$11.27/1000		
steam	169167	lbs	\$4,674	\$34.23

¹Per unit cost estimated from [15]

²Assuming 340 days/yr operation

Table D-13: Production revenue from utility generation

Cash Flow Tables for Different Economic Scenarios

The following sections provide sample cash flow tables for the baseline scenario as well as a scenario where methanol becomes considerably cheaper and the process becomes profitable. This includes production costs, revenue, taxes, depreciation, and a hurdle rate of 11%. To calculate depreciation, we used a 20 year MACRS schedule, depicted below in **Figure D-3**.

Year	3 years	5 years	7 years	10 years	15 years	20 years
1	33.33%	20.00%	14.29%	10.00%	5.00%	3.750%
2	44.45%	32.00%	24.49%	18.00%	9.50%	7.219%
3	14.81%	19.20%	17.49%	14.40%	8.55%	6.677%
4	7.41%	11.52%	12.49%	11.52%	7.70%	6.177%
5		11.52%	8.93%	9.22%	6.93%	5.713%
6		5.76%	8.92%	7.37%	6.23%	5.285%
7			8.93%	6.55%	5.90%	4.888%
8			4.46%	6.55%	5.90%	4.522%
9				6.56%	5.91%	4.462%
10				6.55%	5.90%	4.461%
11				3.28%	5.91%	4.462%
12					5.90%	4.461%
13					5.91%	4.462%
14					5.90%	4.461%
15					5.91%	4.462%
16					2.95%	4.461%
17						4.462%
18						4.461%
19						4.462%
20						4.461%
21						2.231%

Figure D-3. 20-year MACRS schedule that we used to calculate depreciation[91]

We can use the percentages provided in **Figure D-3** to determine depreciation for each year of the project's life. With the depreciation for each year determined, we can find the taxable income by subtracting depreciation from yearly profit. With the taxable income determined, we can then determine the tax due for each year by multiplying taxable income by the tax rate (0% for the baseline scenario, 25% for Scenarios One and Two [40]). We can then determine the undiscounted yearly cash flow for our process. To determine the discounted cash flow, we can multiply the undiscounted cash flow by a discount factor.

$$\text{discount factor} = \frac{1}{(1+a)^N}$$

Where a is the hurdle rate (11% in our case) and N is the year of the project. With the discounted cash flow determined, we then calculate cumulative cash flow for a given project year by summing the discounted cash flows of the years leading up to it. This process was used to determine the values in both of the following cash flow tables.

Baseline Scenario. A cash flow table for our process is depicted below in **Table D-14**.

Year	Revenue (\$MM)	Total Production Costs (\$MM)	Depreciation (\$MM)	Taxable Income (\$MM)	Tax Due (\$MM)	Undiscounted Cash Flow (\$MM)	Discounted Cash Flow (\$MM)	Cumulative Cash Position (\$MM)
0	0	586.73 ¹						-587
1	710.27	1117.16 ²	22.00	-258.75	0	-236.74	-213.28	-970
2	710.27	947.01	42.36	-279.10	0	-236.74	-192.15	-1,183
3	710.27	947.01	39.18	-275.91	0	-236.74	-173.11	-1,376
4	710.27	947.01	36.24	-272.99	0	-236.74	-155.95	-1,549
5	710.27	947.01	33.52	-270.26	0	-236.74	-140.50	-1,705
6	710.27	947.01	31.01	-267.75	0	-236.74	-126.57	-1,845
7	710.27	947.01	28.68	-265.42	0	-236.74	-114.03	-1,972
8	710.27	947.01	26.53	-263.28	0	-236.74	-102.73	-2,086
9	710.27	947.01	26.18	-262.92	0	-236.74	-92.55	-2,188
10	710.27	947.01	26.17	-262.92	0	-236.74	-83.38	-2,281
11	710.27	947.01	26.18	-262.92	0	-236.74	-75.11	-2,364
12	710.27	947.01	26.17	-262.92	0	-236.74	-67.67	-2,440
13	710.27	947.01	26.18	-262.92	0	-236.74	-60.97	-2,507
14	710.27	947.01	26.17	-262.92	0	-236.74	-54.92	-2,568
15	710.27	947.01	26.18	-262.92	0	-236.74	-49.48	-2,623
16	710.27	947.01	26.17	-262.92	0	-236.74	-44.58	-2,673

17	710.27	947.01	26.18	-262.92	0	-236.74	-40.16	-2,717
18	710.27	947.01	26.17	-262.92	0	-236.74	-36.18	-2,757
19	710.27	947.01	26.18	-262.92	0	-236.74	-32.59	-2,793
20	710.27	947.01	26.17	-262.912	0	-236.74	-29.36	-2,826
21	710.27	947.01	13.09	-249.83	0	-236.74	-26.45	-2,855

¹Cost is FCI

²Cost includes startup costs, working capital, and annual operating cost

Table D-14. Cash flow table of the baseline scenario of our current process

Scenario One. Scenario One is the first hypothetical scenario for our process that sees a reduction of methanol cost from \$400/MT to \$265/MT and a 15% decrease in operating costs for utilities and waste treatment. A cash flow table for Scenario One of our process is depicted below in **Table D-15.**

Year	Revenue (\$MM)	Total Production Costs (\$MM)	Depreciation (\$MM)	Taxable Income (\$MM)	Tax Due (\$MM)	Undiscounted Cash Flow (\$MM)	Discounted Cash Flow (\$MM)	Cumulative Cash Position (\$MM)
0		586.73 ¹						-\$587
1	710.27	756.85 ²	22	3.28	0.82	24.46	22.04	-\$735
2	710.27	684.99	42.36	-17.08	0	25.28	20.52	-\$713
3	710.27	684.99	39.18	-13.90	0	25.28	18.48	-\$692
4	710.27	684.99	36.24	-10.96	0	25.28	16.65	-\$674
5	710.27	684.99	33.52	-8.24	0	25.28	15	-\$657
6	710.27	684.99	31.01	-5.73	0	25.28	13.52	-\$642
7	710.27	684.99	28.68	-3.40	0	25.28	12.18	-\$629
8	710.27	684.99	26.53	-1.25	0	25.28	10.97	-\$616

9	710.27	684.99	26.18	-0.90	0	25.28	9.88	-\$605
10	710.27	684.99	26.17	-0.89	0	25.28	8.9	-\$596
11	710.27	684.99	26.18	-0.90	0	25.28	8.02	-\$587
12	710.27	684.99	26.17	-0.89	0	25.28	7.23	-\$579
13	710.27	684.99	26.18	-0.90	0	25.28	6.51	-\$571
14	710.27	684.99	26.17	-0.89	0	25.28	5.86	-\$565
15	710.27	684.99	26.18	-0.90	0	25.28	5.28	-\$559
16	710.27	684.99	26.17	-0.89	0	25.28	4.76	-\$554
17	710.27	684.99	26.18	-0.90	0	25.28	4.29	-\$549
18	710.27	684.99	26.17	-0.89	0	25.28	3.86	-\$545
19	710.27	684.99	26.18	-0.90	0	25.28	3.48	-\$541
20	710.27	684.99	26.17	-0.89	0	25.28	3.14	-\$537
21	710.27	684.99	13.09	12.19	3.05	22.23	2.48	-\$534

¹Cost is FCI

²Cost includes startup costs and working capital

Table D-15. Cash flow table of Scenario One

Scenario Two. Scenario Two is the second hypothetical scenario for our process that sees a 20% reduction of methanol cost from \$265/MT to \$212/MT and a 15% decrease in operating costs for utilities and waste treatment. A cash flow table for Scenario Two of our process is depicted below in **Table D-16**.

Year	Revenue (\$MM)	Total Production Costs (\$MM)	Depreciation (\$MM)	Taxable Income (\$MM)	Tax Due (\$MM)	Undiscounted Cash Flow (\$MM)	Discounted Cash Flow (\$MM)	Cumulative Cash Position (\$MM)
0		564.91						-\$565
1	710.27	756.85	21.18	96.07	24.0175	93.2325	83.9932432	-\$645
2	710.27	593.02	40.78	76.47	19.1175	98.1325	79.6465384	-\$561
3	710.27	593.02	37.72	79.53	19.8825	97.3675	71.1942768	-\$481
4	710.27	593.02	34.89	82.36	20.59	96.66	63.672936	-\$410
5	710.27	593.02	32.27	84.98	21.245	96.005	56.9742948	-\$346
6	710.27	593.02	29.86	87.39	21.8475	95.4025	51.0060724	-\$289
7	710.27	593.02	27.61	89.64	22.41	94.84	45.6804837	-\$238
8	710.27	593.02	25.55	91.7	22.925	94.325	40.9301168	-\$193
9	710.27	593.02	25.21	92.04	23.01	94.24	36.8407505	-\$152
10	710.27	593.02	25.20	92.05	23.0125	94.2375	33.1889848	-\$115
11	710.27	593.02	25.21	92.04	23.01	94.24	29.9007795	-\$82
12	710.27	593.02	25.20	92.05	23.0125	94.2375	26.9369246	-\$52
13	710.27	593.02	25.21	92.04	23.01	94.24	24.2681434	-\$25
14	710.27	593.02	25.20	92.05	23.0125	94.2375	21.8626123	-\$1
15	710.27	593.02	25.21	92.04	23.01	94.24	19.6965696	\$21
16	710.27	593.02	25.20	92.05	23.0125	94.2375	17.7441866	\$41
17	710.27	593.02	25.21	92.04	23.01	94.24	15.9861778	\$59
18	710.27	593.02	25.20	92.05	23.0125	94.2375	14.4015799	\$75

19	710.27	593.02	25.21	92.04	23.01	94.24	12.9747405	\$89
20	710.27	593.02	25.20	92.05	23.0125	94.2375	11.6886453	\$102
21	710.27	593.02	12.60	104.65	26.1625	91.0875	10.178323	\$114

¹Cost is FCI

²Cost includes startup costs and working capital

Table D-16. Cash flow table of Scenario Two

To calculate the IRR of Scenario Two, the cash flow table was entered in Excel, and the NPV was set to 0 using goal seek by manipulating the hurdle rate. The resulting sheet is shown in **Figure D-4**.

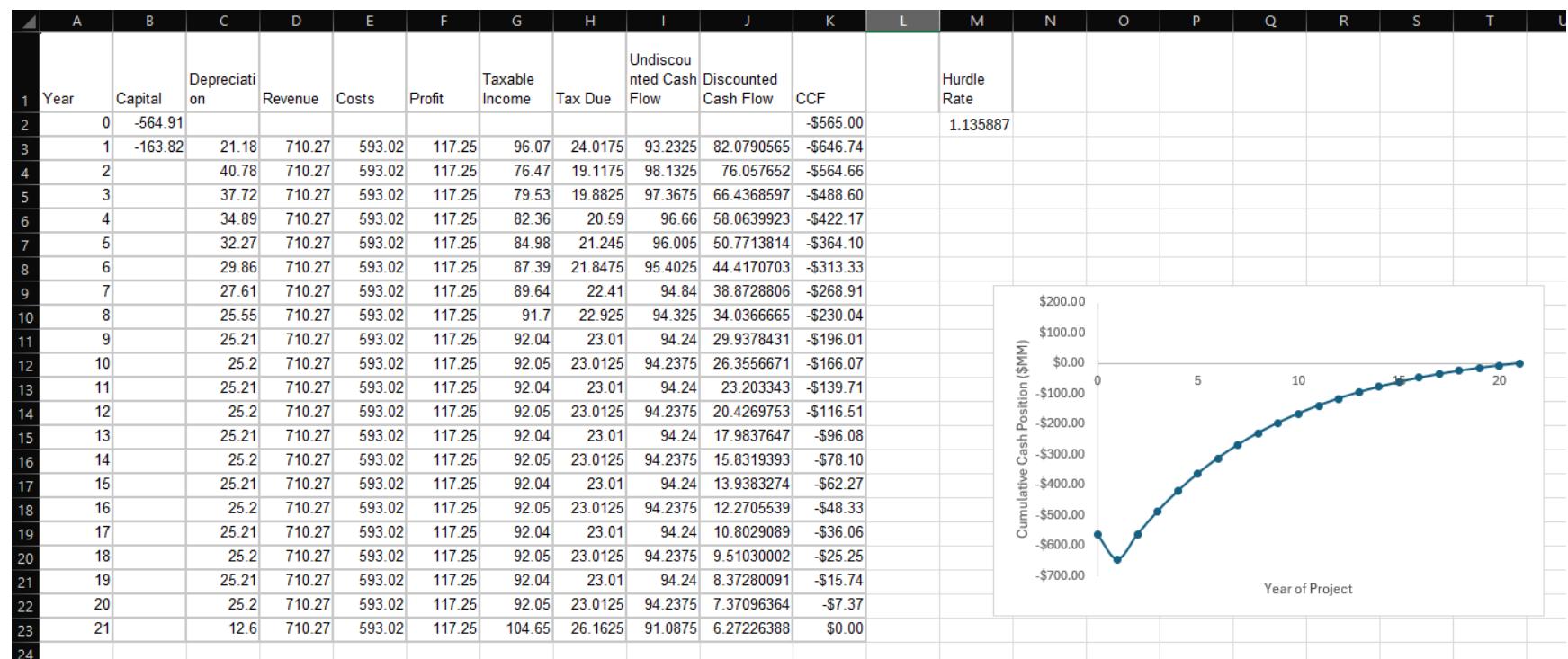


Figure D-4. Cash flow table of Scenario Two after calculating IRR

The goalseek set the hurdle rate (cell M2) to 1.136, corresponding to an IRR of 13.6%. The DPBP is simply the year where the NPV goes from negative to positive, and in **Table D-16** we see that this occurs in year 14.

Revenue Source	\$MM/yr
Materials	667.89
Methane	2.40
Ethane	2.86
Propane	1.93
Ethylene	292.85
Propylene	251.28
Butylene	83.84
C5+	32.72
Utilities	42.52
Low Pressure Steam	8.30
High Pressure Steam	34.23
Total Production Revenue	710.41