MEMORANDUM

DATE: November 30, 2017

TO: Mr. Refinery Person, Plant Manager, Zontec Refining Corp.

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SUBJECT: Successful Preliminary Design Study of Lavender Parish Propylene Fractionation System Modification

As desired, a propylene splitter with a heat pump system was designed to produce both polymer-grade propylene and propane HD-5 at a rate of 5,501 and 576 barrels/day, respectively. The process was modeled, simulated, and optimized using Aspen HYSYS. Hand calculations were used to validate these results.

Economic analyses were conducted based on 10-year MACRS depreciation over a 10-year evaluation life starting in the year 2020. The estimated net present value is $80,400,000 with discounted cash flow rate of return 85%. Based on both NPV and DCFROR analyses, the fractionation system modification was determined to be economically attractive.

A detailed report of the preliminary design study is attached to this memo.

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Tables and Figures

Executive Summary

The objective of the proposed design was to add to the Lavender Parish Refinery's existing fractionation system. The system currently runs a Liquefied Petroleum Gasoline (LPG) feed stream through an alkylation unit to produce propane and high-octane gasoline. In order to meet the US Gulf Coasts' increasing demand for Polymer Grade Propylene, Zontec Refining Corporation has commissioned a design to split the existing LPG feed stream between the current alkylation unit and an additional propylene splitter. The propylene splitter could be designed using either a high pressure system or a heat pump system. The propylene splitter model in this report operates using a heat pump system. Other groups were assigned to model the high pressure system. The proposed design will produce both Polymer Grade Propylene (99.5vol% propylene) and Propane HD-5 (95vol% propane and 5vol% propylene).

Polymer Grade Propylene is of high economic value to the company, and has a sales value of $75/barrel. The proposed design will produce 5501 barrels of Polymer Grade Propylene daily, resulting in an estimated annual revenue of $150,600,000. The company is also expected to profit off of the sales of Propane HD-5, with a sales value of $42/barrel. This system produces 576 barrels of Propane HD-5 daily, resulting in an estimated annual revenue of $8,800,000. Taking into account a service factor of 95%, the total estimated revenue of the proposed design is $151,430,000 annually. The DCFROR of the proposed design was estimated to be 85%, with an NPV of $80,400,000.

Introduction

Zontec Refining Corporation determined that there was an existing need to increase current production of Polymer Grade Propylene in order to meet the US Gulf Coasts' increasing demand. It was decided that this need would be met by expanding on the Lavender Parish's existing fractionation system. The existing process utilized mixed hydrocarbons to produce gasoline. The proposed change entailed the addition of two distillation columns to the existing process. One column would separate a propylene/propane mixture from the liquefied petroleum gas feed stream, and the second would separate propylene and propane. The main focus of this design was the propylene/propane splitter. The preliminary design of this process will include a simulation of the process, an equipment list, design specifications, and an economic analysis.

Polymer Grade Propylene is of high economic value to the company. The intent of the proposed process was to produce propylene with a purity of 99.5vol%. Propylene with this purity is valued at $75/barrel. It was anticipated that the production of Propane HD-5 would also bring economic value to the company. Propane HD-5 is valued at $42/barrel. The production of both of these components was expected to increase the company's annual revenue.

Design Basis

It was the intent of Zontec Refining Company to expand upon the Lavender Parish Refinery's existing fractionation system to increase the production of Polymer Grade Propylene in order to meet increasing demand. In the past, the refinery has produced a light hydrocarbon stream from the use of a Fluidized Catalytic Cracking Unit (FCC). The FCC unit utilized a specific catalyst combined with high temperatures to break apart large molecules into smaller molecules. This resulted in a production of diesel, gasoline, and Liquefied Petroleum Gas (LPG) streams. Normally, the LPG stream has been processed entirely through an alkylation unit and produced products such as propane and high-octane gasoline. The proposed design would split the LPG mix between the existing alkylation unit and a propylene splitter. The focus of this design was on the propylene splitter. The purpose of this splitter was to intake a propylene/propane mixture and produce Polymer Grade Propylene (99.5% propylene by volume) and Propane HD-5 (95% propane, 5% propylene).

Historically, propylene splitters have been designed in one of two ways. These splitters have either utilized a high-pressure system or a heat pump system. The proposed design utilized a heat pump system to produce the desired product. A heat pump system can operate at lower pressures (close to 180 psia) than high pressure systems (which can operate at 300 psia). The lower required pressure decreases the cost of the column itself. It is possible for the column to operate at lower pressures because the heat pump system utilizes heat integration to condense overhead vapors with bottoms product, and vaporizes bottoms product with compressed overhead vapors. This heat integration also lowers required utility costs by avoiding the use of cooling water to condense overhead vapors and steam to vaporize bottoms product. These costs were analyzed in detail in an economic analysis.

The propylene splitter would intake a known feed stream at 6000 barrels per day, 95^{o}F, and 225 psia. The composition of the feed stream was known and is displayed in the following table.

Table 1: PP mix feed composition

Component	Feed (mol%)
Ethane	0.22
Propylene	90.65
Propane	9.13

There was sufficient existing onsite storage for both the propane and propylene products. These products needed to be stored at 20 psi above there bubble point at 100^{o}F. In order to properly account for any pressure drop through the line to storage, the product streams were designed to be delivered at 50 psi above their bubble point at 100^{o}F. The process was modeled using the Peng-Robinson equation of state. The proposed design was modeled and simulated using Aspen HYSYS.

Tentatively, the project was set to begin construction in 3Q 2018. Startup for the project was set for March 31, 2020. The following capital expenditure profile was utilized when analyzing total installed cost of the process.

Table 2: Capital expenditure profile

Year	2017	2018	2019	2020
% of total installed cost	5	35	45	15

A detailed economic analysis was performed on the recommended process design. This analysis utilized a hurdle rate (internal ROR) or 15%. The project was evaluated on a 10-year project life starting in 2020, the year the project is scheduled to start up. The economic analysis also utilized an effective tax rate of 40%.

The following table displays the known prices of each component utilized and produced in the proposed design. Note that the process intakes a feed of mixed propylene/propane, so the displayed number for “Mixed PP Feed” would be a cost to the company. The values displayed for Propane HD-5 and Polymer Grade Propylene would result in revenue for the company.

Table 3: Feed and product prices

Component	Price (2017), $/barrel
Mixed PP Feed	$46
Propane HD-5	$42
Polymer Grade Propylene	$75

Note that because there was existing storage on site for both Propane HD-5 and Polymer Grade Propylene, there was no cost incurred due to distributing and selling these products. The following table displays the specifications for all available utilities at the Lavender Parish Refinery.

Table 4: Utility specifications

Utility	Specification
Maximum Ambient Temperature	117^{o}F
Minimum Ambient Temperature	-22^{o}F
Process Design Dry Bulb (Cooling Towers)	97^{o}F
Process Design Dry Bulb (Air Coolers)	105^{o}F
Process Design Wet Bulb (Cooling Towers)	76^{o}F
Equipment Minimum Design Temperature	-11^{o}F
Absolute Atmospheric Pressure	14.13 psia
Cooling Water Supply Temperature	87^{o}F
Cooling Water Return Temperature	120^{o}F
Cooling Water Supply Pressure	50 psig
Cooling Water Return Pressure	35 psig
Cooling Water Fouling Resistance	0.002 hr-^{o}F-ft^{2}
Instrument Air Pressure	85 psig
Instrument Air Dew Point	-10^{o}F
Elevation Above Sea Level	992.33 ft

The following table displays the utility costs utilized when costing the proposed design.

Table 5: Utility costs

Utility	Cost
Electricity	$0.07/kilowatt hour
High pressure steam (sat)	$4.75/thousand pounds
Med pressure steam (sat)	$3.30/thousand pounds
Low pressure steam (sat)	$1.60/thousand pounds
Boiler feed water	$0.12/thousand pounds (at 180 psig and 275^{o}F)
Fuel gas	$3.80/million Btu (higher heating value)
Cooling water	$120/annual gpm

Note that because the proposed design utilized a heat pump system, the team was able to bypass using many of the available utilities. This reduced annual operating costs. The proposed design utilized both electricity and cooling water, and both of these costs were included in the economic analysis of the project.

Technical Discussion

Design Philosophy

Column

The column was designed to optimize both outlet purities and DCFROR. Plain carbon steel was the preferred material for the column because it was the cheapest, non-reactive material available. The following column aspects were specified in Aspen HYSYS: column efficiency, tray spacing, tray type, number of stages, column pressures, inlet feed temperature, inlet feed pressure, and inlet feed composition. These characteristics were altered with economic feasibility in mind in order to effectively optimize the process. Note that feed stream characteristics were set values and therefore could not be altered when simulating the process.

Traditionally, a range of tray efficiencies of 70-90% is expected for a propylene splitter using a heat pump system. Using this expected range, a tray efficiency value of 85% was specified when simulating the process.

A tray spacing of 1.677 feet (20 inches) was stipulated for the Aspen HYSYS simulation of the process. A range of 20 to 24 inches was specified in the heuristics in chapter 8 of the Turton textbook (1). 20-inch tray spacing was selected because it lowered the height of the column while still producing the desired outlet purities of both propane and propylene.

Four-pass sieve trays were selected in the column design. Sieve trays are commonly used in the refining industry and are capable of handling large liquid flowrates. A four-pass design was selected in order to effectively handle the large liquid flowrate through the column.

The recommended column design has 140 actual stages (140 four-pass sieve trays). The number of stages was altered on the Aspen HYSYS simulation and product purities were analyzed. The purpose of these adjustments was to balance capital and operating costs while still achieving the desired outlet compositions and flowrates.

The column pressures were specified based off of a recommended pressure drop of 0.1 psi per tray. The pressure at the top of the column was known to be 180 psia, so the pressure at the bottom was calculated by adding (140 trays)*(0.1 psia) to achieve a pressure of 194 psia.

Kettle Reboiler

When designing the heat integration system in the process, it was necessary to choose a type of heat exchanger. It was decided that a reboiler would be utilized in the process to ensure that it would act as an equilibrium stage in the column. A kettle reboiler was selected because it is the most common type of reboiler used in the refining industry (2). The kettle reboiler utilized superheated propylene vapor to vaporize the Propane HD-5 bottoms product, while the bottoms product condensed the superheated propylene vapor. The kettle reboiler has 3 “outlet” streams: a liquid bottoms stream of Propane HD-5, a vaporized propane stream that is fed back to the column, and a condensed propylene stream. It is important to note that the vaporized propane and the liquid Propane HD-5 differed in concentration. This difference in concentration allowed the column to run at equilibrium. Note that high flux tubing was used in the design of the kettle reboilers to reduce the area and cost of the equipment.

Knock-out drum

Before compressing the overhead vapors, it was necessary to design a knock-out drum to ensure that no liquid was run through the compressor. A vertical knock-out drum was selected for this process based off of recommended design for vapor-liquid separation in the oil and gas industry (3). When sizing this compressor, the team considered optimal height to diameter ratios based on recommended heuristics in the Turton textbook (1).

Compressor

A centrifugal compressor was selected for the proposed process. Centrifugal compressors are commonly used for vapor compression in the oil and gas industry (4). Additionally, centrifugal compressors are known to be more reliable and require less maintenance than reciprocating compressors (5).

Compressor driver

An explosion-proof electric driver was selected for the compressor in the design. An electric driver was chosen over a driver that required fuel gas or steam in an effort to lower required utility costs. Electric drivers are capable of powering compressors with ratings up to 20,000 horsepower (1). Due to the high power of this driver (4133 horsepower), safety was a serious concern when designing the compressor. At high power, arcing and explosions become a large concern. The team decided to recommend an explosion-proof driver over a totally enclosed driver to create the safest environment possible. It was determined that the safer design was worth the added cost of the explosion-proof driver.

Cooler

The purpose of the cooler in the proposed design was to further cool the condensed propylene stream leaving the kettle reboiler. The stream needed to be cooled to an appropriate temperature prior to being both delivered to storage and returned to the column as recycle. Cooling water was utilized in a shell-and-tube heat exchanger to achieve the desired result because it was readily available and significantly cheaper than any other available utility (such as a refrigerant or chilled water). The heat exchanger was designed using plain carbon steel because it was the cheapest material available and there was no concern of corrosion. In an effort to reduce costs, high flux tubing was used to design and cost the heat exchanger.

A shell-and-tube heat exchanger was selected for this design. The cooling water was run on the tube side, because it was known that this water contained impurities and that the equipment would needed to be cleaned. The shell-and-tube design was chosen to aid in cleaning and servicing of the equipment. It was also a cheaper option than a plate heat exchanger.

Pump

The purpose of the pump in the proposed design was to increase the pressure of the propane delivery stream to storage. The delivery stream was required to be 240 psia (50 psia above the bubble point of propane at 100^{o}F) (6). The pump increased the pressure of the propane product stream by 46 psia. It was desired that the propane product stream be delivered to storage at a constant flowrate, so a centrifugal pump was selected for the design. The team considered both a centrifugal and a reciprocating pump, but a centrifugal pump was recommended due to its design and cost benefits (7). Using a reciprocating pump of this power would have doubled the cost of the pump, which needed to be spared (bringing the total cost of the pumps from $40,000 to $80,000).

Description of the Process

Overall, the intention of this design was to explore the economic viability of propane and propylene separation using a heat pump cycle to lower operating pressure.

The PP mix was fed at a rate of 6,000 actual barrels/day to column T-100 operating with a top and bottom pressure of 180 and 194 psia, respectively. The vapor product leaving the column was propylene with a purity of 99.6vol%. Any liquid in the overhead product was immediately separated via a knock-out drum and C-100 compressed the purified propylene vapors from 180 psia to 320 psia. The heat of compression resulted in a temperature increase in the process stream from 84^{o}F to 148^{o}F. The high energy propylene stream was passed through the kettle reboilers at the base of the column and cooled down to 122^{o}F.

The purified propylene was cooled further via E-102, an additional heat exchanger, before being split into reflux and product streams. Considerably more Polymer-Grade Propylene was returned to the column as reflux than was sent to product storage (83,000 barrels/day returned as reflux and 5501 barrels/day sent to storage). The propylene designated as product encountered a series of three valves that decreased the pressure of the liquid propylene to 270 psi (50 psia above its bubble point at 100^{o}F). The rest of the liquid propylene was throttled to just above the column top stage pressure before re-entering as reflux. This high-purity propylene reflux increased the internal reflux which aided in separation. Simultaneously, liquid Propane HD-5 exited the bottom of the column. Three kettle reboilers at the bottom of the column utilized heat integration to vaporize bottoms product while condensing superheated overhead vapor. The majority of the bottoms product was vaporized and returned to the column as boilup, and 576 barrels/day of liquid propane exited the reboiler and were sent to storage. The product stream (composed of 95vol% propane and 5vol% propylene) encountered a pump to increase the pressure to 240 psia (50 psia above its bubble point at 100^{o}F). The remaining 88,000 barrels/day of purified propane was returned to the tower just above the reboiler stage as vapor to increase internal reflux.

In whole, this process consumed 6,000 barrels/day of low-purity propylene and returned 5,501 barrels/day of polymer-grade propylene along with 576 barrels/day of Propane HD-5. The Process Flow Diagram (PFD) and stream table summary can be found on the following pages.

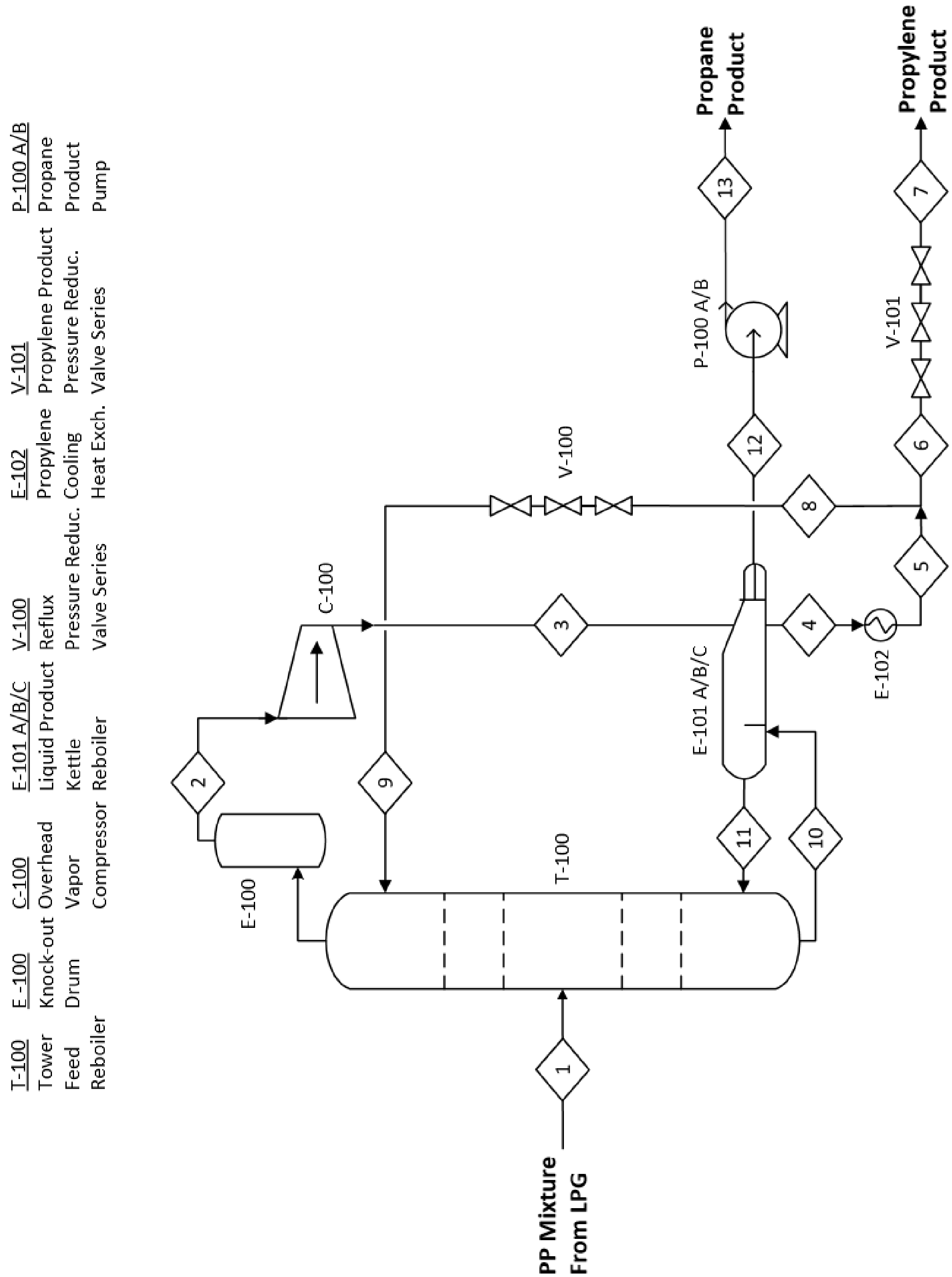


Figure 1: Process flow diagram (PFD)

Table 6: Stream table(1)

Stream Name		Feed	OVHD vapors	Compressed OVHD vapor	Vaporized propane	Cooled Propylene
Stream Number		1	2	3	4	5
Temperature	(°F)	95.0	83.8	147.5	122.2	101.8
Pressure	(psia)	225.0	180.0	319.5	317.5	315.5
Vapor fraction		0.0	1.0	1.0	0.0	0.0
Mass flow rate	(thousand lb/hr)	42.7	66.9	66.9	66.9	66.9
Mole flow rate	(thousand lbmol/hr)	1.0	15.9	15.9	15.9	15.9
Volumetric flow	(thousand barrel/day)	5.6	88.0	88.0	88.0	88.0
Molar enthalpy	(thousand Btu/lbmol)	-2.4	8.1	8.7	3.2	2.5
Component mole frac.						
Ethane		0.0	0.0	0.0	0.0	0.0
Propane		0.1	0.0	0.0	0.0	0.0
Propylene		0.9	1.0	1.0	1.0	1.0
Component Mole Flow	(lbmol/hr)					
Ethane		2.2	26.2	26.3	26.3	26.3
Propane		92.4	43.8	43.8	43.8	43.8
Propylene		917.8	15841.9	15841.9	15841.9	15841.9

Table 7: Stream table(2)

Stream Name		Propylene for product	Propylene to storage	Propylene for reflux	Reflux (after throttle)
Stream Number		6	7	8	9
Temperature	(°F)	101.8	101.5	101.8	83.0
Pressure	(psia)	315.5	270.0	315.5	237.4
Vapor fraction		0.0	0.0	0.0	0.0
Mass flow rate	(thousand lb/hr)	38.8	38.8	630.5	630.5
Mole flow rate	(thousand lbmol/hr)	0.9	0.9	15.0	15.0
Volumetric flow	(thousand barrel/day)	5.1	5.1	82.9	82.9
Molar enthalpy	(thousand Btu/lbmol)	2.5	2.5	2.5	2.5
Component mole frac.					
Ethane		0.0	0.0	0.0	0.0
Propane		0.0	0.0	0.0	0.0
Propylene		1.0	1.0	1.0	1.0
Component Mole Flow	(lbmol/hr)				
Ethane		1.5	1.5	24.7	24.7
Propane		2.5	2.5	41.3	41.3
Propylene		918.8	918.8	14923.1	14923.1

Table 8: Stream table(3)

Stream Name		Bottoms liquid	Boilup	Liquid propane product	Propane to storage
Stream Number		10	11	12	13
Temperature	(°F)	100.8	101.0	101.0	101.8
Pressure	(psia)	194.0	194.0	194.0	240.0
Vapor fraction		0.0	1.0	0.0	0.0
Mass flow rate	(thousand lb/hr)	660.7	656.7	4.0	4.0
Mole flow rate	(thousand lbmol/hr)	15.0	14.9	0.1	0.1
Volumetric flow	(thousand barrel/day)	89.2	88.6	0.5	0.5
Molar enthalpy	(thousand Btu/lbmol)	-47.7	-48.2	-48.2	-48.2
Component mole frac.					
Ethane		0.0	0.0	0.0	0.0
Propane		0.9	0.9	1.0	1.0
Propylene		0.1	0.1	0.1	0.1
Component Mole Flow	(lbmol/hr)				
Ethane		0.0	0.0	0.0	0.0
Propane		14138.7	14053.0	85.7	85.7
Propylene		884.3	879.8	4.5	4.5

Technical Issues and Design Practices

Column

The proposed design of the propylene splitter utilized a heat pump system over a high pressure system. The heat pump utilized a heat integration system that used compressed overhead vapors (superheated propylene vapor) to vaporize the bottoms product (Propane HD-5) in three kettle reboilers. This process also condensed the overhead vapors which would be further cooled before being sent to storage. The use of heat integration reduced operating costs by removing the excessive need for cooling water and any need for steam in the design. The use of a heat pump system was also desirable because it significantly reduced the operating pressure of the column. This lower operating pressure improved the relative volatilities of the components in the mixed PP feed. Additionally, the number of trays within the column could be reduced because the improved relative volatilities allowed for easier separation.

A feed of 90.65 mole% propylene, 9.13 mole% propane, and 0.22 mole% ethane entered the column at 6000 actual barrels per day from an LPG stream. The feed was at a known pressure of 225 psia and a known temperature of 95^{o}F. This feed mixture was purchased at $46/barrel to be converted into polymer-grade propylene and propane HD-5 which would be sold free on board at $75/barrel and $42/barrel respectively.

A bottom operating pressure of 194 psia and a top tray pressure of 180 psia were designated for the column. It should be apparent that the 14 psia pressure drop is the driving force of the column. The reboiler that is attached to this column bears a heat duty for creating this pressure drop. This top pressure was much less than alternative top pressures of 250-300 psia, which benefited the relative volatilities. Overhead vapors were met by a vapor-liquid separator or knock-out drum before being compressed.

Internally, 140 actual four-pass, sieve trays were used, resulting in a 0.1 psi/tray pressure drop from the first equilibrium stage to the reboiler equilibrium stage at the bottom of the tower. Four-pass trays were used over three-, two-, or one-pass trays in order to optimize the liquid handling capacity, which was achieved by reducing both the flow path length as well as the weir crest (6). In addition, sieve or perforated trays are simpler and less expensive than either bubble cap or valve trays, thus sieve trays were used.

Trays were spaced at 1.667 ft in accordance to heuristics outlined in chapter 11 of Turton (1). Using this spacing and the number of trays, a tower height of 243 ft was calculated using the equation below.

$$\text{Column Height (ft)} = (\#\text{ of trays})(\text{tray spacing}) + 7 + (1 + \text{tray spacing})$$

The rationale behind the 7 feet added in the equation was to take into account the space at the bottom of the column below the last stage. Similarly, the last term incorporates the additional space above the first stage at the top of the column.

A column diameter of 13 ft allowed for the appropriate vapor and liquid flow rates to achieve the desired separation. The column diameter determines the tray diameter, which is why tower diameter is designed around the desired surface area for liquid flow, i.e. tray diameter.

The intended material of construction for the column was plain carbon steel, the cheapest option available.

In order to assess actual, rather than theoretical, number of trays, a realistic tray efficiency (ε) of 85% was assumed. The equation below shows how to determine actual from theoretical trays.

$$\text{Number of actual trays} = \frac{\text{Number of theoretical trays}}{\varepsilon}$$

By studying the simulated column component fractions tray-by-tray, it was determined that the optimal feed stage was at stage 73. Recycle streams for the reflux and boil-up were fed at the top and bottom stage respectively. The use of recycles within a process can help increase purity of the end-product downstream, but does decrease the distillate flowrate.

The tops product, purified propylene, was 99.6vol% propylene flowing at 88,000 barrels/day. The entirety of the ethane from the feed was in the vapor product. This high purity propylene was re-fed to the tower as reflux at a flow rate of roughly 83,000 barrels/day at the top stage. The reflux rate was calculated in Aspen HYSYS to be 0.87.

The bottoms product was liquid propane with a purity of 95vol% at 88,500 barrels/day. The liquid propane that left the tower from the bottoms was passed through a reboiler and vaporized before being returned as boilup at a flow rate of 88,000 barrels/day. This resulted in a boilup ratio of 166.

Kettle Reboiler

The main heat exchanger for this system consisted of a combination of three kettle reboilers. The overall heat transfer coefficient for these heat exchangers (U_o) was calculated using the following equation.

$$U_o = \frac{1}{\frac{d_2}{h_i d_1} + R_{fi}\frac{d_2}{d_1} + \frac{d_2 \ln(d_2/d_1)}{2k_w} + R_{fo} + \frac{1}{h_o}}$$

Values of 1.6 inches and 1.7 inches were estimated for the inner and outer diameter of the tubes (8). The values for h_i, h_o, R_{fi}, R_{fo}, and k_w are displayed in the following table.

Table 9: Constant values used in heat transfer calculation

h_i (Btu/ft²*hr*°F)	h_o (Btu/ft²*hr*°F)	R_{fi} (ft²*hr*°F/Btu)	R_{fo} (ft²*hr*°F/Btu)	k_w (Btu/ft*hr*°F)
352.2	616.4	0.001	0.001	29.5

The calculated value of the overall heat transfer coefficient was multiplied by a factor of 2 to account for the use of high flux tubing, resulting in a value of U_o = 307 Btu/ft^2*hr*°F. The heat exchanger was modeled as counter-current, so ΔT_{LMTD} was calculated using the following equation.

$$\Delta T_{LMTD,countercurrent} = \frac{(T_{h,i} - T_{c,o}) - (T_{h,o} - T_{c,i})}{ln\left(\frac{T_{h,i} - T_{c,o}}{T_{h,o} - T_{c,i}}\right)}$$

This value was calculated to be 18.1. The following equation was used to calculate the area of each heat exchanger (8).

$$U_o = qAF\Delta T_{LMTD}$$

A correction factor (F) of 0.9 was selected based off of recommended heuristics (1). Note that the heat duty value (q) given in Aspen HYSYS was 88 million Btu/hr. This heat duty was divided by three to calculate the area for each kettle reboiler. The area of each kettle reboiler was estimated to be 3100 ft^2.

The streams flowing through the reboiler were the superheated propylene stream and the propane stream acting as the boilup. The excess heat from the superheated propylene stream was being used to heat the propane stream. Doing so allowed the propylene stream to condense and the propane stream to vaporize. By using the excess heat generated from compressing the propylene stream to heat the propane stream, the kettle reboilers were able to operate while minimizing the use of excess utilities.

The superheated propylene stream entered the kettle reboilers as a vapor at a pressure of 319.5 psia and a temperature of 147.5°F. The propane stream entered the kettle reboilers as a liquid at a pressure of 194 psia and a temperature of 100.8°F. Exchanging heat between the two streams allowed the propane stream to vaporize and the propylene stream to condense. Exiting the kettle reboilers as a vapor at a pressure of 194 psia and a temperature of 101°F, the vaporized propane is fed back into the column. Exiting the kettle reboilers as a liquid at a pressure of 317.5 psia and a temperature of 122.2°F, the condensed propylene stream was fed to another cooling unit to further lower its temperature. This allowed the propylene to reach appropriate storage conditions.

Knock-out drum

The knock-out drum for this system was designed to prevent the flow of liquid into the compressor. In Aspen HYSYS, the stream entering the compressor was modeled as a completely vapor stream. This was done because if any liquid was allowed to enter the compressor it would damage it.
In the simulation if any liquid was entering the compressor the simulation would not run. However, the simulation is running under ideal conditions with no real atmospheric variations accounted for. In reality there are a number of external issues that could cause a portion of the compressor inlet stream to have partially condensed. Examples of these external circumstances are lower-than-average atmospheric temperatures, or higher-than-normal humidity levels. Both of these issues can draw heat from the pipes causing part of the fluid flowing through them to leave the vapor phase. To prevent this liquid from flowing into the compressor a knock-out drum had to be placed before it to return the liquid to a vapor phase. This knock out drum was sized using heuristics from (1). The remainder of this section will detail the process for sizing the knock-out drum.

The process began by assuming a liquid flow rate for the line entering the knock-out drum. It was assumed that the liquid flow would only account for one percent of the actual flow in the line. Aspen HYSYS listed the actual gas flow of the line as 6851 ACFM. This meant that an actual liquid flow rate of 68.51 ACFM would be used when sizing the drum. The drum was sized as a vertical separator since rule three states that is what is normally used for gas-liquid phase separators (1). According to rule number seven "knock-out drums placed ahead of compressors should hold no less than 10 times the liquid volume passing per minute" (1). Based on this rule the knock-out drum was initially estimated as having a volume of 685.1 cubic feet (19.4 cubic meters) ten times the liquid flow rate that was assumed. Rule four states that the optimum height to diameter ratio is 3 (1). The volume of the knock-out drum was estimated using the equation for the volume of a cylinder.

$$V = \pi * (\frac{D}{2})^2 * h$$

By using the optimal ratio for the height to diameter and the initial estimated volume the equation for the volume of a cylinder can be reduced to a single variable. By inputting the initial volume estimate and substituting 3D for h the needed diameter can be solved for. This resulted in a diameter of 6.62 feet. To allow for the use of standard parts the diameter needs to be rounded up to the nearest 3 inches which causes the diameter to become 6.75 feet. This diameter is technically the diameter of the mist de-entrainer pad; therefore, the diameter of the actual drum still needs to be larger than this. This is accounted for by rounding the diameter of the knock-out drum up another 3 inches to an even 7 feet.

Maintaining the optimal height to diameter ratio of 3 results in a vessel height of 21 feet. The optimal height to diameter ratio was maintained in lieu of the vessels initial volume estimate. Due to this the volume of the knock-out drum has to be recalculated. Plugging a diameter of 7 feet and a height of 21 feet into the equation for the volume of a cylinder results in the vessel having a volume of 808.2 cubic feet (22.9 cubic meters). This volume is greater than the minimum volume required by rule seven of the heuristics of process vessels for drums. This means the dimensions of the knock-out drum meet the necessary specification for keeping liquid from flowing into the compressor (1).

Compressor

The compressor was needed in this design to increase the temperature and pressure of the stream leaving the top of the column. By increasing these two parameters, the stream was able to perform the two functions the system needed it to. First, the stream was able to impart heat energy to other parts of the system. Second, the stream was able to maintain a pressure necessary for the reflux stream to reenter the top of the column.

Before entering the compressor, the stream passed through the knock-out drum to ensure that no liquid entrainment entered the compressor. After passing through the knock-out drum, the stream entered the compressor at 83.8°F and 180 psia. The stream, flowing at a rate of 6581 ACFM, had a mole fraction composition of 0.0016 ethane, 0.0028 propane, and 0.9956 propylene. This was within the desired specifications. The heat duty of the compressor inlet stream was 128,090,000 Btu/hr. The compressor itself was run at 4100 hp with an adiabatic efficiency of 75% and a polytropic efficiency of 76%. This power allowed the compressor outlet stream to leave at 147.5°F and 319.5 psia. This superheated compressed vapor stream was then fed to the cooler.

Compressor Driver

When designing the driver for the compressor in the proposed design, it was necessary to also design a driver. In order to design the driver, the shaft power of the compressor was calculated using the following equation.

$$\text{Shaft power (hp)} = \text{fluid power} + (\text{fluid power})^{0.4}$$

Recall that the required fluid power of the compressor was determined utilizing Aspen HYSYS at a value of 4100 horsepower. With this fluid power, a shaft power of 4133 horsepower was calculated. In order to avoid the costs associated with fuel gas or steam, an electric driver was selected to power the compressor. It was known that electric drivers are designed for services up to 20,000 horsepower (1). An explosion-proof electric motor was selected over a totally enclosed electric motor. At a high shaft power of 4133 horsepower, safety becomes a large concern with this driver. In order to create the safest working environment, the team sacrificed a higher cost to go with what was determined to be the safest design.

Cooler

After the propylene passed through the main heat exchanger, it was further cooled in order to be split into reflux and product streams. To achieve this, a heat exchanger referred to as a cooler was incorporated into the process taking the process stream from 122°F to 102°F using available cooling water. A minimal pressure drop of 2 psi occured across the cooler, changing the pressure from 318 to 316 psia.

This heat exchanger is considerably smaller than the kettle reboilers, at a volume of 3.5ft^3. This was not surprising considering the heat duty of each heat exchanger. The cooler had a duty of 10,000,000 Btu/hr, a small fraction of the heat duty of the main heat exchanger.

The cooler wa exchanging heat between the process stream and cooling water available at the plant. Using heat duty values from the Aspen HYSYS simulation, hand calculations were carried out to determine the required cooling water flow rate. The required cooling flowrate was calculated to be 1900 gallons/minute or 9.9 billion annual gallons/minute. The cooling water that had been heated by the process stream returned at 112.3°F, which was below the maximum cooling water return temperature specified.

In order to size and appropriately cost the heat exchanger, a surface area for conduction was determined from the duty calculated by HYSYS. These calculations yielded an area of 2800 ft^2 and overall heat transfer coefficient of 345 Btu/hr*ft^2*°F. High flux tubes were used in the cooler in an effort to decrease both the area and cost of the heat exchanger.

Pump

Purified propane was to be delivered to a storage tank at 50 psi above the bubble point at 100°F. With the help of the GPSA handbook and the pressure-enthalpy diagrams for propane (found in Appendix III), the bubble point was determined to be 190 psia for propane (6). In order to determine the required pressure, it was assumed that the liquefied propane was pure. The purified Propane HD-5 left the reboiler at 194 psia, so a pump was used in order to increase the pressure to 240 psia. The pump inlet stream entered as liquid propane at 194 psia and 101°F at a flow of 530 barrels/day. The outlet stream was fully liquid at 102°F, so the change in temperature was insignificant.

In determining the type of pump to use, Figure 12-3 GPSA handbook provided insights (6). With a typical operating capacity of 530 barrels/day and a liquid pressure head of 225 ft, determined using Aspen HYSYS, the use of a single-stage, centrifugal pump was preferred.

This relatively small pump is powered by electricity through a motor that was simulated in HYSYS with an assumed 45 percent adiabatic efficiency. Taking this efficiency into account, this pump required a power of 1.34 horsepower. Pumps with this low of power and capacity typically operate at lower efficiencies (1). Similar to other units, the pump will not be encountering any corrosive species under normal operation, so plain carbon steel is used. With such a low rating, due to the small duty required, the cost this pump is very low relative to other units in this process. As a result of the overall low cost of building and operating, this pump is spared.

Capital Cost Estimates

This project was evaluated on a 10-year time frame starting in the year 2020. Note that construction was expected to start in the third quarter of 2018 and the process was set to start up on May 31, 2020. The total capital costs of this project were divided between the years 2017 and 2020 with the following expenditure profile: 5% in 2017, 35% in 2018, 45% in 2019, and 15% in 2020. All dollar values expressed in the economic evaluation of the proposed process are in 2017 dollars.

In order to calculate the capital cost of each piece of equipment, the team began by calculating the "vanilla" purchased cost of the equipment. This cost did not account for factors such as mechanical design pressure, material of construction, or size/power of the equipment. These factors were taken into account to calculate the installed cost of each piece of equipment. These dollar amounts were then escalated to 2017 dollars using CECPI constants. Each piece of necessary equipment was then costed with a contingency rate of 15% and added fees of 3% to allow for any unforeseen costs. These percentages were chosen based off of specified heuristics in the Turton textbook. The following table displays calculated values for the capital cost of each key piece of equipment in the process. These values are in 2017 dollars and include the added cost of contingency and fees (1).

Table 10: Capital costs summary

Equipment	Capital cost
Column	$ 18,300,000.00
Kettle reboiler	$ 1,000,000.00
Knock-out drum	$ 300,000.00
Compressor	$ 2,900,000.00
Electric driver	$ 500,000.00
Cooler	$ 300,000.00
Pump	$ 40,000.00

Column cost calculation

The column was costed as a vertical process vessel using column height, diameter, and mechanical design pressure (allowing for 50 psi above actual operating pressure). The material of construction for this column was plain carbon steel, so it was not necessary to escalate the cost due to the material selected. The cost of the trays in the column are accounted for in the dollar amount displayed in the table above ($18,300,000.00). The trays were costed as 4-pass sieve trays made of plain carbon steel.

Kettle reboiler cost calculation

The kettle reboiler was costed using total heat transfer area, mechanical design pressure, and material of construction (plain carbon steel). Recall that because the total heat transfer area was 9300 ft^2, there are 3 kettle reboilers in this design. The cost displayed in the table above reflects the total cost of the 3 kettle reboilers in the recommended design.

Knock-out drum cost calculation

The knock-out drum was costed as a vertical process vessel using total volume, mechanical design pressure, and material of construction (plain carbon steel). The total capital cost of the knock-out drum was calculated to be $300,000.

Compressor and electric driver cost calculation

The compressor utilized in this design was costed using its fluid power in kilowatts and material of construction (plain carbon steel). Recall that an electric driver was selected for this compressor. The driver was costed using shaft power (fluid power + fluid power$^{0.4}$) in kilowatts.

Cooler cost calculation

The cooler was costed using total heat transfer area, mechanical design pressure, and material of construction. Note that this cooler is considerably smaller than the kettle reboiler, and is therefore considerably cheaper.

Pump cost calculation

The pump was costed using shaft power (in kilowatts), mechanical design pressure, and material of construction (plain carbon steel). Due to the relatively low capital cost associated with this pump ($20,000), the pump was spared. The cost in the table above accounts for both the pump in use and the spared pump.

Re*venue and Operating Expense Estimates*

Revenue Estimates

The annual revenue for this project was calculated using the daily flowrate of Propane HD-5 and Polymer Grade Propylene. The daily flowrate of Propane HD-5 was approximately 576 barrels/day. This value was converted into an annual flowrate of 210,240 barrels/year. Using the known sales value of $42/barrel for Propane HD-5, the team calculated an annual revenue of $8,800,000. The same process was used to calculate the annual revenue from Polymer Grade Propylene at a known sales value of $75/barrel. The team observed a daily flowrate of approximately 5501 barrels/day of Polymer Grade Propylene, resulting in an annual flowrate of 2,007,865 barrels/year and an annual revenue of $150,600,000. The total calculated revenue was $159,400,000. Note that this is total calculated revenue, not actual revenue. Taking into account a service factor of 95%, the actual estimated revenue of the proposed process is $151,430,000.

Operating Expense Estimates

Operating expenses in this process design included utility costs (such as electricity and cooling water) as well as local taxes and insurance.

There was a utility cost associated with the required electric power to operate the compressor in this design. A shaft power of 4133 horsepower was used to cost this utility. This was converted into an annual electricity requirement, then costed with a known electricity cost of $0.07/kilowatt hour. This resulted in an annual cost of $1,900,000 to operate the compressor. The same process was utilized to cost the pump in the process. This pump required a shaft power of 1.34 horsepower, resulting in an annual cost of electricity of $1,000 to operate the pump. Cooling water was required to further cool the propylene stream leaving the kettle reboiler. This cooling water was costed by utilizing the calculated annual volumetric flowrate along with the known cost of $120/annual gpm. This calculation resulted in an annual cooling water cost of $176,000.

In order to estimate the annual cost of local taxes and insurance, the team utilized a relationship of 3.2% of the total capital costs of the proposed design. This relationship was chosen from a recommendation in Chapter 8 of the Turton textbook. It was determined through further analysis that this relationship predicted a reasonable dollar amount for local taxes and insurance of $740,000 annually (1). The predicted operating expenses of the proposed design are summarized in the following table.

Table 11: Operating costs summary

Operating expense	Cost
Electricity	$ 2,000,000.00
Cooling water	$ 200,000.00
Local taxes and insurance	$ 740,000.00
Total	$ 2,940,000.00

It was determined that it was not necessary to account for waste treatment costs in the economic analysis of this project. It was assumed that because cooling water was available in the facility, that there was already a waste treatment system in place and it would therefore not be of any additional cost. Operating labor and direct supervisory/clerical labor were also not included in this analysis. It was assumed that this project would not require any additional personnel at the refinery. Maintenance and repairs were accounted for in capital cost calculations with an added contingency rate of 15%. This percentage was chosen to ensure the economic cost estimate would accurately account for any unforeseen incurred charges. Note that there was not a need to account for laboratory charges, patents and royalties, or research and development costs in this design process, because these charges were not relevant in the addition of a propylene splitter to an existing refinery. Additionally, that there is sufficient storage on-site for the products produced in this process, so there was no need to estimate the cost of selling and distribution.

DCFROR Analysis

A Discounted Flow Rate of Return (DCFROR) analysis was utilized to determine the economic feasibility of the proposed process design. In order to calculate the DCFROR, a cash flow table was utilized. The detailed cash flow table can be found in Appendix I. This cash flow table took into account total sales revenue, raw material costs, and all operating costs highlighted in the previous section. Note that all costs and revenues in the year 2020 were multiplied by 0.75, to account for the expected startup date of March 31. A tax rate of 40% was taken into account, and the total capital costs were depreciated using a 10-year MACRS rate (9). Note that any undepreciated capital costs in 2029 were written off. In order to take into account working capital, a value of 8% of total capital costs was used annually starting in 2020 (the startup year). A working capital value of 15% of total capital costs was found in chapter 9 of the Turton textbook (1). This value represented working capital to start up a plant, so it was concluded that only half this value was necessary for the addition of a process to an existing refinery.

The cash flow analysis resulted in a DCFROR of 85%, showing that the proposed process is very economically attractive. It was requested that the team analyze this project with a hurdle rate of 15%. Using this hurdle rate, the project has a Net Present Value of $80,400,000.

The break-even selling cost of this project was calculated to be $28,500,000. This is the highest cost the project can incur while remaining profitable. This cost was calculated by subtracting the Net Present Value (NPV) at i* (15%) from the sum of the present worth revenue.

Please see Appendix I for the detailed cash flow chart.

Sensitivity Analysis

A sensitivity analysis was performed to analyze the economic feasibility of the proposed design. In order to perform this analysis, values for higher/lower sales revenue, capital costs, raw material costs, and utilities costs were analyzed. The results of this analysis are displayed in the charts below.

Figure 2: DCFROR tornado chart (sensitivity analysis)

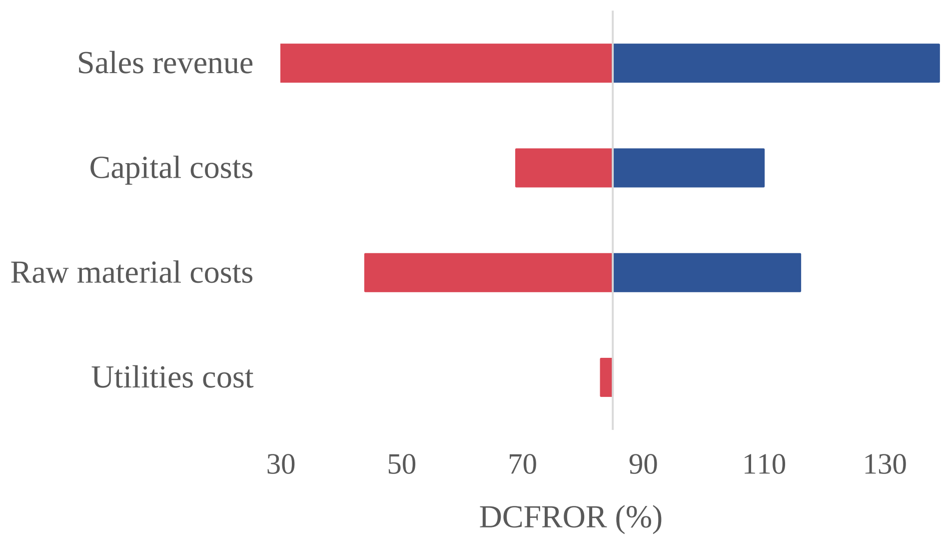
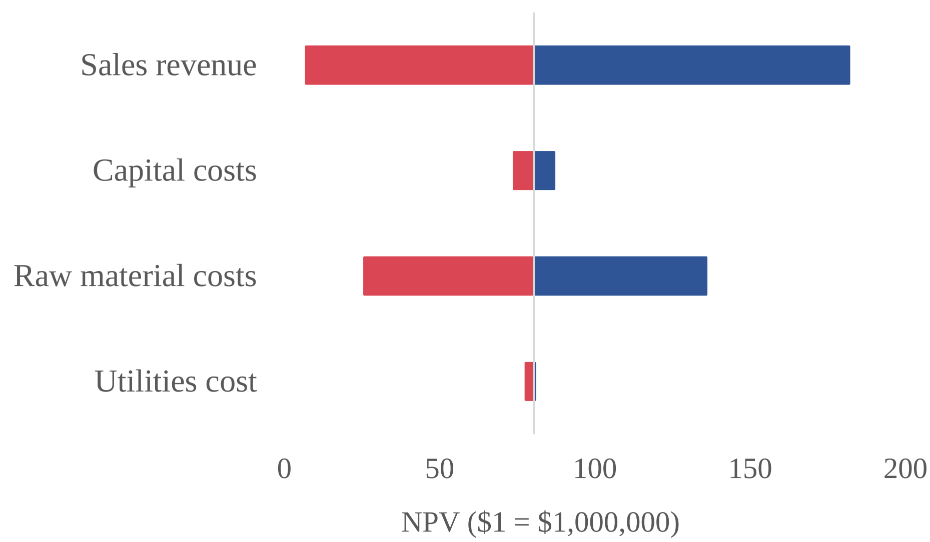


Figure 3: NPV tornado chart (sensitivity analysis)



Sales Revenue Analysis

(FIGUREM1) shows that the sales revenue has a significant impact on the DCFROR. In order to perform this analysis, a literature search was conducted to find historical high and low selling values of both Polymer Grade Propylene and Propane HD-5 (10,11). For Polymer Grade Propylene, a low value of $54/barrel was used and a high value of $88/barrel was used. For Propane HD-5, a low value of $36/barrel was used and a high value of $155/barrel was used.

Recall that the current known selling price of these products is $75/barrel of Polymer Grade Propylene and $42/barrel of Propane HD-5. At the lowest selling value of both Propane HD-5 and Polymer Grade Propylene, the proposed process maintained a DCFROR of 26%, which was still well above the chosen hurdle rate of 15%. This low range produced an NPV of $7,000,000. At the highest selling value of both products, the DCFROR was raised to 139%, and NPV was raised to $182,700,000. Both of the product selling values were varied significantly in this analysis, but the project remained very attractive from an economic standpoint.

Capital Costs Analysis

The total capital costs of the proposed process were analyzed in order to determine the projects economic feasibility at higher and lower capital costs. In order to alter these values, CECPI constants were used to determine a feasible percent change. The CECPI value in 2001 was 397 and in 2017 was 566.6. The team analyzed the percent difference between these values, which was calculated to be approximately 30%. It was concluded that ±30% would be a reasonable variance when analyzing these costs. At -30% of actual capital costs, the DCFROR was raised to 110% and the NPV was raised to $87,000,000. At +30% of actual capital costs, the DCFROR was lowered to 69% and NPV was lowered to $73,900,000. Both of the DCFROR values were well above the hurdle rate of 15% and the project maintained a positive and high Net Present Value.

Raw Material Costs Analysis

In order to analyze the raw material costs of this project, the team used ±30% of the specified raw material cost. This percentage was chosen based on the rationale described in the capital costs analysis above. Recall that the annual cost of raw materials was calculated using a specified value of $46/barrel at 6000 actual barrels per day. This analysis showed a significant impact on the DCFROR. At -30% of actual raw material costs, the DCFROR was raised to 116% and NPV was raised to $136,000,000. At +30% of actual raw material costs, the DCFROR was lowered to 44% and the NPV was lowered to $24,800,000. Note that the NPV remained positive in both scenarios and the DCFROR remained above the hurdle rate of 15%.

Utilities Cost Analysis

Figures 2 and 3 show that changing the cost of utilities (electricity and cooling water) did not significantly alter the DCFROR of this project. Through a literature search, historical values of electricity cost and cooling water costs were found and utilized in this sensitivity analysis (12,13). Electricity was costed at a low value of $.068/kilowatt hour and a high value of $0.125/kilowatt hour, and cooling water was costed at a low value of $75/annual gpm and a high value of $134/annual gpm. The low range of these utility costs did not affect the DCFROR by a percentage, meaning that it remained at 85%. The NPV in this scenario raised slightly to $80,800,000. The high range of utility costs lowered the calculated DCFROR by 2% (to 83%) and lowered the NPV to $77,700,000.

Safety

Employers must ensure that employees are working in a safe, healthy environment. This can be accomplished by following the rules and regulations set forth by the Occupational Health and Safety Administration (14). Training and education are an essential part of making this a successful process.

Employers are required to provide a safe environment, but employees are expected to positively contribute to workplace safety. A major way this is done is by not taking any egregious risks at work due to ignorance or incompetence. Workers need to be attentive to their surroundings so as to be able to notice any safety issues that may arise. In order to properly respond to these issues employees need to be able to satisfactorily execute the safety training they had to undergo to work in this environment. The ability to quickly locate necessary equipment such as fire extinguishers or emergency shut off valves, promptly execute evacuation plans, or carry out other contingencies is essential to the continuation of healthy workplace conditions. Dangers specific to the heat pump based propylene splitter system and the chemicals ethane, propane, and propylene used in it will be laid forth in this section of the report. More information regarding these chemicals can be found on the material safety data sheets supplied by the company.

Two safety aspect that need to be addressed about the system are directly related to the environment that the process is contained in. The first aspect is that it is crucial that a closed system is maintained for the process. Any leaks or spills of the chemicals used will effect production and endanger workers. Through a large portion of this process the chemicals in it are in a liquid form. Exposure to these chemicals in their liquid form can result in frostbite. This is caused by them they quickly sucking heat from their surrounds due to their rapid evaporation (15,16,17). The second aspect is the need to ensure the area in which the process is operating is properly ventilated. In the chance that a loss of containment occurs or internal repairs are needed the use of proper ventilation can prevent suffocation. All three of the components in this process can easily displace the air in an area and lead to oxygen deprivation.

While running this process, care needs to be taken to monitor the pressure and temperature levels throughout the system. The pressures are lower due to running the system with a heat pump; however, the system is still operating well above atmospheric conditions. Pressure systems and relief valves need to be monitored to ensure the process stays within its bounds, and that the system does not rupture.

This portion will provide specific information regarding the handling and interaction with the compounds used during the propylene splitter process laid out in this report. The three chemicals present during the separating process are ethane, propane, and propylene. All three of the chemicals are identified by the National Institute for Occupational Safety and Health as extremely flammable and highly explosive (15,16,17). Due to this classification enormous caution must be taken to prevent them from being exposed to any sort of ignition source. One such source would be the buildup of electrostatic charges in the system which can be prevented through various methods. One method is by regularly cleaning the system to prevent fouling in it from generating electrostatic charges. Another method is by properly grounding the system so that any errant charges can be safely dispersed. Should the system catch fire, operating crews must act prudently, and quickly shut off any feeds to the portion in flames. If possible the fire should be allowed to burn itself out. However, if the fire poses an immediate danger to the surrounding areas then actions need to be taken to combat the fire from a sheltered position. In order to combat the fire, the vessels containing the chemicals may be cooled with a water spray, but due to the presence of propylene in the mixture the actuals flames must be extinguished by means of a powder or carbon dioxide.

Environmental

The cooling water system needs to be properly cared for. The system needs to be regularly cleaned to prevent corrosion in the pipes and to prevent fouling from generating electrostatic charges. Chemicals can be added to the water to delay both of these effects, but they also require the water to undergo specific disposal processes. This is to prevent any negative effects on the environment, and to ensure that the company continues to meet Environmental Protection Agency (EPA) regulations. The system also needs to be regularly tested for microbial activity to prevent the growth of harmful bacteria in the water. Cooling water can be a breeding ground for dangerous bacteria like Legionella which causes Legionnaires' disease (18). Maintaining a disease free environment is an essential part of running the plant.

The specifications which the plants cooling water intake structures were designed for are regulated by the EPA according to the Clean Water Act. Screens had to be placed at the inlets to prevent large numbers of fish and other wildlife from being dragged into the cooling water system. The suction produced from these inlets though can still trap wildlife against the screens potentially injuring or killing them. Precautions such as frequent inspections need to be taken to limit this as much as possible. That way the plant can limit the impact its cooling water inlets have on the surrounding ecosystem. The impact the plants cooling water outlets have on the environment also needs to be monitored. The chemicals alongside the heat that gets added to the cooling water as it travels through the facility can have adverse effects on the environment. These factors can place undue stress on the area around the plant, and have other father reaching consequences (19). Due to the propylene splitter being added onto an existing facility cooling water and the equipment necessary for its disposal are already present on site. The existing equipment when also used to run cooling water for the propylene splitter still meets EPA regulations; therefore no major changes will need to be addressed when obtaining new or revised permits through the National Pollutant Discharge Elimination System (20).

Conclusions

The team was able to effectively model the propylene splitter using a column made of carbon steel with the following specifications.

Table 12: Column design specifications

Diameter (ft)	Tray spacing (ft)	Column height (ft)	Tray type	# of trays	Volume(ft^3)
13	1.667	143	4-pass sieve tray	140	32,000

Recall that a tray efficiency of 85% was assumed in the design of this column. This efficiency was chosen from a given range of 70-90% expected efficiency with a propylene splitter using a heat pump system. An electric explosion-proof driver was selected to run the compressor in the design. This compressor driver had a shaft power of 4133 horsepower, inquiring an annual cost of electricity of $1,900,000. The process was designed using 3 kettle reboilers, each with an area of 850 ft^2. The total cost of the kettle reboilers was $1,000,000. There was no operating cost associated with the kettle reboilers due to the utilization of heat integration. The use of steam/cooling water in these reboilers was avoided because the superheated propylene vapor was used to vaporize the bottoms product, while the bottoms product condensed the superheated propylene. A cooler was used to cool the condensed propylene to an appropriate temperature to be delivered to storage and recycled as reflux to the column. This cooler was modeled as a relatively small shell-and-tube heat exchanger, resulting in a total capital cost of $300,000 and an annual operating cost of $200,000 to pay for the required cooling water. The proposed design included a pump to raise the pressure of the propane delivery stream to 50 psi above the bubble point at 100^oF (240 psia). Accounting for an efficiency of 45%, this pump had a shaft power of 1.34 horsepower. This pump was spared, resulting in a total cost of $40,000. There was also an annual operating cost of $1,000.

Based on the models developed for this project it was determined that it is a technically feasible process. Based off of the economic analysis conducted, it was determined that the project is economically attractive based on the DCFROR of 85% and NPV value of $80,400,000.

Recommendations

After both technical and economic analyses it was determined that this project should be pursued.

During the process of compiling this report, further improvements to the design were realized but not implemented. First, it would have been logical to throttle the purified propylene stream prior to splitting it to reduce the pressure. This would have reduced the number of valves which would reduce total capital costs. Second, it would be logical to consider tray efficiencies of lower than 85% (this is the efficiency that was utilized in the proposed design). Tray efficiencies of as low as 70% could be analyzed in order to propose a more conservative column design. When implementing the proposed design, it is also recommended to analyze the potential need of extra pumps. This could become necessary when pumping the recycled reflux to the top stage of the column.

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Appendix I – Cash Flow Chart (2017-2029)

Please see next page(s) for Cash Flow Chart.

END OF YEAR	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029
Propane production	0	0	0	210240	210240	210240	210240	210240	210240	210240	210240	210240	210240
x Sales price, $/"unit"	42	42	42	42	42	42	42	42	42	42	42	42	42
Propylene production	0	0	0	2007865	2007865	2007865	2007865	2007865	2007865	2007865	2007865	2007865	2007865
x Sales price, $/"unit"	72	72	72	75	75	75	75	75	75	75	75	75	75
Net Revenue	**$ -**	**$ -**	**$ -**	**$ 113,572,500**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**	**$ 151,430,000**
- Raw material costs	0	0	0	(71,800,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)	(95,700,000)
- Other op Costs	0	0	0	(1,700,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)	(2,200,000)
- Depreciation	0	0	0	0	(2,330,000)	(4,194,000)	(3,355,200)	(2,684,160)	(2,148,260)	(1,717,210)	(1,526,150)	(1,526,150)	(1,528,480)
- Manufacturing costs	0	0	0	(555,000)	(740,000)	(740,000)	(740,000)	(740,000)	(740,000)	(740,000)	(740,000)	(740,000)	(740,000)
- Loss forward													
- Writeoff													(764240)
Taxable Income	**0**	**0**	**0**	**39517500**	**50460000**	**48596000**	**49434800**	**50105840**	**50641740**	**51072790**	**51263850**	**51263850**	**50497280**
- Tax @ 40%	0	0	0	(15807000)	(20184000)	(19438400)	(19773920)	(20042336)	(20256696)	(20429116)	(20505540)	(20505540)	(20198912)
Net Income	**0**	**0**	**0**	**23710500**	**30276000**	**29157600**	**29660880**	**30063504**	**30385044**	**30643674**	**30758310**	**30758310**	**30298368**
+ Depreciation	0	0	0	0	2330000	4194000	3355200	2684160	2148260	1717210	1526150	1526150	1528480
+ Loss forward	0	0	0	0	0	0	0	0	0	0	0	0	0
+ Writeoff	0	0	0	0	0	0	0	0	0	0	0	0	764240
- Working capital	0	0	0	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)	(1864000)
- Fixed capital	(1,165,000)	(8155000)	(10485000)	(3495000)									
Cash Flow	**(1165000)**	**(8155000)**	**(10485000)**	**$ 18,351,500**	**$ 30,742,000**	**$ 31,487,600**	**$ 31,152,080**	**$ 30,883,664**	**$ 30,669,304**	**$ 30,496,884**	**$ 30,420,460**	**$ 30,420,460**	**$ 30,727,088**
Discount factor (P/Fi*, n)	1.00	0.87	0.76	0.66	0.57	0.50	0.43	0.38	0.33	0.28	0.25	0.21	0.19
Discounted Cash Flow	**(1165000)**	**(7091304)**	**(7928166)**	**$ 12,066,409**	**$ 17,576,838**	**$ 15,654,902**	**$ 13,467,903**	**$ 11,610,313**	**$ 10,025,849**	**$ 8,669,117**	**$ 7,519,472**	**$ 6,538,671**	**$ 5,743,112**

Appendix II

Please see next page(s) for detailed costing sheets.

Column costing sheet

Volume (m^3)	913.33
D (m)	3.96
Height (m)	74.07

20 in spacing, 140 trays,
1 ft above 7 ft below

For tower	K1	3.4974	Cp^0	$ 584,300	ID # Table A.3	18
	K2	0.4485	CECPI (old)	397	P (barg)	15.8
	K3	0.1074	CECPI (new)	566.6	Fp, vessel	6.79
	Table A.1 - values for vertical Process Vessel		Cp^0 (escalated)	$ 833,915	B1	2.25
					B2	1.82
					FM	1
					FBM	14.61
					CBM	$ 12,186,410

Note that this CBM is the Installed cost of the tower (NOT including trays)

Tower area (m^2)	12.33
K1	2.9949
K2	0.4465
K3	0.3961
Cp^0	8984.83
Cp^0 (escalated)	12823.19
# of trays	140
fQ	1
ID # Table A.6	61
FBM	1.83
CBM, trays	$ 3,285,300

Installed cost of tower and trays (column installed cost)	CBM, tower+trays	$ 15,471,710
	Contingency (15%)	$ 2,320,756
	Fees (3%)	$ 464,151
Total module cost (installed with contingency and fees) of column	CTM, tower+trays	**$ 18,256,618**

Kettle reboiler(s) costing sheet

Capital Cost Calculation					
		(Found in fig 7.4)			
Finding Purchased Cost (C_p^0)	Heat Transfer Area (m^2)	a	b	n	C_p^0
	431.9	29000	400	0.9	$ 123,168.64
Finding bare module cost (C_{BM})	B_1	B_2	F_p	F_M	C_{BM} (2001)
	1.63	1.66	1.104	1	$ 426,582
	(found in Table A.4)^^^				
Pressure factor calculation	C_1	C_2	C_3	P (barg)	
	0.03881	-0.11272	0.08183	26	
	(found in table A.2)^^^				
Calculating actual cost	$C_{TM(2001)}$(15% cont 3% fees)	CECPI 2001	CECPI 2017	C_{TM} (2017)	C_{TM} (2017) (2 HEX's)
	$ 503,367	539.2	566.6	$ 528,946	**$ 1,057,893**

Cooler costing sheet (capital cost)

Capital Cost Calculation					
		(Found in fig 7.4)			
Finding Purchased Cost (C_{p^0})	Heat Transfer Area (m^2)	Purchased cost/unit area	Purchased cost (C_{p^0})		
	280.1	190	$ 53,217		
Finding bare module cost (C_{BM})	B_1	B_2	F_p	F_M	C_{BM} (2001)
	1.63	1.66	1.096	1	$ 183,546
	(found in Table A.4)^^^				
Pressure factor calculation	C_1	C_2	C_3	P	
	0.03881	-0.11272	0.08183	24.3	
	(found in table A.2)^^^				
Calculating actual cost	$C_{TM(2001)}$(15% cont 3% fees)	CECPI 2001	CECPI 2017	C_{TM} (2017)	
	$ 216,585	397	566.6	**$ 309,111**	

Cooler costing sheet (operating cost)

Cooling water cost calculation	Duty (kJ/h)	T(hot, in)degF	T(hot, out)degF	T(cold, in)degF	T(cold, out)degF
	10,840,000.00	323.24	311.94	303.75	317.75
	Cc (Btu/lbm*degF)	Duty (btu/hr)	mass flow (lbm/h)	mass flow (gpm)	annual gpm
	0.997	10,200,000.00	730,763.72	1461.53	768,178,822.18
	Annual cost of CW				
	$ 175,383				

Compressor/compressor driver costing sheet (capital cost)

Compressor cost calculation									
Capital Cost Calculation									
Finding purchase cost C_{p}^{0}	K_1	K_2	K_3	A (fluid power in KW)	C_{p}^{0}				
	2.2897	1.3604	-0.1027	3057	$ 607,532				
Finding bare module cost (C_{BM})	F_{BM}	C_{BM}	C_{TM}	CECPI 2001	CECPI 2017	C_{TM} (2017)			
	2.78	$ 1,688,941	$ 1,992,951	397	566.6	**$ 2,844,347**			
	Fig A.19								
Costing compresser driver (steam)	K_1	K_2	K_3	A (shaft power in KW)	C_{p}^{0}	F_{BM}	C_{BM}	C_{TM}	C_{TM}(2017)
	2.4604	1.4191	-0.1798	3081.78	$ 167,062	1.5	$ 250,594	$ 295,701	**$ 422,026**

Compressor/compressor driver costing sheet (operating cost)

Operating Cost Calculation			
Cost of electricity	$/ KW hour	Shaft Power in KW	Annual Electricity cost
	0.07	3081.78	**$ 1,889,748**

Pump costing sheet (capital cost)

Pump cost analysis

Capital cost calculations

	(Values for centrifugal pump)							
Finding purchase cost (C_p^0)	K_1	K_2	K_3	A (Shaft Power, hp)	A (Shaft Power, KW)	C_p^0		
	3.3892	0.0536	0.1538	1.34	1.00	$ 2,450.19		
Finding bare & total module cost (C_{BM} and C_{TM})	B_1	B_2	F_M	F_P	C_1	C_2	C_3	P
	1.89	1.35	1.55	1.20	-0.3935	0.3957	-0.00226	15.8
	C_{BM}	C_{TM}	CECPI 2001	CECPI 2017	C_{TM} (2017)	Spare pump	Total $C_{TM(2017)}$	
	$ 10,760	$ 12,697	397	566.6	**$ 18,121**	$ 18,121	**$ 36,243**	

Pump costing sheet (operating cost)

Operating cost calculations			
Cost of electricity	$/ KW hour	Shaft Power in KW	Annual Electricity cost
	0.07	1.00	**$ 613.20**

Knock-out drum costing sheet

Knock out drum cost analysis								
Capital cost calculations								
	(Values for centrifugal pump)							
Finding purchase cost (C_p^0)	K_1	K_2	K_3	D (m)	Height (m)	A (volume m^3)	Cp^0	
	3.4974	0.4485	0.1074	2.13	6.40	22.9	$ 20,215	
Finding bare & total module cost (C_{BM} and C_{TM})	P (barg)	P+1	F_p	B_1	B_2	F_{BM}	C_{BM}	C_{TM}
	14.9	15.90	3.48	2.25	1.82	8.6	$ 181,796	$ 214,519
								CTM(2017)
								$ 306,162

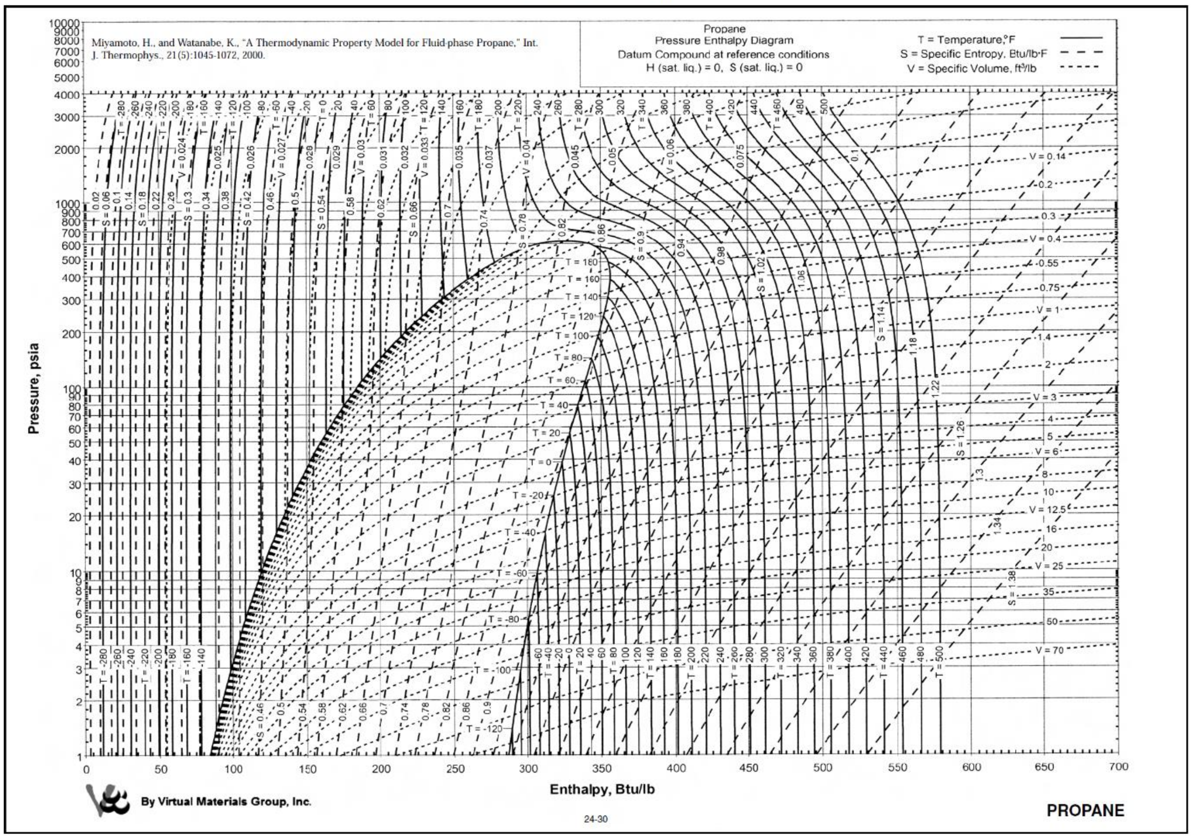
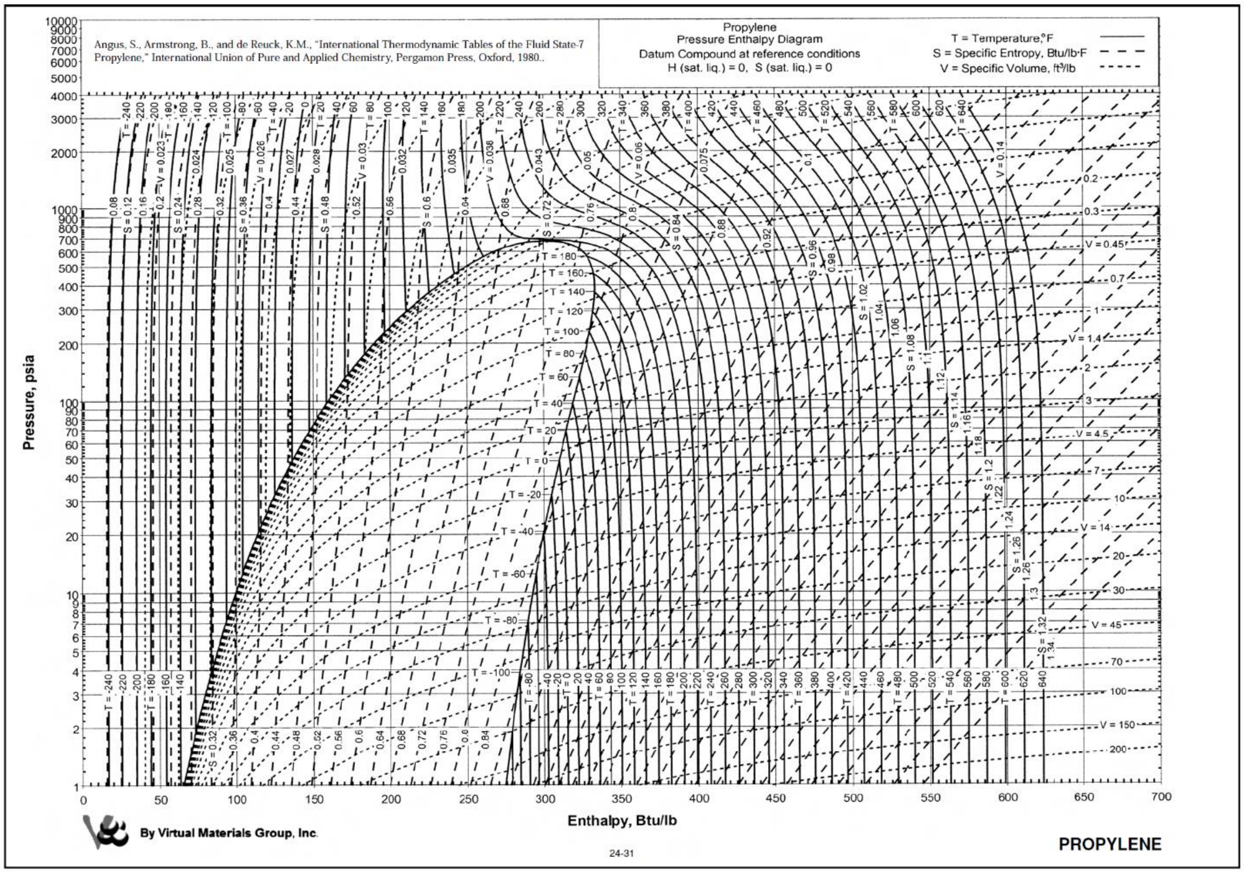

Miyamoto, H., and Watanabe, K., "A Thermodynamic Property Model for Fluid-phase Propane," Int. J. Thermophys., 21(5):1045-1072, 2000.
Propane
Pressure Enthalpy Diagram
Datum Compound at reference conditions
H (sat. liq.) = 0, S (sat. liq.) = 0
T = Temperature, °F
S = Specific Entropy, Btu/lb·F
V = Specific Volume, ft³/lb
Pressure, psia
Enthalpy, Btu/lb
By Virtual Materials Group, Inc.
24-30
PROPANE

FIG. 24-28 Propylene P-H Diagram



Appendix IV

Please see next page(s) for Aspen HYSYS workbook.