Cooper Union for the Advancement of Science and Art

Che 161.2: Process Evaluation and Design II

Ethylene Production Plant Design

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Executive Summary

An ethylene production plant was designed to meet a product specification of 700 metric tons per day. To do so, 140,010 lb/hr of 100% butane is fed to the plant, and 100% of ethane is recycled at a rate of 8,174 lb/hr. This plant process includes: a furnace (to crack the hydrocarbon feed), a quench tower (to cool the exiting stream), a four stage compressor system (to increase the pressure of the cracked gases), a refrigeration section (to cool the stream to cryogenic temperatures), and a five tower fractionation system (to separate the remaining products). The resulting ethylene, propylene, gasoline, and high pressure stream products are subsequently sold.

An economic analysis was performed on the ethylene plant designed, where it is expected that the plant will profit approximately 160 million over a 10 year operation period, with a return on investment of 16%. It was assumed for the economic analysis that the plant will run 8,400 hours a year (0.96 plant operating factor). This promising return on investment makes it sensible to perform more rigorous economic studies, and to potentially go forward in the plant construction. The capital investment for the plant construction was $248,000,000. The utilities to run the compressors and pumps, including low pressure steam, cooling water, and electricity, cost approximately $16,000,000 per year. To reduce unnecessary utility costs, certain process streams were used for heating and cooling. Since it is assumed that wage costs are $4.5 million per year, the total annual operating cost is $20,500,000 per year.

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Introduction

An ethylene production plant was designed for its wide industrial relevance and its great economic potential. A more detailed exploration of the plant process and the motivation for ethylene production are discussed below.

1.1 Motivation for Ethylene Production

Ethylene is a valuable product of this process with a $160 billion per year market alone [1]. Ethylene is produced more than any other organic compound due to its versatile chemical industrical use. Industrial uses of ethylene include polymerization, oxidation, halogenation, alkylation, hydration, oligomerization, and hydroformylation [2]. Specifically, it is used in the production of polyvinyl chloride, and other important plastics [3]. Therefore, in this study, an ethylene production plant was designed for its economic potential. The plant process to produce ethylene is discussed below.

1.2 Plant Process Summary

There are five main sections of the ethylene production plant which must be designed: a furnace, a quench tower, a compressor system, a demethanizer/refrigeration section, and a set of fractionation columns (Figure 1.1).

First, a 100 % n-butane feed enters the furnace, where it is cracked to produce various hydrocarbons, including ethane and ethylene. To be economically efficient, the excess heat is recovered in the convection section, where the diluent steam is superheated (for other processes witin the plant), and the feed is preheated (prior to cracking). Next, the hydrocarbon product is cooled with cooling water in a quench tower to prevent coking and unwanted polymerization reactions. The resulting cooled stream is fed into compressors, where the pressure is increased for the following demethanizer section. In the demethanizer section, the stream is cooled even more, in order to approach cryogenic temperatures by a refrigeration cycle. Finally, using a series of fractionation towers, the product stream is purified to obtain a 99.95 mol% ethylene stream, a propylene stream, and other valuable product streams (i.e. gasoline).

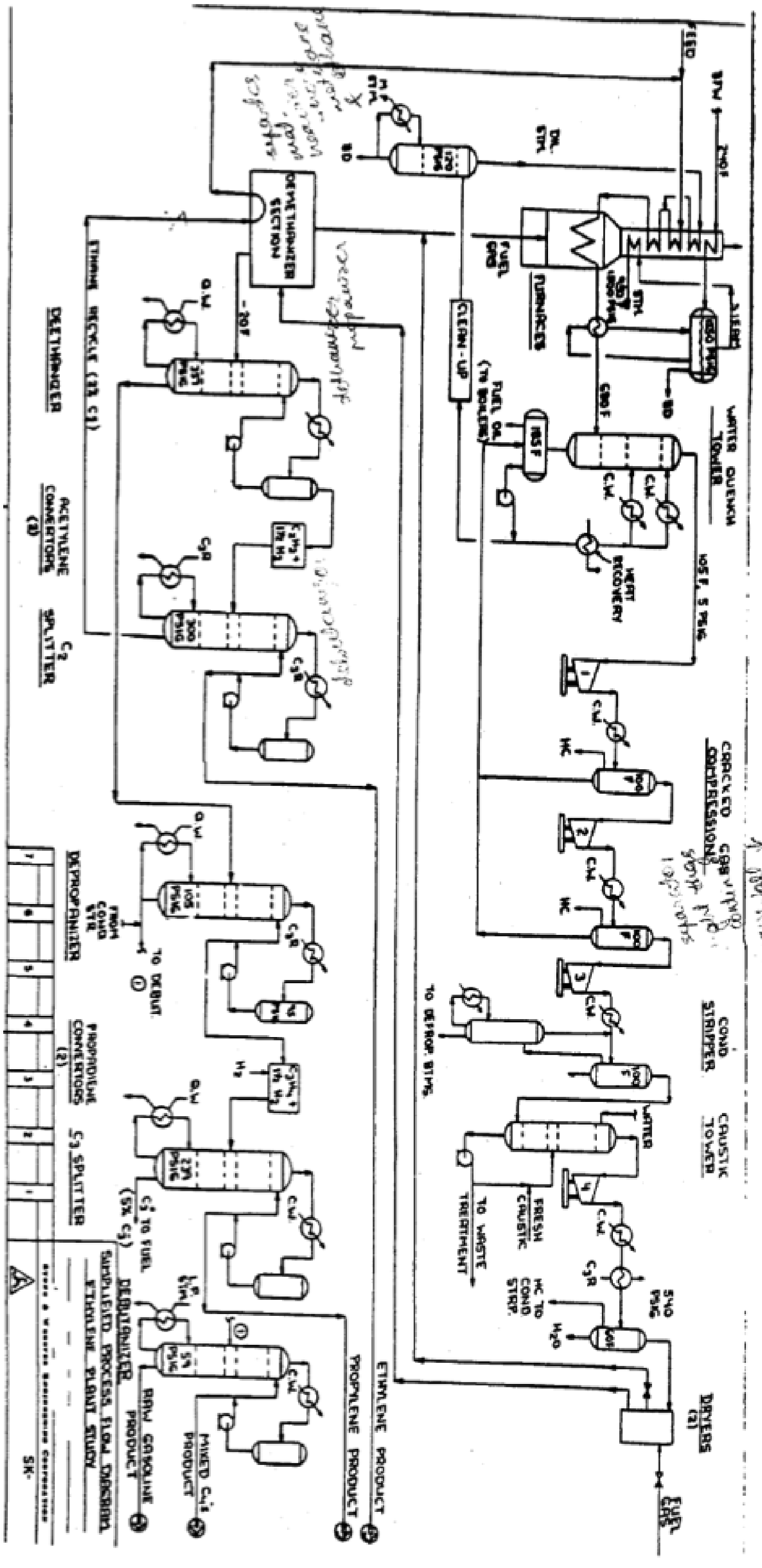

FIGURE 1.1: There are five main sections of the ethylene production plant which must be designed: a furnace, a quench tower, a compressor system, a demethanizer/refrigeration section, and a set of fractionation columns.

Mass Balance

To meet the product specification of 700 metric tons of ethylene per day with 100% ethane recycle, a mass balance was performed on the ethylene plant. As a result, the rate of ethylene production, feed requirements, and ethane recycle were found, as shown below.

2.1 Rate of Ethylene Production

First, the rate of ethylene production was calculated, assuming 99.95% purity of ethylene in the product stream, and assuming 99% recovery of ethylene from the furnace:

$$\text{Plant Capacity} = 700\frac{\text{ton}}{\text{day}} \times 2,204.6\frac{\text{lb}}{\text{ton}} \times \frac{1}{24}\frac{\text{day}}{\text{hours}} = 64,950 \text{ lb/hr} \tag{2.1}$$

Therefore, the plant capacity is to produce 64,950 lb/hr.

2.2 Feed Requirement

By calculating the rate of ethylene production, it was possible to calculate the feed requirement into the furnace. Within the furnace, it was necessary to account for the hydrogenation of acetylene, and the 100% recycle of ethane. The hydrogenation of acetylene is described with the following reaction:

$$C_2H_2 + \frac{3}{2}H_2 \rightarrow \frac{1}{2}C_2H_4 + \frac{1}{2}C_2H_6 \tag{2.2}$$

In terms of mass, the reaction equation becomes:

$$26 \text{ lb } C_2H_2 + 3 \text{ lb } H_2 \rightarrow 14 \text{ lb } C_2H_4 + 15 \text{ lb } C_2H_6 \tag{2.3}$$

The expected composition of the product exiting the furnace, when accounting for ethane recylce, is shown in Table 2.1 [4].

Table 2.1: Ethane Recycle Yields

Component	Weight Percentage
H_2	3.72
CH_4	3.47
C_2H_2	0.42
C_2H_4	47.97
C_2H_6	40.00
C_3H_6	1.29
C_3H_8	0.03
C_4H_6	1.73
Other C_4's	0.60
Other C_5's	0.77
Total	100.00

Using the hydrogenation reaction equation and the ethane recycle yield composition, the product stream from the furnace was determined iteratively (Table 2.2).

For each iteration, the acetylene was hydrogenated, and then 100% of the ethane was recycled. This was repeated until the recycled ethane was extinct. These iterations were performed on

Table 2.2: Furnace Product Stream Composition

Component	Mass Flow Rate (lb/hr)	Mass Percent in Stream
H_2	2,249	1.6
CH_4	26,409	18.9
C_2H_2	844	0.6
C_2H_4	55,590	39.7
C_2H_6	10,600	7.6
C_3H_4	1,582	1.1
C_3H_6	17,523	12.5
C_3H_8	841	0.6
C_4H_6	4,650	3.3
C_4H_8	4,183	3.0
C_4H_{10}	2,814	2.0
Other C_5's	2,576	1.8
Benzene	6,440	4.6
Other C_6's	392	0.3
Toluene	1,120	0.8
Other C_7's	98	0.1
C_8's	392	0.3
Fuel Oil	1,708	1.2
Total	140,010	100.0

Microsoft Excel (for details on Appendix D). Using this information, the ethane recycle rate was determined, assuming 100% ethane recycle.

2.3 Ethane Recycle Rate

To determine the ethane recycle rate, the furnace was illustrated into two parts: a cracking furnace and a recycle furnace (Figure 2.1).

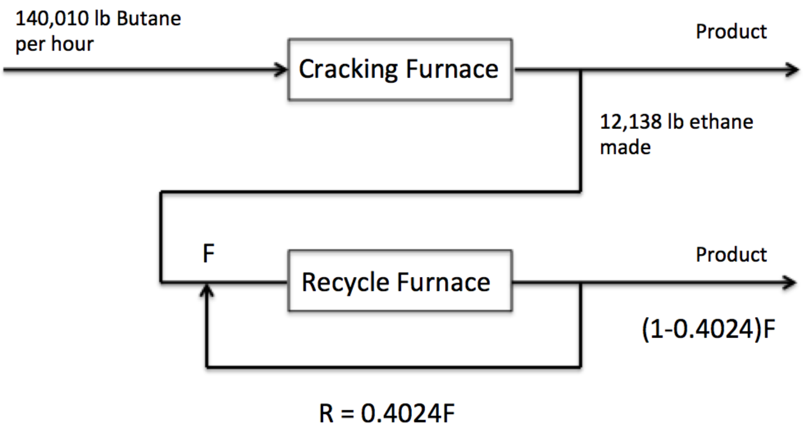

FIGURE 2.1: To determine the ethane recycle rate, the furnace was illustrated into two parts: a cracking furnace and a recycle furnace.

The 12,138 lb ethane made from the cracking furnace, which is subsequently fed into the recycle furnace, was calculated from the feed and the hydrogenation reaction (see Appendix D for more detail). The ethane recycle of interest is R, which is equal to 40.24% (accounting for the recycle yield and the hydrogenation reaction) of the sum of the recycle and the ethane fed into the recycle furnace, F. Using this information, the ethane recycle was found from a mass balance:

$$F = \frac{12,138}{1 - 0.4024} = 20,313 \text{ lb/hr} \tag{2.4}$$

$$R = 0.4024 \times 20,313 = 8,174 \text{ lb ethane/hr} \tag{2.5}$$

Therefore, 8,174 lb ethane are recycled per hour. These calculations were used for the energy balance on the furnace, which is necessary to design, size, and cost the furnace. A summary of the mass balance results on the furnace are shown in Table 2.3.

TABLE 2.3: Summary of Mass Balance

Design Factor	Mass Flow Rate (lb/hr)
Plant Capacity	64,950
Feed Requirement	140,010
Ethane Recycle Rate	8,174

2.4 Feed Cost

By completing the mass balance, the feed operating cost can be determined. Since butane costs \$0.26/lb [5], the operating cost for the feed is \$36,400/hr, which is equivalent to \$305,781,840 per year.

Furnace Energy Balance

An energy balance was performed on the furnace in order to design, size, and cost the furnace. The furnace consists of a radiation section and a convection section, as shown in Figure 3.1. In the convective section (the top of Figure 3.1), the hydrocarbon and steam feeds are preheated prior to entering the radiation section (the bottom of Figure 3.1). In the radiation section, the air and fuel are fed to produce flue gas via combustion. Furthermore, the heat required for cracking the hydrocarbon are supplied, in order to produce the required effluent needed.

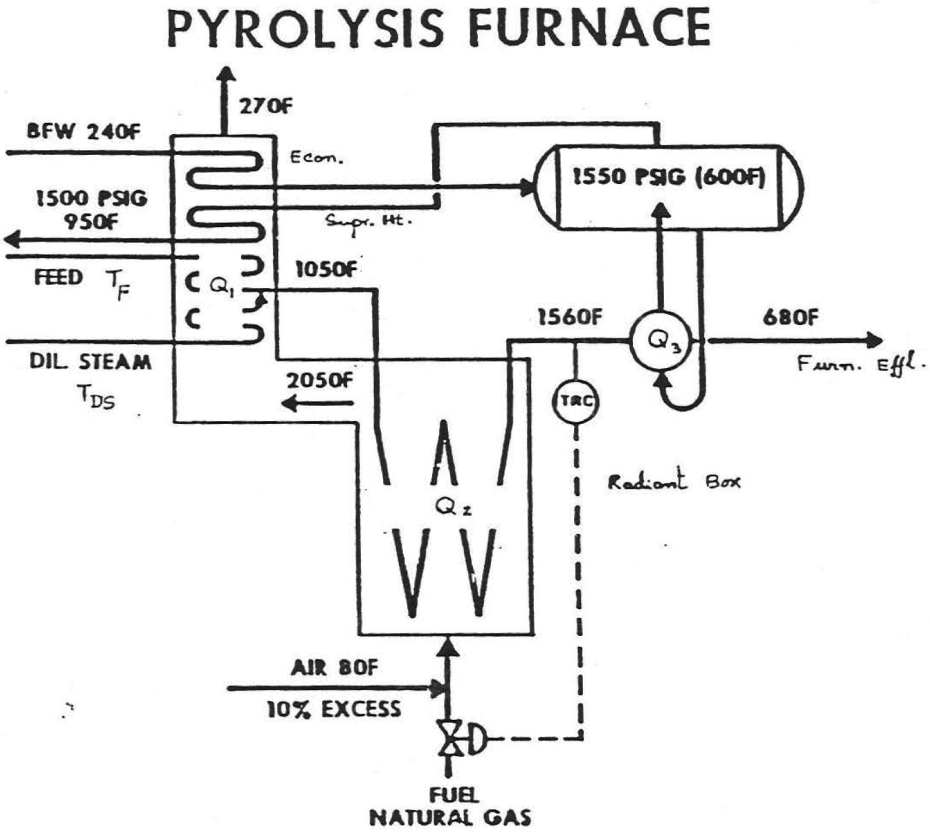

Figure 3.1: A schematic of the pyrolysis furnace. The furnace consists of a radiation section and a convection section. In the convective section, the hydrocarbon and steam feeds are preheated prior to entering the radiation section. In the radiation section, the air and fuel are fed to produce flue gas via combustion. Furthermore, the heat required for cracking the hydrocarbon are supplied, in order to produce the required effluent needed.

3.1 Radiation Section Energy Balance

The heat required for the radiative section of the furnace is found by adding the heat required by the cracking reaction and the sensible heat required to heat the feed stream to the coil outlet temperature (COT) while taking a 1.5% heat loss into account.

3.1.1 Heat of Cracking

First, a heat balance was performed to find the heat of cracking. The heat of cracking can be found from:

$$Q_{\text{cracking}} = \dot{m}(\sum_i H_{f,products} - \sum_j H_{f,reactants}) \tag{3.1}$$

where Q_{cracking} is the heat of cracking, i represents all the product components, j represents all the reactant components, and $\dot{m}$ is the mass flow rate. Using this equation, the heat of cracking is 146.8 MM BTU/hr (for detailed calculations of the heat of cracking, see Appendix D).

3.1.2 Sensible Heat

The sensible heat, or the heat required to increase the temperature of the cracked gas (diluent stream and hydrocarbon stream) from crossover temperature (XOT) to the COT is:

$$Q_{\text{sensible}} = \dot{m}_{steam}(h_{COT,steam} - h_{XOT,steam}) + \dot{m}_{gas}\bar{C}_p(T_{COT} - T_{XOT}) \tag{3.2}$$

$$\begin{aligned} Q_{\text{sensible}} = 56,004\,\frac{\text{lb}}{\text{hr}}(1610.56\,\frac{\text{BTU}}{\text{lb}} - 1448.02\,\frac{\text{BTU}}{\text{lb}}) \\ + 140,010\,\frac{\text{lb}}{\text{hr}} \times 0.91043\frac{\text{BTU}}{\text{lb}^\circ\text{F}}(1550^\circ\text{F} - 1180^\circ\text{F}) = 56.3 \text{ MM BTU/hr} \end{aligned} \tag{3.3}$$

where Q_{sensible} is the sensible heat, $\dot{m}_{steam}$ is the mass flow rate of the steam, $h_{COT,steam}$ is the enthalpy of the steam at COT, $h_{XOT,steam}$ is the enthalpy of the steam at XOT, $\dot{m}_{gas}$ is the mass flow rate of the cracked gas, $\bar{C}_p$ is the average specific heat, T_{COT} is the coil outlet temperature, and T_{XOT} is the crossover temperature.

The mass flow rate of the steam was determined by noting that the steam feed is 40% by mass of the butane feed [4]. The enthalpies of the steam were found at COT and XOT. The average specific heat of the gas was found by linear interpolation at the average temperature of COT and XOT [4]. Therefore, the sensible heat was 56.3 MM BTU/hr.

3.1.3 Process Duty

The process duty is found by adding the sensible heat and the heat of cracking:

$$Q_{\text{Process Duty}} = Q_{\text{cracking}} + Q_{\text{sensible}} \tag{3.4}$$

$$Q_{\text{Process Duty}} = 146.8 \text{ MM BTU/hr} + 56.3 \text{ MM BTU/hr} = 203.0 \text{ MM BTU/hr} \tag{3.5}$$

Therefore, the ideal process duty is 203.0 MM BTU/hr. However, since there is a 1.50% heat loss, the overall process duty within the radiation section (denoted as $Q_{\text{radiation}}$) is 206.1 MM BTU/hr.

3.1.4 Flue Gas Requirements

Next, the flue gas requirements were calculated from the overall process duty within the radiation section, $Q_{\text{radiation}}$, and the change in enthalpy from temperature variations within the furnace:

$$\dot{m}_{\text{flue gas}} = \frac{Q_{\text{radiation}}}{h_{\text{in,flue gas}} - h_{\text{firebox, flue gas}}} \tag{3.6}$$

$$\dot{m}_{\text{flue gas}} = \frac{206.1 \text{ MM BTU/hr}}{1045.4 \text{ BTU/lb} - 616.1 \text{ BTU/lb}} = 480,030 \text{ lb/hr} \tag{3.7}$$

where $h_{\text{firebox,flue gas}}$ is the enthalpy of the flue gas at the firebox, and $h_{\text{in,flue gas}}$ is the enthalpy of the flue gas in. Therefore, the flue gas required was 480,030 lb/hr. The enthalpy of the gas into the furnace was determined by the combustion reaction:

$$1 \text{ lb } CH_4 + 19.66 \text{ lb Air} \rightarrow 20.66 \text{ lb Flue Gas} + \text{Lower Heating Value} \tag{3.8}$$

where the lower heating value (LHV) was 21,720 BTU. Using this combustion reaction, $h_{\text{in,flue gas}}$ was:

$$h_{\text{in,flue gas}} = \frac{\text{Fuel LHV}}{\text{lb Flue Gas/lb Methane}} \tag{3.9}$$

$$h_{\text{in,flue gas}} = \frac{21,720 \text{ BTU/lb Methane}}{20.66 \text{ lb Flue Gas/lb Methane}} = 1045.4 \text{ BTU/lb Flue Gas} \tag{3.10}$$

which is equivalent to the value $h_{\text{in,flue gas}}$ used to determine $\dot{m}_{\text{flue gas}}$. The enthalpy in the firebox, $h_{\text{firebox,flue gas}}$, was determined by linear extrapolation of available data [4].

By calculating the flue gas needed, it was possible to find the fuel needed by using the same combustion Equation 3.8. Therefore, the amount of fuel needed was 23,104 lb/hr. A similar approach was used for the convective section energy balance, and the heat exchangers heat duties.

3.2 Heat Exchangers Energy Balance

The USX and TLX heat exchangers aid in cooling the feed stream after it has left the furnace. The heat duties for both these heat exchangers can be found by adding the heat of the cracked gas, and the heat of the steam. In other words:

$$Q_{\text{Heat Exchangers}} = Q_{\text{cracked gas}} + Q_{\text{steam}} \tag{3.11}$$

where:

$$Q_{\text{steam}} = \dot{m}_{\text{steam}}(h_{COT,steam} - h_{exit,steam}) = 27.5 \text{ MM BTU/hr} \tag{3.12}$$

and:

$$Q_{\text{cracked gas}} = \dot{m}_{\text{cracked gas}}\bar{C}_p(T_{COT} - T_{exit}) = 106.0 \text{ MM BTU/hr} \tag{3.13}$$

Therefore, the heat duty for the heat exchangers is:

$$Q_{\text{Heat Exchangers}} = 106.0 \text{ MM BTU/hr} + 27.5 \text{ MM BTU/hr} = 133.4 \text{ MM BTU/hr} \tag{3.14}$$

This heat exchanger heat duty of 133.4 MM BTU/hr is used to find the total heat available to make steam, as discussed in the following section.

3.3 Convective Section Energy Balance

The amount of heat available in the convective section of the pyrolysis furnace is estimated using the heat of flue gas exiting the radiative section of the furnace, and the given temperature limit on the flue gas exiting the furnace stack. The heat needed to preheat the dilution steam and the feed stream to the cross-over temperature (XOT) is then subtracted from the estimated heat available in the convective section to generate the heat available for preheating boiler feed water (BFW) and superheating high pressure steam to the designated conditions of 875°F at 1750 psig. These calculations are discussed more in detail below.

3.3.1 Heat Available

To find the heat available, it was assumed that 0.50% of the heat was lost in the convective section. Using this assumption, the heat available is:

$$Q_{\text{available}} = 0.9995\dot{m}_{\text{flue gas}}(h_{\text{firebox, flue gas}} - h_{\text{out stack, flue gas}}) \tag{3.15}$$

$$Q_{\text{available}} = (0.9995)480,030\text{lb/hr}(616.1 \text{ BTU/lb} - 54.7 \text{ BTU/lb}) = 269.5 \text{ MM BTU/hr} \tag{3.16}$$

where $h_{\text{out stack, flue gas}}$ is the enthalpy of the flue gas out of the stack in the convective section. Therefore, the heat available in the convective section was 269.5 MM BTU/hr. The enthalpy of the flue gas out of the stack was found from interpolation of available data [4].

3.3.2 Energy Needed to Pre-Heat

Next, the energy needed to pre-heat the dilution steam, from 475 °F to XOT, and the cracked gas, from 60 °F to XOT, was found in order to find the heat of the economizer stream and the superheated stream. The heat needed to pre-heat the dilution steam and the cracked gas is:

$$Q_{\text{pre-heat}} = \dot{m}_{steam}(h_{\text{XOT,steam}} - h_{475^{\circ}\text{F},steam}) + \dot{m}_{hydrocarbon}\bar{C}_p(T_{XOT} - 60^{\circ}\text{F}) \tag{3.17}$$

$$\begin{aligned} Q_{\text{pre-heat}} = {} & 56,004 \text{ lb/hr}(1573.33 \text{ BTU/lb} - 1263.00 \text{ BTU/lb}) \\ & + 140,010 \text{ lb/hr} 0.68 \frac{\text{BTU}}{\text{lb }^{\circ}\text{F}}(1180^{\circ}\text{F} - 60^{\circ}\text{F}) = 123.4 \text{ MM BTU/hr} \end{aligned} \tag{3.18}$$

The enthalpies and specific heats were determined by linear interpolation from available data [4]. The heat needed to pre-heat the dilution steam and the gas in the convective section was 123.4 MM BTU/hr. This was used to find the superheated and economizer steam.

3.3.3 Heats of Superheated and Economizer Stream

Next, the heats of the superheated and economizer streams were found. The heat of the economizer and superheated streams was the excess available heat above the pre-heat energy. In other words.

$$Q_{\text{superheated and economizer}} = Q_{\text{available}} - Q_{\text{pre-heat}} \tag{3.19}$$

$$Q_{\text{super and econ}} = 269.5 \text{ MM BTU/hr} - 123.4 \text{ MM BTU/hr} = 146.2 \text{ MM BTU/hr} \tag{3.20}$$

where $Q_{\text{super and econ}}$ is the heat of the superheated and economizer streams. By finding the heat of the superheated and economizer streams, it is possible to find the total heat available to make steam. Specifically, the heats of the superheated and economizer streams, as well as the heat exchanger heat duties can be used to find the total heat available to make steam:

$$Q_{steam} = Q_{\text{superheated and economizer}} + Q_{\text{Heat Exchangers}} \tag{3.21}$$

$$Q_{steam} = 146.2 \text{ MM BTU/hr} + 133.4 \text{ MM BTU/hr} = 279.6 \text{ MM BTU/hr} \tag{3.22}$$

From the total heat available to make steam, it is possible to find the rate of steam generated by finding the enthalpies of BFW and superheated states. Therefore:

$$\dot{m}_{steam} = \frac{Q_{steam}}{h_{superheated} - h_{BFW}} \tag{3.23}$$

$$\dot{m}_{steam} = \frac{279.6 \text{ MM BTU/hr}}{1461.315\text{BTU/lb} - 212.12 \text{ BTU/lb}} = 223,828 \text{ lb/hr} \tag{3.24}$$

where the enthalpies at the BFW and superheated states were found from available data [6]. Finally, using the rate of steam generated, it was possible to find the superheated stream heat, and the economizer stream heat:

$$Q_{\text{superheated}} = \dot{m}_{superheated}(h_{superheated} - h_{saturated}) = 66.5 \text{ MM BTU/hr} \tag{3.25}$$

$$Q_{\text{economizer}} = Q_{\text{superheated and economizer}} - Q_{\text{superheated}} = 79.7 \text{ MM BTU/hr} \tag{3.26}$$

A summary of all the furnace heat balances are shown in Tables 3.1, 3.2, and 3.3.

Table 3.1: Radiation Section Summary

Energy Balance Factor	Value
Feed Sensible Heat	56.3 MM BTU/hr
Feed Heat of Cracking	146.8 MM BTU/hr
Total Process Duty	203.1 MM BTU/hr
Heat Loss	1.5 %
Radiative Heat Required	206.1 MM BTU/hr
Flue Gas Required	480,030 lb/hr
Fuel Required	23,104 lb/hr
Fuel Heat	501.9 MM BTU/hr

Table 3.2: Heat Exchanger Section Summary

Energy Balance Factor	Value
Cracked Feed Heat	106.0 MM BTU/hr
Steam Heat	27.45 MM BTU/hr
Total Heat Duty	133.4 MM BTU/hr

Table 3.3: Convective Section Summary

Energy Balance Factor	Value
Heat Loss	0.5 %
Heat Available	269.5 MM BTU/hr
Pre-heat Energy Required	146.2 MM BTU/hr
Steam Heat Required	17.4 MM BTU/hr
Steam Heat Available	106.0 MM BTU/hr
Total Heat of Steam	279.6 MM BTU/hr
Steam Made	223,838 lb/hr
Superheating Stream Heat	66.5 MM BTU/hr
Economizer Stream Heat	79.7 MM BTU/hr

Furnace Design

By performing an energy balance on the furnace, it is possible to size and cost the furnace. The sizing and costing of the radiation section, convection section, and heat exchangers is discussed below.

4.1 Radiation Section Design and Sizing

To design the radiation section, it is first necessary to select the average radiant transfer flux, or the heat transferred to the charge stock in the radiant section divided by the total radiant section heat transfer surface. After selecting the average radiant transfer flux, the radiant heat area can be calculated as:

$$A_R = \frac{Q_R}{q} \tag{4.1}$$

where Q_R is the radiant duty, and q is the average radiant flux. Assuming an average radiant flux of 20,000 BTU/(hr·ft^2), the radiant heat area is:

$$A_R = \frac{206.1 \text{ MM BTU/hr}}{20,000 \text{ BTU/(hr}\cdot\text{ft}^2)} = 20,304.6 \text{ ft}^2 \tag{4.2}$$

4.2 Convection Section Design and Sizing

The design of the convection section involves a "shiled or shock" bank, or the first 2 or 3 rows of tubes adjacent to the radiant section. Typically, these tubes are oriented in a staggered triangular pitch normal to the flue gas flow. As a result, they are capable of absorbing and screening the residual radiative comoponent. For bare bank tubes, the convective film coefficient is found by:

$$h_c = \frac{2.14 g^{0.6} T_{ga}^{0.28}}{d_o^{0.4}} \tag{4.3}$$

where T_{ga} is the average flue gas temperature, g is the flue gas mass velocity, and d is the tubes outer diameter. A Manaurite 36 XS pipe was selected, with an outer diameter of 3.18 in [4]. In order to find the gas mass velocity, the diameter of the stack is needed. Therefore, the furnace stack was first designed, as shown below.

4.2.1 Stack Design

To find the gas mass velocity, it was necessary to find the stack height, the friction loss per foot stack, and the diameter of the stack. For a draft specification, Dr, of 2 inches of H_2O, the stack length was determined by:

$$L = \frac{Dr}{0.52\rho(1/T_a - 1/T_{ga})} = 262 \text{ ft} \tag{4.4}$$

The frictional loss per foot stack, f, is therefore 0.008, which is used to calculate the diameter stack diameter d_o:

$$d_o = (\frac{16\dot{m}_{flue}T_{ga}}{211,000\pi^2 f})^{0.2} = 7.8 \text{ ft} \tag{4.5}$$

Finally, the gas mass velocity can be calculated as:

$$g = \frac{\dot{m}_{flue}}{\pi d_o^2/4} = 2.76 \ \frac{\text{lb}}{\text{s} \cdot \text{ft}^2} \tag{4.6}$$

Now, since the gas mass velocity was calculated, the bare bank convective film coefficient can be calculated:

$$h_c = \frac{2.14g^{0.6}T_{ga}^{0.28}}{d_o^{0.4}} = 19.7 \text{ BTU/lb°F} \tag{4.7}$$

The radiant coefficient of the hot gas is calculated as:

$$h_{rg} = 0.0025T_g - 0.5 = 2.5 \text{ BTU/lb°F} \tag{4.8}$$

where T_g is the average flue gas temperature. From the convective film coefficient and the radiant coefficient, the overall film coefficient can be calculated as:

$$h_o = 1.1(h_c + h_{rg}) = 22.2 \text{ BTU/lb°F} \tag{4.9}$$

To find the overall heat transfer coefficient, it was necessary to select a fin surface area, A_f, and a total surface of extended surface tube, A_0. As a result, the in tube film resistance (R_i), the external film resistance (R_o), and the tube-wall resistances (R_w) were calculated. The overall heat transfer coefficient can then be found by:

$$U = \frac{1}{R_i + R_o + R_w} = 7.2 \text{ BTU/hr°Fft}^2 \tag{4.10}$$

which is within the expected range of overall heat transfer coefficient for flue gases in heaters (5-15 BTU/hr°Fft^2). Finally, the convective surface requirement can be calculated by:

$$A_c = \frac{Q_c}{\Delta T_{LM}U} = 39,600 \text{ ft}^2 \tag{4.11}$$

4.3 Heat Exchanger Design and Sizing

To design and size the heat exchanger, a heat balance was applied. The area of the heat exchangers required was:

$$A_{HX} = \frac{Q_{HeatExchanger}}{U \Delta T_{LM}} \tag{4.12}$$

where U is assumed to be 150 BTU/hr°Fft^2 from heuristics by Turton *et al.* Therefore, the heat transfer area is 3,142 ft^2. These sizing calculations can be subsequently used to cost the furnace, as discussed in the following section.

4.4 Furnace Costing

To cost the furnace heat exchanger the computational software CAPCOST was used. The cost of the furnace heat exchanger, convective section, and radiative section are shown in Table 4.1.

Table 4.1: Furnace Cost

Component	FOB Cost	Bare Module Cost
Heat Exchanger	938,000	3,090,000
Convective Section	1,830,000	6,020,000
Radiative Section	145,000	477,000

Quench Tower

To prevent coking and polymerization from occurring, the hot effluent from the furnace is cooled in a quench tower. To do so, USX and TLX heat exchangers are used in series to decrease the temperature of the effluent to 650 °F. Furthermore, the gas exiting from the TLX heat exchanger is cooled to 100 °F. To properly design a quench tower to meet these specifications, an energy balance was first applied.

5.1 Energy Balance

The total cooling duty of the hot effluent includes the condensation duty and the sensible cooling duty, such that:

$$Q_{total\ cooling} = Q_{sensible} + Q_{condensation} \tag{5.1}$$

where $Q_{sensible}$ is the sensible cooling duty, and $Q_{condensation}$ is the condensation cooling duty.

First, the sensible cooling duty, $Q_{sensible}$, was calculated. The sensible cooling duty was the heat removed to decrease the temperature of the cracked gas and steam streams from 650°F to 100°F. Therefore, the sensible cooling duty was calculated as:

$$Q_{sensible} = \dot{m}_{gas}\bar{C}_p(650°\text{F} - 100°\text{F}) + \dot{m}_{steam}\bar{C}_p(650°\text{F} - 100°\text{F}) \tag{5.2}$$

$$\begin{aligned} Q_{sensible} = {} & (56,004 \text{ lb/hr})(0.5486\frac{\text{BTU}}{\text{lb°F}}(650°\text{F} - 100°\text{F}) \\ & + (140,010 \text{ lb/hr})(0.5486\frac{\text{BTU}}{\text{lb°F}}(650°\text{F} - 100°\text{F}) = 58.6 \text{ MM BTU/hr} \end{aligned} \tag{5.3}$$

where the average specific heat, $\bar{C}_p$, was found, using available data [4], at the average temperature of the steam and gas entering and leaving the system. Therefore, the sensible cooling heat of the cracked gas and steam streams was 58.6 MM BTU/hr. Next, the condensation duty was calculated as:

$$Q_{condensation} = \dot{m}_{steam}h_{steam} - \dot{m}_{water}h_{steam} \tag{5.4}$$

where h_{steam} is the enthalpy of the steam at this phase transition temperature and pressure, $\dot{m}_{steam}$ is the mass flow rate of the steam prior to condensation, and $\dot{m}_{water}$ is the mass flow rate of the water after condensation. Assuming thermodynamic equilibrium and ideal conditions, the partition coefficient of water was calculated as 2.6. Therefore:

$$Q_{condensation} = 56,004 \text{ lb/hr} \times 970.1 \text{ BTU/lb} - 21,283 \text{ lb/hr} \times 970.1 \text{ BTU/lb} = 33.7 \text{ BTU/hr} \tag{5.5}$$

Therefore, the total cooling duty was found as:

$$Q_{total\ cooling} = 58.6 \text{ MM BTU/hr} + 33.7 \text{ BTU/hr} = 92.3 \text{ BTU/hr} \tag{5.6}$$

These cooling duties were used to find the coolant mass flow rates, by applying requiring the coolant duty to equal the total cooling duty:

$$Q_{total\ cooling} = \dot{m}_{coolant}\bar{C}_p\Delta T \tag{5.7}$$

where $\dot{m}_{coolant}$ is the coolant stream mass flow rate, $\bar{C}_p$ is the average flow rate of the stream, and ΔT is the temperature rise from the specified inlet temperature of 85°F to 115°F.

Therefore, the coolant mass flow rate could be found as:

$$\dot{m}_{coolant} = \frac{Q_{total\ cooling}}{\bar{C}_p\Delta T} \tag{5.8}$$

$$\dot{m}_{coolant} = \frac{92.3 \text{ BTU/hr}}{1.0\ \frac{\text{BTU}}{\text{lb°F}} \times 30\text{°F}} = 3.1 \text{ MM lb/hr} \tag{5.9}$$

A summary of all the key energy balance values calculated are shown in Table 5.1.

TABLE 5.1: Quench Tower Energy Balance Summary

Energy Balance Factor	Value
Condensation Cooling Duty	33.7 MM BTU/hr
Sensible Cooling Duty	58.6 MM BTU/hr
Total Cooling Duty	92.3 MM BTU/hr
Coolant Mass Flow Rate	3.1 MM lb/hr

These energy balance equations were used in order to design and size the quench tower, as shown in the following section.

5.2 Quench Tower Design and Sizing

To design and size the quench tower, the following initial assumptions were made (Table 5.2).

TABLE 5.2: Quench Tower Sizing Assumptions

Sizing Parameter	Value
Tower Cross Section Area (A_t)	9.62 ft^2
Window Area (A_w)	4.81 ft^2
Curtain Area (A_c)	5.25 ft^2

5.2.1 Window and Curtain Areas

To make sure the sizing assumptions were valid, it is necessary to compare the actual window and curtain velocities to the maximum allowable window and curtain velocities. First, the actual window velocity was compared to the maximum allowable window velocity.
The actual window velocity requires the gas flow rate, which is calculated as:

$$v_{gas} = \frac{\dot{m}_{gas} + \dot{m}_{steam}}{\rho_g} = 85.8 \text{ ft}^3/s \tag{5.10}$$

where ρ_g is the average mass of the cracked gas and steam. The actual window velocity could be subsequently found as:

$$v_{window} = \frac{v_{gas}}{A_w} = 17.8 \text{ ft/s} \tag{5.11}$$

The maximum allowable window velocity can be found from:

$$v_{window,max} = 0.58\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 27.7 \text{ ft/s} \tag{5.12}$$

Therefore, the design window velocity is less than the maximum allowable window velocity, as required. Next, the actual curtain velocity was compared to the maximum curtain velocity. The maximum curtain velocity requires the liquid flow rate, which is calculated as:

$$v_{curtain} = \frac{\dot{m}_{coolant} + \dot{m}_{water}}{\rho_{water}} = 13.8 \text{ ft/s} \tag{5.13}$$

The maximum allowable curtain velocity can be found from:

$$v_{curtain,maximum} = 1.15\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 54.8 \text{ ft/s} \tag{5.14}$$

Therefore, the actual curtain velocity is also well within the maximum allowable curtain velocity. Therefore, the intial sizing assumptions made for the curtain and window area were valid.

Next, it was necessary to confirm the assumed tower cross section area, which is shown in the following section.

5.2.2 Tower Cross Section Area

To confirm the assumed tower cross section area, it is necessary to determine the overall heat transfer coefficient in order to find the contact volume of the quench tower, the total contact area, and finally the height of the tower.

The heat transfer coefficient can be determined from the correlation:

$$U_a = 0.026 G^{0.7} L^{0.4} \tag{5.15}$$

where G is the gas flux in the tower, and L is the liquid flux in the tower. The gas flux in the tower, G, is found as:

$$G = \frac{\dot{m}_{gas} + \dot{m}_{steam}}{A_t} = 20,374.7 \ \frac{\text{lb}}{\text{hr ft}^2} \tag{5.16}$$

The liquid flux in the tower, L, is found as:

$$G = \frac{\dot{m}_{coolant} + \dot{m}_{water}}{A_t} = 323,374.2 \ \frac{\text{lb}}{\text{hr ft}^2} \tag{5.17}$$

Therefore, the heat transfer coefficient is:

$$U_a = 0.026 G^{0.7} L^{0.4} = 4,317.4 \ \frac{\text{BTU}}{\text{lb°F}} \tag{5.18}$$

This heat transfer coefficient can be used to determine the overall heat transfer coefficient:

$$U_a = \frac{1}{\frac{1}{h_L a} + \frac{Q_{sensible}}{Q_{total\ cooling}} \frac{1}{\alpha H_a}} \tag{5.19}$$

Assuming negligible resistance from the $1/(h_L a)$, and $\alpha = 1$, the overall heat transfer coefficient is:

$$U = \frac{1}{\frac{Q_{sensible}}{Q_{total\ cooling}} \frac{1}{H_a}} = 6,800 \ \frac{\text{BTU}}{\text{lb°F}} \tag{5.20}$$

Using the overall heat transfer coefficient, the total contact of the quench tower, the cross sectional area of the contact height, and the number of trays can be found. The total contact area, V_T is:

$$V_T = \frac{Q_{total\ cooling}}{U \Delta T_{LM}} \tag{5.21}$$

$$V_T = \frac{Q_{total\ cooling}}{U \Delta T_{LM}} = 76.4 \text{ ft}^3 \tag{5.22}$$

where ΔT_{LM} is the log-mean temperature difference. The cross sectional area of the contact height, Z_T was:

$$Z_T = \frac{V_T}{A_t} = \frac{76.4 \text{ ft}^3}{9.62 \text{ ft}^2} = 7.9 \text{ ft} \tag{5.23}$$

Furthermore, it is necessary to add 2 ft. above and below the top and bottom trays, respectively, where the total height is:

$$Z_T = 7.9 \text{ ft} + 4 \text{ ft} = 11.9 \text{ ft} \tag{5.24}$$

Finally, the number of trays, N, was found by assuing a tray spacing of 1.5 ft:

$$N = \frac{Z_t}{1.5} = 6 \text{ trays} \tag{5.25}$$

Using Turton *et al* heuristics [7]:

$$N_{actual} = 6 \text{ trays} \times 2.2 = 14 \text{ trays} \tag{5.26}$$

5.2.3 Tower Diameter

The tower diameter, D, is calculated by using the cross sectional area:

$$D = \sqrt{4A_t/\pi} = 3.5 \text{ ft} \tag{5.27}$$

Therefore, the tower diameter is 3.5 ft.

5.2.4 Wall Thickness

To determine the wall thickness of the quench tower, t, the following equation is used:

$$t = \frac{P \times R}{2S \times \epsilon - 0.6P} + C \times A \tag{5.28}$$

where P is the design pressure (50 psia), R is the tower radius (1.75 ft), S is the stress allowance (15,000 psi for carbon steel), ϵ is the joint efficiency (0.85 for carbon steel), and $C \times A$ is the corrosion allowance (0.125 in). Therefore:

$$t = \frac{50 \text{ psia} \times 21 \text{ in}}{2(15,000 \text{ psi}) \times 0.85 - 0.6(50 \text{ psia})} + 0.125 \text{ in} = .17 \text{ in} \tag{5.29}$$

Rounding up to the next 1/8", the tower wall thickness is 0.25". A summary of the quench tower design and sizing is shown in Table 5.3. These values can be used to cost the quench tower, as shown in the following section.

TABLE 5.3: Quench Tower Design Summary

Design Factor	Value
Tower Cross Section Area (A_t)	9.62 ft^2
Window Area (A_w)	4.81 ft^2
Curtain Area (A_c)	5.25 ft^2
Tower Diameter (D)	3.5 ft
Heat Transfer Coefficient (U)	$6,800$ $\frac{\text{BTU}}{\text{lb°F}}$
Total Contact Area (V_T)	76.4 ft^3
Cross Sectional Area Contact Height (Z_T)	7.9 ft
Tower Height (Z_T)	11.9 ft
Tower Thickness (t)	0.25 in

5.3 Quench Tower Costing

Finally, by completing the quench tower design, it is possible to cost the quench tower. To do so, it is necessary to determine the tower weight, which requires the shell and head volumes. The shell volume is given by:

$$V_s = \pi DtL = 2.73 \text{ ft}^3 \tag{5.30}$$

Next, the head volume was found by assuming a spherical head, where:

$$V_h = \pi D^2 t = 0.80 \text{ ft}^3 \tag{5.31}$$

Therefore, the total volume of steel needed for the column is:

$$V_{total} = V_s + V_h = 3.53 \text{ ft}^3 \tag{5.32}$$

Since the density of carbon steel is 0.284 lb per in^3, the total tower weight is 1,733 lb. Adding 25% for nozzles and intervals, the total quench tower weight is 2,311 lb. Since the cost of carbon steel is 50 cents per pound, the tower cost is $1,155.50. To account for installation, the cost is $3,500 for the quench tower.

Next, the tray cost was determined. From the tray cost curve, for a tower diameter of 3.5 feet, the cost of one tray is $850. Therefore, the total tray cost is $11,900. To account for installation, total cost is $35,700.

5.4 Quench Tower Operating Costs

The quench tower requires operating costs for the cooling water. The cooling water operates at a cost of $0.09 per 100 gallons. Therefore, the operating cost of the quench tower is $44 per hour, which is approximately $370,000 per year (assuming 8400 operating hours per year).

A summary of the capital and operating costs of the quench tower are shown below (Tables 5.4 and 5.5, respectively).

Table 5.4: Quench Tower Capital Cost Summary

Item	FOB Cost	Module Cost ($)
Quench Tower	1,160	3,500
Trays	11,900	35,700

Table 5.5: Quench Tower Operating Cost Summary

Item	Cost ($/yr)
Operating Costs (per year)	370,000

Compressor System

After the cooled gas stream leaves the quench tower, it is compressed by a system of four compressors (Figure 6.1). The gas enters the system at 19.7 psia and 100°F, and leaves the system at 565 psia and 60°F. A knockout drum and a heat exchanger follow the first three compressor stages, where water and hydrocarbons are removed. It is assumed that the compression ratio is constant, and that there is a 6 psi pressure drop across each knockout drum.

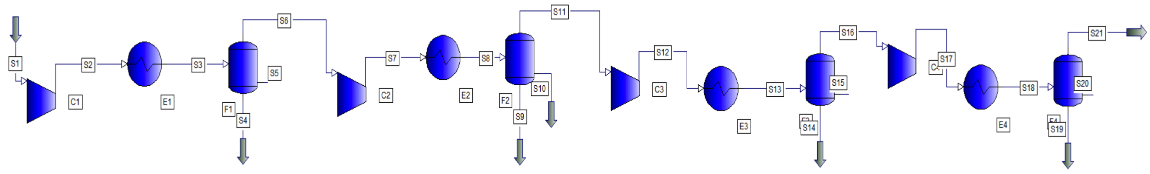

FIGURE 6.1: Schematic of Compressor System. After the cooled gas stream leaves the quench tower, it is compressed by a system of four compressors. The gas enters the system at 19.7 psia and 100°F, and leaves the system at 565 psia and 60°F. A knockout drum and a heat exchanger follow the first three compressor stages, where water and hydrocarbons are removed.

6.1 Compressor Design and Sizing

To design and size the compressor system, the computational tool Pro/II was used. To do so, the following factors were defined within Pro/II:

1. The pressure, temperature, composition, and flowrate of the inlet stream.
2. The hot process outlet fluid temperature of the heat exchangers (100°F).
3. A 6 psi pressure drop for each knockout drum, and that the outlet from the last stream is saturated.
4. The compression ratio for each compressor, γ, where $\gamma = \frac{P_{n+1}}{P_n}$.

The output compressor results, component mass balance, and heat exchanger data is shown in Tables 6.1, 6.2, and 6.3, respectively.

Table 6.1: Compressor Results

Compressor Results	C1	C2	C3	C4
Inlet Temperature (°F)	100.00	100.00	100.00	100.00
Outlet Temperature (°F)	143.48	199.20	201.83	207.68
Inlet Pressure (psia)	20.00	43.10	100.81	242.49
Outlet P (psia)	49.10	105.81	247.49	595.31
Hydrocarbon Flow (lbmol/hr)	6,092.20	5,997.11	5,921.16	5,822.67
Water Content (lb mol/hr)	134.34	56.54	23.21	9.37
Specific Volume (ft^3/lb mol)	137.77	57.99	23.21	8.69
Volumetric Flow (ft^3/min)	13,988.61	5,795.80	2,290.32	843.70
Molecular weight (lb/lb mol)	21.94	23.24	23.10	22.77
Enthalpy (BTU/hr)	6,374,400.65	6,521,938.84	6,313,485.46	5,997,163.16
Horsepower	2,505.14	2,563.12	2,481.20	2,356.89
Frame	46M	46M	29M	29M
Head/Wheel	9,050.00	9,050.00	9,918.00	9,918.00
Polytropic Efficiency	0.77	0.77	0.76	0.76
Adiabatic Efficiency	0.74	0.74	0.73	0.73
Adiabatic Head	23,375.47	26,048.42	25,538.06	24,377.78
Number of Wheels	3.00	3.00	3.00	3.00
Actual Head/Wheel	7,791.82	8,682.81	8,512.69	8,125.93
% Error	-0.14	-0.04	-0.14	-0.18

Table 6.2: Component Mass Balance

Component	C1 (lb/hr)	C2 (lb/hr)	C3 (lb/hr)	C4 (lb/hr)
H_2	1,115.91	1,115.90	1,115.84	1,115.33
CH_4	1,646.21	1,646.08	1,645.37	1,641.35
C_2H_2	32.43	32.41	32.34	32.01
C_2H_4	1,981.51	1,980.95	1,978.03	1,964.00
C_2H_6	352.52	352.37	351.64	348.24
C_3H_4	39.46	39.38	38.94	37.38
C_3H_6	416.35	415.80	412.98	402.06
C_3H_8	19.08	19.05	18.93	18.43
C_4H_6	85.95	85.56	83.71	77.76
C_4H_8	74.49	74.17	72.66	67.72
C_4H_10	48.37	48.14	47.15	44.05
Other C_5's	35.66	35.26	33.51	28.70
Benzene	86.82	80.85	60.67	34.40
Toluene	12.96	10.87	5.49	1.77
C_8's	3.49	2.19	0.57	0.09
Fuel Gas	6.65	1.58	0.11	0.00
Total	5,957.85	5,940.57	5,897.96	5,813.29

Table 6.3: Heat Exchanger Characteristics

Heat Exchanger Properties	E1	E2	E3	E4
Duty (BTU/hr)	22,000.00	19,100.00	18,200.00	19,600.00
Cooling Water Inlet (°F)	80.00	80.00	80.00	80.00
Cooling Water Outlet (°F)	120.00	120.00	120.00	120.00
Process Fluid Outlet Temp (°F)	100.00	100.00	100.00	100.00
Process Fluid Inlet Temp (°F)	143.48	199.20	201.83	207.68
ΔT_{LM}	21.69	43.02	43.89	45.79
U (BTU/(°F $\cdot$ ft^2 $\cdot$ hr)	10.00	10.00	10.00	10.00
Area (ft^2)	101.41	44.40	41.47	42.80
C_p water	1.00	1.00	1.00	1.00
Δ T	40.00	40.00	40.00	40.00
$\dot{m}_{water}$ (lb/hr water)	550.77	478.17	455.64	490.69

6.2 Compressor Costing

By calculating the output compressor results, component mass balance, and heat exchanger data, it was possible to find the compressor costs. The compressor system costs include compressor electricity costs, compressor FOB and installation costs, knockout drum FOB and installation costs, heat exchanger FOB and installation costs, and cooling water costs.

6.2.1 Compressor Costs

The compressor horsepower was used to calculate the compressor electricity costs (Table 6.4).

Table 6.4: Compressor Electricity Costs

Cost	C1	C2	C3	C4
Hourly Electricity Cost ($/hr)	112.08	114.68	111.01	105.45
Yearly Electricity Cost ($)	941,513.41	963,305.14	932,516.10	885,794.58

The compressor capital cost was also calculated, by using the computational software CAPCOST. The compressor capital cost is shown in Table 6.5.

Table 6.5: Compressor Capital Cost

Component	FOB Cost	Bare Module Cost
Compressor 1	635,000	1,740,000
Compressor 2	645,000	1,770,000
Compressor 3	630,000	1,730,000
Compressor 4	608,000	1,670,000

6.2.2 Knockout Drum Costs

Similarly, the knockout drum costs were calculated using the computational software CAPCOST. The knockout drum capital costs are shown in Table 6.6.

Table 6.6: Knockout Drum Capital Cost

Component	FOB Cost	Bare Module Cost
Drum 1	21,100	63,300
Drum 2	11,800	35,500
Drum 3	6,300	18,900
Drum 4	3,300	9,800

6.2.3 Heat Exchanger Costs

The heat exchanger costs were calculated using the computational software CAPCOST. The heat exchanger capital costs are shown in Table 6.7.

Table 6.7: Heat Exchanger Capital Cost

Component	FOB Cost	Bare Module Cost
Heat Exchanger 1	5,360	17,600
Heat Exchanger 2	4,550	15,000
Heat Exchanger 3	4,480	14,700
Heat Exchanger 4	4,510	14,900

6.2.4 Cooling Water Costs

To determine the cooling water costing, the $\dot{m}_{water}$ was used for each heat exchanger. Cooling water is priced at 10 center per 1000 gallons, which resulted in the following cooling water costs (Table 6.8).

Table 6.8: Compressor Cooling Water Costs

Cost	C1	C2	C3	C4
Yearly Water Cost ($)	8,684.56	7,539.78	4,480.00	31,145.98

Demethanizer Section

7.1 Demethanizer Process Summary

The refrigeration section includes two parts (with 3 stages each), which are used to bring the process fluid temperature down from 100°F to -145°F for the cryogentic fractionation processes.

The following steps were followed to design the first section of the refrigeration section:

1. The total heat duty across all 3 heat exchangers was determined in Pro/II by taking the outlet stream from the compressor section and specifying the outlet temperature. This heat duty was divided across the 3 heat exchangeres with exchanger"A" having the largest heat duty and decreasing across each (Figure 7.1).

2. The pressure in the extra tank is the saturation pressure of propylene at 22°F to -145°F because the stream is at the same conditions in the extra tank and entering heat exchanger "A" (Figure 7.1). A 5 psi pressure drop across the extra heat exchanger is assumed to give the pressure exiting compressor "A" (Figure 7.1). The saturation pressure of propylene at -28°F is the pressure of the stream exiting heat exchanger "C", and a 1.5 psi pressure drop is assumed across each knockout tank so which gives the pressure entering compressor "C" (Figure 7.1).

3. The compression ratio from the pressure entering compressor "C" and exiting compressor "A" (Figure 7.1). This is used to calculate the pressure entering and exiting compressor "B" (Figure 7.1). The vapor stream exiting knockout tanks "A" and "B" are assumed to be equal to the stream it is mixing with, and the liquid stream exiting each of these knockout tanks is at the same pressure as the vapor exit stream.

4. The streams entering and exiting each knockout tank are at saturated conditions and their pressures are known, so a Mollier diagram can be used to determine the specific enthalpy of each of these streams.

5. The mass of propylene flowing through each heat exchanger is calculated using the heat duty and change in enthalpy across the heat exchanger.

6. A mass and energy balance around each knock out tank is performed to determine the flow rate of the vapor exit stream of tanks "A" and "B" (Figure 7.1).

7. The specific enthalpy of the stream entering compressor C was determined from a Mollier diagram in Step 4, and the streams entering compressors "A" and "B" are the weighted averages of the specific enthalpies of the stream exiting the previous compressor and of the stream exiting the knockout tank. The enthalpy of the stream exiting each compressor is determined by following a line of constant entropy on the Mollier diagram from the compressor input pressure (saturated conditions) to the output pressure. The enthalpy difference across each compressor can be converted to horsepower and used to cost the compressor and electricity.

8. The compressor design was selected as 38M based on the total flow rate across all compressors giving the desired head per wheel, adiabatic efficiency, and maximum number of wheels.

9. The specific heat ratio is found for propylene, and used with the compression ratio and adiabatic efficiency to find the polytropic efficiency from a chart. These are used to calculate the adiabatic head (W_{ad}) and polytropic head (W_p):

$$W_{ad} = \Delta h_{ad} \times 778 \frac{\text{ft lb}}{\text{BTU}} \tag{7.1}$$

$$W_p = \frac{W_{ad} \eta_p}{\eta_{ad}} \tag{7.2}$$

10. The number of wheels can then be calculated determined from the adiabatic head and the desired head per wheel.

These steps were repeated for the second section of the refrigeration cycle as well (where A, B, and C become D, E, and F, respectively). Figure 7.1 illustrates the first step of this two stage process. The sizing and costing of this process is shown in the following section.

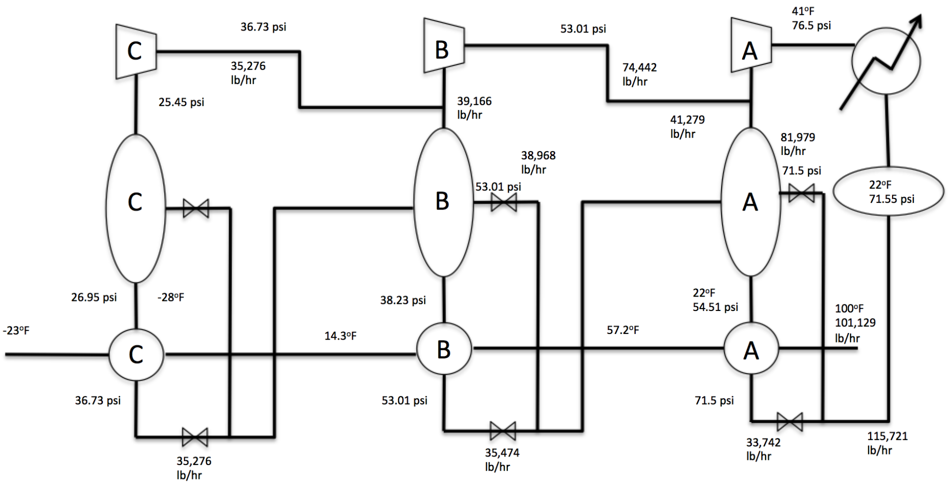

FIGURE 7.1: Schematic of Refrigeration Process. The first cycle of this two stage process is shown, with steam temperatures and pressures.

7.2 Demethanizer Sizing and Costing

Using the methodology previously described, the demethanizer section was sized and costed. The sizing of the tanks, heat exchangers, and compressors is shown in Tables 7.1, 7.2, and 7.3, respectively.

TABLE 7.1: Tank Sizing

Factor	A	B	C	D	E	F
Drum Volume (ft^3)	49,443	23,224	7,852	119,369	64,883	24,718
Drum Diameter (ft)	28	21	15	37	30	22
Drum Length (ft)	80	67	44	111	92	65
Wall Thickness (in)	0.16	0.14	0.13	0.21	0.16	0.13
Rounded Wall Thickness (in)	0.25	0.25	0.25	0.25	0.25	0.25
Shell Volume (ft^3)	147.15	92.16	43.62	268.85	180.23	93.63
Shell Head Volume (ft^3)	51.31	28.86	14.73	89.60	58.90	31.68
Drum Mass (lb)	83,610	69,826	47,185	59,617	43,719	27,737

TABLE 7.2: Heat Exchanger Sizing

Factor	A	B	C	D	E	F	Extra 1	Extra 2
Area (ft^2)	3,449.8	7,409.9	6,954.1	6,073.1	5,545.2	4,125.8	2,711.6	1,796.8

TABLE 7.3: Compressor Sizing

Factor	A	B	C	D	E	F
Flow (cfm)	7,911	3,716	1,256	19,099	10,381	3,955
Horsepower (G HP)	319.81	310.24	110.91	869.67	538.04	262.81

From these sizings, the costs for the tanks, heat exchangers, compressors, and utilities can be found, as shown in Tables 7.4, 7.5, 7.6, and 7.7 respectively.

TABLE 7.4: Tank Costing

Factor	A	B	C	D	E	F
Drum Cost ($)	41,805.45	34,913.22	23,592.75	29,808.78	21,859.56	13,868
Drum Cost Installation ($)	125,416.36	104,739.65	70,778.25	89,426.34	65,578.69	41,606.34

TABLE 7.5: Heat Exchanger Costing

Heat Exchanger	FOD Cost ($)	Bare Module Cost ($)
A	158,000	521,000
B	342,000	1,130,000
C	320,000	1,050,000
D	280,000	921,000
E	255,000	839,000
F	190,000	626,000
Extra 1	125,000	410,000
Extra 2	82,500	272,000

TABLE 7.6: Compressor Costing

Compressor	Electricity Cost ($/hr)	Electricity Cost ($/yr)
A	14.31	120,193.76
B	13.88	116,599.73
C	4.96	41,683.19
D	38.91	326,850.21
E	24.07	202,213.85
F	11.76	98,774.76

Table 7.7: Utility Costing

Cycle	Utility Cost ($/hr)	Utility Cost ($/yr)
1	26.37	221,528.56
2	22.38	187,959.25

Fractionation Columns

8.1 Process Summary

To seperate the valuable products ethylene, propylene, mixed C4's, gasoline, and high pressure stream products, a five tower fractionation system is used. In this process, adeethanizer, depropanizer, and debutanizer are in series, where the subsequent column receives the bottoms of the previous column. Ethylene and propylene towers accept feeds from the deethanizer and depropanizer, respectively (Figure 8.1). To size and cost the fractionation towers, and to meet specifications, the computational tool Pro/II was used.

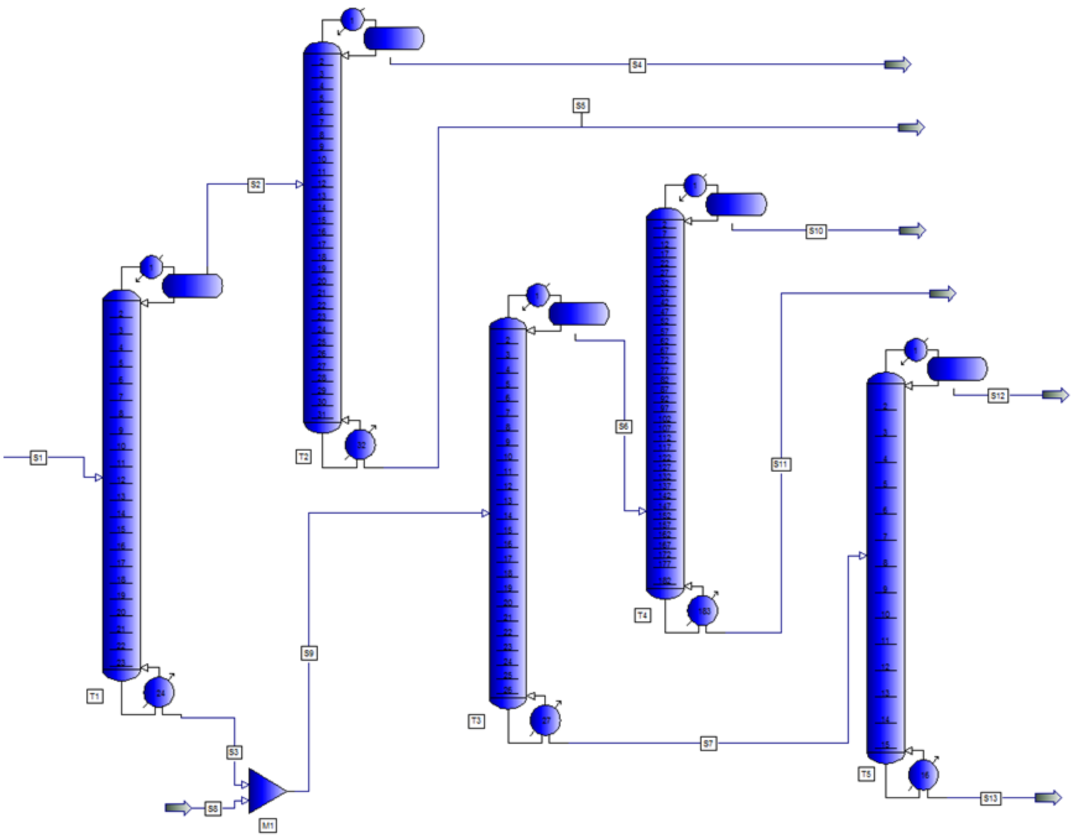

FIGURE 8.1: Schematic of Fractionation Process. A deethanizer, depropanizer, and debutanizer are in series, where the subsequent column receives the bottoms of the previous column. Ethylene and propylene towers accept feeds from the deethanizer and depropanizer, respectively.

The compositions of streams labled S1 and S2 are shown in Table 8.1.

TABLE 8.1: Feed Stream Compositions

Component	S1 Flow Rate (lb/hr)	S2 Flow Rate (lb/hr)
C_2H_4	5370	0
C_2H_6	1620	3
C_3H_6	1430	360
C_3H_8	110	40
C_4's	360	80
C_5's	50	240
Benzenes	10	160
C_7's	0	80

The design specifications for this fractionation process is shown in Table 8.2. These specifications were met by using Pro/II.

TABLE 8.2: Fractionation Process Specifications

Tower	Distillate Specification (mol %)	Bottoms Specification (mol %)
Deethanizer	0.16 C_3H_6	0.26 C_2H_6
Ethylene Tower	99 C_2H_4	99 C_2H_6
Depropanizer	0.04 C_4's	0.2 C_3's
Propylene Tower	0.1 C_3H_8	15 C_3H_6
Debutanizer	0.2 C_5's	1.3 C_4's

8.2 Fractionation Tower Sizing

In order to determine if a single tower or separate rectifying and stripping towers should be used for each distiallation tower, a diameter for each section was calculated. If the two diameters were within 15% of each other, a single tower was used. Otherwise, it was separated into two towers. A distillation column would also be separated into two separate towers if the number of trays was too large, where the tower would be unsafe.

Therefore, the diameters and heights for the rectifying and stripping section were first determined to see whether or not it was necessary to split the tower into two separate towers (Table 8.3).

Table 8.3: Diameter of Rectifying and Stripping Sections for Each Tower

Tower	Diameter (ft)
Deethanizer (Rectifying)	19.3
Deethanizer (Stripping)	28.6
Ethylene Tower (Rectifying)	29.9
Ethylene Tower (Stripping)	25.6
Depropanizer (Rectifying)	9.6
Depropanizer (Stripping)	8.4
Propylene Tower (Rectifying)	29.3
Propylene Tower (Stripping)	29.8
Debutanizer (Rectifying)	6.3
Debutanizer (Stripping)	5.5

As evident from Table 8.3, the ethylene tower must be split into two separate rectifying and stripping towers, as the two diameters were not within 15% of each other. Next, the tower heights were checked to see if it was necessary to split to towers to avoid unsafe conditions (Table 8.4).

Table 8.4: Height of Rectifying and Stripping Sections for Each Tower

Tower	Height (ft)
Deethanizer (Rectifying)	38
Deethanizer (Stripping)	53
Ethylene Tower	98
Depropanizer	88
Propylene Tower	413
Debutanizer	66

Clearly, from Table 8.4, the propylene tower is unsafe. Therefore, it is necessary to split the propylene tower into two separate towers. The resulting sizes for each tower are shown in Table

8.5.

Table 8.5: Fractionation Tower Sizes

Tower	Number of Trays	Feed Tray	Height (ft)	Diameter (ft)
Deethanizer (Rectifying)	11	11	23	19.3
Deethanizer (Stripping)	10	1	38	28.6
Ethylene Tower	32	12	98	29.9
Depropanizer	27	14	88	9.6
Propylene Tower 1	91	91	198	29.3
Propylene Tower 2	92	1	215	29.8
Debutanizer	16	8	66	6.3

8.3 Fractionation Tower Costing

To cost the fractionation towers, the weight of the towers was first calculated. Then, the cost of carbon steel was used to find the total price of each tower. The cost of each tower is shown below:

Table 8.6: Cost of Each Tower

Tower	Module Cost ($)
Deethanizer	20,282,329
Ethylene Tower	35,972,011
Depropanizer	886,607
Propylene Tower	162,150,059
Debutanizer	204,845

8.4 Heat Duties

The reboiler and condenser heat duties were calculated from Pro/II simulations. The condenser and reboiler heat duties, and the associated heat transfer area calculated from an energy balance are shown in Tables 8.7 and 8.8, respectively.

TABLE 8.7: Condenser Heat Duty for Each Tower

Tower	Heat Duty (BTU/hr)	Heat Transfer Area (ft^2)
Deethanizer	6,433,700	11,600
Ethylene Tower	32,205,200	20,500
Depropanizer	6,721,000	8,253
Propylene Tower	42,684,800	24,227
Debutanizer	3,535,600	3,754

TABLE 8.8: Reboiler Heat Duty for Each Tower

Tower	Heat Duty (BTU/hr)	Heat Transfer Area (ft^2)
Deethanizer	27,394,200	2,519
Ethylene Tower	23,557,000	856
Depropanizer	4,694,200	451
Propylene Tower	43,641,800	3,599
Debutanizer	3,031,500	365

8.5 Heat Exchanger Costs

From these heat duties and heat transfer areas, the heat exchanger costs were calculated via CAPCOST. The condenser and reboiler costs are shown in Tables 8.9 and 8.10, respectively.

TABLE 8.9: Condenser Cost for Each Tower

Tower	Bare Module ($)	Utility Cost ($/hr)
Deethanizer	1,770,000	291,425
Ethylene Tower	3,110,000	1,458,786
Depropanizer	1,250,000	304,438
Propylene Tower	544,000	1,933,476
Debutanizer	53,900	160,151

Table 8.10: Reboiler Cost for Each Tower

Tower	Bare Module ($)	Utility Cost ($/hr)
Deethanizer	1,110,000	1,643,471
Ethylene Tower	377,000	1,413,265
Depropanizer	214,000	281,621
Propylene Tower	1,580,000	2,618,220
Debutanizer	183,000	181,870

The remaining heat drum and heat pump costs were calculated in CAPCOST, and were accounted for in the economic analysis, as discussed in the following section.

Costs and Economic Analysis

9.1 Capital and Operational Costs

Each section of the ethylene producing plant was designed, sized, and costed. A summary of the resulting capital and operational costs are shown below (Table 9.1).

Table 9.1: Costing Summary

Equipment	Bare Module Cost ($)	Annual Operational Cost ($)
Furnace	3,567,000	0
Quench Tower	70,700	370,000
Compressor Section	7,099,597	3,754,276
Refrigerant Section	7,811,372	1,315,803
Fractionation Section	229,859,450	10,371,196
Total	248,408,119	15,811,275

A summary of the resulting revenue from the product stream is shown in Table 9.2.

Table 9.2: Revenue Summary

Component	Flow Rate (lb/hr)	Value ($/lb)	Annual Revenue ($)
Ethylene	64,950	0.60	327,348,000
Propylene	16,791	0.16	21,862,077
Gasoline	6,297	0.45	23,802,746
Total			373,012,823

The total bare module costs, the annual costs, and the revenue stream were used to perform an economic analysis on the ethylene producing plant, in order to determine the profitability of this process.

9.2 Economic Analysis

In performing the economic analysis for this ethylene plant, certain assumptions were made:

1. It is assumed that the plant takes two years to be built and three years to reach full production capacity.
2. The plant is assumed to operate for ten years.
3. A building value of $5.5 million and working capital of $3.5 million are used.
4. Fixed costs, which include payroll, salaries, supplies, direct overhead, and indirect overhead, are $4.5 million per year.
5. The variable costs include the purchase of butane, the selling of ethylene, propylene, and fuel, the cost of utilities, and packing and freight of the products.
6. The calculation of all variable costs has been accounted for while the packing and freight of the products have been approximated at $15,000 per ton of ethylene produced.
7. Additionally, an investment grant of 15% of building and plant investment is earned each year.
8. The building is assumed to have a straight line depreciation (5% per year) whereas the plant has a reducing balance depreciation (15% per year).
9. Corporate tax is assumed to be 45%, a tax allowance is earned on the depreciation of the building and plant.

Using the assumptions above, the economic analysis is performed to generate a internal rate of return (IRR) of 16%, indicating that this design is a profitable investment. Over the life of the plant, a profit of approximately $160 million is accumulated. Figure 9.1 illustrates the cumulative cash flow and cumulative discounted cash flow rate over the life of the plant. Based on this economic analysis, the design and construction of this plant is highly profitable.

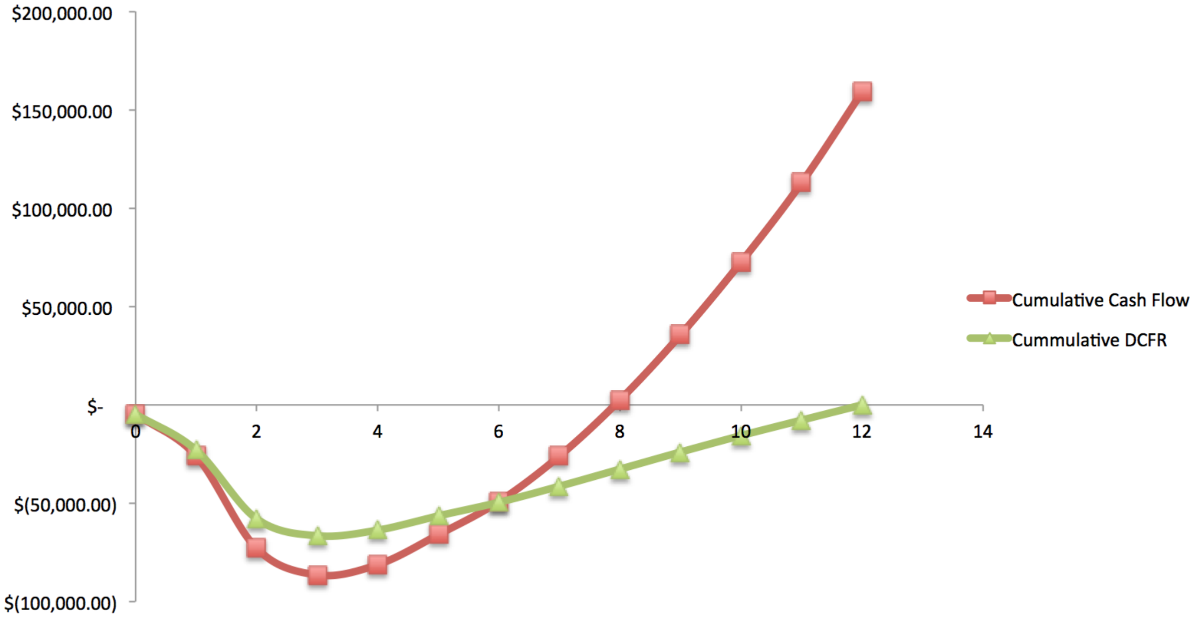

FIGURE 9.1: An economic analysis was performed to find the potential profitability of this plant. Over the life of the plant, a profit of approximately $160 million is accumulated. The cumulative cash flow and cumulative discounted cash flow rates are shown.

Conclusions and Recommendations

10.1 Recommendations

There are numerous recommendations for the design of this chemical plant. As evident from Table 8.3, the ethylene tower must be split into two separate rectifying and stripping towers, as the two diameters were not within 15% of each other. Furthermore, the propylene tower is unsafe due to its height. Therefore, it is recommended to split the propylene tower into two towers. Another design recommendation would be to use more process streams within the plant as utilities for heating and cooling to be economically efficient.

10.2 Conclusion

An ethylene production plant was designed to meet a product specification of 700 metric tons per day. To do so, 140,010 lb/hr of 100% butane is fed to the plant, and 100% of ethane is recycled at a rate of 8,174 lb/hr. This plant process includes: a furnace (to crack the hydrocarbon feed), a quench tower (to cool the exiting stream), a four stage compressor system (to increase the pressure of the cracked gases), a refrigeration section (to separate out methane and hydrogen), and a five tower fractionation system (to separate the remaining products). The resulting ethylene, propylene, mixed C4's, raw gasoline, and high pressure stream products are subsequently sold.

An economic analysis was performed on the ethylene plant designed, where it is expected that the plant will profit approximately 160 million over a 10 year operation period, with a return on investment of 16%. It was assumed for the economic analysis that the plant will run 8,400 hours a year (0.96 plant operating factor). The capital investment for the plant construction was \$248,000,000. The utilities to run the compressors and pumps, including low pressure steam, cooling water, and electricity, cost approximately \$16,000,000 per year. To reduce unnecessary utility costs, certain process streams were used for heating and cooling. Since it is assumed that wage costs are \$4.5 million per year, the total annual operating cost is \$20,500,000 per year.

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Appendix A: Safety Concerns

Considering that modern chemical plants are complex hybrid systems consisting of various process and control equipment, the operation of chemical plants entail diverse risks. Thus, the design of plants is to be accompanied with careful and responsible hazard analysis procedures. Specifically, design and operation of this ethylene production plant requires special measures for protection of personnel as well as the environment. All employees are familiarized with previously established emergency response plans and procedures. Since this plant utilizes flammable materials, sometimes at high pressure and temperature, fire safety precautions must be established. This includes accessibility to fire extinguishers, water sprays, and dry chemical and carbon dioxide systems. Additionally, employees are required to wear goggles, gloves, hardhats, and fire-retardant clothing within the vicinity of the plant. It is highly recommended that exposure to process components be limited. In case of exposure to high-risk chemicals, breathing equipment is provided to employees. A flare system is installed to burn off escaped gases in order to prevent release of harmful gases to the environment. A proper removal system of the carbon monoxide being released as part of the furnace flue gas is also installed.

As in any plant, auxiliary systems have been designed in case of failure. These systems include, but are not limited to, emergency valves, holdup tanks, and compressors. If installation of such systems is not cost-efficient, emergency shut-off systems will be installed instead to ensure safety. This ethylene production plant has been designed and will be constructed with the highest regard for employee safety.

Appendix B: Environmental Implications

The design of this plan will also be accompanied with environmentally friendly procedures. For example, sources of waste are properly dealt with according to environmental considerations and regulations. Recycle and reuse of some materials such as fuel oil are established wherever possible for both environmental friendliness as well as improved plant efficiency. This ethylene production plant has been designed and will be constructed with the highest regard for environment impacts.

Appendix C: Sample Calculations

Plant Capacity Calculations

$$\text{Plant Capacity} = 700\frac{\text{ton}}{\text{day}} \times 2,204.6\frac{\text{lb}}{\text{ton}} \times (1/24)\frac{\text{day}}{\text{hours}} = 64950 \text{ lb/hr} \tag{.1}$$

Mass Balance Calculations

$$F = \frac{12,138}{1-0.4024} = 20,313 \text{ lb/hr} \tag{.2}$$

$$R = 0.4024 \times 20,313 = 8,174 \text{ lb ethane/hr} \tag{.3}$$

Sensible Heat Calculations

$$Q_{\text{sensible}} = \dot{m}_{steam}(h_{COT,steam} - h_{XOT,steam}) + \dot{m}_{gas}\bar{C}_p(T_{COT} - T_{XOT}) \tag{.4}$$

$$\begin{aligned} Q_{\text{sensible}} = 56,004\ \frac{\text{lb}}{\text{hr}}&(1610.56\ \frac{\text{BTU}}{\text{lb}} - 1448.02\ \frac{\text{BTU}}{\text{lb}}) \\ &+ 140,010\ \frac{\text{lb}}{\text{hr}} \times 0.91043\frac{\text{BTU}}{\text{lb}^\circ\text{F}}(1550^\circ\text{F} - 1180^\circ\text{F}) = 56.3 \text{ MM BTU/hr} \end{aligned} \tag{.5}$$

Process Duty Calculations

$$Q_{\text{Process Duty}} = Q_{\text{cracking}} + Q_{\text{sensible}} \tag{.6}$$

$$Q_{\text{Process Duty}} = 146.8 \text{ MM BTU/hr} + 56.3 \text{ MM BTU/hr} = 203.0 \text{ MM BTU/hr} \tag{.7}$$

Flue Gas Requirement Calculations

$$\dot{m}_{\text{flue gas}} = \frac{Q_{\text{radiation}}}{h_{\text{in,flue gas}} - h_{\text{firebox, flue gas}}} \tag{.8}$$

$$\dot{m}_{\text{flue gas}} = \frac{206.1 \text{ MM BTU/hr}}{1045.4 \text{ BTU/lb} - 616.1 \text{ BTU/lb}} = 480,030 \text{ lb/hr} \tag{.9}$$

$h_{in,flue\ gas}$ Calculations

$$h_{\text{in,flue gas}} = \frac{\text{Fuel LHV}}{\text{lb Flue Gas/lb Methane}} \tag{.10}$$

$$h_{\text{in,flue gas}} = \frac{21,720 \text{ BTU/lb Methane}}{20.66 \text{ lb Flue Gas/lb Methane}} = 1045.4 \text{ BTU/lb Flue Gas} \tag{.11}$$

Heat Exchanger Heat Duty Calculations

$$Q_{\text{Heat Exchangers}} = Q_{\text{cracked gas}} + Q_{\text{steam}} \tag{.12}$$

$$Q_{\text{Heat Exchangers}} = 106.0 \text{ MM BTU/hr} + 27.5 \text{ MM BTU/hr} = 133.4 \text{ MM BTU/hr} \tag{.13}$$

Heat Available Calculations

$$Q_{\text{available}} = 0.9995\dot{m}_{\text{flue gas}}(h_{\text{firebox, flue gas}} - h_{\text{out stack, flue gas}}) \tag{.14}$$

$$Q_{\text{available}} = (0.9995)480,030\text{lb/hr}(616.1 \text{ BTU/lb} - 54.7 \text{ BTU/lb}) = 269.5 \text{ MM BTU/hr} \tag{.15}$$

Pre-Heat Energy Calculations

$$Q_{\text{pre-heat}} = \dot{m}_{steam}(h_{\text{XOT,steam}} - h_{475°\text{F},steam}) + \dot{m}_{hydrocarbon}\bar{C}_p(T_{XOT} - 60°\text{F}) \tag{.16}$$

$$\begin{aligned} Q_{\text{pre-heat}} = {} & 56,004 \text{ lb/hr}(1573.33 \text{ BTU/lb} - 1263.00 \text{ BTU/lb}) \\ & + 140,010 \text{ lb/hr}0.68\frac{\text{BTU}}{\text{lb °F}}(1180°\text{F} - 60°\text{F}) = 123.4 \text{ MM BTU/hr} \end{aligned} \tag{.17}$$

Economizer and Superheated Stream Calculations

$$Q_{\text{superheated and economizer}} = Q_{\text{available}} - Q_{\text{pre-heat}} \tag{.18}$$

$$Q_{\text{super and econ}} = 269.5 \text{ MM BTU/hr} - 123.4 \text{ MM BTU/hr} = 146.2 \text{ MM BTU/hr} \tag{.19}$$

$$Q_{steam} = Q_{\text{superheated and economizer}} + Q_{\text{Heat Exchangers}} \tag{.20}$$

$$Q_{steam} = 146.2 \text{ MM BTU/hr} + 133.4 \text{ MM BTU/hr} = 279.6 \text{ MM BTU/hr} \tag{.21}$$

$$\dot{m}_{steam} = \frac{Q_{steam}}{h_{superheated} - h_{BFW}} \tag{.22}$$

$$\dot{m}_{steam} = \frac{279.6 \text{ MM BTU/hr}}{1461.315\text{BTU/lb} - 212.12 \text{ BTU/lb}} = 223,828 \text{ lb/hr} \tag{.23}$$

$$Q_{\text{superheated}} = \dot{m}_{superheated}(h_{superheated} - h_{saturated}) = 66.5 \text{ MM BTU/hr} \tag{.24}$$

$$Q_{\text{economizer}} = Q_{\text{superheated and economizer}} - Q_{\text{superheated}} = 79.7 \text{ MM BTU/hr} \tag{.25}$$

Radiation Section Design and Sizing Calculations

$$A_R = \frac{Q_R}{q} \tag{.26}$$

$$A_R = \frac{206.1 \text{ MM BTU/hr}}{20,000 \text{ BTU/(hr}\cdot\text{ft}^2)} = 20,304.6 \text{ ft}^2 \tag{.27}$$

Stack Design Calculations

$$L = \frac{Dr}{0.52\rho(1/T_a - 1/T_{ga})} = 262 \text{ ft} \tag{.28}$$

$$d_o = (\frac{16\dot{m}_{flue}T_{ga}}{211,000\pi^2 f})^{0.2} = 7.8 \text{ ft} \tag{.29}$$

$$g = \frac{\dot{m}_{flue}}{\pi d_o^2/4} = 2.76 \ \frac{\text{lb}}{\text{s} \cdot \text{ft}^2} \tag{.30}$$

$$h_c = \frac{2.14g^{0.6}T_{ga}^{0.28}}{d_o^{0.4}} = 19.7 \text{ BTU/lb°F} \tag{.31}$$

$$h_{rg} = 0.0025T_g - 0.5 = 2.5 \text{ BTU/lb°F} \tag{.32}$$

$$h_o = 1.1(h_c + h_{rg}) = 22.2 \text{ BTU/lb°F} \tag{.33}$$

$$U = \frac{1}{R_i + R_o + R_w} = 7.2 \text{ BTU/hr°Fft}^2 \tag{.34}$$

$$A_c = \frac{Q_c}{\Delta T_{LM}U} = 39,600 \text{ ft}^2 \tag{.35}$$

Quench Tower Energy Balance Calculations

$$Q_{sensible} = \dot{m}_{gas}\bar{C}_p(650°\text{F} - 100°\text{F}) + \dot{m}_{steam}\bar{C}_p(650°\text{F} - 100°\text{F}) \tag{.36}$$

$$\begin{aligned} Q_{sensible} = {} & (56,004 \text{ lb/hr})(0.5486\frac{\text{BTU}}{\text{lb°F}}(650°\text{F} - 100°\text{F}) \\ & + (140,010 \text{ lb/hr})(0.5486\frac{\text{BTU}}{\text{lb°F}}(650°\text{F} - 100°\text{F}) = 58.6 \text{ MM BTU/hr} \end{aligned} \tag{.37}$$

$$Q_{condensation} = 56,004 \text{ lb/hr} \times 970.1 \text{ BTU/lb} - 21,283 \text{ lb/hr} \times 970.1 \text{ BTU/lb} = 33.7 \text{ BTU/hr} \quad (.38)$$

$$Q_{total\ cooling} = 58.6 \text{ MM BTU/hr} + 33.7 \text{ BTU/hr} = 92.3 \text{ BTU/hr} \quad (.39)$$

$$Q_{total\ cooling} = \dot{m}_{coolant}\bar{C}_p\Delta T \quad (.40)$$

$$\dot{m}_{coolant} = \frac{Q_{total\ cooling}}{\bar{C}_p\Delta T} \quad (.41)$$

$$\dot{m}_{coolant} = \frac{92.3 \text{ BTU/hr}}{1.0\ \frac{\text{BTU}}{\text{lb°F}} \times 30\text{°F}} = 3.1 \text{ MM lb/hr} \quad (.42)$$

Quench Tower Design and Sizing Calculations

$$v_{gas} = \frac{\dot{m}_{gas} + \dot{m}_{steam}}{\rho_g} = 85.8 \text{ ft}^3/s \quad (.43)$$

$$v_{window} = \frac{v_{gas}}{A_w} = 17.8 \text{ ft/s} \quad (.44)$$

$$v_{window,max} = 0.58\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 27.7 \text{ ft/s} \quad (.45)$$

$$v_{curtain} = \frac{\dot{m}_{coolant} + \dot{m}_{water}}{\rho_{water}} = 13.8 \text{ ft/s} \quad (.46)$$

$$v_{curtain,maximum} = 1.15\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 54.8 \text{ ft/s} \quad (.47)$$

$$U_a = 0.026G^{0.7}L^{0.4} \quad (.48)$$

$$G = \frac{\dot{m}_{gas} + \dot{m}_{steam}}{A_t} = 20,374.7\ \frac{\text{lb}}{\text{hr ft}^2} \quad (.49)$$

$$G = \frac{\dot{m}_{coolant} + \dot{m}_{water}}{A_t} = 323,374.2\ \frac{\text{lb}}{\text{hr ft}^2} \quad (.50)$$

$$U_a = 0.026G^{0.7}L^{0.4} = 4,317.4\ \frac{\text{BTU}}{\text{lb°F}} \quad (.51)$$

$$U_a = \frac{1}{\frac{1}{h_L a} + \frac{Q_{sensible}}{Q_{total\ cooling}}\frac{1}{\alpha H_a}} \quad (.52)$$

$$U = \frac{1}{\frac{Q_{sensible}}{Q_{total\ cooling}} \frac{1}{H_a}} = 6,800\ \frac{\text{BTU}}{\text{lb°F}} \tag{.53}$$

$$V_T = \frac{Q_{total\ cooling}}{U \Delta T_{LM}} = 76.4\ \text{ft}^3 \tag{.54}$$

$$Z_T = \frac{V_T}{A_t} = \frac{76.4\ \text{ft}^3}{9.62\ \text{ft}^2} = 7.9\ \text{ft} \tag{.55}$$

$$N = \frac{Z_t}{1.5} = 6\ \text{trays} \tag{.56}$$

$$D = \sqrt{4A_t/\pi} = 3.5\ \text{ft} \tag{.57}$$

$$t = \frac{50\ \text{psia} \times 21\ \text{in}}{2(15,000\ \text{psi}) \times 0.85 - 0.6(50\ \text{psia})} + 0.125\ \text{in} = .17\ \text{in} \tag{.58}$$

Appendix D: Microsoft Excel Spreadsheet Calculations

See the following pages for the Microsoft Excel spreadsheet calculations.