

Development of Process Simulation model of Gas Processing Plant



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Certification

This is to certify that **Osama Masood, 243968, M. Aalishan Tariq, 267908,** and **Syed Shahrukh Haider, 260444** have successfully completed the final project **Development of Process Simulation model of Gas Processing Plant**, at the **National University of Sciences and Technology**, to fulfill the partial requirement of the degree **BE Chemical Engineering**.

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Abstract

The objective of our project is the development of a natural gas processing plant able to process feed from four different wells of various known compositions. Natural gas when excavated is a mixture of gases, oils, water, and solids which need to be processed in order to obtain a refined product that can be further used. Natural gas is not only a great source of low sulfur fuel but is also the feedstock of various businesses for example Petro-chemical and fertilizers industries all use Natural gas to obtain Nitrogen, Hydrogen, Carbon monoxide and ammonia gas. Due to this dependency on Natural gas, it is essential to develop an effective and efficient processing plant that not only extracts a large proportion of gas from the well but also delivers other components as useable by-products. Our goal here is to produce a model that is able to output more than 90% of the total methane present in the feedstock along with producing LPG (C4+C5) and propane recovery (95%) as by-products in an economical manner.

Undertaking

I certify that the project **Development of Process Simulation model of Gas Processing Plant** is our own work. The work has not, in whole or in part, been presented elsewhere for assessment. Where material has been used from other sources it has been properly acknowledged/ referred.

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List of Acronyms

- ❖ C_p Specific heat at constant pressure
- ❖ C_v Specific heat at constant volume
- ❖ P Pressure
- ❖ P Power
- ❖ T Temperature
- ❖ v Volumetric flowrate
- ❖ n number of moles
- ❖ R Universal gas constant
- ❖ ρ Mass density
- ❖ z Gas compressibility factor
- ❖ M Relative molecular mass
- ❖ H Enthalpy
- ❖ H Head
- ❖ Q Heat flow
- ❖ m mass flowrate
- ❖ Δ change in
- ❖ d diameter
- ❖ A Area
- ❖ π pi
- ❖ μ Viscosity
- ❖ k Thermal conductivity
- ❖ l Length
- ❖ U Overall heat transfer coefficient
- ❖ h Heat transfer coefficient
- ❖ G Mass velocity
- ❖ u Velocity
- ❖ g gravitational constant
- ❖ γ Specific gravity
- ❖ u Velocity
- ❖ t time
- ❖ L Flowrate of liquid

- ❖ G Flowrate of gas
- ❖ T_c Critical Temperature
- ❖ P_c Critical Pressure

Chapter 1

1.1 Introduction

Natural gas is the most energy efficient fossil fuel; it offers important energy saving benefits when it is used instead of oil or coal. Although the primary use of natural gas is as a fuel, it is also a source of hydrocarbons for petrochemical feedstocks and a major source of elemental sulfur, an important industrial chemical. Its popularity as an energy source is expected to grow substantially in the future because natural gas can help achieve two important energy goals for the 21st century, providing the sustainable energy supplies and services needed for social and economic development and reducing adverse impacts on global climate and the environment in general. Natural gas consumption and trade have been growing steadily over the past two decades and natural gas has strengthened its position in the world energy mix. Although the demand for natural gas declined in 2009, as a result of the economic slowdown, it is expected to resume growth in both emerging and traditional markets in the coming decades. Such increase in the near future will be driven because of additional demand in current uses, primarily power generation. There is yet a little overlap between the use of natural gas and oil in all large markets. However, there are certain moves in the horizon, including the electrifying of transportation, which will push natural gas use to ever higher levels.

1.2 Natural Gas History

Although natural gas has been known since ancient times, its usage in the commercial sectors is relatively new. In about 1000, B.C., the famous Oracle at Delphi in ancient Greece, was built where natural gas seeped from the ground in a flame. Around 500 B.C., the Chinese started using bamboo pipes or “pipelines” to transport gas that seeped to the surface and to use it to boil sea water to get drinkable water. The first commercialized natural gas occurred in Britain. Around 1785, the British used natural gas produced from coal to light houses and streets. In 1816, Baltimore, Maryland used this type of manufactured natural gas to become the first city in the United States to light its streets with gas. In the United States, the properties of natural gas were discovered by Native Americans, who would ignite the gases that seeped into and around Lake Erie. French

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explorers witnessed this practice around 1626. In 1821, William Hart dug the first successful natural gas well in the U.S. in Fredonia, New York. Eventually, the Fredonia Gas Light Company was formed, becoming the first American natural gas distribution company. In 1836, the City of Philadelphia created the first municipally owned natural gas distribution company. Today, U.S. public gas systems number more than 900, and the Philadelphia Gas Works is the largest and longest operating public gas system in the U.S. During most of the 19th century, natural gas was used almost exclusively as a source of light, but in 1885, Robert Bunsen's invention of what is now known as the Bunsen burner opened vast new opportunities to use natural gas. Once effective pipelines began to be built in the 20th century, the use of natural gas expanded to home heating and cooking, appliances such as water heaters and oven ranges, manufacturing and processing plants, and boilers to generate electricity.

1.3 Natural Gas Origin

There are different theories as to the origins of fossil fuels. The most widely accepted theory of the origin of natural gas states that it came from the remains of land and aquatic plants, animals, and microorganisms that was trapped within sediments as they were deposited and transformed over long periods of time into their present form. Two main mechanisms, namely biogenic and thermogenic, are responsible for the degradation of the original fossil organic material in sediments. Biogenic gas is formed at shallow depths and low temperatures due to the action of bacteria on the organic debris accumulating in the sediments. In contrast, thermogenic gas is formed at greater depths by degradation of organic matter, called kerogen, accumulated in fine-grained sediments, especially clays and shales. This degradation occurs through the combined effects of temperature and pressure. Thermogenic gas is believed to be produced through two mechanisms, namely, direct thermal cracking of sedimentary organic matter and secondary thermal cracking of oil that is formed in the first stage. The former is called the primary thermogenic gas that coexists with oil, while the latter is called secondary thermogenic gas that coexists with an insoluble solid matter, called pyro bitumen. Both mechanisms involve thermal cracking with some degree of sustained pressure, mainly through the weight of the sedimentary formation. Little information is available on the time required to generate thermogenic gas other than

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the general belief that it is a very long time. Hydrocarbons are generated in “source rock” and often migrate to subsurface formations called “Reservoir rock.” The quality of a reservoir rock is determined by its two principal properties: porosity and permeability. Porosity is the void space between the grains, and indicates the rock’s capacity to contain liquid or gaseous hydrocarbons. Permeability is the measurement of the rock’s ability to transmit the oil or gas.

1.4 Types of Natural Gas

The natural gases can be classified as:

- Conventional natural gas
- Un-conventional natural gas

Conventional gas is typically “free gas” trapped in multiple porous zones in naturally occurring rock formations such as carbonates, sandstones, and siltstones. Conventional natural gas generally occurs in deep reservoirs, either associated with crude oil (associated gas) or in reservoirs that contain little or no crude oil (non-associated gas).

Conventional gas is typically found in medium to highly porous reservoirs with a permeability greater than 1 millidarcy. Pressure moves the gas towards the production wells. As such, it can be extracted via traditional techniques that are relatively easy and inexpensive.

The second classification of Natural gas is the un-conventional natural gas. This is the natural gas is that is held in compact formations and is not readily available to exploit rather it demands the use of expensive and modern methods.

This category has further sub branches which include:

- Gas in tight sand/ tight shale
- Coal-bed methane
- Gas in geo-pressurized reservoirs
- Gas hydrates.

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1.5 Properties of Natural Gas

Properties	Value
Relative molar mass	17-20
Carbon content, weight%	73.3
Hydrogen content, weight%	23.9
Oxygen content, weight%	0.4
Hydrogen/carbon atomic ratio	3.0-4.0
Relative density, 15C	0.72-0.81
Boiling point, C	-162
Auto ignition temperature, C	540.560
Octane number	120-130
Methane number	69-99
Stoichiometric air/fuel ratio, weight	19.2
Vapor flammability limits, volume %	5-15
Flammability limits	0.7-2.1
Lower heating/calorific value, MJ/kg	38-50
Stoichiometric lower heating value, MJ/kg	2.75
Methane concentration, volume %	80-99
Ethane concentration, volume %	2.7-4.6
Nitrogen concentration, volume %	0.1-15
Carbon dioxide concentration, volume %	1-5
Sulfur concentration, weight% ppm	<5
Specific CO ₂ formation, g/MJ	38-50

Table 1 Properties of Natural gas

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1.6 Composition of Natural Gas

Compound	Symbol	Percentage (%)
Methane	CH ₄	85-98
Ethane	C ₂ H ₆	1.5-9.0
Propane	C ₃ H ₈	0.1-1.5
Butane	C ₄ H ₁₀	0.1-0.5
Carbon Dioxide	CO ₂	0.05-2.0
Nitrogen	N ₂	0.2-5.0
Hydrogen Sulfide	H ₂ S	0.01-0.05
Rare Gases	A, He	0.1-2.0

Table 2 Composition of natural gas

1.7 Why Process Natural Gas:

Natural gas when excavated is not in a refined condition, it consists of a mixture of gas, oil, water and solid residues. Some of these constituents can be processed to extract valuable products which can be commercially used while others are just contaminants which have to be removed in order to make the products clean, environmentally friendly and more efficient in delivering their total value.

These processing plants are set up in such a manner that they not only separate the undesired components from the Natural gas but are equipped with equipment's that produce useable by products for commercial use.

1.8 Steps of Natural Gas Processing

In general, a natural gas processing plant must be equipped with primary inlet separators in order to separate the crude flowing from the well into gas and condensate/water. The removal of Acid gas (H₂S and CO₂) is the second step in the processing plant to eliminate harmful and corrosive gases from the gas. Dehydration/moisture removal units then follow to removal water content from the mix. The addition of Mercury and Nitrogen Removal for further purification of the final dry gas. Finally, NGL recovery units like De-ethanizer, De-propanizer and De-butanizers are installed to recover LPG and NGLS from the remainder.

Chapter 2

LITERATURE REVIEW

2.1 Types of Flow Regimes

2.1.1 Dispersed Bubble Flow

At higher liquid flow rates and for a wide range of gas flow rates, small gas bubbles are dispersed throughout the continuous liquid phase. Due to the effect of buoyancy these bubbles tend to accumulate in the upper part of the pipe.

2.1.2 Plug (Elongated Bubble) Flow

At relatively low gas flow rates, as the liquid flow rate is reduced, the smaller bubbles of dispersed bubble flow coalesce to form larger bullet shaped bubbles that move along the top of the pipe.

2.1.3 Stratified (Smooth and Wavy) Flow

At low liquid and gas flow rates gravitational effects cause total separation of the two phases. These results in the liquid flowing along the bottom of the pipe and the gas flowing along the top, where the gas liquid surface is smooth. As the gas velocity is increased in a stratified smooth flow the interfacial shear forces increase, rippling the liquid surface and producing a wavy interface.

2.1.4 Slug Flow

As the gas and liquid flow rates are increased further, the stratified liquid level grows and becomes progressively wavier until eventually the whole cross-section of the pipe is blocked by a wave. The resultant "piston" of liquid is then accelerated by the gas flow, surging along the pipe and scooping up the liquid film in front as it progresses. This "piston" is followed by a region containing an elongated gas bubble moving over a thin liquid film. Hence an intermittent regime develops in which elongated gas bubbles and liquid slugs alternately

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surge along the pipe. The major difference between an elongated bubble flow and a slug flow is that in an elongated bubble flow there are no entrained gas bubbles in the liquid slugs.

2.1.5 Annular Flow

When gas flow rates increase, annular (also referred to as annular-mist) flow occurs. During annular flow, the liquid phase flows largely as an annular film on the wall with gas flowing as a central core.

Some of the liquid is entrained as droplets in this gas core. The annular liquid film is thicker at the bottom than at the top of the pipe because of the effect of gravity and, except at very low liquid rates, the liquid film is covered with large waves.

2.1.6 Churn (Transition) Flow

If a change from a continuous liquid phase to a continuous gas-phase occurs, the continuity of the liquid in the slug between successive Taylor bubbles is repeatedly destroyed by a high local gas concentration in the slug. This oscillatory flow of the liquid is typical of churn (froth) flow. It may not occur in small diameter pipes. The gas bubbles may join and liquid may be entrained in the bubbles. In this flow regime, the falling film of the liquid surrounding the gas plugs cannot be observed.

2.2 Natural Gas Properties

2.2.1 Gas Specific Gravity

Specific gravity of a natural gas is defined as the ratio of gas density to the density of air, both defined at the same pressure and temperature. These densities are usually defined at standard conditions (14.7 psia and 60F). Therefore, specific gravity of gas is defined as:

$$\gamma_g = \frac{M}{M_{air}}$$

Where M is the molecular weight of gas and M_{air} is the molecular weight of air that is equal to 29. Once we can calculate the value of molecular weight of a mixture, we can calculate the specific gravity of the mixture. For a gas mixture, we can calculate the molecular weight as:

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$$M = \sum_{i=1}^n y_i M_i$$

Where M_i is the molecular weight of component i , y_i is the mole fraction of component i , and n is the total number of components.

2.2.2 Gas Compressibility Factor

The volume of a real gas is usually less than what the volume of an ideal gas would be, and hence a real gas is said to be super-compressible. The ratio of the real volume to the ideal volume, which is a measure of the amount the gas deviates from perfect behavior, is called the super compressibility factor, sometimes shortened to the compressibility factor. It is also called the gas deviation factor, and given the symbol “ Z ”. The gas deviation factor is by definition the ratio of the volume actually occupied by a gas at a given pressure and temperature to the volume it would occupy if it behaved ideally. The real gas equation of state is then written as:

$$PV = ZnRT$$

The reduced pressure and reduced temperature are defined as:

$$P_r = \frac{P}{P_c} \quad \text{and} \quad T_r = \frac{T}{T_c}$$

Where P_r and T_r are reduced pressure and reduced temperature, respectively; and P_c and T_c are critical pressure and critical temperature of the gas, respectively. The values of critical pressure and critical temperature can be estimated from the following equations if the composition of the gas and the critical properties of the individual components are known.

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$$P_c = \sum_i^n P_{Ci} y_i \quad \text{and} \quad T_c = \sum_i^n T_{Ci} y_i$$

where P_{ci} and T_{ci} are the critical pressure and critical temperature of component i , respectively; and y_i is the mole fraction of component i .

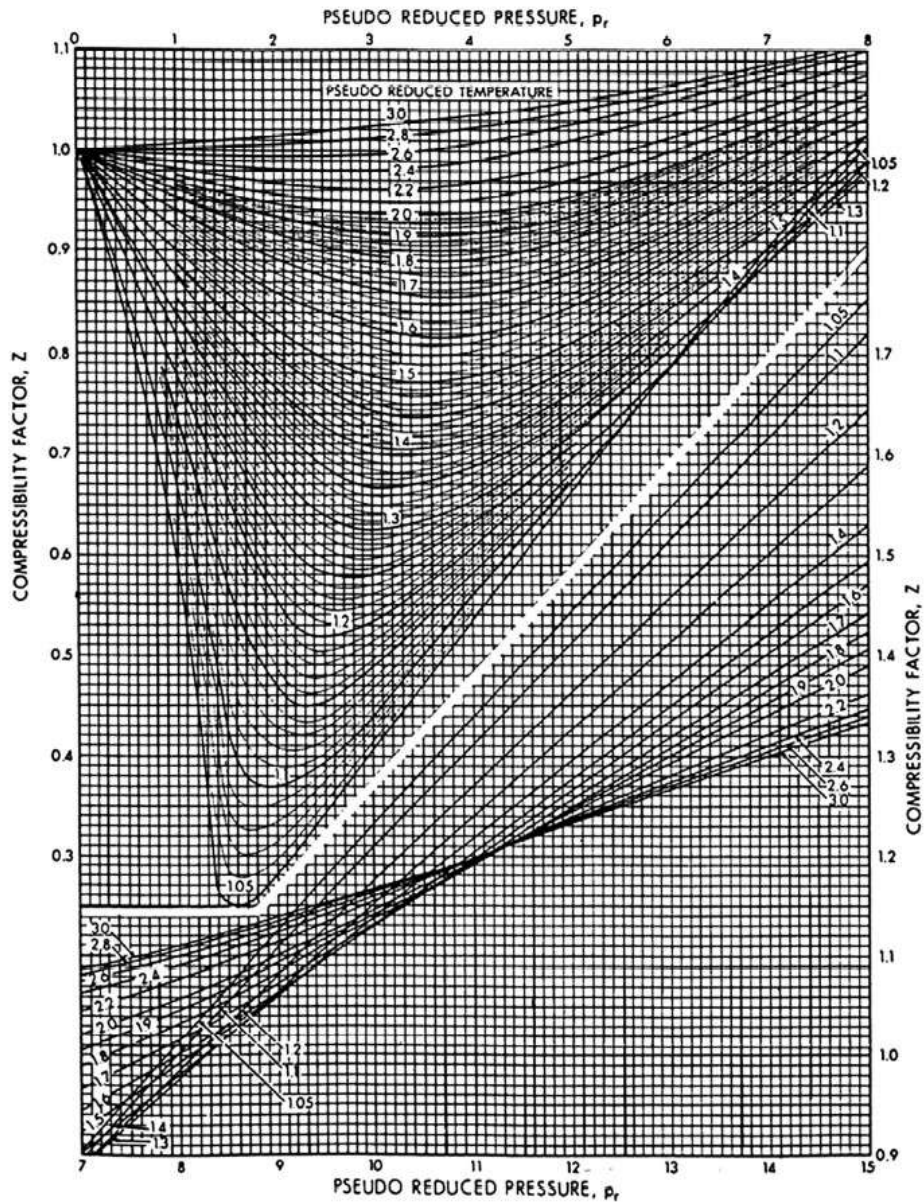


Figure 1 Compressibility factor vs Pseudo Reduced Pressure

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2.2.3 Gas Density

The gas density is defined as mass per unit volume and so can also be derived and calculated from the real gas law:

$$\rho_g = \frac{m}{V} = \frac{PM}{ZRT}$$

Knowing that molecular weight of gas is the product of specific gravity and molecular weight of air, and the value of R is 10.73 in field units [8.314 in SI units], we can write the equation for density as:

$$\rho_g = 2.7 \frac{P \gamma_g}{ZT}$$

where ρ_g is in lbm/ft³, P is in psia, and T is in R. Alternately,

$$\rho_g = 3.49 \frac{P \gamma_g}{ZT}$$

where ρ_g is in kg/m³, P is in kPa, and T is in K.

2.2.4 Specific Heat, Cp

Specific heat is defined as the amount of heat required to raise the temperature of a unit mass of a substance through unity. It is an intensive property of a substance. It can be measured at constant pressure (Cp), or at constant volume (Cv), resulting in two distinct specific heat values. In terms of basic thermodynamic quantities:

$$C_p = \left(\frac{\partial H}{\partial T} \right)_p$$

$$C_v = \left(\frac{\partial U}{\partial T} \right)_v$$

where H is the molal enthalpy (BTU/lbmole), U is the molal internal energy (Btu/lbmole), Cp is the molal specific heat at constant pressure (BTU/lbmole-o R), and Cv is the molal specific heat at constant volume (Btu/lbmole-o R). Both heat capacities are thermodynamically related. It can be proven that this relationship is controlled by the P_eV_eT behavior of the substances through the following relationship:

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$$C_p - C_v = T \left(\frac{\partial V}{\partial T} \right)_p \left(\frac{\partial P}{\partial T} \right)_v$$

For ideal gases:

$$C_p - C_v = R$$

where R is the universal gas constant.

2.3 Natural Gas Processing Plant

Processing consists of separating all of the various hydrocarbons and impurities from the raw natural gas to produce what is termed “pipeline quality” dry natural gas. Pipeline companies impose restrictions on the makeup of natural gas allowed into the pipeline. Pipeline companies require the raw natural gas to be purified by removing “waste products” and associated hydrocarbons such as ethane, propane, butane, i-butane, and pentane plus (natural gasoline).

Associated hydrocarbons termed “natural gas liquids” (NGLs) can be a valuable byproduct of natural gas processing. They can be sold separately and have a variety of different uses, such as enhancing oil recovery in oil wells, providing raw materials for oil refineries and/or petrochemical plants, and as sources of energy.

Some of the needed processing can be accomplished at or near the wellhead (field processing). Complete processing of natural gas takes place at a processing plant, usually located in a natural gas producing region. Extracted natural gas is transported to processing plants through a network of gathering pipelines. A complex gathering system may consist of thousands of miles of pipeline, interconnecting the processing plant to more than 100 wells. In addition to the processing done at the wellhead and centralized processing plants, some final processing is also done at “extraction plants.” These plants are located on major pipeline systems and remove small quantities of NGLs that may still exist in pipeline-quality gas.

Natural gas processing to meet pipeline-quality gas standards involves the following four main steps to remove the various impurities:

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- Oil and condensate removal
- Water Removal
- Separation of NGLs
- Sulfur and carbon dioxide removal

In addition to the four processes above, heaters and scrubbers are installed, usually at or near the wellhead. Scrubbers remove sand and other large-particle impurities. Heaters ensure that the temperature of the gas does not drop below the hydrate formation temperature.

When the gas temperature drops there is a tendency for “hydrates” to form. Hydrates are solid or semi-solid compounds, resembling ice-like crystals. Should these hydrates accumulate, they can impede passage of natural gas through valves and gathering systems, plug process instrumentation, and reduce the capacity of processing vessels. To reduce the occurrence of hydrates, the following equipment may be used:

- Indirect-fired heater
- Hydrate inhibitors
- Dehydration
- Low-temperature units

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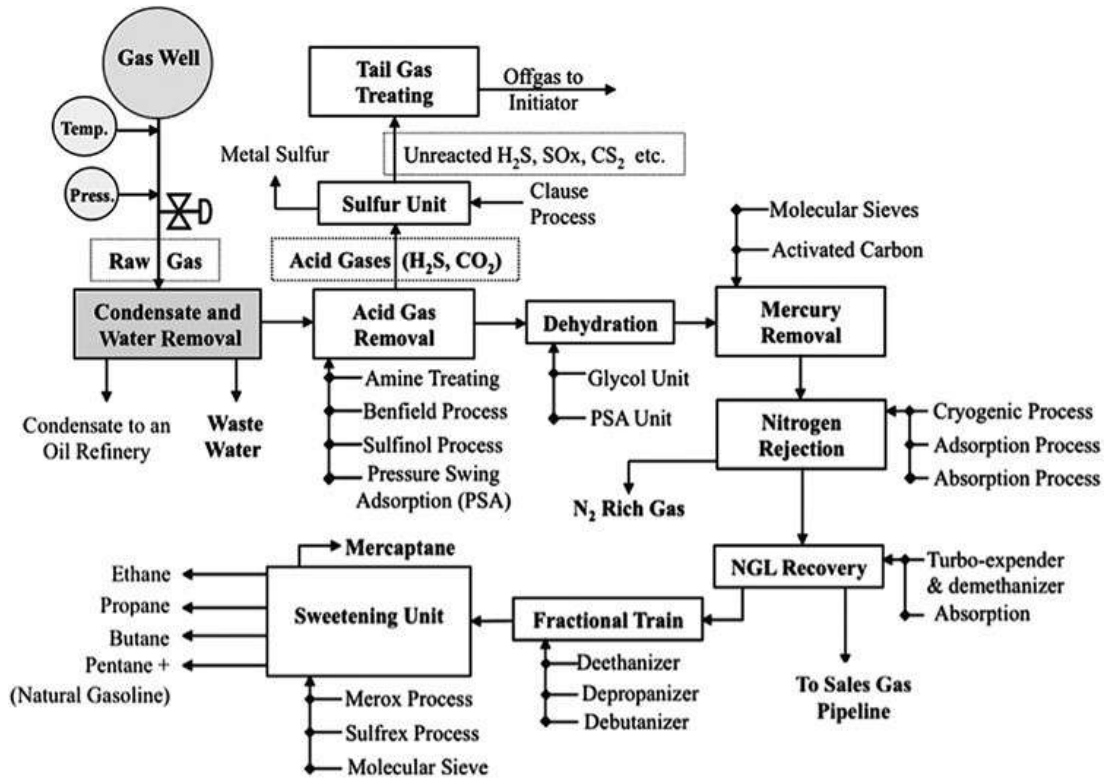


Figure 2 Natural Gas processing plant

Chapter 3

PROCESS DESCRIPTION

3.1 Process Flow Diagram

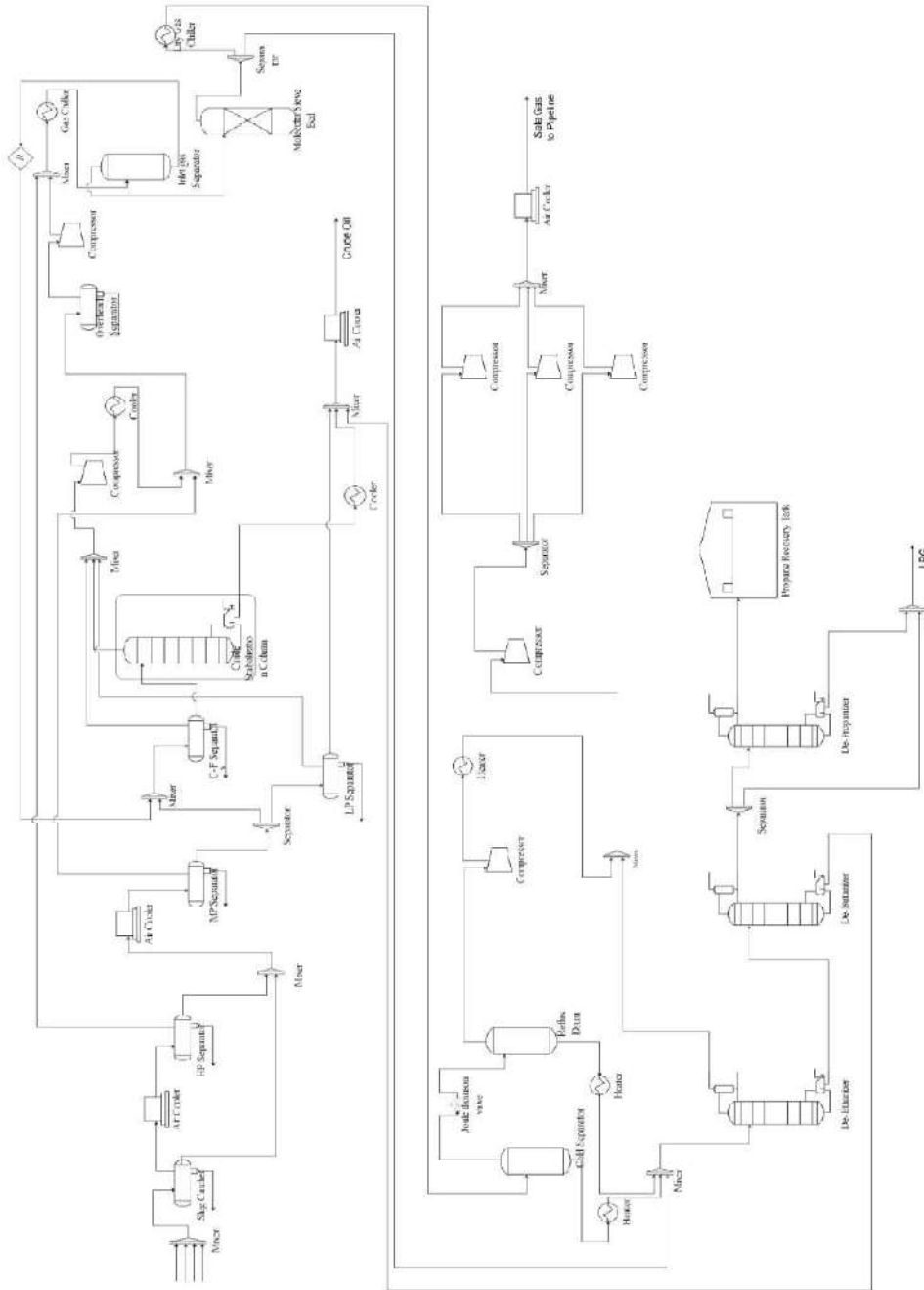


Figure 3 Process Flow Diagram

Chapter 3

3.2 Process Description

Natural Gas is one of the most essential products in use in modern times. It is a mixture of methane and ethane mostly. One such process starts when streams from four wells, namely X-1, X-2, X-3 and X-4, combine and enter a slug catcher. All four wells have different stream compositions, flow rates, and other parameters such as Temperature and Pressure. Furthermore, the streams coming directly from wells vary a lot in terms of all these parameters.

Firstly, A Slug Catcher is used to collect liquids that have settled in flow lines which can overload the gas/liquid handling capacity of the plant especially during pigging operations.

The Slug Catcher helps protect your valuable production equipment from abrupt large bursts of liquids entering your gas stream. It is typically used downstream of pipeline and pigging operations where large volumes of liquid can collect in the pipeline and are pushed through in a short time. The unit has a liquid level controller and liquid outlet control valve, as well as a crash dump connection. This connection allows you the option of adding a crash level controller and secondary liquid outlet to dispose of sudden large slugs.

The top stream from the slug catcher, comprising mostly of gases of lighter components is cooled and enters a three-phase separator at high pressure. The working principle of the three-phase separator is based on the density difference of water/oil/gas. The gas-liquid separation is achieved by centrifugal force and gravity. In simple terms, in a three-phase separator, the separation is performed in the order of gas/liquid separation and then oil/water separation.

The bottom stream from the HP separator and the slug catcher, comprising mostly of heavier liquid components, are mixed and enter a Medium Pressure Separator. The produced water is removed at each three-phase separator as well. The stream is then split into two, and enters two different separators. This allows some of the liquid to be bypassed, reducing the duty on the Crude Stabilization Column (CSC). Similarly, top streams from the separators are bypassed to reduce the duty on the CSC, by reducing the total flow rate.

Chapter 3

The bottom stream of the CSC is cooled and is sent to the crude refinery for further processing, which is not relevant to Natural Gas Processing. The top stream is compressed and cooled in multiple stages, and the bypassed top streams from three previous separators enter step by step. Finally, the gas is chilled using an Ammonia Refrigeration System, and undergoes separation of still allow some of the heavy gas components to be recycled and removed.

Next, a Molecular Sieve Bed (MSB) is used to remove water from the gas to below acceptable levels. The characteristics of these molecular sieves include fast adsorption speed, frequent regeneration ability, good crushing resistance and pollution resistance. These features can improve both the efficiency and lifetime of the sieve.

The gas is then split into two, and only one stream is cooled. This reduces the duty required by the cooler, without disturbing the product quality and specifications in any way. The stream is then subjected to expansion via a Joule-Thompson valve, and a reflux drum recycles it to the fractionation train i.e. a series of distillation columns which is then used to obtain the final products.

JT valve works on the principle that when the pressure of a gas/Liquid changes, its temperature also changes.

The Joule–Thomson effect describes the temperature change of a gas or liquid when it is forced through a valve, while kept insulated. (Insulation is required to avoid influence of surrounding environment effects).

First is the de-ethanizer, which removes Natural Gas from the top. The gas is subjected to compression via both reciprocating and centrifugal compressors, and is cooled. Reciprocating compressors have poor mechanical efficiency due to large sliding parts. Centrifugal compressors have better mechanical efficiency due to absence of sliding parts. Installation cost for setting up reciprocating compressors is higher. Installation cost for setting up centrifugal compressors is lower. Also Reciprocating compressors are used where high pressure is needed while centrifugal are used where high velocity is required

Chapter 3

The bottoms of the de-ethanizer enter the de-butanizer. Butanes, propane, and some left over ethane is separated, while fractions heavier than C5 leave from the bottom to be used in the oil refinery. The top stream is split into two. One enters the de-propanizer, while the other bypasses the column to allow for production of LPG of the desired composition. Hence, the other two products i.e. propane and LPG are obtained.

Chapter 4

MATERIAL BALANCE

4.1 Well Feed Composition

Name	WELL X-1	WELL X-2	WELL X-3	WELL X-4
CO2	1.47E-02	4.20E-03	1.45E-02	1.47E-02
Nitrogen	8.00E-03	6.60E-03	6.00E-03	8.00E-03
Methane	0.6742	0.726	0.6721	0.6723
Ethane	6.41E-02	6.58E-02	6.34E-02	6.40E-02
Propane	3.62E-02	3.48E-02	3.50E-02	3.62E-02
i-Butane	7.90E-03	7.00E-03	8.00E-03	7.90E-03
n-Butane	1.47E-02	1.27E-02	1.50E-02	1.47E-02
i-Pentane	6.50E-03	5.30E-03	6.80E-03	6.60E-03
n-Pentane	6.50E-03	5.40E-03	7.00E-03	6.70E-03
n-Hexane	9.70E-03	7.40E-03	9.80E-03	1.01E-02
n-Heptane	1.47E-02	1.19E-02	1.59E-02	1.56E-02
n-Octane	1.54E-02	1.29E-02	1.77E-02	1.65E-02
n-Nonane	1.00E-02	8.40E-03	1.20E-02	1.08E-02
n-Decane	7.80E-03	6.90E-03	9.20E-03	8.40E-03
n-C11	6.00E-03	5.20E-03	6.90E-03	6.50E-03
n-C12	4.80E-03	4.20E-03	5.60E-03	5.20E-03
n-C13	4.80E-03	4.30E-03	5.80E-03	5.20E-03
n-C14	3.90E-03	3.50E-03	4.70E-03	4.20E-03
n-C15	4.10E-03	3.50E-03	4.60E-03	4.50E-03
n-C16	3.10E-03	2.70E-03	3.60E-03	3.30E-03
n-C17	2.60E-03	2.20E-03	3.00E-03	2.80E-03
n-C18	2.40E-03	2.10E-03	2.90E-03	2.60E-03
n-C19	2.00E-03	1.80E-03	2.40E-03	2.20E-03
n-C20	1.70E-03	1.40E-03	2.00E-03	1.90E-03
n-C21	1.50E-03	1.30E-03	1.80E-03	1.60E-03
n-C22	1.30E-03	1.10E-03	1.60E-03	1.40E-03

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n-C23	1.10E-03	1.00E-03	1.30E-03	1.20E-03
n-C24	1.00E-03	9.00E-04	1.20E-03	1.10E-03
n-C25	9.00E-04	8.00E-04	1.10E-03	1.00E-03
n-C26	8.00E-04	7.00E-04	9.00E-04	9.00E-04
n-C27	8.00E-04	6.00E-04	9.00E-04	8.00E-04
n-C28	7.00E-04	6.00E-04	8.00E-04	8.00E-04
n-C29	6.00E-04	5.00E-04	7.00E-04	7.00E-04
n-C30	6.00E-04	5.00E-04	6.00E-04	6.00E-04
H2O	6.49E-02	4.58E-02	5.52E-02	5.90E-02
Mass Flow Rate (kg/hr)	34781.5	43209.2	40601.0	40120.2

Table 3 Well feed composition

4.2 Raw Feed Mixture

Name	Slug Catcher
CO2	1.15E-02
Nitrogen	7.11E-03
Methane	0.688577481
Ethane	6.44E-02
Propane	3.55E-02
i-Butane	7.66E-03
n-Butane	1.42E-02
i-Pentane	6.24E-03
n-Pentane	6.34E-03
n-Hexane	9.14E-03
n-Heptane	1.44E-02
n-Octane	1.55E-02
n-Nonane	1.02E-02
n-Decane	8.02E-03
n-C11	6.10E-03
n-C12	4.91E-03

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n-C13	4.99E-03
n-C14	4.05E-03
n-C15	4.14E-03
n-C16	3.15E-03
n-C17	2.63E-03
n-C18	2.48E-03
n-C19	2.09E-03
n-C20	1.73E-03
n-C21	1.54E-03
n-C22	1.34E-03
n-C23	1.14E-03
n-C24	1.04E-03
n-C25	9.43E-04
n-C26	8.19E-04
n-C27	7.65E-04
n-C28	7.19E-04
n-C29	6.19E-04
n-C30	5.70E-04
H2O	5.55E-02
Mass Flow Rate (kg/hr)	158711.9

Table 4 Raw fed mixture

4.3 Medium Pressure Separator

Name	MPS_FEED	MPS_TOP	MPS_BOT	PW-3
CO2	0.00705741	0.01747	0.004342	0.00011682
Nitrogen	0.00120902	0.00501	0.000217	1.14E-06
Methane	0.22968906	0.790356	0.083362	3.16E-07
Ethane	0.05554153	0.102587	0.04329	1.18E-09
Propane	0.05706741	0.048932	0.059235	6.21E-12
i-Butane	0.01801269	0.007787	0.020698	4.00E-15
n-Butane	0.03754051	0.012638	0.044075	1.03E-14

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i-Pentane	0.02160782	0.003443	0.02637	8.50E-18
n-Pentane	0.023403	0.002989	0.028754	8.29E-18
n-Hexane	0.04116089	0.00199	0.051426	1.27E-20
n-Heptane	0.07208077	0.001345	0.090617	1.28E-23
n-Octane	0.08195961	0.00059	0.103281	7.21E-27
n-Nonane	0.05534162	0.000159	0.069801	2.13E-30
n-Decane	0.04391232	5.11E-05	0.055405	2.62E-34
n-C11	0.03354465	1.54E-05	0.04233	6.72E-38
n-C12	0.02703424	5.53E-06	0.034117	7.21E-42
n-C13	0.02747221	2.05E-06	0.03467	2.28E-45
n-C14	0.0222833	5.78E-07	0.028122	2.07E-49
n-C15	0.02279706	2.96E-07	0.028771	1.21E-53
n-C16	0.0173476	9.30E-08	0.021893	4.83E-58
n-C17	0.01446229	4.09E-08	0.018252	4.45E-64
n-C18	0.01365622	1.92E-08	0.017235	1.22E-70
n-C19	0.01148459	8.81E-09	0.014494	6.43E-78
n-C20	0.00953747	1.84E-09	0.012037	8.66E-76
n-C21	0.00846457	7.51E-10	0.010683	3.64E-80
n-C22	0.00736329	3.29E-10	0.009293	6.73E-85
n-C23	0.00629294	1.04E-10	0.007942	4.15E-89
n-C24	0.0057423	4.01E-11	0.007247	1.41E-93
n-C25	0.00519166	1.70E-11	0.006552	2.09E-97
n-C26	0.00450888	5.98E-12	0.00569	1.74E-102
n-C27	0.00421114	1.90E-12	0.005315	5.10E-103
n-C28	0.00395824	1.03E-12	0.004995	4.79E-103
n-C29	0.0034076	4.74E-13	0.0043	4.13E-103
n-C30	0.00314078	5.65E-14	0.003964	3.80E-103
H2O	0.00251529	0.004627	0.001225	0.99988172
Mass Flow Rate (kg/hr)	86813.7	3515.9	83289.2	8.562325

Table 5 Medium Pressure Separator

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4.4 Crude Stabilization Column

Name	CFS_BOT	CSC_TOP	CSC_BOT	CSC_RFLX
CO2	0.002213	0.004166	4.99E-38	0.0040686
Nitrogen	3.54E-05	0.000175	2.12E-39	0.0001755
Methane	0.025392	0.079678	5.43E-42	0.0792688
Ethane	0.030175	0.059796	4.07E-30	0.058538
Propane	0.059538	0.126417	1.08E-20	0.1241219
i-Butane	0.024049	0.072693	3.59E-14	0.0722436
n-Butane	0.053553	0.199181	1.98E-11	0.1990054
i-Pentane	0.034961	0.224028	0.000169	0.2260022
n-Pentane	0.03771	0.215728	0.008554	0.2178152
n-Hexane	0.059703	0.005435	0.082775	0.0056981
n-Heptane	0.101881	0.005696	0.138983	0.0059212
n-Octane	0.111789	0.003533	0.151499	0.0036564
n-Nonane	0.073482	0.001195	0.099315	0.0012349
n-Decane	0.057484	0.000461	0.077601	0.000476
n-C11	0.043646	0.000168	0.058891	0.0001728
n-C12	0.035111	6.90E-05	0.047365	7.11E-05
n-C13	0.035654	3.15E-05	0.048093	3.24E-05
n-C14	0.028915	1.09E-05	0.039	1.12E-05
n-C15	0.029581	6.28E-06	0.039898	6.46E-06
n-C16	0.022509	2.37E-06	0.03036	2.44E-06
n-C17	0.018766	1.13E-06	0.02531	1.16E-06
n-C18	0.01772	5.90E-07	0.0239	6.07E-07
n-C19	0.014902	2.95E-07	0.020099	3.03E-07
n-C20	0.012375	8.76E-08	0.016691	9.01E-08
n-C21	0.010983	4.16E-08	0.014814	4.28E-08
n-C22	0.009554	2.07E-08	0.012886	2.13E-08
n-C23	0.008165	8.06E-09	0.011013	8.30E-09
n-C24	0.007451	3.71E-09	0.01005	3.82E-09

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n-C25	0.006736	1.83E-09	0.009086	1.89E-09
n-C26	0.00585	7.79E-10	0.007891	8.02E-10
n-C27	0.005464	3.14E-10	0.00737	3.24E-10
n-C28	0.005136	1.88E-10	0.006927	1.94E-10
n-C29	0.004422	9.81E-11	0.005964	1.01E-10
n-C30	0.004075	1.81E-11	0.005497	1.87E-11
H2O	0.001019	0.001529	7.67E-37	0.0014755
Mass Flow Rate (kg/hr)	42461.3	216794.3	38414.7	212747.7

Table 6 Crude stabilization column

4.5 Molecular Sieve Bed

Name	IGS_TOP	MSB_OUT	WATER
CO2	0.013776	0.013783	0
Nitrogen	0.008624	0.008628	0
Methane	0.833433	0.833833	0
Ethane	0.075139	0.075175	0
Propane	0.037364	0.037382	0
i-Butane	0.00718	0.007184	0
n-Butane	0.012778	0.012784	0
i-Pentane	0.004801	0.004804	0
n-Pentane	0.00392	0.003922	0
n-Hexane	0.001416	0.001417	0
n-Heptane	0.000807	0.000807	0
n-Octane	0.00024	0.000241	0
n-Nonane	3.58E-05	3.59E-05	0
n-Decane	5.47E-06	5.47E-06	0
n-C11	6.43E-07	6.43E-07	0
n-C12	9.78E-08	9.78E-08	0
n-C13	1.08E-08	1.08E-08	0

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n-C14	8.19E-10	8.19E-10	0
n-C15	1.98E-10	1.98E-10	0
n-C16	2.19E-11	2.19E-11	0
n-C17	5.62E-12	5.62E-12	0
n-C18	1.45E-12	1.45E-12	0
n-C19	4.39E-13	4.40E-13	0
n-C20	9.97E-15	9.97E-15	0
n-C21	1.73E-15	1.73E-15	0
n-C22	3.77E-16	3.77E-16	0
n-C23	3.54E-17	3.54E-17	0
n-C24	5.11E-18	5.11E-18	0
n-C25	9.14E-19	9.14E-19	0
n-C26	1.18E-19	1.18E-19	0
n-C27	9.64E-21	9.65E-21	0
n-C28	3.28E-21	3.28E-21	0
n-C29	7.92E-22	7.93E-22	0
n-C30	3.62E-24	3.62E-24	0
H2O	0.00048	0	1
NH3	0	0	0
Mass Flow Rate (kg/hr)	74046.8	74015.1	31.773

Table 7 Molecular Sieve bed

Chapter 4

4.6 De-Propanizer

Table 8 De-Propanizer

Comp	Feed(kmol)	xf	Recovery (%)	Distillate	xd
C2	10.70	0.0876	100	10.70	0.132
C3	71.75	0.5874	95	68.17	0.844
i-C4	14.31	0.1171	5	0.680	8.42e-3
n-C4	25.37	0.2077	5	1.27	0.016
i-C5	0.024	0.0002	0	0	0
Total	122.15	1		80.81	1

$$B = F - D$$

$$122.15 - 80.81 = 41.34 \text{ kmol/hr}$$

4.7 De-Butanizer

Name	D-ETH_TOP	D-BUT_TOP	D-BUT_BOT
CO2	0.016547	1.05E-05	1.45E-17
Nitrogen	0.008473	2.10E-12	1.24E-29
Methane	0.881103	1.59E-06	2.76E-20
Ethane	0.093877	0.087774	1.19E-10
Propane	9.33E-10	0.588016	4.73E-06
i-Butane	1.59E-17	0.117169	0.000789
n-Butane	4.75E-20	0.206875	0.01656
i-Pentane	2.10E-27	0.000147	0.419623
n-Pentane	6.06E-30	7.18E-06	0.343394
n-Hexane	1.00E-30	1.17E-11	0.124129
n-Heptane	1.00E-30	6.10E-17	0.07073

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n-Octane	1.00E-30	2.08E-22	0.021083
n-Nonane	1.00E-30	5.48E-28	0.003143
n-Decane	1.00E-30	9.95E-31	0.000479
n-C11	1.00E-30	1.00E-30	5.64E-05
Mass Flow Rate (kg/hr)	38901.0	10528.7	3214.8

Table 9 De-Butanizer

4.8 Overall Material Balance

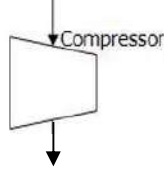
Stream	In(kg/hr)	Out(kg/hr)
Mix Feed	349900	--
Sale Gas	--	132890
Propane	--	7835
LPG	--	15356
Crude	--	181982
Water	--	11837
Total	349900	349900

Table 10 Overall Material Balance

Chapter 5

ENERGY BALANCE

5.1 Compressor



Power of compressor formula.

$$C_p = C_v + R$$

$$\frac{T_2}{T_1} = \left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}$$

$$\frac{C_p}{C_v} = \gamma$$

$$P = 2.31 \left(\frac{\gamma}{\gamma-1}\right) \left(\frac{T_2 - T_1}{M}\right) Q_m$$

With:

P1 = Pressure inlet compressor

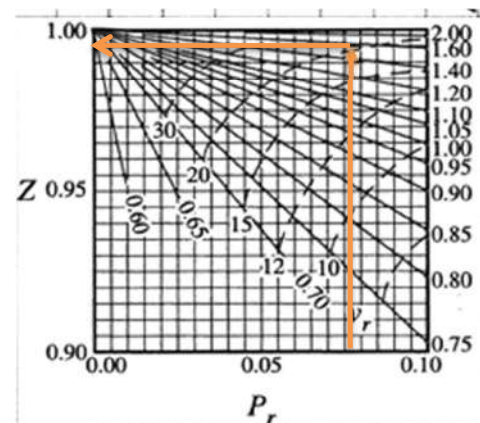
P2= Pressure outlet compressor

γ = Isentropic Coefficient

Cp = Specific heat at constant pressure

Cv = Heat capacity at fixed volume

γ	1.12
P2 (bar)	39.62
P1 (bar)	6.18
Temp in (k)	362.05
Cv (J/kgK)	105.5
Cp (J/kgK)	113.4



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	P in (psig)	P out (psig)	Energy (btu/hr)
Compressor 2	330	1030	1.099e+006
Compressor 3	450	500	2.026e+055
Compressor 4	438	500	9.551e+005
Compressor 5	500	1105	2.187e+006
Compressor 6	500	1105	2.187e+006
Compressor 7	500	1105	2.188e+006

Table 11 Compressor Duties

5.2 Expander/Turbine

P Discharge: 32.04 bar

P Suction: 71.13 bar

T in: 239.26

$\gamma = 2.79$

$$\frac{T_2}{T_1} = \left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}$$

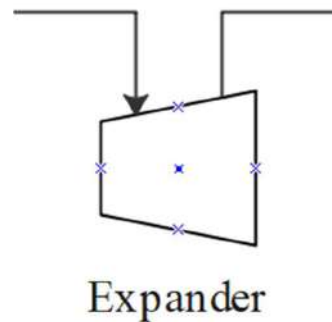
T out: 205.34

Cp: 25.28 J/kgK

Energy: 9.254e+005 Kj/hr

Power: 308KW

$$P = 2.31 \left(\frac{k}{k-1}\right) \left(\frac{T_{dis} - T_{suc}}{M}\right) Q_m$$



Chapter 5

5.3 De-Propanizer

“Feed temperature, $T_f = 53.64\text{ }^\circ\text{C}$

Heat Capacity of Feed, $C_{pf} = 0.42\text{ kcal/kg-}^\circ\text{C}$

Heat Capacity of Distillate, $C_{pd} = 0.44\text{ kcal/kg-}^\circ\text{C}$

Column top temperature, $T_t = -33.27\text{ }^\circ\text{C}$

Column bottom temperature, $T_b = 15.04\text{ }^\circ\text{C}$

Latent heat of top product, $H_D = 94\text{ kcal/kg}$

Latent heat of bottom product, $H_w = 88\text{ kcal/kg}$

Reflux ratio, $R = 1 : 2$

Vapor rates from column will be, $V = D + 2 * D = 3 * 669.34 = 2008.02\text{ kg/h}$

Heat load on reboiler will be,

$$Q_b = F * C_{pf} * (T_b - T_f) + V * H_D$$

$$\text{Reboiler duty } Q_b = 1000 * 0.42 * (104 - 50) + 2008.02 * 94$$

$$\text{Calculated reboiler duty } Q_b = 1.247e+006\text{ kJ/h}$$

So, heat load on condenser will be (assuming distillate is at 5 degC subcooled)

$$\text{Condenser duty } Q_c = V * C_{pd} * (T_t - 5) + V * H_D$$

$$\text{Putting the numbers, } Q_c = 2.301e+006\text{ kJ/h}$$

	De-Ethanizer	De-Butanizer	Crude Stabilization Column
Q _b (kJ/h)	7.222e+006	7.220e+006	1.049e+008
Q _c (kJ/h)	1.107e+007	8.259e+006	NIL

Table 12 Distillation Column Duties

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5.4 Air Cooler

Unrefined gas:

Mass flowrate: 6.859e+004 kg/hr

Cp= 34.11 J/kgK

$\Delta t = 49.5$ °F

$E_{in} = E_{out}$

$mC_p\Delta t$ (hot) = $mC_p\Delta t$ (cold)



Air Cooler

Air Stream:

Mass flowrate: 439200 kg/hr

Cp= 1 J/kgK

$\Delta t = 23.9$ °F

5.5 Heat Exchanger

Equation used: $Q = mC_p\Delta T$

For Heat Exchanger 1:

Mass flowrate: 2.188e+005

Cp = 27.31 j/kgK

$\Delta t = 218.7$ °F

$Q = mC_p\Delta T$

$$= 2.188e+005 * 27.31 * 218.7$$

$$= 1.288e+009 \text{ kJ/h}$$

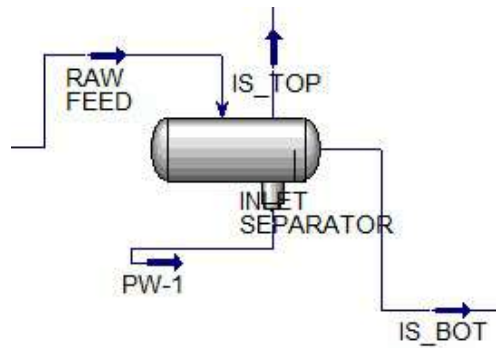
Chapter 6

DESIGN AND CALCULATIONS

6.1 Design of Inlet Separator

Operating Condition:

- $P = 7306 \text{ kPa}$
- Vapor mass flow = $6.86 \times 10^4 \text{ kg/hr}$
- Liquid mass flow = $9.01 \times 10^4 \text{ kg/hr}$
- $\rho_l = 864.7 \text{ kg/m}^3$
- $\rho_v = 59.87 \text{ kg/m}^3$



$$u_t = 0.07 \sqrt{\frac{(\rho_l - \rho_v)}{\rho_v}}$$

$$\rightarrow u_t = 0.07 \sqrt{\frac{(864.7 - 59.87)}{59.87}} = 0.257 \text{ m/s}$$

$$\bullet V_{ap} \text{ Volumetric flow rate} = \frac{68590}{3600 \times 59.87} = 0.323 \text{ m}^3/\text{s}$$

$$\bullet P = 73 \text{ bar}$$

$$\rightarrow L_v/D_v = 5$$

Operating pressure, bar	Length: diameter, L_v/D_v
0-20	3
20-35	4
>35	5

- Cross sectional area of vapor flow

Chapter 6

$$\rightarrow \frac{\pi D^2}{4} \times 0.5 = 0.393D^2$$

- Vapor velocity, $u_v = \frac{\text{Volumetric Flow rate}}{\text{Cross-sectional Area}}$

$$\frac{0.323}{0.393D^2} = \frac{0.82}{D^2}$$

- $h_v = 0.5D_v$
- Vapor resistance time $\rightarrow \frac{h_v}{u_t} \rightarrow \frac{0.5D_v}{0.257}$
- Actual vapor resistance time $\rightarrow \frac{L_v}{u_v} = \frac{5D^3}{0.83} = 6.02D^3$
- For satisfactory separation, residence time = actual

$$\rightarrow D = 1.89m, \text{ taken as } D = 2m$$

- Liquid volumetric flow rate = $\frac{\text{Mass flow rate of liquid}}{3600 \times \rho}$

$$= \frac{9.01 \times 10^4}{3600 \times 864.7} = 0.029 \text{ m}^3/\text{s}$$

- Liquid cross sectional area = $\frac{\pi \times 2^2 \times 0.5}{4} = 1.57 \text{ m}^2$
- Length = $5 \times 2 = 10 \text{ m}$
- Hold-up Volume = CS Area x Length

$$\rightarrow 10 \times 1.57 \rightarrow 15.7 \text{ m}^3$$

- Hold up time = $\frac{\text{Liquid Volume}}{\text{Flow Rate}}$

$$= \frac{15.7}{0.029} = 784s = \mathbf{13 \text{ mins}}$$

Chapter 6

6.2 Design of Compressor

Operating Conditions:

P Discharge: 39.62 bar

P suction: 6.18 bar

T inlet: 362.05K

$\gamma = 1.12$

$C_v = 105.5 \text{ J/kgK}$

$C_p = 113.4 \text{ J/kgK}$

$$C_p = C_v + R$$

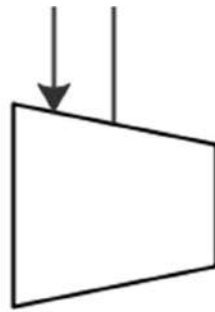
$$\frac{C_p}{C_v} = \gamma$$

$$\frac{T_2}{T_1} = \left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}$$

$$P = 2.31 \left(\frac{\gamma}{\gamma-1}\right) \left(\frac{T_2 - T_1}{M}\right) Q_m$$

T out: 441.79K (335 F)

Power: 6616 kW



Compressor

Chapter 6

6.3 Design of De-Propanizer

Basis:

C_4 = Light Key. $I-C_4$ = Heavy Key.

No more than 3mol/h LK in Bottoms.

No more than 3mol/h HK in Distillate.

Compound	Feed(mol/h)	X_f	Top (mol/h)	Bot (mol/h)	x_d	x_b
C_2	10.7	0.088	10.7	0.0	0.13	--
C_3	71.7	0.587	68.7	3.0	0.83	0.07
$i-C_4$	14.3	0.117	3.0	11.3	0.04	0.28
$n-C_4$	25.3	0.208	0.0	25.3	--	0.65
		$\Sigma=1.000$	$\Sigma=82.14$	$\Sigma=39.6$	$\Sigma=1.0$	$\Sigma=1.0$

Table 13 De-Propanizer light and heavy key

- Pressure = 36 psia
- Bubble Point = 15 C

Compound	K_i	$K_i x_i$
C_2	--	--
C_3	2.7	0.19
$i-C_4$	1.1	0.31
$n-C_4$	0.8	0.52
		$\Sigma=1.02$

Table 14 De-Propanizer $K_i x_i$ value

- Dew Point = -33 C

Chapter 6

Compound	K_i	y_i/K_i
C ₂	3.5	0.04
C ₃	1.2	0.71
i-C ₄	0.16	0.25
n-C ₄	--	--
		$\Sigma=0.99$

Table15 De-Propanizer Kiyi value

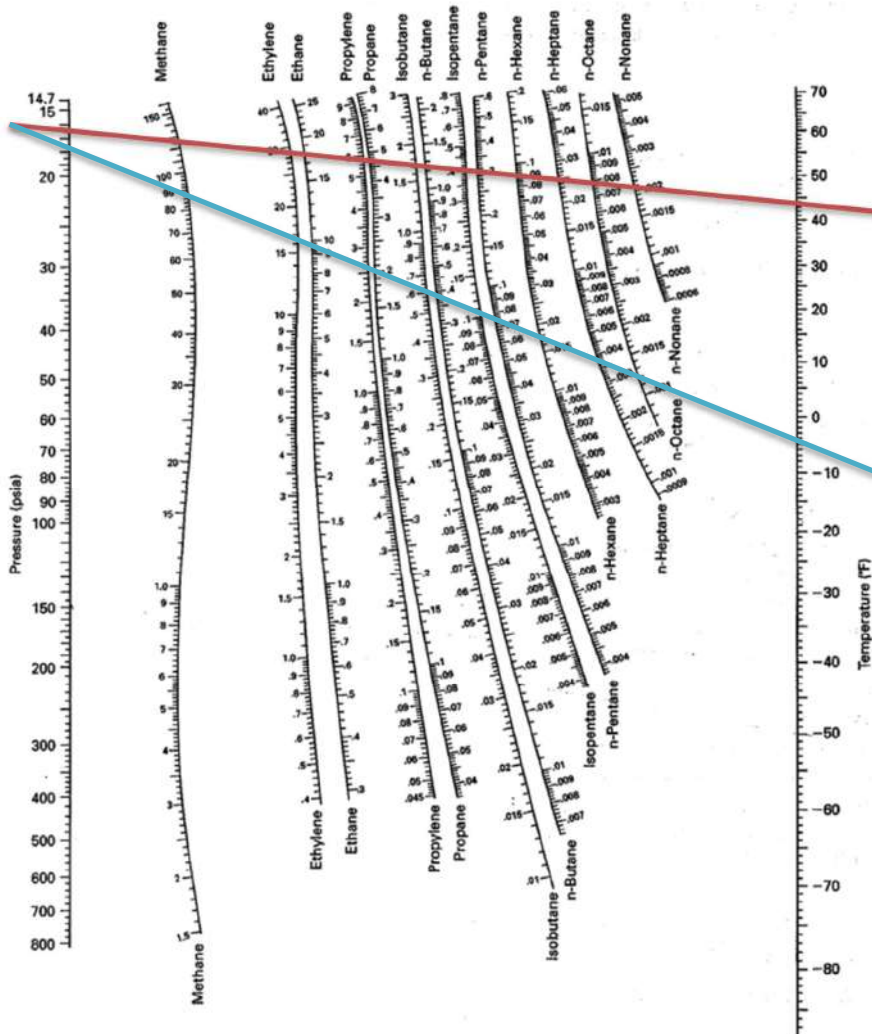


Figure 4 Dew point and bubble point

Chapter 6

- We now calculate average relative volatilities:

Compound	Top(-33°C)	Bot(15°C)	Avg. α
C ₂	8.6	21.8	15.1
C ₃	2.4	7.5	5.0
i-C ₄	1.0	1.0	1.0
n-C ₄	0.7	0.6	0.65

$$\alpha_i = \frac{K_i}{K_{HK}}$$

Table 16 De-Propanizer temperature

- Limiting non-key flows (Rectifying Section):

C ₂	α_i	d _i	$l_i = \frac{d_i}{(\alpha_i - 1)}$	$\underline{v}_i = l_i + d_i$
	15.1	10.7	0.75	11.45

Table 17 Relative Volatility for C2

- Limiting non-key flows (Rectifying Section):

n-C ₄	α_i	b _i	$v_i = \frac{\alpha_i b_i}{(\alpha_{lk} - \alpha_i)}$	$l'_i = v_i + b_i$
	0.65	25.3	3.78	29.08

Table 18 Relative Volatility for n-C4

- Heat to vaporize feed = 1.815×10^6 kJ/hr
- Molar flow rate = 122 mol/hr
- Heat to vaporize 1 mol feed:

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 1 - q$$

$$\rightarrow \frac{1.815 \times 10^6}{1 \text{ hr}} * \frac{1 \text{ hr}}{122 \text{ mol}} \rightarrow 1.49 \times 10^4 \text{ kJ/mol}$$

$$q = \frac{\text{heat to vaporise 1 mol of feed}}{\text{molar latent heat of feed}} = \frac{1.49 \times 10^4}{1.48 \times 10^4} \approx 1.0$$

Chapter 6

$$\sum \frac{\alpha_i x_{i,f}}{\alpha_i - \theta} = 0$$

- For a value of $\theta = 1.68$

Compound	α_i	$x_{i,f}$	$\frac{\alpha_i x_{i,f}}{\alpha_i - \theta}$
C ₂	15.1	0.088	0.144
C ₃	5.0	0.587	0.153
i-C ₄	1.0	0.117	-0.172
n-C ₄	0.65	0.208	-0.127
		$\Sigma=1.00$	$\Sigma \approx 0.00$

Table 19 volatility for feed

Compound	α_i	$x_{i,d}$	$\frac{\alpha_i x_{i,d}}{\alpha_i - \theta}$
C ₂	15.1	0.13	0.219
C ₃	5.0	0.83	1.875
i-C ₄	1.0	0.04	-0.089
n-C ₄	0.65	--	--
		$\Sigma=1.00$	$\Sigma=1.995 \approx 2.00$

Table 20 volatility for distillate

- $R_m + 1 = 2$

→ $R_m = 1$

$$\sum \frac{\alpha_i x_{i,d}}{\alpha_i - \theta} = R_m + 1$$

Chapter 6

- Use equation,

$$N_m = \frac{\log\left(\frac{x_{lk}}{x_{hk}}\right)_d \left(\frac{x_{hk}}{x_{lk}}\right)_b}{\log 1k} = 13$$

$$\frac{R_m}{1 + R_m} = 0.5$$

- Say $R = 2.0$, $\frac{R}{R+1} = 0.67$

R	2	3
N	18.05	17.17

- $N_m/N = 0.73$
- $N = 18.05 = \underline{19 \text{ plates}}$

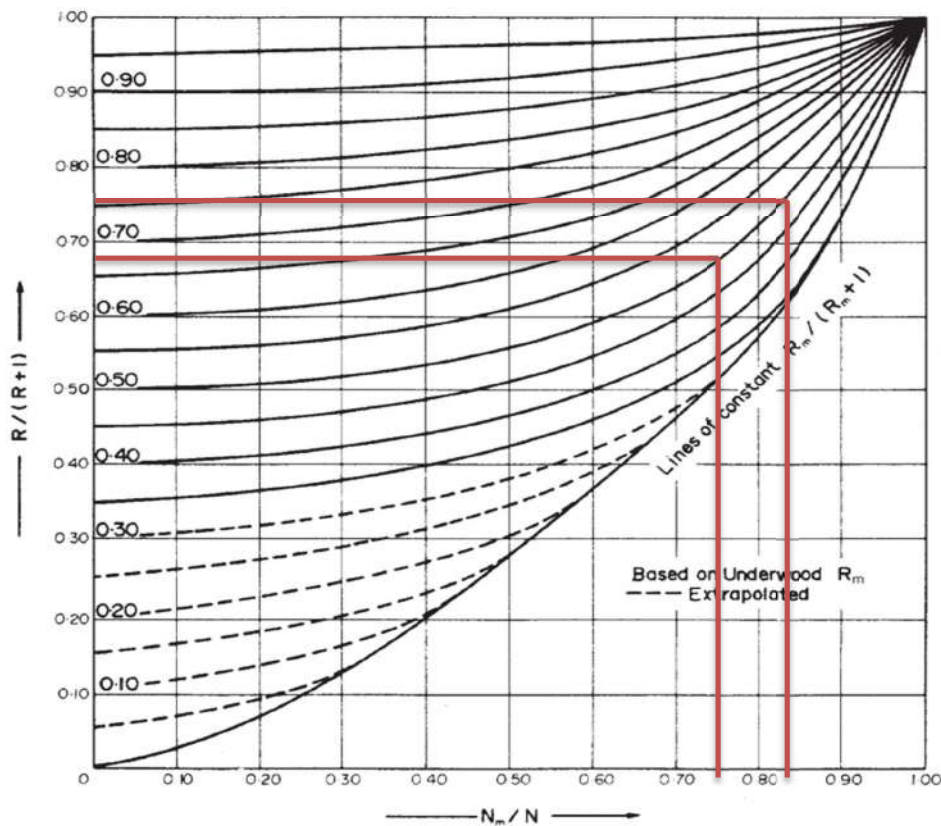


Figure 5 Erbar-Madox correlations

Chapter 6

Efficiency:

- $\mu_{\alpha}\alpha_{\alpha} = 0.196$
- $E = 51 - 32.5\log(0.196)$
- $E = 74\%$
- Actual No. of Plates = $19/0.74 = 25.67 \rightarrow \underline{26 \text{ plates}}$

Feed Point Location:

- Use the empirical equation given by Kirkbride.
- $N_r + N_s = 25$.
- $\frac{N_r}{N_s} =$

$$\log \left[\frac{N_r}{N_s} \right] = 0.206 \log \left[\left(\frac{B}{D} \right) \left(\frac{x_{f, \text{HK}}}{x_{f, \text{LK}}} \right) \left(\frac{x_{b, \text{LK}}}{x_{d, \text{HK}}} \right)^2 \right]$$

Column Diameter:

- Based on the Souders and Brown equation,

$$\mu_v = (-0.171l_t^2 + 0.27l_t - 0.047) \left(\frac{\rho_l - \rho_v}{\rho_v} \right)^{1/2}$$

- $l_t =$ plate spacing = 1.0m

Chapter 6

6.4 Design of Low Pressure Separator:

Settling velocity:

$$u_t = 0.07 * \sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

Liquid Density	ρ_L	1007.356
Vapor Density	ρ_v	1.225213
Settling Velocity	u_t (m/s)	2.005946

Table 21 Densities

Vessel Size:

$$D_v = \sqrt{\frac{4 * V}{\pi * \rho_v * u_t}}$$

Mass Flow Rate of Gas	V (kg/hr.)	572367.1
Mass Flow Rate of Gas	V (kg/sec.)	158.9909
Vapor Density	ρ_v (kg/m ³)	1.225213
Diameter of Vessel	D_v (m)	9.077907

Table 22 Properties of fluids

Chapter 6

Liquid Holdup:

$$V_{holdup} = Q_{liq} \times t$$

$$h_{liq} = \frac{V_{holdup} \times 4}{\pi \times D_v \times s q.}$$

From the book of Richardson Coulson

For a minimum holdup time of 600 seconds

$$Q_{liq} = 0.005413 \text{ m}^3/\text{s}$$

$$V_{holdup} = 3.24779863$$

V_{holdup}	$Q_{liq} \times t$	3.24779863
h_{liq}		0.05020511
h_{vap}	$D_v + 0.4$	9.477907
h_{inlet}	$D_v \times 0.5$	4.538954
h_{total}	$h_{liq} + h_{vap} + h_{inlet}$	14.0670659
$\frac{h_{total}}{D_v}$		1.54959349

Table 23 Specifications of separator

Chapter 6

6.5 Design of Heat Exchanger:

$$\dot{Q}_{shell} = \dot{m} * (c_p * \Delta T + x * h_v)$$

Where,

x = quality of vapor

$$\dot{Q}_{shell} = 60.972 * [2.133 * (182.8 - 137) + 0.17 * 2101] = 27.75 \text{ MW}$$

Similarly, heat load in tube side is:

$$\dot{Q}_{tube} = 11.05 * [1.941 * (131.5 - 122.9) + 1 * 2495] = 27.75 \text{ MW}$$

LMTD

HOT FLUID		COLD FLUID	Difference
182.8	High T	131.5	51.3
137	Low T	122.9	14.1
45.8		8.6	37.2

TABLE 24- DATA FOR LMTD CALCULATIONS

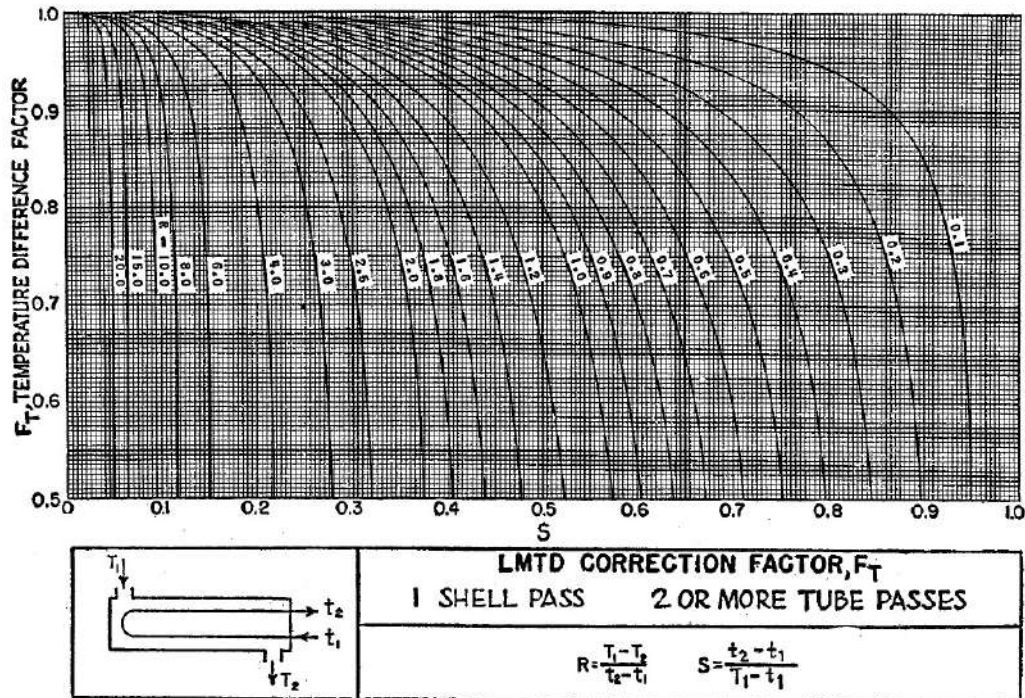
$$LMTD = \frac{51.3 - 14.1}{\ln\left(\frac{51.3}{14.1}\right)} = 28.8 \text{ C}$$

To find true LMTD, we first find R & S factors. Assume 1-6 passes:

$$R = \frac{45.8}{8.6} = 5.33$$

$$S = \frac{8.6}{182.8 - 122.9} = 0.144$$

Chapter 6

FIGURE 6- F_T CORRECTION FACTOR FOR 1-6 HX

From above graph,

$$F_T = 0.99$$

$$\therefore \Delta t_{lm} = 0.99 * 28.8 = 28.51 \text{ C}$$

Shell Side Heat Transfer Coefficient

1. Calculate equivalent diameter:

$$D_e = 4 * \frac{\text{axial flow area}}{\text{wetted perimeter}}$$

From table

$$D_e = 0.99$$

2. Calculate Reynold's No.:

$$Re = \frac{D_e * W * P_t}{ID * C * B * \mu} = \frac{0.99 * 60.972 * 1.25}{25 * 0.25 * 5 * 4.76 * 10^{-5}} = 3.54 * 10^5$$

3. Determine Chilton-Colburn Factor (j_H):

Chapter 6

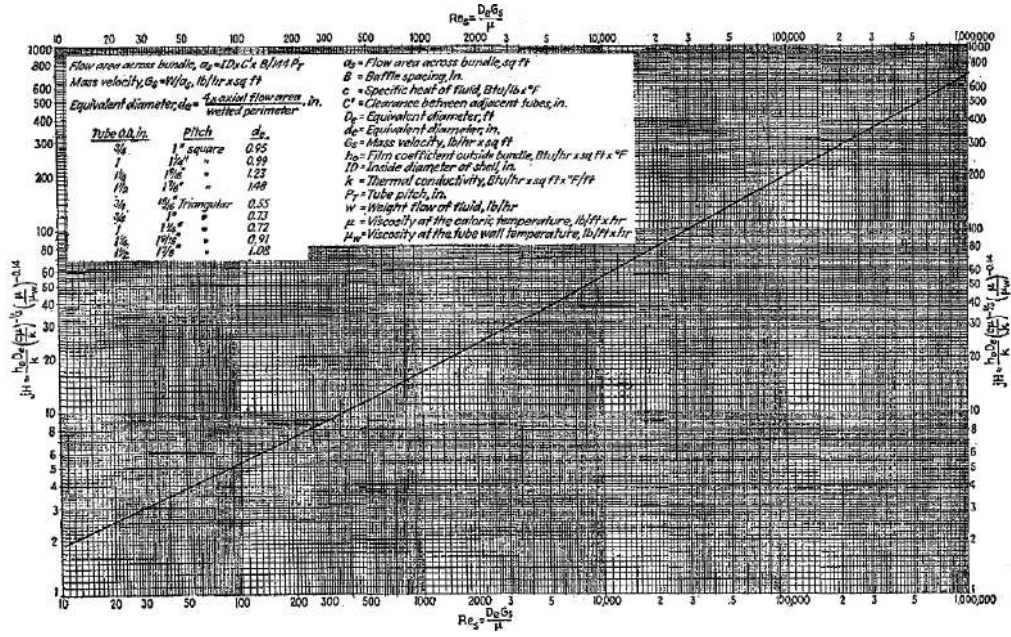


FIGURE 7- JH VS RE CURVE FOR SHELL SIDE

From above graph, for $Re = 3.54 \cdot 10^5$,

$$j_H = 420$$

4. Calculate h_o :

$$h_o = \frac{j_H}{D_e} * k^{\frac{2}{3}} * (c_p * \mu)^{\frac{1}{3}}$$

$$h_o = \frac{420}{0.99 \text{ in}} * 0.013^{\frac{2}{3}} * (2133 * 4.76 * 10^{-5})^{\frac{1}{3}} = 720 \frac{W}{m^2 \cdot K}$$

Chapter 6

Tube Side Heat Transfer Coefficient

APPENDIX OF CALCULATION DATA

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TABLE 10. HEAT EXCHANGER AND CONDENSER TUBE DATA

Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in. ²	Surface per lin ft, ft ²		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876			0.403
	16	0.065	0.370	0.1076			0.329
	18	0.049	0.402	0.127			0.258
¾	20	0.035	0.430	0.145	0.1963	0.1263	0.190
	10	0.134	0.482	0.182			0.965
	11	0.120	0.510	0.204			0.884
	12	0.109	0.532	0.223			0.817
	13	0.095	0.560	0.247			0.727
	14	0.083	0.584	0.268			0.647
	15	0.072	0.606	0.289			0.571
	16	0.065	0.620	0.302			0.520
1	17	0.058	0.634	0.314	0.2618	0.1754	0.469
	18	0.049	0.652	0.334			0.401
	8	0.165	0.670	0.355			1.61
	9	0.148	0.704	0.389			1.47
	10	0.134	0.732	0.421			1.36
	11	0.120	0.760	0.455			1.23
	12	0.109	0.782	0.478			1.11
	13	0.095	0.810	0.515			1.00
1 ¼	14	0.083	0.824	0.546	0.3271	0.2409	0.890
	15	0.072	0.856	0.576			0.781
	16	0.065	0.870	0.594			0.710
	17	0.058	0.884	0.613			0.639
	18	0.049	0.902	0.639			0.545
	8	0.165	0.920	0.665			2.09
	9	0.148	0.954	0.714			1.91
	10	0.134	0.982	0.757			1.75
1 ½	11	0.120	1.01	0.800	0.3271	0.2498	1.58
	12	0.109	1.03	0.836			1.45
	13	0.095	1.06	0.884			1.28
	14	0.083	1.08	0.923			1.13
	15	0.072	1.11	0.960			1.00
	16	0.065	1.13	1.000			0.901
	17	0.058	1.15	1.040			0.810
	18	0.049	1.17	1.080			0.720

FIGURE 8- DIMENSION OF STEEL PIPES IPS

1. Calculate flow area of one pipe, therefore, from above figure:

$$a'_t = 0.2618 \text{ in}^2 = 0.0186 \text{ m}^2$$

$$a_t = \frac{n * a'_t}{\text{passes}} = \frac{252 * 0.0186}{2} = 0.0193 \text{ m}^2$$

2. Calculate Re:

$$Re = \frac{D * W_t}{\mu * a_t} = \frac{0.81 \text{ in} * 11.05}{4.13 * 10^{-6} * 0.0193} = 3.04 * 10^5$$

3. Determine Chilton-Colburn Factor (j_H):

Chapter 6

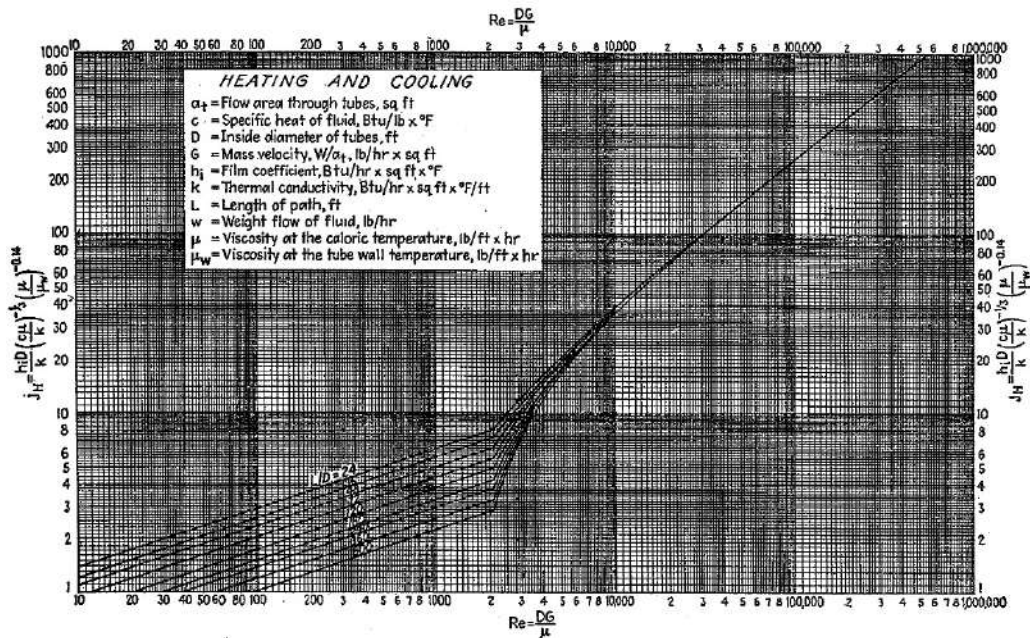


FIGURE 9- JH VS RE CURVE FOR TUBE SIDE

From above graph, for $Re = 3.04 \cdot 10^5$,

$$j_H = 600$$

1. Calculate h_i :

$$h_i = \frac{j_H}{D} * k^{\frac{2}{3}} * (c_p * \mu)^{\frac{1}{3}}$$

$$h_i = \frac{600}{0.81 \text{ in}} * 0.76^{\frac{2}{3}} * (1941 * 4.13 * 10^{-6})^{\frac{1}{3}} = 650 \frac{W}{m^2.K}$$

1. Calculate h_{io} :

$$h_{io} = 650 * \frac{0.81}{1} = 595 \frac{W}{m^2.K}$$

Overall Heat Transfer Coefficient

1. Overall Clean HT Coefficient:

$$U_c = \frac{595 * 720}{595 + 720} = 325.78 \frac{W}{m^2.K}$$

Chapter 6

2. Overall Dirty HT Coefficient:

$$a'' = 0.622 \frac{ft^2}{ft} = 0.289 \frac{m^2}{m}$$

$$A = 252 * 0.622 * 20ft * 2 \text{ passes} = 444 m^2$$

$$U_D = \frac{Q}{A * \Delta t_{lm}} = \frac{27.75MW}{444 * 28.51} = 303.41 \frac{W}{m^2 \cdot K}$$

1. Dirt Factor:

$$R_d = \frac{325.78 - 303.41}{325.78 * 303.41} = 0.00023 m^2 \cdot \frac{K}{W} (< 0.002, \text{Acceptable})$$

Shell Side Pressure Drop

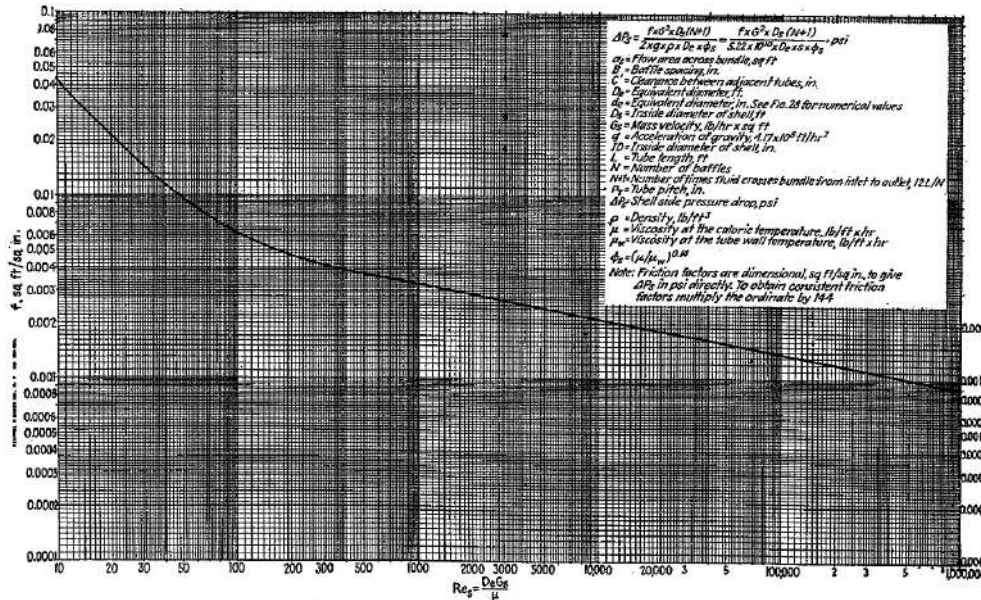


FIGURE 10- SHELL SIDE FRICTION FACTORS

For $Re = 3.54 * 10^5$,

$$f = 0.0014 \frac{ft^2}{in^2} = 0.2016 \frac{m^2}{m^2}$$

$$N + 1 = \frac{L}{B} = \frac{20ft}{5/12ft} = 29.7 \cong 30$$

$$\Delta P_s = \frac{0.2016 * 25 * 30 * 767688^2}{5.22 * 10^{10} * 0.0254 * 0.99} = 8.86. \text{ psi}$$

Chapter 6

Tube Side Pressure Drop

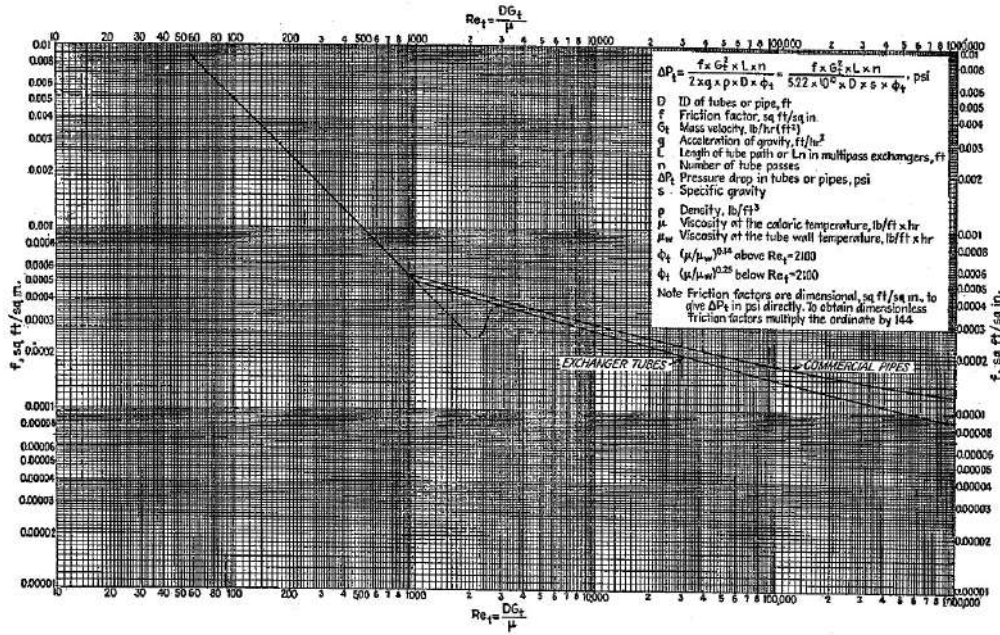


FIGURE 11- TUBE SIDE FRICTION FACTORS

For $Re = 3.04 \times 10^5$,

$$f = 0.00016 \frac{ft^2}{in^2} = 0.01872 \frac{m^2}{m^2}$$

$$\Delta P_t = \frac{0.01872 * 20ft * 2 * 21.47^2}{5.22 * 10^{10} * 0.154 * 0.9805} = 6.7 \text{ psi}$$

Chapter 6

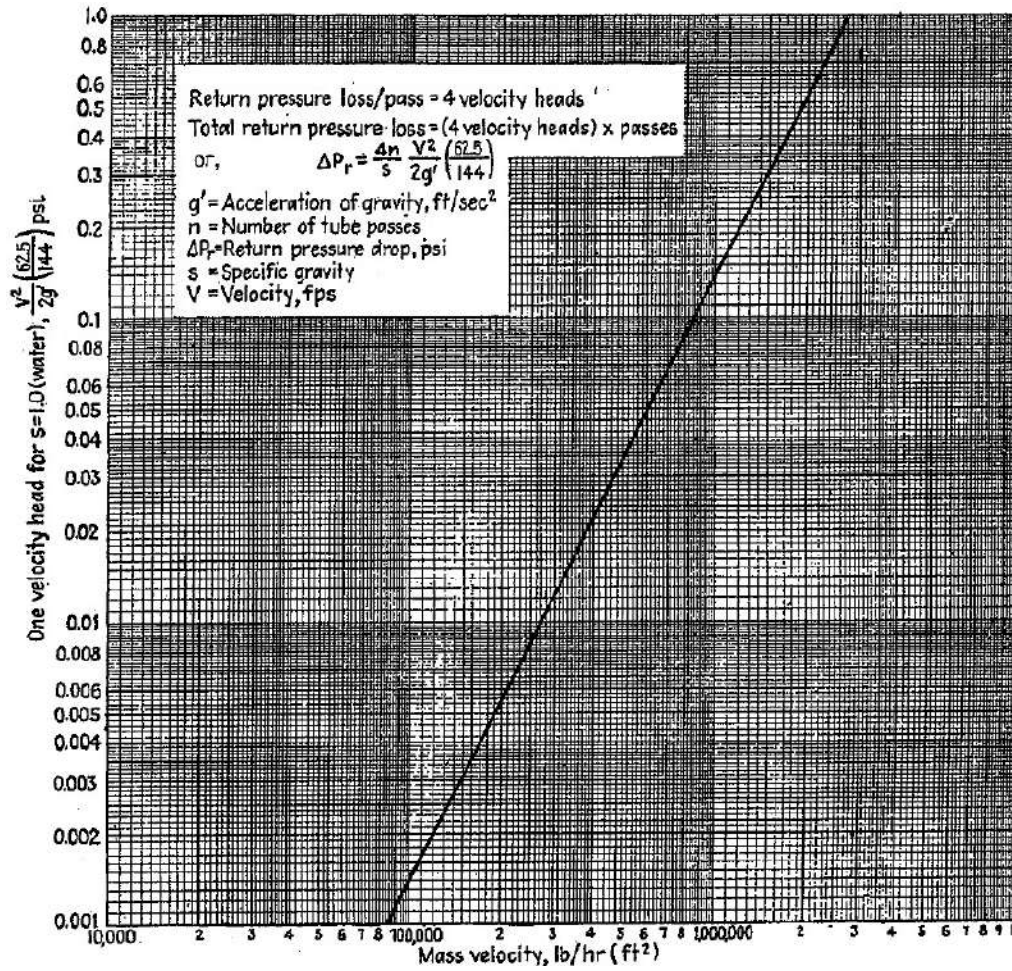


FIGURE 12- TUBE SIDE RETURN PRESSURE LOSS

For mass velocity = 8.5×10^5 ,

$$\frac{v^2}{2g} = 0.09 \text{ psi}$$

$$\Delta P_r = \frac{4 \times 6}{0.9805} \times 0.09 = 2.20 \text{ psi}$$

$$\therefore \Delta P = 2.20 \text{ psi} + 6.7 = 8.9 \text{ psi} (< 10 \text{ psi, Acceptable})$$

Chapter 7

COST ESTIMATION AND ECONOMIC ANALYSIS

7.1 Cost of Equipment: -

7.1.1 Horizontal Vessels:

For Inlet Separator:

Diameter = 2 m

Length = 10 m

Pressure = 70 bar

= \$20,000 x 1.0 x 2.5

= \$50,000

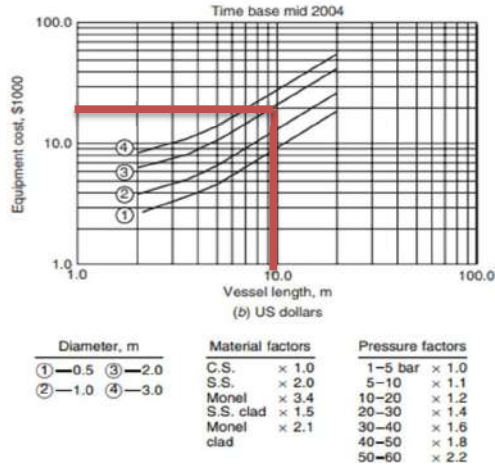


Figure 13 Horizontal Vessel costing

Equipment	Length (m)	Pressure (bar)	Diameter (m)	Cost (\$)
H-P separator	15.0	73	2.0	75,000
M-P separator	7.5	24	1.5	25,200
L-P separator	4.0	10	1.0	6,000
C-F separator	4.0	9	0.6	5,000
Over Head	8.0	24	1.0	28,000

All Material Factors are taken to be that of carbon Steel.

Table 25 Cost of separators

Chapter 7

7.1.2 Vertical Vessel:

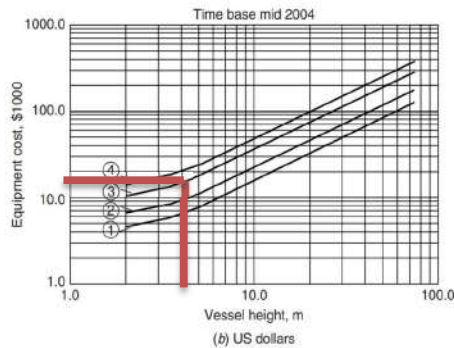
Diameter = 1 m

Height = 3 m

Pressure = 70 bar

$$= \$16,000 \times 1.0 \times 2.5$$

$$= \$ 40,000$$



Diameter, m		Material factors	Pressure factors
①—0.5	③—2.0	C.S. × 1.0	1–5 bar × 1.0
②—1.0	④—3.0	S.S. × 2.0	5–10 × 1.1
		Monel × 3.4	10–20 × 1.2
		S.S. clad × 1.5	20–30 × 1.4
		Monel × 2.1	30–40 × 1.6
			40–50 × 1.8
			50–60 × 2.2

For 3 vessels,

$$= \$40,000 \times 3$$

$$= \$120,000$$

Figure 14 Vertical Vessel Costing

7.1.3 Turbine:

Power = 57kW

$$C_e = CS^n$$

$$= \$3,000 \times 570.5$$

$$= \$48,094$$

Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. C

Equipment	Size unit, S	Size range	Constant C,£	Constant C,\$	Index n
Agitators					
Turbine	driver power, kW	3–75	1200	1900	0.5
Boilers					
Packaged	kg/h steam	(5–50) × 10 ³	70	120	0.8
			60	100	0.8
Centrifuges					
Horizontal basket	dia., m	0.5–1.0	35,000	58,000	1.3
Vertical basket			35,000	58,000	1.0
Compressors					
Centrifugal	driver power, kW	20–500	1160	1920	0.8
Reciprocating			1600	2700	0.8
Conveyors					
0.5 m wide		length, m	2–40	1200	0.75
1.0 m wide				1800	0.75
Crushers					
Cone	t/h	20–200	2300	3800	0.85
Pulverisers	kg/h		2000	3400	0.35
Dryers					
Rotary	area, m ²	5–30	21,000	35,000	0.45
Pan		2–10	4700	7700	0.35

Figure 15 Miscellaneous equipment cost

Chapter 7

7.1.4 Compressor:

Power = 6616kW

Equipment	Size unit, S	Size range	Constant C, £	Constant C, \$	Index n
Agitators					
Propeller	driver power, kW	5-75	1200	1900	0.5
Turbine	driver power, kW	5-75	1800	3000	0.5
Boilers					
Packaged up to 10 bar	kg/h steam	(5-50) × 10 ³	70	120	0.8
Vertical basket			60	100	0.8
Centrifuges					
Horizontal basket	dia., m	0.5-1.0	35,000	58,000	1.3
Vertical basket			35,000	58,000	1.0
Compressors					
Centrifugal	driver power, kW	20-500	1160	1920	0.8
Reciprocating			1600	2700	0.8
Conveyors					
Belt	length, m	2-40			
0.5 m wide			1200	1900	0.75
1.0 m wide			1800	2900	0.75
Crushers					
Cone	t/h	20-200	2300	3800	0.85
Pulverisers	kg/h		2000	3400	0.35
Dryers					
Rotary	area, m ²	5-30	21,000	35,000	0.45
Pan		2-10	4700	7700	0.35

Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. C

Equipment	Size unit, S	Size range	Constant C, £	Constant C, \$	Index n
Agitators					
Propeller	driver power, kW	5-75	1200	1900	0.5
Turbine	driver power, kW	5-75	1800	3000	0.5
Boilers					
Packaged up to 10 bar	kg/h steam	(5-50) × 10 ³	70	120	0.8
Vertical basket			60	100	0.8
Centrifuges					
Horizontal basket	dia., m	0.5-1.0	35,000	58,000	1.3
Vertical basket			35,000	58,000	1.0
Compressors					
Centrifugal	driver power, kW	20-500	1160	1920	0.8
Reciprocating			1600	2700	0.8
Conveyors					
Belt	length, m	2-40			
0.5 m wide			1200	1900	0.75
1.0 m wide			1800	2900	0.75
Crushers					
Cone	t/h	20-200	2300	3800	0.85
Pulverisers	kg/h		2000	3400	0.35
Dryers					
Rotary	area, m ²	5-30	21,000	35,000	0.45
Pan		2-10	4700	7700	0.35

= \$1920 x 66160.8

= \$2,190,000

For 7 compressors,

= \$2,190,000 x 7

= \$15,330,000

Figure 15 Miscellaneous equipment cost

7.1.5 Distillation Column:

Height = 26m

Diameter = 2m

Vessel cost = \$100,000

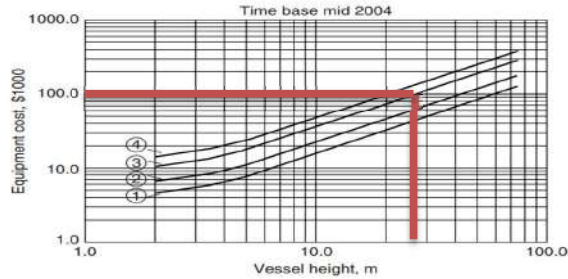


Figure 16 Vessel height costing

Sieve Trays = \$700 x 26

= \$18,200

Total Cost:

= \$100,000 + \$18,200

= \$118,200

For 6 columns,

= \$118,200 x 6

= \$472,800

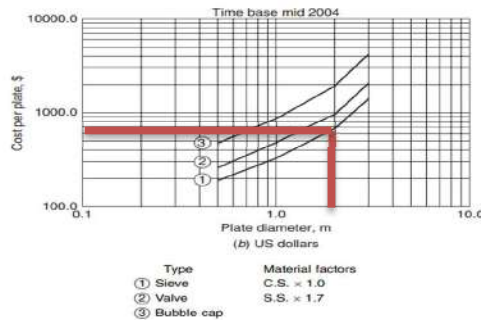


Figure 17 plate dia costing

Chapter 7

7.1.6 Heat Exchanger:

Area = 150m²

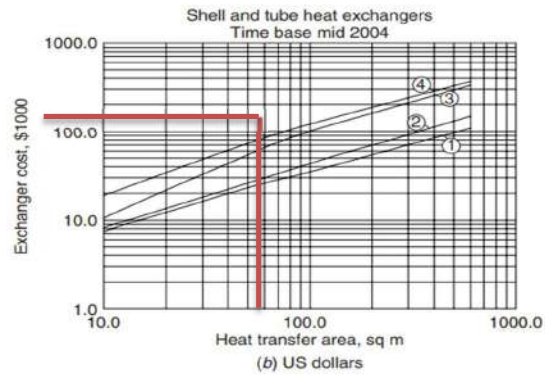
P= 2 bar

= \$150,000

For 9 STHE,

= \$150,000 x 9

= \$1,350,000



Materials		Pressure factors	Type factors
Shell	Tubes		
① Carbon steel	Carbon steel	1-10 bar × 1.0	Floating head × 1.0
② C.S.	Brass	10-20 × 1.1	Fixed tube sheet × 0.8
③ C.S.	Stainless steel	20-30 × 1.25	U tube × 0.85
④ S.S.	S.S.	30-50 × 1.3	Kettle × 1.3
		50-70 × 1.5	

Figure 18 heat transfer area costing

Chapter 7

Total Purchase Cost = \$ 27,620,894

Item	Process type		
	Fluids	Fluids– solids	Solids
1. Major equipment, total purchase cost	PCE	PCE	PCE
f_1 Equipment erection	0.4	0.45	0.50
f_2 Piping	0.70	0.45	0.20
f_3 Instrumentation	0.20	0.15	0.10
f_4 Electrical	0.10	0.10	0.10
f_5 Buildings, process	0.15	0.10	0.05
* f_6 Utilities	0.50	0.45	0.25
* f_7 Storages	0.15	0.20	0.25
* f_8 Site development	0.05	0.05	0.05
* f_9 Ancillary buildings	0.15	0.20	0.30
2. Total physical plant cost (PPC)			
PPC = PCE (1 + f_1 + ... + f_9)			
= PCE ×	3.40	3.15	2.80
f_{10} Design and Engineering	0.30	0.25	0.20
f_{11} Contractor's fee	0.05	0.05	0.05
f_{12} Contingency	0.10	0.10	0.10
Fixed capital = PPC (1 + f_{10} + f_{11} + f_{12})			
= PPC ×	1.45	1.40	1.35

*Omitted for minor extensions or additions to existing sites.

Figure 19 PCE cost

Chapter 7

CEPCI in 2022 = 271

Table 6.2. Purchase cost of miscellaneous equipment, cost factors for use in equation 6.7. Cost basis mid 2004

Equipment	Size unit, S	Size range	Constant		Index	Comment
			C,£	C,\$	n	
Agitators						
Propeller	driver	5-75	1200	1900	0.5	
Turbine	power, kW		1800	3000	0.5	
Boilers						
Packaged						oil or gas fired
up to 10 bar	kg/h steam	$(5-50) \times 10^3$	70	120	0.8	
10 to 60 bar			60	100	0.8	
Centrifuges						
Horizontal basket	dia., m	0.5-1.0	35,000	58,000	1.3	carbon steel
Vertical basket			35,000	58,000	1.0	$\times 1.7$ for ss
Compressors						
Centrifugal	driver	20-500	1160	1920	0.8	electric,
	power, kW					max. press.
Reciprocating			1600	2700	0.8	50 bar
Conveyors						
Belt	length, m	2-40				
0.5 m wide			1200	1900	0.75	
1.0 m wide			1800	2900	0.75	
Crushers						
Cone	t/h	20-200	2300	3800	0.85	
Pulverisers	kg/h		2000	3400	0.35	
Dryers						
Rotary	area, m ²	5-30	21,000	35,000	0.45	direct
Pan		2-10	4700	7700	0.35	gas fired
Evaporators						
Vertical tube	area, m ²	10-100	12,000	20,000	0.53	carbon steel
Falling film			6500	10,000	0.52	
Filters						
Plate and frame	area, m ²	5-50	5400	8800	0.6	cast iron
Vacuum drum		1-10	21,000	34,000	0.6	carbon steel
Furnaces						
Process						
Cylindrical	heat abs, kW	10^3-10^4	330	540	0.77	carbon steel
Box		10^3-10^5	340	560	0.77	$\times 2.0$ ss
Reactors						
Jacketed, agitated	capacity, m ³	3-30	9300	15,000	0.40	carbon steel
			18,500	31,000	0.45	glass lined
Tanks						
Process	capacity, m ³					
vertical		1-50	1450	2400	0.6	atmos. press.
horizontal		10-100	1750	2900	0.6	carbon steel
Storage						
floating roof		50-8000	2500	4350	0.55	$\times 2$ for
cone roof		50-8000	1400	2300	0.55	stainless

Figure 20 Miscellaneous equipment

Chapter 7

(Total Physical Plant Cost) PPC = PCE *3.4	\$93,911,039
(Fixed Capital) FC = PPC*1.45	\$136,171,007
Working Capital (15% of FC)	\$6,808,550
Total Investment Required (Fixed Capital + Working Capital)	\$142,979,557
Operating Time of Plant	365

Table 26 Total Investment

7.2 Operating Cost

Cost Type	Calculation	Cost in dollars
Maintenance	5 % of Fixed Capital	6,808,550
Operating Labour	Estimate	300,000
Laboratory Costs	20 % of Operating Labour	60,000
Supervision	20 % of Operating Labour	60,000
Plant Overheads	50 % of Operating Labour	150,000
Capital Charges	10 % of Fixed Capital	13,617,100
Local Taxes	2 % of Fixed Capital	2,723,420
Total	Sum of all costs	23,659,070

Table 27 Operating Cost

Chapter 7

7.2.1 Variable Cost

Operating Time: 365 days

= 8160 hours

Raw material = $1.587 \times 10^5 \times 0.139 \times 24 \times 365 = \mathbf{\$193239468/year}$

Electricity cost = 0.15\$/MJ

Total electricity consumed = 667kW = 667,000J/s -> $667,000 \times 86400 \times 365 = 2103467 \text{ MJ/y}$

Total cost = $2103467 \times 0.15 = \$315520$

Index in 2004 = 188.9

Index in 2021 = 271

Cost in 2021 = $315520 \times 271/188.9 = \452651

TOTAL OPERATING COST = $\$193239468 + \$452651 + 23659070 = \mathbf{\$217,351,189/y}$

7.3 Payback Period

LPG sale = 6088276/y

Propane sale = 545288/y

Ch4 sale = 22237216/y

Crude sale = 10125974/y

PAYOUT PERIOD = total investment/total sales

= $\$142979557/38345249 = \mathbf{3.73 \text{ years}}$

Chapter 8

SIMULATION

8.1 Selection of Software

We had the choice of performing our simulation on either ASPEN HYSYS or ASPEN PLUS, which are both advanced Software for chemical plant modelling. Both have a diverse bank of fluid packages and equipment selection. The main difference between the two is that ASPEN HYSYS has more fluid packages favoring the Petroleum and Petrochemical Assays while ASPEN PLUS has more fluid packages favoring the Polymer, Chemical and Electrolytic Assays. Since our Project is heavily petroleum related, we have chosen ASPEN HYSYS as our simulation software.

8.2 Selection of Fluid Package

We have selected Peng-Robinson as the fluid package because we were dealing with pressures higher than 10 bar and all components are non-electrolytic (except water).

The screenshot shows the 'Property Package Selection' dialog box. On the left, a list of property packages is shown, with 'Peng-Robinson' selected. On the right, the 'Options' table is displayed, showing the following settings:

Property	Value
Enthalpy	Property Package EOS
Density	Costald
Modify Tc, Pc for H2, He	Modify Tc, Pc for H2, He
Indexed Viscosity	HYSYS Viscosity
Peng-Robinson Options	HYSYS
EOS Solution Methods	Cubic EOS Analytical Method
Phase Identification	Default
Surface Tension Method	HYSYS Method
Thermal Conductivity	API 12A3.2-1 Method

Chapter 8

8.3 Component List

Component	Type		
CO2	Pure Component		
Nitrogen	Pure Component	n-C15	Pure Component
Methane	Pure Component	n-C16	Pure Component
Ethane	Pure Component	n-C17	Pure Component
Propane	Pure Component	n-C18	Pure Component
i-Butane	Pure Component	n-C19	Pure Component
n-Butane	Pure Component	n-C20	Pure Component
i-Pentane	Pure Component	n-C21	Pure Component
n-Pentane	Pure Component	n-C22	Pure Component
n-Hexane	Pure Component	n-C23	Pure Component
n-Heptane	Pure Component	n-C24	Pure Component
n-Octane	Pure Component	n-C25	Pure Component
n-Nonane	Pure Component	n-C26	Pure Component
n-Decane	Pure Component	n-C27	Pure Component
n-C11	Pure Component	n-C28	Pure Component
n-C12	Pure Component	n-C29	Pure Component
n-C13	Pure Component	n-C30	Pure Component
n-C14	Pure Component	H2O	Pure Component

Chapter 8

8.4 Simulation on Aspen HYSYS

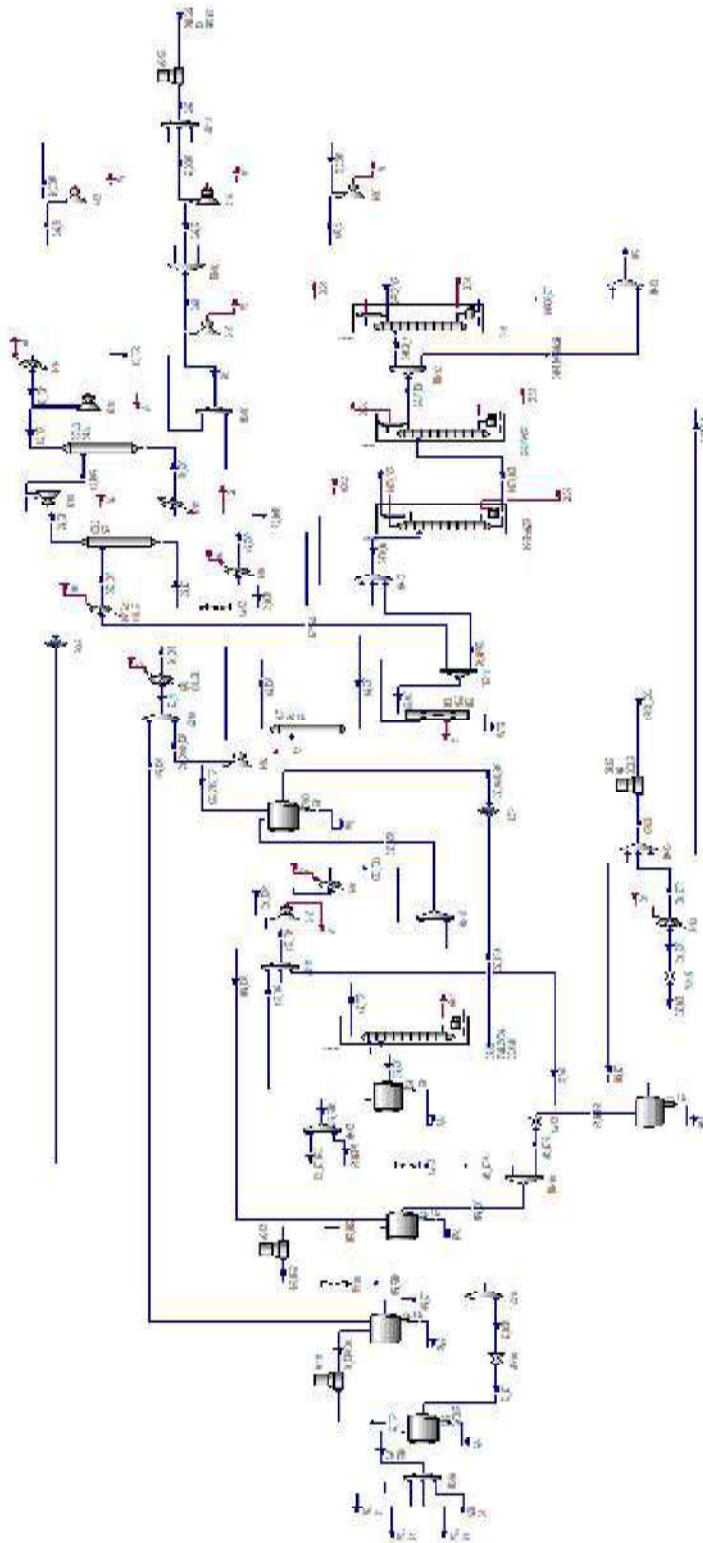


Figure 21 Process Simulation

Chapter 8

8.4.1 Enlarged Simulation

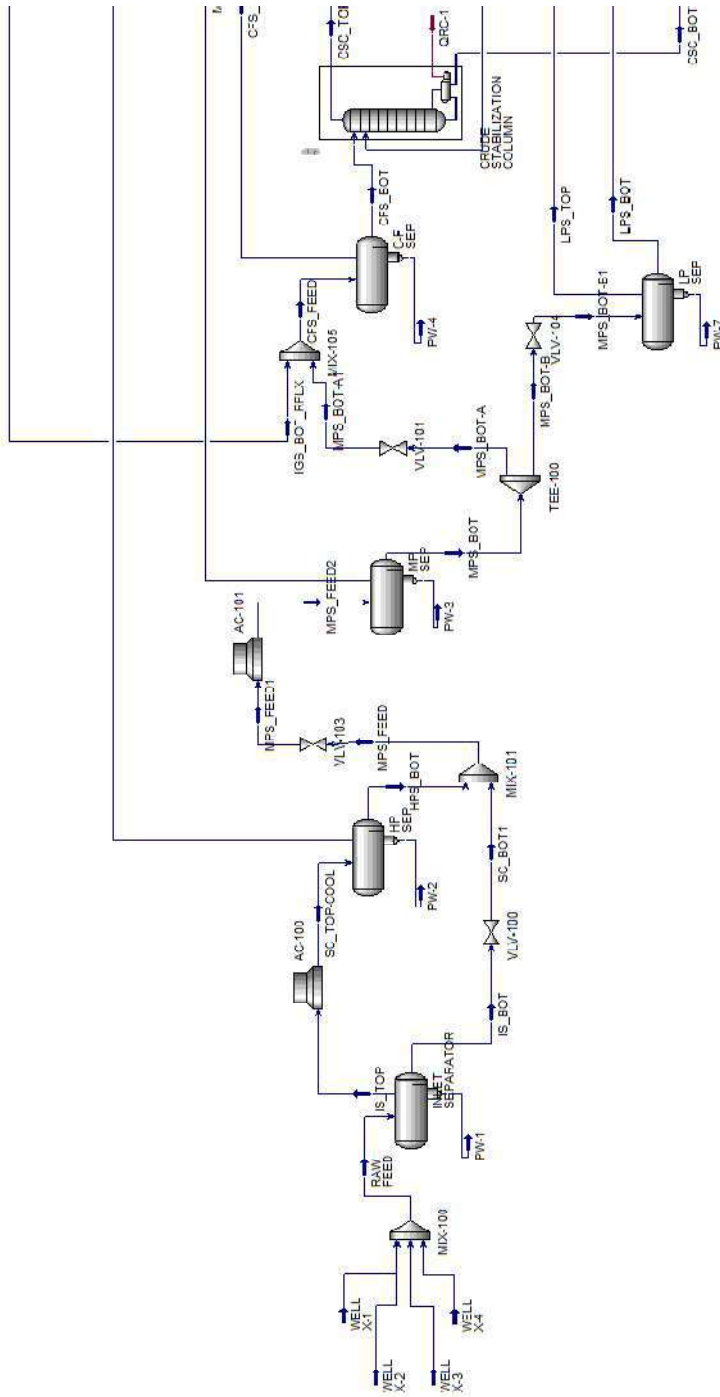


Figure 21.1 Process Simulation

Chapter 8

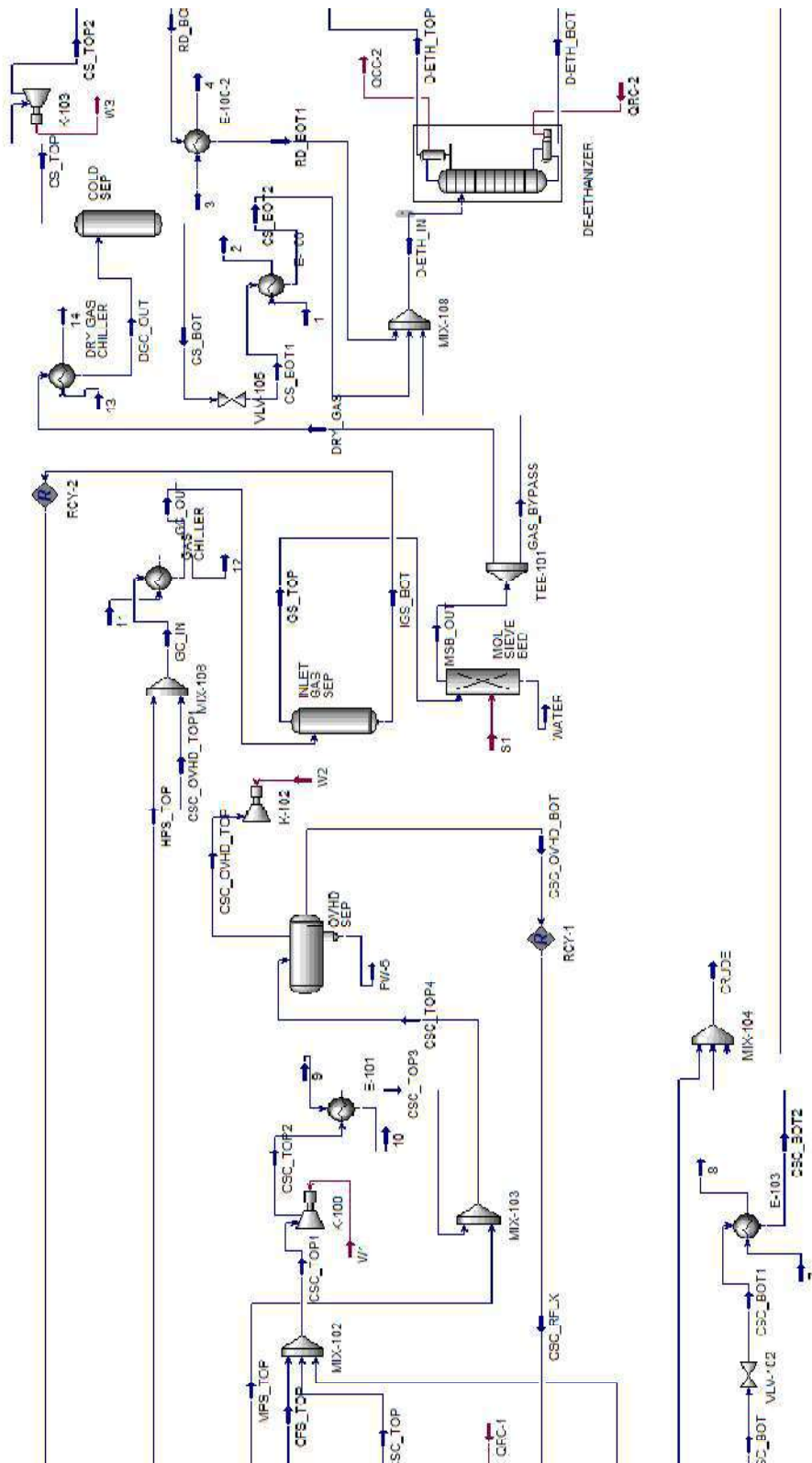


Figure 21.2 Process Simulation

Chapter 8

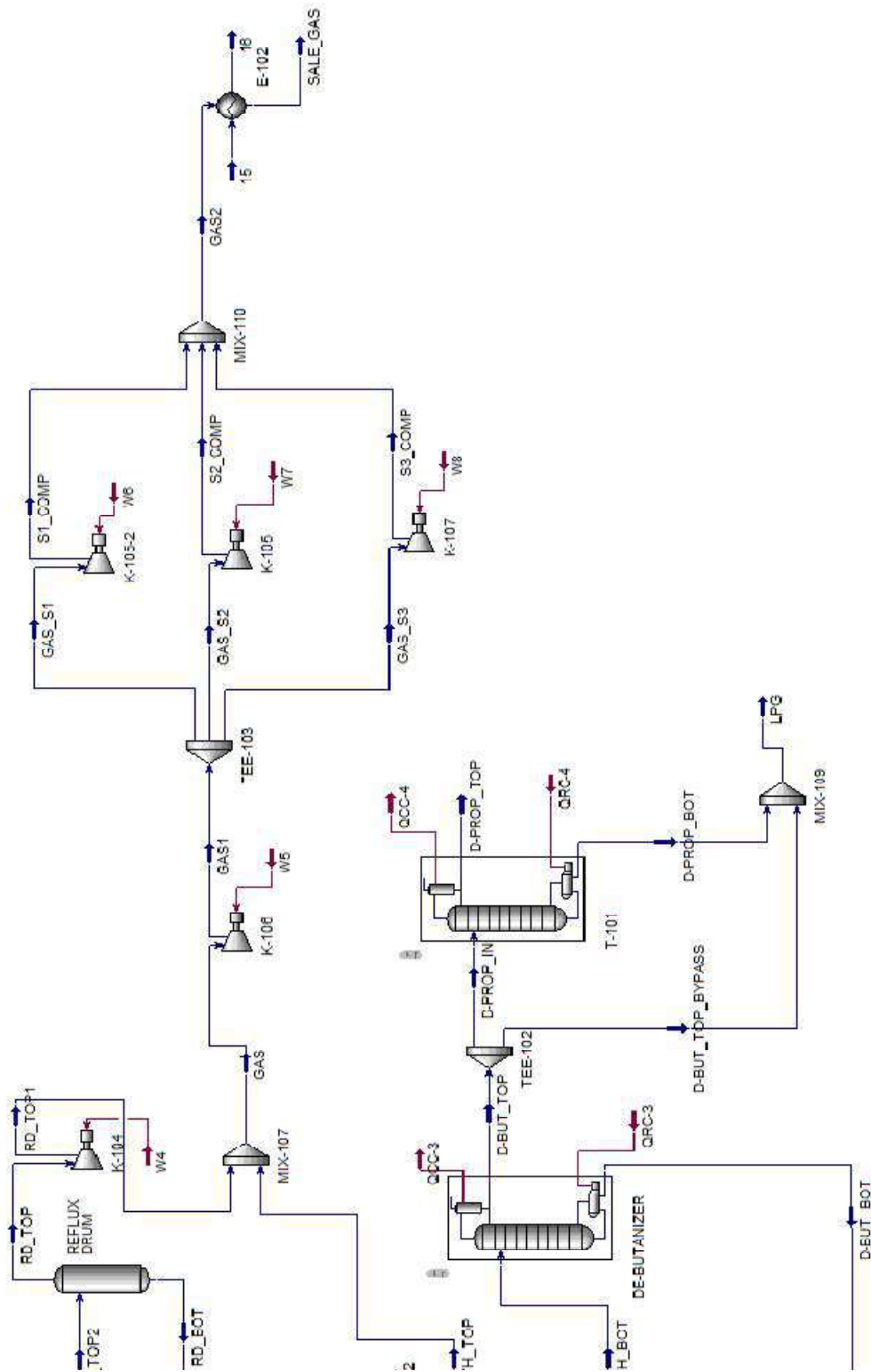


Figure 21.3 Process Simulation

Chapter 8

8.5 Working Parameters

8.5.1 Inlet Separator

3 Phase Separator: SLUG CATCHER

Design Reactions Rating Worksheet Dynamics

Worksheet	Name	RAW FEED	SC_BOT	SC_TOP	PW-1
Conditions	Vapour	0.7696	0.0090	1.0000	0.0000
Properties	Temperature [F]	149.5	149.5	149.5	149.5
Composition	Pressure [psig]	1045	1045	1045	1045
PF Specs	Molar Flow [lbmole/hr]	9830	1757	7566	507.7
	Mass Flow [lb/hr]	3.490e+005	1.895e+005	1.512e+005	9148
	Std Ideal Liq Vol Flow [USGPM]	1451	552.1	880.3	18.31
	Molar Enthalpy [Btu/lbmole]	-5.136e+004	-9.905e+004	-3.557e+004	-1.217e+005
	Molar Entropy [Btu/lbmole-F]	40.78	64.63	36.95	15.10
	Heat Flow [Btu/hr]	-5.049e+008	-1.740e+008	-2.691e+008	-6.176e+007

Delete OK Ignored

8.5.2 High Pressure Separator

3 Phase Separator: HP SEP

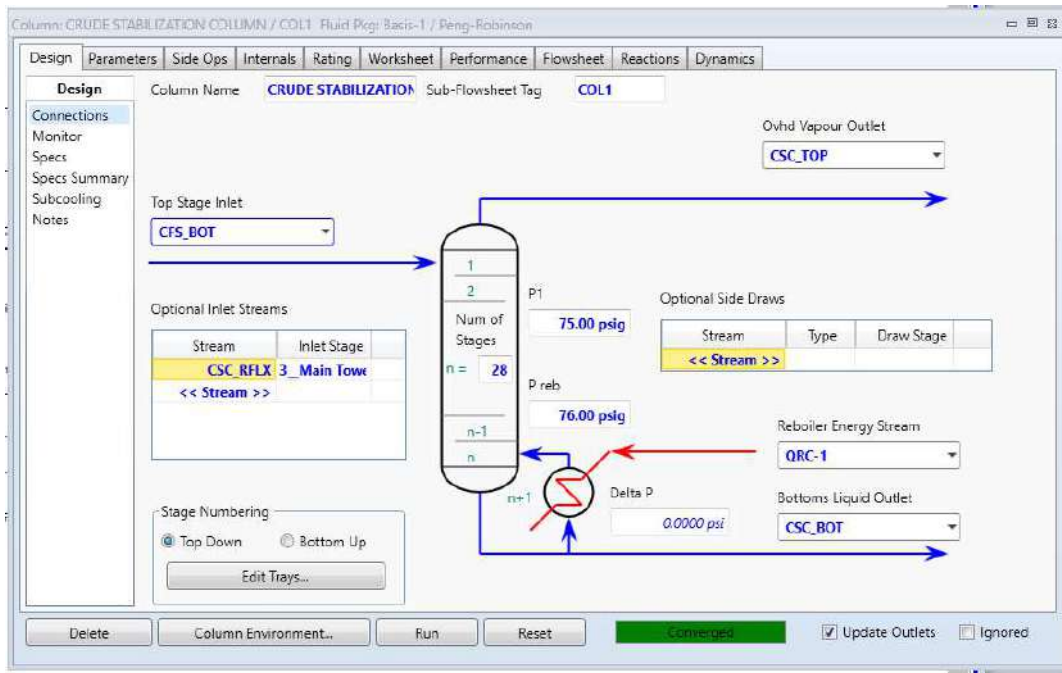
Design Reactions Rating Worksheet Dynamics

Worksheet	Name	SC_TOP-COOL	HPS_BOT	HPS_TOP	PW-2
Conditions	Vapour	0.9931	0.0000	1.0000	0.0000
Properties	Temperature [F]	100.0	100.0	100.0	100.0
Composition	Pressure [psig]	1040	1040	1040	1040
PF Specs	Molar Flow [lbmole/hr]	7566	28.03	7513	24.26
	Mass Flow [lb/hr]	1.512e+005	1859	1.489e+005	437.2
	Std Ideal Liq Vol Flow [USGPM]	880.3	6.169	873.2	0.8751
	Molar Enthalpy [Btu/lbmole]	-3.627e+004	-7.077e+004	-3.586e+004	-1.226e+005
	Molar Entropy [Btu/lbmole-F]	35.76	31.33	35.85	13.59
	Heat Flow [Btu/hr]	-2.744e+008	-1.984e+006	-2.694e+008	-2.974e+006

Delete OK Ignored

Chapter 8

8.5.3 Crude Stabilization Column



Column: CRUDE STABILIZATION COLUMN / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Worksheet Parameters:

Name	CFS_BOT @COL1	CSC_RFLX-2 @COL1	CSC_TOP @COL1	CSC_BOT @COL1
Vapour	0.0000	0.0000	1.0000	0.0000
Temperature [F]	106.9	120.9	192.9	442.6
Pressure [psig]	110.0	330.0	75.00	76.00
Molar Flow [lbmole/hr]	687.9	8048	8225	510.0
Mass Flow [lb/hr]	9.363e+004	4.690e+005	4.780e+005	8.470e+004
Std Ideal Liq Vol Flow [USGPM]	259.4	1627	1660	226.2
Molar Enthalpy [Btu/lbmole]	-1.222e+005	-6.361e+004	-5.355e+004	-1.099e+005
Molar Entropy [Btu/lbmole-F]	67.12	23.58	40.56	127.8
Heat Flow [Btu/hr]	-8.405e+007	-5.119e+008	-4.405e+008	-5.604e+007

Simulation Status: Converged

Chapter 8

Column: CRUDE STABILIZATION COLUMN / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Design Parameters Side Cps Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Design

Connections Monitor Specs Specs Summary Subcooling Notes

Specs Summary

	Specified Value	Active	Current	Fixed/Range	Prim/Alt	Lower	Upper
Btms Prod Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Boilup Ratio	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Temperature	125.0	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Temperature - 2	350.0	<input checked="" type="checkbox"/>	<input checked="" type="checkbox"/>	Fixed	Primary	<empty>	<empty>
CSC_TOP Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Column: CRUDE STABILIZATION COLUMN / COL1 Fluid Pkg: Basis-1 / Peng-Robinson

Design Parameters Side Cps Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Worksheet

Conditions Properties Compositions PF Specs

	CFS_BOT	CSC_RFLX-2	CSC_TOP
CO2	0.0022	0.0041	0.0042
Nitrogen	0.0000	0.0002	0.0002
Methane	0.0254	0.0793	0.0797
Ethane	0.0302	0.0585	0.0598
Propane	0.0595	0.1241	0.1264
i-Butane	0.0240	0.0722	0.0727
n-Butane	0.0536	0.1990	0.1992
i-Pentane	0.0350	0.2260	0.2240
n-Pentane	0.0377	0.2178	0.2157
n-Hexane	0.0597	0.0057	0.0054
n-Heptane	0.1019	0.0059	0.0057
n-Octane	0.1118	0.0037	0.0035
n-Nonane	0.0735	0.0012	0.0012
n-Decane	0.0575	0.0005	0.0005
n-C11	0.0436	0.0002	0.0002
n-C12	0.0351	0.0001	0.0001
n-C13	0.0357	0.0000	0.0000
n-C14	0.0289	0.0000	0.0000
n-C15	0.0296	0.0000	0.0000
n-C16	0.0225	0.0000	0.0000
n-C17	0.0188	0.0000	0.0000
n-C18	0.0177	0.0000	0.0000
n-C19	0.0149	0.0000	0.0000
n-C20	0.0124	0.0000	0.0000
n-C21	0.0110	0.0000	0.0000
n-C22	0.0096	0.0000	0.0000
n-C23	0.0082	0.0000	0.0000
n-C24	0.0075	0.0000	0.0000
n-C25	0.0067	0.0000	0.0000
n-C26	0.0059	0.0000	0.0000
n-C27	0.0055	0.0000	0.0000
n-C28	0.0051	0.0000	0.0000
n-C29	0.0044	0.0000	0.0000
n-C30	0.0041	0.0000	0.0000
H2O	0.0010	0.0015	0.0015

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Chapter 8

8.5.4 Inlet 2-Phase Separator

Separator: INLET GAS SEP

Design Reactions Rating Worksheet Dynamics

Worksheet

Name	GC_OUT	IGS_BOT	IGS_TOP
Vapour	0.9875	0.0000	1.0000
Temperature [F]	70.00	70.00	70.00
Pressure [psig]	1025	1025	1025
Molar Flow [lbmole/hr]	8199	102.1	8097
Mass Flow [lb/hr]	1.681e+005	4855	1.632e+005
Std Ideal Liq Vol Flow [USGPM]	966.4	17.62	948.8
Molar Enthalpy [Btu/lbmole]	-3.680e+004	-6.523e+004	-3.644e+004
Molar Entropy [Btu/lbmole-F]	35.00	25.60	35.12
Heat Flow [Btu/hr]	-3.017e+008	-6.662e+006	-2.950e+008

Delete OK Ignored

8.5.5 Cooler

Cooler: GAS CHILLER

Design Rating Worksheet Performance Dynamics

Design

Connections Parameters User Variables Notes

Delta P: 5.000 psi

Delta T: -47.57 F

Duty: 5.705e+006 Btu/hr

Delete OK Ignored

Chapter 8

Worksheet: GAS CHILLER

Worksheet	Name	GC_IN	GC_OUT	Q4
Conditions	Vapour	1.0000	0.9875	<empty>
Properties	Temperature [F]	117.6	70.00	<empty>
Composition	Pressure [psig]	1030	1025	<empty>
PF Specs	Molar Flow [lbmole/hr]	8199	8199	<empty>
	Mass Flow [lb/hr]	1.681e+005	1.681e+005	<empty>
	Std. Ideal Liq. Vol. Flow [USGPM]	956.4	956.4	<empty>
	Molar Enthalpy [Btu/lbmole]	-3.610e+004	-3.680e+004	<empty>
	Molar Entropy [Btu/lbmole-F]	36.25	35.00	<empty>
	Heat Flow [Btu/hr]	-2.960e+008	-3.017e+008	5.705e+006

Buttons: Delete, OK, Ignored

8.5.6 Compressor

Compressor: K-102

Design

Efficiency

Adiabatic Efficiency	75.000
Polytropic Efficiency	77.165

Polytropic Method

- Schultz
- Huntington
- Reference

Duty: 431.797 hp

Operating Mode

- Centrifugal
- Reciprocating
- Screw
- Wet gas

Curve Input Option

- Single-MW
- Multiple MW
- Multiple IGV
- Non-Dimensional
- Quasi-Dimensionless

Pressure Specs

Delta P: 700.0 psi Pressure Ratio: 3.031

Surge Analysis

To study compressor surge under different emergency scenarios, click the "Surge Analysis" button here.

Buttons: Delete, OK, On, Ignored

Chapter 8

Compressor: K-102

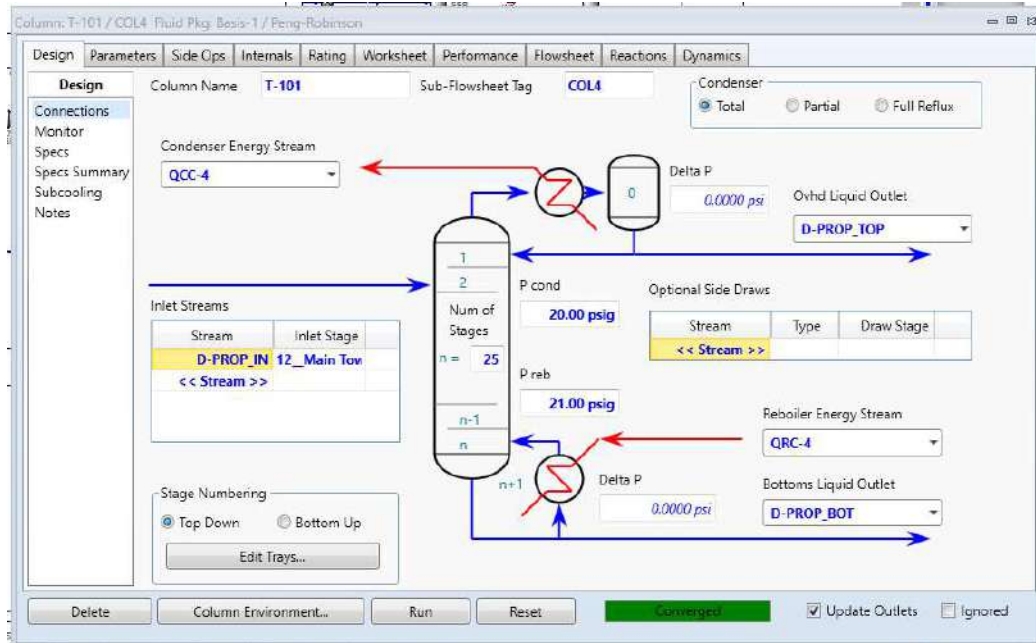
Design Rating Worksheet Performance Dynamics

Worksheet

	CSC_OVHD_TOP	CSC_OVHD_TOP1	W2
Name	1.0000	1.0000	<empty>
Vapour	1.0000	1.0000	<empty>
Temperature [F]	120.9	256.6	<empty>
Pressure [psig]	330.0	1030	<empty>
Molar Flow [lbmole/hr]	686.2	686.2	<empty>
Mass Flow [lb/hr]	1.915e+004	1.915e+004	<empty>
LiqVol Flow [USGPM]	93.22	93.22	<empty>
Molar Enthalpy [Btu/lbmole]	-4.035e+004	-3.875e+004	<empty>
Molar Entropy [Btu/lbmole-F]	38.97	39.53	<empty>
Heat Flow [Btu/hr]	-2.769e+007	-2.659e+007	1.099e+006

Delete On Ignored

8.5.7 De-Propanizer



Chapter 8

Column: T-101 / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Design Parameters Side Ops Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Worksheet

Name	D-PROP_IN @COL4	D-PROP_TOP @COL4	D-PROP_BOT @COL4
Vapour	0.0000	0.0000	0.0000
Temperature [F]	128.6	-27.86	59.08
Pressure [psig]	235.0	20.00	21.00
Molar Flow [lbmole/hr]	269.0	182.4	86.54
Mass Flow [lb/hr]	1.275e+004	7835	4920
Std Ideal Liq Vol Flow [USGPM]	49.07	31.86	17.21
Molar Enthalpy [Btu/lbmole]	-5.322e+004	-5.364e+004	-6.387e+004
Molar Entropy [Btu/lbmole-F]	25.33	18.35	15.59
Heat Flow [Btu/hr]	-1.432e+007	-9.786e+006	-5.527e+006

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Column: T-101 / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Design Parameters Side Ops Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Design

Connections Monitor Specs Specs Summary Subcooling Notes

Specs Summary

	Specified Value	Active	Current	Fixed/Range	Prim/Alt	Lower	Upper
Reflux Ratio	0.5000	<input checked="" type="checkbox"/>	<input checked="" type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Distillate Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Reflux Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Btms Prod Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Comp Recovery	0.9500	<input checked="" type="checkbox"/>	<input checked="" type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Comp Fraction	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>

Delete Column Environment... Run Reset Converged Update Outlets Ignored

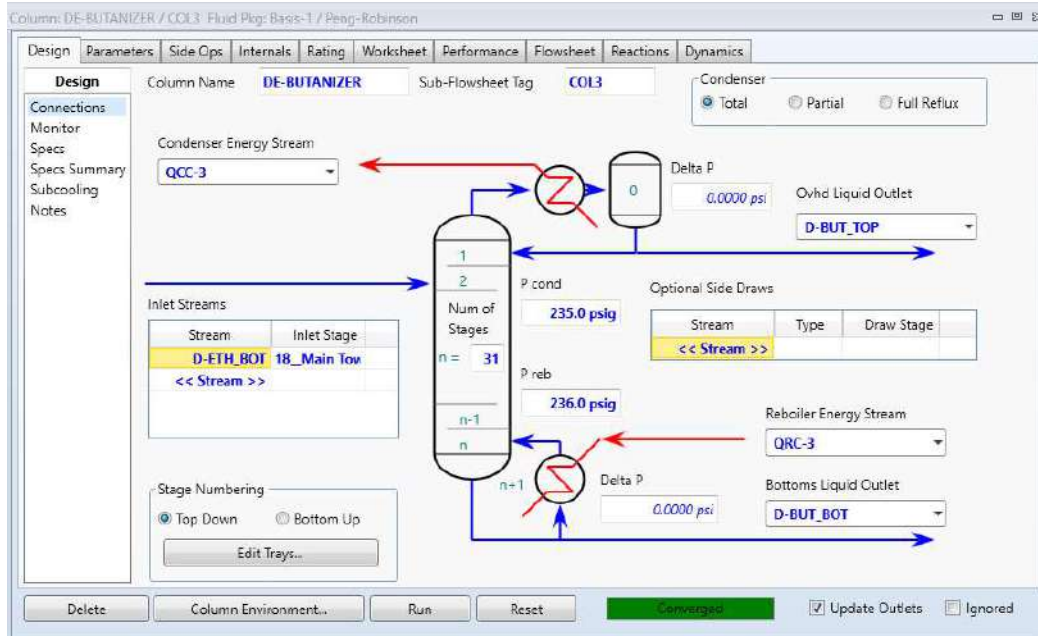
Chapter 8

Column: T-101 / COL4 Fluid Pkg: Basis-1 / Peng-Robinson

Worksheet		D-PROP_IN	D-PROP_TOP	D-PROP_BOT
Conditions	CO2	0.0000	0.0000	0.0000
Properties	Nitrogen	0.0000	0.0000	0.0000
Compositions	Methane	0.0000	0.0000	0.0000
PF Specs	Ethane	0.0877	0.1293	0.0000
	Propane	0.5880	0.8236	0.0913
	i-Butane	0.1172	0.0387	0.2826
	n-Butane	0.2069	0.0084	0.6256
	i-Pentane	0.0001	0.0000	0.0005
	n-Pentane	0.0000	0.0000	0.0000
	n-Hexane	0.0000	0.0000	0.0000
	n-Heptane	0.0000	0.0000	0.0000
	n-Octane	0.0000	0.0000	0.0000
	n-Nonane	0.0000	0.0000	0.0000
	n-Decane	0.0000	0.0000	0.0000
	n-C11	0.0000	0.0000	0.0000
	n-C12	0.0000	0.0000	0.0000
	n-C13	0.0000	0.0000	0.0000
	n-C14	0.0000	0.0000	0.0000
	n-C15	0.0000	0.0000	0.0000
	n-C16	0.0000	0.0000	0.0000
	n-C17	0.0000	0.0000	0.0000

Buttons: Delete, Column Environment..., Run, Reset, Converged, Update Outlets, Ignored

8.5.8 De-Butanizer



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Column: DE-BUTANIZER / COL3 - Fluid Pkg: Basic-1 / Peng-Robinson

Design Parameters Side Ops Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Worksheet

Name	D-ETH_BOT @COL3	D-BUT_TOP @COL3	D-BUT_BOT @COL3
Vapour	0.0000	0.0000	0.0000
Temperature [F]	212.0	128.6	320.3
Pressure [psig]	440.0	235.0	236.0
Molar Flow [lbmole/hr]	581.2	489.1	92.17
Mass Flow [lb/hr]	3.026e+004	2.319e+004	7071
Std Ideal Liq Vol Flow [USGPM]	111.3	89.22	22.12
Molar Enthalpy [Btu/lbmole]	-5.373e+004	-5.322e+004	-6.709e+004
Molar Entropy [Btu/lbmole-F]	30.46	25.33	35.12
Heat Flow [Btu/hr]	-3.123e+007	-2.603e+007	-6.184e+006

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Column: DE-BUTANIZER / COL3 - Fluid Pkg: Basic-1 / Peng-Robinson

Design Parameters Side Ops Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Design

Connections Monitor Specs Specs Summary Subcooling Notes

Specs Summary

	Specified Value	Active	Current	Fixed/Range	Prim/Alt	Lower	Upper
Reflux Ratio	1.500	<input checked="" type="checkbox"/>	<input checked="" type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Reflux Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Btms Prod Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Distillate Rate	<empty>	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Comp Recovery	0.9900	<input checked="" type="checkbox"/>	<input checked="" type="checkbox"/>	Fixed	Primary	<empty>	<empty>
Temperature	350.0	<input type="checkbox"/>	<input type="checkbox"/>	Fixed	Primary	<empty>	<empty>

Delete Column Environment... Run Reset Converged Update Outlets Ignored

Chapter 8

Column: DE-BUTANIZER / COL3 Fluid Pkg: Basis-1 / Peng-Robinson

Design Parameters Side Ops Internals Rating Worksheet Performance Flowsheet Reactions Dynamics

Worksheet

		D-ETH_BOT	D-BUT_TOP	D-BUT_BOT
Conditions	CO2	0.0000	0.0000	0.0000
Properties	Nitrogen	0.0000	0.0000	0.0000
Compositions	Methane	0.0000	0.0000	0.0000
PF Specs	Ethane	0.0738	0.0877	0.0000
	Propane	0.4947	0.5880	0.0000
	i-Butane	0.0988	0.1172	0.0008
	n-Butane	0.1768	0.2069	0.0166
	i-Pentane	0.0667	0.0001	0.4200
	n-Pentane	0.0345	0.0000	0.3436
	n-Hexane	0.0197	0.0000	0.1240
	n-Heptane	0.0112	0.0000	0.0705
	n-Octane	0.0033	0.0000	0.0209
	n-Nonane	0.0005	0.0000	0.0031
	n-Decane	0.0001	0.0000	0.0005
	n-C11	0.0000	0.0000	0.0001
	n-C12	0.0000	0.0000	0.0000
	n-C13	0.0000	0.0000	0.0000
	n-C14	0.0000	0.0000	0.0000
	n-C15	0.0000	0.0000	0.0000
	n-C16	0.0000	0.0000	0.0000
	n-C17	0.0000	0.0000	0.0000

Delete Column Environment... Run Reset Converged Update Cutlets Ignored

Chapter 9

HAZOP ANALYSIS

Hazard and Operability study (HAZOP) is a procedure that systematically and critically examines the operability of a process. Its application to an operating plant gives indication of potential hazards that arise because of deviation from intended design. At the flow-sheet stage, it can be used for preliminary examination of the design and for a detailed study at a later stage.

An Operability study examines the design line by line, and vessel by vessel and uses guidewords to produce thought about the way deviances from planned design can lead to hazardous situations.

Keywords

The following are some of the keywords that are used to carry out HAZOP studies.

No	Keyword	Explanation
1	Intentions	It defines how a specific part of the process was planned to operate.
2	Deviations	These are the aberrations from the designer's intention and are detected using guide words.
3	Causes	These are the reasons that triggered deviations. A deviation is treated as meaningful only if it has a realistic cause.
4	Consequences	Following results from occurrence of meaningful deviances.
5	Hazards	Consequences that can bring about damage or injury.
6	Guide words	These are the words that qualities or quantifies the intention to guide and bring about the thought process so that deviation can be discovered.

Table 28 HAZOP keywords

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Guide Words

Following are some of the guide words that are used most often in HAZOP analysis.

No	Keyword	Explanation
1	No	Negation of the design intent
2	Less	Quantitative decrease
3	More	Quantitative increase
4	Part of	As well as
5	As well as	Logical opposite of the intent
6	Reverse	Reverse
7	Other than	Complete substitution

Table 29 guide words

Guide words application covers both general (react, transfer) and specific (pressure, temperature) parameters. When dealing with a design intention involving a complex set of interrelated plant parameters (e.g., temperatures, reaction rates, composition, or pressure), it may be better to apply the whole sequence of guide words to each parameter individually than to apply each guide word across all the parameters as a group.

Procedure

HAZOP study is usually conducted by a team of experienced professionals, who have complementary skills and comprehensive understanding of the technique and led by an experienced leader.

The team carries out comprehensive study of the process vessel by vessel, and line by line, and then use guide words to detect any hazards. Extent of information depends upon the information required for study.

Description of process and process flowsheets can be used for the preliminary study of Plant. For a detailed, final, study of the design, the flowsheets, piping, and instrument diagram, equipment specifications and layout drawings would be needed.

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Sequence of HAZOP study

Flowing are the steps used to conduct HAZOP study:

1. Selection of the vessel
2. Explanation of intention of vessel and its line
3. Selection of a line
4. Application of guide word
5. Development of meaningful cause
6. Examination of possible causes
7. Examination of possible consequences¹
8. Detection of hazards
9. Making of suitable records

9.1 Three Phase Separator (HP, MP, LP, Inlet Separator)

Study Node	Parameters	Guide Word	Possible Causes	Possible Consequences	Action Required
Feed Inlet line	Flow	No	1. Line blockage for rupture	1. No separation	1. Install no flow alarms. 2. Check flow meters regularly
		High	1. Valve fully opened	1. Separator will overflow due to accumulation	1. Install alarms to prevent flooding of separator 2. Regularly monitor the

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					inlet flow rates
	Temperature	High	1. Air cooler not cooling properly	1. Undesired separation achieved	2. Install temperature alarm
	Pressure	High	Outline blockage of the three phase separator	1. Improper separation 2. Vessel may leak or explode	1. Install high pressure alarm 2. Install PSVs'
		Low	Leakage in the lines	1. Improper separation 2. Low Pressure towards downstream equipment's	1. Install low pressure indicators 2. Regular maintenance to check leakage

Table 30 3-phase separator HAZOP

Chapter 9

9.2 Crude Stabilization column

Study Node	Parameter	Guide word	Possible Cause	Consequence	Action
Feed Inlet	Level	More	<ol style="list-style-type: none"> 1. Level alarm failure 2. Oil line to storage blocked 	<ol style="list-style-type: none"> 1. No production to storage 2. Column flooding 	<ol style="list-style-type: none"> 1. High level alarm with greater sensitivity.
		LESS	<ol style="list-style-type: none"> 1. Level alarm failure 2. Low oil input 	<ol style="list-style-type: none"> 1. Gas to storage release 2. No feed to heat exchanger 3. Poor crude stabilization. 	<ol style="list-style-type: none"> 1. Low level alarm be installed with greater sensitivity.

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Flow	MORE	<ol style="list-style-type: none"> 1. Bypass open 2. Increased temperature 	<ol style="list-style-type: none"> 1. Poor crude stabilization 2. Damage to compressors 3. Pressure increase 4. Pipeline fracture 5. Gas leakage 6. Fire and explosion hazard 	<ol style="list-style-type: none"> 1. Pressure alarm installed 2. PSV installed 3. Explosive alarm
	LESS	<ol style="list-style-type: none"> 1. Low oil flow from CF separator 	<ol style="list-style-type: none"> 1. Column disks not working properly 2. Poor stabilization 	<ol style="list-style-type: none"> 1. Install flow alarm for crude input flow 2. Install pressure alarm.
	REVERSE	<ol style="list-style-type: none"> 1. Pipeline fracture 	<ol style="list-style-type: none"> 1. Pipeline fracture 2. Oil spill 3. Fire and explosion hazard. 	<ol style="list-style-type: none"> 1. Explosive alarm. 2. Check Valve installed

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		NO	<ol style="list-style-type: none"> 1. No oil flow from CF separator 2. Pipeline to storage blocked 	<ol style="list-style-type: none"> 1. No production 	<ol style="list-style-type: none"> 1. Install flow alarm at feed point. 2. Install flow alarm at exit point.
	Pressure	MORE	<ol style="list-style-type: none"> 1. Column tray disks clogged 2. Output line clogged 	<ol style="list-style-type: none"> 1. No crude stabilization 2. Pipeline fracture 3. Oil spill 4. Fire and explosion hazard 	<ol style="list-style-type: none"> 1. Pressure alarm installed 2. PSV checked and installed 3. Explosive alarm
		LESS	<ol style="list-style-type: none"> 1. Pipeline fracture 	<ol style="list-style-type: none"> 1. Pipeline fracture 2. Oil spill 3. Fire and explosion hazard 	<ol style="list-style-type: none"> 1. Low pressure alarm installed 2. Explosive alarm
	Temperature	MORE	<ol style="list-style-type: none"> 1. Air cooler not working properly 	<ol style="list-style-type: none"> 1. More crude vapor in gas stream 2. Increased pressure 	<ol style="list-style-type: none"> 1. Temperature sensitive alarm be installed

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		LESS	1. Air cooler flow valve failure	1. Poor crude stabilization	1. Temperature sensitive sensor should be installed.
	Containment	NO	1. Fractured vessel 2. Vessel open to atmosphere	1. Oil spill 2. Fire hazard 3. Explosion hazard	1. Install flow sensor and level sensor in vessel 2. Allow only qualified personal to do maintenance work

Table 31 Crude Stabilization Column HAZOP

9.3 2-phase separator

Study Node	Parameter	Guide Word	Possible Causes	Possible Consequences	Action Required
Feed Inlet	Flow	No	1. Inlet valve closed 2. Leakage in the lining	1. No separation	1. Regular maintenance of linings 2. Ensure valve is operational

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		High	1. Valve fully opened	Separator will overflow due to accumulation	1. Install alarms to prevent flooding of separator 2. Regularly monitor the inlet flow rates
		Low	1. Valve opening is below required 2. Leakage in the linings	Desired output unattained	1. Install flowmeters to accurately measure flow rates
	Pressure	High	1. Blockage in pipe 2. High temperature leading to high pressure	1. Improper separation 2. Separator may leak or explode	1. Install high pressure alarms 2. Install PSV'S
		Low	Leakage in the lines	1. Improper separation 2. Low pressure towards downstream equipment	1. Install low pressure indicators 2. Regular maintenance to check leakages.

Table 32 2-Phase Separator HAZOP

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9.4 Distillation Column (De-Ethanizer, De-Butanizer, De-Propanizer)

Study Node	Parameter	Guide Word	Possible Causes	Possible Consequence	Action Required
Feed Inlet	Flow	No	<ol style="list-style-type: none"> 1. Pipe blockages 2. Control valve shut 3. Valves fail 4. Tube leakages and blocking 	<ol style="list-style-type: none"> 1. Column dry out 2. Possible dangerous concentration 3. No operation 	<ol style="list-style-type: none"> 1. Install low level alarm 2. Check maintenance procedure and schedule 3. Make bypass 4. Emergency plant shut down
		Less	<ol style="list-style-type: none"> 1. Pipe blockages 2. Control valve shut 3. Valves fail 4. Tube leakages and blocking 	<ol style="list-style-type: none"> 1. Column dry out 2. Changes in product quality 	<ol style="list-style-type: none"> 1. Install low level alarm 2. Check maintenance procedure and schedule 3. Make bypass 4. Emergency plant shut down

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Level	More	<ol style="list-style-type: none"> 1. Control valve is fully opened 2. Increase flow capacity 3. Control valve failure 	<ol style="list-style-type: none"> 1. Flooding in the column 2. Changes in product quality 3. Temperature decrease 4. Rise in bottom 	<ol style="list-style-type: none"> 1. Install high level alarm 2. Install control 3. Check maintenance procedure and schedule
	High	<ol style="list-style-type: none"> 1. Output blockages 	<ol style="list-style-type: none"> 2. Over pressure in column 3. Condensed liquid flow back to distillation 	<ol style="list-style-type: none"> 4. Install high level alarm 5. Check maintenance procedure and schedule
	Low	<ol style="list-style-type: none"> 1. Pipe partial clogged and leakage 	<ol style="list-style-type: none"> 1. Level decrease in the level 2. The valve closed 3. Back flow of material 	<ol style="list-style-type: none"> 1. Install low level alarm 2. Check maintenance procedure and schedule 3. Install valve

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	Temperature	High	<ol style="list-style-type: none"> 1. Heater control failure 2. Heaters operating at more than required energy 	<ol style="list-style-type: none"> 1. Increased vapor content 2. Increased pressure 3. More crude vapor in top product 	<ol style="list-style-type: none"> 1. Install high temperature alarm 2. Check maintenance procedure and schedule
		Low	<ol style="list-style-type: none"> 1. Heater no working at optimized value 	<ol style="list-style-type: none"> 1. Poor distillation separation 2. Tops product condensed into bottom product 	<ol style="list-style-type: none"> 1. Install low temperature alarm 2. Check maintenance procedure and schedule

Table 33 Distillation Column HAZOP

9.5 Heat Exchanger

Study Node	Parameter	Guide Word	Causes	Consequences	Action
Feed Inlet Line	Flow	No	1) Cooling water valve blocked completely	1) Desired temperate is not achieved on shell side fluid	Run Maintenance procedures to remove any

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Temperature			2) Valve closed		blockages.
	Less		1) Blockage of pipe	1) Desired temperate is not achieved on shell side fluid	Indication by higher temperature alarm. Install backup inlet for cold water
	More		1) Cooling water valve failure	Decrease in process fluid temperature beyond required	Indication by low temperature alarm.
	Less		1) Reduced flowrate of process fluid 2) More cold-water flowrate than required	1)Higher outlet temperature on shell side. 2) Tubes and baffles can be corroded from outside by hot fluid	1)Check cold water valve for any problems 2) Check purge stream, line for any blockages
	More		1)Fouling in exchanger 2) Insufficient cold-water flowrate	1)Higher outlet temperature on shell side. 2) Tubes and baffles can be corroded from both sides	1)Maintenance of Heat exchanger to remove any fouling 2)Check cold water supply valve and line for

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					blockages
	Pressure	More	1) Failure of process fluid valve	Tubes can burst	Indication by higher pressure alarm.
	Composition	As well as	1) Tube Leakage 2) Corrosion	1) Shell and tube side fluids get mixed. 2) Corroded tube material contaminates shell side fluid.	1) Perform maintenance of tubes 2) Replace tubes in case of corrosion

Table 34 Heat Exchanger HAZOP

9.6 Potential process modifications

The potential process modifications resulting from this study are the following:

- ❖ Install Flow meter with high/ flow alarm on all the stream
- ❖ Check and update coil of heat exchanger
- ❖ Install High Pressure Alarm and PSV
- ❖ Install Filter in Pipes to ensure there is no plugging
- ❖ Install a high temperature shutdown system (this system would automatically shut down the process in the event of a high equipment temperature)
- ❖ Install a high pressure shut down system
- ❖ Use welded pipes instead of flanged pipes

Conclusion

As stated above, the objective of this project was to deliver a simulation model of a gas processing plant with the following criteria's:

1. Methane recovery 90%
2. Propane recovery 96% and
3. Butane recovery 99%

We have developed a model that is able to sustain the high pressure of the wells without damaging the components and simultaneously installed the equipment in such an order that there is no over-pressure on the components. The addition of multiple bypasses is done to reduce the duty of the equipments thereby reducing the overall cost of the plant. Similarly multiple recycle lines are introduced to enhance the recovery of the different components present in our feed.

Some of the design specifications we have chosen was to introduce 3 separators with reducing pressures instead of a single separator. This was done to firstly, ensure significant hold up time was present to attain adequate separation and secondly, to reduce the load on the separators. Another design alteration that we carried out was the absence of sour gas treating unit/ H₂S removal unit and the mercury guard bed/removal unit. This is because since these components are absent from the reservoir and would only result in adding to the overall plant cost and the operating cost without significant usage.

We have also used the molecular sieves bed dehydration unit instead of more preferred glycol dehydration unit largely because the molecular sieves bed ensures better separation efficiencies and also ease regeneration method. Further down the line, the installation of debutanizer is done for recovering LPG from the remainder to enhance gains in profits.

Conclusion

The plant consists heavily of 4 repeated equipments:

1. 3-phase separators.
2. Distillation columns.
3. 2-phase separators.
4. Compressors.

The developed model was able to deliver the desired results and was not overly costly for installation and maintenance. No chemical reactions were carried out in the process and the entire plant is operation only on physical separation of the crude to deliver Natural gas, LPG, propane (if needed hence depropanizer is installed) and crude from the crude stabilization column.

Results

- Methane recovers 91% of the total reservoir feed.
- The LPG formed 45% propane and 49% butane (iso-butane and n-butane) which is commercially viable.
- The payout period of the plant comes out to be 3.73 years which is acceptable considering the vast plant cost and running cost.

Conclusion

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Conclusion

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