DESIGN OF A VAPOR RECOVERY UNIT FOR HYDROCARBON RECOVERY FROM STORAGE TANK AND OFF-GAS OF A CONDENSATE FRACTIONATION PLANT

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By

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CANDIDATE DECLARATION

It is hereby declared that this project or any part of it has not been submitted elsewhere for the award of any degree or diploma.

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Dedicated to

People who have supported me

throughout my life

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ABSTRACT

A condensate fractionation plant recycles vapors coming from the tank batteries and the fractionation column by a vapor recovery unit (VRU). Otherwise, these valuable hydrocarbons would be lost through flaring or venting. VRU is essentially a multi-stage compressor, with an intercooler and aftercooler. The VRU must be designed such that it is not over-sized or under-sized. The project examined an existing plant of 4,000 barrels/day capacity and attempted to re-design the VRU for a scenario where an additional 6,000 barrels/day will be added to the plant capacity. The major obstacle in this task is to correctly estimate the total volume of vapor to be handled. There is no facility for direct measurements, so the vapor must be estimated from correlations and by process simulation.

The volume of the byproducts from the column was estimated by using software, which simulated the whole fractionation process. With the help of tank data from the site, an analytical calculation was performed to compute the amount of vapor by considering different operating conditions. Following that, a compressor for the VRU system was designed which would be appropriate to handle the volume of hydrocarbon. The benefit of using VRU from an environmental point of view was included in this work. New technologies to minimize tank vapor loss, such as the construction of floating roof tanks, were also considered.

The results obtained from the calculation and simulation steps reflect actual values from the plant with minor deviation, which gave confidence in this design process. The calculations indicate that, for the proposed capacity upgradation, the required VRU compressor should have 2.65 times greater capacity in the case of fixed roof tanks. With a floating roof, the VRU would need only about 3 m3/hour additional capacity. The VRU should also prevent 103.58 tonnes/year of CO2 emission.

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ASTM	American Society for Testing and Materials
BHP	Break Horsepower
BPD	Barrel Per Day
BTU	British Thermal Unit
CFP	Condensate Fractionation Plant
E	Efficiency
EPA	Environment Protection Agency
EIA	Environmental Impact Assessment
GHG	Green House Gas
GWP	Global Warming Potential
IHP	Ideal Horsepower
IPCC	Intergovernmental Panel on Climate Change
HAP	Hazardous Air Pollutant
MMBTU	Million British Thermal Unit
MMSCFD	Million Standard Cubic Feet Per Day
MSCF	Thousand Standard Cubic Feet
MS	Motor Spirit
Mt	Million Tonne
TBP	True Boiling Point
VFD	Variable Frequency Drive
VOC	Volatile Organic Compound
VRU	Vapor Recovery Unit

Chapter 1

INTRODUCTION

Condensate Fractionation is a distillation process where a variety of required products are separated from the raw materials (condensate). In the fractionation column, a heavy part of the condensate is converted into diesel and the lighter parts are divided into Petrol, Kerosene, Octane and other hydrocarbon solvents by utilizing different relative volatility of its components [1]. “Condensate” is a hydrocarbon liquid stream separated from natural gas at the surface condition in separators, field facilities or gas-processing plants and consists of higher molecular weight of hydrocarbons. During distillation, some gases are produced as a by-product from the top of the fractionation column, which is known as off-gas. Generally, this by-product is continuously flared into the environment which produces noise, thermal radiation and air pollution.

Furthermore, in process plants, there are tanks for storing feed and products. From these tanks (specially from condensate and motor spirit tanks), a significant amount of hydrocarbon evaporates. This vapor is vented to maintain the pressure in the tank, which is a waste of valuable fuel [2-6]. Vapor Recovery Unit (VRU), which is a multistage compressor, is an option to prevent this loss and pollution.

A condensate fractionation plant (CFP) in Bangladesh, with 4,000 barrels per day (BPD) condensate handling capacity, produces Motor Spirit (Petrol), Kerosene and Diesel from condensate. Rather than flaring the off-gas from the column and the vapor from the tanks releasing to the atmosphere, in this plant Vapor Recovery Unit (VRU) is used to recover these hydrocarbons. These recovered hydrocarbons are then reused as additional fuel gas to Kerosene stripping Column in the process as well as in the gas generator feed as a utility.

There is a plan to build an additional 6,000 BPD facility to upscale the plant’s overall capacity up to 10,000 BPD. Therefore, the efficiency of the current VRU needs to be examined to see whether it can handle further hydrocarbon vapor generated in the new unit. A new VRU needs to be designed if the existing one is inadequate for the extended capacity.

1.1 Objectives with specific aims and possible outcome

1) Estimate the volume of hydrocarbon vapor lost from the storage tanks.
2) Estimate the quantity of Off-gas lost from the fractionation Column.
3) Design a Vapor Recovery Unit (VRU) to prevent the losses estimated above.
4) Assess the environmental benefits of a VRU.

1.2 Outline of Methodology

1) Collect and analyze relevant data from the existing 4000 BPD Plant.
2) Estimate the volume of vapor from the storage tanks using different correlations.
3) Estimate the amount of off-gas from the process by performing the simulation.
4) To perform these tasks, the simulation software Aspen^{TM} HYSYSv11 will be used.
5) Design a VRU system to recover both hydrocarbons i.e. tank vapor and off-gas.
6) Evaluate the amount of CO_2 emission from flaring if there would be no VRU system.
7) Estimate the heat radiation by the flaring in absence of the VRU system.

Chapter 2

PROCESS DESCRIPTION

The volume of off-gas and tank vapor for a future additional 6,000 BPD capacity unit is required for designing the VRU compressor and associated system of overall 10,000 BPD unit. Prior to that, the present plant is needed to describe and establish a method to estimate the current plant's total recovered hydrocarbon volume by VRU compressor. Based on that calculation, the total hydrocarbon amount can be quantified for an additional 6,000 BPD plant.

2.1 Existing Plant Description

The existing 4,000 BPD plant has a tank and process area. In the tank area, several feed and products are stored. From the tank, feed goes to the process area for distillation. After distillation, products are sent to the designated tanks. The vapor recovery unit located in the process area recycles the vapors to prevent losses. The detailed procedure is described below.

2.1.1 Process Description

The complete process of hydrocarbon production from condensate is shown in Figure 2.1. The condensate is pumped by pump (P-140 A/B) from the feed storage tank to Fractionation Column (C-100) through Condensate-Kerosene Heat Exchanger (E-160), Condensate-Motor Spirit Heat Exchanger (E-100), Condensate-Kerosene Heat Exchanger (E-110) and Condensate-Diesel Heat Exchanger (E-120) to achieve the feed temperature.

The feed enters at the middle section of the Fractionation Column (C-100) and then flows downwards while contacting with the vapor from the bottom of the column. Vapor is injected by a Reboiler (heater H-100) located at the bottom of the column. Then, the highly pure volatile component of liquid which is known as reflux - is injected from the top of the column.

Reflux liquid can be derived by partial condensation of the top product by a series of heat exchange inside Condensate-Column Overhead Heat Exchanger (E-160) and Air-Cooled Overhead Condenser (A-100 A/B). Vapor-liquid mixture is then separated by Reflux Accumulator (D-100). Separated gas as Off-gas will flow to the Vapor Recovery Unit (VRU) and separated liquid reflux liquid is pumped to Fractionation Column (C-100). Motor spirit is taken from Reflux Accumulator and cooled inside Condensate-Motor Spirit Heat Exchanger (E-100) and Motor Spirit Water Cooler (E-150) respectively until reaching a temperature of 40°C before going to the Motor Spirit Production Tanks (T-504, T-505 & T-506).

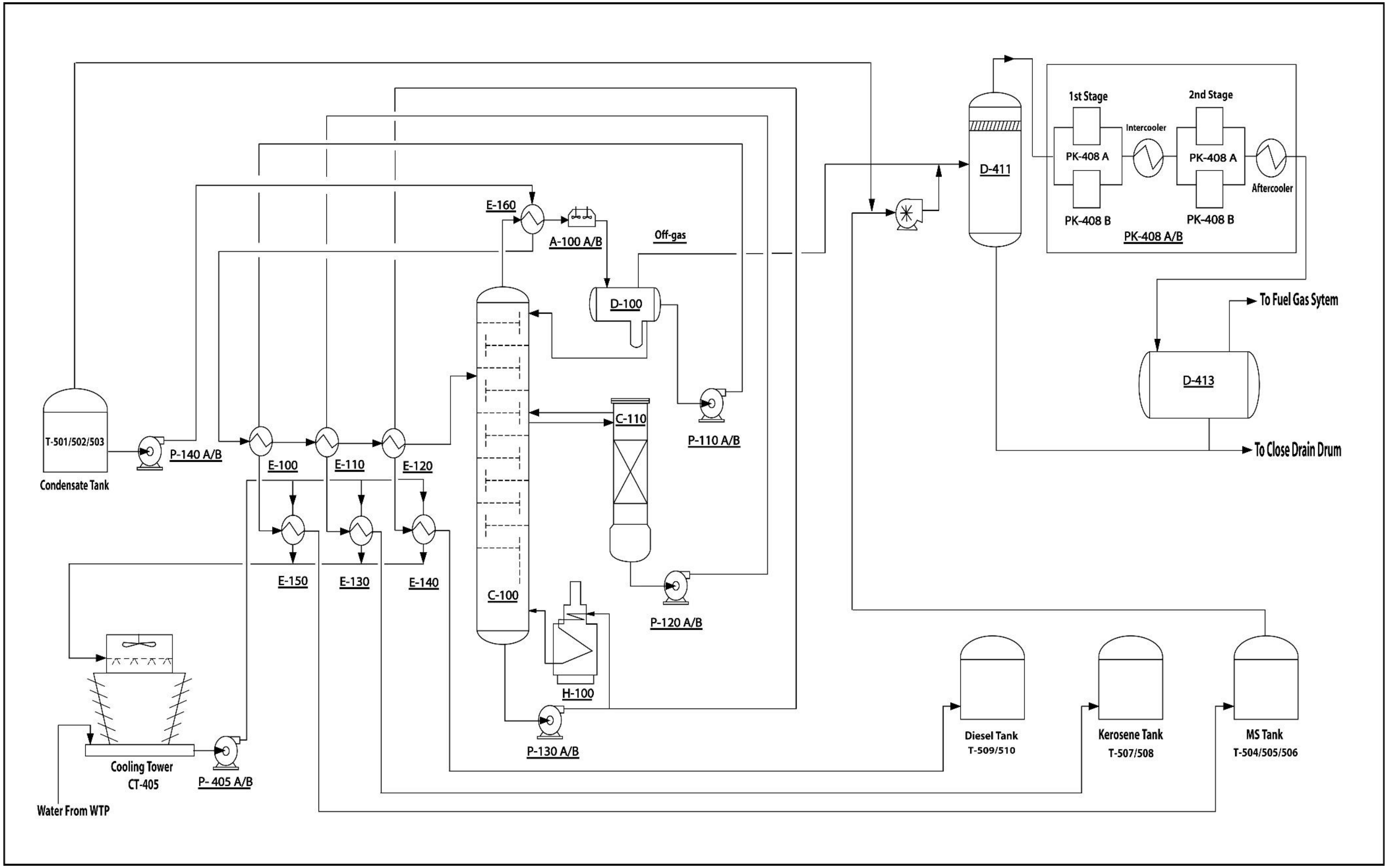


Figure 2.1: Process Flow Diagram of Existing 4,00 BPD Plant

Kerosene is taken from the side of the Fractionation Column (C-100) where the next step is routed to the top section of the Stripping Column (C-101). 35°C of stripping gas is injected from the bottom section of the Stripping Column (C-101. The interior of the Stripping Column (C-101) is packing-type, allowing direct contact between kerosene liquid and stripping gas; resulting in pure kerosene produced from the bottom of the Stripping Column (C-101) and then pumped away by Kerosene Pumps (P-120 A/B) and then cooled inside Condensate-Kerosene Heat Exchanger (E-110) and Kerosene Water Cooler (E-130) respectively until reaching a temperature of 45°C before going to the Kerosene Production Tank (T-507 & T-508). Stripping gas outlet from top of the Stripping Column (C-101) is injected back into Fractionation Column (C-100).

Diesel product is taken from the bottom of the column and pumped away by Diesel Pumps (P-130 A/B) and then routed to Condensate-Diesel Heat Exchanger (E-120) and Diesel Water Cooler (E-140) respectively until reaching temperature of 45°C before going to the Diesel Production Tank (T-509 & T-510). Some portion of diesel is also rerouted into the bottom of the Fractionation Column (C-100) through a fired heater (H-100) to generate vapor.

2.1.2 Storage Tank

There are 10 (Ten) storage tanks in the current 4,000 BPD plant for feed and product storage. The general data of the storage tank are shown in Table 2.1 below.

Table 2.1: Storage Tank General Information for 4,000 BPD Unit

Tank Description	Tank Quantity	Vapor Generation Tank
a. Vertical b. Fixed Roof c. Cone Shaped Roof	**a) Feed Tank:** 1) Condensate Storage Tank -3 EA **b) Product Tank:** 2) MS Storage Tank - 3 EA 3) Kerosene Storage Tank - 2 EA 4) Diesel Storage Tank - 2 EA	1) Condensate Storage Tank 2) MS Storage Tank

a) Feed Storage Tank

Tanks for condensate feed storage are:

1) T-501
2) T-502
3) T-503

The capacity of each tank is 8000 m^2. The vapor space above the liquid is blanketed by gas for safe operation.

The tanks are also equipped with breather valves if blanketing gas is failed. Generated vapor due to daily temperature change will flow to the Vapor Recovery Unit (VRU).

b) Product Storage Tank

Tanks for Motor Spirit storage are:

1) T-504
2) T-505
3) T-506

The capacity of each tank is 6000 m^3. The vapor space above the liquid is blanketed by gas for safe operation. The tanks are also equipped with breather valves for protection if blanketing gas is failed. Generated vapor due to daily atmospheric pressure and temperature will flow to the Vapor Recovery Unit (VRU).

Tanks for Kerosene storage are:

1) T-507
2) T-508

The capacity of each tank is 2000 m^3. The tanks are equipped with breather valves for tank protection.

Tanks for Diesel storage are:

1) T-509
2) T-510

The capacity of each tank is 1000 m^3. The tanks are equipped with breather valves for tank protection.

2.1.3 Vapor Recovery Unit (VRU)

Rather than discharged to the flare system, off-gas produced by Fractionation Column (C-100) and vapor generated in condensate and motor spirit tanks vented to the atmosphere, are recovered by Vapor Recovery Unit (VRU). It will contribute to additional fuel gas for supplying process and utility demand.

The system itself is consisting of:

1) VRU Suction Scrubber (D-411)
2) VRU Compressor Package
3) VRU Discharge Scrubber (D-413)

First, low pressure gas (off-gas & vapor from tanks) is entered into VRU Suction Scrubber for early liquid-gas (if any) separation. Separated liquid is directed to the condensate tank header. Meanwhile, the gas is routed to VRU Compressed Package for compressing and cooling. The final treatment is in VRU Discharge Scrubber for final liquid-gas (if any) separation at higher pressure where the liquid is routed to the condensate tank header and gas is directed to the fuel gas system.

2.2 Design Basis of This Project Work

Analytical calculation of off-gas volume from the column is complicated. But simulation software is widely used in refinery industries and is a powerful approach for modeling to compare with real-time process parameters. On the other hand, the volume of vapor loss determination from the storage tanks requires a lot of variable data and assumptions, which is different from plant to plant depending on locations and weather. And finally, once the hydrocarbon volume will be determined, the capacity and other parameters of the VRU compressor can be computed using equations described in this chapter.

2.2.1 Condensate Fractionation Unit Simulation

The Condensate Fractionation Unit is modeled using a steady-state process simulator Aspen^{TM} HYSYSv11. The distillation unit is one of the most complex unit operations that HYSYS simulates. The Peng-Robinson fluid property package is used for modeling it [7].

Initial data (chemical and physical properties) for condensate from the laboratory of an existing 4,000 BPD plant has been inputted to convert into a series of discrete hypothetical components. These hypothetical components are the base for the property package to predict the remaining thermodynamic and transport properties necessary for the condensate as the feed of the simulation [8]. Once the characterization of the fluid is completed, as per the actual process plant, the equipment and necessary data are added to the simulation environment to run the simulation. After simulation becomes successful, the desired data of off-gas volume as well as other information for tank vapor calculation will be achieved.

Then those data will be compared with the real-time data from the plant to validate the simulation's efficiency. Based on that, the planned 6,000 BPD unit simulation will be conducted.

2.2.2 Vapor Loss from the Storage Tanks

Vapor loss calculation from storage tanks is developed based on equations generated by the US Environment Protection Agency [9]. In the 4,000 BPD plant, the existing tanks are vertical cylindrical designed with a fixed roof cone shape. A typical fixed roof vertical tank is shown in Figure 2.2.

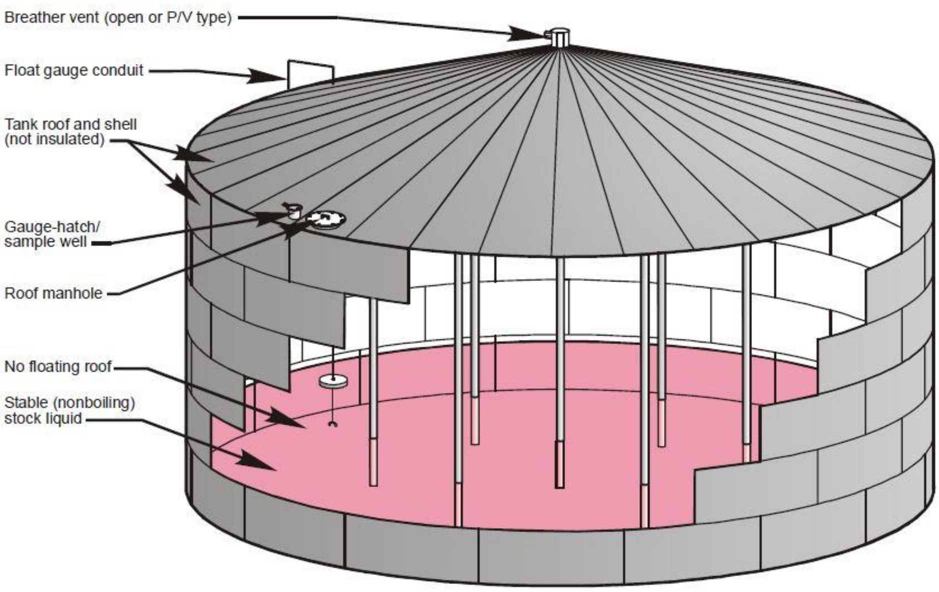


Figure 2.2: Fixed Roof Tank

The fixed roof tanks are the least expensive yet generate the highest amount of vapor as there is no control system to reduce the vapor generation.

On the other hand, floating roof tanks minimize vapor generation because the roof adjusts its height in accordance with the liquid level. Therefore, the void space between the roof and liquid level is insignificant. However, the vapor is produced from deck fittings, nonwelded deck seams, and the annular space between the deck and tank wall. Floating roof tanks can be of different types. One of them is an internal floating roof tank. A typical internal floating roof vertical tank is shown in Figure 2.3:

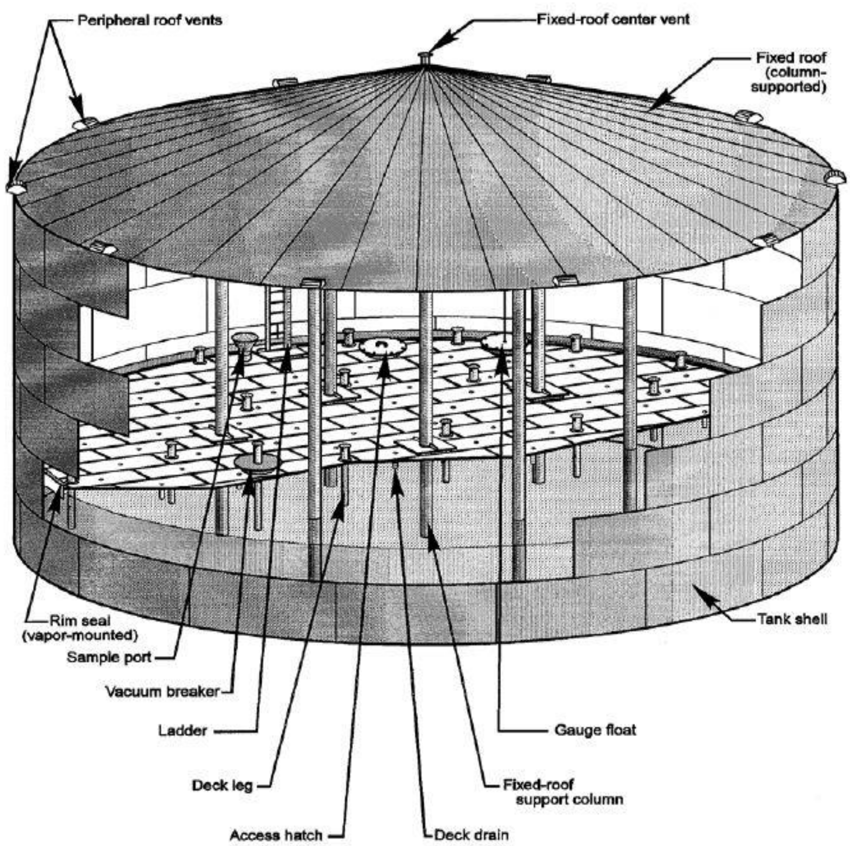


Figure 2.3: Internal Floating Roof Tank

In this project work, for calculating vapor losses from internal floating roof tanks, some assumption has to be made to obtain initial data from the tables listed in Appendix B.

The tank will be equipped with a liquid-mounted primary seal plus a secondary seal. The tank will be a column-supported fixed roof. The tank's deck will be welded and equipped with (1) two access hatches with unbolted, ungasketed covers; (2) an automatic gauge float well with an unbolted, ungasketed cover; (3) a pipe column well with a flexible fabric sleeve seal; (4) a sliding cover, gasketed ladder well; (5) adjusTable deck legs; (6) a slotted sample pipe well with a gasketed sliding cover; and (7) a weighted, gasketed vacuum breaker.

Either Fixed roof tank or floating roof tank, the routine emissions from both are the same and discussed below:

a) Standing losses, L_S,

Standing loss (L_S) refers to the evaporation of hydrocarbons due to daily temperature change without any alteration of liquid level. It is also known as breathing, as vapor ejects from a tank because of the increase and shrinkage of vapor in accordance with temperature escalation or reduction.

b) Working losses, L_W,

Working loss (L_W) refers to the expulsion of vapor when the liquid level inside hydrocarbon changes specifically when a filling operation occurs. The vapor pressure increases with respect to liquid level rise resulting in vapor discharges through vents by the same procedure of standing loss.

Although the types of losses are similar for all types of tanks, the equations are different based on their construction. In this project work, for the future 6,000 BPD unit, tank vapor loss will be calculated for two scenarios - fixed roof tank and internal floating roof tank.

2.2.2.1 Scenario 1: For Fixed Roof Tank

For fixed roof tank the standing loss equation is as bellows:

$$\text{Standing Loss, } L_S = 365 \times V_V \times K_E \times K_S \times W_V \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(2.1)}$$

Where, 365 = constant, the number of daily events in a year, $(\text{year})^{-1}$

V_V = Tank Vapor Space Volume, ft^3

W_V = Stock Vapor Density, lb/ ft^3

K_E = Vapor Space Expansion Factor

K_S = Vented Vapor Saturation Factor

Also, for fixed roof tanks the working loss equation is as bellows:

$$\text{Working Loss, } L_W = 0.0010 \times M_V \times P_{VA} \times Q \times K_N \times K_P \quad \ldots\ldots\ldots\ldots \text{(2.2)}$$

Where,

M_V = vapor molecular weight, lb/lb-mole

P_{VA} = vapor pressure at daily average liquid surface temperature, psia

Q = annual net throughput, bbl/yr

K_N = working loss turnover (saturation) factor,

N = number of turnovers per year, dimensionless [9].

2.2.2.2 Scenario 2: For Internal Floating Roof Tank

Standing losses from internal floating roof tanks are the sum of rim seal, deck fitting and deck seam losses, and may be written as:

Standing Loss, $L_S = L_R + L_F + L_D$……………………….……….… (2.3)

where:

L_S = standing loss, lb/yr

L_R = rim seal loss, lb/yr

L_F = deck fitting loss, lb/yr

L_D = deck seam loss (internal floating roof tanks only), lb/yr

The working loss from floating roof storage tanks, also known as withdrawal loss, can be estimated using the below equation:

Working Loss, $L_w = \frac{0{\cdot}943 Q C_s W_L}{D}\left(1 + \frac{N_C F_C}{D}\right)$………………..… (2.4)

where:

L_W = working (withdrawal) loss, lb/yr

Q = annual net throughput, bbl/yr

C_S = shell clingage factor, bbl/1,000 ft2

W_L = average organic liquid density, lb/gal

D = tank diameter, ft

0.943 = constant, 1,000 $ft^3 gal/bbl^2$

N_C = number of fixed roof support columns, dimensionless

F_C = effective column diameter, ft (column perimeter [ft]/π) [9].

2.2.3 VRU Compressor Design Criteria

VRU compressor design will be conducted based on the total volume of hydrocarbon from tanks and process area. Theoretically, the isothermal compression process needs minimum compression work which is economically favorable. But in practice, this is not feasible. That is why adiabatic/isentropic compression is followed by using multi-stages with intercooler is used between stages to make the process near to isothermal. Several parameters need to take into consideration during compressor design, which is discussed below:

2.2.3.1 Compressor Selection

In a fractionation plant, the steady flow rate of vapor from tanks is not achievable, as stated earlier vapor generation depends on temperature. Reciprocating compressors can handle a wide range of flow rate variations with little loss of efficiency. Reciprocating compressor's performance is not affected by the variation of the properties of vapor, such as molecular weight, specific gravity or density of the inlet vapor. This is a key factor, as there is a possibility of a change of feed composition if the condensate comes from different gas fields or the same gas field over time.

There is another advantage of selecting a reciprocating compressor. In the existing plant, the outlet/discharge pressure is fixed, to meet up the utility fuel gas line pressure. And the inlet pressure is low i.e., the compression ratio is significantly high. A reciprocating compressor is the best choice for these types of operations. VFD can be used for efficient compressors operating. The VFD is an electromechanical driver system whose main purpose is to adjust the speed of the compressor depending on the rate of flow of gas into the VRU.

2.2.3.2 Number of Stages & Compression Ratio

The initial parameter in designing a compressor is to determine the number of stages which depend on the desired compression ratio. In practical terms, a compression ratio ≤ 6 is maintained. With the inlet and discharge pressure data, compression ratio and the number of stages can be calculated by the below equation:

$$r_{opt} = (P_d/P_S)^{1/nS} \quad \text{.......} \quad (2.5)$$

Where, r_{opt} = Optimum Compression Ratio Per Stages

P_d = Final Discharge pressure, psia

P_S = Initial Suction pressure, psia

n_S = Total number of Stages

2.2.3.3 Horsepower Requirement

During compressor design, accuracy depends on the availability of adequate data. That is why for preliminary estimation more than one method is investigated. Here, two methods will be discussed for compressor design.

- Analytical Approach

Based on basic thermodynamic relations, the horsepower requirement for compression work is stated below:

$$\text{Ideal Horsepower, IHP} = \frac{k}{k-1}\frac{3.027p_b}{T_b}qT_1\left[\left(\frac{p_2}{p_1}\right)^{z_1(k-1)/k} - 1\right] \quad \ldots\ldots\ldots\ldots\ldots (2.6)$$

Where,

P_1= Inlet Pressure, psia

T_1= Inlet Temperature, °R

Z_1= Gas Compressibility at Inlet Condition, (From Appendix A)

P_2 = Discharge Pressure, psia

Pb = Base Pressure =14.7 paia

Tb = Base Temperature = 520 °R

k = Isentropic Exponent

q = Gas Flow, MMSCFD

And,

Actual/Break Horsepower, BHP = IHP/E ……………………………….…. (2.7)

Here, E = Efficiency

- Mollier Diagram Method

Mollier diagram (listed in Appendix A) is an Enthalpy- Entropy chart, as a function of pressure and temperature, is a good technique for estimating values of gas for solving compressor related problems. From the Mollier chart, Enthalpy for isentropic compression can be directly gained at the inlet and discharge pressure and temperature of each stage. From that data horsepower requirement for compression work is calculated by the below energy balance equation:

$$\text{Ideal Horsepower, IHP} = \frac{n(h_2 - h_1) \times 778 \cdot 2}{24 \times 60 \times 3300} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(2.8)}$$

Where,

h_1 = Enthalpy of the gas at inlet condition, Btu/lbmole

h_2 = Enthalpy of the gas at discharge condition, Btu/lbmole

n = number of mole of gas being compressed, lbmole/D

= (P x q) / (R x T) (psi x MMSCFD) / °R

Similar to analytical approach,

Actual/Break Horsepower, BHP = IHP/E …………………………………….………. (2.9)

2.2.3.4 Cooling Requirement

Intercooling/ aftercooling heat requirements is an iso-baric process that means the temperature is reduced after compression to its inlet temperature where the pressure remains constant. Similar to horsepower requirements two methods will be discussed for the cooling requirements of compressor design.

- Analytical Approach

The heat removed in the intercooler/aftercooler can be quantified using the below equation:

Heat Removed by Intercooler/Aftercooler, $\Delta H = n\,\bar{C}_P \Delta T$ …………………… (2.10)

Where,

$\bar{C}_P$ = Constant Pressure molal specific Heat of the gas at cooler pressure and avg. cooler Temperature, Btu/lbmole °F

ΔT = Inlet and Outlet Gas Temperature Difference, °F

- Mollier Diagram Method

From the Mollier chart, enthalpy for isentropic cooling can directly be gained at inlet & discharge pressure and inlet temperature. From that data heat removed in the intercooler/aftercooler can be quantified using the below equation:

The ideal compression power (or rate of work) required is given by:

Heat Removed by Intercooler/Aftercooler, $\Delta H = n\,(h_2 - h_3)$ ………………………. (2.11)

Where,

h_2 = Enthalpy of the Gas at Discharge Temperature & Pressure, Btu/lbmole

h_3 = Enthalpy of the Gas at discharge Pressure & Inlet Temperature, Btu/lbmole

n = number of mole of gas being compressed, lbmole/D [10][11]

Based on the description of the current 4,000 BPD plant in the earlier section of this chapter, the design basis has been developed. In the next few chapters, the calculation of the VRU system for the present plant and the new plant will be presented.

Chapter 3

CALCULATION FOR EXISTING 4,000 BPD CAPACITY

Before designing a VRU system for a new plant, off-gas volume, tank vapor loss volume and finally VRU compressor's compulsory parameters for the existing 4,000 BPD plant will be calculated in this chapter. Then it will be matched with the plant's real-time data to verify the validity of the calculated values to continue the same calculation for the new proposed plant of 10,000 BPD.

3.1 Off-Gas Estimation from the Column

Off-gas volume and other necessary data will be obtained by simulating HYSYS software for the current 4,000 BPD fractionation plant. Before running the simulation initially, we have to input some existing data and make some assumptions.

The following pressure drops are assumed in the simulation models:

Table 3.1: Assumed Pressure Drop in Equipment

EQUIPMENT	Pressure Drop, bar	Pressure Drop, psi
Shell and Tube Heat Exchanger	0.5	7.25
Column	0.3	4.35

The method of simulation by HYSYS is to create a pseudo-component from Laboratory generated 'True Boiling Point (TBP)' analysis data based on raw condensate obtained from an existing plant. 'True Boiling Point (TBP)' or ASTM D86 data for condensate are inputted data named as "Input Assay" in the calculation window as oil manager in HYSYS are shown in Table 3.2 to 3.4.

Table 3.2: Input Assay-Bulk Property

Property Name	Value
Standard Density	48.56 lb/ft^3
Viscosity Type	Kinematic
Viscosity 1 Temperature	20.00 C
Viscosity 1	0.7836 cSt
Viscosity 2 Temperature	40.00 C
Viscosity 2	0.6353 cSt

Table 3.3: Input Assay-Distillation

Assay Basis- Mass	
Assay Percent	**Temperature, ℃**
14.40	70.00
19.90	80.00
25.50	90.00
47.30	110.0
57.10	130.0
70.00	150.0
79.30	175.0
82.50	190.0
86.30	210.0
89.60	230.0
92.60	250.0

Table 3.4: Input Assay-Light Ends

Light End Basis – Mass %		
Light Ends	**Composition**	**NBP, ℃**
Methane	0.0000	-161.5
Ethane	0.2900	-88.60
Propane	15.54	-42.10
n-Butane	44.35	-0.5020
i-Butane	32.34	-11.73
n-Pentane	0.3710	36.06
i-Pentane	5.782	27.88
22-Mpropane	1.165	9.498
Cyclopentane	1.6×10^{-2}	49.25
CO_2	0.0000	-78.55
H_2O	0.0000	100.0
Percent of Light Ends in Assay	**1.0000**	

After completing the characterization of fluid, for the next step of the simulation, the achieved feed stream "Condensate" is shown in Table 3.5.

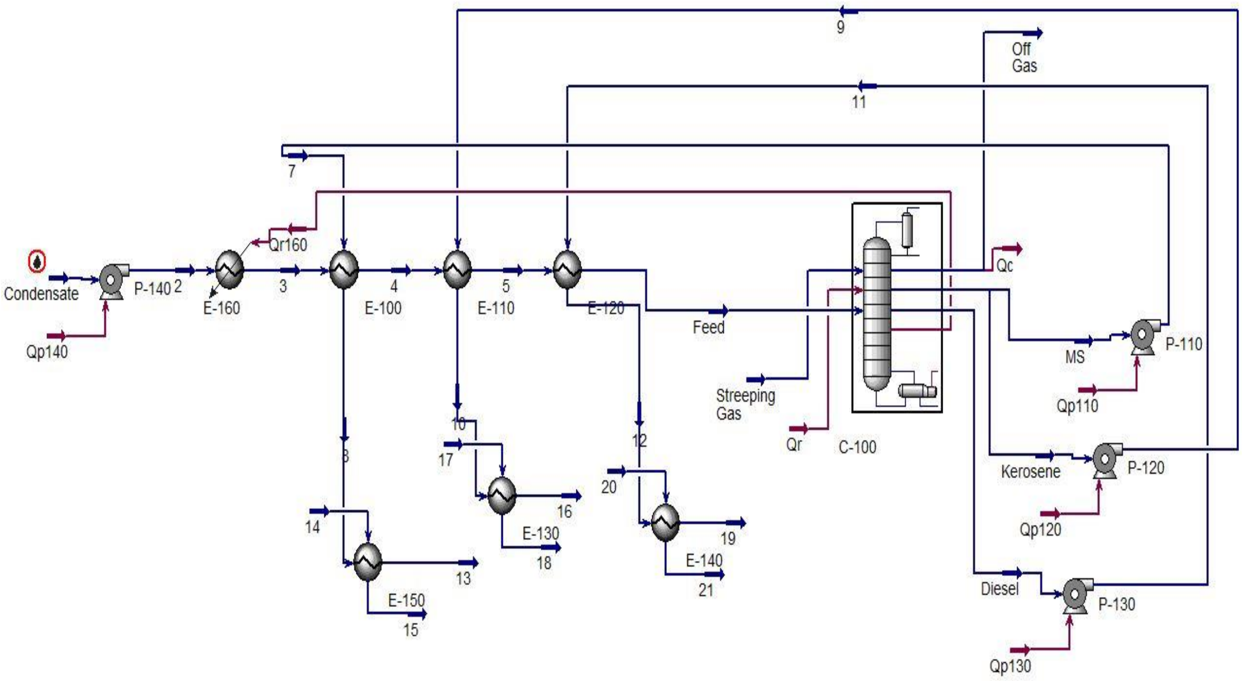
In the Simulation Tab, Condensate is entered to the 12th tray of Fractionation Column (C-100) through multiple equipment. First, the feed is pumped by Condensate Feed Pumps (P-140 A/B) to Condensate -Column Overhead Heat Exchanger (E-160) to absorb heat from top column off-gas then flowing to Condensate - Motor Spirit Heat Exchanger (E-100). After that, condensate the temperature will be increased by absorbing the heat content of kerosene products inside the Condensate-Kerosene Heat Exchanger (E-110) and Condensate-Diesel Heat Exchanger (E-120). Hot condensate is then entering the Fractionation Column (C-100).

Table 3.5: Feed Stream "Condensate"

Stream Name	Condensate	Liquid phase
Vapor/ Phase Fraction	0.0000	1.0000
Temperature, °C	30.00	30.00
Pressure, psig	2.901	2.901
Molar Flow, lb mole/hr	431.8	431.8
Mass Flow, lb/hr	4.544×10^4	4.544×10^4
Std Ideal Liq. Vol, bbl/day	4000	4000
Molar Enthalpy, Btu/lb mole	-9.906×10^4	-9.906×10^4
Molar Entropy, Btu/lb mole-F	7.019	7.019
Heat Flow, Btu/hr	-4.278×10^7	-4.278×10^7
Liq. Vol Flow @ Std Cond, bbl/day	4000	4000
Fluid Package	Basis-1	

As the name suggests, fractionation is also known as the distillation process means the separation method by utilizing different relative volatility of its components. The highest volatile component will be drawn from the top of the column, mild volatile components will be produced as column side draw and the lowest volatile component is a column bottom product. Heat is injected into the column by a reboiler located at the bottom of the column to generate high temperature vapor back to the column (24th tray). Liquid with a high pure volatile component is also injected back to the top of the column (1st tray) which is known as reflux. It can happen by partial condensation of the top product by a series of heat exchanging inside Condensate-Column Overhead Heat Exchanger (E-160) and Air -Cooled Overhead Condenser. The vapor-liquid mixture is then separated by Reflux Accumulator.

Separated vapor off-gas will flow to the Vapor Recovery Unit (VRU). Heat and mass transfer will occur between low temperature liquid and high temperature vapor in several trays (24 trays) installed inside the column. It will allow a light component of liquid to diffuse into vapor, on the other hand also allowing a heavy component of vapor will diffuse into liquid. The result is a vapor with a purer light component and a liquid with a purer heavy component. The HYSYS Scheme for 4,000 BPD Condensate Fractionation Plant is shown in Figure 3.1 below.



Feed		
Temperature	90.00	C
Pressure	43.51	psig
Mass Flow	4.544e+004	lb/hr

MS		
Temperature	76.00	C
Pressure	5.000	psig
Mass Flow	2.942e+004	lb/hr

Kerosene		
Temperature	199.0	C
Pressure	16.00	psig
Mass Flow	1.403e+004	lb/hr

Diesel		
Temperature	350.0	C
Pressure	18.00	psig
Mass Flow	1984	lb/hr

Off Gas		
Temperature	76.00	C
Pressure	5.000	psig
Mass Flow	14.68	lb/hr

Figure 3.1: HYSYS Scheme for 4000 BPD Condensate Fractionation Plant

The sub-flow sheet of the column environment is shown in Figure 3.2.

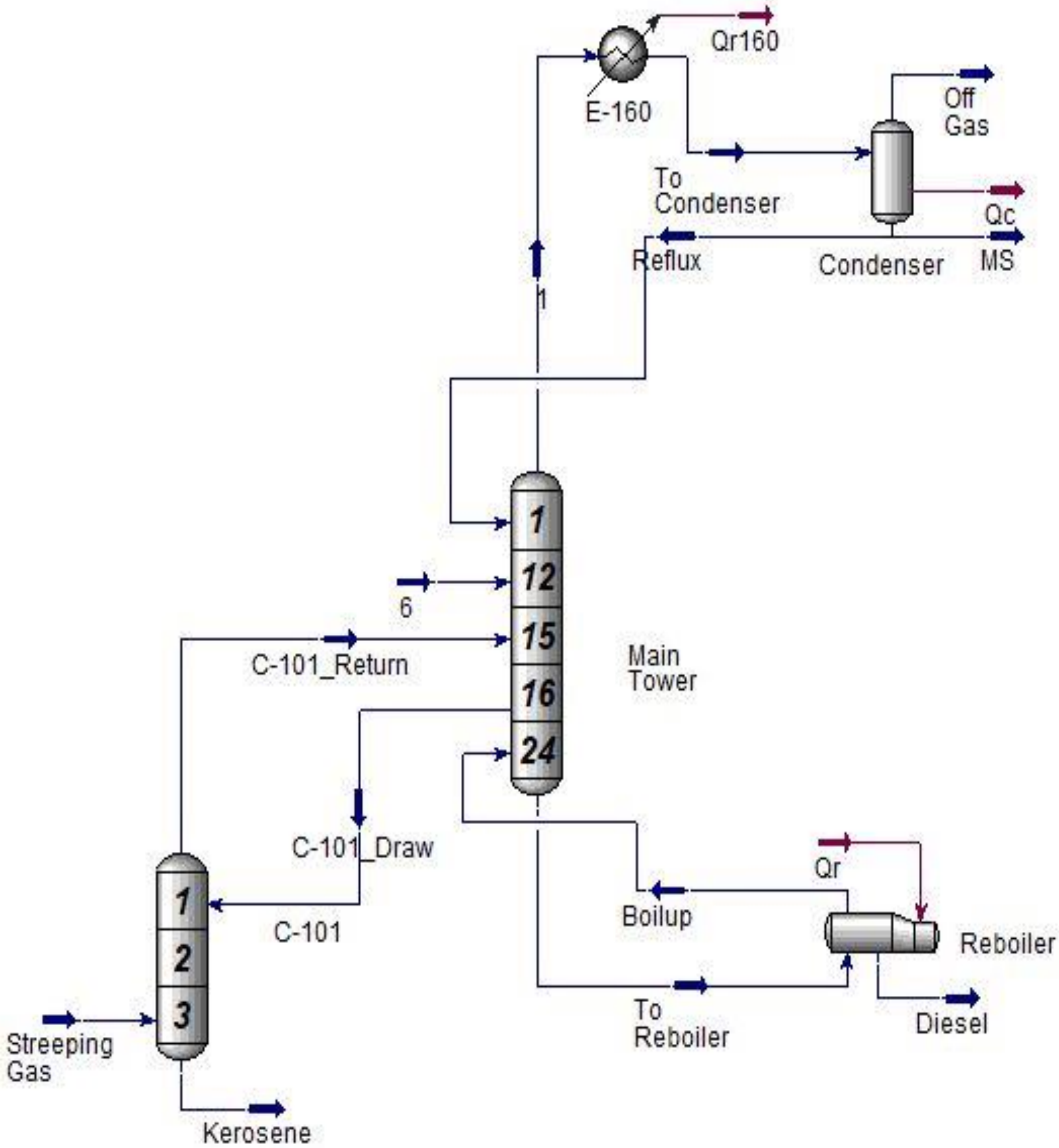


Figure 3.2: Sub-flow Sheet of Column Environment

Products of Fractionation Column (C-100) are:

- Off gas — drawn as column top product
- Motor spirit — drawn as column top product
- Kerosene — produced as column side draw
- Diesel — drawn as column bottom product

The composition of feed and products are tabulated in Table 3.6.

Table 3.6: Composition of Feed and Products

Composition	Condensate	Motor Spirit	Kerosene	Diesel	Off Gas
	Mole Fraction, %				
Methane	0.0000	0.0015	0.0011	0.0000	0.2897
Ethane	0.0001	0.0002	0.0000	0.0000	0.0075
Propane	0.0037	0.0049	0.0000	0.0000	0.0847
n-Butane	0.0080	0.0107	0.0000	0.0000	0.0666
i-Butane	0.0059	0.0078	0.0000	0.0000	0.0644
n-Pentane	0.0001	0.0001	0.0000	0.0000	0.0002
i-Pentane	0.0008	0.0011	0.0000	0.0000	0.0033
22-Mpropane	0.0002	0.0002	0.0000	0.0000	0.0011
Cyclopentane	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0000	0.0000	0.0000	0.0000	0.0002
H_2O	0.0000	0.0000	0.0000	0.0000	0.0000
NBP[0]59*	0.0812	0.1082	0.0000	0.0000	0.1333
NBP[0]71*	0.0886	0.1182	0.0000	0.0000	0.1027
NBP[0]86*	0.1149	0.1533	0.0001	0.0000	0.0873
NBP[0]100*	0.2070	0.2760	0.0005	0.0000	0.1028
NBP[0]113*	0.0935	0.1244	0.0010	0.0000	0.0314
NBP[0]129*	0.0846	0.1102	0.0088	0.0000	0.0165
NBP[0]142*	0.0881	0.0819	0.1154	0.0000	0.0081
NBP[0]157*	0.0595	0.0013	0.2527	0.0000	0.0001
NBP[0]171*	0.0336	0.0000	0.1451	0.0000	0.0000
NBP[0]185*	0.0244	0.0000	0.1055	0.0000	0.0000
NBP[0]201*	0.0192	0.0000	0.0828	0.0000	0.0000
NBP[0]214*	0.0194	0.0000	0.0836	0.0001	0.0000
NBP[0]229*	0.0157	0.0000	0.0678	0.0004	0.0000
NBP[0]243*	0.0132	0.0000	0.0566	0.0027	0.0000
NBP[0]257*	0.0093	0.0000	0.0389	0.0154	0.0000
NBP[0]272*	0.0074	0.0000	0.0242	0.0877	0.0000
NBP[0]286*	0.0062	0.0000	0.0087	0.2086	0.0000
NBP[0]300*	0.0042	0.0000	0.0025	0.1824	0.0000
NBP[0]315*	0.0030	0.0000	0.0014	0.1351	0.0000
NBP[0]330*	0.0024	0.0000	0.0010	0.1055	0.0000
NBP[0]345*	0.0020	0.0000	0.0008	0.0881	0.0000
NBP[0]358*	0.0019	0.0000	0.0008	0.0837	0.0000
NBP[0]373*	0.0020	0.0000	0.0008	0.0903	0.0000
Total	**1.0000**	**1.0000**	**1.0000**	**1.0000**	**1.0000**

Motor spirit (temperature 76°C), as a product of the top of the column, is taken from the condenser and cooled inside E-100 until reaching temperature 55°C before final cooling in E-150 to decrease its temperature to 40°C and then directed to the motor spirit tank header.

Kerosene (temperature 207°C) is drawn from the 16th tray of C-100 which next step is directed to the top section of the Stripping Column (C-101). Stripping gas (35°C) is injected from the bottom of C-101. Internal of C-101 is packing type allowing direct contact between kerosene liquid and stripping gas. The result is 199°C pure kerosene produced from the bottom of C-101 and then pumped to cooling inside E-110 to reach 145°C kerosene before final product cooling in E-130. The final product temperature is 45°C and then directed to the kerosene tank header. Stripping gas from the top of C-101 is injected back into the 15th tray of C-100.

Diesel (temperature 350°C) product is drawn from the bottom of the column and directed to E-120. After releasing its heat content inside E-120, 171°C diesel product is achieved before entering the final product cooling heat exchanger in E-140. The final product temperature is 45°C and then directed to the diesel tank header. Some portions of diesel are injected into the Fired Heater (reboiler) for heating and vaporizing and directed back as vapor into the bottom of C-100.

Off gas will be directed to the Vapor Recovery Unit (VRU) from the top of the column and then utilized as additional fuel gas. The simulated result of products is compared with the plant's production volume to observe the accuracy of the simulation in Table 3.7 below.

Table 3.7: Simulation Value vs Real Plant Value of Feed & Products

Product Name	Simulated Value			4,000 BPD Plant's Value	
	Mass Flow lb/hr	Mass Flow kg/hr	Percent	Mass Flow kg/hr	Percent
Condensate	4.544×10^4	2.061×10^4	100 %	2.115×10^4	100 %
Motor Spirit	2.942×10^4	1.334×10^4	64.73 %	1.427×10^4	67.47 %
Kerosene	1.403×10^4	6364	30.87 %	5916.19	27.98 %
Diesel	1984	900	04.37%	954	04.51 %
Off-Gas	14.68	6.66	0.03%	9.81	0.04 %

From the above table, it is observed that plant's data and simulated data are fairly matched. Which gives the assurance that the simulation can be used for further estimation. Now, for 4,000 BPD plant, the simulated physical properties of the off-gas are shown in below Table 3.8.

Table 3.8: Simulated Physical Properties Off-Gas

Stream Name	Off Gas	Liquid phase
Vapor/ Phase Fraction	1.0000	0.0000
Temperature, °C	76.00	76.00
Pressure, psig	5.000	5.000
Molar Flow, lbmole/hr	0.2572	0.2572
Mass Flow, lb/hr	14.68	14.68
Std Ideal Liq Vol, bbl/day	1.612	1.612
Molar Enthalpy, Btu/lbmole	-5.085×10^4	-5.085×10^4
Molar Entropy, Btu/lbmole-F	35.48	35.48
Heat Flow, Btu/hr	-1.308 x 104	-1.308 x 104
Liq. Vol Flow @ Std Cond, bbl/day	1.610	1.610
Fluid Package	Basis-1	

3.2 Tank Vapor Losses Calculation

Total Vapor Loss will come from the summation of standing loss and working loss. In the present plant, parallel tank operation is carried out for the same type of fuel storage tank. Therefore, during calculation, one tank is assumed to be working and the rest of the other tanks are standing. Preliminary data for calculation are shown in Table 3.9 to 3.13.

3.2.1 Standing Loss Calculation for Condensate & Motor Spirit Tank

By equation (2.1), standing loss for a single tank has several parameters which need to be estimated step by step from the data taken from the plant. The equation used in this section is taken from "Compilation of Air Pollutant Emissions Factors (AP-42)" of the US Environmental Protection Agency [9].

$$\textbf{Standing Loss, } L_s = 365 \times V_V \times K_E \times K_S \times W_V$$

a) Tank Vapor Space Volume, Vv

The tank vapor space volume, Vv for vertical cone roofed tank is calculated by the below equation.

$$V_V = \prod/4 \times D^2 \times H_{vo}$$

$$= \prod/4 \times D^2 \times (H_s - H_L + 1/3\ H_{RO})$$

$$= \prod/4 \times D^2 \times (H_s - H_L + 1/3\ S_R \times R_s) \quad \ldots\ldots (3.1)$$

Here D = Tank Diameter, ft

H_S = Tank Shell Height, ft

H_L = Stock Liquid Height = 1/2 H_S, ft

S_R = Tank Cone Roof Slope

R_S = Tank Shell Radius

Table 3.9: Data for Tank Vapor Space Volume Calculation

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
D = Tank Diameter	25.7	m			84.30	ft
H_S = Tank Shell Height	17.15	m			56.25	ft
H_L = Stock Liquid Height = 1/2 H_S			8.6	m	28.13	ft
S_R = Tank Cone Roof Slope	0.0625				0.0625	
R_S = Tank Shell Radius	42.2	ft			42.20	ft

Therefore, from equation (3.1),

$$\mathbf{V_v} = (3.1416/4) \times (84.30)^2 \times ((56.25-28.13) + (1 \times 0.0625 \times 42.20/3))$$

$$= \mathbf{161875.15\ ft^3}$$

b) Vapor Space Expansion Factor, K_E

The vapor space expansion factor, K_E, is obtained by

$$K_E = \frac{\Delta Tv}{T_{LA}} + \frac{\Delta P_v - \Delta P_B}{P_A - P_{vA}} \text{ (if TVP > 0.1 psia \& Breather Valve vent setting range > ± 0.03)}$$

………… …………...………..…(3.2)

Here,

ΔT_V = Daily Vapor Temperature Range, °R

ΔP_V = Daily Vapor Pressure Range, psia

ΔP_B = Breather Vent Pressure Setting Range, psig

P_A = Atmospheric Pressure, psia

P_{VA} = Vapor Pressure at Daily Average Liquid Surface Temperature, psia

T_{LA} = Daily Average Liquid Surface Temperature, °R

- ***Daily Vapor Temperature Range***, $\Delta T_v = 0.72 \times \Delta T_A + (0.028 \times \alpha \times I)$

$$= 0.72 (T_{AX} - T_{AN}) + (0.028 \times \alpha \times I) \quad \ldots\ldots (3.3)$$

Where

ΔT_V = Daily Vapor Temperature Range, °R

ΔT_A = Daily Ambient Temperature Range, °R

T_{AX} = Daily Maximum Ambient Temperature, °R

T_{AN} = Daily Minimum Ambient Temperature, °R

α = Tank Paint Solar Absorptance, dimensionless

I = daily total solar insolation factor, Btu/ft^2 day

Table 3.10: Data for Daily Vapor Temperature Range Calculation

Description	Data	Unit	Unit Conversion	Unit
T_{AX} = Daily Maximum Ambient Temperature	39	°C	561.87	°R
T_{AN} = Daily Minimum Ambient Temperature	5	°C	500.67	°R
α = Tank Paint Solar Absorptance (From Appendix B, Table B-1)	0.54		-	
I = Daily total solar insolation factor	7920	BTU/ft^2 day	-	

Therefore, in equation (3.3),

$$\Delta Tv = 0.72 \times (561.87-500.67) + (0.028 \times 0.54 \times 7920)$$

$$= 163.8 \text{ °R}$$

- ***Daily Average Liquid Surface Temperature***

$$T_{LA} = 0.44 \times T_{AA} + 0.56 \times T_B + 0.0079 \times \alpha \times I \quad \ldots\ldots (3.4)$$

$$= 0.44 \times ((T_{AX}+T_{AN})/2) + 0.56 \times ((T_{AX}+T_{AN})/2) + 6\alpha-1) + 0.0079 \times \alpha \times I$$

$$= 0.44*((561.87+500.67)/2) +0.56*(((561.87+500.67)/2) +6*0.54-1)$$

$$+0.0079*0.54*7920$$

$$= 566.3 \text{ °R}$$

- ***Daily Vapor Pressure Range,*** $\Delta P_V = P_{VX} - P_{VN}$ ……………………………….… (3.5)

Where

ΔP_V = Daily Vapor Pressure Range, psia

P_{VX} = Vapor Pressure at Daily Maximum Liquid Surface Temperature, psia

P_{VN} = Vapor Pressure at Daily Minimum Liquid Surface Temperature, psia

Presure equation for P_{VX} & P_{VN} is as follows:

$$P = \exp \left\{ \left[0.7553 - \left(\frac{413.0}{T + 469.6}\right)\right] S^{0.5} \log(\mathrm{RVP}) - \left[1.854 - \left(\frac{1{,}042}{T + 469.6}\right)\right] S^{0.5} + \left[\left(\frac{2{,}416}{T + 469.6}\right) \log(\mathrm{RVP}) - \left(\frac{8{,}742}{T + 469.6}\right) + 15.64\right\} \quad \text{………...… (3.6)}$$

Where, Maximum Liquid Temperature,

T_{LX} (@ P_{VX}) = T_{LA} + 0.25 x Δ Tv = 566.3 + 0.25 x 163.8 = 607.26 °R = 147.66 °F

Minimum Liquid Temperature,

T_{LN} (@ P_{VN}) = T_{LA} - 0.25 x Δ Tv = 566.3 - 0.25 x 163.8 = 607.26 °R = 65.67 °F

Table 3.11: Data for Slope of the ASTM Distillation Curve Calculation

Description	Data	Unit	Unit Conversion	Unit
RVP = Raid vapor pressure	7.1	psi		
Temp. ASTM D-86 @ 15% Distillation Curve	100.71	°C	213.28	°F
Temp. ASTM D-86 @ 5% Distillation Curve	60.86	°C	141.55	°F

S = (Temp. ASTM D-86 @ 15% Distillation Curve - Temp. ASTM D-86 @ 5% Distillation Curve) /10 ……………………………………………….... (3.7)

= (213.28 – 141.55)/10 °F/Percent

= 7.17 °F/Percent

Therefore, from equation (3.6),

P_{VX} = EXP ((0.7533- (413/ (147.66 +459.6))) x $7.17^{0.5}$ x $\log_{10}$(7.1) -(1.854-(1042/ (147.66 +459.6))) x $7.17^{0.5}$+ ((2416/ (147.66 +459.6))-2.013) x $\log_{10}$ (7.1) -(8742/ (147.66

+459.6)) +15.64)

= 15.10 psi

P_{VN} = EXP ((0.7533- (413/ (65.67+459.6))) x $7.17^{0.5}$ x $\log_{10}(7.1)$ -(1.854-(1042/ (65.67

+459.6))) x $7.17^{0.5}$+((2416/ (65.67+459.6))-2.013) x $\log_{10}(7.1)$ -(8742/ (65.67+459.6))

+15.64) = 4.36 psia

P_{VA} = EXP ((0.7533- (413/ (566.3 +459.6))) x $7.17^{\wedge 0.5}$ x $\log_{10}(7.1)$ -(1.854-(1042/ (566.3

+459.6))) x $7.17^{0.5}$+((2416/ (566.3 +459.6))-2.013) x $\log_{10}(7.1)$ -(8742/ (566.3 +459.6))

+15.64) = 8.49 psi

And in equation (3.5),

ΔP_V = 15.10 -4.36

= 10.73 psia

- ***Breather Vent Pressure Setting Range,*** $\Delta P_B = P_{BP} - P_{BV}$ ……………………..… (3.8)

Table 3.12: Data for Breather Vent Pressure Setting Range Calculation

Description	Data	Unit	Unit Conversion	Unit
P_{BP} = Breather Vent Pressure Setting	0.10	barg	0.147	psig
P_{BV} = Breather Vent Vacuum Setting	0.006	barg	0.0882	psig
P_A = Atmospheric Pressure	14.7	psia		

From equation (3.10),

ΔP_B = 0.10- 0.006

= 0.094 barg = 1.36 psig

Therefore, in equation (3.2),

$\mathbf{K_E}$ = (163.8/566.3) +((10.73 -1.36)/ (14.7-8.49))

= **1.80**

c) Vented Vapor Saturation Factor, K_S

The equation for vented vapor saturation factor, K_S is as follows

$$\mathbf{K_S} = \frac{1}{1+0.053 P_{vA} H_{v0}} \quad \text{……………………………………………………… (3.9)}$$

= 1/ (1+0.053 x 8.49 x ((56.25-28.13) +(1 x 0.0625 x 42.20/3)))

= 0.071

d) Vapor Density, Wv

The equation for vapor density, Wv is as follows

$W_V = (M_V \times P_{VA}) / (R \times T_{LA})$ ……………………………………………………… (3.10)

Table 3.13: Data for Vapor Density Calculation

Description	Data	Unit
M_V = Molecular weight	67.64	
R = Ideal Gas Constant	10.731	psia ft3/lb-mole °R

Therefore, in equation (3.10)

Wv = (67.64 * 8.49) / (10.731* 566.3)

= **0.094 lb/ft^3**

In equation (2.1),

Standing Loss, Ls = 365 x Vv x KE x Ks x Wv

= 365 x 161875.15 x 1.80 x 0.071 x 0.94 lb/yr

= 714245.07 lb/yr

= 714245.07 / (2.205 x 365 x 24) kg/hr

= 36.98 kg/hr

= (36.98 x 2.205 x 24)/ (0.094 x 1000000)

= 0.021 MMSCFD

The summary view of the standing loss calculation for a single condensate storage tank is shown in following Table 3.14.

Table 3.14: Summary of Standing Loss Calculation for Condensate Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
D = tank diameter	25.7	m			84.30	ft
H_S = tank shell height	17.15	m			56.25	ft
H_L = stock liquid height = 1/2 Hs			8.6	m	28.13	
S_R = tank cone roof slope	0.063				0.0625	ft
R_S = tank shell radius	42.2	ft			42.20	ft
V_V = Tank Vapor Space Volume			161875.15	ft3		
T_{AX} = daily maximum ambient temperature	39	°C			561.87	°R
T_{AN} = daily minimum ambient temperature	5	°C			500.67	°R
α = Paint Factor (For Fixed roof tank)	0.54				-	
I = daily minimum ambient temperature	7920	BTU/ft^2/day			-	
Δ T_V = Daily Vapor Temperature Range,			163.8	°R	-	
T_{LA} = Daily Average Liquid Surface Temperature			566.3	°R		
T_{LX} = Maximum Liquid temperature			607.26	°R	147.66	°F
T_{LN} = Minimum Liquid temperature			525.36	°R	65.76	°F
RVP = Raid vapor pressure	7.1	psia				
Temp. ASTM D-86 @ 15% Distillation Curve	100.71	°C			213.28	°F
Temp. ASTM D-86 @ 5% Distillation Curve	60.86	°C			141.55	°F
S = Slope of the ASTM Distillation Curve at 10% Evaporation			7.17	°F/Percent	-	
P_{VX} = Vapor pressure at T_{LX}			15.10	PSIA	-	
P_{VN} = Vapor pressure at T_{LN}			4.36	PSIA	-	
P_{VA} = Vapor pressure at T_{LA}			8.49	PSIA	-	
ΔP_V = Vapor pressure Range			10.73	PSIA	-	
P_{BP} = Breather Vent Pressure Setting	0.10	barg			1.45	PSIG
P_{BV} = Breather Vent Vacuum Setting	0.006	barg			0.087	PSIG
P_A = Atmospheric Pressure	14.7	PSIA			-	
ΔP_B = Breather Vent Pressure Setting Range			0.094	barg	1.36	PSIG
M_v = Molecular Weight	67.64	lb/lb-mole				
R= Ideal Gas Constant	10.731	psi ft3/lb-mole °R				
K_E = Vapor Space Expansion Factor			1.80			
K_S = Vented Vapor Saturation Factor			0.071			
W_V = Stock Vapor Density			0.094	lb/ft3		
L_S = Standing Loss			**714245.07**	**lb/yr**	**36.98**	**kg/hr**
					0.021	**MMSCFD**

In Table 3.14, the 2nd column represents all the data from Table 3.9-3.13 obtained from the plant. With these data, the calculated values are shown in the 4th column, which is required to calculate the Standing loss. Unit conversion of data is shown in the 6th column.

Similarly, one motor spirit tank vapor is obtained and recorded in Table 3.15 below:

Table 3.15: Summary of Standing Loss Calculation for Motor Spirit Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
D = tank diameter	23.4	m			76.75	ft
H_S = tank shell height	15.66	m			51.36	ft
H_L = stock liquid height = 1/2 Hs			7.8	m	25.68	ft
S_R = tank cone roof slope	0.0625				0.0625	ft
R_S = tank shell radius	38.4	ft			38.40	ft
V_V = Tank Vapor Space Volume			122525.83	ft^3		
T_{AX} = daily maximum ambient temperature	39	°C			561.87	°R
T_{AN} = daily minimum ambient temperature	5	°C			500.67	°R
α = Paint Factor (For Fixed roof tank)	0.54				-	
I = daily minimum ambient temperature	7920	BTU/ft^2/day			-	
ΔT_V = Daily Vapor Temperature Range,			163.8	°R	-	
T_{LA} = Daily Average Liquid Surface Temperature			566.3	°R		
T_{LX} = Maximum Liquid temperature			607.26	°R	147.66	°F
T_{LN} = Minimum Liquid temperature			525.36	°R	65.76	°F
RVP = Raid vapor pressure	9.5	PSIA				
Temp. ASTM D-86 @ 15% Distillation Curve	78.51	°C			173.32	°F
Temp. ASTM D-86 @ 5% Distillation Curve	56.8	°C			134.24	°F
S = Slope of the ASTM Distillation Curve at 10% Evaporation			3.91	°F/Percent		
P_{VX} = Vapor pressure at TLX			20.79	PSIA	-	
P_{VN} = Vapor pressure at T_{LN}			5.59	PSIA	-	
P_{VA} = Vapor pressure at T_{LA}			11.30	PSIA	-	
ΔP_V = Vapor pressure Range			15.20	PSIA	-	
P_{BP} = Breather Vent Pressure Setting	0.10	barg			1.45	PSIG
P_{BV} = Breather Vent Vacuum Setting	0.006	barg			0.087	PSIG
P_A = Atmospheric Pressure	14.7	PSIA			-	
ΔP_B = Breather Vent Pressure Setting Range			0.094	barg	1.36	PSIG
M_V = Molecular Weight	42.29	lb/lb-mole				
R= Ideal Gas Constant	10.731	psi ft3/lb-mole °R				
K_E = Vapor Space Expansion Factor			4.362			
K_S = Vented Vapor Saturation Factor			0.059			
W_V = Stock Vapor Density			0.079	lb/ft3		
L_S = Standing Loss			**909859.49**	**lb/yr**	**47.10**	**kg/hr**
					0.032	**MMSCFD**

3.2.2 Working Loss Calculation for Condensate & Motor Spirit Tank

By equation 2.2, working loss needs to evaluate from the data (in Table 3.16) taken from the plant.

Working Loss, $L_w = 0.0010 \times M_v \times P_{VA} \times Q \times K_N \times K_P$

a) Working Loss Turnover (Saturation) Factor, K_N

$K_N = (180+N) / (6 \times N)$ for $N > 36$ ……………………..… (3.13)

Where,

Number of Turnovers Per Year, $N = 5.615 \times Q / V_L \text{ (year)}^{-1}$ ……….........… (3.14)

Tank normal Liquid Volume, $V_L = 3.1416/4 \times D^2 \times H_L \text{ ft}^3$ …………….......… (3.15)

Table 3.16: Data for Tank Normal Liquid Volume Calculation

Description	Data	Unit	Unit Conversion	Unit
Mv = Vapor Molecular Weight	67.64	lb/lb-mole		
P_{VA} = Vapor pressure at T_{LA}	8.487	psi		
Q= Net Throughput	32	m^3/hr	1,761,675.259	bbl/yr
K_P= working loss product factor for all organic liquid	1			

From equation (3.14) & (3.15)

$$V_L = 3.1416/4 \times D^2 \times H_L$$

$$= 3.1416/4 \times 84.30^{2 \times} 28.13$$

$$= 156968.61 \text{ ft}^3$$

$$N = 5.615 \times 1761675.259 / 15668.61 = 63 \text{ (year)}^{-1}$$

Therefore, in equation (3.13)

$$\mathbf{K_N} = (180+63) / (6 \times 63)$$

$$\mathbf{= 0.64}$$

In equation (2.2),

Working Loss, L_w = 0.0010 x 67.64 x 8.487 x 1761675.26 x 0.64 x 1

= 649982.50 lb/yr

= 649982.50 / (2.205 x 365 x 24) kg/hr

= 33.65 kg/hr

= (33.65*2.205*24)/(67.64*1000000) MMSCFD

= 0.019 MMSCFD

The summary view of working loss calculation for condensate storage tank is shown in the following Table 3.17:

Table 3.17: Summary of Working Loss Calculation for Condensate Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
M_V = Vapor Molecular Weight	67.64					
P_{VA} = Vapor pressure at T_{LA}			8.487	psi	-	
Q= Net Throughput	32	m^3/hr			1,761,675.26	bbl/yr
V_L = Tank normal Liquid Volume			156,968.61	ft3		
N= number of turnovers per year			63			
K_P= working loss product			1			
K_N = Working Loss Turnover (saturation) Factor			0.64			
Lw = Working Loss			**649,982.504**	**lb/yr**	**33.65**	**kg/hr**
					0.019	**MMSCFD**

Similarly, motor spirit tank vapor is obtained and recorded in the below Table 3.18:

Table 3.18: Summary of Working Loss Calculation for Motor Spirit Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
Mv = Vapor Molecular Weight	42.29					
P_{VA} = Vapor pressure at T_{LA}			11.3019	psi	-	
Q= Net Throughput	18	m^3/hr			973,876.10	bbl/yr
V_L = Tank normal Liquid Volume			118,824.48	ft3		
N= number of turnovers per year			43			
K_P= working loss product			1			
K_N = Working Loss Turnover (saturation) Factor			0.866			
Lw = Working Loss			**381,014.89**	**lb/yr**	**19.73**	**kg/hr**
					0.013	**MMSCFD**

3.2.3 Overall Vapor Loss Quantity from Tank

The summary view of standing loss and working loss from tanks of 4,000 BPD Condensate fractionation are tabulated here:

Table 3.19: Overall Vapor Loss Quantity from Tank

Sl No.	Source of Emission	Stock Content	Standing Losses, Kg/hr	Working Losses, Kg/hr
1	T-501	Condensate	36.98	
2	T-502	Condensate	36.98	
3	T-503	Condensate		33.65
4	T-504	Motor Spirit	47.10	
5	T-505	Motor Spirit	47.10	
6	T-506	Motor Spirit		19.73
Total			168.16	53.38
		Grand Total	**221.54**	**Kg/hr**
			136,945.00 x 10^{-6}	**MMSCFD**

3.3 Total Recoverable Hydrocarbon from Plant

From Simulation and tank loss calculation the summation of hydrocarbon volume to recover is as below:

Table 3.20: Volume of Recoverable Hydrocarbon for 4,000 BPD

Description	Mass Flow lb/hr	Volume
Off-Gas Volume	14.68	2,062.76 x 10^{-6} MMSCFD
Vapor Losses from Tank	-	136,945.00 x 10^{-6} MMSCFD
Total Recoverable Hydrocarbon Volume	-	**139,007.76 x 10^{-6} MMSCFD**

3.4 Vapor Recovery Unit (VRU) Design

The off-gas from the process have pressure 5 psig and low-pressure vapor from tanks with pressure 0.9 psig are entered in VRU system common header. As tank vapor pressure is low that is why a booster fan is used to increase the pressure same as off-gas. The three streams (Off-gas, MS Vapor & Condensate Vapor) passed through Mixer M-100 in HYSYS to obtain the VRU Pre Inlet Vapor, which is then connected to a 2-phase separator as, VRU Suction Scrubber for early liquid-gas separation (fig 4.3). The gas from the scrubber as "VRU Com Inlet Vapor" then feeds to VRU Compressor which is a reciprocating compressor. And the liquid goes to the condensate tank through a Close Drain Drum.

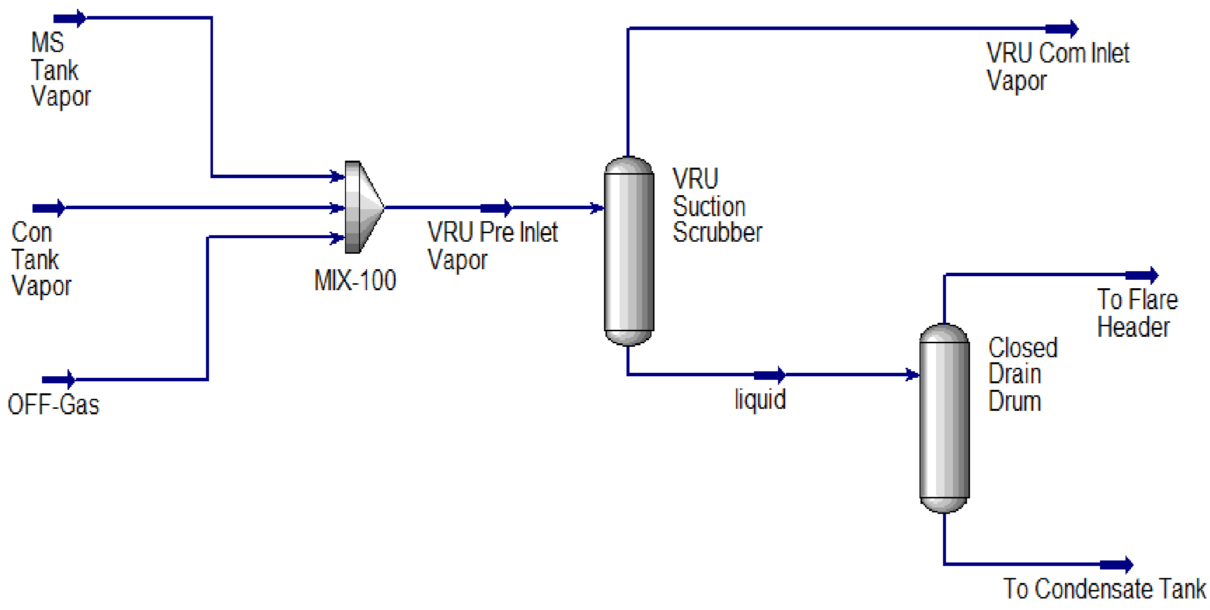


Figure 3.3: 4,000 BPD Plant VRU Compressor Inlet Scheme

The preliminary data gathered from the simulation for the design of the compressor are listed below:

Table 3.21: General Data for Compressor Design

Description	Data	Unit	Unit Conversion	Unit
Mv = Fluid Molecular Weight	51.84			
K @ Inlet Condition	1.10	Btu/lbmole-F		
Q = Flow Rate	0.139	MMSCFD		
E = Volumetric Eff. %	0.8		0.8	
P_S = Suction Pressure,	0.35	barg	19.845	psia
P_D = Discharge Pressure	6.5	barg	110.25	psia
P_b = Base Pressure,	14.7	psia		
T_b = Base Temperature	520	°R		

3.4.1 Vapor Recovery Unit (VRU) Compressor Capacity

The capacity of VRU $= 139{,}007.76 \times 10^{-6}$ MMSCFD

$= 165.22\ m^3/hr$

10% Margin for design Capacity $= 181.74\ m^3/hr$

3.4.2 Compression Ratio & Number of Stages

By the inlet and discharge pressure data from Table 3.21, compression ratio and number of stages can be calculated with equation (2.5):

$$r_{opt} = (P_d/P_S)^{1/n_S}$$

$$= (6.5/.35)^{1/2}$$

$$= 4.31\text{, which} < 6$$

As $n_s = 2$;

Therefore, Compressor is Double Stages

3.4.3 Horsepower Requirement

- Analytical Method:

From equation (2.6) & (2.7),

$$\text{Ideal Horsepower, IHP} = \frac{k}{k-1}\frac{3.027p_b}{T_b}qT_1\left[\left(\frac{p_2}{p_1}\right)^{z_1(k-1)/k}-1\right]$$

Actual/Brake Horsepower, BHP = IHP/E

Table 3.22 listed the preliminary data of pressure and temperature and calculated values to determine horsepower. The value of compressibility is gained from Appendix A.

Table 3.22: Inlet & Discharge Condition 1st Stage & 2nd Stage

1st Stage					
Name	**Description**	**Data**	**Unit**	**Calculated Value**	**Unit**
Inlet Condition	Pressure, P_1/Ps	0.35	barg	5.15	psig
	Temperature, T_1	68.01	°C	614.09	°R
	Pseudo-critical Pressure, P_{PC1}			512	psia
	Pseudo-critical Temperature, T_{PC1}			552	°R
	Pseudo-reduced Pressure, P_{PR1}			0.04	psia
	Pseudo-reduced Temperature, T_{PR1}			1.11	°R
	Compressibility, Z_1			0.989	
Discharge Condition	Pressure, P_2			1.51	barg
	Temperature, T_2			697.77	°R
2nd Stage					
Inlet Condition	Pressure, P_3=P_2	1.51	barg	22.17	psig
	Temperature, T_3= T_1	68.01	°C	614.09	°R
	Pseudo-critical Pressure, P_{PC3}			512	psia
	Pseudo-critical Temperature, T_{PC3}			552	°R
	Pseudo-reduced Pressure, P_{PR3}			0.07	psia
	Pseudo-reduced Temperature, T_{PR3}			1.11	°R
	Compressibility, Z_3			0.981	
Discharge Condition	Pressure, P_4	6.50	barg		
	Temperature, T_4			697.05	°R

1st Stage,

Ideal Horsepower, IHP_1 = 11.01 hp

Actual/Brake Horsepower, BHP_1 = 13.77 hp

2nd Stage,

Ideal Horsepower, IHP_2 = 10.92 hp

Actual/Brake Horsepower, BHP_2 = 13.65 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = $IHP_1 + IHP_2$

= 11.01+ 10.92 hp

= 21.93 hp

Total Break Horsepower, BHP = $BHP_1 + BHP_2$

= 13.77 + 13.65 hp

= 27.41 hp

- *Mollier Diagram Method:*

From equation (2.8) & (2.9),

Ideal Horsepower, IHP = $\frac{n(h_2 - h_1) \times 778 \cdot 2}{24 \times 60 \times 3300}$

Actual/Brake Horsepower, BHP = IHP/E

From the Enthalpy-Entropy diagram (Mollier Diagram) in Appendix A, Table 3.23 of data on the next page are obtained.

Table 3.23: Enthalpy-Entropy Data

Name	Description	Data	Unit
First Stage Constant Entropy	h_1 @ P1 & T1	1300	Btu/lbmole
	T2	210	°F
	h_2 @ P2 & T2	2400	Btu/lbmole
Intercooler at Constant Pressure	h_3 @ P2 & T1/3	1200	Btu/lbmole
Second Stage Constant Entropy	T4	260	°F
	h4 @ P4 & T4	3000	Btu/lbmole
Intercooler at Constant Pressure	h_5 @ P4 & T5/1/3	1100	Btu/lbmole

1st Stage,

Ideal Horsepower, IHP_1 = 7.38 hp

Actual/Brake Horsepower, BHP_1 = 9.22 hp

2nd Stage,

Ideal Horsepower, IHP_2 = 12.07 hp

Actual/Brake Horsepower, BHP_2 = 15.09 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = IHP_1 + IHP_2

= 7.38 + 12.07 hp

= 19.45 hp

Total Break Horsepower, BHP = BHP_1 + BHP_2

= 9.22 + 15.09 hp

= 24.31 hp

The calculated results of compressor capacity and horsepower are compared with the plant's actual compressor's data for comparison Table 3.24 below.

Table 3.24: Calculated Value vs Real Plant Value of Compressor

Description	Method	4000 BPD Capacity	
		Existing Field	Calculated
Capacity of VRU Compressor, m^3/hr	-	185	181.74
Compressor Stage	-	2	2
Ideal Horsepower Required for Compression, hp	Analytical	23	22.39
	Mollier Diagram		19.85
Break Horsepower Required for Compression, hp	Analytical	29	27.99
	Mollier Diagram		24.82

3.4.4 Cooling Requirement

- Analytical Method:

From equation (2.10),

Heat Removed by Intercooler/Aftercooler, $\Delta H = n\ \bar{C}_P \Delta T$

Table 3.25 & 3.26 shows the initial data of and calculated values to determine heat removed by intercooler and aftercooler. The value of average specific heat data is gained from Appendix A.

Table 3.25: Data for Inter Cooler Load Calculation

Name	Description	Data	Unit	Calculated Value	Unit
Intercooler	Number of Mole, n			409.52	lbmole/D
	Average Specific Heat $\overline{C_P}$ @ $(T_2+T_1)/2$ & P_2			22.64	Btu/ lbmole-F
	Inlet Temperature, T_2	697.77	°R	237.77	°F
	Outlet Temperature, $T_3=T_1$	614.09	°R	154.09	°F
	$\Delta T= (T_3-T_2)$			83.68	°F

Heat Removed by Intercooler, **ΔH_1 = 7.76 x 10^5 Btu/D**

Table 3.26: Data for After Cooler Load Calculation

Name	Description	Data	Unit	Calculated Value	Unit
Aftercooler	Number of Mole, n	409.52	lbmole/D		
	$\overline{C_P}$ @ $(T_4+T_1)/2$ & P4			23.16	Btu/ lbmole-F
	Inlet Temperature, T_4/T_3	697.05	°R	237.05	°F
	Outlet Temperature, $T_5=T_1$	614.09	°R	154.09	°F
	$\Delta T=(T_3-T_2)$			82.96	°F

Heat Removed by Aftercooler, $\mathbf{\Delta H_2 = 7.87 \times 10^5}$ **Btu/D**

- Mollier Diagram Method:

From equation (2.11),

Heat Removed by Intercooler/Aftercooler, $\Delta H = n\,(h_2-h_3)$

From Enthalpy-Entropy data in Table 3.23 & 3.25

Heat Removed by Intercooler, $\mathbf{\Delta H_1 = 4.91 \times 10^5}$ **Btu/D**

Heat Removed by Aftercooler, $\mathbf{\Delta H_2 = 7.78 \times 10^5}$ **Btu/D**

Off-gas volume from simulation matched with real plant volume. Tanks' vapor amount from a real plant could not be obtained as it varies day to day due to the fluctuation of temperature and unloading operation. That is why, at first the volume of vapor for 4,000 BPD is calculated analytically. With that volume and amount of off-gas the VRU compressor capacity, required horsepower are calculated. The power requirement for the compressor is maximum as in the calculation, option to use a VFD was not included. Then it is matched with the currently running VRU compressor's data from the site (as shown in Table 3.24) for comparison. The similarity of value gives the confidence to carry out further calculations.

Chapter 4

CALCULATION FOR ADDITIONAL 6,000 BPD CAPACITY

Based on the 4,000 BPD simulation of feed, the off-gas quantity for 6,000 BPD unit will be calculated. In the later section in this chapter prior to estimation of tank vapor loss required assumption and initial data will be discussed

4.1 Off-Gas Estimation from Column

Because of the increased capacity of the plant to run the simulation, several adjustments have to be made, which is different from the previous. When the column converges and all other connections of equipment are fitting, the required data of off-gas and other parameters are obtained. The HYSYS Scheme for 6,000 BPD Condensate Fractionation Plant is shown in Figure 4.1.

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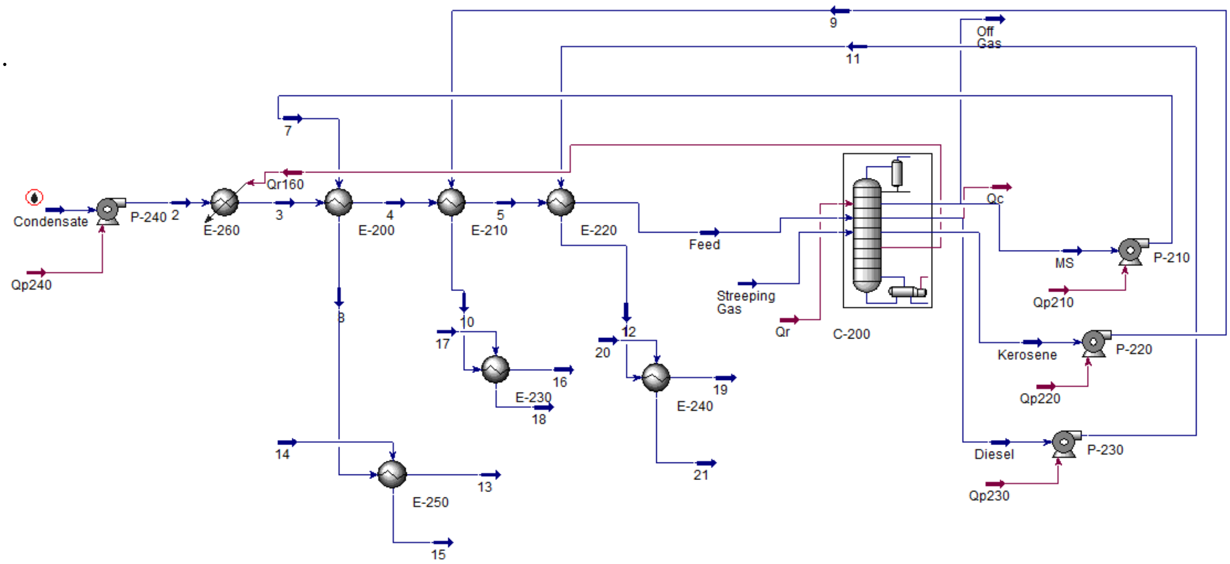

Feed		
Temperature	90.00	C
Pressure	43.51	psig
Mass Flow	6.815e+004	lb/hr

MS		
Temperature	76.00	C
Pressure	5.000	psig
Mass Flow	4.413e+004	lb/hr

Kerosene		
Temperature	199.0	C
Pressure	16.00	psig
Mass Flow	2.105e+004	lb/hr

Diesel		
Temperature	350.0	C
Pressure	18.00	psig
Mass Flow	2976	lb/hr

Off Gas		
Temperature	76.00	C
Pressure	5.000	psig
Mass Flow	22.02	lb/hr

Figure 4.1: HYSYS Scheme for 6,000 BPD Condensate Fractionation Plant

The sub-flow sheet of the column environment is shown in Figure 4.2.

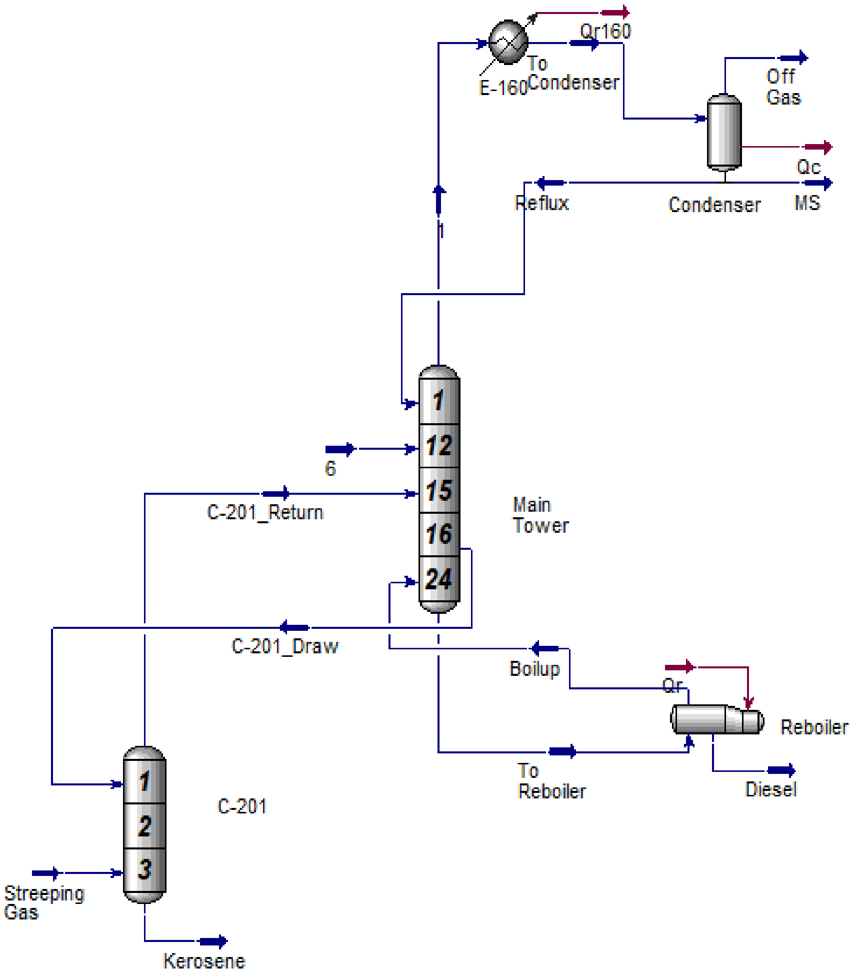

Figure 4.2: Sub-flow Sheet of Column Environment

The simulated result for products is in Table 4.1 below.

Table 4.1: Simulation Product Value of 6,000 BPD Plant

Product Name	**Simulated Value**		
	Mass Flow lb/hr	**Mass Flow kg/hr**	**Percent**
Condensate	6.815×10^4	3.091×10^4	100 %
Motor Spirit	4.413×10^4	2.002×10^4	64.76 %
Kerosene	2.105×10^4	9546	30.88 %
Diesel	2976	1350	43.68 %
Off-Gas	22.02	9.99	0.32 %

Table 4.2 exhibits off-gas estimation after simulation.

Table 4.2: Exhibits Off-Gas Estimation

Stream Name	Off Gas	Liquid phase
Vapor/ Phase Fraction	1.0000	0.0000
Temperature, °C	76.00	76.00
Pressure, psig	5.000	5.000
Molar Flow, lbmole/hr	0.3858	0.3858
Mass Flow, lb/hr	22.02	22.02
Std Ideal Liq Vol, bbl/day	2.418	2.418
Molar Enthalpy, Btu/lbmole	-5.085×10^4	-5.085×10^4
Molar Entropy, Btu/lbmole-F	35.48	35.48
Heat Flow, Btu/hr	-1.962 x 104	-1.962 x 104
Liq. Vol Flow @ Std Cond, bbl/day	2.416	2.416
Fluid Package	Basis-1	

4.2 Tank Vapor Losses Calculation

At present, the size and quantity of the storage tanks can provide the feed to process and store the products from the process for 38 days if the feed intake operation interrupts. Based on that viewpoint, the quantity of the storage tank for 6,000 BPD are shown in below Table 4.3:

Table 4.3: Tank Preliminary Data for 6,000 BPD

Tank Quantity		Vapor Generation Tank
1) Condensate Storage Tank	- 5 nos (T 601- T 605)	1) Condensate Storage Tank
2) MS Storage Tank	- 4 nos (T 606- T 610)	2) MS Storage Tank
3) Kerosene Storage Tank	- 7 nos (T 611- T 618)	
4) Diesel Storage Tank	- 1 nos (T 609)	

Based on the discussion in chapter 3, we will calculate tank vapor loss if the tanks are internal floating roof type and if fixed roof type in 2 (two) scenarios.

4.2.1 Scenario 1: Fixed Roof tank

In this scenario, we will consider the newly constructed tanks that can be fixed roof tanks similar to the existing plant.

4.2.1.1 Standing Loss Calculation for Condensate & Motor Spirit Tank

For fixed roof tanks in 6,000 BPD unit, all the parameters will be the same to calculate the standing loss, except the quantity of the tank will increase. Therefore, standing loss for one condensate storage tank will be:

Table 4.4: Standing Loss for Condensate Fixed Roof Tank

Description	Calculated Value	Unit	Unit Conversion	Unit
L_S = Standing Loss (for Condensate)	**714,245.07**	**lb/yr**	**36.98**	**kg/hr**
			0.021	**MMSCFD**

Similarly, for one motor spirit tank vapor is recorded in the below Table 4.5:

Table 4.5: Standing Loss for Motor Spirit Fixed Roof Tank

Description	Calculated Value	Unit	Unit Conversion	Unit
L_S = Standing Loss (for Motor Spirit)	**909,859.49**	**lb/yr**	**47.10**	**kg/hr**
			0.032	**MMSCFD**

4.2.1.2 Working Loss Calculation for Condensate & Motor Spirit Tank

For working loss calculation, all the parameters will be the same as the present unit except the net throughput and the quantity of the tank will increase. Therefore, working loss for one condensate storage tank is shown in the below Table 4.6:

Table 4.6: Working Loss for Condensate Fixed Roof Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
M_V = Vapor Molecular Weight	67.64					
P_{VA} = Vapor pressure at T_{LA}			8.487	psi	-	
Q= Net Throughput	40	m^3/hr			2,202,094.07	bbl/yr
V_L = Tank normal Liquid Volume			156,968.61	ft3		
N= number of turnovers per year			79			
K_P= working loss product			1			
K_N = Working Loss Turnover (saturation) Factor			0.55			
L_W = Working Loss			**692,119.785**	**lb/yr**	**35.83**	**kg/hr**
					0.020	**MMSCFD**

Similarly, working loss of motor spirit tank vapor is recorded in Table 4.7 below:

Table 4.7: Working Loss for Motor Spirit Fixed Roof Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
M_V = Vapor Molecular Weight	42.29					
P_{VA} = Vapor pressure at T_{LA}			11.30	psi	-	
Q= Net Throughput	27	m^3/hr			1,486,413.50	bbl/yr
V_L = Tank normal Liquid Volume			118,824.48	ft^3		
N= number of turnovers per year			70			
K_P= working loss product factor			1			
K_N = Working Loss Turnover (saturation) Factor			0.59			
L_W = Working Loss			**421,843.47**	**lb/yr**	**21.84**	**kg/hr**
					0.015	**MMSCFD**

4.2.2 Scenario 2: Internal Floating Roof tank

In this scenario, initially, to reduce the vapor loss, the tanks for the new unit can be internal floating roof types. Preliminary data for calculation are shown in the table below (4.8 to 4.11). The equations used in this section are taken from "Compilation of Air Pollutant Emissions Factors (AP-42)" of The US Environmental Protection Agency [9].

4.2.2.1 Standing Loss Calculation for Condensate & Motor Spirit Tank

From equation (2.3), the standing loss has several parameters which need to be evaluated step by step.

$$\textbf{Standing Loss, } L_S = L_R + L_F + L_D$$

a) Rim Seal Loss, L_R

Rim seal loss from floating roof tanks can be estimated using the following equation:

$$L_R = (K_{Ra} + K_{Rb} \times v^n) D \times P^* \times M_V \times K_C \text{ ....(4.1)}$$

Here,

L_R = rim seal loss, lb/yr

K_{Ra} = zero wind speed rim seal loss factor, lb-mole/ft * yr

K_{Rb} = wind speed dependent rim seal loss factor, lb-mole/(mph) nft * yr

v = average ambient wind speed at tank site, mph

n = seal-related wind speed exponent, dimensionless

P* = vapor pressure function, dimensionless

$$= \frac{\frac{P_{vA}}{P_A}}{\left(1+\left[1-\left(\frac{p_{vA}}{P_A}\right)\right]^{0.5}\right)^2} \quad \text{.. (4.2)}$$

Where,

P_{VA} = vapor pressure at average daily liquid surface temperature, psia;

PA = atmospheric pressure, psia

D = tank diameter, ft

M_V = average vapor molecular weight, lb/lb-mole

K_C = product factor

Now, from equation (3.8)

- ***Vapor Pressure at Average Daily Liquid Surface Temperature***

$$P = \exp\left\{\left[0.7553 - \left(\frac{413.0}{T+469.6}\right)\right] S^{0.5}\log(RVP) - \left[1.854 - \left(\frac{1{,}042}{T+469.6}\right)\right] S^{0.5} + \left[\left(\frac{2{,}416}{T+469.6}\right)\log(RVP) - \left(\frac{8{,}742}{T+469.6}\right) + 15.64\right\}\right.$$

- ***Daily Average Liquid Surface Temperature***

$T_{LA} = 0.33 \times T_{AA} + 0.7 \times T_B + 0.004 \times \alpha \times I$

$= 0.33 \times ((T_{AX}+T_{AN})/2) + 0.7 \times ((T_{AX}+T_{AN})/2) + 6\alpha - 1) + 0.004 \times \alpha \times I$

..................................(4.3)

Table 4.8: Data for Daily Average Liquid Surface Temperature Calculation

Description	Data	Unit	Unit Conversion	Unit
T_{AX} = Daily Maximum Ambient Temperature	39	°C	561.87	°R
T_{AN} = Daily Minimum Ambient Temperature	5	°C	500.67	°R
α = Paint Factor	0.54		-	
I = Daily Minimum Ambient Temperature	7920	BTU/ft^2 day	-	
RVP = Raid vapor pressure	7.1	psi		
S = Slope of the ASTM Distillation Curve at 10% Evaporation	7.17	°F/Percent		

Therefore,

T_{LA} = 0.33*((561.87+500.67)/2) +0.7*(((561.87+500.67)/2) +6*0.54-1)

+0.004*0.54*7920

= 565.9 °R

P_{VA} = EXP ((0.7533- (413/ (565.9 +459.6))) x 7.17$^{\wedge 0.5}$ x $\log_{10}$(7.1) -(1.854-(1042/

(565.9 +459.6))) x 7.17$^{0.5}$+((2416/ (565.9 +459.6))-2.013) x $\log_{10}$(7.1) -(8742/

(565.9 +459.6)) +15.64) = 8.43 psia

Table 4.9: Data for Vapor Pressure Function Calculation

Description	Data	Unit
P_{VA} = Vapor pressure at T_{LA}	8.43	psia
P_A = Atmospheric Pressure	14.7	psi
K_{Ra} = zero wind speed rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.3	lb-mole/ft-yr
K_{Rb} = wind speed dependent rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.6	lb-mole/((mph)n-ft-yr))
K_c = product factor (for all other organic liquids, except crude oil)	1	
M_V = Molecular weight	67.64	
v = Average ambient wind speed	0	
n = Seal-related wind speed exponent (For Liquid-mounted seal & Rim-mounted secondary)	0.3	
D = Tank Diameter	84.30	ft

Therefore, **P*** = (8.49/14.7) / (1+(1- (8.49/14.7))$^{0.5}$)^2

= 0.210

And, **LR =** (0.3+ 0.6 X 0^0.3) x 84.30 x 0.210 x 67.64 x 1

= 359.109 lb/yr

= 0.0.19 kg/hr

= 10.42 x 10^{-6} MMSCFD

b) Deck Fitting Loss, LF

Deck fitting losses from floating roof tanks can be estimated by the following equation:

$L_F = F_F \times P^* \times M_V \times K_C$ ……………………………………………………………(4.4)

Where,

L_F = the deck fitting loss, lb/yr

F_F = total deck fitting loss factor, lb-mole/yr

$= [(N_{F1} \times K_{F1}) + (N_{F2} \times K_{F2}) + \ldots\ldots\ldots + (N_{Fnf} \times K_{Fnf})]$ ………………….(4.5)

Here:

N_{Fi} = number of deck fittings of a particular type (i = 0,1,2,...,nf) dimensionless

K_{Fi} = deck fitting loss factor for a particular type fitting (i = 0,1,2,...,nf), lbmole/yr

nf = total number of different types of fittings, dimensionless

Also,

$N_{F5} = (5 + D/10 + D^2/600)$ …………………… ………………………(4.6)

Table 4.10: Data for Deck Fitting Loss Calculation

Description	Data	Unit
K_{Fa1} = Access hatches for Unbolted cover, ungasketed	36	lb-mole/yr
N_{F1} = number of deck fittings of a K_{Fa1}	2	
K_{Fa2} = Gauge float well for Unbolted cover, ungasketed	14	lb-mole/yr
N_{F2} = number of deck fittings of a K_{Fa2}	1	
K_{Fa3} = pipe column well for round pipe, flexible fabric sleeve seal	10	lb-mole/yr
N_{F3} = number of deck fittings of a K_{Fa3}	1	
K_{Fa4} = Ladder well for Sliding cover, gasketed	56	lb-mole/yr
N_{F4} = number of deck fittings of a K_{Fa4}	1	
K_{Fa5} = Deck legs for Adjustable, internal floating deck	7.9	lb-mole/yr
K_{Fa6} = Sample pipe well for Un-gasketed or gasketed sliding cover	43	lb-mole/yr
N_{F6} = number of deck fittings of a K_{Fa6}	1	
K_{Fa7} = Vacuum breaker Weighted mechanical actuation, gasketed	6.2	lb-mole/yr
N_{F7} = number of deck fittings of a K_{Fa7}	1	

From Equation (4.6),

$$N_{F5} = (5 + 84.30/10 + 84.30^2/ 600)$$

$$= 25.27$$

From equation (4.5),

$$F_F = 36 \text{ x } 2 + 14 \text{ x } 1 + 10 \text{ x } 1 + 56 \text{ x } 1 + 7.9 \text{ x } 25.27 + 43 \text{ x } 1 + 62 \text{ x } 1$$

$$= 400.85 \text{ lb-mole/yr}$$

Therefore, in equation (4.4),

$$\mathbf{L_F} = 4000.85 \text{ x } 0.210 \text{ x } 1 \text{ x } 67.64$$

$$= 5692.250 \text{ lb/yr}$$

$$= 0.295 \text{ kg/hr}$$

$$= 165.10 \text{ x } 10^{-6} \text{ MMSCFD}$$

c) Deck Seam Loss, LD

Deck seam loss can be estimated by the following equation:

$$L_R = K_D \times S_D \times D^2 \times P^* \times M_V \times K_C \text{ ....(4.7)}$$

where:

K_D = deck seam loss per unit seam length factor, lb-mole/ft-yr

S_D = deck seam length factor, ft/ft^2

Table 4.11: Data for Deck Seam Loss Calculation

Description	Data	Unit
K_D = deck seam loss per unit seam length factor (For welded deck)	0	lb-mole/ft-yr

In equation (4.7),

Therefore, $\mathbf{L_R}$ **=** 0 lb/yr

In equation (2.3),

Standing Loss, Ls = (359.109 + 5692.250 + 0) lb/yr

= 6051.359 lb/yr

= 0.313 kg/hr

= 165.10 x 10^{-6} MMSCFD

The summary view of standing loss calculation for condensate storage tank is shown in the following Table 4.12 in the next page:

Table 4.12: Summary of Standing Loss Calculation for Condensate Internal Floating Roof Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
T_{AX} = Daily Maximum Ambient Temperature	39	°C			561.87	°R
T_{AN} = Daily Minimum Ambient Temperature	5	°C			500.67	°R
α = Paint Factor (For Fixed Roof Tank)	0.54				-	
I = Daily Minimum Ambient Temperature	7920	BTU/ft^2 day			-	
RVP = Raid vapor pressure	7.1	psi				
S = Slope of the ASTM Distillation Curve at 10% Evaporation	7.17	°F/Percent				
T_{LA} = Daily Average Liquid Surface Temperature			565.9	°R		
P_{VA} = Vapor pressure at T_{LA}			8.43	psia		
P_A = Atmospheric Pressure	14.7	psi				
P* = Vapor Pressure Function			0.210			
K_{Ra} = zero wind speed rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.3	lb-mole/ft-yr				
K_{Rb} = wind speed dependent rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.6	lb-mole/((mph) n-ft-yr))				
Kc = product factor (for all other organic liquids, except crude oil))	1					
M_V = Molecular weight	67.64					
v = average ambient wind speed	0					
D = Tank Diameter	84.30	ft				
n = seal-related wind speed exponent (For Liquid-mounted seal & Rim-mounted secondary)	0.3					
K_{Fa1} = Access hatches for Unbolted cover, ungasketed	36	lb-mole/yr				
N_{F1} = number of deck fittings of a K_{Fa1}	2					
K_{Fa2} = Gauge float well for Unbolted cover, ungasketed	14	lb-mole/yr				
N_{F2} = number of deck fittings of a K_{Fa2}	1					
K_{Fa3} = pipe column well for Round pipe, flexible fabric sleeve seal	10	lb-mole/yr				
N_{F3} = number of deck fittings of a K_{Fa3}	1					
K_{Fa4} = Ladder well for Sliding cover, gasketed	56	lb-mole/yr				
N_{F4} = number of deck fittings of a K_{Fa4}	1					
K_{Fa5} = Deck legs for AdjusTable, internal floating deck	7.9	lb-mole/yr				
N_{F5} = number of deck fittings of a K_{Fa5}			25.27			
K_{Fa6} = Sample pipe well for Ungasketed or gasketed sliding cover	43	lb-mole/yr				
N_{F6} = number of deck fittings of a K_{Fa6}	1					

Table 4.12: Summary of Standing Loss Calculation for Condensate Internal Floating Roof Tank (contd.)

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
K_{Fa7} = Vacuum breaker Weighted mechanical actuation, gasketed	6.2	lb-mole/yr				
N_{F7} = number of deck fittings of a K_{Fa7}	1					
F_F = total deck fitting loss factor			400.85	lb-mole/yr		
K_D = deck seam loss per unit seam length factor (For welded deck)	0	lb-mole/ft-yr				
L_R = Rim Deck Loss			359.11	lb/yr	0.019	kg/hr
					10.48×10^{-6}	MMSFCD
L_F = Deck Fitting Losses			5692.25	lb/yr	0.29	kg/hr
					166.16×10^{-6}	MMSFCD
L_S = Standing Losses			**6051.36**	**lb/yr**	**0.31**	**kg/hr**
					166.16×10^{-6}	**MMSFCD**

Similarly, motor spirit tank vapor is obtained and recorded in the below Table 4.13:

Table 4.13: Summary of Standing Loss Calculation for Motor Spirit Internal Floating Roof Tank

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
T_{AX} = Daily Maximum Ambient Temperature	39	°C			561.87	°R
T_{AN} = Daily Minimum Ambient Temperature	5	°C			500.67	°R
α = Paint Factor (For Fixed Roof Tank)	0.54				-	
I = Daily Minimum Ambient Temperature	7920	BTU/ft^2 day			-	
RVP = Raid vapor pressure	9.5	psi				
S = Slope of the ASTM Distillation Curve at 10% Evaporation	3.91	°F/Percent				
T_{LA} = Daily Average Liquid Surface Temperature			565.9	°R		
P_{VA} = Vapor pressure at T_{LA}			11.22	psia		
P_A = Atmospheric Pressure	14.7	psi				
P* = Vapor Pressure Function			0.346			
K_{Ra} = zero wind speed rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.3	lb-mole/ft-yr				

Table 4.13: Summary of Standing Loss Calculation for Motor Spirit Internal Floating Roof Tank (contd.)

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
K_{Rb} = wind speed dependent rim seal loss factor (For Liquid-mounted seal & Rim-mounted secondary)	0.6	lb-mole/((mph)n-ft-yr))				
M_V = Molecular weight	42.29					
v = average ambient wind speed	0					
D = Tank Diameter	76.75	ft				
Kc = product factor (for all other organic liquids, except crude oil)	1					
n = seal-related wind speed exponent (For Liquid-mounted seal & Rim-mounted secondary)	0.3					
K_{Fa1} = Access hatches for Unbolted cover, ungasketed	36	lb-mole/yr				
N_{F1} = number of deck fittings of a K_{Fa1}	2					
K_{Fa2} = Gauge float well for Unbolted cover, ungasketed	14	lb-mole/yr				
N_{F2} = number of deck fittings of a K_{Fa2}	1					
K_{Fa3} = pipe column well for Round pipe, flexible fabric sleeve seal	10	lb-mole/yr				
N_{F3} = number of deck fittings of a K_{Fa3}	1					
K_{Fa4} = Ladder well for Sliding cover, gasketed	56	lb-mole/yr				
N_{F4} = number of deck fittings of a K_{Fa4}	1					
K_{Fa5} = Deck legs for AdjusTable, internal floating deck	7.9	lb-mole/yr				
N_{F5} = number of deck fittings of a K_{Fa5}			25.27			
K_{Fa6} = Sample pipe well for Ungasketed or gasketed sliding cover	43	lb-mole/yr				
N_{F6} = number of deck fittings of a K_{Fa6}	1					
K_{Fa7} = Vacuum breaker Weighted mechanical actuation, gasketed	6.2	lb-mole/yr				
N_{F7} = number of deck fittings of a K_{Fa7}	1					
F_F = total deck fitting loss factor			378.90	lb-mole/yr		
K_D = deck seam loss per unit seam length factor (For welded deck)	0	lb-mole/ft-yr				
L_R = Rim Deck Loss			225.29	lb/yr	0.012	kg/hr
					7.90×10^{-6}	MMSFCD

Table 4.13: Summary of Standing Loss Calculation for Motor Spirit Internal Floating Roof Tank (contd.)

Description	Data	Unit	Calculated Value	Unit	Unit Conversion	Unit
L_F = Deck Fitting Losses			5539.47	lb/yr	0.29	kg/hr
					194.29×10^{-6}	MMSFCD
L_D = Deck Seam Losses			0	lb/yr		
L_S = Standing Losses			**5764.75**	**lb/yr**	**0.30**	**kg/hr**
					202.19×10^{-6}	**MMSFCD**

4.2.2.2 Working Loss Calculation for Condensate & Motor Spirit Tank

The working loss from floating roof storage tanks, also known as withdrawal loss, can be estimated using Equation (2.4) in Table 4.14

$$\textbf{Working Loss, } L_W = \frac{0{\cdot}943 Q C_s W_L}{D}\left(1 + \frac{N_C F_C}{D}\right)$$

where:

L_W = working (withdrawal) loss, lb/yr

Q = annual net throughput, bbl/yr

C_S = shell clingage factor, bbl/1,000 ft2

W_L = average organic liquid density, lb/gal

D = tank diameter, ft

0.943 = constant, 1,000 ft3gal/bbl^2

N_C = number of fixed roof support columns, dimensionless

F_C = effective column diameter, ft (column perimeter [ft]/π)

Table 4.14: Summary of Working Loss Calculation for Condensate Internal Floating Roof Tank

Description	Data	Unit	Unit Conversion	Unit
Nc = number of fixed roof support columns	1			
Fc = effective column diameter (Assumed)	1	ft (column perimeter [ft]/π)		
CS = shell clingage factor	0.0040	bbl/1,000 ft2		
WL = average organic liquid density	6.49	lb/gal		
Q= Net Throughput	40	m3/hr	2202094.07	bbl/yr
LWD = Withdrawal Loss	**647.09**	**lb/yr**	**0.34**	**kg/hr**
			18.89 x 10^{-6}	**MMSCFD**

Similarly, for motor spirit tank working loss is obtained and recorded in the below Table 4.15:

Table 4.15: Summary of Working Loss Calculation for Motor Spirit Internal Floating Roof Tank

Description	Data	Unit	Unit Conversion	Unit
Nc = number of fixed roof support columns	1			
Fc = effective column diameter (Assumed)	1	ft (column perimeter [ft]/π)		
CS = shell clingage factor	0.0040	bbl/1,000 ft2		
WL = average organic liquid density	6.09	lb/gal		
Q= Net Throughput	27	m3/hr	1486413.50	bbl/yr
LWD = Withdrawal Loss	**450.67**	**lb/yr**	**0.023**	**kg/hr**
			15.81 x 10^{-6}	**MMSCFD**

The standing loss and working loss quantity of individual tanks for 6,000 BPD will be used to determine the overall hydrocarbon volume for 10,000 BPD condensate fractionation plant in the next chapter. The extended unit installation can be either with the fixed roof or internal floating roof type tanks. It is observed that there is a significant difference in vapor generation between these two types, which is an important factor for the calculation of the VRU capacity and other calculations.

Chapter 5

VRU DESIGN FOR 10,000 BPD CAPACITY

This chapter would reflect the key objective of this project work (design a vapor recovery unit for hydrocarbon recovery from the storage tank and off-gas of a condensate fractionation plant of 10,000 BPD capacity) by discussing and summarizing the values from chapter 3 and 4.

5.1 Off-Gas Estimation from Column

From chapter 3 & 4, the quantity of off-gas generated from the column are listed in Table 5.1 below:

Table 5.1: Volume of Recoverable Hydrocarbon for 6,000 BPD Unit

Sources	Off-Gas Volume
4,000 BPD Unit (Simulated)	2,062.76 x 10^{-6} MMSCFD
6,000 BPD Unit (Simulated)	3,094.14 x 10^{-6} MMSCFD
For 10,000 BPD Plant	**5,156.90 x 10^{-6} MMSCFD**

5.2 Tank Vapor Losses Calculation

As discussed in the previous chapters, parallel tank operation is carried out for the same type of fuel storage tank. Therefore, the same procedures would be followed in this chapter i.e.one tank is assumed as working and the rest of the tanks are standing. Also, two scenarios will be discussed for both fixed roof and internal floating roof tanks.

5.2.1 Scenario 1: Fixed Roof Tank

In chapter 3, the working loss amount for each tank of 4,000 BPD unit is less than the amount calculated in chapter 4 which was for each tank of 6,000 BPD. Although the construction and all data were similar, net throughput was higher for 6,000 BPD unit than 4,000 BPD. In scenario 1, the working loss amount will be used from 6,000 BPD unit to design the VRU because the maximum volume of hydrocarbon needs to be calculated as a common practice. The total volume of fixed roof tank vapor for 10,000 BPD will be the summation of 4,000 BPD and 6,000 BPD unit standing loss and working loss amount, which is tabulated here:

Table 5.2: Overall Vapor Loss Quantity from Tank

Sl No.	Source of Emission	Unit Name	Stock Content	Standing Losses, Kg/hr	Working Losses, Kg/hr
1	T-501	4,000 BPD	Condensate	36.98	
2	T-502		Condensate	36.98	
3	T-503		Condensate	36.98	
4	T-601	6,000 BPD	Condensate	36.98	
5	T-602		Condensate	36.98	
6	T-603		Condensate	36.98	
7	T-604		Condensate	36.98	
8	T-605		Condensate		35.83
9	T-504	4,000 BPD	Motor Spirit	47.10	
10	T-505		Motor Spirit	47.10	
11	T-506		Motor Spirit	47.10	
12	T-606	6,000 BPD	Motor Spirit	47.10	
13	T-607		Motor Spirit	47.10	
14	T-608		Motor Spirit	47.10	
15	T-609		Motor Spirit		21.84
Total				541.47	57.67
Grand Total				**599.14**	**Kg/hr**
				369,946.24 x 10^{-6}	**MMSCFD**

5.2.1.1 Total Recoverable Hydrocarbon from Plant

From Table 5.1 & 5.2 the summation of Hydrocarbon volume to be recovered is as below:

Table 5.3: Volume of Recoverable Hydrocarbon for 6,000 BPD

Description	Volume
Off-Gas Volume	5,156.90 x 10^{-6} MMSCFD
Vapor Losses from Tank	369,946.24 x 10^{-6} MMSCFD
Total Recoverable Hydrocarbon Volume	**375,103.14 x 10^{-6} MMSCFD**

5.2.2 Scenario 2: Internal Floating Roof Tank

In this scenario, the working loss amount will be used from 4,000 BPD unit to design the VRU because the amount is larger than 6,000 BPD internal floating roof tanks. Similar to the fixed roof tank, for 10,000 BPD the total volume of the internal floating roof tank vapor will be the summation of 4,000 BPD & 6,000 BPD unit standing loss and working loss amount, which is tabulated below in Table 5.4:

Table 5.4: Overall Vapor Loss Quantity from Tank

Sl No.	Source of Emission	Unit Name	Stock Content	Standing Losses, Kg/hr	Working Losses, Kg/hr
1	T-501	4,000 BPD	Condensate	36.98	
2	T-502		Condensate	36.98	
3	T-503		Condensate		33.65
4	T-601	6,000 BPD	Condensate	0.31	
5	T-602		Condensate	0.31	
6	T-603		Condensate	0.31	
7	T-604		Condensate	0.31	
8	T-605		Condensate	0.31	
9	T-504	4,000 BPD	Motor Spirit	47.10	
10	T-505		Motor Spirit	47.10	
11	T-506		Motor Spirit		19.73
12	T-606	6,000 BPD	Motor Spirit	0.30	
13	T-607		Motor Spirit	0.30	
14	T-608		Motor Spirit	0.30	
15	T-609		Motor Spirit	0.30	
Total				170.91	53.38
Grand Total				**224.29**	**Kg/hr**
				138,620.71 x 10^{-6}	**MMSCFD**

5.2.2.1 Total Recoverable Hydrocarbon from Plant

From Table 5.1 & 5.4, the summation of hydrocarbon volume to be recovered is as below:

Table 5.5: Volume of Recoverable Hydrocarbon for 6,000 BPD

Description	Volume
Off-Gas Volume	5,156.90 x 10^{-6} MMSCFD
Vapor Losses from Tank	138,620.71 x 10^{-6} MMSCFD
Total Recoverable Hydrocarbon Volume	**143,777.61 x 10^{-6} MMSCFD**

5.3 Vapor Recovery Unit (VRU) Design

Hydrocarbon volume to be recovered by VRU for 10,000 BPD unit has been estimated in Table 5.3 and 5.5 for two different scenarios. The streams from 4,000 BPD and 6,000 BPD unit will go to the VRU compressor through mixer 200 and suction scrubber as shown in below Figure 5.1.

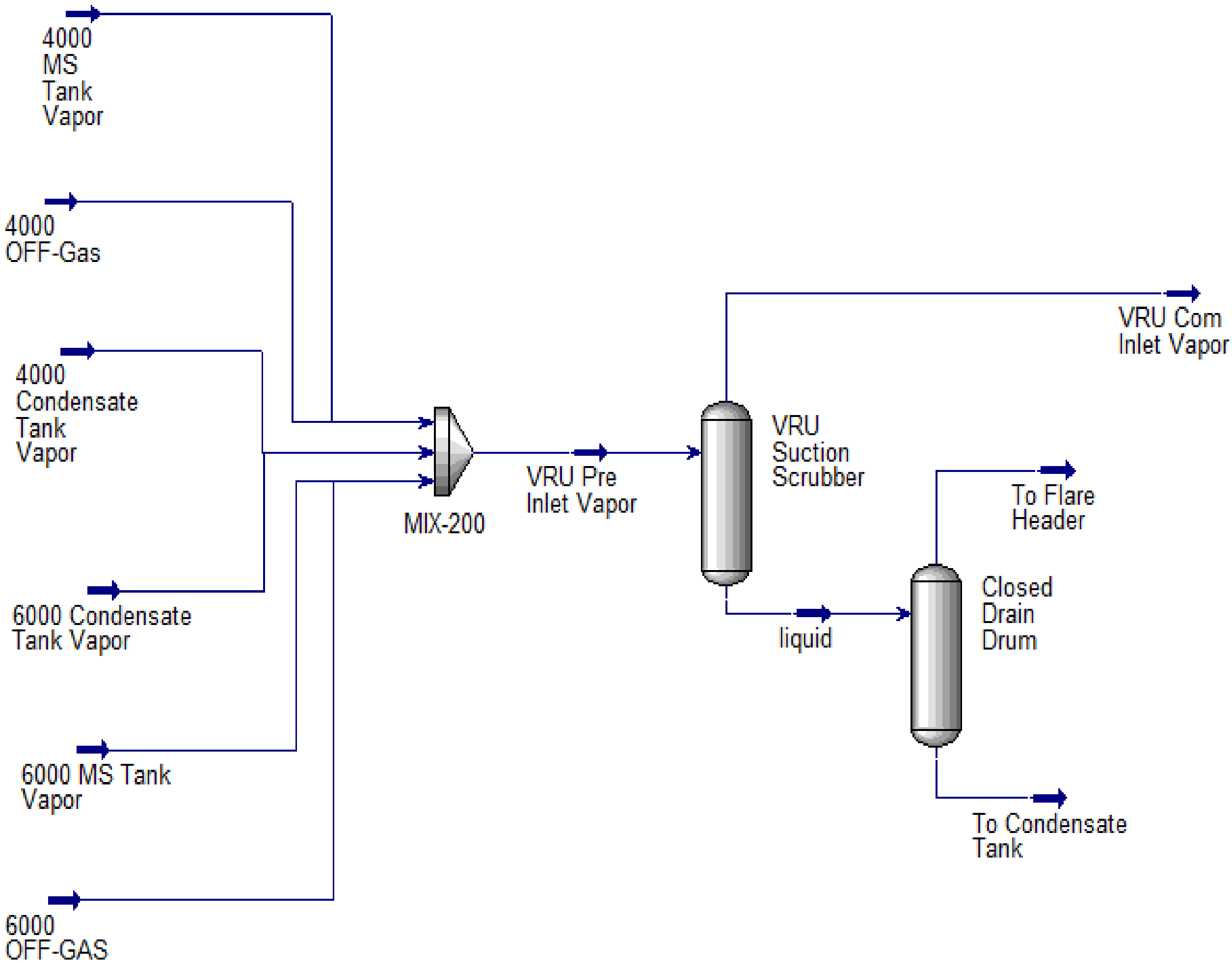

Figure 5.1: 10,000 BPD Plant VRU Compressor Inlet Scheme

As the estimation was calculated for two scenarios, the VRU compressor design would also be described for those two scenarios.

5.3.1 Scenario 1: Fixed Roof tank

The preliminary data gathered from the simulation for the design of the compressor are listed below:

Table 5.6: General Data for 10,000 BPD Unit Compressor Design (Scenario-1)

Description	Data	Unit	Unit Conversion	Unit
Mv = Fluid Molecular Weight	53.09			
K @ Inlet Condition	1.12	Btu/ lbmole-F		
Q = Flow Rate	0.376	MMSCFD		
E = Volumetric Eff. %	0.8		0.8	
P_S = Suction Pressure,	0.35	barg	19.845	psia
P_D = Discharge Pressure	6.5	barg	110.25	psia
P_b = Base Pressure,	14.7	psia		
T_b = Base Temperature	520	°R		

5.3.1.1 Vapor Recovery Unit (VRU) Compressor Capacity

The capacity of VRU = 375,103.14 x 10^{-6} MMSCFD

= 445.84 m^3/hr

10% Margin for design Capacity = 490.43 m^3/hr

5.3.1.2 Compression Ratio & Number of Stages

Compression Ratio & stages are same as previous capacity plant:

r_{opt} = 4.31, which < 6

n_s = 2; therefore, Compressor is Double Stages

5.3.1.3 Horsepower Requirement

- Analytical Method:

Table 5.7 listed the preliminary data of pressure and temperature and calculated values to determine horsepower. The value of compressibility is gained from Appendix A.

Table 5.7: Inlet & Discharge Condition 1st Stage & 2nd Stage (Scenario-1)

1st Stage					
Name	**Description**	**Data**	**Unit**	**Calculated Value**	**Unit**
Inlet Condition	Pressure, P_1/Ps	0.35	barg	5.15	psig
	Temperature, T_1	69.82	°C	617.35	°R
	Pseudo-critical Pressure, P_{PC1}			620	psia
	Pseudo-critical Temperature, T_{PC1}			640	°R
	Pseudo-reduced Pressure, P_{PR1}			0.03	psia
	Pseudo-reduced Temperature, T_{PR1}			0.96	°R
	Compressibility, Z_1			0.980	
Discharge Condition	Pressure, P_2			1.51	barg
	Temperature, T_2			715.57	°R
2nd Stage					
Inlet Condition	Pressure, P_3=P_2	1.51	barg	22.17	psig
	Temperature, T_3= T_1	69.82	°C	617.35	°R
	Pseudo-critical Pressure, P_{PC3}			620	psia
	Pseudo-critical Temperature, T_{PC3}			640	°R
	Pseudo-reduced Pressure, P_{PR3}			0.06	psia
	Pseudo-reduced Temperature, T_{PR3}			0.96	°R
	Compressibility, Z_3			0.990	
Discharge Condition	Pressure, P_4	6.50	barg		
	Temperature, T_4			716.65	°R

From equation (2.6) & (2.7),

1st Stage,

Ideal Horsepower, IHP_1 = 30.56 hp

Actual/Brake Horsepower, BHP_1 = 38.20 hp

2nd Stage,

Ideal Horsepower, IHP_2 = 30.90 hp

Actual/Brake Horsepower, BHP_2 = 38.62 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = $IHP_1 + IHP_2$

= 30.56 + 30.90 hp

= 61.46 hp

Total Break Horsepower, BHP = $BHP_1 + BHP_2$

= 38.20 + 38.62 hp

= 76.82 hp

- Mollier Diagram Method:

From the Enthalpy-Entropy diagram (Mollier Diagram) in Appendix A, Table 5.8 of data are obtained.

Table 5.8: Enthalpy-Entropy Data (Scenario-1)

Name	Description	Data	Unit
First Stage Constant Entropy	h_1 @ P_1 & T_1	1600	Btu/lbmole
	T_2	210	°F
	h_2 @ P_2 & T_2	2900	Btu/lbmole
Intercooler at Constant Pressure	h_3 @ P_2 & $T_{1/3}$	1300	Btu/lbmole
Second Stage Constant Entropy	T_4	255	°F
	h4 @ P_4 & T_4	3000	Btu/lbmole
Intercooler at Constant Pressure	h_5 @ P_4 & $T_{5/1/3}$	1200	Btu/lbmole

From equation (2.8) & (2.9),

1st Stage,

Ideal Horsepower, IHP_1 = 23.91 hp

Actual/Brake Horsepower, BHP_1 = 29.90 hp

2nd Stage,

Ideal Horsepower, IHP_2 = 31.27 hp

Actual/Brake Horsepower, BHP_2 = 39.09 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = $IHP_1 + IHP_2$

= 23.91 + 31.27 hp

= 55.18 hp

Total Break Horsepower, BHP = $BHP_1 + BHP_2$

= 29.90 + 39.09 hp

= 69.98 hp

5.3.1.4 Cooling Requirement

- Analytical Method:

Table 5.9 & 5.10 shows the initial data and calculated values to determine heat removed by intercooler and aftercooler. The value of average specific heat data is gained from Appendix - A.

Table 5.9: Data for Inter Cooler Load Calculation (Scenario-1)

Name	Description	Data	Unit	Calculated Value	Unit
Intercooler	Number of Mole, n			1123.24	lbmole/D
	Average Specific Heat $\overline{C_P}$ @ $(T_2+T_1)/2$ & P_2			18.56	"Btu/
	Inlet Temperature, T_2	lbmole-F"			
	Outlet Temperature, $T_3=T_1$	715.57	°R	255.57	°F
	$\Delta T= (T_3-T_2)$	617.35	°R	157.35	°F

From equation (2.10)

Heat Removed by Intercooler, $\mathbf{\Delta H_1 = 20.47 \times 10^5}$ **Btu/D**

Table 5.10: Data for Aftercooler Load Calculation (Scenario-1)

Name	Description	Data	Unit	Calculated Value	Unit
Aftercooler	Number of Mole, n	1,123.24	lbmole/D		
	Average Specific Heat $\overline{C_P}$ @ $(T_4+T_1)/2$ & P_4			16.76	"Btu/
	Inlet Temperature, T_4	lbmole-F"			
	Outlet Temperature, $T_5=T_1$	716.65	°R	256.65	°F
	$\Delta T= (T_3-T_2)$	617.35	°R	157.35	°F

From equation (2.11)

Heat Removed by Aftercooler, $\mathbf{\Delta H_2 = 18.69 \times 10^5}$ **Btu/D**

- ***Mollier Diagram Method:***

From the Enthalpy-Entropy data in Table 5.6 & 5.8 in equation (2.11),

Heat Removed by Intercooler, $\mathbf{\Delta H_1 = 17.97 \times 10^5}$ **Btu/D**

Heat Removed by Aftercooler, $\mathbf{\Delta H_2 = 20.22 \times 10^5}$ **Btu/D**

5.3.2 Scenario 2: Internal Floating Roof tank

The preliminary data gathered from the simulation for the design of the compressor are listed below:

Table 5.11: General Data for 10,000 BPD Unit Compressor Design (Scenario-2)

Description	Data	Unit	Unit Conversion	Unit
Mv = Fluid Molecular Weight	37.66			
K @ Inlet Condition	1.14	Btu/ lbmole-F		
Q = Flow Rate	0.144	MMSCFD		
E = Volumetric Eff. %	0.8		0.8	
P_S = Suction Pressure,	0.35	barg	19.845	psia
P_D = Discharge Pressure	6.5	barg	110.25	psia
P_b = Base Pressure,	14.7	psia		
T_b = Base Temperature	520	°R		

5.3.2.1 Vapor Recovery Unit (VRU) Compressor Capacity

The capacity of VRU = 143,777.61 x 10^{-6} MMSCFD

= 170.89 m^3/hr

10% Margin for design Capacity = 187.98 m^3/hr

5.3.2.2 Compression Ratio & Number of Stages

Compression Ratio & stages are same as previous capacity plant:

r_{opt} = 4.31, which < 6

n_s = 2; therefore, Compressor is Double Stages

5.3.2.3 Horsepower Requirement

- Analytical Method:

Table 5.12 listed the preliminary data of pressure and temperature and calculated values to determine horsepower. The value of compressibility is gained from Appendix A.

Table 5.12: Inlet & Discharge Condition 1st Stage & 2nd Stage (Scenario-2)

1st Stage					
Name	**Description**	**Data**	**Unit**	**Calculated Value**	**Unit**
Inlet Condition	Pressure, P_1/Ps	0.35	barg	5.15	psig
	Temperature, T_1	69.82	°C	617.35	°R
	Pseudo-critical Pressure, P_{PC1}			633	psia
	Pseudo-critical Temperature, T_{PC1}			570	°R
	Pseudo-reduced Pressure, P_{PR1}			0.03	psia
	Pseudo-reduced Temperature, T_{PR1}			1.08	°R
	Compressibility, Z_1			0.980	
Discharge Condition	Pressure, P_2			1.51	barg
	Temperature, T_2			733.57	°R

Table 5.12: Inlet & Discharge Condition 1st Stage & 2nd Stage (Scenario-2) (Contd.)

2nd Stage					
Description	**Data**	**Unit**	**Calculated Value**	**Unit**	
Inlet Condition	Pressure, P_3=P_2	1.51	barg	22.17	psig
	Temperature, T_3= T_1	69.82	°C	617.35	°R
	Pseudo-critical Pressure, P_{PC3}			633	psia
	Pseudo-critical Temperature, T_{PC3}			570	°R
	Pseudo-reduced Pressure, P_{PR3}			0.06	psia
	Pseudo-reduced Temperature, T_{PR3}			1.08	°R
	Compressibility, Z_3			0.990	
Discharge Condition	Pressure, P_4	6.50	barg		
	Temperature, T_4			734.86	°R

From equation (2.6) & (2.7),

1st Stage,

Ideal Horsepower, IHP_1 = 11.89 hp

Actual/Brake Horsepower, BHP_1 = 14.86hp

2nd Stage,

Ideal Horsepower, IHP_2 = 12.02 hp

Actual/Brake Horsepower, BHP_2 = 15.02 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = IHP_1 + IHP_2

= 11.89 + 12.02 hp

= 23.90 hp

Total Break Horsepower, BHP = BHP_1 + BHP_2

= 14.86+ 15.02 hp

= 29.88 hp

- Mollier Diagram Method:

From the Enthalpy-Entropy diagram (Mollier Diagram) in Appendix A, Table 5.13 of data are obtained.

Table 5.13: Enthalpy-Entropy Data (Scenario-2)

Name	Description	Data	Unit
First Stage Constant Entropy	h_1 @ P_1 & T_1	1600	Btu/lbmole
	T_2	210	°F
	h_2 @ P_2 & T_2	2900	Btu/lbmole
Intercooler at Constant Pressure	h_3 @ P_2 & $T_{1/3}$	1300	Btu/lbmole
Second Stage Constant Entropy	T_4	255	°F
	h_4 @ P_4 & T_4	3000	Btu/lbmole
Intercooler at Constant Pressure	h_5 @ P_4 & $T_{5/1/3}$	1200	Btu/lbmole

From equation (2.8) & (2.9),

1st Stage,

Ideal Horsepower, IHP_1 = 9.18 hp

Actual/Brake Horsepower, BHP_1 = 11.48 hp

2nd Stage,

Ideal Horsepower, IHP_2 = 12.01 hp

Actual/Brake Horsepower, BHP_2 = 15.01 hp

Therefore, from 1st & 2nd Stage the total horsepower will be:

Total Ideal Horsepower, IHP = $IHP_1 + IHP_2$

= 9.18 + 12.01 hp

= 21.19 hp

Total Break Horsepower, BHP = $BHP_1 + BHP_2$

= 11.48 + 15.01 hp

= 26.49 hp

5.3.2.4 Cooling Requirement

- Analytical Method:

Table 5.14 & 5.15 shows the initial data of and calculated values to determine heat removed by intercooler and aftercooler. The value of average specific heat data is gained from Appendix A.

Table 5.14: Data for Inter Cooler Load Calculation (Scenario-2)

Name	Description	Data	Unit	Calculated Value	Unit
Intercooler	Number of Mole, n			431.32	lbmole/D
	Average Specific Heat $\overline{C_P}$ @ $(T_2+T_1)/2$ & P2			15.78	"Btu/
	Inlet Temperature, T_2	lbmole-F"			
	Outlet Temperature, $T_3=T_1$	733.57	°R	273.57	°F
	$\Delta T=(T_3-T_2)$	617.35	°R	157.35	°F

From equation (2.10)

Heat Removed by Intercooler, $\mathbf{\Delta H_1 = 7.91 \times 10^5}$ **Btu/D**

Table 5.10: Data for Aftercooler Load Calculation (Scenario-2)

Name	Description	Data	Unit	Calculated Value	Unit
Aftercooler	Number of Mole, n	431.32	lbmole/D		
	Average Specific Heat $\overline{C_P}$ @ $(T_4+T_1)/2$ & P_4			13.98	"Btu/
	Inlet Temperature, T_4	lbmole-F"			
	Outlet Temperature, $T_5=T_1$	734.86	°R	274.86	°F
	$\Delta T=(T_3-T_2)$	617.35	°R	157.35	°F

From equation (2.11)

Heat Removed by Aftercooler, $\mathbf{\Delta H_2 = 7.09 \times 10^5}$ **Btu/D**

- ***Mollier Diagram Method:***

From the Enthalpy-Entropy data in Table 5.11 & 5.13 in equation (2.11),

Heat Removed by Intercooler, $\boldsymbol{\Delta H_1}$ = $\mathbf{6.90 \times 10^5}$ **Btu/D**

Heat Removed by Aftercooler, $\boldsymbol{\Delta H_2 = 7.76 \times 10^5}$ **Btu/D**

A model has been developed to estimate the hydrocarbon volume for the 4,000 BPD capacity plant and to design a VRU compressor in chapter 3. Based on that model, in chapter 4, the hydrocarbon volume has been determined for an additional 6,000 BPD capacity plant. With all the data from previous chapters, the overall calculation to the design of a VRU system for 10,000 BPD plant has been made in this chapter. The results of these calculations will be discussed in chapter 7.

Chapter 6

ENVIRONMENTAL BENEFITS OF VRU SYSTEM

Flaring and venting of associated gas impact local, environmental and public health, as well as contributes to potential climate change. VRU System prevents the potential release of hydrocarbons and CO_2 to the environment.

6.1 Environmental Concerns

Gas flaring is a major environmental concern the world is facing today as it generates a significant amount of greenhouse gases specially CO_2, which contributes to the overall burden of global warming. The amount of flared global gas was 150 billion cubic meters in 2019 which is the highest level in more than a decade. In 2018, flared gas amount (145 billion cubic meter) worldwide was roughly equal to the total gas demand of the Africa Continent with emissions of approximately 275 Mt CO_2 [12][13].

Venting of unburned hydrocarbons also releases other chemical compounds into the atmosphere specially methane CH_4, which pollutes the air and may result in health risks to the local population as well as increase the greenhouse effect. Methane is 25 times more impactful as a greenhouse gas than CO_2 over a 100 years' timeline [14]. As per the information from U.S Energy Information Administration, in the USA, 538,479 million cubic feet of gas was released into the environment in 2019 through venting and flaring [15].

Vent gas may contain volatile organic compounds (VOCs), Hazardous air pollutants (HAPs), Greenhouse gases (GHGs), Toxic gases. Flaring produces CO_2 and methane when burnt, the major two GHG gasses which have a serious impact on Global Warming Potential (GWP). Also flaring causes some local environmental problems, such as thermal and heat radiating, agricultural impact, etc. [16].

6.1.1 Estimation of CO2 Emissions and Heat Radiation by Flaring

According to the United States Environmental Protection Agency, Carbon dioxide emissions per MMBTU of gas are determined by multiplying the carbon coefficient times the fraction oxidized times the ratio of the molecular weight of carbon dioxide to carbon (44/12). The average carbon coefficient of pipeline natural gas burned in 2017 is 14.43 kg carbon per MMBTU (EPA 2019). The fraction oxidized to CO_2 is assumed to be 100 percent (IPCC 2006).

Then Carbon dioxide emissions per MMBTU flared gas can be converted to carbon dioxide emissions per thousand cubic feet (MSCF) using the average heat content of natural gas in 2017, 10.37 therms/Mcf (EIA 2019) times 0.1 MMBTU/1 therms [17].

CO_2 Emission = 14.43 kg C/MMBTU × 44 kg CO_2/12 kg C × 1 metric ton/1,000 kg × 10.37 therms/MSCF × 0.1 MMBTU/1 therms

= 0.0549 metric tons/MSCF Flared gas ..……………..…………………… (6.1)

Heat radiation in the vicinity of the flare area can be calculated by considering the heat content of natural gas 10.37 therms/MSCF (although the vented gas contains higher heat content).

Heat Radiation Amount by flare = Flared Gas Volume x 10.37 therms/MSCF gas x 0.1 MMBTU/1therms…………………………..……. (6.2)

For 4000 BPD Plant (theoretically, in absence of VRU)

The existing plant has a VRU system that recovers hydrocarbons from the column and tank area. But if there would be no VRU system the off-gas would go to flare and produce CO_2. To understand the amount of emitted CO_2 from flare the below calculation would be is presented.

From equation (6.1)

CO_2 Emission = 0.0549x 2062.76 x 10^{-3} MSCFD x metric tons/MSCF x 365 D/year

= 40.15 metric tons/year

Also, by burning hydrocarbons, the generated heat amount is calculated from equation. (6.2)

Heat Radiation Amount by flare = 2062.76 x 10^{-3} MSCFD x 10.37 therms/MSCF x 0.1 MMBTU/1therms x 365 D/year

= 7.81 x 10^{8} BTU/year

For 10,000 BPD Plant (theoretically, in absence of VRU)

Similar to 4,000 BPD the CO_2 emission volume in absence of VRU will be:

CO_2 Emission = 0.0549 x 5,156.90 x 10^{-3} MSCFD x metric tons/MSCF x 365 D/year

= 103.58 metric tons/year

From equation (6.2)

Heat Radiation Amount by flare = 5,156.90 x 10^{-3} MSCFD x 10.37 therms/MSCF x 0.1 MMBTU/1 therms x 365 D/year

= 19.52 x 10^{8} BTU/year

By installing the VRU system, venting is eliminated and emissions of dangerous pollutants are reduced, which demonstrates a commitment to social responsibility and environmental protection. Also resulting in increased revenue or cost reduction for the facility.

Chapter 7

RESULTS & DISCUSSIONS

7.1 Summary of Results

Results from chapter 4,5 and 6 are summarized in Table 7.1:

Table 7.1: Results in Tabular Form

Description	Method	4000 BPD Capacity		10,000 BPD Capacity (Proposed)	
		Existing Field	Simulated / Calculated	Fixed Roof	Internal Floating Roof
Off-Gas Volume, MMSCFD	-	3,066.01 x 10^{-6}	2,062.76 x 10^{-6}	5,156.90 x 10^{-6}	-
Vapor Losses from Tank, MMSCFD	-	-	136,945.00 x 10^{-6}	369,946.24 x 10^{-6}	138,620.71 x 10^{-6}
Capacity of VRU Compressor, m^3/hr	-	185	181.74	490.43	187.98
Compressor Stage	-	2	2	2	2
Ideal Horsepower Required for Compression, hp	Analytical	23	22.39	61.46	23.90
	Mollier Diagram		19.85	55.18	21.19
Break Horsepower Required for Compression, hp	Analytical	29	27.99	76.82	29.88
	Mollier Diagram		24.82	69.98	26.49
Heat Removed by Intercooler, Btu/D	Analytical	-	7.92 x 10^5	20.47 x 10^5	7.91 x 10^5
	Mollier Diagram		5.012 x 10^5	17.97 x 10^5	6.90 x 10^5
Heat Removed by Aftercooler, Btu/D	Analytical	-	8.03 x 10^5	18.69 x 10^5	7.09 x 10^5
	Mollier Diagram		7.94 x 10^5	20.22 x 10^5	7.76 x 10^5
CO_2 Emission, metric tons/year	-	-	40.15	103.58	
Amount of Heat Radiation by Flare, Btu/year	-	-	7.81 x 10^8	19.52 x 10^8	

7.2 Discussion

For the sake of discussion, it is important to keep in mind that the current plant capacity is 4,000 BPD. The main plan is to increase the capacity to 10,000 BPD. Therefore, an additional 6,000 BPD unit will be constructed. The philosophy is to uphold zero loss of hydrocarbons by using the VRU system. To develop a VRU system for a fractionation plant the key component is the compressor, where off-gas and tank vapor enters for pressurization. The amount of these hydrocarbons and other design criteria of VRU compressor are discussed below:

- In the above Table 8.1, a model has been developed to estimate the hydrocarbon volume for 4,000 BPD capacity plant and to design a VRU compressor in chapter 3. Based on that model, in chapter 4, the hydrocarbon volume has been determined for an additional 6,000 BPD capacity plant. 4,000 BPD plant's off-gas volume and the volume generated in the simulation are nearly the same, which verifies the further estimation for larger capacity. To obtain the total amount of off-gas for 10,000 BPD, the quantity for 6,000 BPD unit is simulated based on the same assumptions and same preliminary data of feed used for 4,000 BPD simulation. Therefore, the off-gas for the new unit is 2.5 times larger by quantity than the 4,000 BPD plant as predicted.
- It is worth mentioning that, in simulation, although the feed amount increased from 4,000 BPD to 6,000 BPD, the simulation ran successfully with the same tray number (24 trays). For any engineering design, efficiency and cost-effectiveness are the primary concerns. Therefore, sieve trays are used for existing plant and also will be considered for the future unit [18].
- For a 10,000 BPD plant, the vapor generated from the tank batteries varies significantly based on the types of tanks that will be constructed in an additional 6,000 BPD unit. If tanks are made as fixed roof type, same as the old unit, the volume of vapor for 10,000 BPD unit will be roughly 2.7 times higher than 4,000 BPD plant, whereas if floating roof type tanks are constructed the volume will be 1.01 times higher than the existing plant, which almost the same amount.
- Tank vapor losses will be reduced by approximately 2.69 times in the internal floating roof type than the current tank (fixed roof) type. From a technical point of view, floating roof tanks will be preferable as slightly increased vapor will be generated. But as the economic aspect is not the scope of this work, the construction cost can be studied as future work.

- For fixed roof type, there is an important fact that although the plant's capacity will be upgraded to 2.5 times, the vapor generation will be 2.7 times. The main reason is that the working loss volume will be increased as it depends on the net throughput of the tank. And for additional 6,000 BPD unit net throughput is higher than 4,000 BPD unit with the same diameter tanks.
- Based on the data presented in the above Table, the capacity of the compressor at present will not be able to handle the volume of gas for the proposed 10,000 BPD capacity. A new compressor needs to be installed either for fixed roof type or internal floating roof type tanks.
- Similar to the nonlinearity of tank vapor's amount to the plant's overall capacity change, the compressor's capacity modification is also not linear. Because 2.5 times larger volume off-gas combined with 2.7 times increased tank vapor will be the inlet gas for the compressor.
- For a fixed roof tank, 2.65 times the higher capacity with 2.64 - 2.41 times higher horsepower (than the existing one) compressor is needed for compression work of 10,000 BPD Plant. That means an additional 305.43 m^3/hr hydrocarbon will be required to compress by a new compressor along with the existing compressor.
- On the other hand, if an internal floating roof tank will be installed, then only 3 m^3/hr additional capacity compressor needs to be added. Also, the horsepower requirement will be very negligible.
- The main reason for this significant difference of capacity between the two scenarios is mainly the vast difference of vapor loss generation from tanks based on their construction.
- If there was no VRU system, by flaring the 4,000 BPD unit would produce the same amount of CO_2 roughly as 9 (nine) average cars produce yearly [19]. For 10,000 BPD unit, therefore, it will be the same as 23 (twenty-three) average cars' yearly CO_2 emission amount.
- Another important fact is that if there were no VRU system, the volume of vented gas from the tanks would be 26-72 times larger than the volume of off-gas produced from the column which would be flared. And it is already discussed that vented gas emission is 25 times more hazardous than CO_2 emission. Therefore, VRU will provide a cleaner environment for on-site personnel and inhabitants near the vicinity.

Chapter 8

CONCLUSIONS AND RECOMMENDATIONS

Quantity of the two main components of the vapor recovery unit- off-gas and tank vapors have been estimated. Rather than an analytical approach to calculate the volume of off-gas, process simulation software is used to make a model similar to the existing operational process. Then data from simulation and real plant data are compared to validate the simulation model.

Tank vapor volume is calculated analytically initially for the current plant. During loss estimation, the inputted data were - daily tank operation procedure, tank mechanical data, meteorological conditions etc. After that, these two values are fed to the VRU compressor design calculation for the present plant. Then the results are confirmed with the existing (at the site) compressor's data.

Once the overall procedure to design a VRU system has been developed and justified, the whole method has been repeated for the proposed larger capacity plant which is the main objective of this project. During design for the future plant, two separate VRU system calculations have been shown for two different tank types.

It is observed that, only due to this different tank construction method, vapor generation volume varies significantly which has a substantial impact on the calculation of total VRU compressor capacity, horsepower requirement and on other values. Last but not least, the environmental importance of installing a VRU system in a plant has been estimated by several factors which have direct impacts on the surroundings.

From an economic point of view, a huge number of hydrocarbons would be wasted every year as a byproduct if VRU is not installed. Therefore, it is recommended to conduct a proper economic analysis of the VRU system for this proposed new plant as future work.

Another probable recommendation might be the generation of tank vapor in the laboratory for analysis, as the amount calculated in this work could not be verified with real plant data. It can be performed by taking the real plant's feed and products, producing analogous tank vapor by creating a plant environment in the laboratory to verify with the amount discussed in this project work

Moreover, the detailed calculation and effectiveness of using variable frequency drive (VFD) in the VRU compressor for production proficiencies and energy savings can be discussed.

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APPENDIX A

Gas Deviation Factor

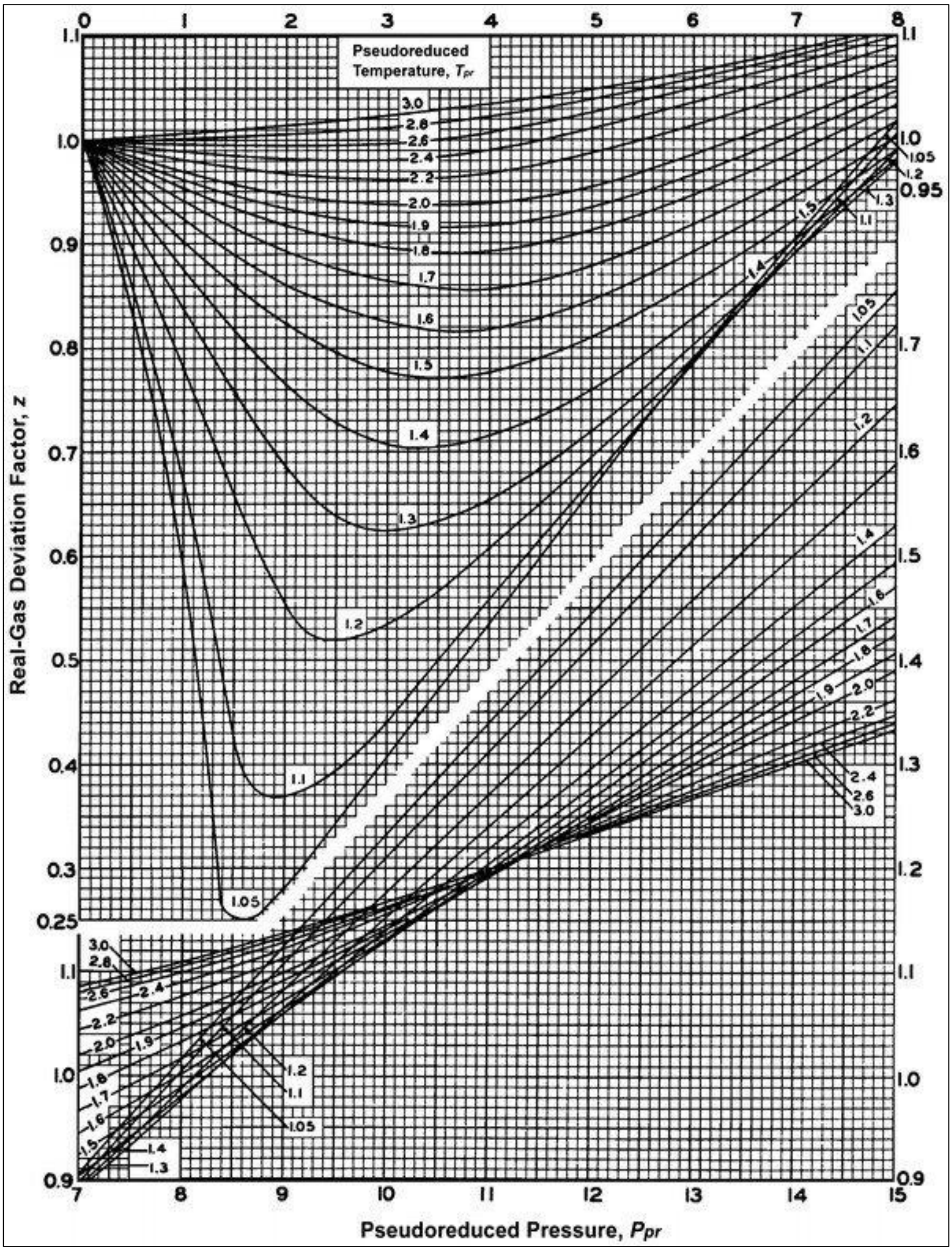

Figure A-1: Gas deviation factor for natural gas

(Source: Chi U. Ikoku, 1998, Natural Gas Production Engineering)

$$P_{PC} = 709.604 - 58.718\ \gamma_g$$

$$T_{PC} = 170.491 + 307.344\ \gamma_g$$

Where,

P_{PC} **= Pseudocritical Pressure, psia**

T_{PC} **= Pseudocritical Temperature, °R**

γ_g **= Specific Gravity**

Figure A-2: Equation for Pseudocritical Pressure & Temperature

(Source: Chi U. Ikoku, 1998, Natural Gas Production Engineering)

$$P_{PR} = P / P_{PC}$$

$$T_{PR} = T / T_{PC}$$

Where,

P_{PR} **= Pseudocritical Pressure, psia**

T_{PR} **= Pseudocritical Temperature, °R**

Figure A-3: Equation for Pseudocritical Pressure & Temperature

(Source: Chi U. Ikoku, 1998, Natural Gas Production Engineering)

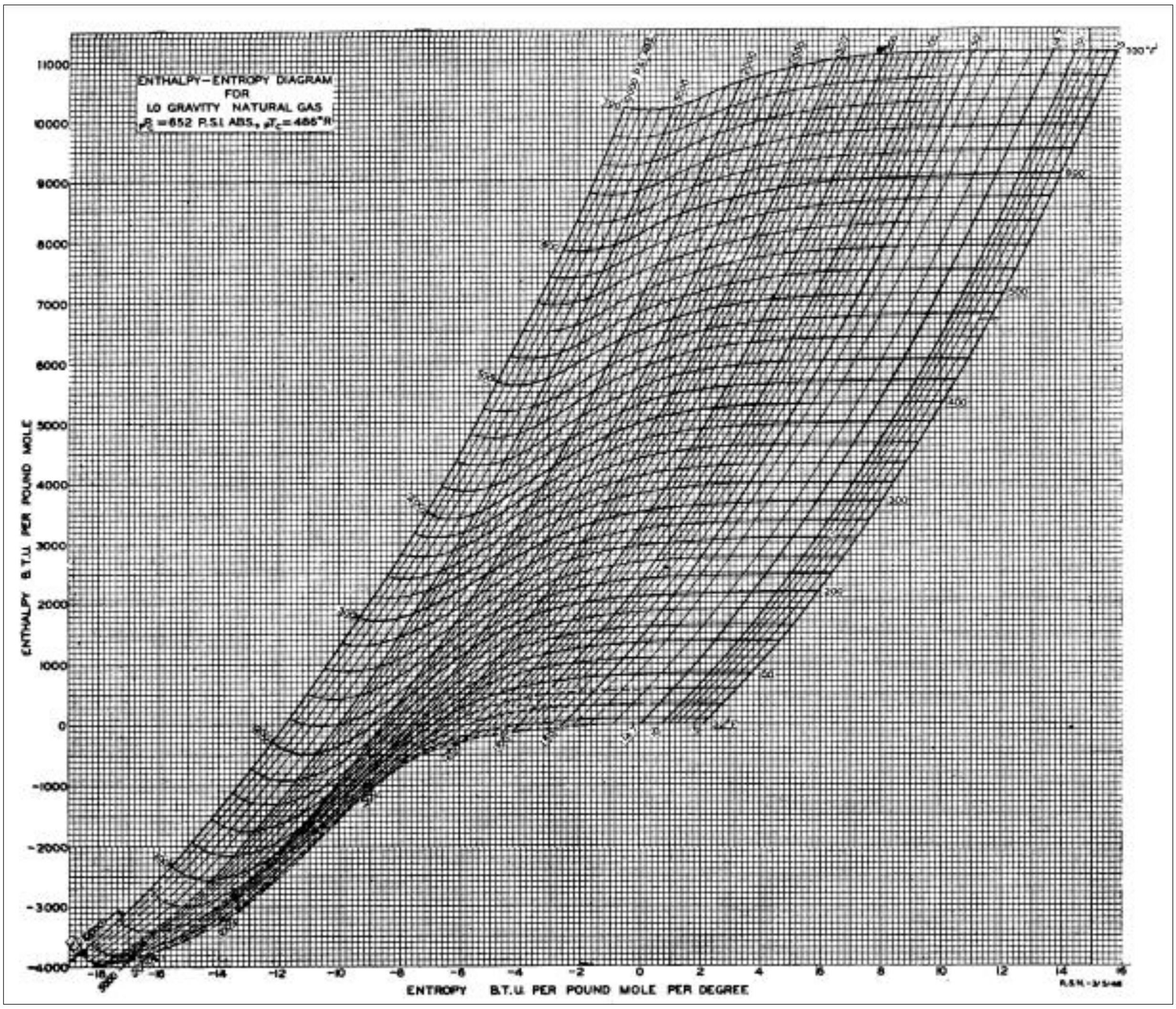

Figure A-4: Mollier Diagram

(Source: Chi U. Ikoku, 1998, Natural Gas Production Engineering)

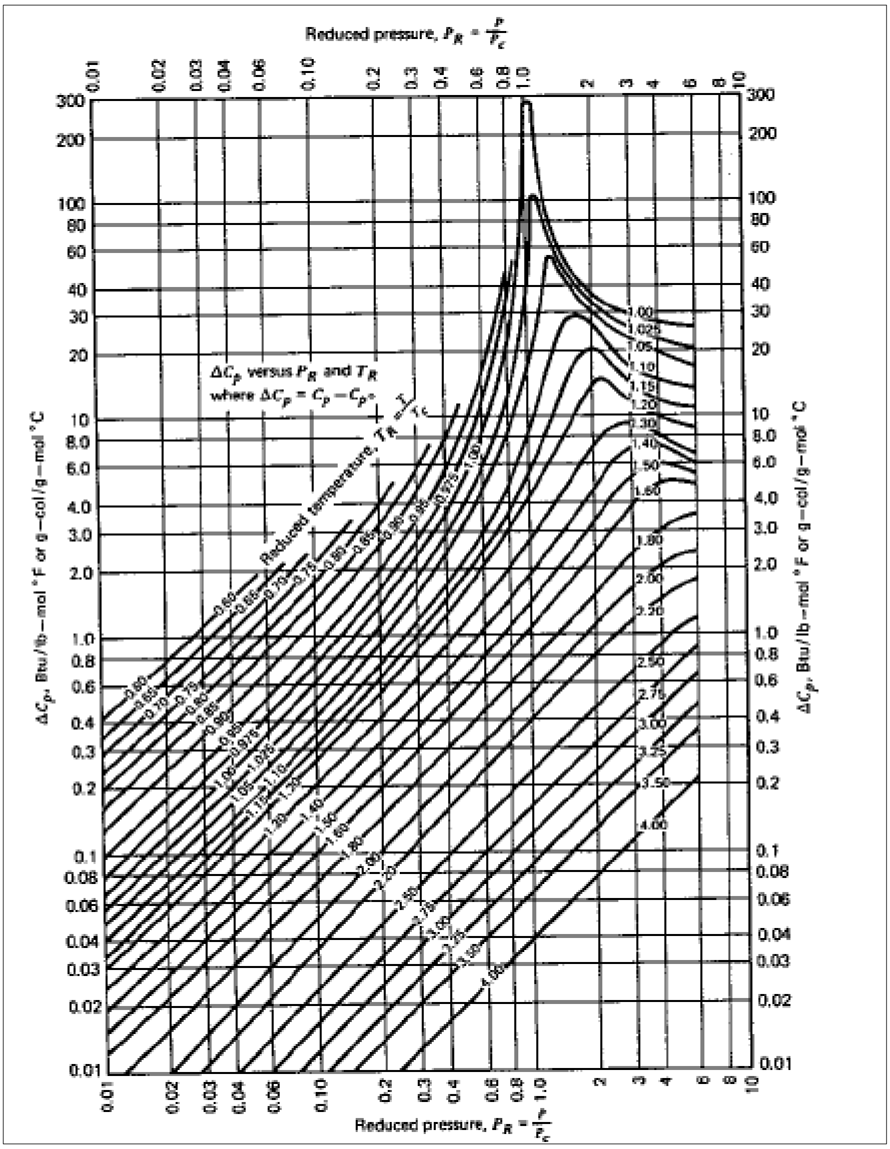

Figure A-5: Isothermal Pressure Correction of Heat Capacity of vapors

(Source: Chi U. Ikoku, 1998, Natural Gas Production Engineering)

APPENDIX B

Table B-1: Paint Solar Absorptance

Surface Color	Shade or Type	Reflective Condition (see Note 1)		
		New	Average	Aged
White		0.17	0.25	0.34
Aluminum	Specular	0.39	0.44	0.49
Aluminum	Diffuse	0.60	0.64	0.68
Beige/Cream		0.35	0.42	0.49
Black		0.97	0.97	0.97
Brown		0.58	0.62	0.67
Gray	Light	0.54	0.58	0.63
Gray	Medium	0.68	0.71	0.74
Green	Dark	0.89	0.90	0.91
Red	Primer	0.89	0.90	0.91
Rust	red iron oxide	0.38	0.44	0.50
Tan		0.43	0.49	0.55
Aluminum (see Note 2)	mill finish, unpainted	0.10	0.12	0.15

NOTE 1 Reflective condition definitions:

New: For paint, paint still retains the fresh shine of having been recently applied; for mill-finish aluminum, surface is shiny. This was previously labeled "Good."

Average: For paint, paint is in good condition, but the initial shine has faded; for mill-finish aluminum, surface is oxidized but still bright. The value given in each case is the average of the New and the Aged values for that case, and does not represent new data.

Aged: For paint, paint is noticeably faded and dull; for mill-finish aluminum, surface is dull. This was previously labeled "Poor."

NOTE 2 This refers to aluminum as the base metal, rather than aluminum-colored paint.

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf

Table B-2: Rim-Seal Loss Factors, K_{Ra}, K_{Rb}, and n, for Floating Roof Tanks

Tank Construction And Rim-Seal System	Average-Fitting Seals		
	K_{Ra} (lb-mole/ft-yr)	K_{Rb} [lb-mole/(mph)^n-ft-yr]	n (dimensionless)
Welded Tanks			
Mechanical-shoe seal			
Primary only[b]	5.8	0.3	2.1
Shoe-mounted secondary	1.6	0.3	1.6
Rim-mounted secondary	0.6	0.4	1.0
Liquid-mounted seal			
Primary only	1.6	0.3	1.5
Weather shield	0.7	0.3	1.2
Rim-mounted secondary	0.3	0.6	0.3
Vapor-mounted seal			
Primary only	6.7[c]	0.2	3.0
Weather shield	3.3	0.1	3.0
Rim-mounted secondary	2.2	0.003	4.3
Riveted Tanks			
Mechanical-shoe seal			
Primary only	10.8	0.4	2.0
Shoe-mounted secondary	9.2	0.2	1.9
Rim-mounted secondary	1.1	0.3	1.5

Tank Construction And Rim-Seal System	Tight-Fitting[d] Seals		
	K_{Ra} (lb-mole/ft-yr)	K_{Rb} [lb-mole/(mph)^n-ft-yr]	n (dimensionless)
Welded Tanks			
Mechanical-shoe seal			
Primary only	1.5	0.4	1.9
Shoe-mounted secondary	1.0	0.4	1.5
Rim-mounted secondary	0.4	0.4	1.0
Liquid-mounted seal			
Primary only	1.0	0.08	1.8
Weather shield	0.4	0.2	1.3
Rim-mounted secondary	0.2	0.4	0.4
Vapor-mounted seal			
Primary only	5.6	0.2	2.4
Weather shield	2.8	0.1	2.3
Rim-mounted secondary	2.2	0.02	2.6

Note: The rim-seal loss factors K_{Ra}, K_{Rb}, and n may only be used for wind speeds below 15 miles per hour.

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf)

Table B-3: Average Clingage Factors, C_S (bbl/103 ft2)

Product Stored	Shell Condition		
	Light Rust	Dense Rust	Gunite Lining
Gasoline	0.0015	0.0075	0.15
Single-component stocks	0.0015	0.0075	0.15
Crude oil	0.0060	0.030	0.60

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf)

Table B-4: Typical Number of Columns as A Function of Tank Diameter for Internal Floating Roof Tanks with Column- Supported Fixed Roofs

Tank Diameter Range D, (ft)	Typical Number Of Columns, N_C
$0 < D \leq 85$	1
$85 < D \leq 100$	6
$100 < D \leq 120$	7
$120 < D \leq 135$	8
$135 < D \leq 150$	9
$150 < D \leq 170$	16
$170 < D \leq 190$	19
$190 < D \leq 220$	22
$220 < D \leq 235$	31
$235 < D \leq 270$	37
$270 < D \leq 275$	43
$275 < D \leq 290$	49
$290 < D \leq 330$	61
$330 < D \leq 360$	71
$360 < D \leq 400$	81

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf)

Table B-5: Deck-Fitting Loss Factors, K_{Fa}, K_{Fb}, and m, and Typical Number of Deck Fittings, N_F

Fitting Type And Construction Details[q]	Loss Factors			Typical Number Of Fittings, N_F
	K_{Fa} (lb-mole/yr)	K_{Fb} (lb-mole/(mph)^m-yr)	m (dimensionless)	
Access hatch				1
Bolted cover, gasketed[b]	1.6	0	0	
Unbolted cover, ungasketed	36[c]	5.9	1.2	
Unbolted cover, gasketed	31	5.2	1.3	
Fixed roof support column well[d]				N_C (Table 7.1-11)
Round pipe, ungasketed sliding cover	31	p		
Round pipe, gasketed sliding cover	25	p		
Round pipe, flexible fabric sleeve seal	10	p		
Built-up column, ungasketed sliding cover[c]	51	p		
Built-up column, gasketed sliding cover	33	p		
Unslotted guidepole and well				f
Ungasketed sliding cover	31	150	1.4	
Ungasketed sliding cover w/pole sleeve	25	2.2	2.1	
Gasketed sliding cover	25	13	2.2	
Gasketed sliding cover w/pole wiper	14	3.7	0.78	
Gasketed sliding cover w/pole sleeve	8.6	12	0.81	
Slotted guidepole/sample well[e]				f
Ungasketed or gasketed sliding cover	43	270	1.4	
Ungasketed or gasketed sliding cover, with float[g]	31	36	2.0	
Gasketed sliding cover, with pole wiper	41	48	1.4	
Gasketed sliding cover, with pole sleeve	11	46	1.4	
Gasketed sliding cover, with pole sleeve and pole wiper	8.3	4.4	1.6	
Gasketed sliding cover, with float and pole wiper[g]	21	7.9	1.8	
Gasketed sliding cover, with float, pole sleeve, and pole wiper[h]	11	9.9	0.89	
Flexible enclosure[i]	21	7.9	1.8	
Gauge-float well (automatic gauge)				1
Unbolted cover, ungasketed[b]	14[c]	5.4	1.1	
Unbolted cover, gasketed	4.3	17	0.38	
Bolted cover, gasketed	2.8	0	0	
Gauge-hatch/sample port				1
Weighted mechanical actuation, gasketed[b]	0.47	0.02	0.97	
Weighted mechanical actuation, ungasketed	2.3	0	0	
Slit fabric seal, 10% open area[c]	12	p		
Vacuum breaker				N_{vb} (Table 7.1-13)[j]
Weighted mechanical actuation, ungasketed	7.8	0.01	4.0	
Weighted mechanical actuation, gasketed[b]	6.2[c]	1.2	0.94	

Table B-5: Deck-Fitting Loss Factors, K_{Fa}, K_{Fb}, and m, and Typical Number of Deck Fittings, N_F (Contd.)

Fitting Type And Construction Details[q]	Loss Factors			Typical Number Of Fittings, N_F
	K_{Fa} (lb-mole/yr)	K_{Fb} (lb-mole/(mph)^m-yr)	m (dimensionless)	
Deck drain (3-inch diameter)				
Open[b]	1.5	0.21	1.7	N_d (Table 7.1-13),
90% closed	1.8	0.14	1.1	
Stub drain (1-inch diameter)[k]	1.2	p		N_d (Table 7.1-15)
Deck leg, IFR-type (total sleeve length approx. 12 inches)[m]				N_l (Table 7.1-15)
Adjustable[c]	7.9	p		
Deck leg, EFR-type (pontoon area of pontoon roofs; total sleeve length approx. 30 inches)				N_l (Table 7.1-14)
Adjustable - ungasketed[b]	2.0	0.37	0.91	
Adjustable - gasketed	1.3	0.08	0.65	
Adjustable - sock	1.2	0.14	0.65	
Deck leg, EFR-type (double-deck roofs and center area of pontoon roofs, total sleeve length approx. 48 inches)				N_l (Table 7.1-14)
Adjustable - ungasketed[b]	0.82	0.53	0.14	
Adjustable - gasketed	0.53	0.11	0.13	
Adjustable - sock	0.49	0.16	0.14	
Deck leg or hanger (no opening through deck)				N_l may be set as 0
Fixed	0	0	0	(no openings)
Rim vent[n]				1
Weighted mechanical actuation, ungasketed	0.68	1.8	1.0	
Weighted mechanical actuation, gasketed[b]	0.71	0.10	1.0	
Ladder well				1[d]
Sliding cover, ungasketed[c]	98	p		
Sliding cover, gasketed	56	p		
Ladder-slotted guidepole combination well				1[d]
Sliding cover, ungasketed	98	p		
Ladder sleeve, ungasketed sliding cover	65	p		
Ladder sleeve, gasketed sliding cover	60	p		

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf)

Table B-6: Internal Floating Roof Tanks: Typical Number of Deck Legs, N1, And Stub Drains, Nd

Deck fitting type	Typical Number Of Fittings, N_F
Deck leg or hanger well[b]	$(5+\frac{D}{10}+\frac{D^2}{600})$
Stub drain (1-inch diameter)[b,c]	$(\frac{D^2}{125})$

(Source: https://www3.epa.gov/ttnchie1/ap42/ch07/final/ch07s01.pdf)