

ENHANCING LPG PRODUCTION CAPACITY

Submitted in partial fulfillment of the requirements for the award of bachelor of
Technology Degree in Chemical Engineering

by

JITENDRA YADAV (Reg. No.40690008)



**DEPARTMENT OF CHEMICAL ENGINEERING
SCHOOL OF BIO AND CHEMICAL ENGINEERING
SATHYABAMA**

**INSTITUTE OF SCIENCE AND TECHNOLOGY
(DEEMED TO BE UNIVERSITY)**

Accredited with Grade "A++" by NAAC

JEPPIAAR NAGAR, RAJIV GANDHI SALAI, CHENNAI - 600 119

OCTOBER 2023



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BONAFIDE CERTIFICATE

This is to certify that this Project Report is the bonafide work of **JITENDRA YADAV, (Reg.No.:40690008)** who carried out the entitled "**ENHANCING THE LPG PRODUCTION CAPACITY**" under my supervision from July 2023 to October 2023.

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(इंडियन ऑयल की ग्रुप कंपनी)
Chennai Petroleum Corporation Limited
(A group company of IndianOil)



TRG PW ENGG PG

July 18, 2023

The HOD,
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Dear Sir / Madam,

Sub: Project Work acceptance approval req

We are pleased to permit our employee Mr. JITENDRA YADAV, Pr No: 4714 to do Project Work in our Organization. During the period from 17.07.2023 to 30.09.2023 subject to the following Conditions:

1. The Employee will be governed by the rules and regulations of the Organization, including all safety precautions and procedures and the Employee will scrupulously adhere to the same with utmost diligence
2. He will not be eligible for any remuneration / stipend.
3. He will not be provided with boarding / lodging facility.
4. He should report to Mr. Monidip Pal Pr No: 804725 - Project Internal guide for the completion of the Project Work titled on "Enhancing LPG Production Capacity" in Ref - 1.
5. After the successful completion of the Project, the employee should submit the Project attendance report along with the Project report (Format as specified in CPCL Training Division) signed by Internal & External guide for obtaining the completion certificate.

Yours truly,
For Chennai Petroleum Corporation Ltd.

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DECLARATION

I, **JITENDRA YADAV**, (Reg.No.:40690008) hereby declare that the project Report entitled "**ENHANCING THE LPG PRODUCTION CAPACITY**" done by me under the guidance of **Dr. D. VENKATESAN**, M. Tech., Ph. D is Submitted in partial fulfillment of the requirements for the award of Bachelor of Technology Degree in Chemical Engineering.

DATED: 02/11/23
PLACE: Chennai


SIGNATURE OF THE CANDIDATE

External Examiner

चेन्नै पेट्रोलियम कॉर्पोरेशन लिमिटेड
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For Chennai Petroleum Corporation Ltd.,

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I convey my thanks to **Dr. S SATHISH, M.E., Ph. D.** Head of the department of chemical Engineering for providing us necessary support and details at the right time during progressive reviews.

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I thank **Mr. JAI PRAKASH B.Tech.** Engineer of CHENNAI PETROLEUM CORPORATION LIMITED, CHENNAI who helped us to complete the project successfully. We would also like to thank our friends and family who have contributed directly or indirectly for doing this project work.

I wish to express my thanks to all Teaching and Non-teaching staff members of the DEPARTMENT OF CHEMICAL ENGINEERING who were helpful in many ways for the completion of the work.

Finally we would like to thank the almighty for giving us this great opportunity and for leading us in the right path with right thought and deed of excellence.

ABSTRACT

CPCL intends to increase the yield of the Liquefied Petroleum Gas (LPG) from the stabilizer tower in Vapor Recovery Unit (plant 2) in Refinery 1 complex. The scope of the project deals with the design considerations, economics and safety aspects of the process.

To achieve an increase in the production of LPG, the impact of new absorption and scrubbing equipment's on the off gas was studied. In the design aspect, an alternate mechanism for LPG maximization was suggested and compared with the existing mechanism. For the same, new equipment designs were done. For the modified plant, the cost of the plant was estimated and HAZOP study of the plant was carried out.

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LIST OF ABBREVIATIONS

ADU	ATMOSPHERIC DISTILLATION UNIT
LPG	LIQUIDIFIED PETROLIUM GAS
MEA	MONOETHANOLAMINE
LMDT	LOG MEAN TEMPERATURE DIFFERENCE
NTU	NUMBER OF TRANSFER UNITS
HTU	HEIGHT OF TRANSFER UNITS

LIST OF SYMBOLS

Q_v	heat transfer from vapor
M_w	cooling water flow rate
A_c	Cross area of condenser
A_T	Area offer by tube
d_o	Tube outer diameter
d_i	Tube inner diameter
N_T	Number of tube
D_B	Bundle of diameter
D_S	Shell diameter
h_i	Tube side coeffiecient
h_o	shell side coeffient
U_o	Over all heat transfer
G_1	Inlet flow rate of gas
G_2	Outlet flow rate of gas
L_1	Solvent outlet flow rate
L_2	Solvent inlet flow rate

CHAPTER 1

1. INTRODUCTION

Crude oil is divided into multiple fractions, each necessitating additional processing or mixing to adhere to quality standards. This separation occurs within distillation column an atmospheric pressure and a two-stage vacuum column located in crude distillation unit. The resulting product streams include upward side distillate naphtha, high naphtha, kerosene, and diesel from the distillation column at atmospheric pressure, as well as lightest oil, weak(spindle) oil, and light oil from the first vacuum stage tower. The second vacuum tower stage yields heavy oil and vacuum residue.

The overhead naphtha is processed in a Vapor Recovery Unit (VRU), where it undergoes separate into gas, Liquefied Petroleum Gas (LPG), (LSRG), and light naphtha. The primary objective of this unit is:

- Producing LPG according to specification from distillation column feed.
- For obtaining raw hexane by feeding the treated light straight run gasoline to hexane unit.
- To feed untreated LSRG and Light Straight Run Naphtha (LSRN) to Catalytic Reforming Unit (CRU) in another refinery complex when required.
- To supply hot LSRN feed to Unifier and plat former section when required.

The basic process for the recovery of LPG from unsterilized naphtha involves the separation C3 and C4 components in the stabilizer column under temperature and pressure controlled operation. Hence in order to maximize the LPG production the following parameters of the stabilizer should be monitored on a regular basis:

- The Top and the Bottom temperature.

- The Column pressure. Low pressure causes more propane and butane to escape into the vapor state.
- The vapor draw.

In the existing scheme in vapor recovery and treatment unit, the LPG vapor is drawn off at the 45th and 47th tray in addition to the overhead draw-off. This necessitates separate condensation at each level. Coupled with high energy consumption and pumping costs, a loss of nearly 20% of LPG in the off-gas is observed in this method.

An attempt has been made to condense C_3 to iC_5 components in the overhead vapor in a single stage, thereby eliminating the need for a side-draw. For the stabilizer column, overhead partial condenser, re-boiler and the reflux pump, the following was considered based on the revamp calculations:

- The feasibility of retaining the existing equipments as they are without any modifications
- The feasibility of retaining the equipments with slight modifications.

If it is not possible to retain, then that particular equipment is replaced by new equipment, which satisfies the required specifications.

Considering the fact that there is an increased demand for LPG in the present scenario and keeping in view the limitations in the existing system, an attempt has been made to select and explore factors that could provide the means to the refinery for improvement in the profitability.

1.1 Product Characteristics

Major product characteristics to be fulfilled are outlined in this section. Target recovery of the various components in the distillate product stream (LPG) from the stabilizer column shall be as follows:

C ₃ recovery	:	100% (mole)
iC ₄ recovery	:	100% (mole)
nC ₄ recovery	:	98% (mole)

The impurity levels in the distillate product shall be limited to the following:

iC ₅ impurity	:	2% (mole) of that in feed
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The impurity levels in distillate product to the LPG stripper shall not be relaxed.

CHAPTER 2

LITERATURE SURVEY

Detailed plant design information for the existing facility was extracted from the plant data sheet. The references utilized in this project were organized based on the type of information they provided. References related to the Stabilizer/Distillation column encompass a wide range of topics related to tray and column design. These references cover various of column and trays design. Additionally, classified related to heat exchanger and re-boiler specifically address their formation and operational aspects.

Stabilization is the process of eliminating lighter components from a hydrocarbon stream that is heavier in nature. In vapour recovery unit Propane, Butane and other lighter ends are separated in overhead stream and side cut of Propane and Butane (LPG).it is essential to prevent the heavier components from mixing with the lighter cut, just as it is important to avoid contamination of the heavier cut by the lighter components. Therefore, it is advisable to limit the withdrawal of side streams in order to maintain the desired separation and purity. In line with this recommendation, we suggest withdrawing a single overhead stream for condensation, and the column is designed to facilitate this operation.

Designing a column encompasses the process of establishing the operating conditions for the column, as well as determining vapor and liquid flow rates, tray count, tray diameter and area, downcomer area, and the preferred operating mode.

In the initial design phase for determining the tray count using the McCabe - Thiele method in a system with a multi-component mixture, it's often practical to assume a pseudo-binary system. It is observed that the accuracy of the product estimation may lesser than actual value and more complication in calculation especially in multi component system.

Fenske - Underwood-Gilliland correlation is used to calculate min. number of theoretical plates.

$$N_{mt} = \frac{[\ln(x_{LK}/x_{HK})_D (x_{HK}/x_{LK})_B]}{\ln \alpha}$$

Where

x_{LK} = Mole fraction of lighter end

x_{HK} = Mole fraction of heavier end

Where

D is Distillate

and B is bottom product.

α is the relative volatility.

The number of theoretical stages under actual reflux conditions can be determined using the Gilliland correlation, which has demonstrated its effectiveness for both binary and multi-component mixtures across a broad range of reflux ratios and tray counts. This method is based on Figure 6, where the abscissa provides the Underwood constant, enabling the determination of the number of theoretical stages under actual reflux conditions. The number of theoretical stages is depend on column design and its efficiency. The column efficiency is calculated by Drickamer and Bradford correlation method by using feed conditions specially molal average viscosity. This approach has demonstrated its reliability for hydrocarbons and related compounds. While occasional deviations of 15-20% were observed in certain cases, comparisons suggest that these deviations are generally conservative or on the safe side.

Once you've estimated the number of trays required, you can determine the column's height by multiplying it by the tray spacing. Tray spacing is typically determined to minimize entrainment. In the case of an atmospheric column, a common practice is to use a spacing of 500 mm. However, it's crucial to consider that there is a maximum height restriction for a column. Once this limit is reached, then the column design and construction cost will high as so many parameters to be considered like column foundation strength and wind velocity etc.

The maximum height of a column is contingent upon various factors, including local weather conditions, susceptibility to events such as hurricanes, typhoons, or earthquakes, and the foundation's ground conditions. Typically, the maximum height is capped at 80 meters, but it could be considerably lower in areas prone to adverse conditions. As the column's height increases, there is a stronger motivation to reduce tray spacing in order to minimize the overall column height. However, it's important to note that there is a practical limit to how small tray spacing can be, with the usual minimum tray spacing falling in the range of approximately 0.35 meters.

The sizing of the column's area and diameter depends on the specific conditions of the column. In cases where operating pressures are moderate to low, typically below 150 psig or 10.34 bars, the column diameter is primarily determined by vapor flow considerations. The above calculation based on column top conditions. The liquid and vapour velocity factor FLV is consider in to account to determine the column pressure and tray fixation.. This factor, is correlated with the flooding velocity, to calculate the actual free area. When free area is added to the downcomer area, total area of the column, can be calculated. After calculating column area, column diameter can be calculated.

The thermosyphon re-boiler is choosen effective separation. The separation or disengagement of phases occurs within the column itself. The re-boiler can be connected in either a once-through or a recirculating liquid arrangement, depending on the specific design and operational requirements of the system.

In horizontal re-boilers of this type, the film coefficient for boiling is not influenced by the velocity of the fluid at low velocities. The rate of heat transfer during boiling is primarily driven by the temperature gradient, with higher temperature differences resulting in more efficient heat transfer, rather than the fluid velocity.

The liquid from stabilizer column bottom is entered into a thermosyphon re-boiler and flows the entire length of the shell. However, because the mass velocity in each instance is calculated based on one-half of the total flow, it is

appropriate to treat the pressure drop in a manner consistent with a shell that lacks baffles and features axial flow.

Condensing the overhead mixed vapors in a partial horizontal shell and tube condenser can lead to the separation of condensate from the vapor phase.

The design of the partial condenser involves various factors, including the heat transfer coefficient and over all heat transfer coefficient and these factors are important in the efficiency and effectiveness of the condensation process within the condenser.

CHAPTER – 3

EXISTING PROCESS EXPLANATION

3.1 FEED PREHEATING

The feedstock, from ADU overhead condensed liquid @ 40° C that is unstablized naphtha is fed in to the stabilizer column after preheating in a heat exchanger up to 80° C. In this heat exchanger, it comes into contact with the stabilizer bottoms, which flow in a countercurrent manner to the feed. The outlet temperature of the exchanger is controlled and monitored by a control valve.

3.1.1 Stabilizer Column Overhead/Sour Gas

Stablizer overhead stream contains more H₂S called as Sour gas undergoes partial condensation to form liquid. The mixture comprising condensate and uncondensed gas is directed into the stabilizer overhead accumulator, where separation processes take place in accumulator sour water is separated from bottom by means of density difference and condensed hydrocarbon liquid send back to tower as reflux by using a pump. Reflux flow is controlled by a flow controller. This controller is acting based on tower top temperature. A pressure controller valve is employed to maintain the tower pressure by adjusting the off gas flow. Off Gas is nothing but uncondensed vapour in accumulator.

The LPG from the column is flows into the LPG stripper. The purpose of the stripper is to remove lighter ends from the LPG. This lighter ends mainly Methane and Ethane components that could be carried within the LPG stream. Both the vessel and the trays within the stripper are constructed from carbon steel. Following the stripper, the LPG from the bottom section is cooled in two parallel water coolers. It is then channeled to the LPG MEA (Methyl Ethyl Alcohol) scrubber under flow control. A flow controller at the stripper outlet governs the total PG flow, diverting a portion of the LPG stream to the LPG vaporizer as needed to maintain control.

3.1.2 Stabilizer Bottoms

The tower bottoms are re-boiled in the stabilizer re-boiler, which is of the horizontal thermosyphon type. The heat required for re-boiler is supplied by a hot slipstream obtained from the bottom pump-around pump of the atmospheric tower. This slipstream primarily contains diesel oil. To maintain the desired temperature, a temperature controller is installed on the liquid at the fourth tray. The controller operates a three-way control valve to regulate the heat input as needed. The tower bottoms, after being re-boiled, exit the system through the feed pre-heater. Following this, they exchange heat with the stabilizer feed before proceeding to the naphtha splitter column. This configuration helps efficiently utilize the heat in the system and optimize. The feed from the bottom of the stabilizer is introduced into the naphtha splitter tower between the fourth and fifth trays. The naphtha splitter tower is used to separate the stabilizer bottom product into two main fractions: C5+95 °C Light Straight Run Gasoline (LSRG) and C6+138 °C light naphtha. This separation process helps refine and separate the different components within the product for further processing and utilization.

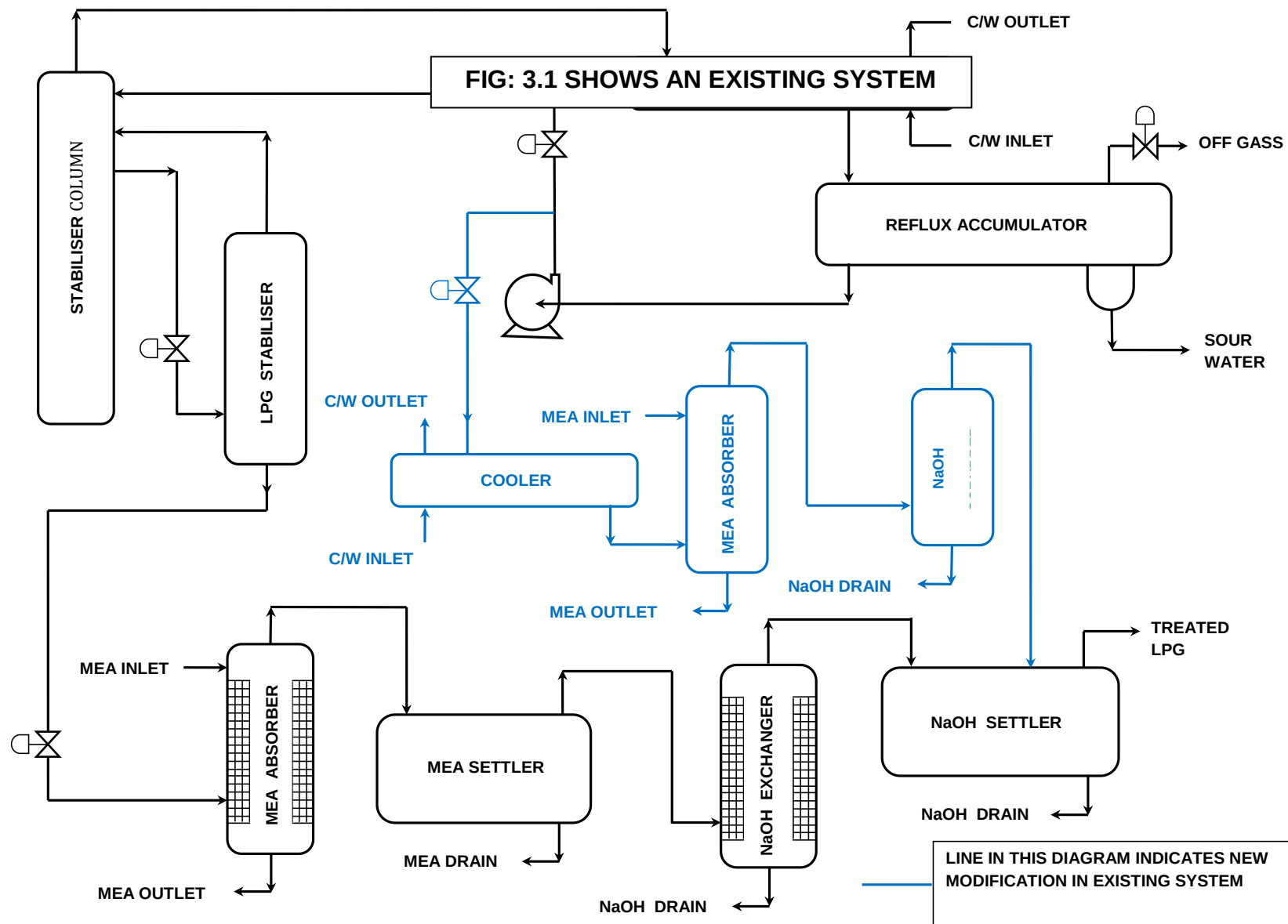


Figure: 3.2 Shows An Revamped System

3.2 REVAMP PROCESS DESCRIPTION

3.2.1 Feed Preheating

The atmospheric distillation column overhead condensed liquid is @ 40 °C is fed to stabilizer column after heating it to a heat exchanger named 2E1 up to 80°C. Stabilizer bottom material which is called stabilized naphtha is re-boiled in a re-boiler named 2E3. Stabiliser based temperature 150°C is controlled by a three way control valve and heating fluid is bottom pump around stream from ADU.

3.2.2 Stabilizer Column Overhead/Sour Gas

During the process, sour gas exits from the upper part of the tower. A portion of the vapor undergoes partial condensation, resulting in the formation of liquid. This mixture of condensate and uncondensed gas then enters the stabilizer overhead accumulator, where a separation process occurs. The primary purpose of the stabilizer overhead accumulator is to function as a vessel specifically designed for collecting the water that has been separated from the hydrocarbon stock in this phase of the process.

The stabilizer reflux pump takes suction from reflux drum and pumps it to stabilizer column to control the tower top temperature as per process requirement. The reflux rate is meticulously managed via a three- way control valve positioned on the bypass of the stabilizer overhead pa condenser. To uphold the system's pressure at the desired level, a pressure controller valve is used. This valve governs the flow of sour gas from the overhead accumulator to the Amine gas Treating Unit, ensuring that the system pressure remains within the specified range.

The flow within the system is precisely monitored and regulated by a specialized controller. Any water separated in the condenser collects in the lower portion of the accumulator. The level of this accumulated water is meticulously controlled by a dedicated level controller. To monitor the process conditions, two temperature controllers are utilized. One of these controllers measures the temperature of the vapor leaving the tower, while the other, positioned on the discharge line from the accumulator, indicates the temperature of the reflux. These

controllers play a crucial role in ensuring the system operates effectively and efficiently.

The condensed LPG from the condenser is routed under a flow control to cooler where the condensed LPG is cooled to the required 40 °C, from the cooler the cooled condensate is routed to the LPG MEA absorber where the Hydrogen Sulphide impurity present in the condensate is removed. From the MEA absorber for further purification the condensate is further fed to the caustic scrubber where the rest of the impurities are removed to the possible extent. From here the LPG is sent to a LPG settling tank where heavier substances which are carried along with the stream is settled. LPG is an odorless gas, in an instance of leakage it is undetectable, hence for safety purposes 20 ppm of mercaptans is added to the final product. This gives LPG a particular smell. This final product is sent to the storage tank.

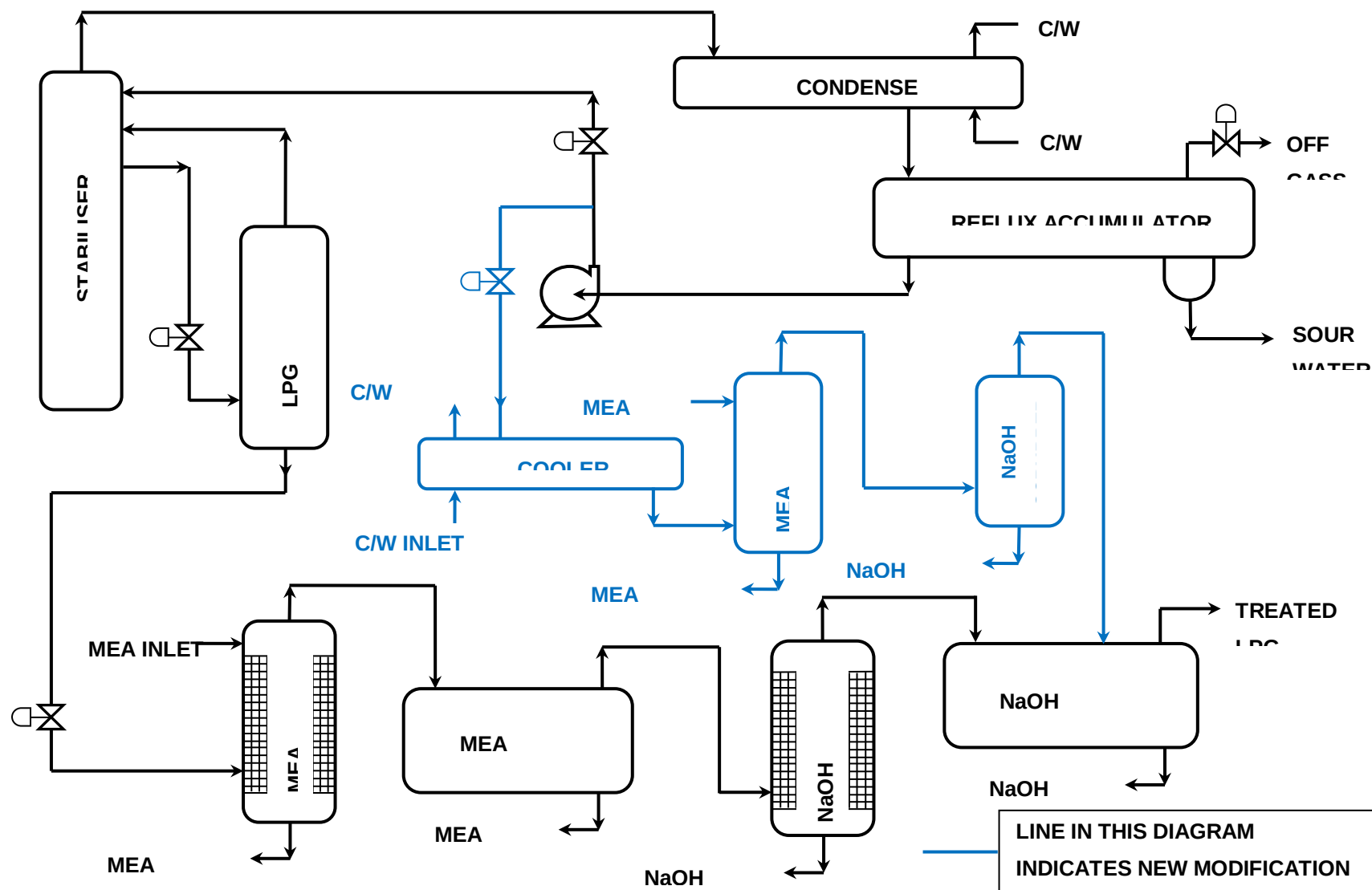


Figure 3.2 : Shows An Revamped System

CHAPTER 4

4.1 MATERIAL AND ENERGY BALANCE

4.1.1 Compositions Of Various Streams

Basis: 30.969 kmol / h of feed

The feed and product composition are given below in table.

Table 4.1: Feed and Product Composition

Feed Component	Kmol / hr	Average molecular Weight	Weight of Components Kg/hr	Product Out kg/hr
H ₂ O	0.231	18.015	4.161	0.041
H ₂ S	4.851	34.082	165.331	Nil
H ₂	0.319	2.016	0.643	0.643
C ₁	2.002	16.040	32.112	32.112
C ₂	1.518	30.070	45.646	45.646
C ₃	5.093	44.093	224.566	224.566
iC ₄	2.948	58.120	171.337	171.337
nC ₄	13.388	58.120	778.110	778.110
iC ₅	0.619	72.150	44.660	44.660
Total	30.969	36.967	1466.566	1297.115

4.1.2 Operating Parameters of the Process

It is consider that the temperature of reflux stream is same as condenser outlet at 318⁰ K and also supposed to be at its bubble point.

Operating pressure is 12.40 kg/cm². The pressure of overhead is fixed as

$$12.40 + 0.5 = 12.45 \text{ kg/cm}^2$$

Temperature and pressure conditions of various streams are given below-

Table 4.2: Temperature and Pressure Conditions of Various Streams

Feed	Operating Temperature(⁰ k)	Operating Pressure(Kg/cm ²)
Overhead condensed liquid	318	12.45

4.1.3 Over All Material Balance

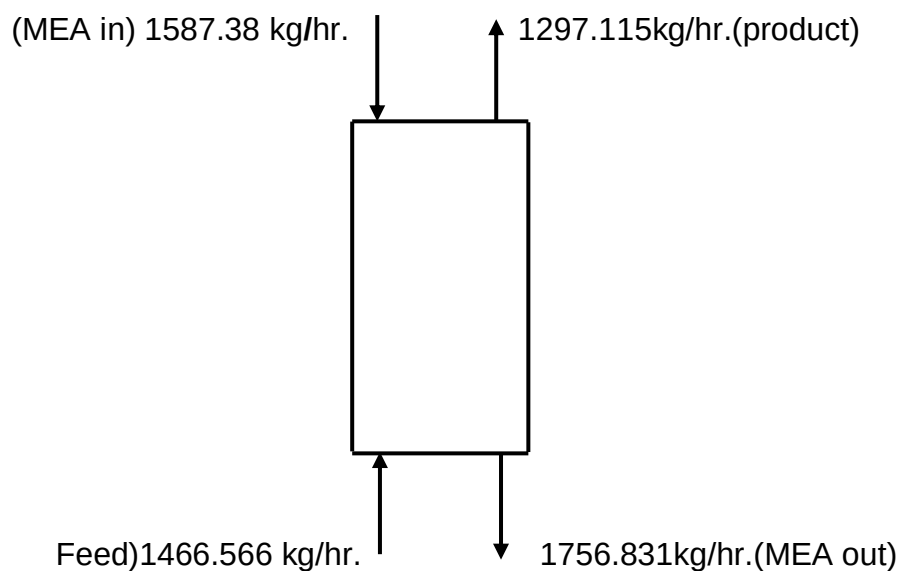


Figure 4.1 : Over All Material Balance Diagram

General Mass Balance Equation,

Mass In + Generation = Mass Out + Consumption

(Since nil generation & consumption here)

Mass In = Mass Out

1466.566 kg/hr. + 1587.38 kg/hr. = 1756+.831 kg/hr. + 1297.115 Kg/hr.

3053.946 kg/hr. = 3053.946 kg/hr.

4.1.4 Material Balance for Mea Absorber

Table 4.3: Composition At Inlet And Outlet of Mea Absorber

COMPOSITON	INLET OF MEA ABSORBER	OUTLET OF MEA ABSORBER
	Kg/hr	Kg/hr
H ₂ O	4.161	4.158
H ₂ S	165.331	6.59 @96% removed
H ₂	0.643	0.643
C ₁	32.112	32.112
C ₂	45.646	45.646
C ₃	224.566	224.566
iC ₄	171.337	171.337
nC ₄	778.110	778.110
iC ₅	44.660	44.660
Total	1466.566	1307.822

Table 4.4: Composition At Inlet And Outlet of Absorber Solvent

COMPOSITON	INLET SOLVENT (kg/hr.)	OUTLETSOLVENT (kg/hr.)
MEA	1587.38	1587.38
H ₂ S	0	158.741
TOTAL	1587.38	1746.121

4.1.5 Material Balance Across Mea Absorber

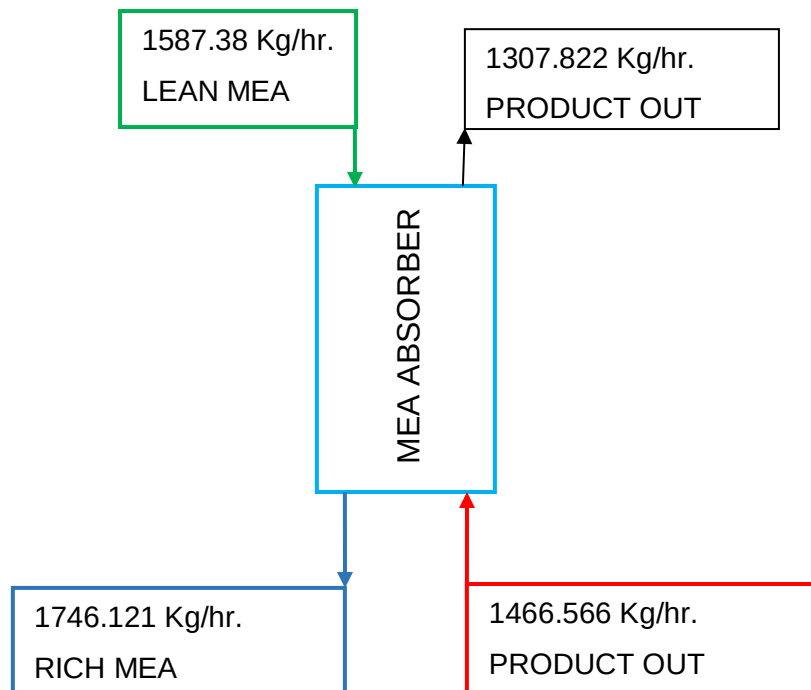


Figure 4.2 : Diagram of Material Balance Across Mea Absorber

General Mass Balance Equation,

Mass In + Generation = Mass Out + Consumption

(Since nil generation & consumption here)

Mass In = Mass Out + Accumulation (H₂S+Water absorbed in NaOH)

1587.38 Kg/hr + 1466.566 Kg/hr = 1307.822 Kg/hr + 1746.121 Kg/hr + 0.003

3053.946 Kg/hr = 3053.946 Kg/hr

4.1.6 Material Balance of NaOH Scrubber

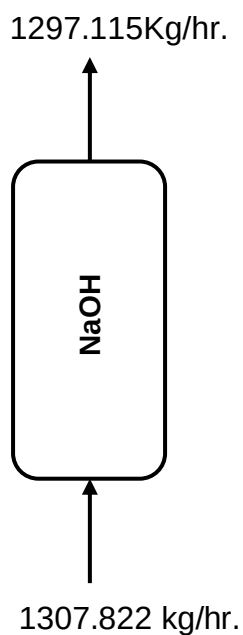


Figure 4.3 : Diagram Of Material Balance Across NaOH Scrubber

General Mass Balance Equation,

Mass In + Generation = Mass Out + Consumption

(Since nil generation & consumption here)

Mass In = Mass Out + Accumulation (H₂S+Water absorbed in NaOH)

$$1307.822 = 1297.115 + 10.706 = 1307.821$$

4.1.7 Individual Material Balance

Table 4.5: Composition At Inlet and Outlet of NaOH Scrubber

COMPOSITON	INLET OF NaOH SCRUBBER	OUTLET OF NaOH SCRUBBER
	Kg/hr.	Kg/hr.
H ₂ O	4.158	0.0415
H ₂ S	6.59	Nill
H ₂	0.643	0.643
C ₁	32.112	32.112
C ₂	45.646	45.646
C ₃	224.566	224.566
iC ₄	171.337	171.337
nC ₄	778.110	778.110
iC ₅	44.660	44.660
Total	1307.822	1297.115

4.1.8 Energy Balance for Cooler

Table 4.6: Energy Balance for Cooler

Component	Molar Flow Rate	Molecular Weight	Mass Flow Rate	Temp.	Specific Heat	Enthalpy	Heat
	Kmol/hr.	Kg/kmol	Kg/hr.	K	J/kg k	J/kg	MJ/hr.
Heat In							
Feed In	30.969	-----	1466.566	318	1482	471276	691.157
Water	36.269	18.015	653.403	301	4184	1259384	822.885
						TOTAL	1514.042
Heat Out							
Feed	30.969	-----	1466.566	313	1482	463866	680.290
Water	36.269	18.015	653.403	305	4184	1276120	833.820
						TOTAL	1514.11

From the above table cooler duty = (691.157-680.290) MJ/hr
= 10867KJ/hr.

4.1.9 Energy Balance Around the Cooler

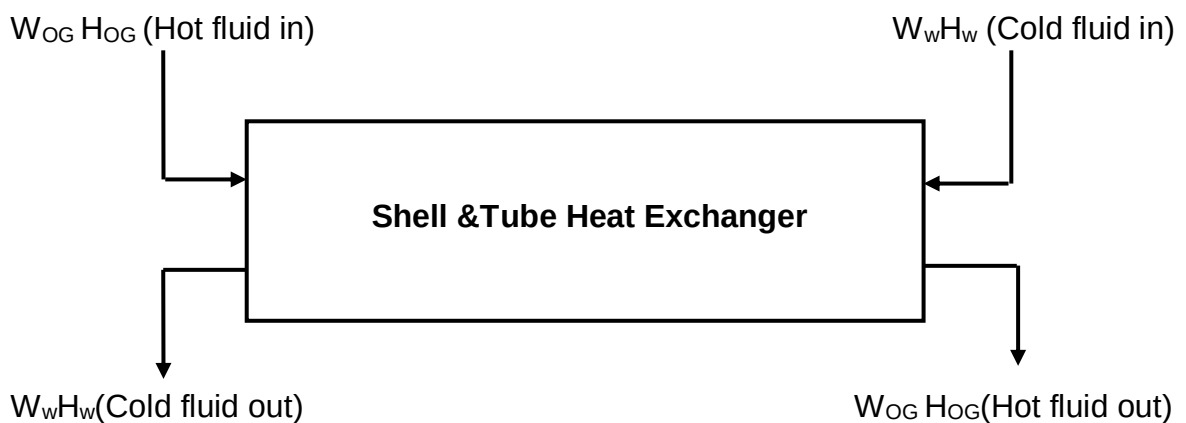


Figure 4.4 : Diagram of Energy Balance around the Cooler

$$\begin{aligned}
 \text{INPUT} &= W_{OG} H_{OG} (H) + W_w H_w (C) \\
 &= (1466.566 \times 471276) \text{J/hr.} + (653.403 \times 1259384) \text{J/hr.} \\
 &= 691.116 \text{ (MJ/hr)} + (822.885 \text{ MJ/hr.}) \\
 &= 1514.042 \text{ MJ/hr.}
 \end{aligned}$$

$$\begin{aligned}
 \text{OUTPUT} &= (1466.566 \times 463866) \text{J/hr.} + (653.403 \times 1276120) \text{J/hr.} - Q_c \\
 &= (680.290 \text{ MJ/hr)} + (834.219 \text{ MJ/hr.}) + (0.068 \text{ MJ/hr.}) \\
 &= 1514.042 \text{ MJ/hr.}
 \end{aligned}$$

Therefore the EB is Satisfied.

CHAPTER 5

DESIGN CALCULATIONS

5.1 HEAT EXCHANGER DESIGN

Shell-side fluid: LPG

Tube-side fluid: Water

Given data:

Flow of the LPG (m_v) = 1466.566 kg/hr

Temperature of feed in ($T_{h \text{ in}}$) = 45°C

Outlet temperature of the condensate ($T_{h \text{ out}}$) = 40°C

Cooling water in inlet ($T_{c \text{ in}}$) = 28°C

Cooling Water Temp. of outlet ($T_{c \text{ out}}$) = 32°C

Tube outer dia. (d_o) = 1 inch = 0.020 m

Specific heat, LPG (C_p) = 1.482 kJ/kgK

Length of the tube (L) = 2.5 m

Number of passes (N_p) = 4

Inner Diameter of the tube = 0.016m

$$\begin{aligned}\text{Heat transferred from vapor } (Q_v) &= m_v \times (C_p) \times (T_{h \text{ in}} - T_{h \text{ out}}) \\ &= (1466.566 / 3600) \times (45 - 40) \times (1.482) \\ &= 3.039 \text{ kW}\end{aligned}$$

$$\begin{aligned}\text{Cooling water flow in kg } (M_w) &= Q_v / ((T_{c \text{ out}} - T_{c \text{ in}}) \times C_p) \\ &= 3.039 / ((32 - 28) \times 4.186) \\ &= 0.182 \text{ kg/s} \\ &= 653.403 \text{ kg/hr}\end{aligned}$$

Supposed overall coefficient (Appendix 1, Table 14)

$$U_{\text{assumed}} = 300 \text{ W/m}^2 \text{ } ^\circ\text{C}$$

Mean Temperature difference:

$$\begin{aligned}R &= (T_{h \text{ in}} - T_{h \text{ out}}) / (T_{c \text{ out}} - T_{c \text{ in}}) \\ &= (45 - 40) / (32 - 28) \\ &= 1.25\end{aligned}$$

$$\begin{aligned}
 S &= (T_{c\ ou} - T_{c\ in}) / (T_{h\ in} - T_{c\ in}) \\
 &= (32 - 28) / (45 - 28) \\
 &= 0.235
 \end{aligned}$$

DESIGN OF CONDENSER

- 1) Load on condenser Q = 10867KJ/hr.
- 2) Q = $UA\Delta T_{LMTD}$
Take U = 300 w/m²°C
- 3) ΔT_{LMTD} = $\frac{(Thi - Tco) - (Tho - Tci)}{\ln(Thi - Tco)/(Tho - Tci)}$
= $(318-305)-(313-301)/\ln(96/70)$
= 12.50°C
 ΔT_{LMTD} = 12.50°C
- 4) Area of the Condenser (A_C) = 10867 / (300 × 12.50)
 A_C = 2.8978m²
- 5) Area offered by tube A_T = $\pi d_o L$
Specification of tubes d_o = 20mm
 d_i = 16mm
L = 2.5m
 A_T = $\pi d_o L$
= 3.14×0.02×2.5
= 0.157m²
- 6) Number of tubes N_T = A_C/A_T
= 2.897/0.157
= 18.45
 N_T = 20 tubes consider

7) Bundle diameter $D_B = d_o (N_T/K_1)^{(1/n_1)}$
 From the data table For Square pitch
 $K_1 = 0.156$
 $N_1 = 2.291$
 $D_B = 0.02(20/0.156)^{(1/2.296)}$
 $D_B = 0.1659\text{m}$

8) The tube pitch can be calculated by considering
 $P_T = 1.25 d_o$
 $= 1.25 \times 0.02$
 $P_T = 0.025\text{m}$

9) Shell diameter D_s
 $D_s - D_B = C \text{ (clearance)}$
 From the chart C value found as 10 mm.
 $0.01 = D_s - 0.1659$
 $D_s = 0.01 + 0.1659$
 $D_s = 0.1669\text{m}$

10) Tube side coefficient:
 Mean Water Temperature $= (32+28)/2$
 $= 30^\circ\text{C}$
 Tube cross sectional area $= \pi/4 d_i^2$
 $= 3.14 \times 16^2/4$
 $= 0.000200\text{m}^2$
 Tubes per pass $= 20/2$
 $= 10$
 Total flow area $= 10 \times 0.000200$
 $= 0.00200\text{m}^2$
 Water mass velocity $= 653.403/(3600 \times 0.0020)$
 $= 90.75 \text{ Kg/sm}^2$
 Density of water $= 995 \text{ Kg/m}^3$
 Water linear velocity $= 90.75/995$
 $= 0.0912 \text{ m/s}$

Thermal conductivity of water	=	614mw/m ⁰ C
Viscosity of water	=	0.7191 mNs/m ²
Reynolds No.	=	$\rho v d_i / \mu$
	=	$(995 \times .0912 \times .016) / .0007191$
	=	2019.05
Consider Reynolds No.	=	2019
Prandtl No.	=	$C_p k / \mu$
	=	$4184 \times 0.0007191 / 0.614$
	=	4.9
From data table(j_h	=	1.5×10^{-1}
Tube side coefficient h_i	=	$0.15 \times 0.614 \times 2019 \times 4.9^{.333} / .016$
	=	19729.28 w/m ² °C
11) Shell side coefficient:		
Baffle spacing	=	$D_s / 5$
	=	$.01659 / 5$
	=	.03318 m
Cross flow area A_s	=	$(d_o - d_i / d_o) D_s B_s$
	=	$[(20-16)/20] \times 0.1659 \times 0.03318$
	=	.0011 m ²
Mass velocity G_s	=	$1466.566 / (3600 \times .0011)$
	=	370.344 Kg/sm ²
Equivalent diameter D_e	=	$1.1(p_t - 0.917 d_o^2) / d_o$
	=	$(1.1/0.020)(0.025^2 - .917 \times 0.020^2)$
	=	0.0142 m
Mean sell side temperature	=	$(45+40)/2$
	=	42.5 ⁰ C
Mixture Density LPG	=	0.55 Kg/m ³
Viscosity of mixture (LPG)	=	0.08 mNs/m ²
Heat Capacity	=	1482KJ/Kg ⁰ C
Thermal conductivity	=	0.614 w/mK
Reynolds No. (Re)	=	$\rho v d_i / \mu$
	=	$(.55 \times 370.344 \times 0.0142) / .00008$
	=	36154.8

$$\begin{aligned} \text{Re} &= 36155 \\ \text{Prandtl No. Pr} &= (.00008 \times 1482 \times)/.017 \\ \text{Pr} &= 6.97 \end{aligned}$$

Choose 25% baffle cut from chart

$$J_h = 4.5 \times 10^{-3}$$

$$\begin{aligned} \text{Shell side coefficient } h_s &= J_h \times K \times N_{Re} \times (N_{Pr}^{.333})/d_i \\ &= (0.017 \times .0045 \times 36155 \times 6.97^{.333})/.0142 \\ &= 371.82 \text{ w/m}^2 \text{ } ^\circ\text{C} \end{aligned}$$

12) Over all Heat Transfer coefficient U_o ;

Thermal conductivity of metal $K_w = 43.6 \text{ w/m } ^\circ\text{C}$ for carbon steel

Take the fouling coefficients from data table

$$\text{for Flue gases (hdo)} = 6000 \text{ w/m}^2 \text{ } ^\circ\text{C}$$

Over all co – eff. $1/U_o$

$$\begin{aligned} &= (1/h_s) + (1/h_{do}) + (d_o/2 \times K) \ln[(d_o/d_i)] \\ &+ [(1/h_d)(d_o/d_i)] + [(1/h_i)(d_o/d_i)] \end{aligned}$$

Over all coefficient $1/U_o$

$$\begin{aligned} &= 1/371.82 + 1/6000 + (.020/(2 \times 43.6)) \ln[(20/16)] \\ &+ [(20/(16 \times 6000))] [(20/(16 \times 19729.28))] \\ &= 3.169 \times 10^{-3} \text{ m}^2 \text{ } ^\circ\text{C} / \text{w} \end{aligned}$$

$$1/U = 3.169 \times 10^{-3} \text{ m}^2 \text{ } ^\circ\text{C} / \text{w}$$

Over all heat transfer coefficient $U_o = 315.55 \text{ w/m}^2 \text{ } ^\circ\text{C}$

This U_o value is well above the assumed value of $300 \text{ w/m}^2 \text{ } ^\circ\text{C}$. So design has adequate area for the duty required.

5.2 CONDENSER DESIGN SUMMARY

Table 5.1: Condenser Design Summary

Parameter	Value Units
Flow rate of required water	653.403 Kg/hr
Number of tubes required	20
Heat transfer coefficient	300 W/m ² K
Heat transfer area calculated	2.9176 m ²
Pitch, pitch diameter	0.025 m
Shell diameter	0.1669m
Tube bundle diameter	0.1569 m
Baffle spacing	0.03318 m
Number of tube passes	2

5.3 DESIGN OF PACKED BED ABSORPTION COLOUMN GIVEN

Feed (LPG) $G_1 = 30.738 \text{ kmol/hr.}$ from material balance data

Feed out Absorber Efficiency= 96%

$G_2 = 25.8875 \text{ kmol/hr.}$ (calculated)

Feed (H_2S) = 4.851 kmol/hr.

Inlet MEA = 25.68 kmol/hr.

Outlet MEA = 34.56 kmol/hr.

Remaining $\text{H}_2\text{S} = 0.1941 \text{ kmol/hr.}$

Absorbed $\text{H}_2\text{S} = 4.65469 \text{ kmol/hr.}$

Density of LPG $\{\rho_A\} = 550 \text{ kg/m}^3$

1 atm. = 1.033 kgf/cm^2

$11.50 \text{ kgf/cm}^2 = 11.1307 \text{ atm.}$

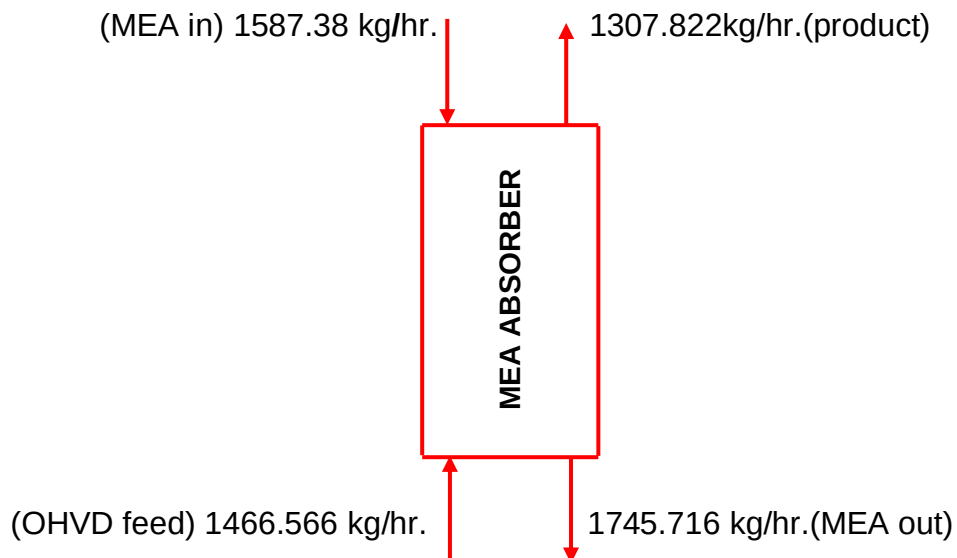


Figure 5.1: Calculation Diagram for Composition of MEA Absorber

G_s = flow of a LPG with salute free,

G_1 = Flow of Feed. (In overhead)

Y_1 = mole ratio of solute (H_2S) in Feed (overhead)

$$Y_1 = \frac{y_1 \text{ kmol of } H_2S}{(1-y_1) \text{ kmol of feed}}$$

$$G_1 = 30.738 \text{ kmol/hr.}$$

$$= 1466.566 \text{ kg/hr.}$$

G_2 = flow of in outlet feed

Y_2 = mole ratio of H_2S in outlet feed

$$Y_2 = \frac{y_2 \text{ kmol of } H_2S}{(1-y_2) \text{ kmol of outlet feed}}$$

X_2 = mole ratio of H_2S in MEA = 0 (salute free)

$$X_1 = \frac{x_1 \text{ kmol of } H_2S}{(1-x_1) \text{ kmol absorbed in MEA}}$$

$L_2 = L_s$ = Inlet flow of MEA with Salute free, kmol / hr.

L_1 = outlet flow of MEA with salute absorbed, kmol / hr.

X_1 = mole fraction of H₂S in liquid solvent leaving the absorber.

$$\begin{aligned}\text{Total feed} &= 30.738 \text{ kmol / hr. (Exclusive water content)} \\ &= 1462.405 \text{ Kg/hr}\end{aligned}$$

$$\begin{aligned}\text{H}_2\text{S Presence in feed} &= 4.851 \text{ kmol / hr.} \\ &= 165.331 \text{ Kg/hr}\end{aligned}$$

96% H₂S is absorbed in MEA and remaining 4% H₂S is carried forward to NaOH scrubber.

$$\begin{aligned}4\% \text{ H}_2\text{S} &= 0.194 \text{ kmol/hr.} \\ &= 6.59 \text{ kg/hr.}\end{aligned}$$

$$\begin{aligned}y_1 &= 4.851/30.738 \\ &= 0.1578\end{aligned}$$

$$\begin{aligned}Gs &= G1(1 - y_1) \\ &= 30.738 (1 - 0.1578) \\ &= 25.8875 \text{ kmo l/ hr.}\end{aligned}$$

$$\begin{aligned}Y_1 &= 0.1578 / (1 - 0.1578) \\ &= 0.1870\end{aligned}$$

$$\begin{aligned}Y_2 &= 0.1870 \times 4/100 \\ &= 0.00748\end{aligned}$$

$$\begin{aligned}Y2 &= Y2/1 + Y2 \\ &= 0.00748/(1 + 0.00748) \\ &= 0.00742\end{aligned}$$

$$\begin{aligned}\text{Partial pressure of H}_2\text{S} &= 0.5 \text{ Kpa} \\ &= 0.5 \times 7.5\end{aligned}$$

$$= 3.75 \text{ mmHg}$$

$$\begin{aligned}\text{Mole fraction in vapor} &= 3.75/760 \\ &= 0.00493\end{aligned}$$

$$\begin{aligned}\text{Mole fraction in liquid} &= \left(\frac{1}{34.083} \right) / \left[\frac{96}{34.083} + 4/61.083 \right] \\ &= 0.0101\end{aligned}$$

$$\begin{aligned}m &= \text{Mole fraction in vapor} / \text{Mole fraction in liquid} \\ &= 0.00493/0.0101 \\ &= 0.488\end{aligned}$$

Equilibrium equation-

$$y = 0.488x$$

Material balance;

$$Ls/Gs = (y1 - y2)/(x1 - x2)$$

$$x1 = m[(Gs/Ls)(y1 - y2)]/m$$

It can be seen that the “optimum” will be between m(Gm/Lm) 0.6 to 0.8,

$$\begin{aligned} x_1 &= 0.60(0.1578 - 0.00748)/0.488 \\ &= 0.1848 \end{aligned}$$

$$X1 = x1/((1 - x1))$$

$$X_1 = 0.226$$

$$X_2 = 0 \text{ (Solute free)}$$

We know that;

$$\begin{aligned} (Ls/Gs)_{min} &= (Y1 - Y2)/(X1 - X2) \\ &= (0.1870 - 0.00748)/0.226 \\ &= 0.794 \end{aligned}$$

$$\begin{aligned} (L_s)_{min} &= Gs \times 0.794 \\ &= 20.55 \text{ kmol / hr.} \end{aligned}$$

The tower is operated at the (L_s/G_s) ratio of 1.2 to 2.0 times the (L_s/G_s)_{min} ratio.

$$\begin{aligned} (Ls/Gs)_{act} &= 1.25 (Ls/Gs)_{min} \\ &= 1.25 \times 0.794 \\ &= 0.9925 \\ (L_s)_{act} &= 25.88 \times 0.9925 \\ &= 25.6859 \text{ kmol / hr.} \\ &= 1587.38 \text{ kg/hr.} \end{aligned}$$

$$\begin{aligned} (Ls/Gs)_{act} &= (Y1 - Y2)/(X_{act} - X2) \\ 0.9925 &= (0.1870 - 0.00748)/ X_{act} \\ X_{act} &= 0.1808 \\ X_{act} &= 0.1808/(1 + 0.1808) \\ &= 0.1531 \end{aligned}$$

From equilibrium eqⁿ :-

$$\begin{aligned} y_{e1} &= 0.488 \times x_{a1} \\ &= 0.488 \times 0.1531 \\ &= 0.0747 \end{aligned}$$

$$\begin{aligned} \frac{(y_1 - y_2)}{[(y_1 - y_{e1}) - (y_2 - y_{e2})]} &= \frac{(0.1578 - 0.00748)}{\{(0.1578 - 0.0747) - (0.00742 - 0)\}} \\ &= 1.987 \end{aligned}$$

$$\begin{aligned} \ln \left[\frac{(y_1 - y_{e1})}{(y_2 - y_{e2})} \right] &= \ln \frac{(0.1578 - 0.0747)}{(0.00742 - 0)} \\ &= 2.415 \end{aligned}$$

$$\begin{aligned} \text{Number of transfer unit (NTU)} &= \frac{\frac{(y_1 - y_2)}{[(y_1 - y_{e1}) - (y_2 - y_{e2})]}}{\ln \left[\frac{(y_1 - y_{e1})}{(y_2 - y_{e2})} \right]} \\ &= 0.822 \end{aligned}$$

$$\begin{aligned} \text{Gas flow rate} &= 1466.566 \text{ kg/hr.} \\ &= 0.407 \text{ kg/sec.} \end{aligned}$$

$$KGa = 6.84 \times 10^{-6} (G/360)^{0.72}$$

$$\begin{aligned} K_{Ga} &= 6.84 \times 10^{-6} (0.407/360)^{0.72} \\ &= 5.169 \times 10^{-8} \end{aligned}$$

$$\begin{aligned} G_{avg} &= (G_1 + G_2)/2 \\ &= (1466.566 + 1307.822)/2 \\ &= 1387.194 \text{ kg/hr.} \\ &= 0.3853 \text{ kg/sec.} \end{aligned}$$

$$\begin{aligned} HTU &= \frac{G_{avg}}{K_{Ga} P} \\ &= 0.3853 / (5.169 \times 10^{-8} \times 1245000) \\ &= 5.987 \end{aligned}$$

$$\begin{aligned} Z &= NTU \times HTU \\ &= 0.822 \times 5.987 \\ &= 4.92 \text{ m.} \end{aligned}$$

Absorption tower Diameter;

$$\begin{aligned} \text{Gas flow rate (G)} &= 1466.566 \text{ kg/hr.} \\ &= 1466.566/3600 \\ &= 0.407 \text{ kg/hr.} \end{aligned}$$

$$\begin{aligned}
 \text{Liquid flow rate} &= 1587.38 \text{ kg/hr.} \\
 &= 1587.38/3600 \\
 &= 0.4409 \text{ kg/sec.} \\
 \text{Density of LPG} &= 550 \text{ kg / hr.;} \\
 \text{and Density of MEA} &= 1061 \text{ kg/hr.}
 \end{aligned}$$

$$\begin{aligned}
 \text{Flooding velocity (F}_{LV}) &= \frac{L}{G} \sqrt{\rho(\text{gas})/\rho(\text{liquid})} \\
 &= \frac{0.4409}{0.407} \sqrt{550/1061} \\
 &= 0.7814
 \end{aligned}$$

From the fig.11.44

$$\begin{aligned}
 K_4 &= 85 \text{ at flooding pressure} \\
 y_1/y_2 = P_1/P_2 &= 21.26 \sim 25 \\
 K_4 &= 0.40 \text{ at design pressure drop.}
 \end{aligned}$$

$$\begin{aligned}
 \text{Flooding\%} &= \sqrt{k_4 \text{ at design pressure drop} / K_4 \text{ at flooding pressure}} \\
 &= \sqrt{(40/85)} \times 100 \\
 &= 68\%
 \end{aligned}$$

From the table 11.2.

Packing factor $F_p = 310$ for 1.5 inch Raching ring:-

$$\begin{aligned}
 \text{H}_2\text{S density at } 45^\circ\text{C } (\rho_v) &= \frac{34.083}{22.4} \times \frac{273}{318} \\
 &= 1.3062 \text{ kg/m}^3
 \end{aligned}$$

$$\begin{aligned}
 \text{Weeping velocity (V}_w^*) &= [\rho_v \times K_4 \times (\rho_L - \rho_v) / 13.1 \times F_p (\mu_L \div \rho_L)]^{1/2} \\
 &= [1.306 \times 0.40 \times (1061 - 1.306) / 13.1 \times 310 (1.67 \div 1061)]^{1/2} \\
 &= 0.5098 \text{ m/s}
 \end{aligned}$$

$$\text{Absorption column area required} = G / V * w$$

$$= 0.407 / 0.5098$$

$$= 0.798 \text{ m}^2$$

$$\begin{aligned}
 \text{Diameter of column (D)} &= \sqrt{4/\pi} \times 0.798 \\
 &= 1.00779
 \end{aligned}$$

$$\begin{aligned}
 \text{Column area (A)} &= \pi/4 \times D^2 \\
 &= 0.785 \times 1.0156 \\
 &= 0.7972 \text{ m}^2
 \end{aligned}$$

5.4 RESULTS

Table 5.2: Result for Condenser Design

Number of Transfer Units	Height of Transfer Units	Height of the tower	Cross sectional area	Diameter
NTU	HTU (m)	Z (m)	A (m ²)	D (m)
0.822	5.987	5	0.7972	1.0078

CHAPTER 6

COST ESTIMATION

6.1 ESTIMATION OF EQUIPMENT COST

The equipments involved in the process are,

1. Cooler
2. Amine absorber column
3. Caustic scrubber

6.1.1 Total Cost of Equipments used

Table 6.1: Equipments Cost

S.No	Equipment Details	No. of units	Price (Rs.)
1	Cooler	1	5000000
2	Amine absorber coloumn	1	9000000
3	Caustic scrubber	1	9000000
Total Equipment Cost (Rs.)			23000000

6.2 TOTAL CAPITAL INVESTMENT (I)

This is defined as follows,

$$I = F.I + A.I + W.I$$

Where,

I = Total Capital Investment

F.I = Fixed Investment

A.I = Investment in Auxiliary Services

W.I = Investment in Working Capital

6.2.1 FIXED INVESTMENT

This is defined as the cost on all processing equipments within the processing area limit. It is a sum of direct and indirect cost.

Total fixed Investment in plan = Direct Plan Cost + indirect Plan Cost

Direct Cost

$$\begin{aligned}\text{Installation for equipments} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 10) / 100 \\ &= \text{Rs. } 2300000\end{aligned}$$

$$\begin{aligned}\text{Land and Building} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 42) / 100 \\ &= \text{Rs.}9660000\end{aligned}$$

$$\begin{aligned}\text{Foundation} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (2300000 \times 10) / 100 \\ &= \text{Rs. } 2300000\end{aligned}$$

$$\begin{aligned}\text{Purchased Equipment Cost} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 100) / 100 \\ &= \text{Rs. } 23000000\end{aligned}$$

$$\begin{aligned}\text{Piping} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 13) / 100 \\ &= \text{Rs. } 2990000\end{aligned}$$

$$\begin{aligned}\text{Instrumentation} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 9) / 100\end{aligned}$$

= Rs. 2070000

Civil = (Equipment Cost×Direct Cost Factor)/100
= (23000000×10)/100
= Rs. 2300000

Clean Up = (Equipment Cost×Direct Cost Factor)/100
= (23000000×3)/100
= Rs. 690000

Installation = (Equipment Cost×Direct Cost Factor)/100
= (23000000×15)/100
= Rs. 3450000

Total Direct Cost = Rs. 48760000

Indirect Cost

Table 6.2 : Indirect Cost Factors

S.No	Item	% Indirect Plan cost
1	Overall Engineering	20
2	Contingency	20
3	Contractor's fee	30
	Total	70

Indirect Cost = (Direct Cost × Indirect Cost Factor)/100
= (48760000×70)/100
= Rs.34132000

Total Indirect Cost = Rs. 34132000

Total Fixed Investment = Direct Cost + Indirect Cost

Total Fixed Investment = Rs. 82892000

6.2.2 Auxiliary Investment

Table 6.3: Auxiliary Cost Factors

S.No.	Item	% of Fixed Investment
1	Auxiliary Building	5.00
2	Roads	6.00
3	Water Supply	6.00
4	Sanitary & Water Disposal	1.00
5	Process Water System	2.00
Total		20.00

Auxiliary Cost = (Total Fixed Investment × Auxiliary Cost Factor)/100
= (82892000.00×20)/100
= Rs.16578400

Total auxiliary cost = Rs.16578400

Total Installed Cost = Total Fixed Investment + Auxiliary Cost
= 82892000 + 16578400

Total Installed Cost = Rs. 99470400

6.2.3 Working Capital

It is the fixed cost saved as cash for providing operation expenses, seller of raw materials and products. This may be conveniently supposed as 12% of total fixed investment during the installed plant.

$$\begin{aligned}\text{Working Capital} &= \text{Total Installed cost} \times 12 / 88 / 100 \\ &= (99470400 \times 12) / 88 / 100\end{aligned}$$

$$\text{Working Capital} = \text{Rs. } 11936448$$

$$\begin{aligned}\text{Total Capital Investment} &= \text{Total Installed Cost} + \text{Working Capital} \\ &= 99470400 + 11936448\end{aligned}$$

$$\text{Total Capital Investment} = \text{Rs. } 111406848$$

6.3 ESTIMATION OF TOTAL PRODUCT COST

Sales Fixation

$$\text{Market Price of LPG} = \text{Rs. } 40000 \text{ per ton}$$

$$\text{LPG production increased per annum} = 1297 \times 8000$$

$$= 10376000 \text{ kg}$$

$$= 10376 \text{ ton}$$

Thus, the total product cost of manufacturing ton of LPG

$$= (10376 \times 40000)$$

$$= \text{Rs. } 415040000$$

$$\text{Total Product Cost} = \text{Rs. } 415040000 \text{ per year}$$

6.4 ESTIMATION OF MANUFACTURING COST

The Manufacturing Cost can be classified,

1. Direct Production cost
2. Fixed charges
3. Plant Overhead Costs

6.4.1 DIRECT PRODUCTION COST

$$\begin{aligned}\text{Raw Materials} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\ &= (415040000 \times 20)/100 \\ &= \text{Rs. } 415040000\end{aligned}$$

$$\begin{aligned}\text{Operating Labour} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\ &= (415040000 \times 15)/100 \\ &= \text{Rs. } 62256000\end{aligned}$$

$$\begin{aligned}\text{Direct Supervisory} &= (\text{Operating labour Cost} \times \text{Manufacturing cost factor})/100 \\ \text{\& Clerical Labour} & \\ &= (62256000 \times 15)/100 \\ &= \text{Rs. } 9338400\end{aligned}$$

$$\begin{aligned}\text{Utilities} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\ &= (415040000 \times 10)/100 \\ &= \text{Rs. } 415040000\end{aligned}$$

Maintenance &

$$\begin{aligned}\text{Repair} &= (\text{Equipment Cost} \times \text{Manufacturing cost factor})/100 \\ &= (230000000 \times 10)/100 \\ &= \text{Rs. } 2300000\end{aligned}$$

$$\begin{aligned}
 \text{Operating Supplies} &= (\text{Cost of Maintenance \& Repair} \times \text{Manufacturing cost factor})/100 \\
 &= (2300000 \times 15)/100 \\
 &= \text{Rs. 345000}
 \end{aligned}$$

$$\begin{aligned}
 \text{Laboratory Charges} &= \text{factor})/100 \\
 &= (62256000 \times 15)/100 \\
 &= \text{Rs. 9338400}
 \end{aligned}$$

$$\begin{aligned}
 \text{Patent \& Royalties} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\
 &= (415040000 \times 5)/100 \\
 &= \text{Rs. 20752000}
 \end{aligned}$$

$$(\text{Operating Labour Cost} \times \text{Manufacturing cost})$$

Total Direct Production Cost = Rs. 228841800

6.4.2 Fixed Charges

$$\begin{aligned}
 \text{Fixed charges} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\
 &= (415040000 \times 10)/100
 \end{aligned}$$

$$\text{Fixed charges} = \text{Rs. 41504000}$$

6.4.3 Plant Overhead Cost

$$\begin{aligned}
 \text{Total expense of operating labour, supervision \& maintenance} & \\
 &= (62256000 + 9338400 + 2300000) \\
 &= \text{Rs. 73894400}
 \end{aligned}$$

Plant Overhead Cost = 50% of total expense for operating labour, supervision, and maintenance

$$\begin{aligned} &= 0.5 \times (73894400) \\ &= \text{Rs. } 36947200 \end{aligned}$$

Total Manufacturing Cost = Direct Production Cost + Fixed Charges + Plant Overhead cost

$$\begin{aligned} &= 228841800 + 41504000 + 36947200 \\ &= \text{Rs. } 307293000 \end{aligned}$$

Total Manufacturing Cost = Rs. 307293000 per year

6.5 PROFIT ANALYSIS DEPRECIATION

Sinking Fund method for depreciation is,

$$D = [(V - V_w) \times i] / [(1 + r)^i - 1]$$

Where;

D = dresses amount given in every year in end.

V = placed cost of the plant = Rs. 99470400

V_w = Scrap or salvage value of plant after 'n' years = 0

N = Long lived (supposed will 15yrs.

r = Yearly interest rate ($\approx 15\%$)

$$\begin{aligned} \text{Depreciation, D} &= [(V - V_w) \times r] / [(1 + r)^n - 1] \\ &= [(99470400 - 0.15) / [(1 + 0.15)^{15} - 1]] \end{aligned}$$

$$\text{Depreciation, D} = \text{Rs } 1833668.428$$

$$\text{Depreciation} = \text{Rs. } 1833668.428$$

GROSS PROFIT

$$\begin{aligned} \text{Gross Profit} &= [\text{Net Annual Sales}] - [\text{Annual Manufacturing Cost}] \\ &= [415040000 - 307293000] \end{aligned}$$

$$\text{Gross Profit} = \text{Rs. } 107747000$$

NET PROFIT

$$\begin{aligned}\text{Net benefits} &= [\text{Gross Profit}] - [\text{Depreciation}] - [\text{Gross Profit} \times \text{Tax Rates}] \\ &= [107747000 - 1833668.428 - [107747000 \times 0.05]]\end{aligned}$$

$$\text{Net Profit} = \text{Rs. } 52039831.57$$

YEALY GET BACK RATE

Yearly percentage on the total investment after income taxes is,

$$\text{Annual Rate of Return} = (P/I) \times 100$$

Where;

P is Net Profit

I is installed cost

$$\begin{aligned}\text{Annual Rate of Return} &= (P/I) \times 100 \\ &= (52039831.57 / 99470400) \times 100\end{aligned}$$

$$\text{Annual Rate of Return} = 52.3169\%$$

PAY BACK PERIOD

$$\text{Pay Back Period} = \frac{[\text{Depreable fixed investment}]}{[\text{profit per year}] + [\text{Depreciation per year}]}$$

$$= \frac{111406848}{(52039831.57 + 1833668.428)}$$

$$\text{Pay Back Period} = 02.06 \text{ Years}$$

CHAPTER 7

PLANT LAYOUT

Plant layout involves strategically positioning the appropriate equipment in the correct locations to enable the efficient and safe processing of a product unit, minimizing both distance and period. The purpose of plant layout cover with the following:

- ❖ Providing nice working environment.
- ❖ Safely material handling.
- ❖ Giving proper space for moving of men and materials
- ❖ To make easier for future plan expansion, bottle necks and revamp.

In line with the points mentioned above, a plant layout has been proposed for the manufacturing of propylene oxide, and key aspects of the plant layout are outlined below.

- ❖ Raw materials storages are located near the production plant area.
- ❖ Loading, unloading and transport provision of the products are placed near the storage area.
- ❖ Raw material and product dispatch are kept Eastern side of the plant. So vehicular movement is controlled.
- ❖ Services department like laboratory, R&D, workshop engineering are placed alongside of the plant for decreasing the time.
- ❖ Water storage pond, surge tank, Fire stations, hydrants and fire pumps are placed within plant for safety purpose.
- ❖ For better communication, consuming time and increasing productivity; change room, control-room, washroom and canteen are provided for each department near production plant.

- ❖ Kept three gate with guards for moving of employees and other activities as vehicles, contractors and transport activities etc.
- ❖ ETP, VOC unit flares are provided in toward north.
- ❖ To main plant, sections are arranged to have minimum piping.

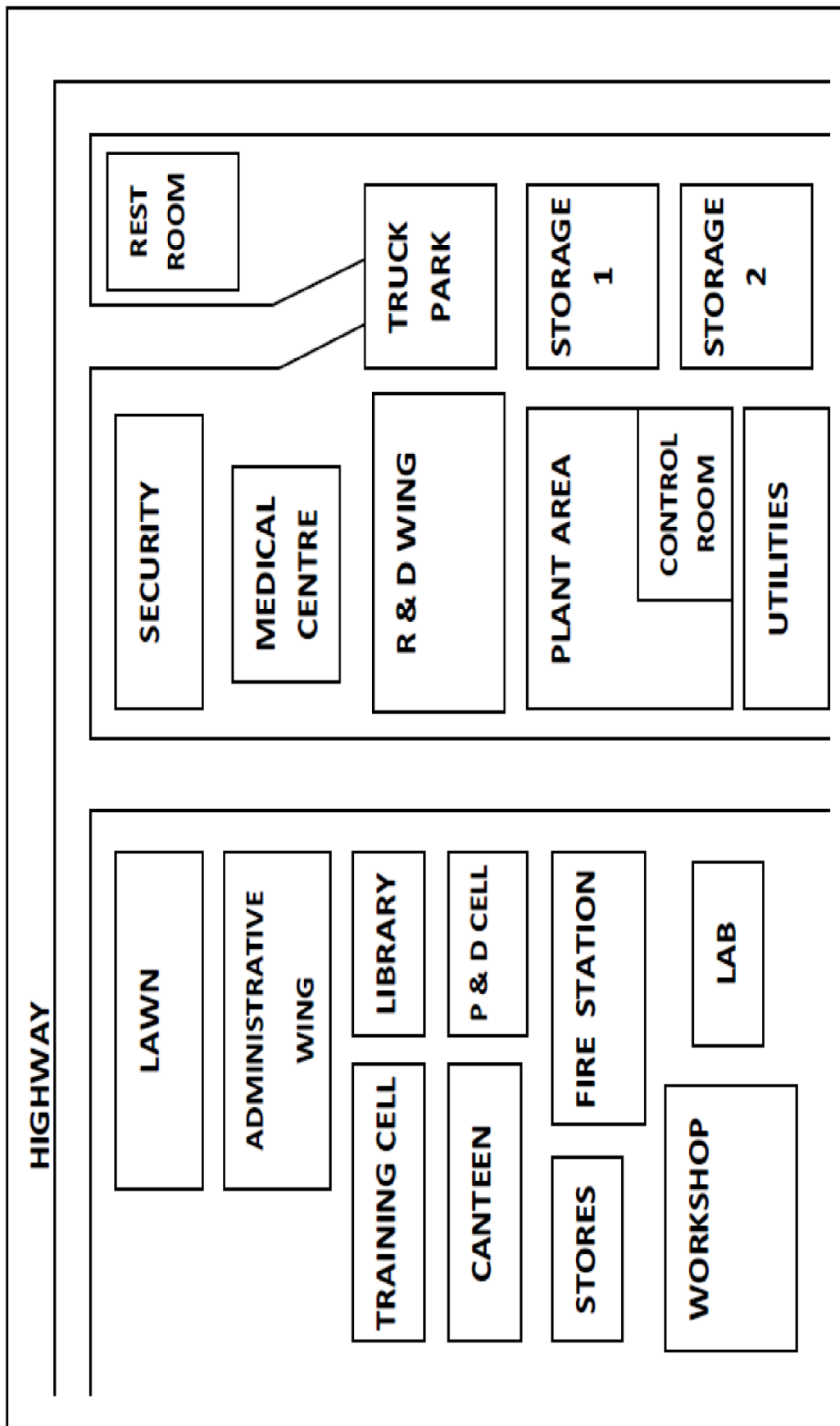


Figure 7.1: Location and Plant Layout

CHAPTER 8

SAFETY AND ENVIRONMENTAL ASPECTS

8.1 HAZOP STUDY FOR SAFETY RELATED ISSUES

A Hazard and Operability (HAZOP) review study is a methodical and structured analysis of a planned or existing process or operation. Its primary objective is to systematically identify and assess potential issues that may present risks to personnel, equipment, or disrupt the smooth functioning of the process. A HAZOP study is ideally conducted as early as possible in the design phase to influence the design and identify potential hazards. However, conducting a HAZOP requires a fairly complete design. As a practical compromise, HAZOP studies are typically performed as a final check once the detailed design has been completed. This approach helps ensure that potential risks and issues are addressed before the process becomes operational while still benefiting from a detailed design.

In this chapter, we will discuss safety in process design, with a focus on performing HAZOP (Hazard and Operability) studies to address safety-related issues:

- identifying and environment assessment of the risk
- Prevention of the risk. As flammable, harmful and toxic materials.
- Prevent of the process. For example, provision of automatic control systems and alarms in order to prevent hazardous deviations in process variables (pressure, temperature, flow).

Table 8.1 and Table 8.2 will detail the HAZOP study carried out for the system in consideration

Table 8.1 : Safety Aspects in Heat Exchangers

PARAMETER	GUIDE WORD	DEVIATION	CAUSES	CONSEQUENCES	PREVENTIVE SYSTEM	ACTIONS
Flow of heating fluid	NONE	No flow	1. Pipeline to the heating fluid side choked. 2. Pump stopped working due to Power or mechanical failure.	Affects the downstream units	Alternate pump must be provided	Pipelines and heat exchanger must be checked for choking
	MORE	High flow	1. Incorrect impeller installed to pump. 2. Pump malfunctioning	Undesired degree of cooling of product.	Alarm and flow controller must be provided	Proper control of flow and regulation of pressure
	LESS	Low flow	1. Slow running of the pump 2. Low pressure in hot fluid header	Downstream units may be affected	Alternate pump must be provided	Pipelines & heat exchanger must be checked for any choking.

Temp of heating fluid	MORE	High T	1. Heating fluid entering at higher temperature 2. Feed stream entering at higher temperature.	Pressure builds up may cause rupture of heat exchanger	Alarm and temperature controller must be provided	Proper Control and maintenance of feed stream and hot fluid flow rate
	LESS	Low T	1. High flow rate of feed stream. 2. Heating fluid entering at lower temperature	Due to low temp of hot fluid, viscosity of feed increases & affects down stream	-	Proper Control and maintenance of feed stream and steam flow rate
Pressure of heating fluid	MORE	High P	1. Pump over speeding. 2. Heating fluid side partially or fully plugged	High pressure on hot fluid side may cause rupture which may cause material mixing with other stream	Alarm and pressure controller must be provided	Pipelines and heat exchanger must be checked at regular periods for any choking
	LESS	Low P	1. Leakage in hot fluid side. 2. Pump running slowly	Leakage in heat exchanger causes mixing of hot fluid with feed and damages equipment.	-	Heat exchanger must be checked to avoid leakage

Flow of cold fluid	NONE	No flow	1. Pump stopped working due to power or mechanical failure 2. Pipeline to the feed side choked	Material entering at high temperature may affect downstream units	Alternate pump must be provided	Pipelines and heat exchanger must be checked for any choking
	MORE	High flow	1. Pump malfunctioning. 2. Pump over speeding.	High flow rate may rupture the equipment	Alarm and flow controller must be provided	Proper control of flow and regulation of pump is recommended
	LESS	Low flow	1. Incorrect impeller installed to pump 2. Pump running slowly	Affects the downstream units	Alternate pump must be provided	Pipelines and heat exchanger must be checked for any choking

Temp of cold fluid	MORE	High T	1. Hot stream entering at higher temperature 2. Low flow rate of feed at high T	Undesired degree of cooling of hot steam	Alarm and temp controller must be provided in feed stream.	Control and maintenance of feed and steam flow rate
	LESS	Low T	1. Hot stream entering at lower temperature. 2. High flow rate of feed.	Viscosity of feed increases & affects the downstream units	-	Control and maintenance of feed and fluid flow rate
Pressure of cold fluid	MORE	High P	1. Feed side partially or fully plugged. 2. Pump over speeding.	Because of high pressure feed side may rupture & causes material spillage	Alarm and pressure controller must be provided.	Pipelines & heat exchanger must be checked for choking
	LESS	Low P	1. Pump running slowly 2. Leakage in feed side	Leaking of feed may catch fire or release toxics	-	Pipelines and heat exchanger must be checked to avoid leakage

Table 8.2: Safety Aspects in Distillation Column

PARAMETER	GUIDE WORD	DEVIATION	CAUSES	CONSEQUENCES	ACTIONS
FLOW	NONE	No flow	1. Pump stopped advertently/ mechanical failure or power failure. 2. Valve choking. 3. Leakage in the pipeline.	1. Fractionation stops 2. Other downstream processes may stop working and damage some units	1. Regulation and control of crude flow rate to the column 2. Pipelines and packing must be checked for any choking
	MORE	High flow	1. Malfunctioning of pump 2. Increased vapor flow rate from the re-boiler due to high flow rate of hot fluid to re-boiler	1. Fractionation may not be of desirable degree. 2. High flow rate may damage packing of the column.	1. Regulation and control of crude flow rate to the column 2. Control of reflux ratio to the column
	LESS	Low flow	1. Decreased flow rate due to malfunctioning of pump 2. Decreased reflux flow rate	1. Fractionation may not be of desirable degree. 2. Choking may cause pressure to build up	1. Regulation of crude flow rate to the column 2. Pipelines and packing must be checked for any choking.

TEMP	MORE	High T	1. Increased flow rate of fluid of re-boiler 2. Low flow rate of cold water supply to the condenser	1. Fractionation may not be of desirable degree. 2. High temperature may damage the column and results in fire and explosion	1. Proper control of hot fluid is recommended. 2. Proper control of cold water supply the condenser
	LESS	Low T	1. Decreased flow rate of fluid of re-boiler 2. More flow rate of cold water supply to the condenser	1. Lighter components start going to top product 2. Fractionation may not be of desirable degree	1. Proper control of hot fluid is recommended 2. Proper control of cold water
PRESSURE	MORE	High P	1. Vapor because of increased flow rate of liquid to the re-boiler 2. Increased temperature of to the column	1. Boiling point of fluid in the column increases due to high pressure in the column 2. Fractionation may not be of desired degree	1. Proper control of hot fluid 2. Control and maintenance of reflux ratio to the column is required
	LESS	Low P	1. Vapor because of decreased flow rate of fluid to the re-boiler 2. Low temperature of the column	1. Components starts going to top may flash due to lower pressure resulting in fire 2. Fractionation may not be of desirable degree	1. Proper control of hot fluid is recommended 2. Maintenance of tray for any kind of channeling

CHAPTER 9

CONCLUSION

- ❖ The study of existing and revamp process has been done.
- ❖ Design of absorption tower has been completed.
- ❖ And the LPG productions were maximized by 43%.
- ❖ Costs of the equipment were estimated.
- ❖ The goal should be to always make better environmental conditions.

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Enhancing LPG Production Capacity

7

Submitted in partial fulfillment of the requirements for the award of
Bachelor of technology Degree in chemical Engineering.

By

JITENDRA YADAV (40690008)



8

DEPARTMENT OF CHEMICAL ENGINEERING

SATHYABHAMA UNIVERSITY
JEPPIAAR NAGAR, RAJIV GANDHINSALAI,
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DEPARTMENT OF CHEMICAL ENGINEERING

BONAFIDE CERTIFICATE

This is to certify that this Project Report is the bonafide work of Jitendra Yadav, (40690008) who carried out the entitled "Enhancing the LPG Production capacity" under my supervision from July 2023 to October 2023.

Internal Guide

Dr.D.Venkatesan

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DR.Sathish

Submitted For Viva Vice Examination held on _____

Internal Examiner

External Examiner

Declaration

I **Jitendra Yadav** hereby declare that the project Report entitled “**Enhancing the LPG Production capacity**” done by me under the guidance of **Dr.D.Venkatesan** is Submitted in partial fulfillment of the requirements for the award of Bachelor of Technology degree in chemical Engineering.

DATED :

PLACE:Chennai

SIGNATURE OF THE CANDIDATE

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- STUDY OF ENVIRONMENTAL SAFETY ASPECT

ABSTRACT

CPCL intends to increase the yield of the Liquefied Petroleum Gas (LPG) from the stabilizer tower in Vapor Recovery Unit (plant 2) in Refinery 1 complex. The scope of the project deals with the design considerations, economics and safety aspects of the process.

To achieve an increase in the production of LPG, the impact of new absorption and scrubbing equipment's on the off gas was studied. In the design aspect, an alternate mechanism for LPG maximization was suggested and compared with the existing mechanism. For the same, new equipment designs were done. For the modified plant, the cost of the plant was estimated and HAZOP study of the plant was carried out.

CHAPTER 1

1. INTRODUCTION

Crude oil is divided into multiple fractions, each necessitating additional processing or mixing to adhere to quality standards. This separation occurs within distillation column an atmospheric pressure and a two-stage vacuum column located in crude distillation unit. The resulting product streams include upward side distillate naphtha, high naphtha, kerosene, and diesel from the distillation column at atmospheric pressure, as well as lightest oil, weak(spindle) oil, and light oil from the first vacuum stage tower. The second vacuum tower stage yields heavy oil and vacuum residue.

The overhead naphtha is processed in a Vapor Recovery Unit (VRU), where it undergoes separate into gas, Liquefied Petroleum Gas (LPG), (LSRG), and light naphtha. The primary objective of this unit is:

- ❖ Producing LPG according to specification from distillation column feed.
- ❖ For obtaining raw hexane by feeding the treated light straight run gasoline to hexane unit.
- ❖ To feed untreated LSRG and Light Straight Run Naphtha (LSRN) to Catalytic Reforming Unit (CRU) in another refinery complex when required.
- ❖ To supply hot LSRN feed to Unifier and plat former section when required.

The basic process for the recovery of LPG from unsterilized naphtha involves the separation C3 and C4 components in the stabilizer column under temperature and pressure controlled operation. Hence in order to maximize the LPG production the following parameters of the stabilizer should be monitored on a regular basis

The top and the bottom temperature

The column pressure. Low pressure causes more propane and butane to escape into the vapor state.

→ **The vapor draw.**

In the existing scheme in vapor recovery and treatment unit, the LPG vapor is drawn off at the 45th and 47th tray in addition to the overhead draw-off. This necessitates separate condensation at each level. Coupled with high energy consumption and pumping costs, a loss of nearly 20% of LPG in the off-gas is observed in this method.

An attempt has been made to condense C3 to iCs components in the overhead vapor in a single stage, thereby eliminating the need for a side-draw. For the stabilizer column, overhead partial condenser, reboiler and the reflux pump, the following was considered based on the revamp calculations:

- ❖ The feasibility of retaining the existing equipments as they are without any modifications
- ❖ The feasibility of retaining the equipments with slight modifications.

If it is not possible to retain, then that particular equipment is replaced by new equipment, which satisfies the required specifications.

Considering the fact that there is an increased demand for LPG in the present scenario and keeping in view the limitations in the existing system, an attempt has been made to select and explore factors that could provide the means to the refinery for improvement in the profitability.

1.1 PRODUCT CHARACTERISTICS

Major product characteristics to be fulfilled are outlined in this section.

Target recovery of the various components in the distillate product stream (LPG) from the stabilizer column shall be as follows:

C ₃ recovery	:	100% (mole)
iC ₄ recovery	:	100% (mole)
nC ₄ recovery	:	98% (mole)

The impurity levels in the distillate product shall be limited to the following:

iC ₅ impurity	:	2% (mole) of that in feed
--------------------------	---	---------------------------

The impurity levels in distillate product to the LPG stripper shall not be relaxed.

LITERATURE SURVEY

CHAPTER 2

2. LITERATURE SURVEY

Detailed plant design information for the existing facility was extracted from the plant data sheet. The references utilized in this project were organized based on the type of information they provided. References related to the Stabilizer/Distillation column encompass a wide range of topics related to tray and column design. These references cover various of column and trays design. Additionally, classified related to heat exchanger and reboiler specifically address their formation and operational aspects. Stabilization is the process of eliminating lighter components from a hydrocarbon stream that is heavier in nature. In vapor recovery unit Propane, Butane and other lighter ends are separated in overhead stream and side cut of Propane and Butane (LPG).it is essential to prevent the heavier components from mixing with the lighter cut, just as it is important to avoid contamination of the heavier cut by the lighter components. Therefore, it is advisable to limit the withdrawal of side streams in order to maintain the desired separation and purity. In line with this recommendation, we suggest withdrawing a single overhead stream for condensation, and the column is designed to facilitate this operation.

Designing a column encompasses the process of establishing the operating conditions for the column, as well as determining vapor and liquid flow rates, tray count, tray diameter and area, downcomer area, and the preferred operating mode.

In the initial design phase for determining the tray count using the McCabe- Thiele method in a system with a multi-component mixture, it's often practical to assume a pseudo-binary system. It is observed that the accuracy of the product estimation may lesser than actual value and more complication in calculation especially in multi component system.

Fenske-Underwood-Gilliland correlation is used to calculate min. number of theoretical plates.

$$N_{mt} = \frac{[\ln(x_{LK}/x_{HK})_D (x_{HK}/x_{LK})_B]}{\ln \alpha}$$

Where

x_{LK} =Mole fraction of lighter end

x_{HK} =Mole fraction of heavier end

Where;

D is Distillated

and B is bottom product.

α is the relative volatility.

The number of theoretical stages under actual reflux conditions can be determined using the Gilliland correlation, which has demonstrated its effectiveness for both binary and multi-component mixtures across a broad range of reflux ratios and tray counts. This method is based on Figure 6, where the abscissa provides the Underwood constant, enabling the determination of the number of theoretical stages under actual reflux conditions. The number of theoretical stages is depend on column design and its efficiency. The column efficiency is calculated by Drickamer and Bradford correlation method by using feed conditions specially molal average viscosity. This approach has

demonstrated its reliability for hydrocarbons and related compounds. While occasional deviations of 15-20% were observed in certain cases, comparisons suggest that these deviations are generally conservative or on the safe side.

Once you've estimated the number of trays required, you can determine the column's height by multiplying it by the tray spacing. Tray spacing is typically determined to minimize entrainment. In the case of an atmospheric column, a common practice is to use a spacing of 500 mm. However, it's crucial to consider that there is a maximum height restriction for a column. Once this limit is reached, then the column design and construction cost will high as so many parameters to be considered like column foundation strength and wind velocity etc.

The maximum height of a column is contingent upon various factors, including local weather conditions, susceptibility to events such as hurricanes, typhoons, or earthquakes, and the foundation's ground conditions. Typically, the maximum height is capped at 80 meters, but it could be considerably lower in areas prone to adverse conditions. As the column's height increases, there is a stronger motivation to reduce tray spacing in order to minimize the overall column height. However, it's important to note that there is a practical limit to how small tray spacing can be, with the usual minimum tray spacing falling in the range of approximately 0.35 meters.

The sizing of the column's area and diameter depends on the specific conditions of the column. In cases where operating pressures are moderate to low, typically below 150 psig or 10.34 bars, the column diameter is primarily determined by vapor flow considerations. The above calculation based on column top conditions. The liquid and vapour velocity factor FLV is consider in to account to determine the column pressure and tray fixstation.. This factor, is correlated with the flooding velocity, to calculate the actual free area. When free area is added to the downcomer area, total area of the column, can be calculated. After calculating column area, column diameter can be calculated.

The thermosyphon reboiler is choosen effective sepatation . The separation or disengagement of phases occurs within the column itself. The reboiler can be connected

in either a once-through or a recirculating liquid arrangement, depending on the specific design and operational requirements of the system.

In horizontal reboilers of this type, the film coefficient for boiling is not influenced by the velocity of the fluid at low velocities. ¹⁸ The rate of heat transfer during boiling is primarily driven by the temperature gradient, with higher temperature differences resulting in more efficient heat transfer, rather than the fluid velocity.

The liquid from stabilizer column bottom is entered into a thermosyphon reboiler and flows the entire length of the shell. However, because ³ the mass velocity in each instance is calculated based on one-half of the total flow, it is appropriate to treat the pressure drop in a manner consistent with a shell that lacks baffles and features axial flow.

Condensing the overhead mixed vapors in a partial horizontal shell and tube condenser can lead to the separation of condensate from the vapor phase.

The design of the partial condenser involves various factors, including the heat transfer coefficient and over all heat transfer coefficient and these factors are important in the efficiency and effectiveness of the condensation process within the condenser.

EXISTING PROCESS explanation

CHAPTER – 3

3. EXISTING PROCESS EXPLANATION

3.1 FEED PREHEATING

The feedstock, from ADU overhead condensed liquid @ 40 °C that is unstabilized naphtha is fed in to the stabilizer column after preheating in a heat exchanger up to 80°C. In this heat exchanger, it comes into contact with the stabilizer bottoms, which flow in a countercurrent manner to the feed. The outlet temperature of the exchanger is controlled and monitored by a control valve.

3.2 STABILIZER COLUMN OVERHEAD/SOUR GAS

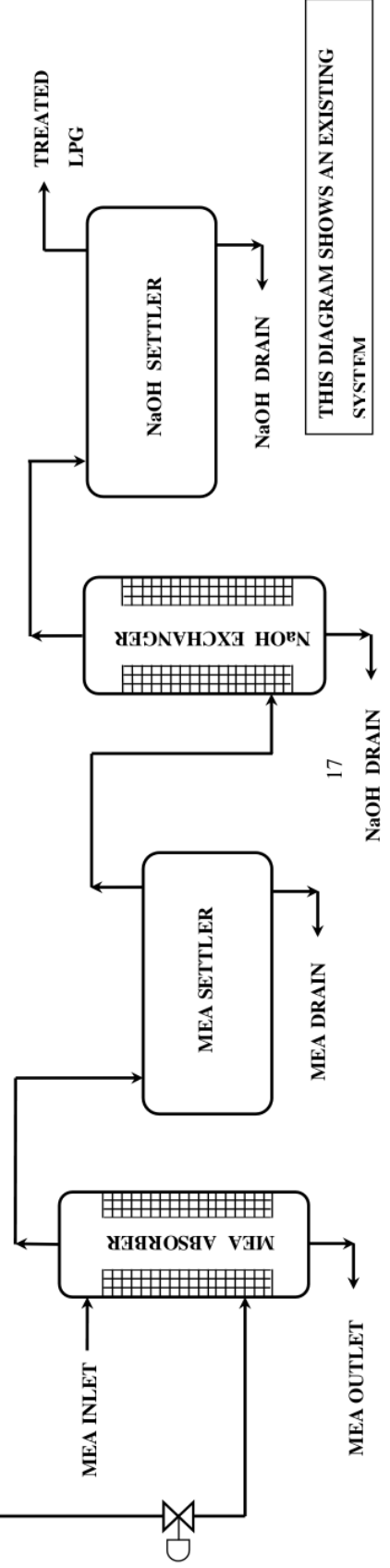
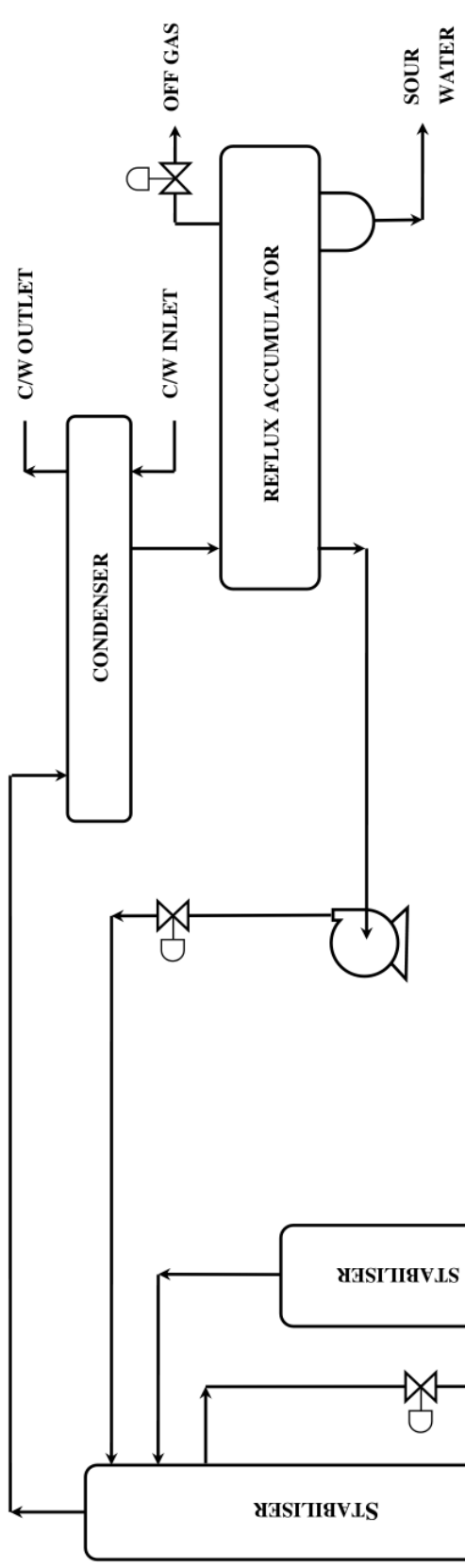
Stabilizer overhead stream contains more H₂S, called as Sour gas undergoes partial condensation to form liquid. The mixture comprising condensate and uncondensed gas is directed into the stabilizer overhead accumulator, where a separation process takes place. In accumulator, sour water is separated from the bottom by means of density difference and condensed hydrocarbon liquid is sent back to the tower as reflux by using a pump. Reflux flow is controlled by a flow controller. This controller is acting based on tower top temperature. A pressure controller valve is employed to maintain the tower pressure by adjusting the off gas flow. Off Gas is nothing but uncondensed vapor in accumulator.

The LPG from the column is flows into the LPG stripper. The purpose of the stripper is to remove lighter ends from the LPG. This lighter ends mainly Methane and Ethane components that could be carried within the LPG stream. Both the vessel and the trays within the stripper are constructed from carbon steel. Following the stripper, the LPG from the bottom section is cooled in two parallel water coolers. It is then channeled to the LPG MEA (Methyl Ethyl Alcohol) scrubber under flow control. A flow controller at the stripper outlet governs the total LPG flow, diverting a portion of the LPG stream to the LPG vaporizer as needed to maintain control.

3.3 STABILIZER BOTTOMS

The tower bottoms are reboiled in the stabilizer reboiler, which is of the horizontal thermosyphon type. The heat required for reboiler is supplied by a hot slipstream obtained from the bottom pump-around pump of the atmospheric tower. This slipstream primarily contains diesel oil. To maintain the desired temperature, a temperature controller is installed on the liquid at the fourth tray. The controller operates a three-way control valve to regulate the heat input as needed. The tower bottoms, after being reboiled, exit the system through the feed pre-heater. Following this, they exchange heat with the stabilizer feed before proceeding to the naphtha splitter column. This configuration helps efficiently utilize the heat in the system and optimizThe feed from the bottom of the stabilizer is introduced into the naphtha splitter tower between the fourth and fifth trays. The naphtha splitter tower is used to separatethe stabilizer bottom product into two main fractions: C5+95 °C Light Straight Run Gasoline (LSRG) and C6+138 °C

light naphtha. This separation process helps refine and separate the different components within the product for further processing and utilization.



THIS DIAGRAM SHOWS AN EXISTING SYSTEM

3.2 REVAMP PROCESS DESCRIPTION

3.2.1 FEED PREHEATING

The atmospheric distillation column overhead condensed liquid is @ 40 °C is fed to stabilizer column after heating it to a heat exchanger named 2E1 up to 80°C. Stabilizer bottom material which is called stabilized naphtha is reboiled in a reboiler named 2E3. Stabiliser based temperature 150°C is controlled by a threeway control valve and heating fluid is bottom pump around stream from ADU.

3.2.2 STABILIZER COLUMN OVERHEAD/SOUR GAS

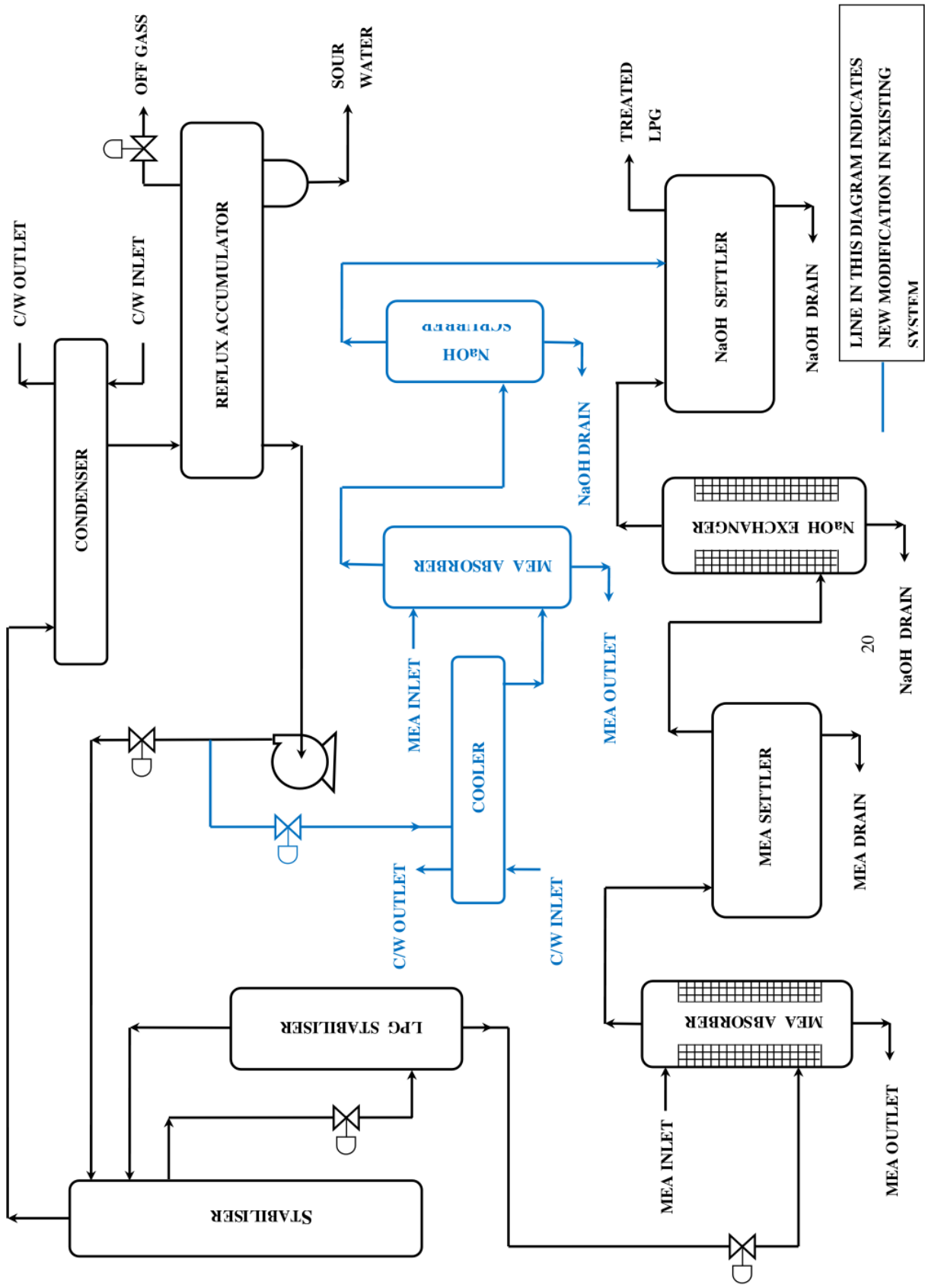
During the process, sour gas exits from the upper part of the tower. A portion of the vapor undergoes partial condensation, resulting in the formation of liquid. This mixture of condensate and uncondensed gas then enters the stabilizer overhead accumulator, where a separation process occurs. The primary purpose of the stabilizer overhead accumulator is to function as a vessel specifically designed for collecting the water that has been separated from the hydrocarbon stock in this phase of the process.

The stabilizer reflux pump takes suction from reflux drum and pump it to stabilizer column to control the tower top temperature as per process requirement. The reflux rate is meticulously managed via a three- way control valve positioned on the bypass of the stabilizer overhead pa condenser. To uphold the system's pressure at the desired level, a pressure controller valve is used. This valve governs ³the flow of sour gas from the overhead accumulator to the amine gas treating unit, ensuring that ⁶the system pressure remains within the specified range.

The flow within the system is precisely monitored and regulated by a specialized controller. Any water separated in the condenser collects in the lower portion of the accumulator. The level of this accumulated water is meticulously controlled by a dedicated level controller. To monitor the process conditions, two temperature controllers are utilized. One of these controllers measures ⁶the temperature of the vapor leaving the

tower, while the other, positioned on the discharge line from the accumulator, indicates the temperature of the reflux. These controllers play a crucial role in ensuring the system operates effectively and efficiently.

The condensed LPG from the condenser is routed under a flow control to a cooler where the condensed LPG is cooled to the required 40 °C, from the cooler the cooled condensate is routed to the LPG MEA absorber where the hydrogen sulphide impurity present in the condensate is removed. From the MEA absorber for further purification the condensate is further fed to the caustic scrubber where the rest of the impurities are removed to the possible extent. From here the LPG is sent to a LPG settling tank where heavier substances which are carried along with the stream are settled. LPG is an odorless gas, in an instance of leakage it is undetectable, hence for safety purposes 20 ppm of mercaptans is added to the final product. This gives LPG a particular smell. This final product is sent to the storage tank.



MATERIAL AND ENERGY BALANCE**4.1 MATERIAL BALANCE****4.1.1 COMPOSITIONS OF VARIOUS STREAMS:**

Basis: 30.969kmol/h of feed

The feed and product composition are given below in table.

Feed Component	Kmol/hr.	Average molecular Weight	Weight of Components Kg/hr.	Product Out kg/hr.
H ₂ O	0.231	18.015	4.161	0.041
H ₂ S	4.851	34.082	165.331	Nil
H ₂	0.319	2.016	0.643	0.643
C ₁	2.002	16.040	32.112	32.112
C ₂	1.518	30.070	45.646	45.646
C ₃	5.093	44.093	224.566	224.566
iC ₄	2.948	58.120	171.337	171.337
nC ₄	13.388	58.120	778.110	778.110
iC ₅	0.619	72.150	44.660	44.660
Total	30.969	36.967	1466.566	1297.115

4.1.2 OPERATING PARAMETERS OF THE PROCESS:

It is consider that the temperature of reflux stream is same as condenser outlet at 318⁰K and also supposed to be at its bubble point.

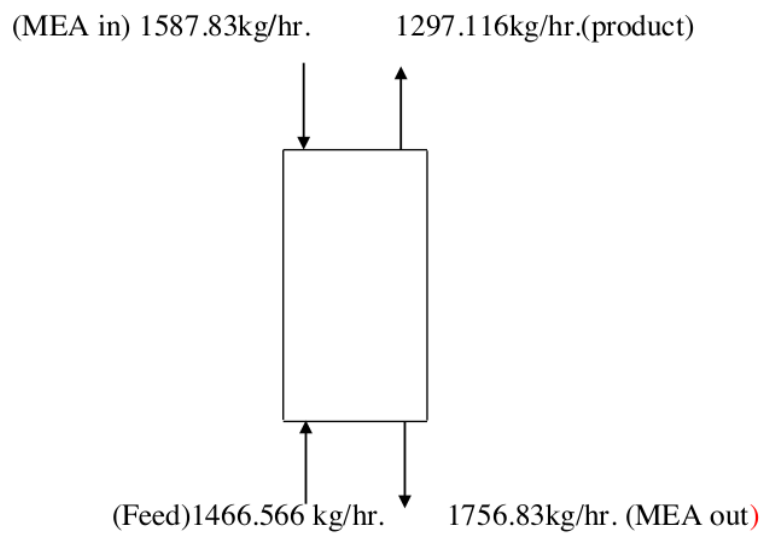
Operating pressure is 12.40kg/cm². The pressure of overhead is fixed as

$$12.40+0.5 = 12.45 \text{ kg/cm}^2$$

Table 4.3 Temperature and pressure conditions of various streams are given below-

Feed	Operating Temperature(⁰ k)	Operating Pressure (Kg/cm2)
Overhead condensed liquid	318	12.45

4.1.3 OVER ALL MATERIAL BALANCE:



General Mass Balance Equation,

Mass In + Generation = Mass Out + Consumption

(Since nil generation & consumption here)

Mass In = Mass Out

1466.566kg/hr. + 1587.38kg/hr.= 1756.83 kg/hr. + 1297.116Kg/hr.

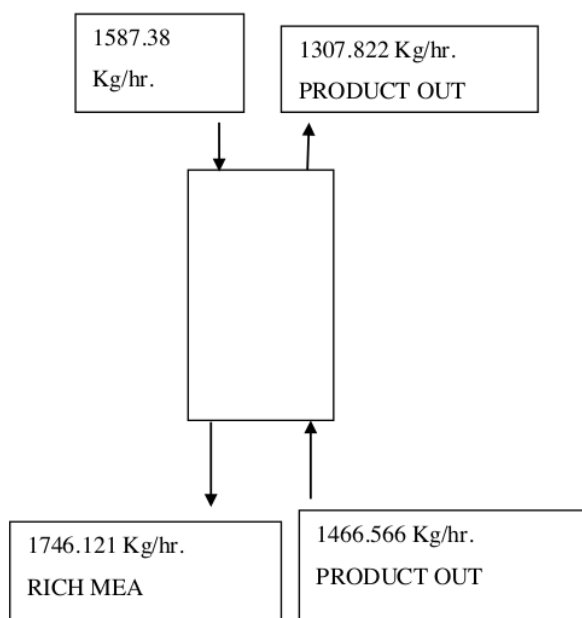
3053.946 kg/hr. = 3053.945 kg/hr.

4.1.4 MATERIAL BALANCE FOR MEA ABSORBER:

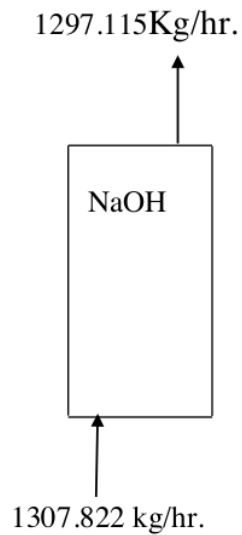
COMPOSITON	INLET OF MEA ABSORBER	OUTLET OF MEA ABSORBER
	Kg/hr.	Kg/hr.
H ₂ O	4.161	4.158
H ₂ S	165.331	6.59 @96% removed
H ₂	0.643	0.643
C ₁	32.112	32.112
C ₂	45.646	45.646
C ₃	224.566	224.566
iC ₄	171.337	171.337
nC ₄	778.110	778.110
iC ₅	44.660	44.660
Total	1466.566	1307.822

COMPOSITON	INLET SOLVENT (kg/hr.)	OUTLET SOLVENT (kg/hr.)
MEA	1587.38	1587.38
H ₂ S	0	158.741
TOTAL	1587.38	1746.121

MATERIAL BALANCE ACROSS MEA ABSORBER:



4.1.5 MATERIAL BALANCE OF NaOH SCRUBBER



General Mass Balance Equation,

Mass In + Generation = Mass Out + Consumption

(Since nil generation & consumption here)

Mass In = Mass Out + Accumulation (H₂S+Water absorbed in NaOH)

$$1307.822 = 1297.115 + 10.706 = 1307.821$$

4.1.6 INDIVIDUAL MATERIAL BALANCE

COMPOSITON	INLET OF NaOH SCRUBBER	OUTLET OF NaOH SCRUBBER
	Kg/hr.	Kg/hr.
H ₂ O	4.158	0.0415
H ₂ S	6.59	Nill
H ₂	0.643	0.643
C ₁	32.112	32.112
C ₂	45.646	45.646
C ₃	224.566	224.566
iC ₄	171.337	171.337
nC ₄	778.110	778.110
iC ₅	44.660	44.660
Total	1307.822	1297.115

ENERGY BALANCE

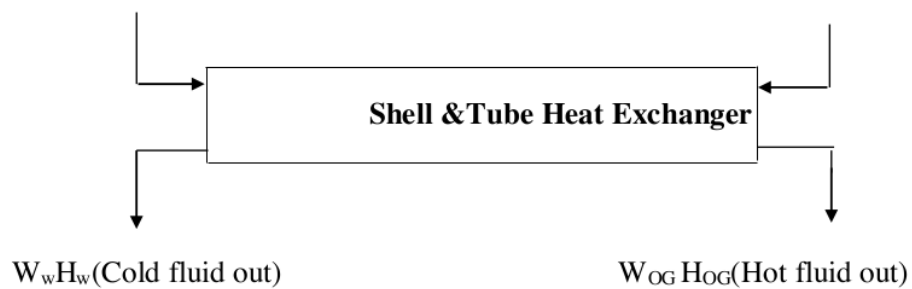
4.1.7 ENERGY BALANCE FOR COOLER

Components	Molar Flow Rate	Molecular Weight	Mass Flow Rate	Temp.	Specific Heat	Enthalpy	Heat
	Kmol/hr.	Kg/kmol	Kg/hr.	K	J/kg k	J/kg	MJ/hr.
Heat In							
Feed In	30.969	-----	1466.566	318	1482	471276	691.157
Water	36.269	18.015	653.403	301	4184	1259384	822.885
						TOTAL	1514.042
Heat Out							
Feed	30.969	-----	1466.566	313	1482	463866	680.290
Water	36.269	18.015	653.403	305	4184	1276120	833.820
						TOTAL	1514.11

. From the above table cooler duty = (691.157-680.290) MJ/hr.
= 10867KJ/hr.

4.1.8 ENERGY BALANCE AROUND THE COOLER

$W_{OG} H_{OG}$ (Hot fluid in) $W_w H_w$ (Cold fluid in)



$$\begin{aligned}
 \text{INPUT} &= W_{OG} H_{OG} (H) + W_w H_w (C) \\
 &= (1466.566 \times 471276) \text{ J/hr.} + (653.403 \times 1259384) \text{ J/hr.} \\
 &= 691.116 \text{ (MJ/hr.)} + (822.885 \text{ MJ/hr.}) \\
 &= 1514.042 \text{ MJ/hr.}
 \end{aligned}$$

$$\begin{aligned}
 \text{OUTPUT} &= (1466.566 \times 463866) \text{ J/hr.} + (653.403 \times 1276120) \text{ J/hr.} - Q_C \\
 &= (680.290 \text{ MJ/hr.}) + (834.219 \text{ MJ/hr.}) + (0.068 \text{ MJ/hr.}) \\
 &= 1514.042 \text{ MJ/hr.}
 \end{aligned}$$

Therefore; the EB is satisfied.

CHAPTER 5

DESIGN CALCULATIONS

² Heat Exchanger Design

Shell-side fluid: LPG

Tube-side fluid: Water

Given data:

Flow of the LPG (mv) = 1466.566 kg/hr

Temperature of feed in (Th in) = 45°C

Outlet temperature of the condensate (Th out) = 40°C

Cooling water in inlet (Tc in) = 28°C

Cooling water temp. of outlet (Tc out) = 32°C

Tube outer dia. (do) = 1 inch = 0.020 m

Specific heat, LPG(Cp) = 1.482 kJ/kg K

Length of the tube (L) = 2.5 m

Number of passes (Np) = 4

Inner Diameter of the tube = 0.016m

Heat transferred from vapor (Qv) = mv x(Cp) x (Th in - Th out)

$$\begin{aligned} &= 1466.566 / 3600) \times (45 - 40) \times (1.482) \\ &= 3.039 \text{ kW} \end{aligned}$$

Cooling water flow in kg (Mw) = Qv / ((Tc out - Tc in) x Cp)

$$= 3.039 / ((32 - 28) \times 4.186)$$

$$= 0.182 \text{ kg/s}$$

$$= 653.403 \text{ kg/hr}$$

Supposed ³ overall coefficient (Appendix 1, Table 14)

$$U_{\text{assumed}} = 300 \text{ W/m}^2 \text{ }^{\circ}\text{C}$$

Mean Temperature difference:

$$R = (T_{h \text{ in}} - T_{h \text{ out}}) / (T_{c \text{ out}} - T_{c \text{ in}})$$

$$= (45 - 40) / (32 - 28)$$

$$= 1.25$$

$$S = (T_{c \text{ out}} - T_{c \text{ in}}) / (T_{h \text{ in}} - T_{c \text{ in}})$$

$$= (32 - 28) / (45 - 28)$$

$$= 0.235$$

DESIGN OF CONDENSER

1) Load on condenser Q = 10867 KJ/hr.

2) Q = $UA\Delta T_{\text{LMTD}}$

Take U = 300 w/m²⁰C

3) $\Delta T_{\text{LMTD}} = \frac{\overset{19}{(Thi - Tco) - (Tho - Tci)}}{\ln(Thi - Tco)/(Tho - Tci)}$

$$= (318-305) - (313-301)/\ln(96/70)$$
$$= 12.50^{\circ}\text{C}$$

$$\Delta T_{\text{LMTD}} = 12.50^{\circ}\text{C}$$

4) Area of the condenser(A_C) = 10867/(300 × 12.50)

$$A_C = 2.8978 \text{ m}^2$$

5) Area offered by tube $A_T = \pi d_o L$

Specification of tubes

$$d_o = 20\text{mm}$$

$$d_i = 16\text{mm}$$

$$L = 2.5\text{m}$$

$$A_T = \pi d_o L$$

$$= 3.14 \times 0.02 \times 2.5$$

$$= 0.157\text{m}^2$$

6) Number of tubes

$$N_T = A_c / A_T$$

$$= 2.897 / 0.157$$

$$= 18.45$$

$$N_T = 20 \text{ tubes consider}$$

7) ²⁴ Bundle diameter

$$D_B = d_o (N_T / K_1)^{(1/n_1)}$$

From the data table

For Square pitch

$$K_1 = 0.156$$

$$N_1 = 2.291$$

$$D_B = 0.02(20/0.156)^{(1/2.296)}$$

$$D_B = 0.1659\text{m}$$

8) The tube pitch can be calculated by considering

$$P_T = 1.25 d_o$$

$$= 1.25 \times 0.02$$

$$P_T = 0.025\text{m}$$

9) Shell diameter D_s

$$D_s - D_B = C \text{ (clearance)}$$

From the chart C value found as 10mm.

$$0.01 = D_s - 0.1659$$

$$D_s = 0.01 + 0.1659$$

$$D_s = 0.1669\text{m}$$

10) ¹⁶ Tube side coefficient:

$$\text{Mean Water Temperature} = (32 + 28)/2$$

$$= 30^\circ\text{C}$$

$$\text{Tube cross sectional area} = \pi/4 d_i^2$$

$$= 3.14 \times 16^2/4$$

$$= 0.000200\text{m}^2$$

$$\text{Tupe per pass} = 20/2$$

$$= 10$$

$$\text{Total flow area} = 10 \times 0.000200$$

$$= 0.00200\text{m}^2$$

$$\text{Water mass velocity} = 653.403/(3600 \times 0.0020)$$

$$= 90.75 \text{ Kg/sm}^2$$

$$\text{Density of water} = 995 \text{ Kg/m}^3$$

$$\text{Water linear velocity} = 90.75/995$$

$$= 0.0912 \text{ m/s}$$

$$\begin{aligned}
 \text{Thermal conductivity of water} &= 614 \text{ mW/m}^0\text{C} \\
 \text{Viscosity of water} &= 0.7191 \text{ mNs/m}^2 \\
 \text{Reynolds No.} &= \frac{\rho v d_i}{\mu} \\
 &= \frac{(995 \times 0.0912 \times 0.016)}{0.0007191} \\
 &= 2019.05 \\
 \text{Consider Reynolds No.} &= 2019 \\
 \text{Prandtl No.} &= \frac{C_p k}{\mu} \\
 &= 4184 \times 0.0007191 / 0.614 \\
 &= 4.9 \\
 \text{From data table(} j_h &= 1.5 \times 10^{-1} \\
 \text{Tube side coefficient hi} &= 0.15 \times 0.614 \times 2019 \times 4.9^{.333} / 0.016 \\
 &= 19729.28 \text{ W/m}^2\text{ }^0\text{C}
 \end{aligned}$$

2
11) Shell side coefficient:

$$\begin{aligned}
 \text{Baffle spacing} &= D_s / 5 \\
 &= 0.01659 / 5 \\
 &= 0.003318 \text{ m} \\
 \text{Cross flow area } A_s &= (d_o - d_i / d_o) D_s B_s \\
 &= [(20 - 16) / 20] \times 0.1659 \times 0.003318 \\
 &= 0.0011 \text{ m}^2 \\
 \text{Mass velocity } G_s &= 1466.566 / (3600 \times 0.0011) \\
 &= 370.344 \text{ Kg/sm}^2
 \end{aligned}$$

15

$$\text{Equivalent diameter } D_e = \frac{1.1(p_r - 0.917 d_o^2)}{d_o}$$

$$= (1.1/0.020)(0.025^2 - .917 \times 0.020^2)$$

$$= 0.0142 \text{ m}$$

$$\text{Mean shell side temperature} = (45+40)/2$$

$$= 42.5^\circ\text{C}$$

$$\text{Mixture Density LPG} = 0.55 \text{ Kg/m}^3$$

$$\text{Viscosity of mixture(LPG)} = 0.08 \text{ mNs/m}^2$$

$$\text{Heat Capacity} = 1482 \text{ KJ/Kg}^\circ\text{C}$$

$$\text{Thermal conductivity} = 0.614 \text{ w/mK}$$

$$\text{Reynolds No. Re} = \rho v d_i / \mu$$

$$= (.55 \times 370.344 \times 0.0142) / .00008$$

$$= 36154.8$$

$$\text{Re} = 36155$$

$$\text{Prandtl No. Pr} = (.00008 \times 1482 \times) / .017$$

$$\text{Pr} = 6.97$$

Choose 25% baffle cut from chart

$$J_h = 4.5 \times 10^{-3}$$

$$\text{Shell side coefficient } h_s = J_h K \times N_{Re} \times (N_{Pr}^{.333}) / d_i$$

$$= (.017 \times .0045 \times 36155 \times 6.97^{.333}) / .0142$$

$$= 371.82 \text{ w/m}^2^\circ\text{C}$$

12) Over all Heat Transfer coefficient U_o :

$$\text{Thermal conductivity of metal Kw} = 43.6 \text{ w/m}^\circ\text{C for carbon steel}$$

Take the fouling coefficients from data table

for Flue gases (h_{do}) = 6000 W/m²°C

Over all co-eff. $\frac{1}{U_o} = \frac{1}{h_s} + \frac{1}{h_{do}} + \frac{d_o}{2K} \ln\left(\frac{d_o}{d_i}\right) + \frac{1}{h_d} \left(\frac{d_o}{d_i}\right) + \frac{1}{h_i} \left(\frac{d_o}{d_i}\right)$

Over all coefficient $\frac{1}{U_o} =$

$$= \frac{1}{371.82} + \frac{1}{6000} + \left(\frac{0.020}{2 \times 43.6}\right) \ln\left(\frac{20}{16}\right) + \left(\frac{20}{16 \times 6000}\right) + \left(\frac{20}{16 \times 19729.28}\right)$$

$$= 3.169 \times 10^{-3} \text{ m}^2 \text{ } ^\circ\text{C} / \text{W}$$

$$\frac{1}{U_o} = 3.169 \times 10^{-3} \text{ m}^2 \text{ } ^\circ\text{C} / \text{W}$$

Over all heat transfer coefficient $U_o = 315.55 \text{ W/m}^2 \text{ } ^\circ\text{C}$

This U_o value is well above the assumed value of 300 W/m²°C. So design has adequate area for the duty required.

DESIGN SUMMARY

Table 6.6: Condenser Design Summary

Parameter	Value Units
Flow rate of required water	653.403 Kg/hr
Number of tubes required	20
Heat transfer coefficient	300 W/m ² K
Heat transfer area calculated	2.9176 m ²
Pitch, pitch diameter	0.025 m
Shell diameter	0.1669m
Tube bundle diameter	0.1569 m
Baffle spacing	0.03318 m
Number of tube passes	2

5.1 DESIGN OF PACKED BED ABSORPTION COLOUMN

GIVEN :

Feed (LPG) $G_1 = 30.738 \text{ Kmol/hr.}$ from material balance data

Feed out Absorber Efficiency = 96%

(G_2) = 25.8875 kmol/hr. (calculated)

Feed (H_2S) = 4.851 Kmol/hr.

Inlet MEA = 25.68 Kmol/hr.

Outlet MEA = 34.56 Kmol/hr.

Remaining H_2S = 0.1941 Kmol/hr.

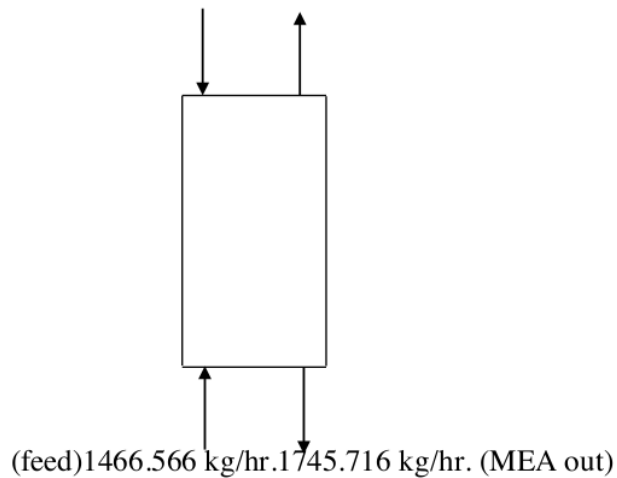
Absorbed H_2S = 4.65469 kmol/hr.

Density of LPG $\{\rho_A\} = 550 \text{ kg/m}^3$

1 atm = 1.033 kgf/cm²

11.50 kgf/cm² = 11.1307 atm

(MEA in) 1587.38 kg/hr. 1307.822 kg/hr. (product)



Fig; 1

G_s = flow of a LPG with salute free,

G_1 = Flow of Feed. (In overhead)

Y_1 = mole ratio of solute (H_2S) in Feed (overhead)

$$Y_1 = \frac{y_1 \text{ kmol of H}_2\text{S}}{(1-y_1) \text{ kmol of feed}}$$

$$G_1 = 30.738 \text{ kmol/hr.}$$

$$= 1466.566 \text{ kg/hr.}$$

G_2 = flow of in outlet feed

Y_2 = mole ratio of H_2S in outlet feed

$$Y_2 = \frac{y_2 \text{ kmol of H}_2\text{S}}{(1-y_2) \text{ kmol of outlet feed}}$$

X_2 = mole ratio of H_2S in MEA = 0

$$X_1 = \frac{x_1 \text{ kmol of H}_2\text{S}}{(1-x_1) \text{ kmol absorbed in MEA}}$$

$L_2 = L_s$ = Inlet flow of MEA with Salute free, kmol / hr.

L_1 = outlet flow of MEA with salute absorbed ,kmol / hr.

X_1 = ²² mole fraction of H_2S in liquid solvent leaving the absorber.

Total feed = 30.738 kmol/hr. (Exclusive water content)
= 1462.405 Kg/hr

H_2S Presence in feed = 4.851 kmol/hr.
= 165.331 Kg/hr

96% H_2S is absorbed in MEA and remaining 4% H_2S is carried forward to NaOH Scrubber.

4% H_2S = 0.194 kmol/hr.
= 6.59 kg/hr.

y_1 = 4.851/30.738
= 0.1578

$$\begin{aligned}
 \text{Salute free flow rate of feed (G}_s\text{)} &= G_1 (1-y_1) \\
 &= 30.738 (1-0.1578) \\
 &= 25.8875 \text{ kmol/hr.} \\
 Y_1 &= 0.1578/(1-0.1578) \\
 &= 0.1870 \\
 Y_2 &= 0.1870 \times 4/100 \\
 &= 0.00748 \\
 y_2 &= Y_2/1+Y_2 \\
 &= 0.00748/(1+0.00748) \\
 &= 0.00742
 \end{aligned}$$

$$\begin{aligned}
 \text{Partial pressure of H}_2\text{S} &= 0.5 \text{ Kpa} \\
 &= 0.5 \times 7.5 \\
 &= 3.75 \text{ mmHg}
 \end{aligned}$$

$$\begin{aligned}
 \text{Mole fraction in vapor} &= 3.75/760 \\
 &= 0.00493
 \end{aligned}$$

$$\begin{aligned}
 \text{Mole fraction in liquid} &= \left(\frac{1}{34.083} \right) / \left[\frac{96}{34.083} + \frac{4}{61.083} \right] \\
 &= 0.0101 \\
 m &= \frac{0.00493}{0.0101} \\
 &= 0.488
 \end{aligned}$$

$$\begin{aligned}
 \text{Equilibrium equation-} \\
 Y &= 0.488x
 \end{aligned}$$

Material balance;

$$\begin{aligned}
 L_s/G_s &= [(y_1 - y_2)/(x_1 - x_2)] \\
 x_1 &= [m (G_s/L_s)] [(y_1 - y_2)/m]
 \end{aligned}$$

9 It can be seen that the “optimum” will be between $m(G_m/L_m)$ 0.6 to 0.8,

$$\begin{aligned}
 X_1 &= 0.60(0.1578 - 0.00748)/0.488 \\
 &= 0.1848
 \end{aligned}$$

$$\begin{aligned}
 X_1 &= \frac{x_1}{(1-x_1)} \\
 X_1 &= 0.226 \\
 X_2 &= 0 \text{ (Solute free)}
 \end{aligned}$$

We know that;

$$\begin{aligned}
 (L_s/G_s)_{\min} &= (Y_1 - Y_2)/(X_1 - X_2) \\
 &= (0.1870 - 0.00748)/0.226 \\
 &= 0.794 \\
 (L_s)_{\min} &= G_s \times 0.794 \\
 &= 20.55 \text{ kmol/hr.}
 \end{aligned}$$

6 The tower is operated at the (L_s/G_s) ratio of 1.2 to 2.0 times the $(L_s/G_s)_{\min}$ ratio.

$$\begin{aligned}
 (L_s/G_s)_{\text{act}} &= 1.25 (L_s/G_s)_{\min} \\
 &= 1.25 \times 0.794 \\
 &= 0.9925 \\
 (L_s)_{\text{act}} &= 25.88 \times 0.9925 \\
 &= 25.6859 \text{ kmol/hr.} \\
 &= 1587.38 \text{ kg/hr.}
 \end{aligned}$$

$$\begin{aligned}
 (L_s/G_s)_{\text{act}} &= (Y_1 - Y_2)/(X_{\text{act}} - X_2) \\
 0.9925 &= (0.1870 - 0.00748)/X_{\text{act}} \\
 X_{\text{act}} &= 0.1808 \\
 x_{\text{act}} &= 0.1808/(1 + 0.1808) \\
 &= 0.1531
 \end{aligned}$$

From equilibrium eqⁿ;

$$\begin{aligned}
 y_{e1} &= 0.488 x_{\text{act}} \\
 &= 0.488 \times 0.1531
 \end{aligned}$$

$$\begin{aligned}
 &= 0.0747 \\
 \text{NTU} &= \frac{\frac{(y_1 - y_2)}{[(y_1 - y_{e1}) - (y_2 - y_{e2})]}}{\ln\left[\frac{(y_1 - y_{e1})}{(y_2 - y_{e2})}\right]} \\
 \text{Number of transfer unit (NTU)} &= \frac{\frac{(0.1578 - 0.00748)}{[(0.1578 - 0.0747) - (0.00742 - 0.0)]}}{\ln\left[\frac{(0.1578 - 0.0747)}{(0.00742 - 0)}\right]} \\
 &= 0.822 \\
 \text{Gas flow rate} &= 1466.566 \text{ kg/hr.} \\
 K_{Ga} &= 0.407 \text{ kg/sec.} \\
 &= 6.84 \times 10^{-6} (G/360)^{0.72} \\
 K_{Ga} &= 6.84 \times 10^{-6} (0.407/360)^{0.72} \\
 &= 5.169 \times 10^{-8} \\
 G_{avg} &= (G_1 + G_2)/2 \\
 &= (1466.566 + 1307.822)/2 \\
 &= 1387.194 \text{ kg/hr.} \\
 &= 0.3853 \text{ kg/sec.} \\
 \text{HTU} &= \frac{G_{avg}}{K_{Ga} P} \\
 &= \frac{0.3853}{5.169 \times 10^{-8} \times 1245000} \\
 &= 5.987 \\
 Z &= \text{NTU} \times \text{HTU} \\
 &= 0.822 \times 5.987 \\
 &= 5 \text{ m.}
 \end{aligned}$$

Absorption tower Diameter;

$$\text{Gas flow rate (G)} = 1466.566 \text{ kg/hr.}$$

$$= 1466.566/3600$$

$$= 0.407 \text{ kg/hr.}$$

$$\text{Liquid flow rate} = 1587.38 \text{ kg/hr.}$$

$$= 1587.38/3600$$

$$= 0.4409 \text{ kg/sec.}$$

Density of LPG = 550 kg/hr. and Density of MEA = 1061 kg/hr.

$$\text{Velocity factor (F}_{LV}) = \frac{L}{G} \sqrt{\rho(\text{gas})/\rho(\text{liquid})}$$

$$= \frac{0.4409}{0.407} \sqrt{550/1061}$$

$$= 0.7814$$

From the fig.11.44

$$K_4 = 85 \text{ at flooding pressure}$$

$$y_1/y_2 = P_1/P_2$$

$$= 21.26 \sim 25$$

$K_4 = 0.40$ at design pressure drop.

$$\text{Flooding\%} = \sqrt{k_4 \text{ at design pressure drop} / K_4 \text{ at flooding pressure}}$$

$$= \sqrt{(40/85)} \times 100$$

$$= 68\%$$

From the table 11.2.

Packing factor $F_p = 310$ for 1.5 inch Rasching ring.

$$\text{H}_2\text{S density at } 45^\circ\text{C} (\rho_v) = \frac{34.083}{22.4} \times \frac{273}{318}$$

$$= 1.3062 \text{ kg/m}^3$$

$$\text{Weeping velocity (V}_w^*) = [\rho_v \times K_4 \times (\rho_L - \rho_v) / 13.1 \times F_p (\mu_L \div \rho_L)]^{1/2}$$

$$= [1.306 \times 0.40 \times (1061 - 1.306) / 13.1 \times 310 (1.67 \div 1061)]^{1/2}$$

$$= 0.5098 \text{ m/s}$$

$$\begin{aligned}
 \text{Absorption column area required} &= G / V_w^* \\
 &= 0.407 / 0.5098 \\
 &= 0.798 \text{ m}^2 \\
 \text{Diameter of column (D)} &= \sqrt{4/\pi} \times 0.798 \\
 &= 1.00779 \\
 \text{Column area (A)} &= \pi/4 \times D^2 \\
 &= 0.785 \times 1.0156 \\
 &= 0.7972 \text{ m}^2
 \end{aligned}$$

Results

Number of Transfer Units	Height of Transfer Units	Height of the tower	Cross sectional area	Diameter
NTU	HTU (m)	Z (m)	A (m ²)	D (m)
0.822	5.987	5	0.7972	1.0078

CHAPTER 6

COST ESTIMATION

6.1 ESTIMATION OF EQUIPMENT COST

The equipments involved in the process are,

1. Cooler
2. Amine absorber column
3. Caustic scrubber

6.1 Total Cost of Equipments used

S.No	Equipment Details	No. of units	Price (Rs.)
1	Cooler	1	5000000
2	Amine absorber column	1	9000000
3	Caustic scrubber	1	9000000
Total Equipment Cost (Rs.)			23000000

6.2 TOTAL CAPITAL INVESTMENT (I)

This is defined as follows,

$$I = F.I + A.I + W.I$$

Where,

I = Total Capital Investment

F.I = Fixed Investment

A.I = Investment in Auxiliary Services

W.I = Investment in Working Capital

6.2.1 FIXED INVESTMENT

This is defined as the cost on all processing equipments within the processing area limit. It is a sum of direct and indirect cost.

$$\text{Total fixed Investment in plan} = \text{Direct Plan Cost} + \text{indirect Plan Cost}$$

DIRECT COST

$$\begin{aligned}\text{Installation for equipments} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 10) / 100 \\ &= \text{Rs. } 2300000\end{aligned}$$

$$\begin{aligned}\text{Land and Building} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 42) / 100 \\ &= \text{Rs. } 9660000\end{aligned}$$

$$\begin{aligned}\text{Foundation} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (2300000 \times 10) / 100 \\ &= \text{Rs. } 2300000\end{aligned}$$

$$\begin{aligned}\text{Purchased Equipment Cost} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 100) / 100 \\ &= \text{Rs. } 23000000\end{aligned}$$

$$\begin{aligned}\text{Piping} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 13) / 100 \\ &= \text{Rs. } 2990000\end{aligned}$$

$$\begin{aligned}\text{Instrumentation} &= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100 \\ &= (23000000 \times 9) / 100\end{aligned}$$

$$= \text{Rs. } 2070000$$

Civil

$$= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100$$

$$= (23000000 \times 10) / 100$$

$$= \text{Rs. } 2300000$$

Clean Up

$$= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100$$

$$= (23000000 \times 3) / 100$$

$$= \text{Rs. } 690000$$

Installation

$$= (\text{Equipment Cost} \times \text{Direct Cost Factor}) / 100$$

$$= (23000000 \times 15) / 100$$

$$= \text{Rs. } 3450000$$

Total Direct Cost **= Rs. 48760000**

INDIRECT COST

Table 6.2 Indirect Cost Factors

S.No	Item	% Indirect Plan cost
1	Overall Engineering	20
2	Contingency	20
3	Contractor's fee	30
	Total	70

$$\begin{aligned}
 \text{Indirect Cost} &= (\text{Direct Cost} \times \text{Indirect Cost Factor}) / 100 \\
 &= (48760000 \times 70) / 100 \\
 &= \text{Rs. } 34132000
 \end{aligned}$$

Total Indirect Cost = Rs. 34132000

Total Fixed Investment = Direct Cost + Indirect Cost

Total Fixed Investment = Rs. 82892000

6.2.2 AUXILIARY INVESTMENT

Table 6.3 Auxiliary Cost factors

S.No.	Item	% of Fixed Investment
1	Auxiliary Building	5.00
2	Roads	6.00
3	Water Supply	6.00
4	Sanitary & Water Disposal	1.00
5	Process Water System	2.00
Total		20.00

$$\begin{aligned}
\text{Auxiliary Cost} &= (\text{Total Fixed Investment} \times \text{Auxiliary Cost Factor}) / 100 \\
&= (82892000.00 \times 20) / 100 \\
&= \text{Rs. } 16578400
\end{aligned}$$

Total auxiliary cost = Rs. 16578400

$$\begin{aligned}
\text{Total Installed Cost} &= \text{Total Fixed Investment} + \text{Auxiliary Cost} \\
&= 82892000 + 16578400
\end{aligned}$$

Total Installed Cost = Rs. 99470400

6.2.3 WORKING CAPITAL

It is the fixed cost saved as cash for providing operation expenses, seller of raw materials and products. This may be conveniently supposed as 12% of total fixed investment during the installed plant.

$$\begin{aligned}
\text{Working Capital} &= \text{Total Installed cost} \times 12 / 88 / 100 \\
&= (99470400 \times 12) / 88 / 100
\end{aligned}$$

Working Capital = Rs. 11936448

$$\begin{aligned}
\text{Total Capital Investment} &= \text{Total Installed Cost} + \text{Working Capital} \\
&= 99470400 + 11936448
\end{aligned}$$

Total Capital Investment = Rs. 111406848

6.3 ESTIMATION OF TOTAL PRODUCT COST

Sales Fixation

Market Price of LPG = Rs. 40000 per ton

LPG production increased per annum = 1297×8000

= 1037600kg

= 10376ton

Thus, the total product cost of manufacturing ton of LPG

= (10376×40000)

= Rs. 415040000

Total Product Cost = Rs. 415040000 per year

6.4 ¹¹ ESTIMATION OF MANUFACTURING COST

The Manufacturing Cost can be classified,

1. Direct Production cost

2. Fixed charges

3. Plant Overhead Costs

6.4.1 DIRECT PRODUCTION COST

Raw Materials = $(\text{Total Product Cost} \times \text{Manufacturing cost factor}) / 100$

= $(415040000 \times 20) / 100$

= Rs. 415040000

Operating Labour = $(\text{Total Product Cost} \times \text{Manufacturing cost factor}) / 100$

= $(415040000 \times 15) / 100$

= Rs. 62256000

$$\begin{aligned}
 \text{Direct Supervisory} & \quad (\text{Operating labour Cost} \times \text{Manufacturing cost} \\
 \text{\& Clerical Labour} & \quad = \text{factor})/100 \\
 & \quad = (62256000 \times 15)/100 \\
 & \quad = \text{Rs. 9338400}
 \end{aligned}$$

$$\begin{aligned}
 \text{Utilities} & \quad = (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\
 & \quad = (415040000 \times 10)/100 \\
 & \quad = \text{Rs. 415040000}
 \end{aligned}$$

$$\begin{aligned}
 \text{Maintenance \&} & \\
 \text{Repair} & \quad = (\text{Equipment Cost} \times \text{Manufacturing cost factor})/100 \\
 & \quad = (23000000 \times 10)/100 \\
 & \quad = \text{Rs. 2300000}
 \end{aligned}$$

$$\begin{aligned}
 \text{Operating Supplies} & \quad = (\text{Cost of Maintenance \& Repair} \times \text{Manufacturing cost} \\
 & \quad = \text{factor})/100 \\
 & \quad = (2300000 \times 15)/100 \\
 & \quad = \text{Rs. 345000}
 \end{aligned}$$

$$\begin{aligned}
 \text{Laboratory Charges} & \quad = \text{factor})/100 \\
 & \quad = (62256000 \times 15)/100 \\
 & \quad = \text{Rs. 9338400}
 \end{aligned}$$

$$\begin{aligned}
 \text{Patent \& Royalties} & \quad = (\text{Total Product Cost} \times \text{Manufacturing cost factor})/100 \\
 & \quad = (415040000 \times 5)/100 \\
 & \quad = \text{Rs. 20752000}
 \end{aligned}$$

$$(\text{Operating Labour Cost} \times \text{Manufacturing cost})$$

Total Direct Production Cost = Rs. 228841800

6.4.2 Fixed Charges

$$\begin{aligned}\text{Fixed charges} &= (\text{Total Product Cost} \times \text{Manufacturing cost factor}) / 100 \\ &= (415040000 \times 10) / 100\end{aligned}$$

$$\text{Fixed charges} = \text{Rs. } 41504000$$

6.4.3 Plant Overhead Cost

$$\begin{aligned}\text{Total expense of operating labour, supervision \& maintenance} \\ &= (62256000 + 9338400 + 23000000) \\ &= \text{Rs. } 73894400\end{aligned}$$

Plant Overhead Cost = 50% of total expense for operating labour, supervision, and maintenance

$$\begin{aligned}&= 0.5 \times (73894400) \\ &= \text{Rs. } 36947200\end{aligned}$$

14
Total Manufacturing Cost = Direct Production Cost + Fixed Charges + Plant Overhead cost

$$\begin{aligned}&= 228841800 + 41504000 + 36947200 \\ &= \text{Rs. } 307293000\end{aligned}$$

Total Manufacturing Cost = Rs. 307293000 per year

6.5 PROFIT ANALYSIS DEPRECIATION

Sinking Fund method for depreciation is,

$$D = [(V - V_w) \times i] / [(1 + r)^i - 1]$$

Where;

D = dresses amount given in every year in end.

V = placed cost of the plant = Rs.99470400

V_w = Scrap or salvage value of plant after 'n' years = 0

n = Long lived (supposed will 15 years)

r = Yearly interest rate ($\approx 15\%$)

$$\begin{aligned} \text{Depreciation, } D &= [(V - V_w) \times r] / [(1 + r)^n - 1] \\ &= [(99470400 - 0.15) / [(1 + 0.15)^{15} - 1]] \end{aligned}$$

$$\text{Depreciation, } D = \text{Rs } 1833668.428$$

$$\text{Depreciation} = \text{Rs. } 1833668.428$$

GROSS PROFIT

$$\begin{aligned} \text{Gross Profit} &= [\text{Net Annual Sales}] - [\text{Annual Manufacturing Cost}] \\ &= 415040000 - 307293000 \end{aligned}$$

$$\text{Gross Profit} = \text{Rs. } 107747000$$

NET PROFIT

$$\text{Net benefits} = [\text{Gross Profit}] - [\text{Depreciation}] - [\text{Gross Profit} \times \text{Tax Rates}]$$

$$= [107747000 - 1833668.428 - [107747000 \times 0.05]]$$

Net Profit = Rs. 52039831.57

YEALY GET BACK RATE

Yearly percentage on the total investment after income taxes is,

$$\text{Annual Rate of Return} = (P/I) \times 100$$

Where;

P is Net Profit

I is installed cost

$$\text{Annual Rate of Return} = (P/I) \times 100$$

$$= (52039831.57 / 99470400) \times 100$$

Annual Rate of Return = 52.3169%

PAY BACK PERIOD

$$\begin{aligned} \text{Pay Back Period} &= \frac{[\text{Depreable fixed investment}]}{[[\text{profit per year}] + [\text{Depreciation per year}]]} \\ &= 111406848 / (52039831.57 + 1833668.428) \end{aligned}$$

Pay Back Period = 02.06 Years

CHAPTER 7

PLANT LAYOUT

Plant layout involves strategically positioning the appropriate equipment in the correct locations to enable the efficient and safe processing of a product unit, minimizing both distance and period. The purpose of plant layout cover with the following, →

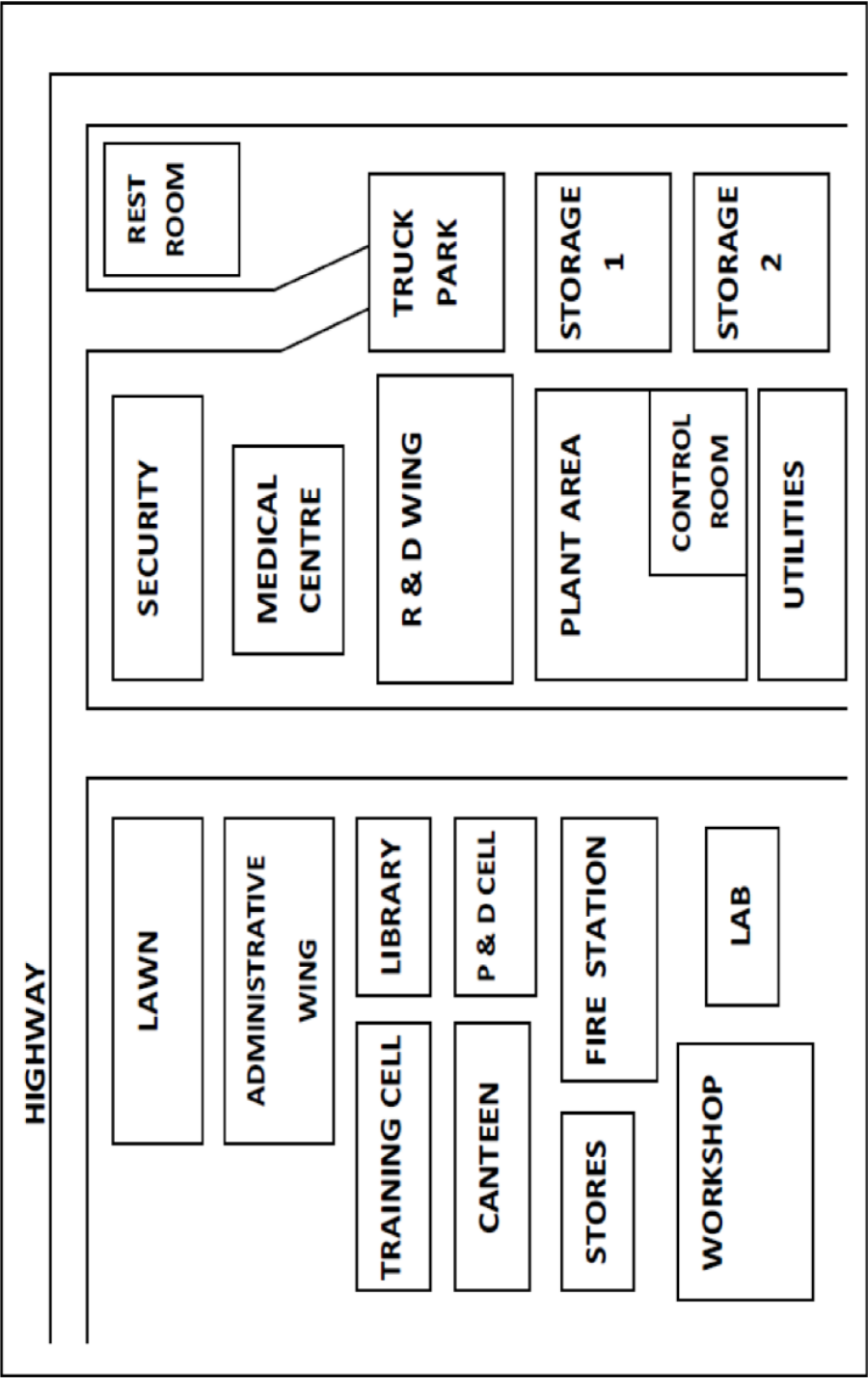
- ❖ Providing nice working environment.
- ❖ Safely material handling.
- ❖ Giving proper space for moving of men and materials
- ❖ To make easier for future plan expansion,bottle necks and revamp.

In line with the points mentioned above, a plant layout has been proposed for the manufacturing of propylene oxide, and key aspects of the plant layout are outlined below.

- ❖ Raw materials storages are located near the production plant area.
- ❖ Loading,unloadingand transportprovision of the products are placed near the storage area.
- ❖ Raw material and product dispatch are kept Eastern side of the plant. So vehicular movement is controlled.
- ❖ Services department like laboratory, R&D, workshop engineering is placed alongside of the plant for decreasing the time.

- ❖ Water storage pond, surge tank, Fire stations, hydrants and fire pumps are placed within plant for safety purpose.
- ❖ For better communication, consuming time and increasing productivity; change room, control-room, washroom and canteen are provided for each department near production plant.

- ❖ Kept three gate with guards for moving of employees and other activities as vehicles, contractors and transport activities etc.
- ❖ ETP, VOC unit flares are provided in toward north.
- ❖ To main plant, sections are arranged to have minimum piping.



CHAPTER 8

SAFETY AND ENVIRONMENTAL ASPECTS

8.1 HAZOP STUDY FOR SAFETY RELATED ISSUES

⁴ A Hazard and Operability (HAZOP) review study is a methodical and structured analysis of a planned or existing process or operation. Its primary objective is to systematically identify and assess potential issues that may present risks to personnel, equipment, or disrupt the smooth functioning of the process.

A HAZOP study is ideally conducted as early as possible in the design phase to influence the design and identify potential hazards. However, conducting a HAZOP requires a fairly complete design. As a practical compromise, HAZOP studies are typically performed as a final check once the detailed design has been completed. This approach helps ensure that potential risks and issues are addressed before the process becomes operational while still benefiting from a detailed design.

In this chapter, we will discuss safety in process design, with a focus on performing HAZOP (Hazard and Operability) studies to address safety-related issues:

- fingering and environment assessment of the risk.
- Prevention of the risk. As flammable, harmful and toxic materials.
- Prevent of the process. For example, provision of automatic control systems and alarms in order to prevent hazardous deviations in process variables (pressure, temperature, flow).

Table 8.1 and Table 8.2 will detail the HAZOP study carried out for the system in consideration.

8.1.1 Table 8.1 Safety Aspects in Heat Exchangers

PARAMETER	GUIDE WORD	DEVIATION	CAUSES	CONSEQUENCES	PREVENTIVE SYSTEM	ACTIONS
Flow of heating fluid	NONE	No flow	1. Pipeline to the heating fluid side choked. 2. Pump stopped working due to Power or mechanical failure.	Affects the downstream units	Alternate pump must be provided	Pipelines and heat exchanger must be checked for choking

	MORE	High flow	1. Incorrect impeller installed to pump. 2. Pump malfunctioning	Undesired degree of cooling of product.	Alarm and flow controller must be provided	Proper control of flow and regulation of pressure
	LESS	Low flow	1. Slow running of the pump 2. Low pressure in hot fluid header	Downstream units may be affected	Alternate pump must be provided	Pipelines & heat exchanger must be checked for any choking.

Temp of heating fluid	MORE	High T	1. Heating fluid entering at higher temperature 2. Feed stream entering at higher temperature.	Pressure builds up may cause rupture of heat exchanger	Alarm and temperature controller must be provided	Proper Control and maintenance of feed stream and hot fluid flow rate
	LESS	Low T	1. High flow rate of feed stream. 2. Heating fluid entering at lower temperature.	Due to low temp of hot fluid, viscosity of feed increases & affects down stream	-	Proper Control and maintenance of feed stream and steam flow rate

Pressure of heating fluid	MORE	High P	1. Pump over speeding. 2. Heating fluid side partially or fully plugged	High pressure on hot fluid side may cause rupture which may cause material mixing with other stream	Alarm and pressure controller must be provided	Pipelines and heat exchanger must be checked at regular periods for any choking
	LESS	Low P	1. Leakage in hot fluid side. 2. Pump running slowly	Leakage in heat exchanger causes mixing of hot fluid with feed and damages equipment.	-	Heat exchanger must be checked to avoid leakage

Flow of cold fluid	NONE	No flow	1. Pump stopped working due to power or mechanical failure 2. Pipeline to the feed side choked	Material entering at high temperature may affect downstream units	Alternate pump must be provided	Pipelines and heat exchanger must be checked for any choking
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Temp of cold fluid	MORE	High T	1. Hot stream entering at higher temperature 2. Low flow rate of feed at high T	Undesired degree of cooling of hot steam	Alarm and temp controller must be provided in feed stream.	Control and maintenance of feed and steam flow rate
	LESS	Low T	1. Hot stream entering at lower temperature. 2. High flow rate of feed.	Viscosity of feed increases & affects the down stream units	-	Control and maintenance of feed and fluid flow rate
Pressure of cold fluid	MORE	High P	1. Feed side partially or fully plugged. 2. Pump over	Because of high pressure feed side may rupture & causes material spillage	Alarm and pressure controller must be provided.	Pipelines & heat exchanger must be checked for

			speeding.			Choking
			1. Pump running slowly 2. Leakage in feed side	Low P		Pipelines and heat exchanger must be checked to avoid leakage
					-	
					Leaking of feed may catch fire or release toxics	

8.1.2 Table 8.2: Safety aspects in distillation column

PARAMETER	GUIDE WORD	DEVIATION	CAUSES	CONSEQUENCES	ACTIONS
FLOW	NONE	No flow	1. Pump stopped advertently/ mechanical failure or power failure. 2. Valve choking. 3. Leakage in the pipeline.	1. Fractionation stops 2. Other downstream processes may stop working and damage some units	1. Regulation and control of crude flow rate to the column 2. Pipelines and packing must be checked for any choking
	MORE	High flow	1. Malfunctioning of pump 2. Increased vapor flow rate from the reboiler due to high flow rate of hot fluid to	1. Fractionation may not be of desirable degree. 2. High flow rate may damage packing of the column.	1. Regulation and control of crude flow rate to the column 2. Control of reflux ratio to the column

			reboiler		
	LESS	Low flow	1. Decreased flow rate due to malfunctioning of pump 2. Decreased reflux flow rate	1. Fractionation may not be of desirable degree. 2. Choking may cause pressure to build up	1. Regulation of crude flow rate to the column 2. Pipelines and packing must be checked for any choking.

TEMP	MORE	High T	1. Increased flow rate of fluid of reboiler 2. Low flow rate of cold water supply to the condenser	1. Fractionation may not be of desirable degree. 2. High temperature may damage the column and results in fire and	1. Proper control of hot fluid is recommended. 2. Proper control of cold water supply the condenser
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PRESSURE					explosion	
	LESS	Low T	1.Decreased flow rate of fluid of reboiler 2.More flow rate of cold water supply to the condenser	1.Lighter components start going to top product 2.Fractionation may not be of desirable degree	1.Proper control of hot fluid is recommended 2.Proper control of cold water	
	MORE	High P	1.Vapor because of increased flow rate of liquid to the reboiler2.Increased temperature of to the column	1.Boiling point of fluid in the column increases due to high pressure in the column 2.Fractionation may not be of desired degree	1.Proper control of hot fluid 2.Control and maintenance of reflux ratio to the column is required	

	LESS	Low P	1.Vapor because of decreased flow rate of fluid to the reboiler 2.Low temperature of the column	1.Components starts going to top may flash due to lower pressure resulting in fire 2.Fractionation may not be of desirable degree	1.Proper control of hot fluid is recommended 2.Maintenance of tray for any kind of channeling
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Conclusion

- The study of existing and revamp process has been done.
- Design of absorption tower has been completed.
- And the LPG production were maximized by 43%.
- Cost of the equipment were estimated.
- The goal should be to always make better environmental conditions.

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