# PROCESS DESIGN OF GAS DEHYDRATION UNIT

A thesis submitted in partial fulfillment of the requirements for the Degree of
Master of Technology
in
Refining & Petrochemical Engineering

By
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Under the guidance of

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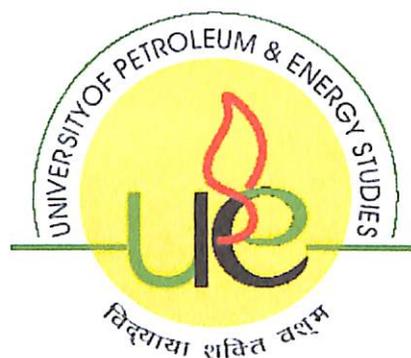
College of Engineering
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Dehradun
May, 2009

CERTIFICATE

This is to certify that the work contained in this thesis titled "**PROCESS DESIGN OF GAS DEHYDRATION UNIT**" has been carried out by Mr. VITHAL REDDY under my supervision and has not been submitted elsewhere for a degree.

7.5.09

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Abstract

Dehydration is an essential step in the processing of natural gas. A preliminary design of the dehydration process has been carried out using try ethylene glycol (TEG) as a desiccant.

The work comprised of analysis of overall dehydration process flow and a preliminary design of the absorber unit to achieve the product specification. Empirical approach is followed in most of the calculations. The work also covers manual calculations for equipment sizing to find the major equipment parameters. Line sizing procedure has been discussed briefly.

Finally the calculated design parameters are compared with design values. A satisfactory match between the calculated results and design values validate the calculation procedures.

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NOMENCLATURE

V	Mass flow rate Vapor, kg / hr
L	Mass flow rate Liquid, kg / hr
ρ_v	Density of the Vapor, kg / m^3 or lb / ft^3
ρ_l	Density of the Liquid, kg / m^3 or lb / ft^3
γ	Kinematic Viscosity of the liquid , cSt
γ_A	Activity coefficient, From figure 5.1
K	Equilibrium constant for water in a TEG – water system.
E_a	Absorption Efficiency
A_T	Absorption factor
Y_{n+1}	flow rate of water in the entering rich gas , kg / hr
Y_1	flow rate of water in the dry gas, kg / hr
n	Number of stages
ΔP	pressure drop, in of H$_2$O / ft of packing
G	Gas mass velocity, kg / m^2 sec or lb / ft^2 sec
F_P	packing factor
D	diameter of the Tower, in or m
W	Water content on mass/vol basis, at saturation, as found from regular water content correlation
B	Water content on mass/ vol basis
Q_V	Volumetric flow rate of gas, ft^3 / hr
Q_{LL}	Volumetric flow rate of Light Liquid, ft / hr
Q_{HL}	Volumetric flow rate of Heavy Liquid, ft^3 / hr
ρ_H	Density of Gas, lb / ft^3
ρ_w	Density of Water, lb / ft^3
μ_L	Viscosity of the Light Liquid, cp
θ_{LL}, θ_{HH},	Residence time for Light and Heavy liquids, min
T_H, T_S	Hold up Time and Surge time, min
P	Pressure in Psi a
K	Terminal velocity constant, ft / sec
U_T	Terminal Velocity , ft / sec
U_v	Vapor Velocity , ft / sec
D	diameter of the vessel, ft or in
L	length of the vessel, ft
A_T	total cross sectional area, ft^2
H	height, ht or in
H_V	vapor space area, ft
L_1	Length upto first bucket, ft
L_{min}	Vapor / liquid separation minimum length, ft
U_{VA}	actual vapor velocity, ft / sec
ϕ	Liquid dropout time, sec
H_{LL}	light layer thickness, in

ΔSG	Specific Gravity between light and heavy liquids
D_P	Droplet size, microns
Δ_H	Height difference between light and heavy liquid weirs, ft
L_2	length of the first bucket, ft
L_3	length between two buckets, ft
L_4	length of the second bucket, ft
P_A, P_B	vapour pressure of water and TEG, mm of Hg
$X_{LD}, X_{LB}, X_{HD} \& X_{HB}$	composition of more and less volatile component in distillate and in residue
$\alpha_1, \alpha_2 \& \alpha_{AVG}$	Relative volatility at top, bottom and average of the column
S_M	Minimum number off stages.
d	Pipe inside diameter, in
Q	Gas flow rate, ft^3 / hr
V	average flow velocity, ft / sec
ΔP	pressure drop, psi
P_I	inlet pressure, psi
T_I	Inlet temperature, deg R
Z	Compressibility factor
L	Length, ft
f	Moody Friction factor
S	Specific gravity
Re	Reynolds Number
μ	Liquid Viscosity, lb / ft $-$ sec

Chapter 1
INTRODUCTION

Natural Gas:

A naturally occurring mixture of hydrocarbon and non-hydrocarbon gases found in porous geological formations beneath the earth's surface, often in association with oil is known as natural gas. The principal constituent in natural gas is methane (CH_4).

Natural gas is formed from the buried remains of tiny plants and sea animals that died more than 200 million years ago in mother earth. Under the heat and pressure inside the earth, these energy rich organic material slowly decayed, then changed form until all that was left were concentrations of natural gas in layers of rocks.

Natural gas is removed from the earth by drilling wells into the rock, then using pipes to bring gas to the surface. In most wells, the pressure of the gas is enough to force it to the surface and then to the central collection points. When natural gas comes out along with crude oil, it is known as associated gas.

A typical composition of dry natural gas is as shown in table 1.1.

Composition	Mole %
Methane	78.14
Ethane	8.63
Propane	6.13
i – Butane	1.23
n – Butane	1.69
i – Pentane	0.4
n – Pentane	0.39
Hexane	0.43
CO_2	2.52
N_2	0.20

Table 1.1   Dry Natural Gas Compositions

World Scenario:

Col. Drake well drilled world's first on-land in the Pennsylvania State of USA in 1859. The first offshore field was discovered in the Gulf of Mexico in 1938 and a well was drilled up to 9000 feet off the Texas coast. Enormous gas fields are found in Russia, Iran and Qatar. Saudi Arabia has tremendous gas potential. Through a program of exploration in the 1990's, Saudi Arabia has increased its gas reserves to 235 trillion cubic feet (tcf), including more than 90 tcf of non-associated gas.

Indian Scenario:

Mr. Goodenough of Mckillip Stewart Company first drilled a well near Jaipur in Upper Assam and oil was struck at a depth of 36 m, just seven years after Col. Drake discovered oil in Pennsylvania. The first well drilled by ONGC at Jwalamukhi in 1958 struck gas in small quantity. In the same year oil was struck in Cambay. Thereafter followed the discovery of Ankleshwar (1960), Sanand (1962), Geleki (1968) etc. Indian offshore covers a sedimentary area of about 3,20,000 Km^2, out of which Bombay offshore basin is about 1,40,000 sq. km. upto water depth of 200 M. Total of the sedimentary basin in about 1,000,000 sq. Km. i.e. approximately 40% of total land area of India. Indian Offshore has been divided into 10 sedimentary basins, Kutch, Saurashtra, Bombay, Daman, Kerala, Cauvery, Godavari, Mahanadi, Bengal and Andaman & Nicobar islands. The first process platform (NF or BHF) was commissioned on 13.06.78 and the crude oil and gas were sent to the shore terminal at Uran by two separate subsea pipe trunk lines 30" x 203 km long line and 26" x 203 km long line respectively. The first major process platform (BHN) was commissioned on 11.02.81 in the Bombay High North. The first major process platform in the South (BHS), was commissioned on 25.08.82 which facilitated transportation of oil and gas from south field to Uran. Right now there are 14 process platforms (BHF, BHN, WIN, NQG, NQO & NQP in Bombay High North; BHS, WIS, SCA, SHP, SHG, SHW, ICP & ICW in Bombay High South, 4 well cum process platforms (NA, NC, SA and SHD),83 well platforms and 3 SBMs (NA, SA, SHD) in Bombay High. Natural Gas is found onshore in the North East, Rajasthan and Gujarat and offshore in Bombay High. Recently, Reliance Industries have reported gas findings in the Krishna, Godavari Basin.

Gas Processing and Conditioning:

Water vapour and acid gas (H_2S ans CO_2) are the most common undesirable impurities found in gas streams. Presence of water vapors besides corroding the pipeline, leads to formation of solid gas hydrates which adversely affects the transportation. Free water has to be removed from the produced gas before it enters a pipeline for the following fundamental reasons:

1. To prevent the formation of hydrates that will restrict or block the flow of gas in the line, plug valves, fittings, etc...

2. To prevent corrosion in the line.

3. To prevent slugging and erosion.

4. To prevent increase in Specific volume and decrease in heating value of gas.

5. To minimize free water condensing in the pipeline thereby reducing the internal cross – sectional area of the pipe available for flow and causing partial blockage and consequentially reduced gas flow throughput

To prevent such difficulties, essentially gas stream, which is transported in transmission lines, is dehydrated as per pipeline specifications. Most of the liquid free water associated with extracted natural gas is removed by simple separation methods at or near the wellhead. However, the removal of the water vapor requires more complex treatment that is gas dehydration.

There are various methods used for gas dehydration such as

1. Absorption using liquid desiccants

2. Adsorption using solid desiccants

3. Dehydration using $CaCl_2$

This project work mainly discuss about the unit operations encountered in the Absorption using liquid desiccant method and design of the major equipments.

Chapter 2
LITERATURE REVIEW

Absorption, [7]

In this operation, a soluble component in the gas phase is dissolved in the liquid phase. This is done by either physical absorption or by chemical reaction.

Stripping

This is the opposite of absorption. That is, it is the removal by the application of heating or by the application of a lean gas of the soluble liquid components. It is actually distillation when heat is applied to generate the stripping medium.

Tower Sizing Packed Columns

For the design of packed columns, the generalized pressure drop correlation can be used. It is a plot of the gas flow multiplied by several factors pertaining to the type of packing and the fluids being fractionated against a liquid to gas ratio. This plot is given in Figure 7.3 This plot is shown in Perry, This correlation shows pressure drop lines. The design procedure is to size the tower to operate at about 80% of flooding which is about 0.25 to 0.5 inch of H_2O/ft. For foaming solutions use 0.1 inch of H_2O/ft of packing.

$$X - axis: \frac{L}{V}\sqrt{\frac{\rho_V}{\rho_L}}$$

$$Y - Axis: \frac{G^2 F_p \gamma^{0.1}}{\rho_V(\rho_L - \rho_V)}$$

Where,

L = liquid mass flow lb/hr/ft2

G=gas flow mass velocity lb/hr/ft2

ρ_v = gas density lb/ft3

ρ_L = liquid density lb/ft3

ρ_w = density of water

μL = liquid viscosity in centipoise

Fp = packing factor as given in the following table

Tower Height Absorption

Let us consider what happens in a tower when absorption takes place. Consider first a two-component system in both gas and liquid phases. In gas, assume pure gas with solute and in liquid assume pure solvent and solute. If we allow gas phase and liquid to come together they will eventually reach equilibrium. Thus we can, for fixed temperature and pressure conditions, get an equilibrium curve showing the relation between the concentration of the solute in the gas phase and in the liquid phase.

For example, here are some curves for NH_3 in air and water:

Solute - material being absorbed

Solvent - carrier fluid
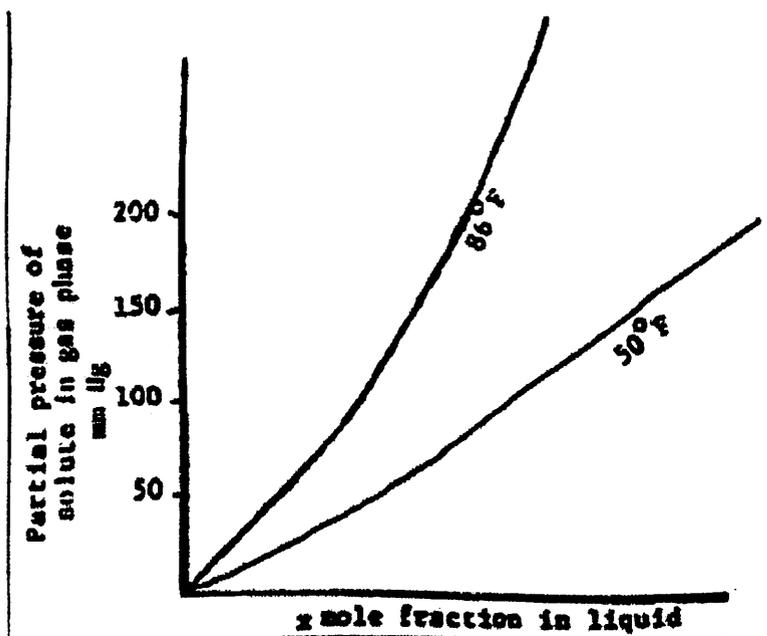
Figure 2.1   Absorption curves

These curves can sometimes be expressed by:

Raoult's Law (when have ideal solution and gas)

$P_y = P_{vp} \cdot x$

Where,

P_v = partial pressure of solute in the gas

x = mole fraction of the solute in solvent.

P_{vp} = vapour press.

Henry's Law.

$P_v = m'. x$, m' is Henry's law constant

This law is more often applicable.

Absorption of a Single Component

As in distillation, we can apply a graphical method for solving the number of theoretical plates required in an absorber absorbing one component from an inert carrier gas and with a non-volatile solvent. We plot an equilibrium curve with ordinates of
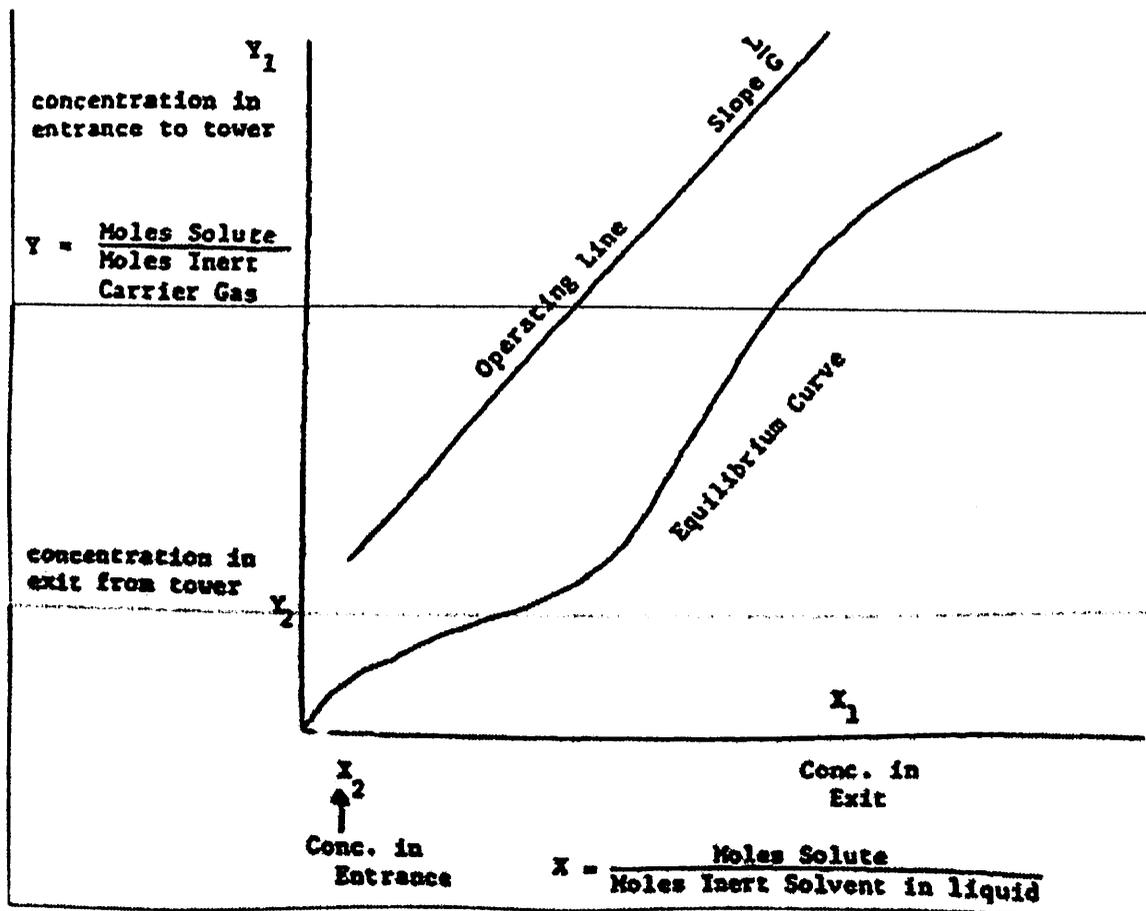
Figure 2.2   More absorption curves

Figure 2.3 shows the tower to which figure 2.2 refers. Let us consider a counter current absorber
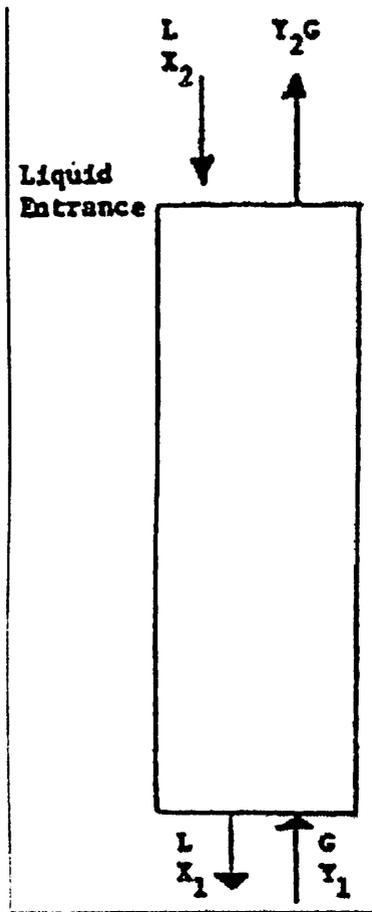
Figure 2.3   Counter current absorption

A stripper is just an inverted absorber as shown in figure 2.4
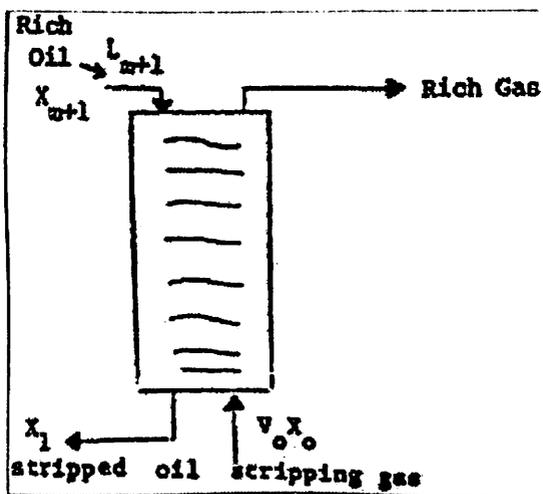
Figure 2.4   Stripper operation

Simplified methods for finding the number of trays in absorption column:

Absorption (Short Cut Method)

Given the gas rate, the oil rate and the % absorption of the key component

1. Calculate from rich gas rate and lean oil rate.

2. From operating temperature and pressure find K for key component and thus calculate L/VK

3. Go to chart figure 15-50 in GPSA Data book and read the number of theoretical trays required for the desired key component recovery.

4. This gives a very conservative value (a larger number of trays than probably required), thus it is probably better to use one of the methods discussed further on

Kremser Brown Method of Finding the Number of Trays for Absorption

Let us consider the Kremser Brown method of finding the number of trays for a specific absorption. Given the gas rate, the oil rate and % absorption of the key component.

1. Calculate L/V from rich oil rate (have to estimate amount of liquid absorbed) and rich gas rate

2. From operating temperature and pressure find K for the key component and thus calculate A= L/KA (Key comp. one being absorbed)

3. Go to chart Figure 15-51 in GPSA Data Book and read the number of theoretical trays

4. Calculate % recovery of other components for this number of trays and the same L/V ratio but a different K value because of a different component

5. Check to see that L assumed is OK - as long as it is not too much lower than that assume that it is probably OK to use.

1. Simulation is carried out in HYSYS. The Fluid package used here is Peng-Robinson Equation. This is based on [2, 12]

2. Absorption column, stripping column and still column are packed bed towers. Generally, structured packing is used. The sizing procedure was given in many literatures. [1,3,4] Industrial practice for sizing these equipments are based on [1]

3. Flash tank is generally Three phase Bucket and Weir separator. The sizing is based on [6]

4. Line sizing is based on [9, 10]

Chapter 3

GAS DEHYDRATION

Gas dehydration is the process of removing water vapor from natural gas. In the dehydration process, wet gas is contacted with suitable solvent/absorbent or adsorbent to remove the water from the natural gas.

3.1 Methods used for Gas dehydration:

1. Absorption using liquid desiccants

2. Adsorption using solid desiccants

3. Dehydration with $CaCl_2$.

Usually the choice of dehydration method is between liquid and solid desiccants.

Advantages of absorption using liquid desiccants

1. Lower installed cost

2. Lower pressure drop

3. Process is continuous

4. Glycol make up is easily accomplished

5. Glycol units require less regeneration heat per pound of water removed

6. Liquid desiccants dehydration equipment is simple to operate and maintain

Advantages of solid desiccants:

1. Dew points as low as -73 °C.

2. They are less affected by small changes in gas pressure, temperature and flow rate.

3. They are less susceptible to corrosion or foaming.

Economics frequently favor liquid desiccants dehydration when it will meet required dehydration specification.

Nature of liquid desiccant:

1. Very hygroscopic

2. Inexpensive

3. Non- corrosive

4. Stable when in contact with gaseous hydrocarbons

5. Capable of regeneration

6. Stable whilst being heated to strip – off water before recirculation and reuse.

7. Low moderate viscosity

8. Low vapour pressure at contact temperature.

9. Low solubility in hydrocarbon fluid

10. Low in foaming and emulsifying tendencies

Glycols are extremely stable to thermal and chemical decomposition, readily available at moderate cost, useful for continuous operation and are easy to generate. These properties make glycol as obvious choice as dehydrating agents.

Four glycols have been successfully used to dry natural gas; ethylene glycol, diethylene glycol (DEG), triethylene glycol (TEG), tetraethylene glycol (TREG). Among which TEG is used, since boiling point, regeneration temperature are high as well as decomposition temperature is also high. TEG has gained universal acceptance as the most cost effective choice. TEG is an important non-volatile industrial solvent. It is also useful in the manufacture of insecticides and in the synthesis of some organic derivatives. Pure TEG is useful in the production of plasticizers for cellophane, glue, cork, powdered ceramics and some plastics. TEG is a component in the formulation of some pigments, printing dyes, inks and pastes.

Advantage of TEG:

1. Does not solidify in concentrated solution

2. Stable in sulfur, oxygen, CO_2 at normal operating temperature

3. Highly hygroscopic

4. Carry over is small

5. High dew point depression (-10^0C to 29^0C)

6. Thermal degradation starts at 206^0C.

Disadvantage of TEG:

1. High initial cost

2. Exhibits some foaming tendency in light hydrocarbon liquids. Antifoaming agent sometimes added.

3. It is a searching liquid. It can find gaps to pass through very easily. A percentage of Glycol loss can always be attributed to this property of searching

Advantage TEG over DEG:

1. More easily regenerated to effective purity for recycling.
2. Higher temperature of decomposition
3. Vaporization loss is low
4. Capital and operating cost are lower

Properties of TEG:

Formula : $C_6H_{14}O_4$

Boling point : 288^0C

Vapour pressure at 25^0C : 0.010 mm of Hg

Density at 25^0C : $1.119\ g/cm^3$

Freezing point : -7^0C

Pour point : -58^0C

Viscosity at 25^0C : 373 cp

Flash point : $160\ ^0C$

Fire point : 166^0C

Specific gravity : 1.100

p^H : 8.8 – 9.2

Decomposition starts : 206^0C

Disadvantage of absorption using liquid desiccants.

1. Water dew points below -10 °C require stripping gas.
2. H_2S and CO_2 are present then we require a water dew point of about 100^0C.
3. Glycol is susceptible to contamination.
4. Glycol is corrosive when contaminated or decomposed.

3.2 Absorption using TEG:

Feed gas is scrubbed to remove free liquids (water and condensate) and solid particles before entering the glycol contactor. The separator may stand-alone or form an integral part of the glycol contactor vessel.

The saturated feed gas is introduced at the bottom of the contactor and rises up through the column where it encounters lean glycol, which is injected at the top of the column. On its way down, the column the glycol absorbs water and the rich glycol

collects at the bottom of the contactor. Dry gas leaves the column via a de-entrainment device, e.g. a demister mat, to remove entrained droplets. The lean glycol temperature should always be higher than the gas temperature by at least 3^0C. A higher lean glycol temperature would raise the gas dew point temperature i.e. it would increase the water content of the dry gas and increase the glycol losses in the gas stream leaving the glycol contactor. Too cold a lean glycol temperature would condense hydrocarbons in the upper section of the glycol contactor, which would create foaming inside the glycol contactor, and thereby increase glycol losses. Hydrocarbon condensate would also entrain in the glycol and would overload equipment in the glycol regenerator.

3.3 Glycol regeneration unit:

Glycol regeneration is the process of removing water from the glycol after gas dehydration.

Need for glycol regeneration are:
1. To remove water
2. To remove dissolved gases
3. To remove liquid hydrocarbon
4. To remove solid contaminants.

The rich glycol is first preheated in the reflux condenser located at the top of the glycol still (65^0C) again the glycol is heated in the glycol flash pre-heater (80^0C). Heated rich glycol flows to glycol flash drum where the low pressure (around 4 bar g) and high temperature facilitate the release of gas dissolved in the glycol. The glycol Flash drum is also where most of the liquid hydrocarbons are separated from the rich glycol stream. These liquid hydrocarbons would otherwise be vaporized in the glycol regenerator causing an overload, coking and foaming problems, resulting in an increase in glycol losses and a risk of glycol still column flooding. The liquid residence time of the glycol in the flash drum is about 30 min at operating level. The flash drum used here is bucket and weir type separator.

The rich glycol leaves the glycol flash drum and flows through glycol filter to eliminate solid particles. Since this is the closed circuit some amount solid formation is possible. Regeneration systems contain various types of filters and strainers. A particle filter or fine mesh strainer is required to protect the pump. To reduce foaming, an activated carbon filter is installed to remove heavy hydrocarbons from the glycol.

Then the glycol is heated in lean – rich heat exchanger. Lean – rich glycol heat exchanger serves for two purposes.

1. to cool the lean glycol.

2. to conserve energy by reducing the heat duty in the reboiler.

After heating, the rich glycol is send to the still. The rich glycol flows downward through a bed of random packing in the column. The water vapor is removed at the top of the still column while lean glycol is removed at the bottom. The lean glycol is stored in a surge drum. Now from the surge drum lean glycol is pumped to glycol contactor through a lean – rich heat exchanger.

3.4 Process Flow scheme:

The Process Flow Scheme is shown in fig. 3.1. The purity of the regenerated lean glycol would be 98.5% only according to this flow scheme. With this purity, we cannot achieve the dry gas specification.

The purity of the lean glycol may be improved by stripping. The idea is to reduce the partial pressure of the water vapor in the still by adding Stripping gas inside the still. Stripping gas may be allowed to trip the semi lean glycol. Stripping gas may be fuel gas or the dry natural gas itself.

Stripping gas may be sent in the reboiler or the semi lean glycol may be stripped in a separate stripper. The addition of stripping gas in reboiler. The addition of stripping gas either inside the reboiler or in the separate stripper reduces the partial pressure of the water vapor. Thereby we can achieve 99.5% glycol purity.

The three cases

1. Without stripping gas

2. With stripping gas in reboiler

3. With stripping gas in a separate stripper

3.5 PROCESS DESCRIPTION

3.5.1 Coalescer separator

Coalescer separator is a vertical separator, installed upstream of the glycol contactor to remove hydrocarbon liquid, water mist and dissolved solids from compressed natural gas stream of compressor. Natural gas enters the coalescer through the hand

Figure 3.1

valve. The level of the free liquid separated is transmitted to the level controller valves, through a level indicating controller.

The coalesce separator is equipped with the following,

1. Pressure safety valve

2. Pressure switch high

3. Level switch high

4. Level Switch low

5. Pressure differential transmitter along with pressure differential indicator

6. Pressure transmitter along with pressure indicator

7. Temperature transmitter and thermo well

3.5.2 Glycol Contactor

Glycol contactor is nothing but a absorber, installed downstream of the coalesce separator to remove water from the natural gas using glycol as a liquid desiccant. Natural gas enters the contactor at the bottom of the contactor through a fine mesh. Fine mesh is provided to remove hydrocarbon liquid, water mist and dissolved solids in case of coalesce separator fails. The lean glycol enters at the top of the contactor through a shut down valve. The dry gas is withdrawn from the top of the contactor, which passes through a fine mesh. This fine mesh is to prevent the glycol carry over. The rich glycol is withdrawn from the bottom of the contactor through a shut down valve. The level of the free liquid collected at the bottom is transmitted to the level controller valve, through a level indicating controller by level transmitter (LT), which controls the level of the free liquid.

3.5.3 Scrubber

Scrubber is a vertical separator, installed downstream of the glycol contactor to remove hydrocarbon liquid, water mist and dissolved solids from dry gas stream of contactor. Dry gas enters the scrubber through the dry gas – lean glycol heat exchanger. The pressure control valve (PCV), according to the pressure indicator in the contactor, controls the outlet stream, dry gas of the Scrubber.

3.5.4 Dry gas – lean glycol heat exchanger

Dry gas – lean glycol heat exchanger serves the transfer of heat from lean glycol to the dry gas. The heat exchanger is equipped with the following
1. Temperature transmitter with temperature indicator
2. Thermo well

3.5.5 Flash tank

The flash tank is the horizontal Three-phase Bucket and Weir separator, which separates the rich glycol stream into gas, hydrocarbon liquid and glycol. Necessary internals are provided in the flash tank for effective separation of the three phases. Gas is withdrawn from the top under pressure control valve. The separated hydrocarbon liquid from the separation compartment flows over the weir into the hydrocarbon compartment where its level is transmitted to the level controller valve,

through a level-indicating controller by LT, which controls the level. A level control valve controls the glycol level in the separation compartment.

3.5.6 Lean – rich glycol heat exchanger

The heat exchanger is equipped with the following

1. Thermo wells

3.6 Filters

A particle filter or fine mesh strainer is required to protect the pump. To reduce foaming, an activated carbon filter is installed to remove heavy hydrocarbons from the glycol.

The flash tank is equipped with the following:

1. Pressure differential transmitter (PDT) along with pressure differential indicator

3.7 Glycol recirculation pump

The glycol recirculation pump is nothing but a reciprocating pump, which discharges the lean glycol at the pressure of 88 kg / cm^2 g.

3.8 Still column and reboiler

Still column and reboiler is often called as regenerator or reconcentrator. This where the concentration of the lean glycol is increased to the requirement. The rich glycol from the heat exchanger flows inside the packing section through a level control valve is controlled by LT, which is placed in the flash tank. Reboiler is maintained at the temperature of 204^{0}C, since beyond 204^{0}C glycol degrades thermally. The heating source for the reboiler is hot oil system. The stripping gas from the scrubber through pressure control valve (PCV) gets heated in the reboiler. The Semi Lean glycol flows over the weir in the reboiler. Then the semi lean glycol is routed to the stripping column. The semi lean glycol level is controlled by level control valve LCV, which is placed in the semi lean glycol inlet to the stripping column. The rich stripping gas from the stripping column enters the reboiler.

3.9 Reflux condenser

Reflux condenser is suited on the top of the still column. The vapor coming out of the still column is condensed in this reflux condenser. The non – condensable vapor are routed to the vent header. The cooling medium used here is rich glycol from the contactor. The temperature control valve (TCV), according to the temperature in the reflux condenser, controls the flow of the rich glycol. The reflux condenser is equipped with the following,

3.10 Surge drum

The surge drum is nothing but a storage vessel. It stores the lean glycol from the stripping column. Here a gas driven pump adds fresh glycol, if required, from the glycol drum. The relived glycol from PSV is also routed into the surge drum.

3.11 Stripping column

The semi lean glycol from the reboiler enters the column at the top. According to the level in the reboiler the flow into the column is controlled. The outlet, lean glycol is send back to the surge drum. The stripping gas from the reboiler enters the column in the bottom. The rich stripping gas is then send again to the reboiler.

Chapter 4
PROCESS STEADY STATE SIMULATION

At plant level simulators are normally used to obtain flow rates, compositions and flowing medium properties at various operating conditions in the process. Typically, simulation outputs would contain mass flow rate, volume flow rate at standard or operating conditions and vapour fraction, apart from composition, density, viscosity, specific heat, thermal conductivity, surface tension, etc., of each phase in each stream. For simulations, the input source is the process design basis (PDB). The inputs themselves are typically compositions, pressures, temperatures and main flow rates of product and / or raw material, at various points in the process, for various operating conditions etc.

ASPEN-HYSYS or simply HYSYS is used as simulation software. With HYSYS, one can create rigorous steady state and dynamic models for plant design and trouble shooting. Through the completely interactive HYSYS interface, one has the ability to easily manipulate process variables and unit operation.

For doing HYSYS simulation, following steps are carried out.

1. All the components involved in the scheme are selected

2. Suitable equation of state is chosen. Generally Peng – Robinson equation favours for Hydrocarbons. [12]

3. Then we enter into the Simulation Environment.

The saturation of natural gas with water is simulated in HYSYS as follows.

1. The pressure, temperature, flow rate & composition of the natural gas is entered in one stream.

2. Another stream, water is alone taken with same condition as the natural gas, with some flow rate.

3. The two streams are mixed in a hypothetical mixer.

4. The outlet stream from the mixer is separated in the Separator.

5. The top stream is the wet gas saturate with water and the bottom stream is the free liquid.

6. The flow rate of the inlet water stream is changed until we get the constant composition in the gas outlet stream.

Thus, the natural gas is saturated with water.

4.1 Process design basis (PDB)

Process Design Basis is the input required.

The following data form the design basis.

Wet gas

Flow rate	:	2.1 MMSCMD (dry basis)
Temperature	:	54.4^0C
Pressure	:	88 kg / cm^2 g
Moisture content	:	saturated with moisture at inlet

conditions

Gas molecular weight	:	21.86 (dry basis)

Dry gas:

Water content	:	112 kg / MMSCM

Lean Glycol:

Minimum concentration	:	99.44%
Glycol loss	:	13.4 1 / MMSCM (both dehydration and regeneration)

Hot oil	:	HYTHERM – 500

Reboiler:

Temperature	:	204^0C

Composition of saturated natural gas   :

Composition	Mole %
Methane	78.64
Ethane	8.69
Propane	6.17
i – Butane	1.24
n – Butane	1.70
i – Pentane	0.40
n – Pentane	0.39
Hexane	0.43
CO_2	1.9
N_2	0.2
H_2O	0.24

Table 4.1 Saturated gas compositions from simulation at existing conditions

By simulation, following data have been taken.

Wet gas

Flow rate	:	73.8613 MMSCFD
		2.0916 MMSCMD
Temperature	:	54.4^0C
Pressure	:	88 kg / cm^2 g
Moisture content	:	3.7949 m^3 / hr
		3783.0072 kg / hr
Gas molecular weight	:	21.57

Dry gas:

Water content	:	112 kg / MMSCM

Gas Composition is shown in Table 3.1.

Calculation of TEG requirement:

Inlet gas:

Water rate	=	3783.0072 kg / day
Gas rate	=	73.8613 MMSCFD
	=	2.0916 MMSCMD
Water content	=	3783.0072 / 2.0916
	=	1808.6667 kg / MMSCM
Water content, W_{in}	=	1808.6667 mg / Sm^3

Outlet gas:

Water content, W_{out}	=	112 kg / MMSCM
	=	112 mg / Sm^3

$$\frac{W_{IN} - W_{OUT}}{W_{IN}} = \frac{1808.6667 - 112}{1808.6667}$$

$$\frac{W_{IN} - W_{OUT}}{W_{IN}} = 0.9381 \cong 0.94$$

Water to be removed = $W_{in} - W_{out}$

$$= 1808.6667 - 112$$

$$= 1696.6667 \text{ kg /MMSCM}$$

Total water to be removed = 1696.6667 x 2.0916

= 3548.7481 kg/ day

4.2 Process Flow Diagram (PFD)

The process flow diagram for the TEG dehydration unit is shown in fig 3.1

PFD is the overview diagram that describes the actual process flow in the entire process plant. It shows main equipment and critical instruments like Control valves in schematic form. They may not contain every "operation" from simulations, because some of them might have been hypothetical. PFD serve mainly to show the stream numbers (or names) of important streams in the plant. From the simulation stream tables or the material and energy balance (MEB), the data pertaining to these important streams can be correlated. Typically, MEB would contain only composition, mass flow rates and energy flow rates of incoming and outgoing streams. Often, the MEB is presented in tabular form on the PFD themselves.

Either the tabulation of the MEB data on PFD, or a separate MEB document, would be used to list, and highlight the differences between, various operating and design cases applicable to the plant. If MEB data are given only in the simulation stream tables, then the operating and design cases would be listed and explained in a write-up attached with the simulation stream tables, as a simulation report. If MEB data are presented on PFD, the write-up could be brief and placed in the notes column.

AutoCAD is a convenient tool for drafting of PFD.

For PFDs, the input source is the process description in the PDB and possibly a preliminary equipment list in which the main plant equipment items are identified with tag numbers.

Chapter 5
PROCESS DESIGN

5.1 Mass and Heat Balance

Material and Heat balance would contain appropriate data as indicated by the title, about material mass flows in and out and Energy stream flows in and out. This may be in tabular form leaving out the property data for convenience.

For Material & Heat balance, the inputs are the stream numbers and stream data from simulations, as well as the selected stream numbers from the PFDs.

5.2 Equipment sizing

The major equipments involved in the Dehydration unit are

1. Absorption Column

2. Flash Tank (Three – phase separator)

3. Stripping Column

4. Still Column

The equipment Size has been carried out for the above-mentioned Equipments.

5.2.1 Design of Absorption column

Packed Bed Tower, [1, 3]

Symbols Used:

V - Mass flow rate Vapor, kg / hr

L - Mass flow rate Liquid, kg / hr

ρ_v - Density of the Vapor, kg / m^3 or lb / ft^3

ρ_l - Density of the Liquid, kg / m^3 or lb / ft^3

γ - Kinematic Viscosity of the liquid , cSt

γ_A – Activity coefficient , From figure 5.1

K – Equilibrium constant for water in a TEG – water system.

E_a - Absorption Efficiency

A_T – Absorption factor

Y_{n+1} – flow rate of water in the entering rich gas , kg / hr

Y_1 – flow rate of water in the dry gas , kg / hr

n – Number of stages

ΔP – pressure drop, in of H_2O / ft of packing

G – gas mass velocity, kg / m^2 sec or lb / ft^2 sec

F_P – packing factor

D – diameter of the Tower, in or m

W – Water content on mass/vol basis, at saturation,as found from regular water

     content correlation

B – Water content on mass/ vol basis : $=1.314 \times 10^{-6}$ when W is in kg/10^6 m^3

                                      $=2.109 \times 10^{-5}$ when w is in lbm/MMscf

Activity Coefficient for H_2O Concentration at Various Temperatures[[8]]

**Figure 5.1**

Input Data

Mass flow rate Vapor, V = 79457.41 kg / hr

Density of the Vapor, ρ_v = 87.34 kg / m^3 = 5.452 lb / ft^3

Density of the Liquid, ρ_l = 1107.84 kg / m^3 = 69.16 lb / ft^3
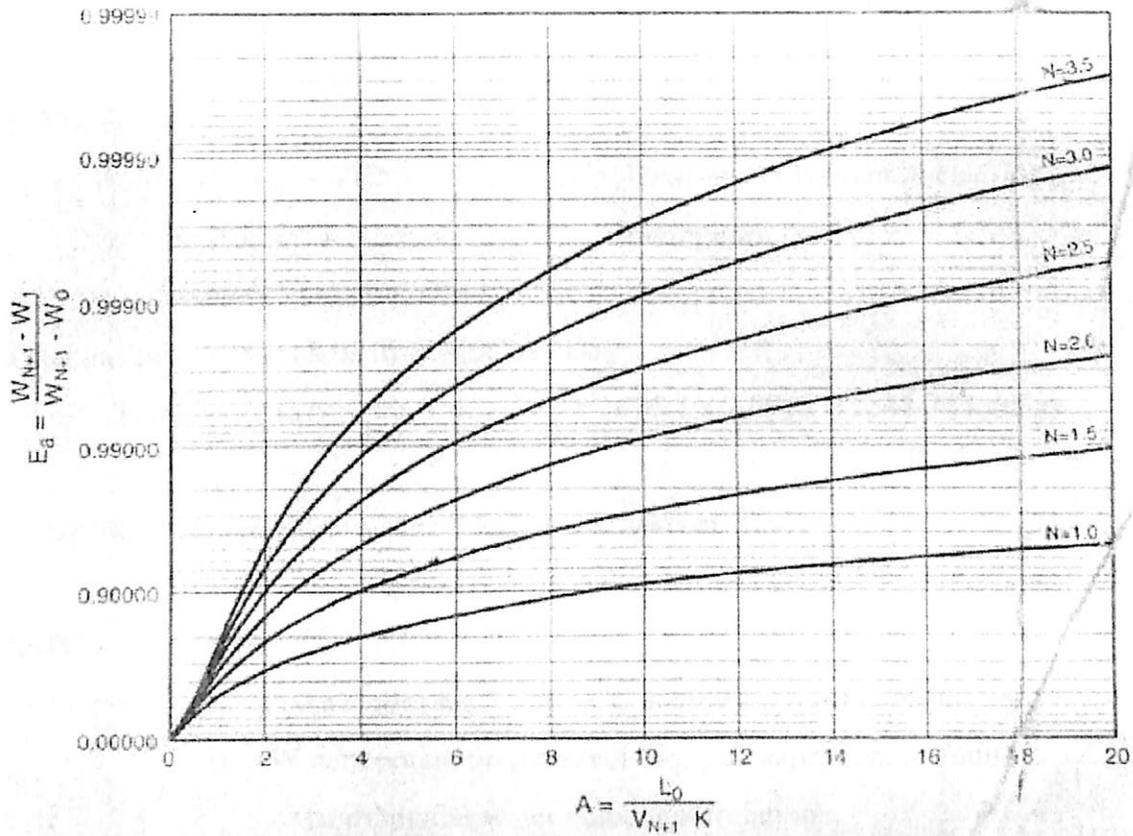
Efficiency of Absorption, E_a vs. Absorption Factor, A

Figure 5.2
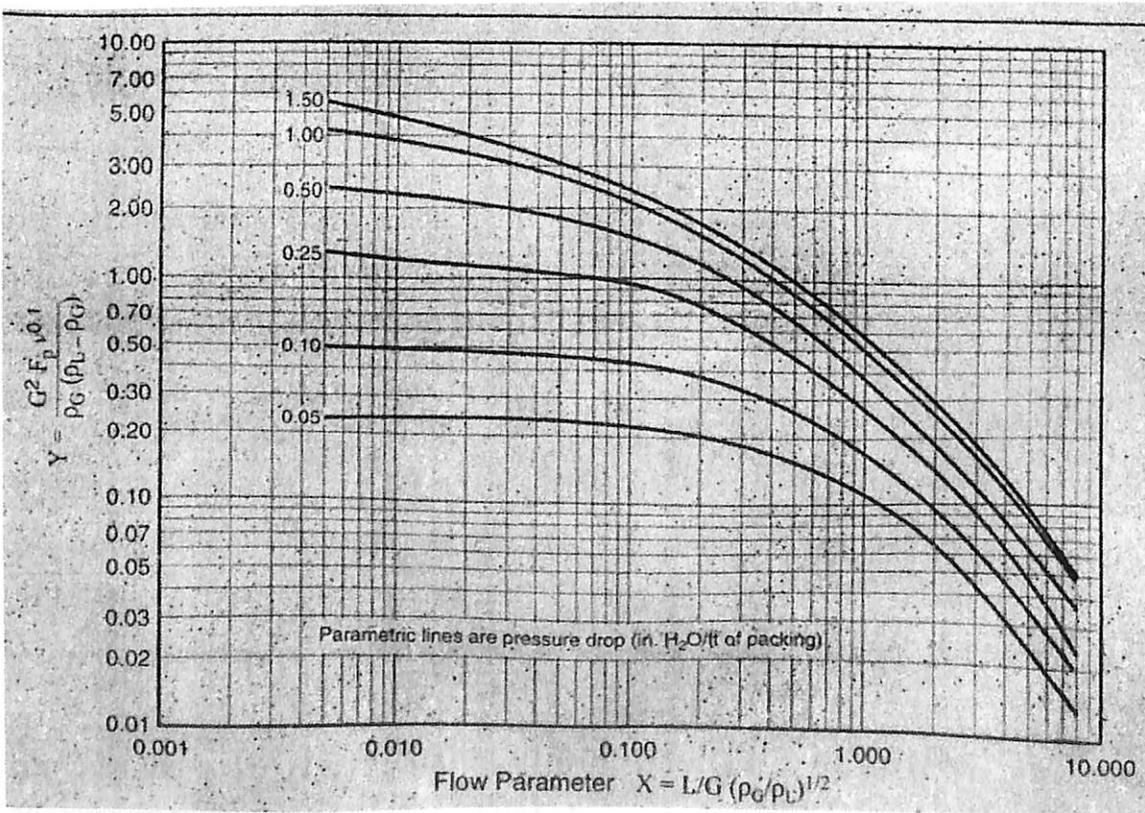
Figure 5.3

Kinematic Viscosity of the liquid, $\gamma = 9.509$ cSt

Dry gas Specification = 112kg/ MMSCM

1. Water recovery

$Win - Wout = 1808.67 - 112$ Where, Win, $Wout$ -water content in entering and

leaving gas.

$= 1696.7$ kg/MMSCM

Total gas processed is 2.0916 MMSCM / Day

⇨   Total water to be recovered / Day = 1676.7 x 2.0916 = 3548.74 kg/Day

2. Equilibrium relation (K) for TEG-water system

$K = B.W. \gamma_A$,   where

γ_A is taken from fig 5.1 at 58^0C at 99.9% glycol concentration

W – Water content on mass/vol basis, at saturation, as found

from regular water content correlation

B – Water content on mass/ vol. basis:

$= 1.314 \times 10^{-6}$ when W is in $kg/10^6 \ m^3$

$= 2.109 \times 10^{-5}$ when w is in lbm/MMscf

$= 1314 \times 10^{-6} \times 1808.67 \times 0.61$

$= 1.467 \times 10^{-3}$

3. Absorption factor and absorption efficiency

$Ea = \dfrac{W_{N+1} - W_1}{W_{N+1} - W_0}$,

Where,

W_{N+1} – Water content of entering wet gas

W_1 – Water content of leaving dry gas

W_0 – Water content of gas in equilibrium with lean TEG

X_0 = mol fraction of H_2O = $[(100-X_{gl})/18] / [((100-X_{gl})/18) + (X_{gl}/150)]$

$X_{gl} = 99.96$ Wt%   => $X_0 = 0.003323$

$W_0 = W \, \gamma_A \, X_0$

$= 1808.67 \times 0.61 \times 0.003323$

$= 3.6$

Therefore $Ea = [1808.67 - 112] / [1808.67 - 3.6]$

$Ea = 0.94$

From figure 5.2

At, $Ea = 0.94$ and $N = 1.5$, Number of theoretical contactors in the absorber

(Normally Assumed for TEG- gas dehydration)

$A = 5.5$

4. Lean glycol circulation rate

$L_0 = A \, K \, V_{N+1}$

$= 5.5 \times 1.426 \times 10^{-3} \times (79457/21.57)$

$= \mathbf{28.89}$ kmol/hr

Lean Glycol average molecular weight: $= 0.9996 \times 150 + 0.0004 \times 18$

$= 149.94$

Kg Lean Glycol / hr $= 28.89 \times 149.94 = 4332$ kg/hr

5. HETP:

For glycol dehydration, a HETP of 1.5-2.0 m can be used to estimate the height for both random and structured packing.

$$E_a = \frac{Y_{n+1} - Y_1}{Y_{n+1}}$$

$$E_a = \frac{157.80 - 1.95}{157.80} = 0.98764$$

$$A_T = \frac{L}{KV} = 4\frac{4094.21}{2.379x10^{-3}x79457.41}$$

$$A_T = 21.6591$$

$$E_a = \left[\frac{A_T^{n+1} - A_T}{A_T^{n+1} - 1}\right]$$

Where, n- theoretical No. of stages. Putting n=1, 2, we get **Ea** as fallows.

n = 1	Ea = 0.9559
n = 2	Ea = 0.998

Theoretical No. of stages = 2

Taking HETP of 2 m ;[1], (for Glycol dehydration, an HETP of 1.5 – 2.0 m can be

Used to estimate contactor height both for random and

structured packing).

Packing Height = n * HETP = 2 * 1.5 = 3 m

At Pressure drop = 0.25 in of H_2O / ft   [1], page 323.

6. Diameter:

From figure 5.3

X-axis

$$\frac{L}{V}\sqrt{\frac{\rho_V}{\rho_L}} = \frac{4332}{79457.41}\sqrt{\frac{87.34}{1107.84}} = 0.0153$$

Y-axis   G – gas mass velocity, kg / m^2 sec or lb / ft^2 sec

V - Mass flow rate Vapor, kg / hr

$$\frac{G^2 F_p \gamma^{0.1}}{\rho_V(\rho_L - \rho_V)} = 1.3$$

Packing - Pall Rings (Plastic) - 2" size.

F_P = 26 for Pall Rings – 2"   [1], page 326.

$$\frac{G^2 x26x(9.509)^{0.1}}{5.453(69.16-5.452)} = 1.3$$

$$G^2 = 13.8645$$

$$G = 3.7235 \frac{lb}{ft^2.sec}$$

$$\boxed{G = 17.8874 \text{ kg} / \text{m}^2}$$

$$Area = \frac{V}{G} = \frac{79457.41}{17.8874 * 3600} = 1.233m^2$$

D = 1.251 m ≈ 50 in

Following are the additional parameters to be added to find the total height of contactor for the corresponding diameter calculated.

1.	Packing height = 3 m		= 118.1088 in
2.	Tower Cover		= 18 in
3.	Packing Support (Chemical Porcelain)	= 8.5 + 8.5	= 17 in
4.	Liquid distributor (Weir Flow)		= 48 in
5.	Space between Packing and packing support = 15 in x 2		= 30 in
6.	Redistribution height		= 4 in
7.	Liquid hold Down plate	= 4 in x 3	= 12 in
8.	Stacked Layers of large & intermediate Sized Packing		= 10 in
9.	Liquid clearance		= 12 in
			= 269.1088 in

Tower height   = 6.8354 m

Tower diameter   = 1.251 m

5.2.2 Flash Tank:

Bucket and weir type 3 phase separator (Horizontal), [6]

Gas – Condensate (Hydro Carbon) – Rich Glycol

Symbols used:

Q_V, - Volumetric flow rate of gas, ft^3 / hr

Q_{LL} - Volumetric flow rate of Light Liquid, ft^3 / hr

Q_{HL} - Volumetric flow rate of Heavy Liquid, ft^3 / hr

ρ_v - Density of Gas, lb / ft^3

ρ_L - Density of Light Liquid, lb / ft^3

ρ_H - Density of Gas, lb / ft^3

ρ_w - Density of Water, lb / ft^3

μ_L - Viscosity of the Light Liquid, cp

θ_{LL}, θ_{HH},- Residence time for Light and Heavy liquids, min

T_H, T_S - Hold up Time and Surge time, min

P – Pressure in Psi a

K – Terminal velocity constant, ft / sec

U_T – Terminal Velocity , ft / sec

U_v – Vapor Velocity, ft / sec

D – diameter of the vessel, ft or in

L – length of the vessel, ft

A_T - total cross sectional area, ft^2

H – height, ht or in

H_V - vapor space area, ft

L_1 - Length upto first bucket, ft

L_{min} - Vapor / liquid separation minimum length, ft

U_{VA} - actual vapor velocity, ft / sec

ϕ - liquid dropout time, sec

H_{LL} - light layer thickness, in

ΔS_G – Specific Gravity between light and heavy liquids

D_P – Droplet size , microns

ΔH - Height difference between light and heavy liquid weirs, ft

L_2 – length of the first bucket, ft

L_3 – length between two buckets, ft

L_4 – length of the second bucket, ft

Subscripts

H, HLL – heavy liquids

HLL – high liquid level

L , LL – light liquid

LLL – low level liquid

V – vapor

Input data:

Volumetric flow rate of gas, $Q_v = 331.7$ ft^3 / hr

Volumetric flow rate of Light Liquid, $Q_{LL} = 10$ ft^3 / hr

Volumetric flow rate of Heavy Liquid, $Q_v = 139.6$ ft^3 / hr

Density of Gas, $\rho_v = 0.142$ lb / ft^3

Density of Light Liquid, $\rho_L = 50.759$ lb / ft^3

Density of Gas, $\rho_H = 67.23$ lb / ft^3

Density of Water, $\rho_v = 60.54$ lb / ft^3

Viscosity of the Light Liquid, $\mu_L = 2$ cp

Residence time for Light and Heavy liquids, $\theta_{LL} = \theta_{HH} = 10$ min

Hold up Time + Surge time = $T_H + T_S = 10$ min

Droplet size, $D_P = 150$ microns

Calculation:

1. Vapor velocity:

$$U = k\sqrt{\frac{\rho_L - \rho_v}{\rho_v}}$$

$$K = 0.35 + .0001(P - 100)$$

P – Pressure in Psi a

P = 3.5 kg / cm^2 g = 64.4771 Psi a

K = 0.3464 ft / sec

For glycol or amine solution multiply K by 0.6 – 0.8

Assume 0.7

K = 0.3464 * 0.7 = 0.2425

$$U_T = 0.2425\sqrt{\frac{50.759 - 0.1412}{0.1412}}$$

$= 4.5765$ ft / sec

Vapor velocity, $U_V = 0.75 * 4.5765 = 3.4324$ ft / sec

2. Diamete | $U_V = 3.4324$ ft / sec |

$$D = \left[\frac{4(\vartheta_{LL}Q_{LL} - \vartheta_{HL}Q_{HL})}{0.7\pi\left(L/D\right)}\right]^{\frac{1}{3}}$$

$$L/D = 1.5$$

$$D = \left[\frac{4(10*10 - 139.6*10)}{0.7\pi*1.5}\right]^{\frac{1}{3}}$$

| $D = 3.1153$ ft |

$$A_T = \frac{\pi}{4}D^2 = \frac{\pi}{4}*3.1153^2$$

| $A_T = 7.6224$ ft^2 |

3. Vapor area, A_V

$$\frac{Hv}{D} = 0.2$$

$$Hv = 0.2xD = 0.2x3.1153 = 0.6231ft$$

| $H_V = 0.6231$ ft |

$$y = \frac{a + cx + ex^2 + gx^3 + ix^4}{1.0 + bx + dx^2 + fx^3 + hx^4}$$

Where, $y = Av / A_T, x = Av / D$

$a = -4.75593x10^{-5}, b = 3.924091, c = 0.174875$

$d = -6.358805, e = 5.668973, f = 4.018448$

$g = -4.91641, h = -1.801705, i = -0.145348$

$$\frac{Av}{A_T} = 0.1425, \quad A_V = 0.1425x7.6224$$

$A_v = 1.0862 \text{ ft}^2$

4. Length upto first bucket , L_1

$$L_1 = \frac{\left(Q_{HL} x \theta_{HL} + Q_{LL} \theta_{LL}\right)}{A_T - A_V}$$

$$= \frac{139.6 x 10 + 10 x 10}{7.6224 - 1.0862} = 3.8147 ft$$

$$\boxed{L_1 = 3.8147 \text{ ft}}$$

5. Vapour / Liquid separation minimum length , Lmin

$$Q = \frac{Hv}{Uv} = \frac{0.6231}{3.4324} = 0.1815 \sec$$

Actual vapour velocity, $\qquad UvA = \dfrac{Qv}{Av} = \dfrac{331.7}{1.0862} = 305.3765 \dfrac{ft}{hr}$

$UvA = 0.0848 \; ft/\sec$

$L\min = \phi x UvA = 0.1815 x 0.0848$

$L\min = 0.0154 ft$

$L_1 > L\min,$

6. Height difference between light and heavy liquid weirs, ΔH

$$H_{LL} = \frac{0.00128 x \theta_{LL} x \Delta S_G x D_P{}^2}{\mu_L}$$

$$\Delta S_G = \frac{(67.23 - 50.729)}{60.54} = 0.2721$$

$$H_{LL} = \frac{0.00128 x 10 x 0.2721 x 150^2}{2}$$

$H_{LL} = 39.1824 \text{ in}$

$$\boxed{H_{LL} = 3.2652 \text{ ft}}$$

$$\Delta H = H_{LL}\left(1 - \frac{\rho_L}{\rho_H}\right)$$

$$\Delta H = 3.2652\left(1 - \frac{50.759}{67.23}\right)$$

$$\Delta H = 0.8\,ft$$

7. Dimension of first bucket

Top of the bucket = D-H$_V$

$$= 3.1153-0.6231$$

$$= 2.4922\ \text{ft}$$

Bottom of the Bucket = 0.125D = 0.125x3.1153

$$= 0.3894\ \text{ft}$$

HLL = 2.4922-0.5 = 1.9922 ft

LLL = 0.3894+0.5 = 0.8894 ft

$$\frac{H_{HLL}}{D} = \frac{1.9922}{3.1153} = 0.6395$$

$$\frac{H_{LLL}}{D} = \frac{0.8894}{3.1153} = 0.2855$$

$$\frac{A_{HLL}}{A_T} = 0.6802,$$

$$\frac{A_{LLL}}{A_T} = 0.2359$$

$$A_{HLL} = 0.6802x7.6224 = 5.1848\,ft^2$$

$$A_{LLL} = 0.2359x7.6224 = 1.7981\,ft^2$$

$$L_2 = \left(\frac{(T_H + T_S)xQ_{LL}}{A_{HLL} - A_{LLL}}\right) = \frac{10x10}{60(5.1848 - 1.7981)}$$

T_H+T_S=10 min

8. Length between two buckets, L$_3$

$$L_3 = D/12 = \frac{3.1153}{12} = 0.2596\,ft$$

9. Dimension of second Bucket

Top of the Bucket $= D - H_v - \Delta H$

$$= 3.1153 - 0.6231 - 0.8$$
$$= 1.6922 ft$$

HLL = 1.6922-0.5 = 1.1922 ft

LLL = 0.5 ft

$$\frac{H_{HLL}}{D} = \frac{1.1922}{3.1153} = 0.3827$$

$$\frac{H_{LLL}}{D} = \frac{0.5}{3.1153} = 0.1605$$

$$\frac{A_{HLL}}{A_T} = 0.3529,$$

$$\frac{A_{LLL}}{A_T} = 0.1038$$

$$A_{HLL} = 0.3529 x 7.6224 = 2.6899 ft^2$$

$$A_{LLL} = 0.1038 x 7.6224 = 0.7912 ft^2$$

$$L_4 = \left(\frac{(T_H + T_S) x Q_{HL}}{A_{HLL} - A_{LLL}} \right) = \frac{10 x 139.6}{60(2.6899 - 0.7912)}$$

$T_H + T_S = 10$ min $\qquad \boxed{L_4 = 12.254 ft}$

10. Total Length

Total Length L = L1+L2+L3+L4

$$= 3.8147+0.4921+0.2596+12.254$$
$$= 16.8204 \text{ ft}$$

$$\frac{L}{D} = \frac{16.8204}{3.1153} = 5.3993$$

which is within the limit, but the Length of the separation is too large hence diameter of the Vessel is increased to 4 ft and the Length of the Vessel is recalculated.

$$\boxed{D = 4.00 \text{ ft}}$$

$$A_T = \frac{\pi}{4} x 4^2 = 12.5664 ft^2$$

1. Vapor area, A_V

$$\frac{H_v}{D} = 0.2 \quad , \quad Hv = 0.2 \times 4 = 0.8$$

$$\frac{Av}{A_T} = 0.1425,$$

$A_v = 0.1425 \times 7.6224 = 1.7907 \, ft^2$

2. Length upto first bucket , L_1

$$L_1 = \frac{10x10 + 139.6x10}{60(12.5664 - 1.7907)} = 2.3138 \, ft$$

3. Vapour / liquid separation minimum length , Lmin

$$Q = \frac{Hv}{Uv} = \frac{0.8}{3.4324} = 0.2331 \sec$$

$$UvA = \frac{Qv}{Av} = \frac{331.7}{1.7907x3600} = 0.0515 \, ft\!\!\bigg/\!\!_{\sec}$$

$L \min = \phi x U_{vA} = 0.0515x0.2331$

$L \min = 0.012 \, ft$

$L_1 > L \min, Acceptable$

4. Height difference between light and heavy liquid weirs, ΔH

DH=0.8 ft

5. Dimension of first bucket

Top of the bucket = D-H_V

= 4.0 - 0.8

= 3.2 ft

Bottom of the Bucket = 0.125D = 0.125x4

= 0.5 ft

HLL = 3.2-0.5 = 2.7 ft

LLL = 0.5+0.5 = 1.00 ft

$$\frac{H_{HLL}}{D} = \frac{2.7}{4} = 0.675$$

$$\frac{H_{LLL}}{D} = \frac{1}{4} = 0.25$$

$$\frac{A_{HLL}}{A_T} = 0.7242,$$

$$\frac{A_{LLL}}{A_T} = 0.1958$$

$$A_{HLL} = 0.7242 x 12.5664 = 9.1006\, ft^2$$

$$A_{LLL} = 0.1958 x 12.5664 = 2.4605\, ft^2$$

$$L_2 = \left(\frac{(T_H + T_S) x Q_{LL}}{A_{HLL} - A_{LLL}} \right) = \frac{10 x 10}{60(9.1006 - 2.4605)}$$

$T_H + T_S = 10$ min $\boxed{L_2 = 0.251 ft}$

6. Length between two buckets, L₃

$$L_3 = D/12 = \frac{4}{12} = 0.3333\, ft$$

5. Dimension of second bucket

Top of the Bucket $= D - H_v - \Delta H$

$$= 4.0 - 0.8 - 0.8$$

$$= 2.4\, ft$$

HLL = 2.4-0.5 = 1.9 ft

LLL = 0.5 ft

$$\frac{H_{HLL}}{D} = \frac{1.9}{4} = 0.475$$

$$\frac{H_{LLL}}{D} = \frac{0.5}{4} = 0.125$$

$$\frac{A_{HLL}}{A_T} = 0.47,$$

$$\frac{A_{LLL}}{A_T} = 0.0721$$

$$A_{HLL} = 0.47 x 12.5664 = 5.9062\, ft^2$$

$$A_{LLL} = 0.0721 x 12.5664 = 0.906\, ft^2$$

$$L_4 = \left(\frac{(T_H + T_S) x Q_{HL}}{A_{HLL} - A_{LLL}} \right) = \frac{10 x 139.6}{60(5.9062 - 0.906)}$$

$T_H + T_S = 10$ min $\boxed{L_4 = 4.6528 ft}$

8. Total length

Total Length L = L1+L2+L3+L4

$\qquad\qquad = 2.3138 + 0.251 + 0.3333 + 4.6528$

$\qquad\qquad = 7.5509$ ft

$$\frac{L}{D} = \frac{7.5509}{4.00} = 1.8877 > 1.5$$

Result:

Length of the vessel   **= 2.3014 m**

Diameter of the vessel   **= 1.2192 m**

The Vessel is shown pictorially in figure 5.4
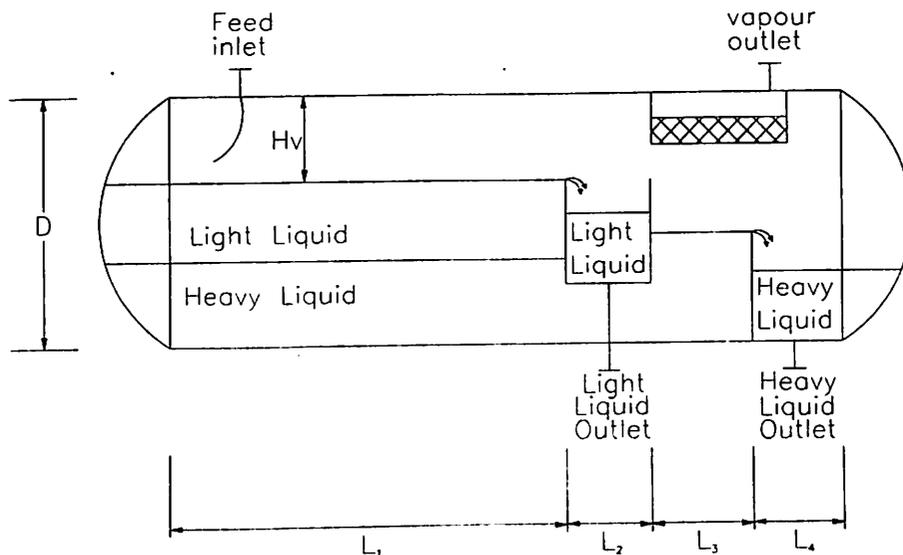
Figure 5.4 Pictorial representation of three-phase separator

5.2.3. Design of stripping column

Packed bed tower, [4, 1]

Symbols used:

V - Mass flow rate vapor, kg / hr

L - Mass flow rate liquid, kg / hr

ρ_v - Density of the vapor, kg / m^3 or lb / ft^3

ρ_l - Density of the liquid, kg / m^3 or lb / ft^3

γ - Kinematic viscosity of the liquid , cSt

K – Equilibrium constant for water in a TEG – water system.

E_a - Stripping efficiency

S_T – Stripping factor

ΔP – pressure drop, in of H_2O / ft of packing

Y_{n+1} – flow rate of water in the entering rich gas , kg / hr

Y_1 – flow rate of water in the dry gas , kg / hr

n – Number of stages

G – gas mass velocity, kg / m² sec or lb / ft² sec

F_P – packing factor

D – diameter of the tower, in or m

Input Data:

Mass flow rate vapor, V = 510 kg / hr

Mass flow rate liquid, L = 4229.499 kg / hr

Density of the vapor, ρ_v = 3.472E-02 lb / ft³

Density of the liquid, ρ_l = 62.21 lb / ft³

Kinematic viscosity of the liquid , γ = 0.939 cSt

K = 33.33 (from HYSYS)

$$E_a = \frac{Y_{n+1} - Y_1}{Y_{n+1}}$$

$$E_a = \frac{21.20324 - 0.15}{21.20324} = 0.9929$$

$$S_T = \frac{KV}{L} = \frac{33.33x510}{4229.499}$$

$$S_T = 4.019$$

From GPSA, No. of theoretical stages = 3

HETP = 0.75 m , Packing Height = n * HETP = 3 * 0.75 = 2.25 m

$$\frac{L}{V}\sqrt{\frac{Pv}{P_L}} = \frac{4229.499}{510}\sqrt{\frac{3.472x10^{-2}}{62.21}} = 0.1959$$

ΔP = 0.25 in of H_2O / ft

Packing - Pall Rings (Plastic) - 2" size.

F_P = 26 for Pall Rings – 2"   [3], Page 323.

From Figure,

$$\frac{G^2 F_p \gamma^{0.1}}{P_V (P_L - P_V)} = 0.8.$$

$$\frac{G^2 x 26 x (0.9395)^{0.1}}{3.472 * 10^{-2} (62.21 - 3.472 * 10^{-2})} = 0.8$$

$$G^2 = 0.0668$$

$$G = 0.2585 \frac{lb}{ft^2.sec}$$

$$\boxed{G = 1.2416 \frac{Kg}{m^2.sec}}$$

$$Area = \frac{V}{G} = \frac{510}{1.2416 x 3600} = 0.114 m^2$$

$$D = 381 \text{ mm} \approx 15 \text{ in}$$

Results:

1. Packing height   = 2.25 m   = 118.1088 in
2. Tower cover   = 10 in
3. Packing Support (chemical porcelain)   = 6.875 in
4. Liquid distributor (weir flow)   = 13 in
5. Space between packing and packing support = 10 in x 2   = 20 in
6. Redistribution height   = 4 in
7. Liquid hold down plate   = 4 in x 2   = 8 in
8. Stacked layers of large & intermediate sized packing   = 10 in
9. Liquid clearance   = 12 in

                                                = 172.4566 in

Tower height   **= 4.3805 m**

Tower diameter   **= 381 mm**

5.2.4. Design of still column

Packed bed tower,[13]

Symbols used:

V - Mass flow rate vapor, kg / hr

L - Mass flow rate liquid, kg / hr

ρ_v - Density of the vapor, kg / m^3 or lb / ft^3

ρ_l - Density of the liquid, kg / m^3 or lb / ft^3

γ - Kinematic viscosity of the liquid , cSt

K – Equilibrium constant for water in a TEG – water system.

ΔP – pressure drop, in of H$_2$O / ft of packing

P_A, P_B - vapor pressure of water and TEG, mm of Hg

X_{LD}, X_{LB} , X_{HD} & X_{HB} - composition of more and less volatile component in distillate and in residue

α_1, α_2 & α_{AVG}– Relative volatility at top, bottom and average of the column.

S_M - Minimum number of stages

G – gas mass velocity, kg / m^2 sec or lb / ft^2 sec

F_P – packing factor

D – diameter of the tower, in or m

Input Data:-

Feed Rate, F = 79.33 lb mol/hr + 56.39 lb mol/hr

Vapor rate, V = 71.31 lb mol/hr = 669.0 kg/hr

Liquid rate,L = 64.41 lb mol/hr = 4229 kg/hr

Composition of more and less volatile component in distillate and in residue

X_{LD} = 0.2664, X_{LB} = 0.0403

X_{HD} = 0.0001, X_{HB} = 0.9590

Vapour pressure of water at 66.7°C = 201.6 mmHg

Vapour pressure of water at 204°C = 12570 mmHg

Vapour pressure of TEG at 66.7°C = 0.1 mmHg

Vapour pressure of TEG at 204°C = 23 mmHg

Density of the vapor, ρ_v = 0.7441 kg / m^3 = 0.0472 lb / ft^3

Density of the liquid, ρ_l = 964.1 kg / m^3 = 61.1712 lb / ft^3

Kinematic viscosity of the liquid, γ = 0.5951 cSt

$$\alpha_1 = \frac{P_A}{P_B} = \frac{201.6}{0.1} = 2016$$

$$\alpha_2 = \frac{P_A}{P_B} = \frac{12570}{85.23} = 147.4832$$

$$\alpha_{AVG} = \sqrt{\alpha_1 \alpha_2} = \sqrt{2016 x 147.4832}$$

$$\alpha_{AVG} = 545.276$$

$$S_M = \frac{\log\left[\left(\dfrac{X_{LD}}{X_{HD}}\right)\left(\dfrac{X_{HB}}{X_{LB}}\right)\right]}{\log \alpha_{avg}}$$

$$S_M = \frac{\log\left[\left(\dfrac{0.2664}{0.0001}\right)\left(\dfrac{0.9590}{0.0403}\right)\right]}{\log 545.276}$$

$$\boxed{S_M = 1.75}$$

HETP = 0.75 m , Packing Height = S_M * HETP = 1.75 * 0.75 = 1.3125 m

ΔP = 0.25 in of H_2O / ft

Packing - Pall Rings (Plastic) - 2" size.

F_P = 26 for Pall Rings – 2"

$$X = \frac{L}{V}\left(\frac{P_Y}{P_L}\right)^{0.5}$$

$$\gamma = \frac{G^2 F_P \gamma^{0.1}}{P_Y\left(P_L - P_Y\right)}$$

$$X = \frac{4229}{669}\left(\frac{0.0472}{61.1712}\right)^{0.5} = 0.1756$$

$$\frac{G^2 F_P \gamma^{0.1}}{P_V\left(P_L - P_V\right)} = 0.23$$

$$\frac{G^2 x26x(0.595)^{0.1}}{0.0472(61.1712 - 0.0472)} = 0.8$$

$$G^2 x8.5562 = 0.8$$

$$G = 0.3058\frac{lb}{ft^2 . \sec}$$

$$G = 1.469\frac{Kg}{m^2 . \sec}$$

$$Area = \frac{V}{G} = \frac{669}{3600 x 1.469} = 0.1265 m^2$$

$$D = 0.401m \approx 16 \text{ in}$$

Results:

1.	Packing height	= 1.3125 m	= 51.6726 in
2.	Tower cover		= 10 in
3.	Packing Support (chemical porcelain)		= 7 in
4.	Liquid distributor (weir flow)		= 14 in
5.	Space between packing and packing support = 10 in x 2		= 20 in
6.	Redistribution height		= 4 in
7.	Liquid hold down plate	= 4 in x 2	= 8 in
8.	Stacked layers of large & intermediate sized packing		= 10 in
9.	Liquid clearance		= 12 in
			= 136.6726 in
	Tower height		**= 3.4715 m**
	Tower diameter		**= 0.401 m**

5.3. Line sizing

The Process has too many lines. The Line Sizing is carried out only for the process lines. The Line Sizing procedures for different phase are discussed below.

5.3.1. Liquid line sizing

The flowing step by step procedure where carried for sizing Liquid lines.

1. Calculate the diameter of the pipe $d = \sqrt{\dfrac{0.012Q}{V}}$

2. Calculate Moody friction factor f, f is the function of Reynolds number. From the graph shown in [10], Figure 2.3 , page no - 19

$$Re = \frac{\rho * d * V}{12\mu}$$

3. Calculate the pressure drop in the pipe $\Delta P = \dfrac{0.00115 f Q^2 S}{d^5}$

4. Calculate the Testing pressure of the pipe

Testing pressure = 1.5 * Design Pressure

Design pressure = 1.1 * Operating pressure

5. Find the Pipe Class and rating

6. Find the Nominal diameter and Schedule number of the pipe.

Where

d - Pipe inside diameter, in

Q - Liquid flow rate, Barrels / day

V - average flow velocity, ft / sec

ΔP - pressure drop , psi / 100 ft

f - Moody Friction factor

S - Specific gravity

Re – Reynolds Number

ρ – Liquid Density, lb / ft^3

μ - Liquid Viscosity, lb / ft – sec

5.3.2. Gas line sizing

The flowing step-by-step procedure where carried for sizing Gas lines.

1. Calculate the diameter of the pipe $d = 12 * \sqrt{\dfrac{4*Q}{\pi*3600*V}}$

2. Calculate Moody friction factor f, f is the function of Reynolds number. From the graph shown in [10] Figure 2.3 , page no - 19

$Re = \dfrac{\rho*d*V}{12\mu}$

3. Calculate the pressure drop in the pipe $\Delta P = \dfrac{12.6*Z*T_1 fQ^2 SL}{P_1 d^5}$

4. Calculate the Testing pressure of the pipe

Testing pressure = 1.5 * Design Pressure

Design pressure = 1.1 * Operating pressure

5. Find the Pipe Class and rating

6. Find the Nominal diameter and Schedule number of the pipe.

Where

d - Pipe inside diameter, in

Q - Gas flow rate, ft^3 / hr

Q_l – Gas flow rate, MMSCFD

V - average flow velocity, ft / sec

ΔP - pressure drop, psi

P_l – inlet pressure, psi

T_l – Inlet temperature, deg R

Z – Compressibility factor

L – Length , ft

f - Moody Friction factor

S - Specific gravity

Re – Reynolds Number

ρ – Gas Density, lb / ft^3

μ - Liquid Viscosity, lb / ft – sec

5.3.3. Two phase line sizing

The flowing step by step procedure where carried for sizing Two phase lines.

1. Calculate the Fluid erosion velocity, $V_e = \dfrac{c}{\sqrt{\rho m}}$

2. Calculate the pipe inside diameter, $d = \sqrt{\dfrac{4*A}{\pi}}$

$$A = \dfrac{9.35 + \dfrac{ZRT}{21.25P}}{Ve}$$

3. Calculate Moody friction factor f, f is the function of Reynolds number. From the graph [10], Figure 2.3, page no - 19

$\text{Re} = \dfrac{\rho m * d * V}{12\mu}$

3. Calculate the pressure drop in the pipe $\Delta P = \dfrac{0.00036 f W^2}{d^5 \rho m}$

4. Calculate the Testing pressure of the pipe

    Testing pressure = 1.5 * Design Pressure

    Design pressure = 1.1 * Operating pressure

5. Find the Pipe Class and rating

6. Find the Nominal diameter and Schedule number of the pipe.

Where

d - Pipe inside diameter, in

V_e – errosional velocity, ft / sec

W – Total Liquid plus Vapor rate, lb / hr

ΔP - pressure drop, psi / 100 ft

f - Moody Friction factor

Re – Reynolds Number

ρm – Mixture Density, lb / ft^3

μ - Liquid Viscosity, lb / ft – sec

5.4 Piping and instrumentation diagram

Piping and instrumentation diagram (P&ID) are made for process plants. P&ID is a schematic representation of the plant, showing the following.

1. Legend sheet which shows the numbering system adopted for equipment, lines, instruments etc, symbols used, piping class used etc

2. Interconnections between equipment, platform etc., by means of piping, fittings, conveyor etc.

3. Instrumentation and controls such as actuators for control and safety valves, etc

4. General and special process requirements such as insulation type, thickness, slope requirements, etc

PFD is the base document for generating P&ID and other inputs for P&ID are listed below

1. Process design basis

2. Process description, operating and control philosophy

3. Set of symbols and notations (legends)

4. Sizing calculations (Line/Equipment)

Chapter 6

RESULTS AND DISCUSSION

RESULTS

For individual units, major parameters have been manually calculated and compared with the design values, the percentage deviations is tabulated as shown below.

1. Absorption column

Parameter	Calculated value	Design value	% Deviation
Lean Glycol Circulation, kg/hr	4332	4094	+5.49
Water Recovery, kg/day	3548	3548	0.0
No. of Stages	3	3	0.0
Column Height, mm	6835.4	6841	-0.08
Column Diameter, mm	1251	1253	-0.16

Table 6.1

2. Stripping Column

Parameter	Calculated value	Design value	% Deviation
Column Height, mm	3810	3800	-0.06
Column Diameter, mm	4380.5	4500	-2.72

Table 6.2

3. Three Phase Separator

Parameter	Calculated value	Design value	% Deviation
Vessel Length, mm	1219.2	1220	-0.06
Vessel Diameter, mm	2340.4	2340	0.02

Table 6.3

4. Still Column

Parameter	Calculated value	Design value	% Deviation
Column Height, mm	401	400	+0.25
Column Diameter, mm	3471.5	3570	-2.8

Table 6.4

DISCUSSION

The major variation notable out of results is the lean glycol circulation, which might have resulted out of manual errors in reading graphs. There can still be improvement in the results of lean glycol requirement if suitable assumptions are made.

Column diameter is calculated and based on diameter column height is calculated suitably referring to the normal design practices. Hence, this may vary from one original equipment manufacturer (OEM) to other.

Chapter 7

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSION

Process design of gas-TEG contact column and other calculated results found are in good agreement with design values. A good match of calculated results with the actual industrial data validates the empirical methods used.

RECOMMENDATIONS

A converged Hysys process model can be generated which gives process conditions and material properties for all the streams. An overall material balance spreadsheet can be imported from Hysys and shown into an Excel file.

Line sizing procedures can be fallowed and calculated for all the streams.

Using AutoCAD PFD and P and ID's can be generated.

A safety chart, showing the control logics involved in operation and maintenance of entire unit, can be made.

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Appendices

1. Physical Properties of glycols

	EG	DEG	TEG
Formula	$C_2H_6O_2$	$C_4H_{10}O_3$	$C_6H_{14}O_4$
Molecular Weight	62.1	106.1	150.2
Boiling Point at 760 mm Hg °F	387.1	472.6	545.9
Boiling Point at 760 mm Hg °C	197.3	244.8	288
Vapor Pressure at 25°C, mm Hg	0.12	0.01	0.01
Density at 25°C , g/cm³	1.110	1.113	1.119
at 60°C , g/cm³	1.085	1.088	1.092
Pounds per Gallon at 25°C	9.26	9.29	9.34
Freezing Point, °C	-13	-8	-7
Pour Point, °C	-	-54	-58
Viscosity in Centipoises at 25°C	16.5	28.7	37.3
At 60°C	4.68	6.99	8.77
Surface Tension at 25°C Dynes/cm	47	44	45
Refractive Index at 25°C	1.430	1.446	1.454
Specific Heat at 25°C	0.56	0.55	0.53
Flash Point, °C (COC)	116	138	160
Fire Point, °C (COC)	119	143	166

2. Temperture vs vapour pressure for Glycol
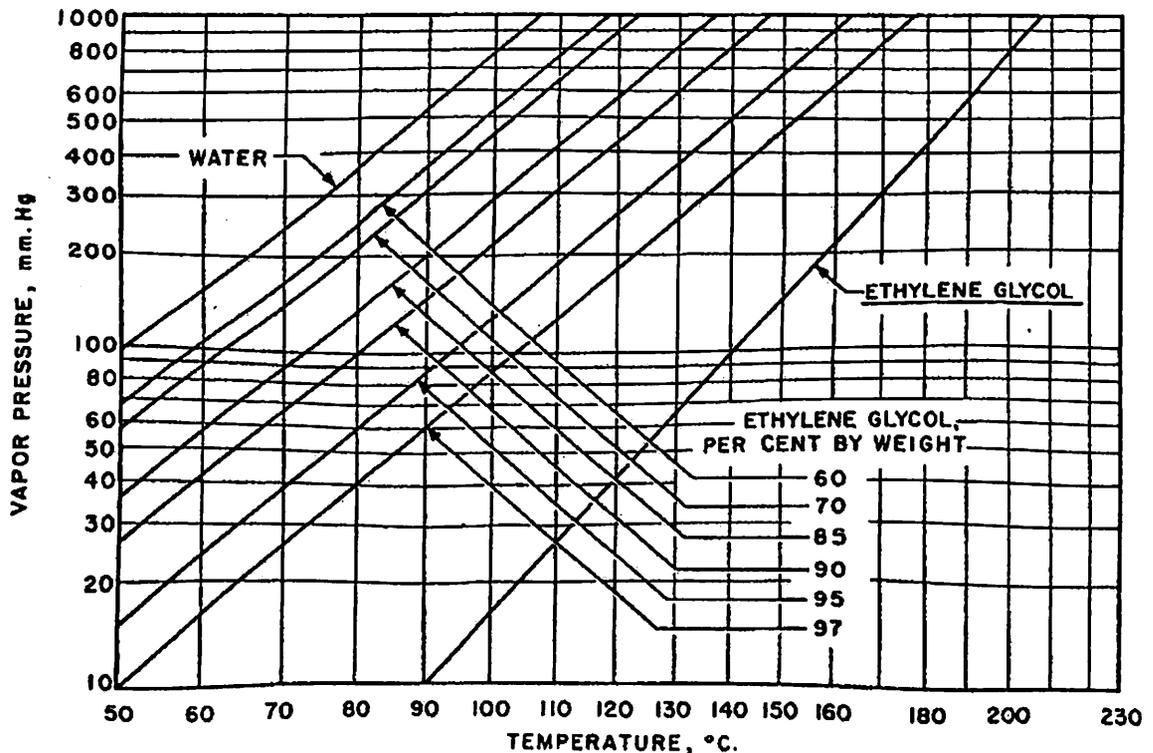
3. Water content of Sweet Natural Gas
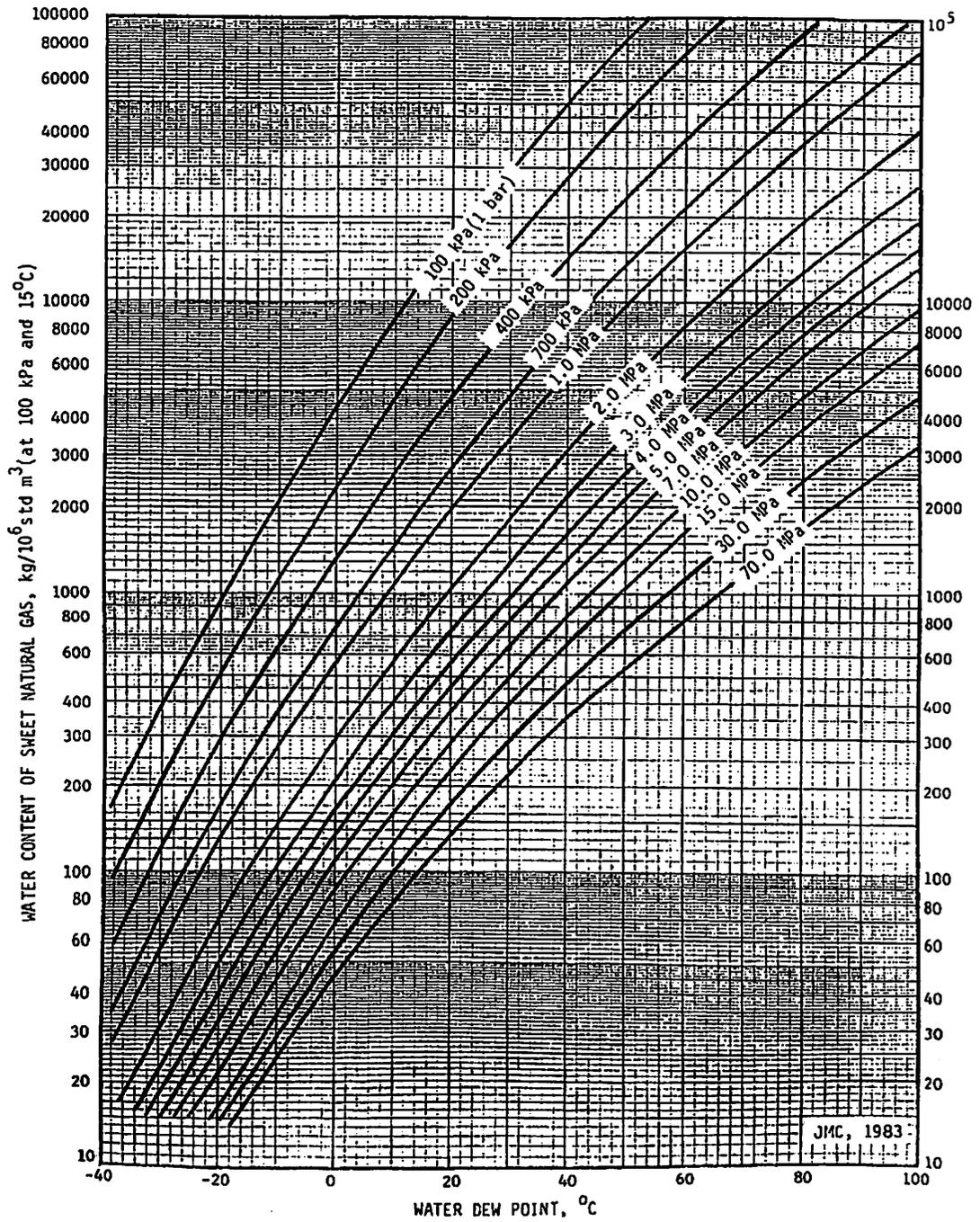
Water Content of Sweet, Lean Natural Gas. (Metric)