ENGINEERING DESIGN OF GAS GATHERING STATION

By

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Major Project

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

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Declaration

This is to certify that

1. The thesis comprises my original work towards the degree of Master of Technology in Chemical Process & Plant Design at Nirma University and has not been submitted elsewhere for a degree.
2. Due acknowledgment has beenmade in the text to all other material used.

Kedar Nath

Certificate

This is to certify that the Major Project report entitled "**Engineering Design of Gas Garhering Station**" submitted by **Mr. Kedar Nath (11MCHC06)**, towards the partial fulfillment of the requirements for the degree of Master of Technology in Chemical Engineering (Chemical process and plant design) of Nirma University of Science and Technology, Ahmedabad is the record of work carried out by him under my supervision and guidance. In my opinion, the submitted work has reached a level required for being accepted for examination. The results embodied in this report, to the best of my knowledge, haven't been submitted to any other university or institution for award of any degree or diploma.

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Abstract

The project involves developing an Excel model and HYSYS simulation of gas gathering station and gas dehydration process to obtain optimum parameters for the process, the product purity and recovery of the gas being separated.The dehydration is an important process in gas processing. The gas is dehydrate to avoid dangers associated with pipeline transport and processing of wet gas. The problem includes corrosion, water condensation and plug created by ice or gas hydrates.Gathering system is a typical representation of complex heterogeneous industrial system, which is not only an important oilfield production facility system for gathering and transporting oil and gas, but also a key section to realize the function of oil and water separation. Crude Oil is separated here into oil, gas and water.Received gas is further treated by gad dehydration process.

Nomenclature

A	Vertical vessel cross sectional area
A_B	Downcomer cross sectional area ft^2
D	vessel diameter
A_r	Total cross sectional area
C_d	drag coefficient dimensionless
D_n	dia ,in (iner or outlet vapor/liquid as specified)
D_p	droplet dia ft or microne
D_{vd}	Vapor disengagement dia ft
ge	gravitational constant 32.17 ft/s2
H	height, ft
H_a	liquid level above baffle in or ft
H_h	holdup height ft
H_L	height from liquid interface to LL nozzle ft
H_{LLt}	light layer thickness in
H_{LLv}	light liquid height in vessel
H_{me}	mist eliminator to top tan. Height ft.
K_s	Stokes law terminal velocity constant (in/min) (cp)(lb/ft^3)
K	Terminal velocity constant
L_{min}	Vapor /liquid separation minimum length ft
p	Pressure ,psig or psia

Abbreviation

GGS	Gas gathering Station
CTF	Combine tank form
GCP	Gas compressor plant
GCS	Gas collection station
ETP	Effluent treatment plant
FWKOD	Free Water Knock out Drum
IDBH	Indrict bath heater for LP Header
BTEX	Benzene, Toluene, Ethyl benzene and Xylene

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Chapter 1

INTRODUCTION

GGS (gas gathering station) is an installation which receives oil through manifolds from its different assigned fields. Oil is separated here in the form of oil, gas and water. Separated oil is sent to CTF for further treatment, separated gas is sent to GCP through GCS in order to receive back compressed gas for gas injection programme for gas lift wells. Gas from GGS is sent at 2-3 kg/m2 and received back at 41 kg/m2. Effluent water is sent to ETP and then to CWIP for its further treatment. This treated water finally from CWIP is sent back to GGS after its treatment to impart water injection programme for different wells. Both water injection and gas injection programmes are carried out and controlled by GGS [1].

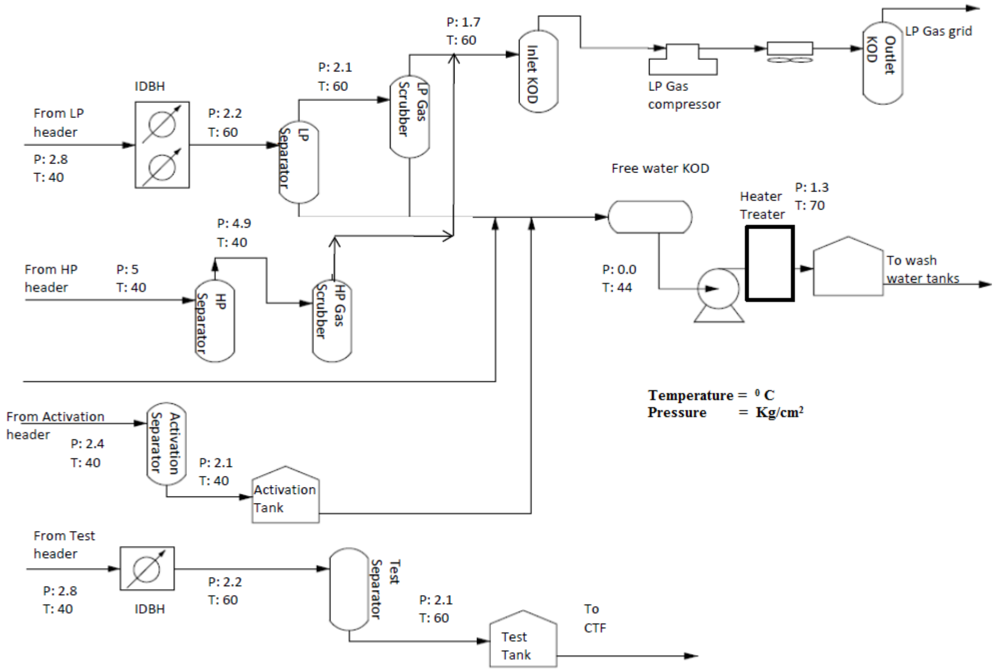


Figure 1.1: Process Flow Diagram of GGS

1.1 Capacity of the gas gathering system:

- Total oil processing capacity is 1900-2000m3/day
- It receives 20000 m3/day of gas at 6 kg/m2 for gas injection in gas lift wells.
- It receives oil from 110 wells.
- Crude oil is dispatched at the pressure of 5-10kg/cm2 to CTF.

1.2 Manifold

Purpose:

1. To group the wells based on their pressure.
2. To group the wells based on quality of oil, i.e. pure or emulsion
3. To isolate any wells for testing purpose.
4. To divert wells to the required header through operation of valves

1.3 Separator

Purpose:

To flash the well fluid to separate into liquids and gas at a controlled pressure.Fluid enters tangentially and due to the sudden pressure drop to the set level the fluid gets separated into liquid and gases. Baffles are fitted inside the separator to help in better separation of fluid. The fluid is given residence time to allow better separation.

1.4 Storage system

Purpose:

1. To store oil before pumping
2. To measure oil produced

Process: From heater treater and separator is taken into overhead cylindrical tanks for measurement

1.5 Oil despatch system

Purpose:

To dispatch oil from GGS to CTF.

Process: Oil is pumped from storage tanks to CTF by reciprocating pumps.

1.6 Chemical dosing system

Purpose:

To mix demulsifier chemicals into emulsion oil before it enters heater treater.

Process: Demulsifier chemical is pumped into emulsion oil stream before heater treater.to obtain optimum results proper dosing element should be injected.

1.7 Heater treater

Purpose:

Demulsification of oil into oil and water by chemical thermal and electrical means.

Process: Basically there are three sections in heater treater.

1. Heating chamber
2. Middle chamber
3. Electrical chamber

1. Heating chamber- the fire tubes which extends up to this section are in a submerged condition in emulsion oil. The heating of oil emulsion oil decreases the viscosity of oil and water and reduce the resistance of water movement. The heat further reduces the surface tension of individual droplets by which when they collide form bigger droplets. This progressive action results in the separation of oil and free water.
2. Middle chamber- the fluids from heating enter into this chamber through fixed water. It does not allow gas pass into electrical chamber. The gas which enters heating chamber leaves from top through mist extractor. The oil in this chamber is controlled by oil level controller.
3. Electrical chamber- In this section constant level of water is maintained so that oil is washed and free water droplets of water are eliminated before fluid proceeds towards electrode plates. These plates are connected with high voltage supply of 10,000 to 25,000 volts. When fluid passes through these electrodes the droplets are polarizes and extract each other. This attraction causes the droplets to combine; they become large enough to settle into oil and water layers by the action of gravity.

1.8 Effluent disposal system

Purpose:

To dispose effluent water from effluent handling tanks to effluent treatment plant.

Process: Effluent water is disposed by pumping effluent to effluent treatment plant through pipeline.

1.9 Water injection manifold

Process:

To divert water in water injection wells.

To monitor water injection in water injection wells.

Purpose: Water coming from water injection plant is diverted to different water injection wells through different lines of wells.

1.10 LP Header

In the proposed GGS receives well fluid from the wells. The LP header consists of 70 fingers (60 nos. : 4” & 10 nos. : 8”). For better flow & separation, the well fluid from LP header is heated in Indirect Bath Heaters i.e. part of Indirect Bath Heater Package prior to separation process. Heated well fluid from the Indirect Bath Heaters are routed to two phase LP Separators where gas and liquid are separated. Separated liquid (Oil + Water) is collected in Free Water Knock out Drum (FWKOD) to separate free water before being fed to the Heater Treaters Package . Gas from the LP Separators is passed through LP Gas Scrubber to separate the entrained liquid. The gas from the separators is also utilized as fuel gas for blanketing in Heater treaters. Gas from LP Gas Scrubber is routed to internal consumers, Gas Compressors. , Condensate from LP Compressor Package, Liquid from LP Scrubber and liquid from Skim oil tanks are collected in Free Water Knock Out Drum (FWKOD) mainly to provide retention time for separation of free water if any before being fed to the Heater Treaters Package .

1.11 HP Header

In the proposed set up, the HP header consists of 60 fingers (60 nos. : 4”). The well fluid received through HP header are routed to vertical two phase HP separator where gas and liquid are separated. Gas from the HP Separator is passed through HP Gas Scrubber to separate the entrained liquid. Gas from HP Gas Scrubber is routed to internal consumers. Provision is also kept for routing the gas to, external consumers & LP gas Compressor Package. Liquid from the HP Separator & HP Gas Scrubber is routed to LP Separatorsand thus joins the LP stream for further processing.

1.12 Test Header

In the proposed set up test header consists of 60 fingers (60 nos. : 4”). For better flow & separation, the well fluid through two streams from test header is heated in dual coil Indirect Bath Heater i.e. part of Indirect Bath Heater Package prior to separation process. Heated well fluid of 1st stream from dual coil Indirect Bath Heater is routed to two phase test separator where gas and liquid are separated. Provision is also kept to bypass bath heater and route the fluid directly to the test separator. Separated liquid (Oil + Water) is collected in Test tanks . Similarly, heated well fluid of 2nd stream from dual coil Indirect Bath Heater is routed to two phase test separator where gas and liquid are separated. Provision is also kept to bypass bath heater and route the fluid directly to the test separator. Separated liquid (Oil + Water) is collected in Test tanks . From Test Tanks , the liquid is pumped by Recycle pumps to Free Water Knock out Drum (FWKOD) .Gas from the Test Separators is routed to LP Gas Scrubber to separate the entrained liquid. Gas from LP Gas Scrubber is routed to internal consumers and LP gas Compressor Package .

1.13 Activation Header

In the proposed set up, new Activation header i.e. part of well manifold receives well fluidfrom the wells field. The activation header consists of 60 fingers (60 nos. : 4").The well fluid from activation header is routed to two phase activation separator where gas and liquid are separated. Separated liquid (Oil + Water) is collected in activationtank. From activation Tank, the liquid is pumped by Recycle pumpsto Free Water Knock out Drum (FWKOD).Gas from the activation Separator is routed to flare.

1.14 Oil/Water/Gas Processing

Separated liquid (Oil + Water) from LP separators, Liquid emulsion , Recycle from activation tank, Recycle from Test Tanks, Condensate from LP Compressor Package, Liquid from LP Scrubber and liquid from Skim oil tanks are collected in Free Water Knock Out Drum (FWKOD) mainly to provide retention time for separation of free water if any before being fed to the Heater Treaters Package. Free water separated in 1st chamber of FWKOD is pumped by Effluent Transfer Pumps to Effluent Wash Tanks located in ETP for further processing. The oil+ water emulsion from 1st chamber flows towards 2nd chamber of FWKOD from where it is pumped by Heater Treater Feed pumps to Heater Treater Package. Provision is also kept to bypass FWKOD in case of emergency power failure to Effluent transfer pumps and Heater treater feed pumps. Further separation of water is achieved in Heater treaters where oil & water emulsion get breaks and desired phase separation is achieved. Produced Water separated in the Heater Treater Package is also routed to Effluent Wash Tanks. Oil layer accumulated in wash tanks is routed by gravity to skim oil tanks. Oil from skim oil tanks is routed back to FWKOD by Skim Oil Transfer Pumps . Oil from Heater Treater Package is routed to Oil Storage Tanks. The Storage tanks is having one discharge headers i.e. Oil Dispatch Header. During normal operation oil is pumped to CTF by Oil Dispatch Pumps through pipeline. If off-spec. oil is received in oil storage tanks from Heater Treaters, the off-spec. oil is recycled back to Heater Treater Package by using one of the oil dispatch Pumps . Demulsifier dosing is done at the downstream of the Indirect Bath Heaters and inlet of the Heater Treaters. AntiScalent dosing is done at the manifold.

1.15 Water in gas

Water is a problem in the gas phase, both in gas processing and in pipeline transport.

The main problems with water in gas are:

- Corrosion

- Liquid water formation
- Ice formation
- Hydrate formation

In pipelines where it is known that the gas is wet, the problem can be countered. If it is known in the design phase the pipeline can be designed with more corrosion resistant materials or increased material thickness. If the problem occurs during production, the problem can be minimized by injecting inhibitors into the gas. In dry gas pipelines the problems ought not to occur, but can occur in case of insufficient dehydration. If not discovered the problems are more serious here, because the pipelines are not designed for these conditions. When discovered inhibitors can be added until adequate dehydration is available again. Liquid water in the pipeline is a problem, not only concerning liquids in compressors, but also a problem because the liquid water can create liquid plugs and increase corrosion. Ice formation is only a problem when the temperatures are adequately low for ice to form. Ice is especially a problem in process equipment and valves, where the ice can create blockages. Ice are manly a problem in low temperature gas treatment like NGL recovery and gas liquefaction (see section 2.3). When low temperature gas treatment is utilized ultralow water contents are required, making the requirements for the dehydration process more stringent. Although ice is a problem, gas hydrates are often more troublesome.

1.16 Gas hydrates

Gas hydrates are crystals of natural gas and water which can appear fare above the temperature where ice is formed. Gas hydrates are a caged structure containing a gas molecule like methane, the cage is formed by water through hydrogen bonding, as illustrated in Figure 1-2. Because the gas hydrate crystals are similar to ice crystals, the problems with gas hydrates are similar to those with ice, although gas hydrates are more troublesome because of the higher formation temperature. Because hydrates can form in pipelines, large amounts of hydrates can be in the gas Simultaneously; this can create plugs in the pipeline. Because of the potentially high hydrate contents in the gas the blockage can arise within minutes without any prior warning.

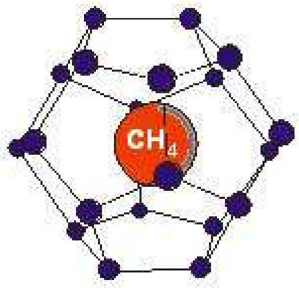


Figure 1.2: Gas Hydrates

1.17 Prevention

Because of the potential dangers from gas hydrates they must be prevented.

There are several methods to prevent gas hydrate formation, they are:

- Gas dehydration
- Raising the temperature
- Reducing the pressure
- Adding inhibitors

Gas dehydration is the most efficient way to prevent hydrate formation, but there may be practical limitation to the use of dehydration, e.g. one central dehydration unit. Gas dehydration will be treated further in section 2. If the gas stream cannot be dehydrated, one of the other prevention methods must be used. Raising the temperature of a pipeline is very impractical, likewise is reducing the pressure, because such a reduction will reduce the pipeline flow. The only practical solution is therefore to ad inhibitors to the gas.

1.18 Inhibitors

Inhibitors acts as antifreeze in the gas, the usual inhibitors are:

- Alcohols
- Glycols

Methanol and monoethylene glycol (MEG) are the most commonly used inhibitors, low doses are often injected continuously in pipeline where hydrate formation is a problem. Higher doses of especially methanol are used temporally to dissolve hydrate plugs. MEG is more viscous than methanol, but has the advantage of being easier to regenerate from the gas than methanol, because methanol regeneration is usually not feasible. MEG is the most commonly used glycol, because it is more efficient at a given mass concentration than diethylen glycol (DEG). DEG may nevertheless be used as inhibitor in the pipeline, but only if DEG also is the glycol used in the dehydration process afterwards. The different glycols are treated more thoroughly in section 2.4.

There are other possible inhibiters that prevent hydrate formation they are:

- Salts
- Ammonia
- Monoethanolamine

Salts are very rarely used because of the risk of corrosion and deposits. Ammonia is corrosive, toxic and can form solid deposits of carbonates obtained with carbon dioxide and water. Monoethanolamine is only attractive if it after pipe transport is used (and thereby recovered) for gas sweetening.

1.19 Gas treatment

The purpose of gas treatment is to clean the gas for unwanted impurities and get it to the desired condition before it is exported. The composition of the gas is the decisive factor for which gas treatment procedures that are used. The most common cleaning procedures are gas sweetening, dehydration and hydrocarbon recovery; more seldom treatments can be removal of inorganic elements. The purpose of cleaning the gas of its impurities is to improve the gas quality, avoid dangers to the process plant or pipeline from e.g. corrosion or enable the gas to be brought to the desired export condition. After purification usually only compression is required, for the gas to reach its desired export condition. In rare cases the desired export condition could require liquefaction of the gas.

1.20 Gas sweetening

To minimize corrosion it is often necessary to remove acid components in the gas. It is manly CO2 and H2S that are removed, although in some cases other sulphur components are present in the gas and must therefore also be removed. The most common sweetening procedure is absorption of the acid, with amines in an aqueous solution. Afterwards the rich amine solution is regenerated before it can be reused. Because the amines are in an aqueous solution, the sweet gas will be water saturated. Amine sweetening must therefore be conducted before gas dehydration. Absorption is the most common procedure, but other procedures can also be used. e.g. membrane processes if only carbon dioxide are to be removed.

1.21 Dehydration

The problems with wet gas have already been described in section 2.1, where dehydration was deemed to be the most efficient way to solve the problems associated with wet gas. Dehydration is usually done by absorption, although other processes like adsorption, membrane processes and refrigeration may be used. The dehydration process will be described in section 2.

1.22 Hydrocarbon recovery

In gas with a high content of C2+ components, there is a risk of NGL (Natural Gas Liquids) formation. NGL may be removed from the gas to avoid liquid in the pipeline or to sell the more expensive NGL separately, instead of as a part of the gas. Hydrocarbon recovery is performed by cooling the gas below its dew point temperature, condensing the more heavy hydrocarbons in the gas, the condensed liquid is then removed in a separator. The easiest way to cool the gas is in heat exchangers; this is most efficient at high pressure. Hydrocarbon recovery by cooling with heat exchangers may not yield the desired gas purity depending on the initial composition. In these cases the temperature can be lowered further by flashing the gas in a Jules-Thompson valve or in a turbo-expander. Because of the low temperatures achieved by flashing the gas, low water content is essential to prevent ice formation. Further improvements in hydrocarbon recovery can be achieved by distilling the liquid from the NGL recovery, thus recovering the methane condensed in this treatment.

1.23 Inorganic contents

If the gas quality is below pipeline quality because of contamination by inorganic elements, it is necessary to remove these impurities. Some of the inorganic components are only present in trace amounts, but can none the less create problems. The most common inorganic component is nitrogen, the nitrogen contents might be high, either naturally or if nitrogen is used for injection into the reservoir to improve hydrocarbon recovery. Nitrogen can be recovered by cryogenic distillation, adsorption or membrane separation. Radon may be present in the gas, it is radioactive, but with a half-life of 3.8 days the health problems from radon is minimal. The problem is that it decays into radioactive lead, which eventually will turn into non-radioactive lead. The result is that low-level radioactive materials will sediment in the process equipment and pipes; this constitutes a problem because cleaning produces radioactive waste.

1.24 Other contaminants

Benzene, Toluene, Ethyl benzene and Xylene (BTEX) are a problem because of environmental concerns. BTEX is removed from the gas during glycol dehydration, a smaller amount BTEX may also be removed during gas sweetening. When the glycol is regenerated the BTEX will be removed with the water, and thereby be vented to the atmosphere. BTEX are also a problem in cryogenic gas treatment because they can freeze like water. BTEX can not be removed from the gas before the dehydration. The BTEX problem can be reduced by using a light glycol, because BTEX is more solvable in larger glycols. Alternatively the vented gas from the glycol regenerator can be flared or treated to remove the BTEX before it is vented to the atmosphere.

1.25 Compression

The gas is compressed from the process pressure to the pipeline pressure in one or more steps, depending on the pressure difference. After each compression the gas is cooled and condensed liquids are separated off.

1.26 Liquefaction of the gas

Liquefied natural gas is an advantage when gas is stored or transported by non pipeline transport. Liquefaction of methane requires extensive refrigeration to temperatures as low as -161 °C. A very low water contents are therefore required. Liquefied natural gas is an advantage when gas is stored or transported by non pipeline transport. Liquefaction of methane requires extensive refrigeration to temperatures as low as -161 °C. A very low water contents are therefore required.

1.27 Water treatment

Unlike oil , gas and water treatment is an environmental issue. Water is a waste product in oil and gas production; therefore it is used for well injection. When water is separated off in the three-phase separation it still has a small hydrocarbon contents. This hydrocarbon contents constitutes no problem when the water is used for well injection, only when it is released into the well. Because of environmental concerns the hydrocarbons needs to be removed from the water so the contents is below the threshold limit value for water released into the well. The hydrocarbons in the water are oil that did not separate off in the separators and dissolved gas. First the oil is removed using hydro cyclones; the oil is lead back to the separator system. The gas is removed from the water by decreasing the pressure thus decreasing the solvability in the water. The gas is separated off before the water is released into the well.

1.28 Hysys simulation of GGS

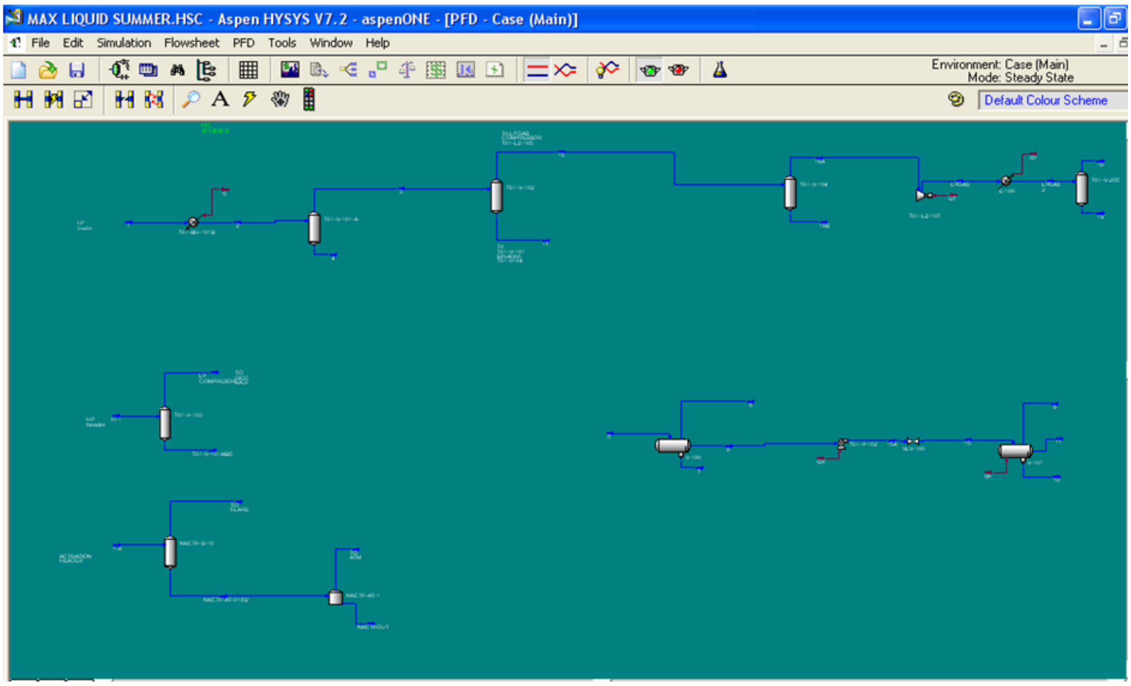


Figure 1.3: Gas gathering station simulation

1.29 Problem Identification

Natural gas that comes from oil wells is not totally pure but there are contaminants or mixtures in gas or typically termed ’associated gas’ like water vapor, hydrogen sulfide (H2S), carbon dioxide, helium, nitrogen, and other. These mixtures in natural gas can cause the problems for the production operation, transportation, storage and use of the gas. One of those contaminants is water content. This water can result in corrosion of pipeline and fittings in gas transmission systems and the formation of ice or hydrates that causing flow restriction, with resulting consequences in terms of plant operating efficiency.

1.30 Objective:

The main objective of the study are:

- Engineering design of process equipment
- Horizontal & vertical separator design
- Indirect oil bath heater design
- Analyze the water content is removed in natural gas.
- Process simulation of the plant “gas dehydration process” by hysys 3.2

Chapter 2

LITERATURE SURVEY

There are four methods that are used for gas dehydration; they vary in efficiency and cost.

2.1 Dehydration methods

The methods used for gas dehydration are absorption, adsorption, membrane processes and refrigeration. The methods may be used by themselves or be combined to reach the desired water contents.

In dehydration by absorption water is removed by a liquid with strong affinity for water, glycols being the most common. The lean (dry) glycol removes the water from the gas in an absorption column known as a contactor. After the contactor the rich (wet) glycol must be regenerated before it can be reused in the contactor. The regeneration is done by distilling the glycol thus removing the water. With glycol absorption it is possible to lower the water contents down to approximately 10 ppm vol, depending on the purity of the lean glycol. Gas dehydration by glycol absorption will be treated more thoroughly in section 2.4.

Dehydration by adsorption is done with a two bed system, where the beds are filled with adsorbents e.g. silica gel. The gas is lead through one of the adsorbers, where water is removed. Meanwhile the other adsorber is regenerated by blowing hot dry gas through it, this gas is then cooled and the water condenses. The Water is separated off and the gas is lead back to the wet gas, this is illustrated in Figure 2-1.

The efficiency of the adsorption process depends on the adsorbent used; there are several types of adsorbents available. The most efficient adsorbents are molecular sieves, this is aluminosilicates that have been altered to improve the adsorption characteristics, achieving a water contents as low as below 0.1 ppmvol.

In membrane processes the gas passes through a membrane that separates of the water. Membrane processes yields water content between 20-100 ppmvol . The problem with membrane processes are that they only become economically viable compared to glycol absorption at flows below 1.5-106 Nm^3/d (56 MMscfd).

Gas dehydration by refrigeration is a low cost dehydration method. Water condenses when the gas is cooled; the water is then removed in a separator. The separation method can be conducted numerous times. The method is most efficient at high pressure. The amount of water removed in the refrigeration process is often insufficient. Becrause of the low cost the refrigeration process are often used before the other dehydration processes.

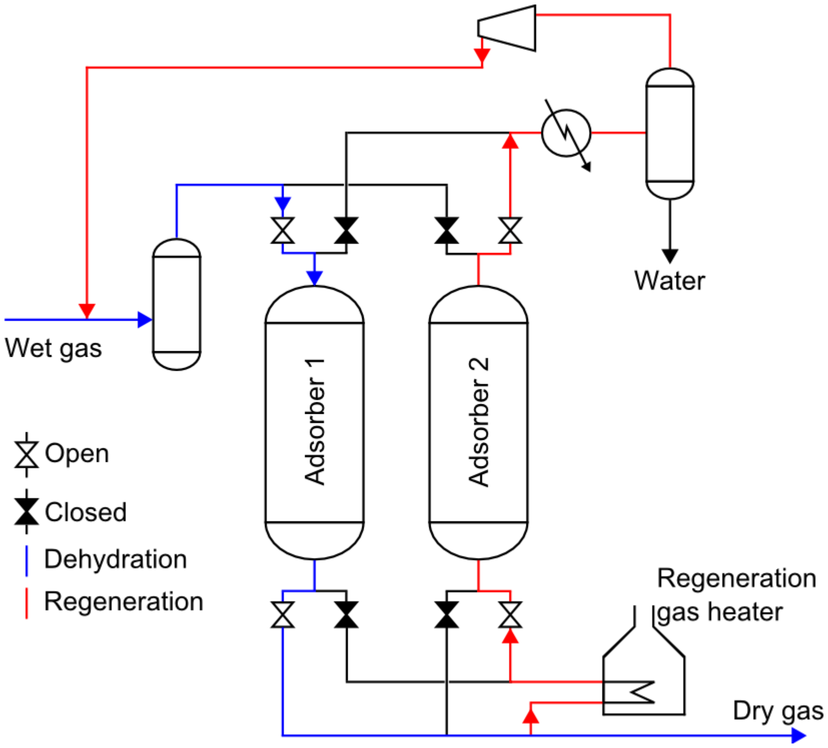


Figure 2.1: Absorption Method

2.2 Comparison of the methods

The two most efficient dehydration methods are absorption and adsorption. Absorption with glycol is the preferred dehydration method because it is more economical than adsorption. This is due to the following differences between absorption and adsorption:

- Adsorbent is more expensive than glycol.
- It requires more energy to regenerate adsorbent than glycol.
- Replacing glycol is much cheaper than replacing an adsorption bed.
- Glycol can be changed continuously, while changing an adsorption bed requires a shutdown.

Some low temperature treatment like liquefaction requires water content below what glycol plants can achieve. In these cases an adsorption plant is required, to minimize the cost this can be combined with a glycol plant that removes the majority of the water.

2.3 Water absorption

The basis for gas dehydration by absorption is the absorbent; there are certain requirements for absorbents for gas dehydration:

- Strong affinity for water to minimize the required amount of absorbent.
- Low affinity for hydrocarbons to minimize hydrocarbon loss during dehydration.
- Low volatility at the absorption temperature to minimize vaporization losses.
- Low solubility in hydrocarbons, to minimize losses during absorption.
- Low tendency to foam and emulsify, to avoid reduction in gas handling capacity and minimize losses during absorption and regeneration.
- Low viscosity for ease of pumping and good contact between the gas and liquid phases.
- Large difference in volatility and boiling point compared to water to minimize vaporization losses during regeneration.
- Good thermal stability to prevent decomposition during regeneration.
- Low potential for corrosion.

2.4 Glycols used for dehydration

Glycol is a common name for diols; with the two alcohols these substances have a high affinity for water. In dehydration 1,2-ethandiol also known as Monoethylen glycol (MEG) and the small polymers of MEG (diethylen glycol (DEG), triethylen glycol (TEG) and tetraethylen glycol (TREG)) are the most commonly used for absorbents. Higher polymers than TREG is usually not used for dehydration because they become too viscous compared to the smaller polymers.

Properties for MEG, DEG, TEG, TREG and water are compared in Table 2-1.

In Table 2-1 the important values are the normal boiling point, vapor pressure, viscosity, maximum recommended regeneration temperature and the onset of decomposition.

The normal boiling point and vapor pressure has an influence in the distillation. The greater the difference for these properties between the top and bottom product, the easier it is to separate the components. The separation between glycol and water is important because the water contents in the lean glycol determine the amount of water the glycol can remove from the gas.

The larger polymers TEG and TREG have the best properties for dehydration. TREG has slightly better properties than TEG, but because of the additional cost of TREG, TEG offers the best cost/benefit compromise and is therefore the most commonly used glycol.

The decomposition temperature is the point where DEG, TEG and TREG begin to react with the water and decompose into MEG. The temperatures in (240 °C) originate from manufacturer data. These temperatures are just below and above the maximum

Table 2.1: Properties of MEG, DEG, TEG, TREG and water

	MEG	DEG	TEG	TREG	WATER
Formula	$C_2H_6O_2$	$C_4H_{10}O_3$	$C_6H_{14}O_4$	$C_8H_{18}O_5$	H_2O
Molar mass [kg/kmol]	62.07	106.12	150.17	194.23	18.015
Normal boiling point [°C]	197.1	245.3	288.0	329.7	100.0
Vapor pressure @ 25 °C [Pa]	12.24	0.27	0.05	0.007	3170
Density @ 25 °C [kg/m3]	1110	1115	1122	1122	55.56
Viscosity @ 25 °C [cP]	17.71	30.21	36.73	42.71	0.894
Viscosity @ 60 °C [cP]	5.22	7.87	9.89	10.63	0.469
Maximum recommended regeneration temperature [°C]	163	177	204	224	-
Onset of decomposition [°C]	-	240	240	240	-

recommended regeneration temperature of 204 °C. This indicates that some TEG will decompose at 204 °C. At this temperature there will be some hot-spots in the boiler where the temperature will exceed 207 °C. When TEG decomposes it becomes MEG and DEG, therefore it will not influence the dehydration process, only give a slightly larger glycol loss because MEG and DEG are more volatile than TEG.

2.5 Dry Gas

The efficiency of the dehydration is measured on the water contents in the dry gas. The dew-point temperature for the water in the gas is often a more useful parameter than the total water contents. The dew-point temperature must be below the minimum pipeline temperature, to avoid liquid in the gas pipeline. Figure 2-2 shows the relation between dew-point temperature and the water contents in the lean TEG at different temperatures.

A dew-point temperature of 6 to 11 °C below the desired dew-point may be used to insure against non-ideal situations. The water dew-point may differ from the gas dew-point; the total gas dew-point may be influenced by other hydrocarbons in the gas. This can result in condensation of hydrocarbons in the gas pipeline; this is also undesirable but much less so than water condensation.

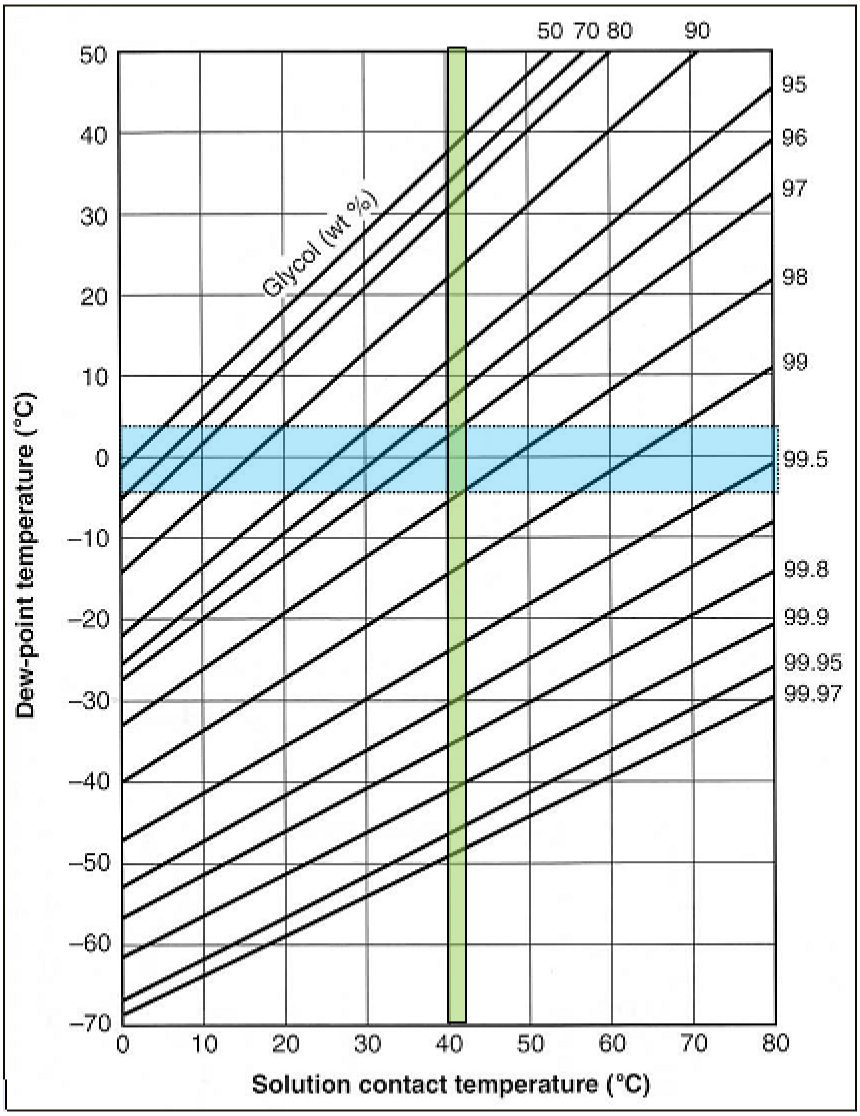


Figure 2.2: Water due point, after dehydration with TEG[8]

Chapter 3

Process Design Calculation

1. Hydraulics
2. Vertical Separator Design
3. Horizontal Separator Design
4. Oil Bath Heater Design

3.1 Hydraulics

The term hydraulics is just used to describe the pressure changes in a fluid as it flows from a given source to a given destination through piping or open channels, overground or underground or both, using the pressure of a pump or compressor, or just flowing by gravity. A fundamental thing to remember is that line sizing and hydraulics is related mainly to economics. There is no technical significance except for erosional velocity limit and velocity of sound. It is done mainly to lower the operating and initial costs. There is no point using a larger, more expensive pipe in place where a smaller, more cost effective pipe can be fitted.

Maybe it is worth mentioning that on offshore oil & gas wellhead platforms, in general, sufficient upstream pressure is available and so a mere check that the velocity is less than erosional limit, should suffice. But on production (process- or processing-) platforms, full hydraulics should be done, not mere line size check as mentioned above.

By far the most important aspect to remember in line sizing/hydraulics is that if line sizes are chosen reasonably, the pressure drop contribution of pipes, bends and elbows, and tees, is usually a negligible part of the overall hydraulics. Even valves, if there are not too many in series in a given circuit, do not contribute significantly, unless they are smaller in size than the line, or reduced bore by two sizes or more.

Factors that contribute to hydraulics are (in decreasing order of importance):

- Elevation;
- Pressure drop in Control Valves;

- Pressure drop in Equipment;
- Pressure drop in Line Fittings

Elevation: Even if you do not have a plant layout, you have to estimate the elevation differences involved, e.g. based on size and relative positions of equipment, or based on pre-determined floor levels.

Control valves: Control valves are usually sized to have a pressure drop of 30 % of total pressure drop of the circuit excluding elevation differences, or 50 % of the pressure drop in all other items than control valves, as before excluding elevation differences. Control valves are always sized deliberately to have somewhat high pressure drop, or else they will not control effectively since other resistances in the circuit would govern. Since control valve is a variable resistance because it opens and closes, when checking hydraulics, especially at late stage of project, the actual size of valve selected by Instrumentation department, and its CV at full opening should be re-checked. In ideal situations the control valve pressure drop at 60 to 70 % opening should be considered. But in non-ideal (re-check, revamp) situations, opening of down to 10 to 20 % and up to 90 to 95 % should be accepted while considering pressure drop.

Equipment: Equipment usually have relatively low pressure drop individually, but their contribution increases as more equipment are added in series. For equipment, we take the maximum (fouled, choked) pressure drop for hydraulics. But if examining different flow rate cases for hydraulics, the appropriate pressure drop for each flow case is taken.

Line fittings: This refers to strainers, drainers, steam traps, static mixers, etc.

3.1.1 Line Sizing

Line Sizing is only one of the steps involved in hydraulics, but often it is considered an exercise in itself. Pipe sizing is one of the most important job of a process engineer. Piping is a major cost component in any process plant, it may vary between 5 to 15% of total plant invest ment and optimum design may result in considerable cost reduction. Therefore it becomes necessary that extreme care should be taken while sizing process piping. Line sizing includes calculation of pressure drop at valves, bends, strainers and other accessories. It also includes calculation of frictional losses for particular pipe diameters. Selecting a particular pipe diameter depends on these calculations.

3.1.2 Bernoulli's theorem

The Bernoulli Theorem is a mathematical derivation based on the law of conservation of energy. This theorem states that the total energy of a fluid at any particular point above a datum plane is the sum of the elevation head, the pressure head, and the velocity head[17].

$$H = Z_e + \frac{P}{\rho} + \frac{V^2}{2g} \tag{3.1}$$

If there are no friction losses and no energy is added to or taken from the system, H is constant for any point in the fluid. In reality, whenever fluid is moving there is friction loss (hL). This loss describes the difference in total energy at two points in the system. Expressing the energy levels at Point 1 versus Point 2 then becomes[17]:

$$Z_{e1}+\frac{P1}{\rho 1}+\frac{V1^2}{2g}=Z_{e2}+\frac{P2}{\rho 2}+\frac{V2^2}{2g}+hl \tag{3.2}$$

3.1.3 Flow in pipes and Reynolds number

At low velocities, fluid molecules or particles carried by the fluid move in a reasonably straight line. Velocity of the fluid is maximum at the center of the pipe and zero at the pipe wall. This flow pattern is referred to as laminar. If the velocity is increased it will reach a critical point where fluid particles begin to show a random motion transverse to the direction of flow. This is the critical velocity. This random motion is typical of what is referred to as turbulent flow. Above the critical velocity the flow is considered to be completely turbulent even though there is always a boundary layer at the pipe wall where flow is laminar. In the turbulent zone the velocity profile is more nearly straight across the face of the pipe.

Reynolds developed a dimensionless number that may be considered as the ratio of the dynamic forces of mass flow to the shear stress due to viscosity.

$$Re=\frac{DV\rho}{\mu} \tag{3.3}$$

If the Reynolds number is less than 2000, flow may be considered laminar. If it is above 4000, the flow is turbulent. In the zone between 2000 and 4000 the flow could be either turbulent or laminar, but cannot be predicted by the Reynolds number.

If a non-circular conduit is encountered, the Reynolds number can be approximated by using an equivalent diameter for D. The equivalent diameter would equal four (4) times the hydraulic radius.

Hydraulic Radius = Area of flowing fluid / Wettedperrimete

This conversion would not apply to extremely narrow shapes where the width is small relative to the length. In such cases an approximation may be used wherein one-half the width of the passage is equal to the hydraulic radius.

3.1.4 Pressure loss due to friction

Flow is always accompanied by friction. This friction results in a loss of energy available for work. A general equation for pressure drop due to friction is the Darcy-Weisbach (often referred to as simply the Darcy) equation. This equation can be rationally derived by dimensional analysis, with the exception of the friction factor, fm, which must be determined experimentally[17].

$$H_l = \frac{F_m L V^2}{2GD} \tag{3.4}$$

Converting to kpa, eqution becomes

$$\Delta P_f = \frac{0.5\rho F_m L V^2}{d} \tag{3.5}$$

The Moody friction factor, fm, is used in the equations above. Some equations are shown in terms of the Fanning friction factor, ff, which is one fourth of fm (fm = 4.0 ff). A graph of both Fanning and Moody friction factors as a function of Reynolds number appears in Figure 1.1 (Appendix-1)

3.1.5 Friction factor and Pipe Roughness

When the fluid flow is laminar (Re<2000), the friction factor has a direct relationship to the Reynolds number, such that

$$F_m = \frac{64}{Re} or F_f = \frac{16}{Re} \tag{3.6}$$

Pipe roughness has no effect on the friction factor in laminar flow.

When the flow is turbulent, the friction factor depends on the Reynolds number and the relative roughness of the pipe, e/D, which is the roughness of the pipe, e, over the pipe diameter, D.

Figure 1.1 (Appendix-1) incorporates the relative roughness of the pipe into the determination of the friction factor. Figure 1.2 (Appendix) indicates relative roughness and friction factors for various piping materials.

$$\frac{1}{\sqrt{F_m}} = -2log_{10}\left(\frac{\varepsilon}{3.7D} + \frac{2.52}{R_e\sqrt{F_m}}\right) \tag{3.7}$$

3.1.6 General Considerations

3.1.6.1 Economic considerations

Economics play a major role in pipe sizing. As pipe size increases, investment for piping increases but pipe has less pressure drop resulting in less pumping cost. Generally economic velocity and pressure drop for various service conditions should be used to get an economic pipe size. However, in case of long lines, special material of construction etc, may become necessary to do economic evaluation.

3.1.6.2 Velocity and Pressure Drop Consideration

These considerations become important when fluid flows under gravity, under its own pressure or certain pressure has to be dropped in the flow system. Sometimes it may not be possible to absorb the entire pressure in pipeline because of excessive velocity and additional restrictions to flow may be necessary. In such cases, the size of the line is not governed by economic considerations.

3.1.7 Special Considerations

3.1.7.1 Structural and Mechanical Requirement

Mechanical strength of pipe reduces as the diameter decrease. Small size pipes have greater tendency to sag when they are erected and additional supports become necessary to keep them in position

3.1.7.2 Unusual Flow Situations

1. Vortexing: Liquid drawn from a vessel can entrap vapours by a vortex if velocity in the pipeline is more than a certain limit. Vortexing is undesirable because it not only affects the performance of components like pumps and control valves but can damage them also. In general, vortexing occurs when the static head above a liquid draw off nozzle is less than two velocity head.

2. Water hammer: Shock due to water hammer generally prevails in lines equipped with check valves, quick closing valves (plug, butterfly, ball valves) or in lines connected to reciprocating pumps. Improper sizing of condensate lines, malfunctioning of steam traps can cause water hammering.

3. Vibration: Higher velocity will favour vibrations. The most common instances of vibration in piping occur in the following:

- Lines with higher velocity
- Fluctuating flow lines
- Transfer line from heaters to column
- A bend very close to pump discharge line

3.1.7.3 Velocity limitations

In certain situations, minimum and maximum velocity limits are dictated by process conditions rather than economics etc. These situations are described below:

1. Erosion velocity: Velocity in any piping system should be below erosion velocity limit, otherwise pipe material will start eroding and will soon lead to a failure. Erosion velocity for a single phase line is too high to occur in practice. For two phase hydrocarbon liquid and vapour lines, erosion velocity is given by
$\mathrm{v}_{ero} = 122/\sqrt{Density\,(kg/m^3)}\ (m/s)$, for continuous service
$\mathrm{v}_{ero} = 195/\sqrt{Density\,(kg/m^3)}\ (m/s)$, for intermittent service

2. Erosion of protective layer: Sometimes certain corrosion inhibitors are used in flowing material which forms a protective layer on the inner surface of pipe. If velocity exceeds certain limits, the protective layer gets eroded and corrosion inhibitors become ineffective. Velocity limitations in such cases come from the manufacturers of corrosion inhibitors.

3. Settling of solid particles: If solid particles are suspended in a liquid phase, they should not be allowed to settle down in the pipeline. To avoid settling a minimum velocity of about 0.9 m/sec should be maintained in the pipe.

3.1.7.4 Equipment limitations

Sometimes equipment specifications govern line size. For example, pump suction lines should be sized to take care of pump NPSH requirements.

3.1.7.5 Space limitations

In some cases, where a new pipe is to be provided in an existing plant, the available space may dictate pipe size. Or in cases like offshore, where space is a constraint, pipe size selection plays a vital role.

3.1.7.6 Codes/statutory requirements

Some times velocities and pressure drops in pipelines are dictated by various engineering codes or statutory rules. Some cases are described below:

- Safety valve inlet piping :Maximum pressure drop allowable in the inlet line of the safety valve should not exceed 3% of the set pressure of the safety valve, This is as per API recommendation.

- ii. Limitation on Noise level :Whenever a velocity approaches sonic velocity, intolerable noise is generated. This normally does not occur for liquid lines because sonic velocity is too high to occur in process plant. But for gases, whenever the pressure drop across the restriction is about 50% of upstream pressure, sonic flow occurs. To avoid this, whenever a lot of pressure reduction is required, it should be done in stages. For example, steam venting or air venting.

3.1.8 Sizing Criteria

3.1.8.1 Sizing criteria for Liquid flow lines

Single-phase liquid lines should be sized primarily on the basis flow velocity. For lines transporting liquids in single phase from one pressure vessel to another by pressure differential, the flow velocity should not exceed 15 feet/second at maximum flow rate to minimize flashing ahead of the control valve. If practical, flow velocity should not be less than 3 feet/second to minimize deposition of sand and other solids. At these flow velocities the overall pressure drop in the piping will usually be small.

- Flow velocities in liquid lines may be read from Figure 1.3 (Appendix-1) or may be calculated using the following derived equation:

$$V = \frac{0.12Q_l}{d_l^2} \tag{3.8}$$

Where, V$_l$ = average liquid flow velocity, feet/second
Q$_l$ = liquid flow rate, barrels/day.
d$_i$ = pipe inside diameter, inches.

- Pressure drop (psi per 100 feet of flow length) for single phase liquid lines may be read from Figure 1.4 (Appendix-1) or may be calculated using the following (Fanning) equation:

$$\Delta P = \frac{.00115 f Q_l^2 S_l}{d_l^5} \tag{3.9}$$

where: ΔP = pressure drop, psi/100 feet.
f = Moody friction factor, dimensionless.
Q$_l$ = liquid flow rate, barrels/day.
S$_l$ = liquid specific gravity (water = 1).
d$_i$ = pipe inside diameter, inches.

- The Moody friction factor, f, is a function of the Reynolds number and the surface roughness of the pipe. The modified Moody diagram, Figure 1.1 (appendix-1), may be used to determine the friction factor once the Reynolds number is known. The Reynolds number may be determined by the following equation:

$$R = \frac{\rho d_f V_l}{\mu} \tag{3.10}$$

where: R$_e$ = Reynolds number, dimensionless.
ρ_l = liquid density, lb/ft3.
d$_f$ = pipe inside diameter, ft.
V$_l$ = liquid flow velocity, ft/sec.
μ_l= liquid viscosity, lb/ft-sec,
or = centipoise divided by 1488,
or = (centistokes times specific gravity) divided by 1488.

3.1.8.2 Sizing criteria for Gas flow lines

Single phase gas lines should be sized so that the resulting end pressure is high enough to satisfy the requirements of the next piece of equipment. Also velocity may be a noise problem if it exceeds 60 feet/second; however, the velocity of 60 feet/second should not be interpreted as absolute criteria. Higher velocities are acceptable when pipe routing, valve choice and placement are done to minimize or isolate noise.

The design of any piping system where corrosion inhibition is expected to be utilized should consider the installation of additional wail thickness in piping design and/or reduction of velocity to reduce the effect of stripping inhibitor film from the pipe wall. In such systems it is suggested that a wall thickness monitoring method be instituted.

- General Pressure drop equation

$$P_1^2 - P_2^2 = 25.2\frac{SQ_g^2 ZT_1FL}{d^5} \tag{3.11}$$

Where, P1 = upstream pressure, psia
P2 = downstream pressure,. psia
S = gas specific gravity at standard conditions
Qg = gas flow rate. MMscfd (at 14.7 psig and 60 F)
Z = compressibility factor for gas (Refer to GPSA handbook)
T1 = flowing temperature, oR
f = moody friction factor, dimensionless
d = pipe ID, in.
L = length, feet
When, the change in pressure, is less than 10% of the inlet pressure, then,

$$\Delta P = 12.6\frac{SQ_g^2 ZT_1FL}{P_1 d^5} \tag{3.12}$$

3.1.9 Procedure for pipe sizing for single phase flow

1. Basic Data required temperature, pressure, flow rate, physial properties.

2. **State of Fluid**:Designer should know if the fluid is liquid or vapour or mixed phase. It is important because the physical properties are widely different depending on the state of the fluid. Separate charts are to be used for liquid, vapor, and two phase.

3. **Flow rate**:The flow rate for line sizing shall correspond to the maximum fluid quantities established by the process design condition plus 10% over-design factor. Flow rate shall be expressed in m3/hour for liquid and in kg-s/hour for vapor lines. If line has to handle different flow rates or different materials at different times, line size shall be calculated for all cases independently and then the proper size that will suit to all conditions shall be selected.

4. **Fluid properties**:The fluid properties namely, density and viscosity shall be specified at the flowing conditions. These shall be average values for the length of the line in general cases. In cases where extreme variation occur in these properties

over the pipe length, the total length of pipe shall be divided appropriately in small sections for purpose of line sizing calculations.This is required generally in cases of compressible fluids. Viscosity expresses the readiness with which a fluid flows when it is acted upon by an external force. Two types of viscosity measurements are used, absolute and kinematic. Absolute viscosity is a measure of a fluid's internal resistance to deformation or shear. Kinematic viscosity is the ratio of absolute viscosity to mass density. Viscosity is temperature dependent. The viscosity of most liquids decreases with an increase in temperature, whereas that of gases increases. Pressure has almost no effect on the viscosity of liquids or near perfect gases. On the other hand, the viscosity of saturated or slightly superheated vapors is changed appreciably by pressure changes. Specific volume is the inverse of density. Relative density of a liquid is the ratio of the density of the liquid at a specified temperature to the density of water at 15°C. The relative density of gas is defined as the ratio of the molecular mass of the gas to the molecular mass of air.

5. **Recommended pipe sizing criteria**:Analyze each individual case carefully and select appropriate sizing criteria to decide allowable velocity and pressure drop.

6. **Equivalent length of pipeline**:To calculate the total pressure drop in pipeline, its equivalent length is required. It consists of two components namely, straight length of pipe and equivalent straight length of various pipe fittings and valves. The latter component takes care of the additional pressure drop due to valves and pipe fittings.

3.1.10 Method followed for Line sizing

1. Assume or select a nominal pipe size (NPS). Select the schedule number and pipe I.D. most appropriate to the NPS.

2. Calculate the velocity of the flowing medium at the selected NPS. (equation 3.8)

3. Check whether the calculated velocity of the flowing medium is within the allowable limit velocity range. If it is higher than the allowable velocity, increase the NPS to the next available size. If it is lower than the allowable velocity, proceed to the next step.

4. Calculate the pressure drop in the selected NPS at the calculated velocity. If the pressure drop is higher than the allowable pressure drop, then increase the NPS. If the calculated pressure drop is lower than the allowable pressure drop, then the selected pipe size is appropriate. (equation 3.9)

3.2 Separator Design

Separators are basically drums used to separate heterogeneous mixtures. Heterogeneous mixtures consist of two or more different constituents, which generally do no chemically react with each other. The components of a heterogeneous mixture can be separated using one or more appropriate techniques. Extreme care is to be taken while designing and specifying process vessels, not only because their contribution to the total plant investment may vary between 10-30%, but also because smooth operation of several critical equipments in a process plant largely depends on the performance and adequacy of process vessels.

Out of a number of options available, the aim is to choose a vessel which is best suited to the requirements. Firstly, we define the mandatory requirements which the separator shall satisfy. The different factors that may play a role are cited below. Secondly, we decide the orientation of the vessel. Finally, we make the final choice by comparing the different options we have for the best results.

3.2.1 Separator Classification Based on Service

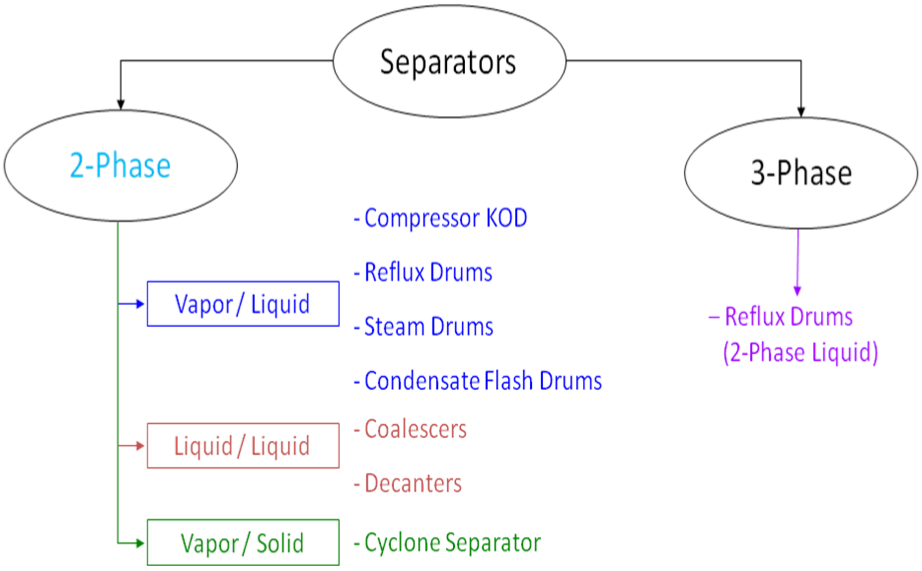


Figure 3.1: Classification of Separator

3.2.2 Selection Strategy

To facilitate the choice of a separator type for an application, the following points are compared:

1. Duty:Whether we want to achieve bulk separation or complete separation.
2. Gas Handling Capacity:This includes the maximum capacity that the separator may be needed to handle under the most severe conditions and the turndown ratio (ratio of maximum to minimum flow).

3. Liquid Removal Efficiency:The liquid removal efficiency needs to be checked against the overall efficiency required and with respect to the removal of mist.
4. Liquid Handling Capacity:This mainly includes handling the slugs that may form.
5. Fouling Tolerance:The main fouling agents that may be present include sand and other sticky materials.
6. Pressure Drop:Pressure drop in different vessels vary, and hence the chosen vessel should comply with the pressure drop available.

3.2.3 Vapor Liquid Separators

The separation of gases and liquids needs to be carried out at different stages in a chemical process or operation. The wet gas, more commonly known as the gas/liquid stream is made to enter the separating vessel and the gas and liquid streams emerge separately. The efficiency with which the separation needs to be carried out depends upon the future usage of streams. For example, a simple knock out vessel will suffice if the stream is heading towards a heat exchanger where condensation will take place, ensuring that sufficient bulk separation happens. On the other hand, if the separator lies upstream of a gas compressor, the efficiency will need to be very high, so as to not cause any damage to the forthcoming equipment. Such factors will determine the internals that need to be provided in the vessel.

3.2.4 Principles of Separation

Three principles used to achieve physical separation of gas and liquids or solids are momentum, gravity settling, and coalescing. Any separator may employ one or more of these principles, but the fluid phases must be "immiscible" and have different densities for separation to occur.

1. **Momentum**:Fluid phases with different densities will have different momentum. If a two phase stream changes direction sharply, greater momentum will not allow the particles of the heavier phase to turn as rapidly as the lighter fluid, resulting in separation. Separation using momentum forms the first stage if a number of stages are used to facilitate separation and accomplishes bulk separation of the two phases in a stream. An inlet diverter is used, which causes the liquid droplets entrained in the vapor to impinge the diverter and then drop by gravity.
2. **Gravity** Settling:This forms the secondary stage, where the liquid droplets settle by gravity as the vapor flows through the disengagement area. When the net gravitational force is equal to or greater than the drag force of the gas flowing around the droplet, the heavier droplets will settle down. The vapor velocity needs to be less than the terminal velocity for the separation to occur.
3. **Coalescing**:Very small droplets such as mist cannot be separated by gravity alone. These droplets can be coalesced to form larger droplets that will settle by gravity. Coalescing devices in separators force the gas to follow a tortuous path. The momentum of the droplets causes them to collide with other droplets or the coalescing

device, forming larger droplets. These larger droplets can then settle out of the gas phase with gravity. Wire mesh screens and vane packs are typical examples of coalescing devices.

3.2.5 Orientation

Table 3.1: Orientation

Vertical vessel is preferred if:	**Horizontal vessel is preferred if:**
Vapor liquid ratio is high	Vapor liquid ratio is low
Area available for gas flow has to be constant	Large liquid slugs have to be accommodated
Re-vaporization or re-mixing of fluids is to be avoided	A large liquid surface area Separator is required for de-gassing purposes
Stable vapor liquid separation is required even if liquid level fluctuates frequently	A low downward liquid velocity is required (like handling a foaming liquid)

3.2.5.1 Feed Inlet

The feed comes through the upstream piping, passes through the inlet nozzle and the inlet device. The major factor that comes into picture while deciding the inlet device and the diameter of the feed nozzle is the momentum of the incoming feed. The momentum of the incoming feed is a deciding factor since too much momentum can lead to impingement of the vessel upon feed's inlet.

Feed Momentum = $\rho_m V^2 in$

Where,

ρ_m: mean density of the mixture in the feed pipe i.e. (M$_g$ + M$_l$) / (Q$_g$ + Q$_l$), kg/m^3

v$_m in$, velocity of the mixture in inlet nozzle i.e. (Q$_g$ + Q$_l$) / Area of Inlet Nozzle, m/s

3.2.5.2 Purposes Served by Inlet Device

- It reduces the inlet stream's momentum
- It facilitates the bulk separation or gas and liquid
- It enhances flow distribution of gas and liquid phases

Commonly used inlet devices include half-open pipe and Schoepentoeter.

3.2.5.3 Vertical Knock-out Drum

The knock-out drums are typically characterized by absence of internals. The vessel diameter is kept sufficiently large to facilitate low gas velocity in the vessel, hence a majority of droplets settle by gravity. Basic features are given below:

Application Bulk separation of gas and liquid

Characteristics

- Unlimited turndown
- High slug handling capacity
- Liquid removal efficiency typically 90%
- Very low pressure drop
- Insensitive to fouling

Recommended Use

- Vessels where internals have to be kept to a minimum
- Fouling service
- Foaming service

Non-recommended Use Where efficient demisting of gas is required

Typical Process Applications

- Vent and flares stack knock out drums
- Production separator
- Bulk separator (e.g. upstream of gas coolers)
- Flash vessel

3.2.5.4 Horizontal Knock-out Drum

This is similar in basic characteristics to the vertical knock-out drum.

Application Bulk separation of gas and liquid

Characteristics

- Can handle large liquid fractions
- Unlimited turndown
- High slug handling capacity
- Liquid removal efficiency typically 90%
- Very low pressure drop
- Insensitive to fouling

Recommended Use

- Vessels where internals have to be kept to a minimum and where there are height limitations
- Slug catchers - fouling service e.g. wax, sand
- For foaming or very viscous liquids

Non-recommended Use Where efficient demisting of gas is required

Typical Process Applications

- Vent and flares stack knock out drums
- Production separator for low gas/oil ratio
- Bulk separator (e.g. upstream of gas coolers)
- Slug catcher

3.2.6 Designing of Vertical Separator

Input Required The following input data is required for designing of a vessel

1. Inlet and outlet gas and liquid flow rates
2. Physical properties such as liquid and gas densities and liquid viscosity
3. Diameter of inlet and outlet pipes and thickness of the demister mat
4. Hold-up times for control and operator intervention
5. Slug and foaming allowances

Output After the designing has been done, we will have the following as our output:

1. Inlet device
2. Vessel diameter
3. Vessel height
4. Pressure drop in the vessel
5. Inlet and outlet nozzle diameters

3.2.6.1 Designing Procedurem

Step I: Inlet Device

Calculate the momentum of the inlet feed, Feed Momentum = ρ_m $v_m{}^2$ in, Pa

Where: ρ_m: mean density of the mixture in the feed pipe i.e. (M_g + M_l) / (Qg + Q_l), kg/m^3

$v_m$in: velocity of the mixture in inlet nozzle i.e. (Qg + Q_l) / Area of Inlet Nozzle, m/s

The internal nozzle diameter, d1, may be taken equal to that of the feed pipe, but also a momentum criterion (dependent on the inlet device, if any) shall be satisfied:

If

- ρ_m v_m^2 in $\leq$ 1400, no inlet device is used
- $\rho_m v_m^2$, in $\leq$ 2100, half-open pipe is used
- $\rho_m v_m^2$, in $\leq$ 8000, Schoepentoeter is used

Using d1, find area of the inlet nozzle and subsequently the velocity. Calculating the momentum, finalize the inlet device to be used.

Step II**: Vessel Diameter**

Calculate Q^*_{max} (highest value of volumetric gas load factor) to design a vessel large enough to handle the gas flow rate under the most severe condition[12].

$$Q^*_{max}=Q^G_{max}\sqrt{\frac{\rho\_g}{\rho\_l-\rho\_g}}\text{ , m}^3/sec$$

Where:

Q^G_{max}=Highest envisaged gas flow rate and includes a margin for surging, uncertainties in basic data, etc. and is expressed in m3/s. In refineries and chemical plants the design margin ranges typically from 1.15 to 1.25

ρ_g= Density of gas phase, kg/m3

ρ_l= Density of liquid phase, kg/m3

The minimum required vessel cross-sectional area for gas flow, A_G min (m^2), is determined by the following formula[12]:

$$A_g min=Q^*_{max}/\lambda_{max}$$

Where λmax is the maximum allowable gas load factor, which is a measure of the gas handling capacity of the selected separator. It is also referred to as k-factor or Souders-Brown velocity

$$\lambda=(Q_g-A_g)/\sqrt{\frac{\rho\_g}{\rho\_l-\rho\_g}}$$

This is a superficial gas velocity modified with a gas density scaling factor which accounts to a large extent for the effect of operating pressure. In a vertical vessel AG, min is the cross-sectional area of the vessel. However, if a wire mesh is used, it is the cross-sectional area of the wire mesh, which can be much smaller than the vessel's cross sectional area.

For a wiremesh demister, the vessel diameter shall satisfy the gas handling capacity criterion to limit the liquid carry-over:

$$\lambda_{max}=\left(Q^{*}_{max}/\Pi * D^{2}_{min}/4\right) = 0.105 f_{\eta} f_{\Phi}, \text{ m/sec}$$

or

$$\text{D}\geqslant 3.48\sqrt{\frac{Q^{*}_{max}}{f_{\eta} f_{\Phi}}}$$

Where,
D: Vessel diameter, m
f_{η}:Derating factor allowing for the liquid phase viscosity (ηL)
If η_l > 0.001 Pa.s ≥ fη = (0.001 / ηL)0.04
If ηL ⩽ 0.001 Pa.s ≥ fη = 1
f: Derating factor related to the flow parameter at the $faceofthewiremesh, \Phi wm$
If $\Phi wm \leq 0.1 \geq f\Phi = 1/(1+10\Phi wm)$
In practice,Φwm will not exceed 0.1
Φwm: It is a function of the flow parameter of feed entering the vessel and the inlet device.
If a Schoepentoeter is used, $\Phi wm = 0.05$ $\Phi feed$, taking Schoepentoeter efficiency 95%
If a half-open is used,$\Phi wm = 0.2$ Φfeed, taking half-open pipe efficiency 80%

To limit the vapor carry-under, the vessel diameter shall satisfy the liquid de-gassing criterion: Assuming that if bubbles larger than 200 µm in size are able to escape, the carry-under will be negligible, we can state the following criteria based on Stoke's Law:
VL ≤ QL, max / (π D2 / 4) = 2.2 * 10-8 (ρL – ρG), m/s
Or D ≥ 7608$\sqrt{\frac{Q_l max\eta \text{l}}{\text{ρ_l-ρ_g}}}$
The diameter should also satisfy the de-foaming criterion:
$D \geq 95Q0.5L, max\eta L/(\rho L – \rho G)0.14, m$

Step III: Vessel Height

First, the height required for liquid hold-up (h) is determined based on the following criteria :

1. The level gauge (LG) span is supposed to extend 0.15 m above the BTL.
2. LZA(LL), i.e. low level trip should be at least 0.1 m above the LG level
3. LA(L), i.e. low level pre-alarm should be at least 0.10 m above LZA(LL) or, should be located so that there is sufficient liquid hold-up time between the two levels for operator intervention (this is to be specified, but is typically 1-2 min).
4. LA(H) i.e. high level pre alarm should be placed such that the minimum distance between LA(H) and LA(L) 0s 0.35 m or, it can be located so that there is sufficient hold-up time between the two levels for control purposes. The hold-up time is 3 min for total product outflow for separator trains, 5 min for total product outflow for columns to furnaces or other columns and 10 min for total product outflow for feed surge drums.

5. The slugs are expected, they are accommodated between NL and LA (H). if the volume of the slug to be expected is not known, the volume is taken as 2 s to 5 s of flow with maximum feed (gas + liquid) velocity. So the total volume to be provided between LA(L) and LA(H) is the sum of the required control volume and the volume of the anticipated slug.
6. LZA(HH) i.e. high level trip should be at least 0.10 m above LA(H) or, should be located so that there is sufficient liquid hold-up time between the two levels for operator intervention (this is to be specified, but is typically 1-2 min).
7. If the liquid has a foaming tendency, the distance between LA (H) and LZA (HH) is increased by further 0.25 m.

The total tangent-to-tangent vessel height (H, m) is[13]:

$$H = h + X_1 + X_2 + X_3 + t_{wm} + X_4$$

Where:
t_{wm}: thickness of demister mat, usually 0.15 m
X_1: 0.05 D with a minimum of 0.15 m (Schoepentoeter),0.3 D with a minimum of 0.3 m (Half-open pipe)
X_2:$d_1 + 0.02m$ (Schoepentoeter, and d_1 being internal diameter of inlet nozzle) d_1 (m) (Half-open pipe)
X_3: d_1 with a minimum of 0.3 m (Schoepentoeter) ,0.45 D with a minimum of 0.9 m (Half-open pipe)
X_4: 0.15 D with a minimum of 0.15 m (Schoepentoeter), 0.15 D with a minimum of 0.15 m (Half-open pipe)

Step IV: Nozzle Sizing

The sizing of the nozzles is based on the maximum flow rates, including the appropriate design margin.

The feed inlet nozzle diameter, d1, has been explained above in the inlet device selection section.

The gas outlet nozzle diameter, d2, is normally taken to be equal to that of the outlet pipe, but the following criterion should be satisfied:

$$\rho_G \ V_G^2, \text{out} \leq 4500, \text{Pa}$$

The liquid outlet nozzle diameter, d_3, is selected such that the liquid velocity does not exceed 1 m/s. The minimum diameter is 0.05 m.

Step V: Pressure Drop

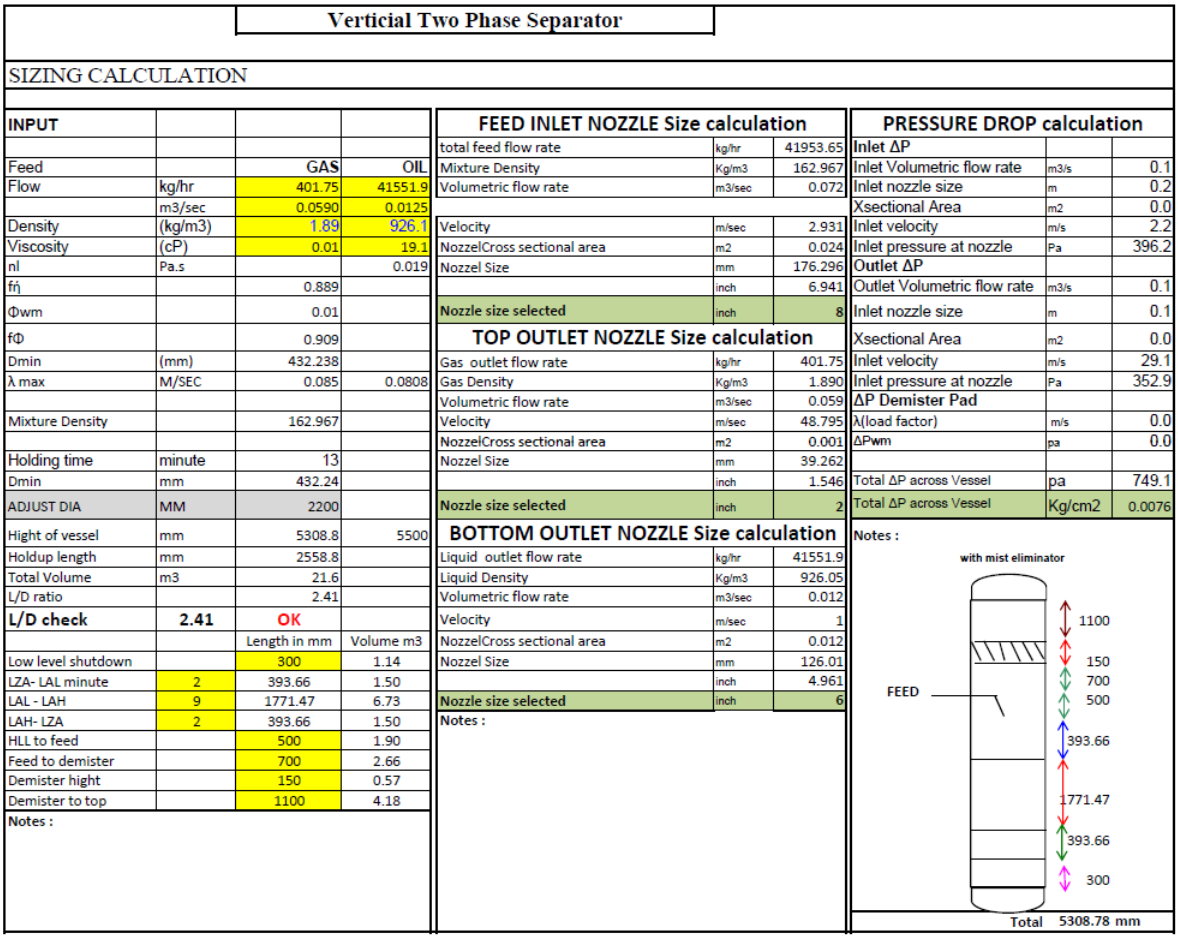
The pressure differential between inlet and vapor outlet is:

$$P_{in} - P_{out} = 0.5 \ \rho_m \ V_m^2, \text{in} + 0.22 \ \rho_G \ V_G^2, \text{out} + \Delta p_{wm} + \Delta p_{sch}, \text{Pa}$$

Where,
ρ_m: Mean density of the mixture in the feed pipe, kg/m^3
V_m^2, in: Velocity of the mixture in inlet nozzle, m/s
ρ_G: Density of gas phase, kg/m^3
V_G^2, out: Velocity of gas in gas outlet nozzle, m/s
Δp_{wm}: Pressure drop across the mistmat, Pa = 200 * (ρL – ρG) λ^2 t_{wm}
t_{wm}: Thickness of demister mat, usually 0.15 m
λ: Souders Brown Velocity, m/s = (QG / AG) / √(ρG/(ρL- ρG))
Δp_{sch}: Pressure drop across Schoepentoeter (if present), Pa = 0.08 * ρ_m $v^2{}_m$, in

3.2.7 Vertical Two Phase Separator Calculation

Table 3.2: Vertical Two Phase Separator Calculation



Verticial Two Phase Separator

SIZING CALCULATION

INPUT			
Feed		**GAS**	**OIL**
Flow	kg/hr	401.75	41551.9
	m3/sec	0.0590	0.0125
Density	(kg/m3)	1.89	926.1
Viscosity	(cP)	0.01	19.1
nl	Pa.s		0.019
fή		0.889	
Φwm		0.01	
fΦ		0.909	
Dmin	(mm)	432.238	
λ max	M/SEC	0.085	0.0808
Mixture Density		162.967	
Holding time	minute	13	
Dmin	mm	432.24	
ADJUST DIA	MM	2200	
Hight of vessel	mm	5308.8	5500
Holdup length	mm	2558.8	
Total Volume	m3	21.6	
L/D ratio		2.41	
L/D check	**2.41**	OK	
		Length in mm	Volume m3
Low level shutdown		300	1.14
LZA- LAL minute	2	393.66	1.50
LAL - LAH	9	1771.47	6.73
LAH- LZA	2	393.66	1.50
HLL to feed		500	1.90
Feed to demister		700	2.66
Demister hight		150	0.57
Demister to top		1100	4.18

Notes :

FEED INLET NOZZLE Size calculation		
total feed flow rate	kg/hr	41953.65
Mixture Density	Kg/m3	162.967
Volumetric flow rate	m3/sec	0.072
Velocity	m/sec	2.931
NozzelCross sectional area	m2	0.024
Nozzel Size	mm	176.296
	inch	6.941
Nozzle size selected	inch	8
TOP OUTLET NOZZLE Size calculation		
Gas outlet flow rate	kg/hr	401.75
Gas Density	Kg/m3	1.890
Volumetric flow rate	m3/sec	0.059
Velocity	m/sec	48.795
NozzelCross sectional area	m2	0.001
Nozzel Size	mm	39.262
	inch	1.546
Nozzle size selected	inch	2
BOTTOM OUTLET NOZZLE Size calculation		
Liquid outlet flow rate	kg/hr	41551.9
Liquid Density	Kg/m3	926.05
Volumetric flow rate	m3/sec	0.012
Velocity	m/sec	1
NozzelCross sectional area	m2	0.012
Nozzel Size	mm	126.01
	inch	4.961
Nozzle size selected	inch	6

Notes :

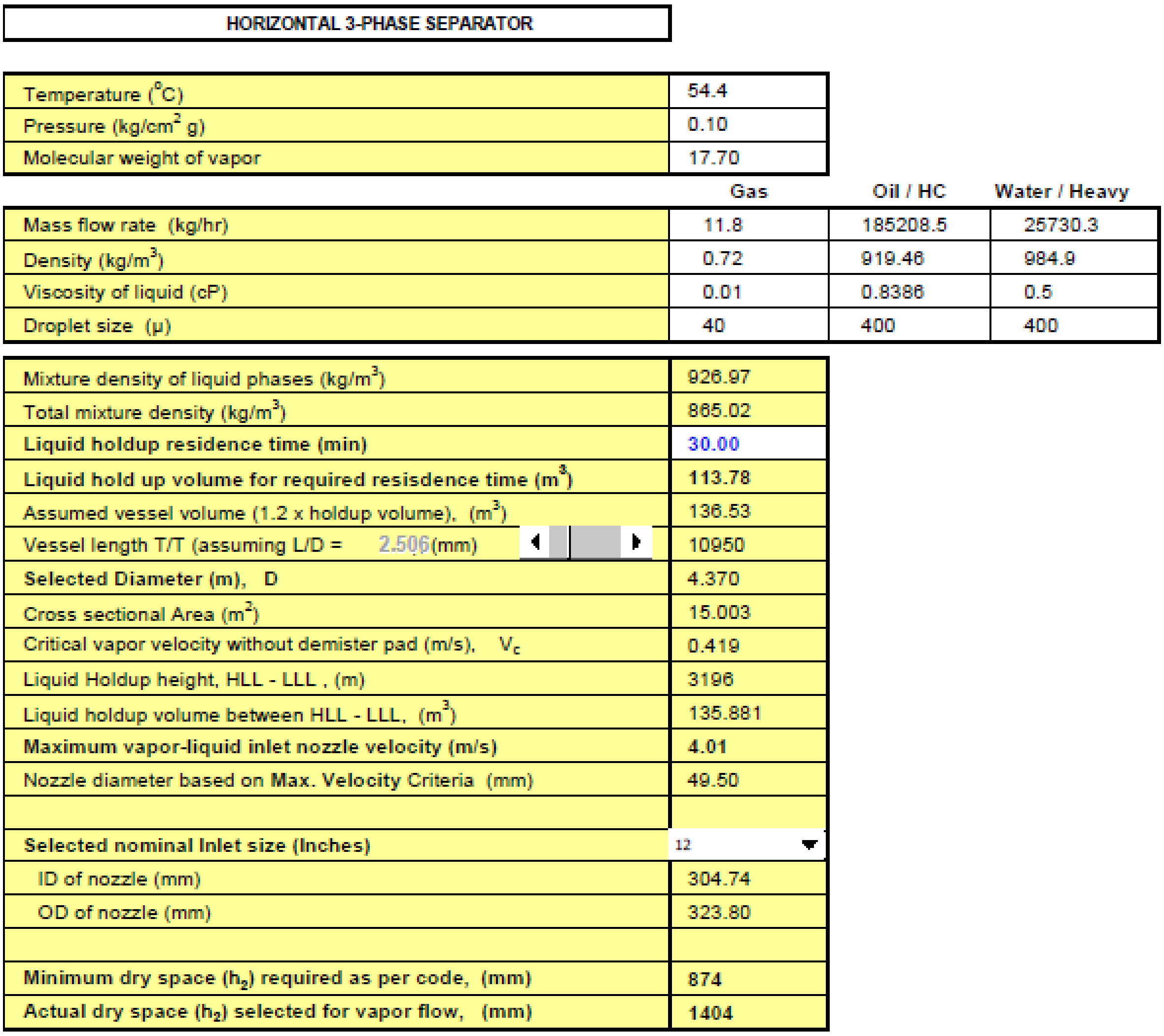
PRESSURE DROP calculation		
Inlet ΔP		
Inlet Volumetric flow rate	m3/s	0.1
Inlet nozzle size	m	0.2
Xsectional Area	m2	0.0
Inlet velocity	m/s	2.2
Inlet pressure at nozzle	Pa	396.2
Outlet ΔP		
Outlet Volumetric flow rate	m3/s	0.1
Inlet nozzle size	m	0.1
Xsectional Area	m2	0.0
Inlet velocity	m/s	29.1
Inlet pressure at nozzle	Pa	352.9
ΔP Demister Pad		
λ(load factor)	m/s	0.0
ΔPwm	pa	0.0
Total ΔP across Vessel	pa	749.1
Total ΔP across Vessel	Kg/cm2	0.0076

Notes :

with mist eliminator

Total 5308.78 mm

3.2.8 Horizontal three phase separator

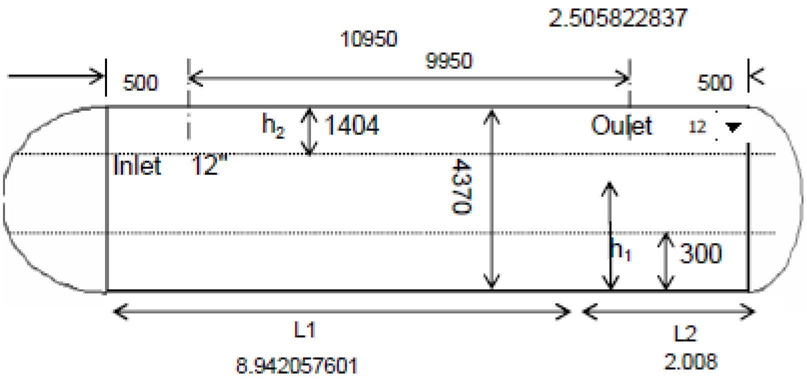


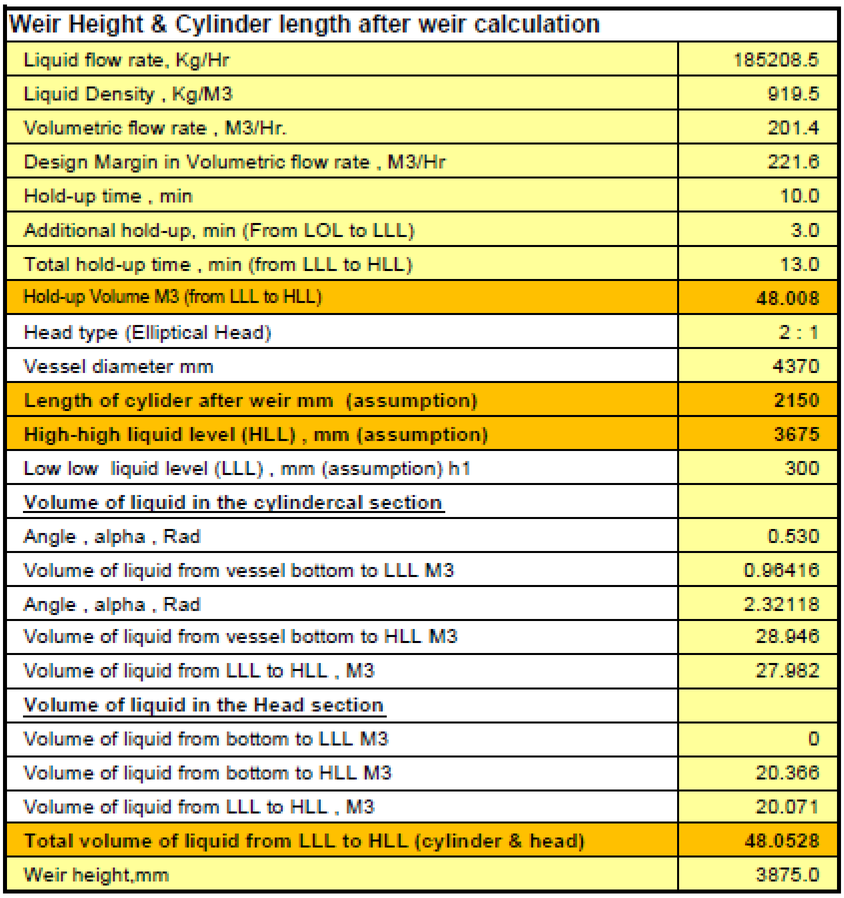
HORIZONTAL 3-PHASE SEPARATOR

Temperature (°C)	54.4
Pressure (kg/cm^2 g)	0.10
Molecular weight of vapor	17.70

	Gas	Oil / HC	Water / Heavy
Mass flow rate (kg/hr)	11.8	185208.5	25730.3
Density (kg/m^3)	0.72	919.46	984.9
Viscosity of liquid (cP)	0.01	0.8386	0.5
Droplet size (μ)	40	400	400

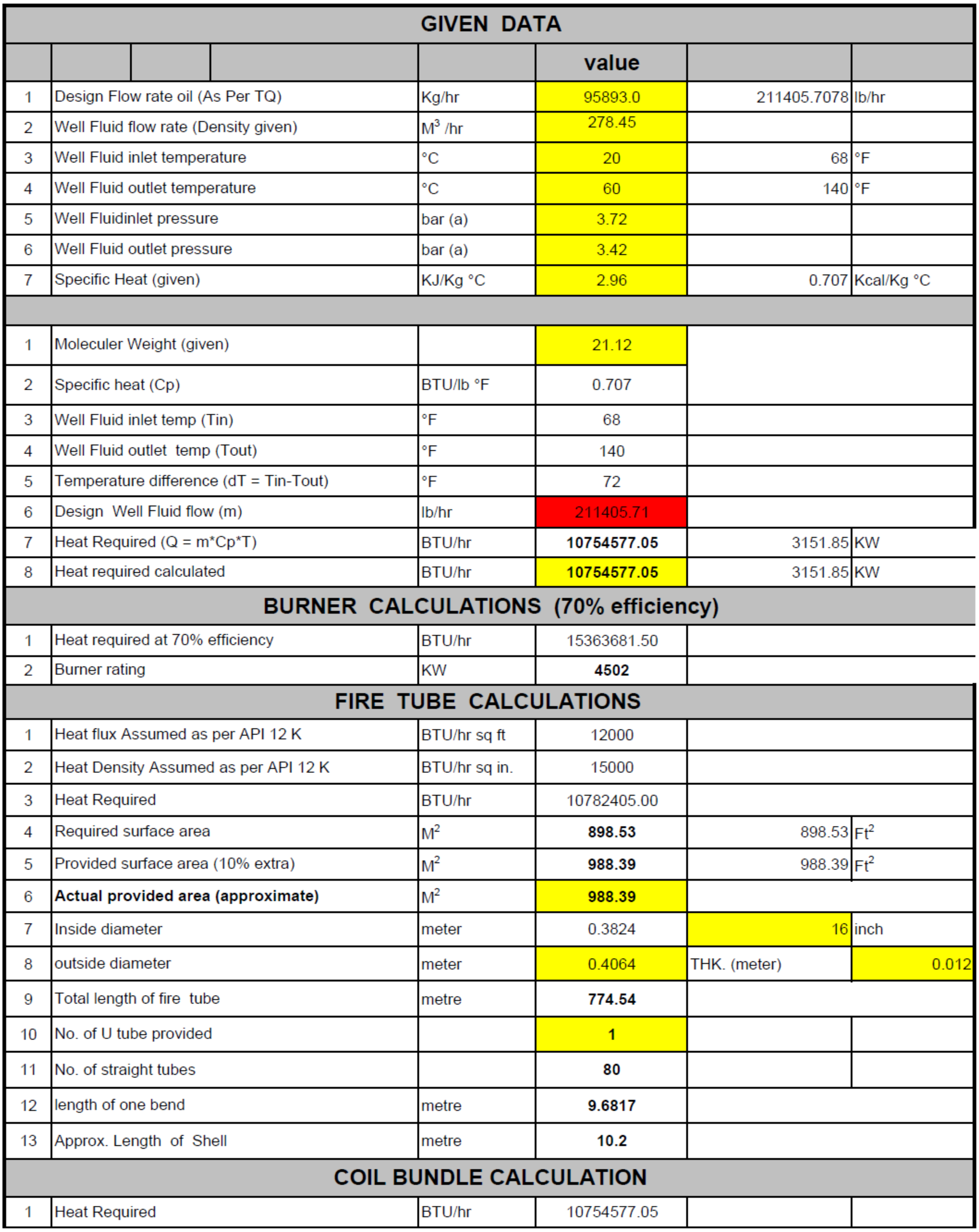
Mixture density of liquid phases (kg/m^3)	926.97
Total mixture density (kg/m^3)	865.02
Liquid holdup residence time (min)	30.00
Liquid hold up volume for required resisdence time (m^3)	**113.78**
Assumed vessel volume (1.2 x holdup volume), (m^3)	136.53
Vessel length T/T (assuming L/D = 2.506 (mm)	10950
Selected Diameter (m), D	4.370
Cross sectional Area (m^2)	15.003
Critical vapor velocity without demister pad (m/s), V_c	0.419
Liquid Holdup height, HLL - LLL , (m)	3196
Liquid holdup volume between HLL - LLL, (m^3)	135.881
Maximum vapor-liquid inlet nozzle velocity (m/s)	**4.01**
Nozzle diameter based on Max. Velocity Criteria (mm)	49.50
Selected nominal Inlet size (Inches)	12
ID of nozzle (mm)	304.74
OD of nozzle (mm)	323.80
Minimum dry space (h_2) required as per code, (mm)	**874**
Actual dry space (h_2) selected for vapor flow, (mm)	**1404**



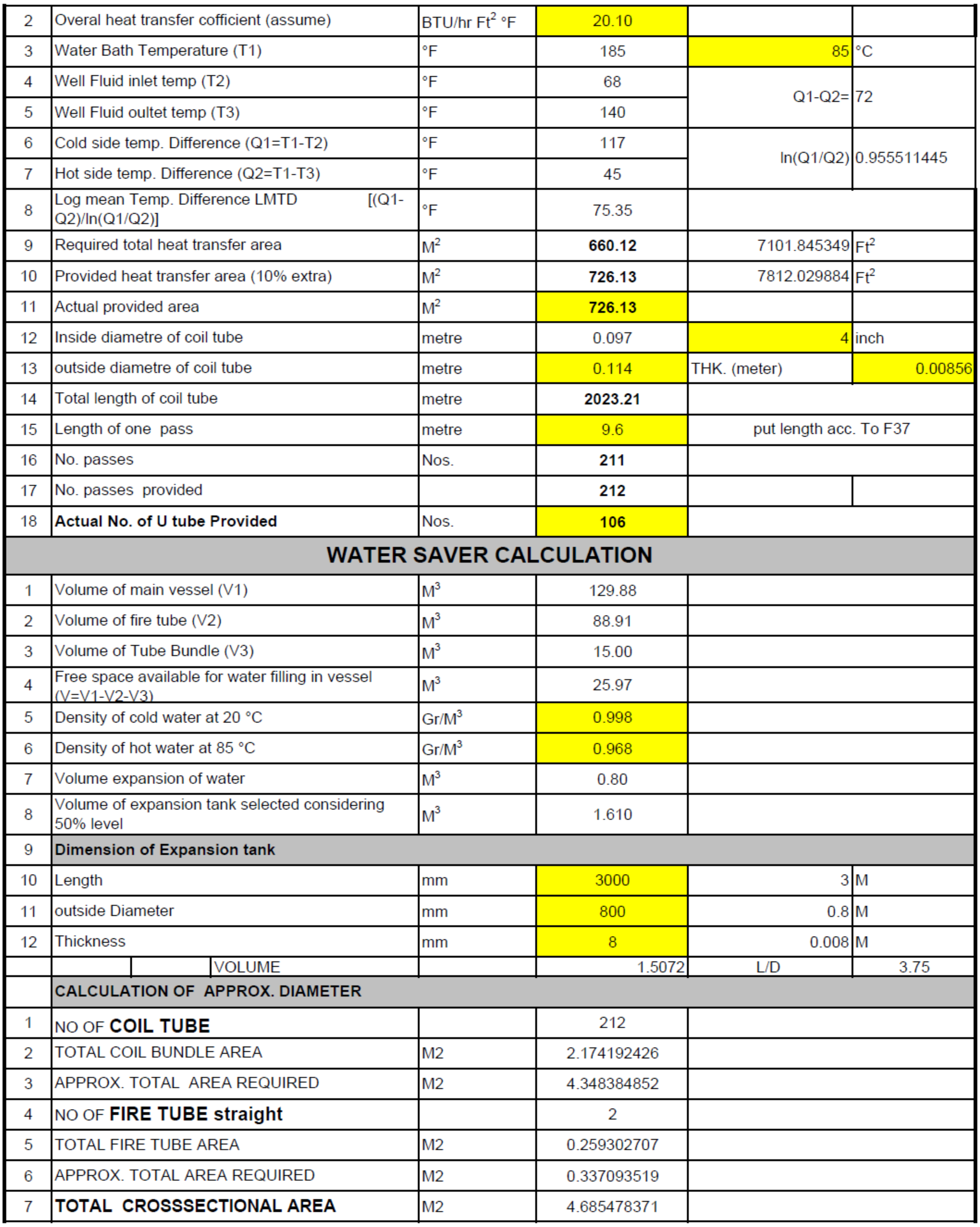


Weir Height & Cylinder length after weir calculation	
Liquid flow rate, Kg/Hr	185208.5
Liquid Density , Kg/M3	919.5
Volumetric flow rate , M3/Hr.	201.4
Design Margin in Volumetric flow rate , M3/Hr	221.6
Hold-up time , min	10.0
Additional hold-up, min (From LOL to LLL)	3.0
Total hold-up time , min (from LLL to HLL)	13.0
Hold-up Volume M3 (from LLL to HLL)	**48.008**
Head type (Elliptical Head)	2 : 1
Vessel diameter mm	4370
Length of cylider after weir mm (assumption)	**2150**
High-high liquid level (HLL) , mm (assumption)	**3675**
Low low liquid level (LLL) , mm (assumption) h1	300
Volume of liquid in the cylindercal section	
Angle , alpha , Rad	0.530
Volume of liquid from vessel bottom to LLL M3	0.96416
Angle , alpha , Rad	2.32118
Volume of liquid from vessel bottom to HLL M3	28.946
Volume of liquid from LLL to HLL , M3	27.982
Volume of liquid in the Head section	
Volume of liquid from bottom to LLL M3	0
Volume of liquid from bottom to HLL M3	20.366
Volume of liquid from LLL to HLL , M3	20.071
Total volume of liquid from LLL to HLL (cylinder & head)	**48.0528**
Weir height,mm	3875.0

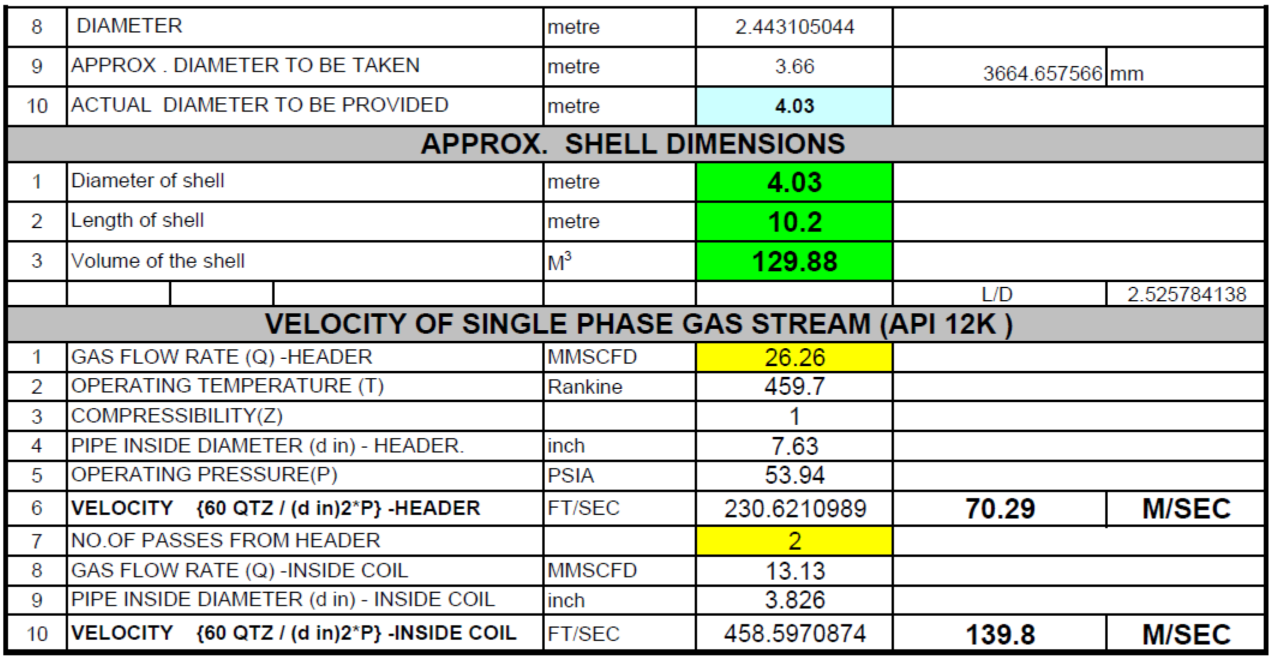
3.2.9 Oil Bath Heater Design Calculation



GIVEN DATA					
			value		
1	Design Flow rate oil (As Per TQ)	Kg/hr	95893.0	211405.7078	lb/hr
2	Well Fluid flow rate (Density given)	M^3 /hr	278.45		
3	Well Fluid inlet temperature	°C	20	68	°F
4	Well Fluid outlet temperature	°C	60	140	°F
5	Well Fluidinlet pressure	bar (a)	3.72		
6	Well Fluid outlet pressure	bar (a)	3.42		
7	Specific Heat (given)	KJ/Kg °C	2.96	0.707	Kcal/Kg °C
1	Moleculer Weight (given)		21.12		
2	Specific heat (Cp)	BTU/lb °F	0.707		
3	Well Fluid inlet temp (Tin)	°F	68		
4	Well Fluid outlet temp (Tout)	°F	140		
5	Temperature difference (dT = Tin-Tout)	°F	72		
6	Design Well Fluid flow (m)	lb/hr	211405.71		
7	Heat Required (Q = m*Cp*T)	BTU/hr	**10754577.05**	3151.85	KW
8	Heat required calculated	BTU/hr	**10754577.05**	3151.85	KW
BURNER CALCULATIONS (70% efficiency)					
1	Heat required at 70% efficiency	BTU/hr	15363681.50		
2	Burner rating	KW	**4502**		
FIRE TUBE CALCULATIONS					
1	Heat flux Assumed as per API 12 K	BTU/hr sq ft	12000		
2	Heat Density Assumed as per API 12 K	BTU/hr sq in.	15000		
3	Heat Required	BTU/hr	10782405.00		
4	Required surface area	M^2	**898.53**	898.53	Ft^2
5	Provided surface area (10% extra)	M^2	**988.39**	988.39	Ft^2
6	**Actual provided area (approximate)**	M^2	**988.39**		
7	Inside diameter	meter	0.3824	16	inch
8	outside diameter	meter	0.4064	THK. (meter)	0.012
9	Total length of fire tube	metre	**774.54**		
10	No. of U tube provided		**1**		
11	No. of straight tubes		**80**		
12	length of one bend	metre	**9.6817**		
13	Approx. Length of Shell	metre	**10.2**		
COIL BUNDLE CALCULATION					
1	Heat Required	BTU/hr	10754577.05		



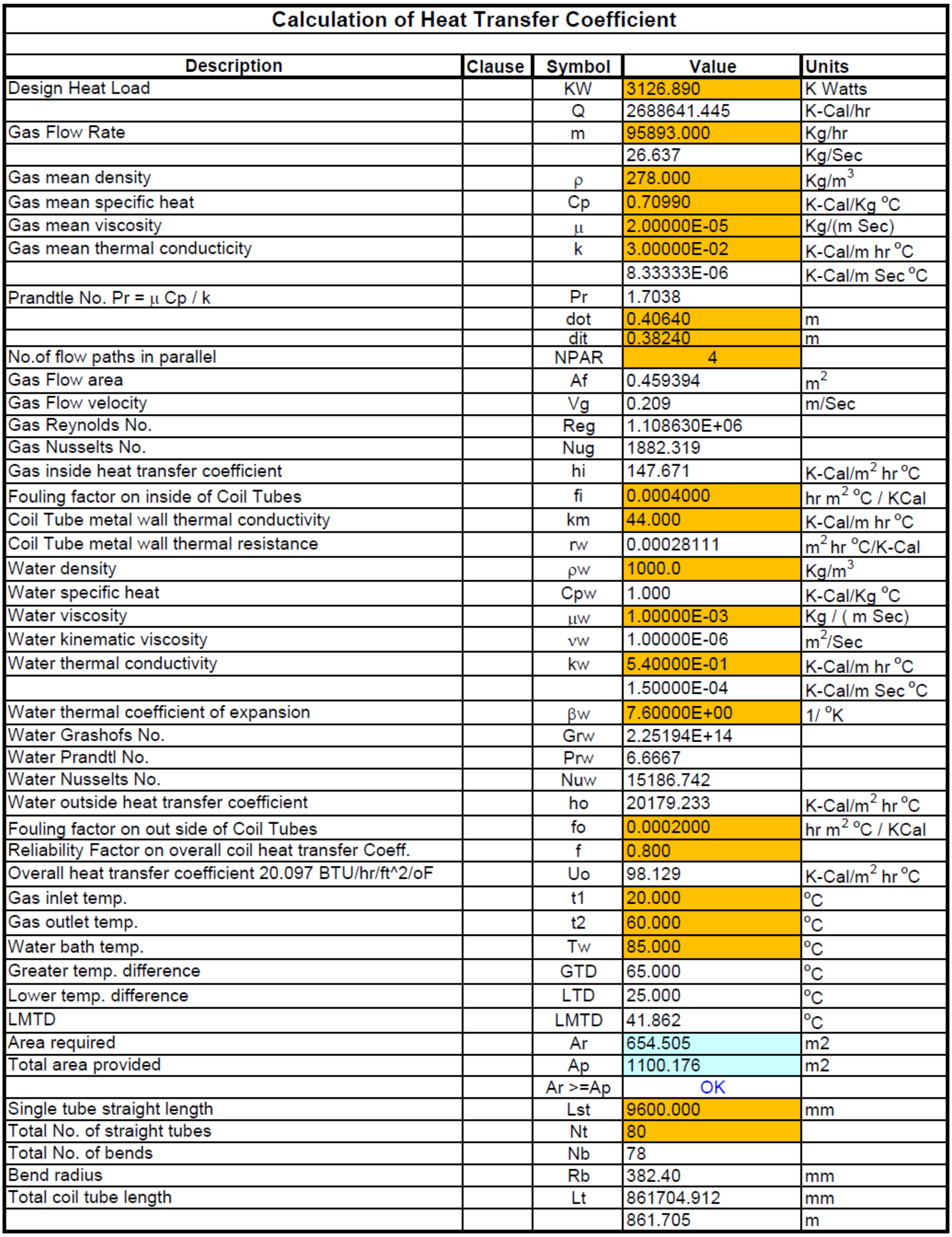
2	Overal heat transfer cofficient (assume)	BTU/hr Ft^2 °F	20.10		
3	Water Bath Temperature (T1)	°F	185	85	°C
4	Well Fluid inlet temp (T2)	°F	68	Q1-Q2=	72
5	Well Fluid oultet temp (T3)	°F	140		
6	Cold side temp. Difference (Q1=T1-T2)	°F	117	ln(Q1/Q2)	0.955511445
7	Hot side temp. Difference (Q2=T1-T3)	°F	45		
8	Log mean Temp. Difference LMTD [(Q1-Q2)/ln(Q1/Q2)]	°F	75.35		
9	Required total heat transfer area	M^2	**660.12**	7101.845349	Ft^2
10	Provided heat transfer area (10% extra)	M^2	**726.13**	7812.029884	Ft^2
11	Actual provided area	M^2	**726.13**		
12	Inside diametre of coil tube	metre	0.097	4	inch
13	outside diametre of coil tube	metre	0.114	THK. (meter)	0.00856
14	Total length of coil tube	metre	**2023.21**		
15	Length of one pass	metre	9.6	put length acc. To F37	
16	No. passes	Nos.	**211**		
17	No. passes provided		**212**		
18	**Actual No. of U tube Provided**	Nos.	**106**		
	WATER SAVER CALCULATION				
1	Volume of main vessel (V1)	M^3	129.88		
2	Volume of fire tube (V2)	M^3	88.91		
3	Volume of Tube Bundle (V3)	M^3	15.00		
4	Free space available for water filling in vessel (V=V1-V2-V3)	M^3	25.97		
5	Density of cold water at 20 °C	Gr/M^3	0.998		
6	Density of hot water at 85 °C	Gr/M^3	0.968		
7	Volume expansion of water	M^3	0.80		
8	Volume of expansion tank selected considering 50% level	M^3	1.610		
9	**Dimension of Expansion tank**				
10	Length	mm	3000	3	M
11	outside Diameter	mm	800	0.8	M
12	Thickness	mm	8	0.008	M
	VOLUME		1.5072	L/D	3.75
	CALCULATION OF APPROX. DIAMETER				
1	NO OF **COIL TUBE**		212		
2	TOTAL COIL BUNDLE AREA	M2	2.174192426		
3	APPROX. TOTAL AREA REQUIRED	M2	4.348384852		
4	NO OF **FIRE TUBE straight**		2		
5	TOTAL FIRE TUBE AREA	M2	0.259302707		
6	APPROX. TOTAL AREA REQUIRED	M2	0.337093519		
7	**TOTAL CROSSSECTIONAL AREA**	M2	4.685478371		



8	DIAMETER	metre	2.443105044		
9	APPROX . DIAMETER TO BE TAKEN	metre	3.66	3664.657566	mm
10	ACTUAL DIAMETER TO BE PROVIDED	metre	**4.03**		
APPROX. SHELL DIMENSIONS					
1	Diameter of shell	metre	**4.03**		
2	Length of shell	metre	**10.2**		
3	Volume of the shell	M^3	**129.88**		
				L/D	2.525784138
VELOCITY OF SINGLE PHASE GAS STREAM (API 12K)					
1	GAS FLOW RATE (Q) -HEADER	MMSCFD	26.26		
2	OPERATING TEMPERATURE (T)	Rankine	459.7		
3	COMPRESSIBILITY(Z)		1		
4	PIPE INSIDE DIAMETER (d in) - HEADER.	inch	7.63		
5	OPERATING PRESSURE(P)	PSIA	53.94		
6	**VELOCITY {60 QTZ / (d in)2*P} -HEADER**	FT/SEC	230.6210989	**70.29**	**M/SEC**
7	NO.OF PASSES FROM HEADER		2		
8	GAS FLOW RATE (Q) -INSIDE COIL	MMSCFD	13.13		
9	PIPE INSIDE DIAMETER (d in) - INSIDE COIL	inch	3.826		
10	**VELOCITY {60 QTZ / (d in)2*P} -INSIDE COIL**	FT/SEC	458.5970874	**139.8**	**M/SEC**

Table 3.3: Horizontal 3-Phase Separator

Table 3.4: Heat Transfer Calculation



Calculation of Heat Transfer Coefficient

Description	Clause	Symbol	Value	Units
Design Heat Load		KW	3126.890	K Watts
		Q	2688641.445	K-Cal/hr
Gas Flow Rate		m	95893.000	Kg/hr
			26.637	Kg/Sec
Gas mean density		ρ	278.000	Kg/m^3
Gas mean specific heat		Cp	0.70990	K-Cal/Kg °C
Gas mean viscosity		μ	2.00000E-05	Kg/(m Sec)
Gas mean thermal conducticity		k	3.00000E-02	K-Cal/m hr °C
			8.33333E-06	K-Cal/m Sec °C
Prandtle No. Pr = μ Cp / k		Pr	1.7038	
		dot	0.40640	m
		dit	0.38240	m
No.of flow paths in parallel		NPAR	4	
Gas Flow area		Af	0.459394	m^2
Gas Flow velocity		Vg	0.209	m/Sec
Gas Reynolds No.		Reg	1.108630E+06	
Gas Nusselts No.		Nug	1882.319	
Gas inside heat transfer coefficient		hi	147.671	K-Cal/m^2 hr °C
Fouling factor on inside of Coil Tubes		fi	0.0004000	hr m^2 °C / KCal
Coil Tube metal wall thermal conductivity		km	44.000	K-Cal/m hr °C
Coil Tube metal wall thermal resistance		rw	0.00028111	m^2 hr °C/K-Cal
Water density		ρw	1000.0	Kg/m^3
Water specific heat		Cpw	1.000	K-Cal/Kg °C
Water viscosity		μw	1.00000E-03	Kg / (m Sec)
Water kinematic viscosity		νw	1.00000E-06	m^2/Sec
Water thermal conductivity		kw	5.40000E-01	K-Cal/m hr °C
			1.50000E-04	K-Cal/m Sec °C
Water thermal coefficient of expansion		βw	7.60000E+00	1/ °K
Water Grashofs No.		Grw	2.25194E+14	
Water Prandtl No.		Prw	6.6667	
Water Nusselts No.		Nuw	15186.742	
Water outside heat transfer coefficient		ho	20179.233	K-Cal/m^2 hr °C
Fouling factor on out side of Coil Tubes		fo	0.0002000	hr m^2 °C / KCal
Reliability Factor on overall coil heat transfer Coeff.		f	0.800	
Overall heat transfer coefficient 20.097 BTU/hr/ft^2/oF		Uo	98.129	K-Cal/m^2 hr °C
Gas inlet temp.		t1	20.000	°C
Gas outlet temp.		t2	60.000	°C
Water bath temp.		Tw	85.000	°C
Greater temp. difference		GTD	65.000	°C
Lower temp. difference		LTD	25.000	°C
LMTD		LMTD	41.862	°C
Area required		Ar	654.505	m2
Total area provided		Ap	1100.176	m2
		Ar >=Ap	OK	
Single tube straight length		Lst	9600.000	mm
Total No. of straight tubes		Nt	80	
Total No. of bends		Nb	78	
Bend radius		Rb	382.40	mm
Total coil tube length		Lt	861704.912	mm
			861.705	m

Chapter 4

The glycol dehydration process

The dehydration process can be divided into two parts, gas dehydration and glycol regeneration. In dehydration water is removed from the gas using glycol and in regeneration water is removed from the glycol, before it can be reused for dehydration.

Dehydration: Dehydration always consists of an inlet scrubber and a contactor. Sometime it might be preferable to lower the gas inlet temperature before the dehydration, so an inlet cooler might also be used.

4.1 Process description

The process is described by the equipment used in the glycol plant.

4.1.1 Inlet cooler

An inlet cooler may be used because dehydration is more efficient at low temperatures. Another benefit of inlet cooling is that some water (and hydrocarbons) in the gas will condense, and be removed in the inlet scrubber, instead of in the contactor. An inlet cooler is used when the inlet gas temperature is higher than the desired temperature in the contactor. It is also a helpful tool in simulation if the temperature in the contactor needs to be optimized.

4.1.2 Feed gas KOD

The inlet feed gas KOD removes free liquid and liquid droplets in the gas, both water and hydrocarbons. Removing liquid water in the KOD decreases the amount of water that has to be removed in the contractor. This decreases the size of the contactor and the glycol needed in it, to reach the required conditions for the outlet gas. Liquid hydrocarbons are also a problem in the contactor because they increase the glycols tendency to foam, thereby decreasing the contactors efficiency and increasing the glycol loss in the contractor and from the regeneration system. Another problem is that hydrocarbons can be accumulated in the glycol polluting it and thereby decreasing the dehydration efficiency.

4.1.3 Contactor

The contactor is the absorption column where the gas is dried by the glycol. The lean glycol enters at the top of the contactor while the rich glycol is collected at the bottom of the contactor and sent to regeneration. The wet gas enters the contactor at the bottom, while the dry gas leaves at the top.

The required water dew-point of the dry gas dictates the lean glycol temperature and purity. This is illustrated in Figure 2-2. The glycol temperature into the contactor must be 35 to 45 °C higher than the gas entering the contactor to minimize hydrocarbon condensation into the glycol.

At contactor temperatures below 10 °C TEG becomes too viscous, thus reducing the column efficiency. The contactor temperature may be as high as 66 °C, but glycol vaporization loss is often deemed unacceptably high above 50 °C.

The glycol flow into the contactor is dictated by the water content in the gas and numbers of trays in the column. A usual glycol flow is 16 to 14 liters Lean TEG per kg water in the gas. Contactor columns with nine trays usually operate with 16 to 14 liters TEG/kg Water.

4.1.4 Flash valve

After the contactor column the pressure is reduced to the regeneration pressure by a flash valve. The pressure drop over this valve depends on the pressure in the contactor and the pressure loss in the pipes and equipment until the regeneration column. Two places in the system unwanted gas is vented off the system, in the flash separator and the regenerator. To prevent blowback the pressure in these units must be higher than where they vent to. The slightly higher pressure also acts as a propellant in transporting the gas from the dehydration system.

4.1.5 Flash separator

It is a good idea to install a separator after the flash valve. Because of the decreased pressure hydrocarbons absorbed in the glycol will be released. Without a separator the gas in the glycol will be released together with water in the regenerator. In the regenerator the water vapour is usually just vented to the atmosphere, thus increasing the plants emission of hydrocarbons. With a flash separator the hydrocarbon rich gas, can be used as process gas in the plant. The pressure in the flash separator must be above the pressure in the system that the gas is vented too; the separator pressure will therefore differ between plants.

4.1.6 Filters

Filters are only necessary if there is a problem with solid particles or liquid hydrocarbons in the glycol. Solid particles in the glycol accumulate, increasing the wear on the equipment and can create plugs in heat exchangers. Solid particles can easily be removed with sock filters, which can be made of cloth fabrics, paper or fiberglass. Liquid hydrocarbons like condensate and BTEX can be removed from the glycol by activated carbon filters.

4.1.7 Heat exchangers

The numbers of heat exchangers varies with the design of the process plant. Because of the large temperature difference between the contactor and regenerator column, rich glycol needs to be heated while lean glycol must be cooled. With proper design of heat exchangers between the rich and lean glycol most of the energy can be conserved.

Rich glycol may be heated before and/or after the flash separation. Heating before the flash separator increases hydrocarbon recovery along with glycol loss. Heating before the flash separator is preferable if hydrocarbon contents in the rich glycol after the separation are too high.

Besides the heat exchangers the glycol is heated in the regenerator boiler. The lean glycol temperature may also need to be adjusted before it enters the contactor, this can be done with the dry gas or a cooler.

4.1.8 Regenerator

The regenerator is a distillation column, where glycol and water is separated. The rich glycol is preheated in heat exchangers before it is feed to the regenerator column.

At the top of the column is a partly condenser, this provide reflux thus improving the separation between water and glycol. The condenser also minimizes glycol loss from the regenerator. The remaining water vapour leaves the condenser and is vented to the atmosphere. The temperature in the condenser is given as 94°C .

The energy required to separate glycol and water is supplied by the reboiler at the regenerator column. The reboiler temperature is dictated by the glycol used for the dehydration as described in Table 2-1. For TEG the recommended maximum temperature in the reboiler is 204 °C. The Lean glycol is taken from the reboiler and is transferred to a storage tank before it is recycled or is recycled directly from the reboiler.

The pressure in the regeneration system is just above atmospheric pressure, this is to insure that no air can enter the system from the atmospheric vent. The operating conditions for the regenerator influence the purity of glycol. At 204 °C TEG yields a lean glycol concentration of 99.7 % wt .

4.1.9 Glycol storage tank

This is an optional instalment that ensures a constant glycol flow to the contactor column. Because there will be a loss of glycol in the dehydration system, a storage tank can act a buffer to prevent insufficient glycol flow, and also be used to measure the glycol contents in the system.

4.1.10 Glycol circulation pump

Because of the pressure difference between the regenerator and the contactor, the glycol pressure needs to be increased. This is done with the glycol regeneration pump. The glycol is cooled below 80 °C before pumping to protect the pump.

4.2 Process plant

A possible design of a dehydration plant is given in Figure 4-1.

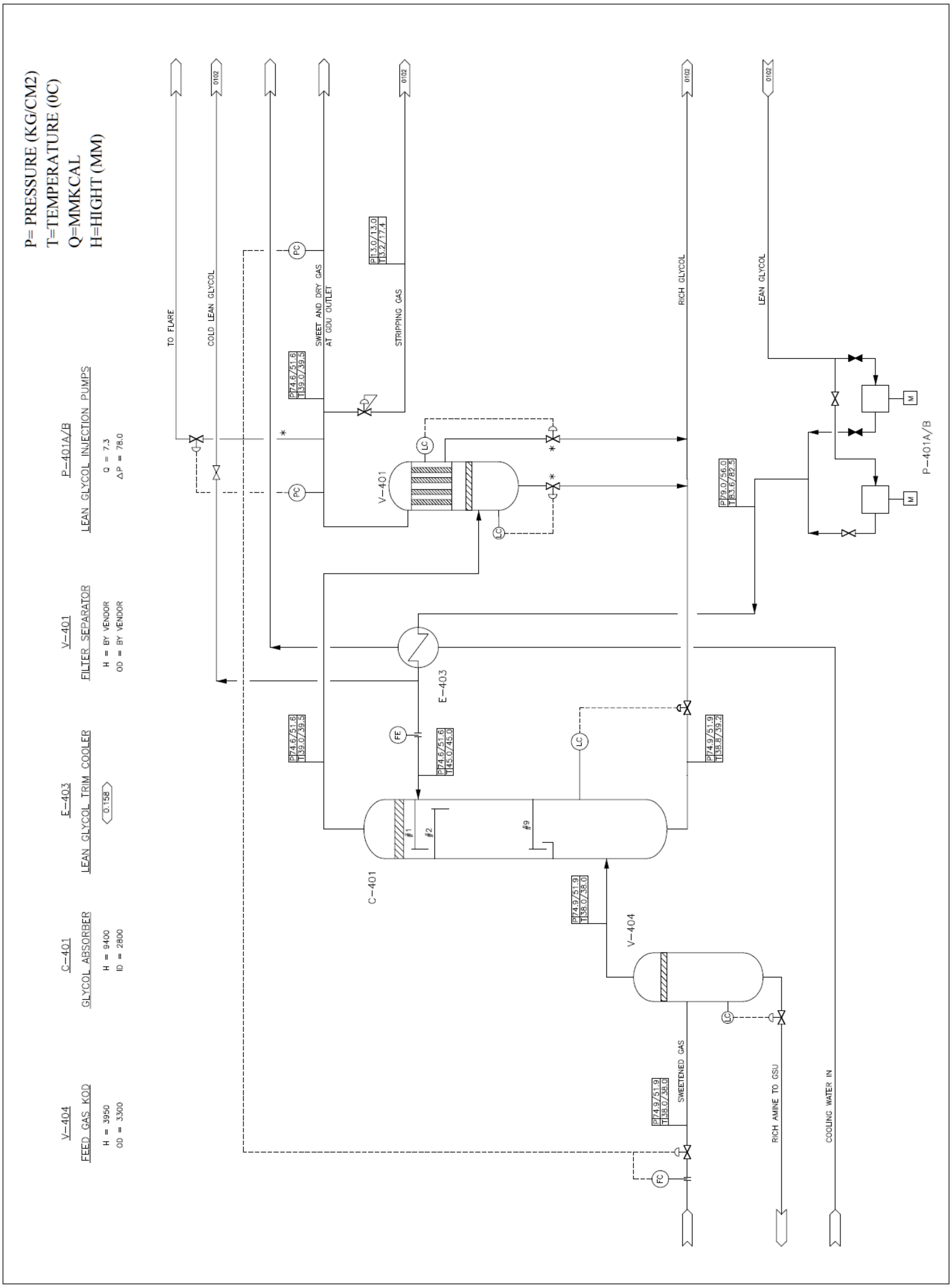


Figure 4.1: Gas Dehydration Plant

The design in Figure 4-1 incorporates most of the units described in section 4-1.

4.3 Glycol Regeneration

A possible design of glycol regeneration plant is given in Figure 4.2

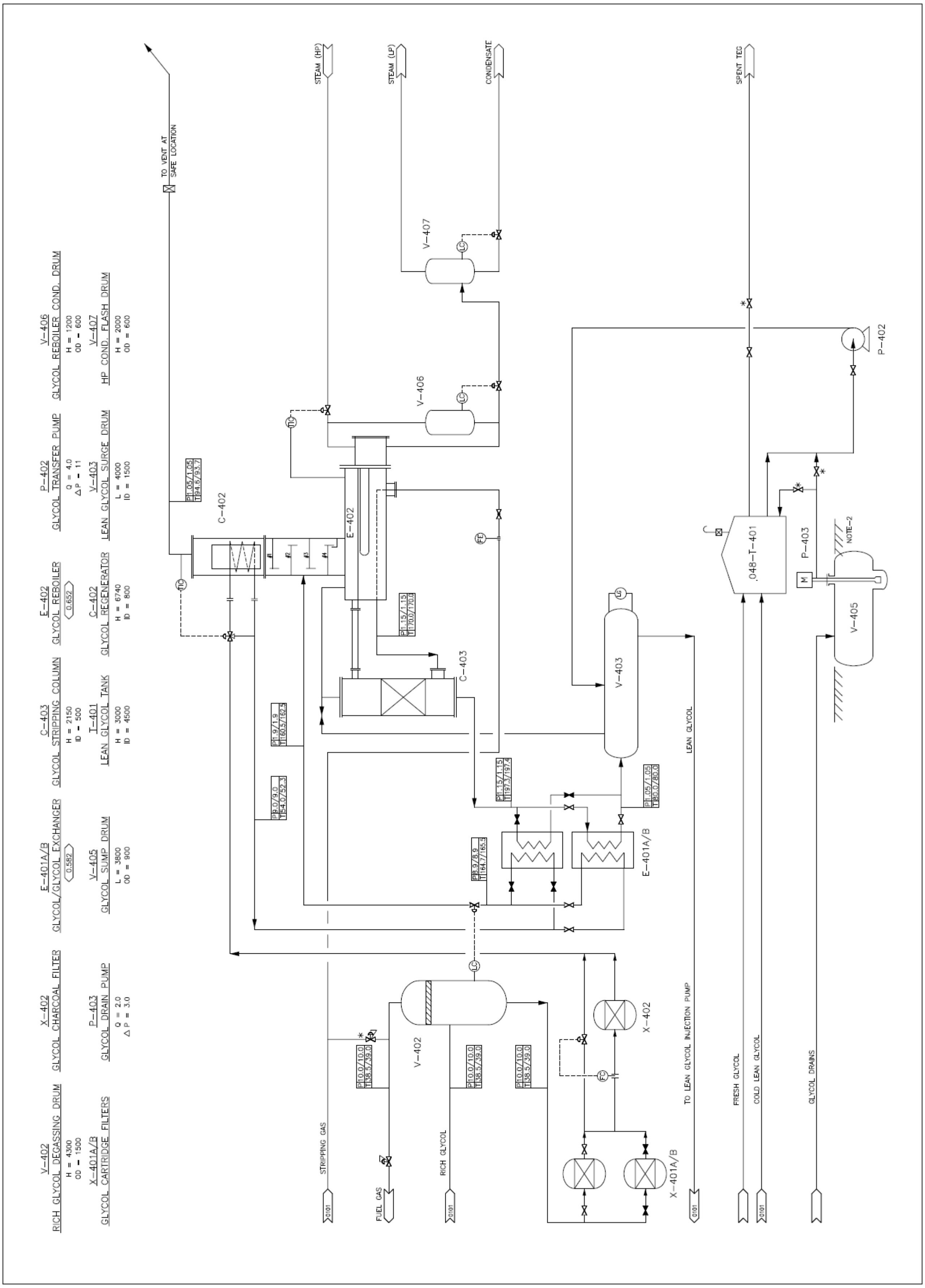


Figure 4.2: Glycol regeneration plant

The main function in the glycol regeneration system can be divided into three:

1. Achieve the optimal pressure and temperature conditions for regeneration of the rich glycol.
2. Glycol regeneration.
3. Readjust glycol temperature and pressure for optimal dehydration conditions in the contactor.

The degassed glycol leaving (v-402) enters then the regeneration section where it is frist filtered. Tryethylene glycol will not exhibit here high degree of foaming if it is free of surfactant type material. These materials may be introduced through compressor oil, plug cock lubricant, and corrosion inhibitors used either in the formation or in the gas gathering system. So such product must be chosen carefully.

Before entering the still column to be regenerated, glycol is preheated in two steps. First, in heating coil a top the regeneration column, providing the desired overhead reflux which is controlled by 3 ways overhead temperature control valve .Second, in a plate heat exchanger with regenerated glycol.

The glycol regeneration column fitted with 4 bubble cap trays is top mounted on the reboiler. The bath temperature is controlled at 204 0C by the action on the HP steam flow rate to the reboiler.

Lean TEG leaving E-402 is afterward stripped by dry fuel gas in the packed stripping column C-403 to reach the final concentration of 99.7% wt. Then it returns to the glycol surge drum V-403, after being cooled to 80 0C in the glycol-glycol exchanger E-401 A/B. From the surge drum V-403, it is pumped by the reciprocating pump P-401 A/B to the trim cooler E-403 for cooling the lean glycol to 45 0C and further send back to the glycol absorber C-401.

Because of these considerations the design of the regeneration process varies with the design of the plant. The integration of heat exchangers is especially important, because this reduces the overall energy consumption of the plant.

4.4 Summary:

There are four models for gas dehydration. They are refrigeration, membrane processes, adsorption and absorption. Refrigeration does in many case not remove enough water from the gas, it is however often used in combination with the other dehydration methods. Membrane processes is only economical for small gas flows, which excludes it in most dehydration cases. The adsorption yields the lowest water contents in the gas, dependent on the adsorbent. Even though the absorption process can not remove as much water as adsorption it is often the preferred method. This is because it removes sufficient water to reach the required criteria for the dry gas, as well as gives a better cost/benefit result than the adsorption process. In some cases where low temperature gas treatment is involved adsorption dehydration is required. In those cases the cost is often reduced by combining adsorption plant with an absorption plant. The efficiency of a dehydration process is evaluated by the water contents in the gas after the dehydration. The water contents after the dehydration is often given as the water dew-point, this is to insure that no water will condense in the pipeline. The water dew-point is therefore more practicable because it is directly comparable with the pipeline operating conditions.

The glycol dehydration process can be divided into two parts. First lean glycol dries the wet gas, thereby making the glycol rich. In the second part of the process water is removed from the rich glycol making it lean once again. In the second part of the plant pressure and temperature is change to achieve the optimal operating conditions both in the contactor and the regenerator. The changing of pressure and temperature creates a wide range of possibilities of the final design of the plant.

Chapter 5

Process Simulation

When a process plant is simulated with process simulation programs there are several things that must be taken into account. This includes the settings for the simulation program, design and settings for the process plant. When the simulation is being created additional problems might arise, because variables needs to be defined or values estimated before the simulation can be calculated.

5.1 Dehydration plant specifications

The specifications consist of the composition, flow, temperature and pressure for the wet gas, and the required purity of the lean glycol. Finally the operation conditions for the contactor are given.

Gas

Flow(mmscfd)	10
Temperature(^{0}C)	29.23
Pressure(bar)	62

Molar composition

Component	Mole fraction
Nitrogen	0.00099
H2S	0.01548
CO2	0.0283
Methane	0.898
Ethane	0.0309
Propane	0.0147
i-Butane	0.0058
n-Butane	0.0029
i-Pentane	0.0009
n-Pentane	0.0004
H2O	0.0009
TEGlycol	0

Glycol

Type	TEG
Lean TEG purity	99.05 wt%
Lean TEG temperature	50
Lean TEG pressure	62.05

Contactor

Pressure(bar)	62
Gas temperature(^{0}C)	29.23
Glycol Temperature(^{0}C)	50

The lean TEG purity must be equal to or higher than 99.05 wt%, to insure this demand is met at all times the value used in the dehydration plant will be 99.6 wt%. The specifications for the dehydration plant will be incorporated in the final plant design.

5.2 Dehydration plant design

The design of the dehydration plant has its origin in the processes described in section 4.3, and the requirements for the plant given in section 5.1. A description of the dehydration plant will be given in this here along with some plant specifications.

The gas inlet temperature differs from the temperature in the contactor. The gas is therefore cooled and the liquids removed in the inlet scrubber. The TEG temperature and pressure differs from the contactor specification, first the pressure is raised by a pump and then the temperature is reduced.

The contactor is a column with eight trays. The TEG inlet and gas outlet is in the top of the column, the TEG outlet and gas inlet is in the bottom of the column. The TEG flow is dependent on the water contents in the wet gas, with a flow of 0.025 m3 TEG/kg Water.

After the contactor the pressure of the now rich TEG is reduced, by a valve. The temperature is increased before the rich TEG is flashed to remove hydrocarbons dissolved in the TEG in the contactor. The temperature is increased once more before the regenerator.

The regenerator is a column with two trays plus a condenser and a boiler. The rich TEG enters the regenerator on the middle tray. Because a TEG purity of 99.6 wt% is required stripping gas is added to the boiler of the column. The stripping gas is pure nitrogen with a flow of 28.3 Nm3 gas/m3 rich TEG. The temperatures given for the boiler and condenser are 205 °C and 98.9 °C respectably.

After the regenerator the once again lean TEG is cooled to 50 °C to save the pump. At this point there will also be some kind of TEG makeup system to replace the TEG lost to the gas phases in the dehydration plant. Finally the lean TEG is recycled to the start point.

The described dehydration plant is illustrated with operating conditions in Figure 4-1 & 4.2.

5.2.1 Creating the simulation model

The HYSYS simulation being created here uses the glycol thermodynamic package, and the MKS unit set.

Dehydration The first step in the simulation is to create the gas streams into the contactor column. The gas is cooled and flashed to remove the condensed components. The water content in the flashed gas is used to calculate the TEG flow. In this case the water flow is 107.7 kg/h, thus making the TEG flow 2.69 m3/h.

Now the size of the TEG flow is known, the TEG pressure and temperature must be adjusted to required criteria for the contactor. When the gas and TEG outlet streams are added to the contactor column it can now be calculated.

Regeneration The next step is to reduce the TEG pressure, and increase the temperature before it is flashed to remove dissolved water and hydrocarbons released at this new temperature and pressure. After the separator the temperature is increased before the TEG enters the regenerator.

The flow of stripping gas into the regenerator is dependent on the glycol flow. In this case the flow is calculated to 80 Nm3/h. This flow goes into the reboiler, even though HYSYS illustrates the stream on the side of the regenerator column.

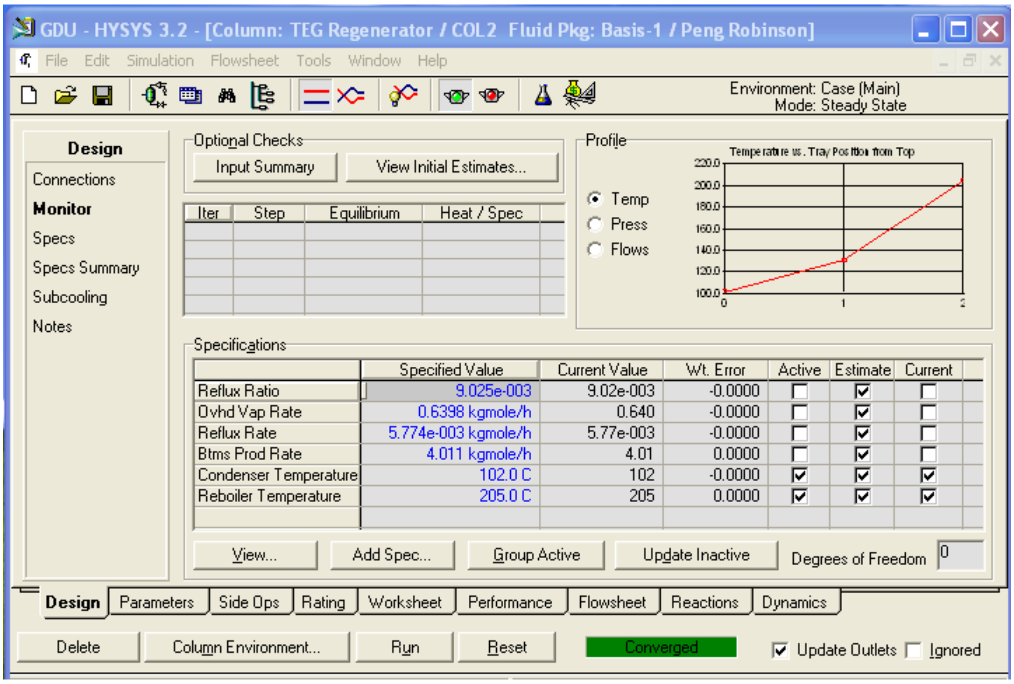
In the regenerator there are two degrees of freedom, which signify that two variables must be defined. This is perfect since two design criteria are given namely the condenser and reboiler temperature. When these criteria are entered into HYSYS the column control screen in not converg.

HYSYS calculation of the column is unable to converge with only the condenser and reboiler temperature given. The design parameters can be supplemented by estimated start values that help to solve the calculations. Alternatively other values may be defined that describes the distillation better than the temperatures. These new designed values could be some of the estimates previously used only to guide the calculation.

Column control conditions The first estimate is the lean TEG purity, the TEG mass fraction in the liquid phase in the reboiler is set to 0.996.

A second estimate is the total vapour flow from the condenser. This estimate is calculated as the total flow of stripping gas into the regenerator plus the flow of water in the TEG. The total vapour estimate is calculated as mass flow, giving an estimated flow of 217 kg/h.

A third estimate is given by the reflux ratio from the condenser. This value is not estimated but calculated by HYSYS. The third estimate is actually redundant in most cases because the first two estimates are sufficient to describe the column. But in some rare cases when the column is reset, it will not converge again. To avoid these situations this third initial estimate are added to the column. This gives the column control screen in converg.



Specifications

	Specified Value	Current Value	Wt. Error	Active	Estimate	Current
Reflux Ratio	9.025e-003	9.02e-003	-0.0000	☐	☑	☐
Ovhd Vap Rate	0.6398 kgmole/h	0.640	-0.0000	☐	☑	☐
Reflux Rate	5.774e-003 kgmole/h	5.77e-003	-0.0000	☐	☑	☐
Btms Prod Rate	4.011 kgmole/h	4.01	0.0000	☐	☑	☐
Condenser Temperature	102.0 C	102	-0.0000	☑	☑	☑
Reboiler Temperature	205.0 C	205	0.0000	☑	☑	☑

Degrees of Freedom 0

Figure 5.1: Regenerator control screen in HYSYS

With these estimates the column now converges.

TEG makeup The TEG from the regenerator is cooled and recycled back to the TEG inlet stream. To do this a logical recycle operator must be inserted between the two streams. There is a problem with the recycling of the TEG; this is that small amounts of TEG are lost from the system in the gas flow from the contactor, separator and regenerator. The lost TEG must be replaced, or less TEG than required is recycled.

To makeup the lost TEG the gas stream from the contactor, separator and regenerator is entered into a component splitter. In the component splitter the TEG is separated from the gas, creating a stream of pure TEG that is transferred back to the TEG stream. A mixer is required to mix the recovered TEG with the TEG from the regenerator.

5.2.2 The final HYSYS model

The final HYSYS model of the dehydration plant is illustrated in Figure 5-2

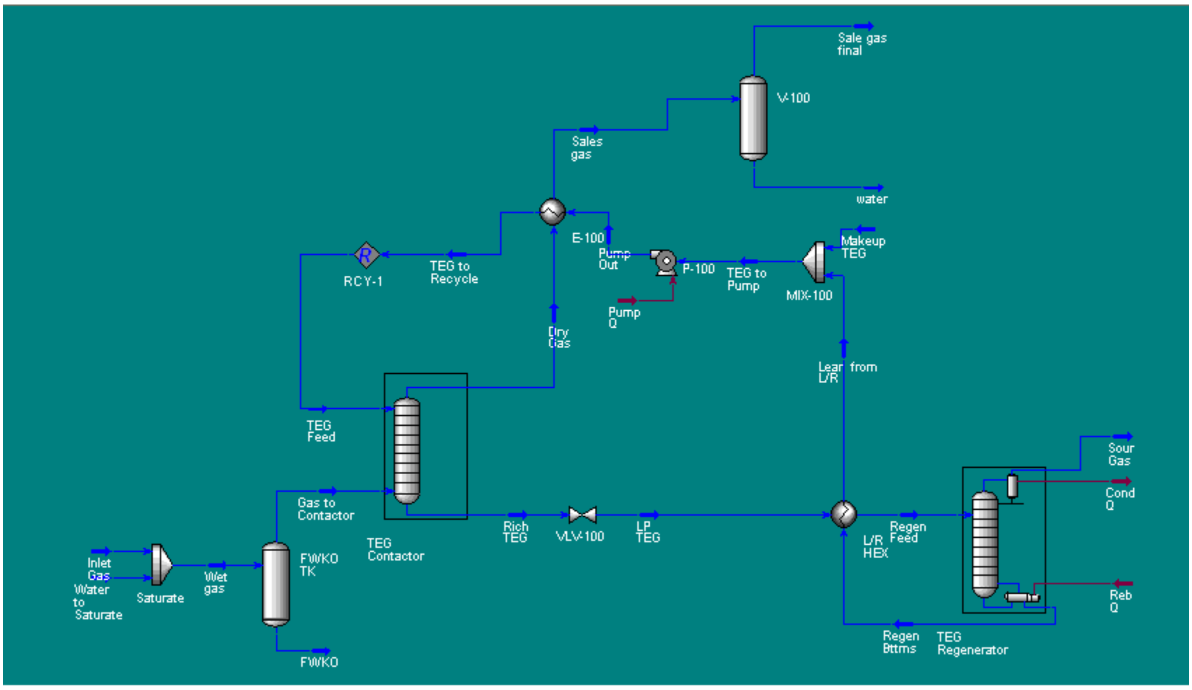


Figure 5.2: Hysys Model

5.3 Simulation results

water content in the gas before and after the contactor is given in Table 5.1

	Mole fraction [mole%]	Mass fraction [wt%]	Mole flow [kmole/h]	Mass flow [kg/h]
WET GAS	0.000999	0.00097	0.5	9.007
SALE GAS	0.0000514	0.0000502	0.0256900	0.270210
DIFFERENCE	-	-	-	8.73679

Table 5.1: water content in the gas before and after the contactor

From the data in Table 5.1 it can be concluded that the contactor removes 97% of all the water from the wet gas.

Water flow: When evaluating a dehydration process it is important to investigate the water flow in the simulation. The water mass balance is therefore given in Table 5.2.

The three main exit points for the water is in the liquid flow from the inlet scrubber (8%) and the vapour flow from the regenerator (35%) & with sour gas (49%). The most cost efficient dehydration method is the inlet scrubber; this decreases the required water removal capacity for the dehydration plant. With the actual water flow in the inlet scrubber, it would perhaps be more appropriate to call it a separator. The main idea with the inlet scrubber is to remove liquid droplets from the gas.

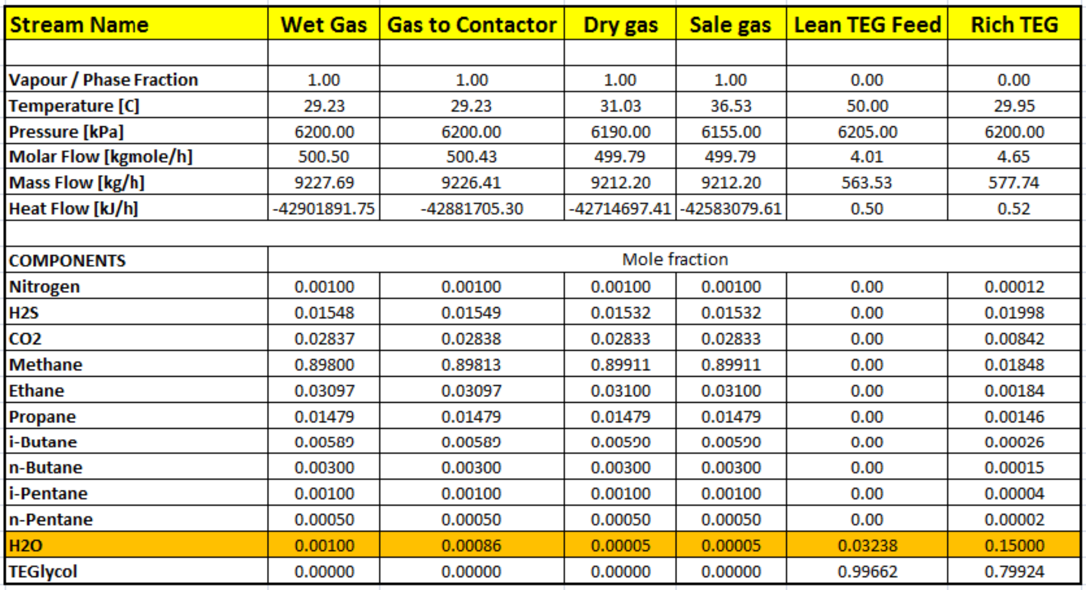
	Stream	Flow [kg/h]	Total [kg/h]
in	wet gas	9.008	14.305
	TEG to pump	5.298	
out	FWKOD	1.275	14.305
	Dry gas	0.463	
	flash separator	0.463	
	Regeneration bottom	5.066	
	Sour Gas	7.039	

Table 5.2: Water mass balance

Glycol purity: The last parameter to be investigated is the efficiency of the regenerator. The dehydration unit is design with a lean TEG purity of 99.6 wt%. The lean TEG purity in stream regenerator is 99.62 wt%, after the makeup TEG is added the purity becomes 99.63 wt%. The result of the TEG regeneration is as expected.

Glycol loss: The total loss of TEG in the dehydration plant equals the size of the makeup TEG stream. TEG loss is 0.0158 kg/h, compared to the total TEG flow of 558.22 kg/h, giving a TEG loss of 0.003%.

Material Balance: The results given in the report are limited to the results of interest. All the results are given as pdf-files on the CD in the folder \HYSYS\.



Stream Name	Wet Gas	Gas to Contactor	Dry gas	Sale gas	Lean TEG Feed	Rich TEG
Vapour / Phase Fraction	1.00	1.00	1.00	1.00	0.00	0.00
Temperature [C]	29.23	29.23	31.03	36.53	50.00	29.95
Pressure [kPa]	6200.00	6200.00	6190.00	6155.00	6205.00	6200.00
Molar Flow [kgmole/h]	500.50	500.43	499.79	499.79	4.01	4.65
Mass Flow [kg/h]	9227.69	9226.41	9212.20	9212.20	563.53	577.74
Heat Flow [kJ/h]	-42901891.75	-42881705.30	-42714697.41	-42583079.61	0.50	0.52
COMPONENTS	Mole fraction					
Nitrogen	0.00100	0.00100	0.00100	0.00100	0.00	0.00012
H2S	0.01548	0.01549	0.01532	0.01532	0.00	0.01998
CO2	0.02837	0.02838	0.02833	0.02833	0.00	0.00842
Methane	0.89800	0.89813	0.89911	0.89911	0.00	0.01848
Ethane	0.03097	0.03097	0.03100	0.03100	0.00	0.00184
Propane	0.01479	0.01479	0.01479	0.01479	0.00	0.00146
i-Butane	0.00589	0.00589	0.00590	0.00590	0.00	0.00026
n-Butane	0.00300	0.00300	0.00300	0.00300	0.00	0.00015
i-Pentane	0.00100	0.00100	0.00100	0.00100	0.00	0.00004
n-Pentane	0.00050	0.00050	0.00050	0.00050	0.00	0.00002
H2O	0.00100	0.00086	0.00005	0.00005	0.03238	0.15000
TEGlycol	0.00000	0.00000	0.00000	0.00000	0.99662	0.79924

Table 5.3: Material Balance

Chapter 6

Conclusion

There are several processes involved in processing the reservoir fluid into oil, gas and water. One of the most important processes is gas dehydration by TEG, because wet gas increases corrosion and can course plugs from ice or gas hydrate in pipeline.

Absorption with TEG offers the best cost benefit choice for the dehydration process. The dehydration process is divided into two parts, the dehydration and the regeneration. In the dehydration part, gas is dried by the glycol. In the regeneration part the water is removed from the glycol so it can be used for dehydration once more. There are several possibilities in the design of the dehydration plant. The design options include the integration of heat exchangers and the recovery rate of the glycol.

The objective of this report was design and simulation of the dehydration process by TEG. This requires calculations of the interaction between the components in the dehydration process; these are done with engineering design and calculations.

Chapter 7

APPENDIX

7.1 Appendix-1

7.2 Appendix - 2

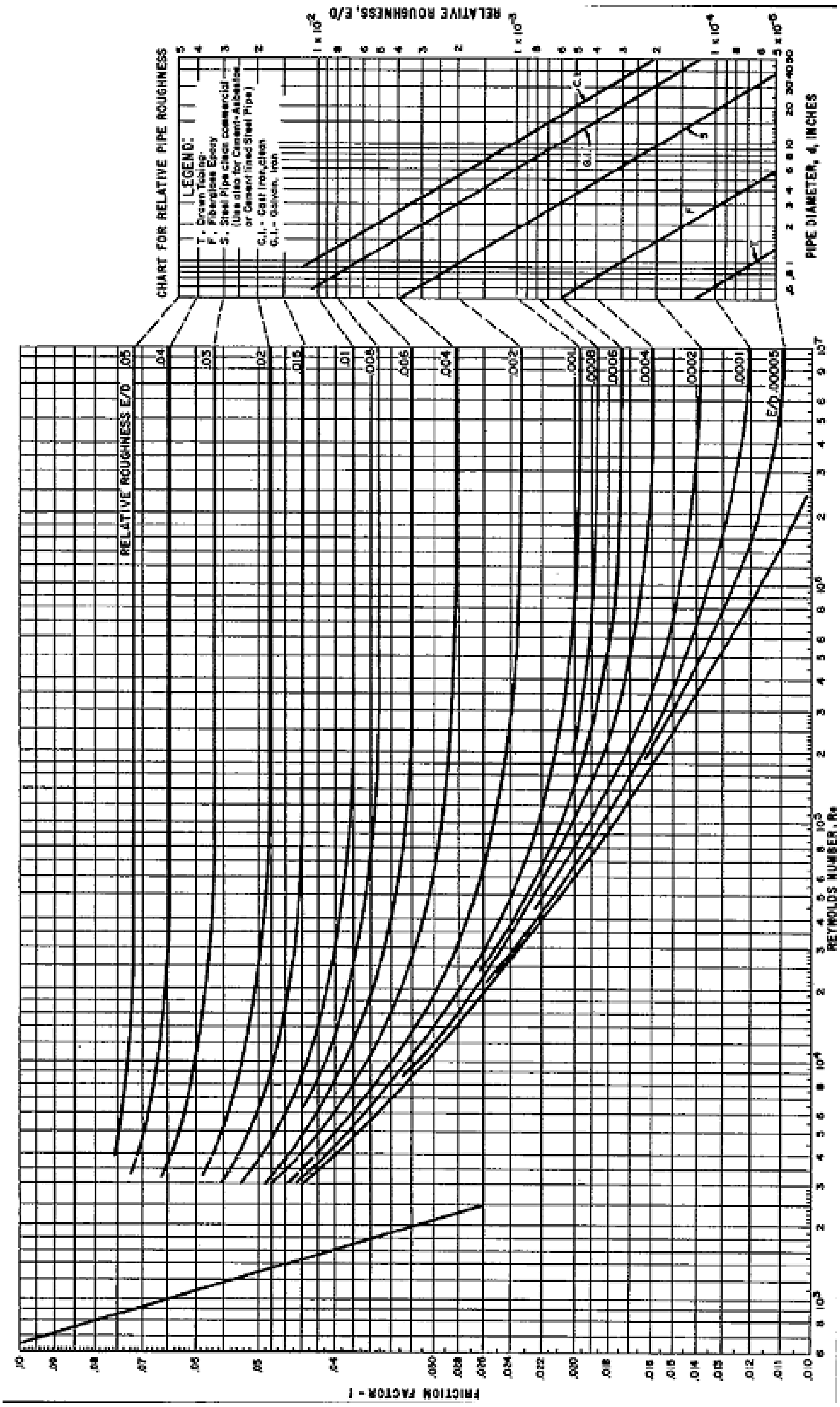


Figure 7.1: Friction Factor Diagram (Modified Moody Diagram)

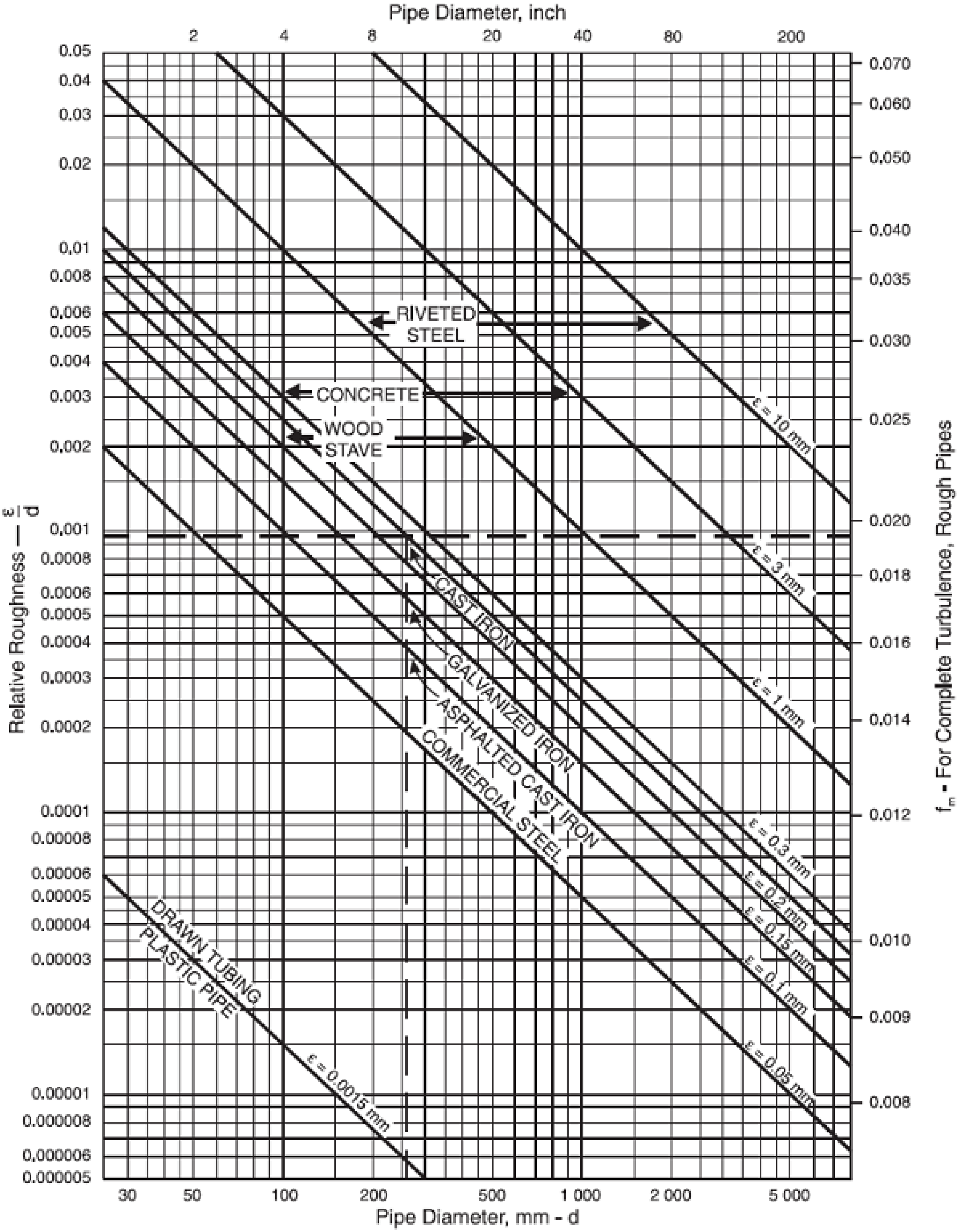


Figure 7.2: Relative Roughness of Pipe Materials and Friction Factors for Complete Turbulence

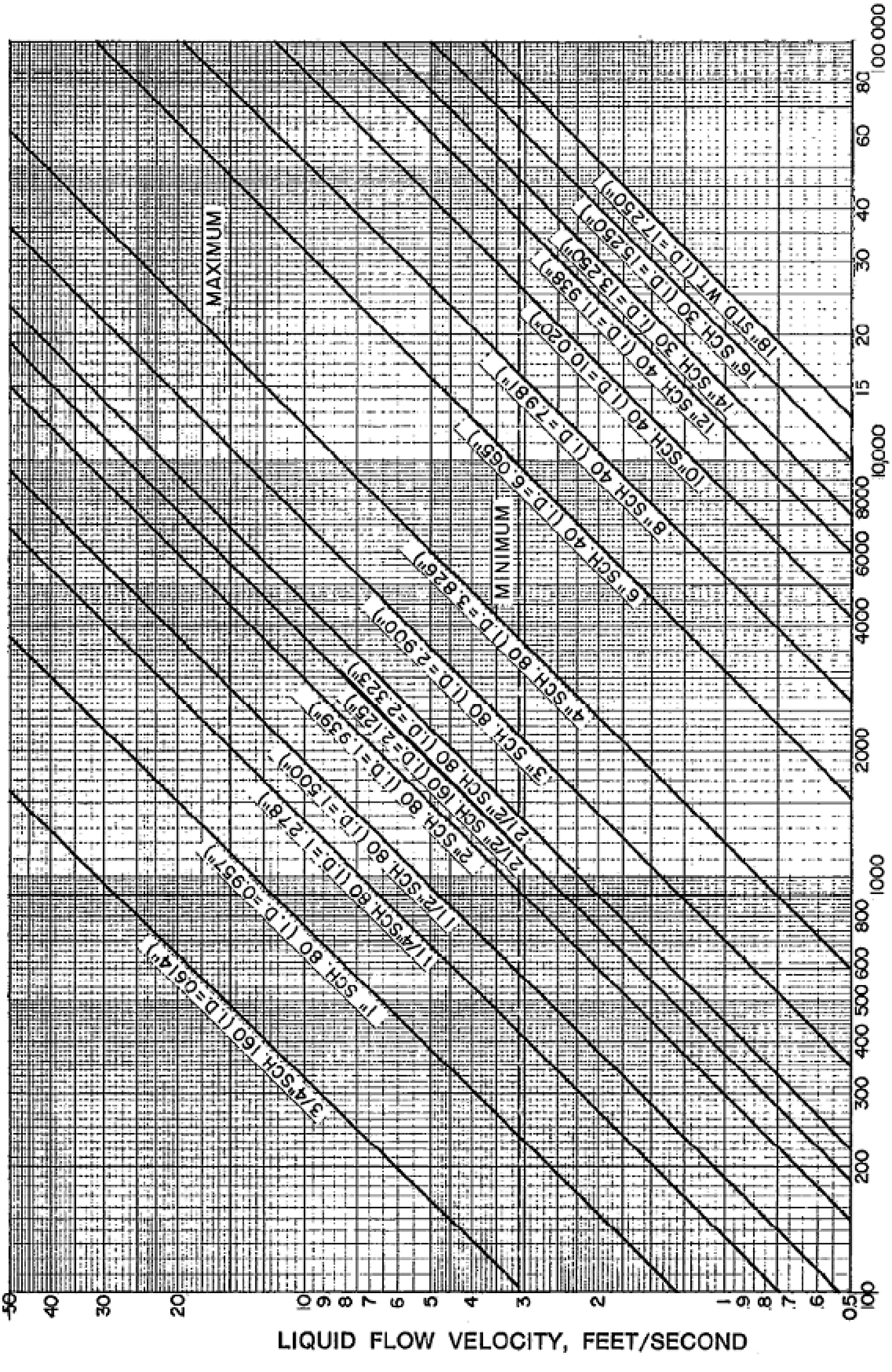
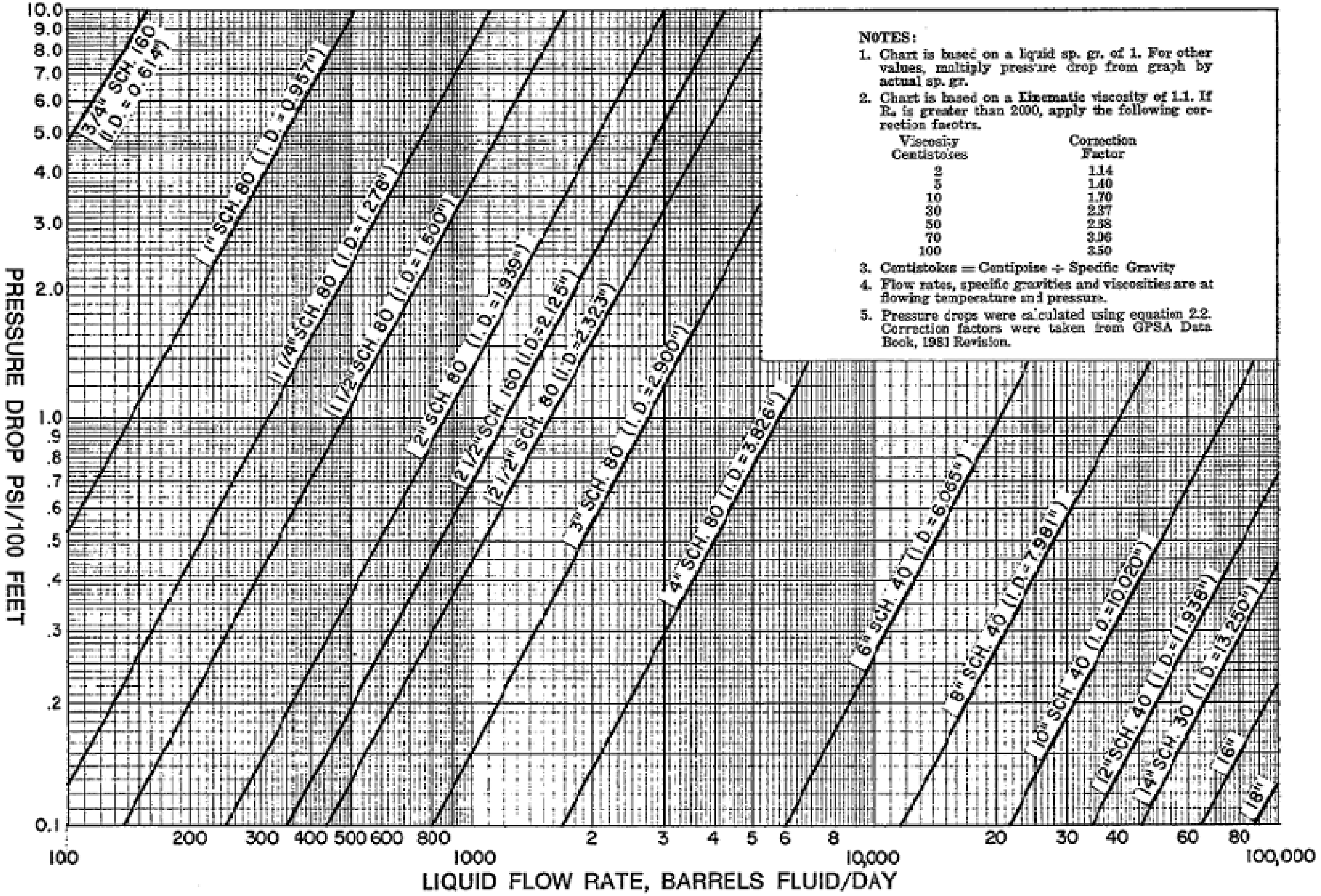


Figure 7.3: Velocity in Liquid lines



NOTES:

1. Chart is based on a liquid sp. gr. of 1. For other values, multiply pressure drop from graph by actual sp. gr.
2. Chart is based on a Kinematic viscosity of 1.1. If R_e is greater than 2000, apply the following correction factors.

Viscosity Centistokes	Correction Factor
2	1.14
5	1.40
10	1.70
30	2.37
50	2.38
70	3.06
100	3.50

3. Centistokes = Centipoise ÷ Specific Gravity
4. Flow rates, specific gravities and viscosities are at flowing temperature and pressure.
5. Pressure drops were calculated using equation 2.2. Correction factors were taken from GPSA Data Book, 1981 Revision.

Figure 7.4: Pressure drop in liquid lines

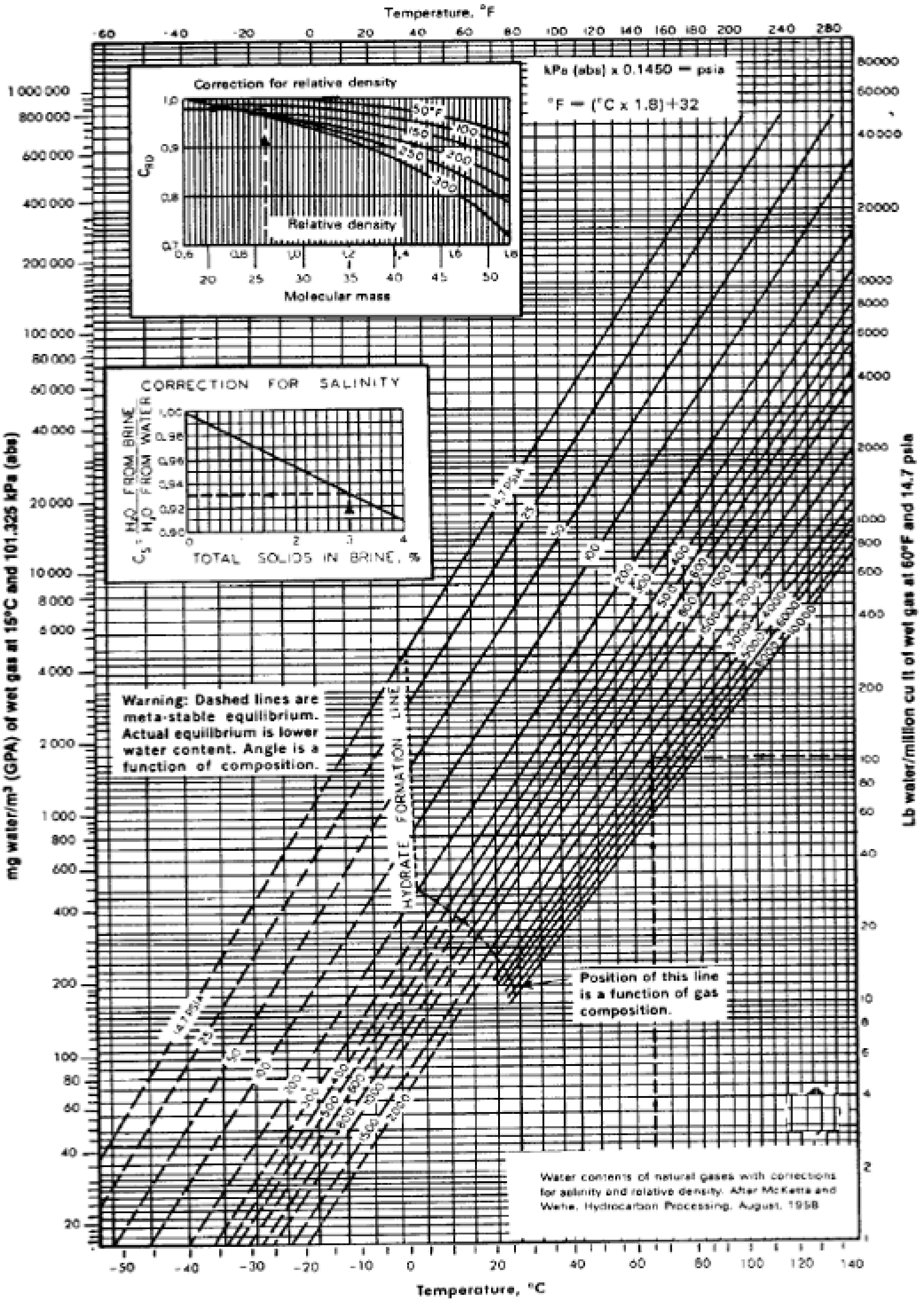


Figure 7.5: water content of natural gas

Bibliography

[1] Offshore Book – An introduction to the offshore industry; Offshore Center Danmark; January 2008.

[2] Natural Gas – Production, Processing, Transport; A. Rojey et.al; Editions Technip; 1997; ISBN: 2-7108-0693-2.

[3] GPSA Engineering Data Book, 11th Edition (Electronic), Volume I & II

[4] API Recommended practice 14E, Recommended Practice for Design and Installation of Offshore Production Platform Piping Systems, Fifth edition, March 2007

[5] F. S. Manning and R. E. Thompson;Oilfield Processing of Petroleum - Vol. 1 Natural Gas; 1st ed.; Pennwell Publishing Company 1991

[6] Robert H. Perry and Don W. Green;Perry's Chemical Engineers' Handbook; 8th ed.; McGraw-Hill 2007

[7] Specification for Glycol-Type Gas Dehydration units – API Specification 12GDU; American Petroleum Institute; 1990.

[8] Glycol - type gas dehydration system-design engineering practice (DEP 20.04.10.10-Gen)

[9] API Recommended practice 12K: indirect Type Oil- Field Heaters.

[10] Design guide for OIL & Gas Separators (Engineers India Ltd).

[11] J. D. Seader and E. J. Henley;Separation Process Principles; 1st ed.; John Wiley & Sons Inc. 1998; ISBN: 0-471-58626-9.

[12] Wayne D. Monnery and William Y. Svrcek, October 1993, Successfully specify three phase separator, University of Calgary, Page 29-40.

[13] W.Y.Svrcek, W.D. Monnery, October 1993, Design Two Phase Separators Within the Right Limits, University of Calgary, Page 53-59.

[14] Jing Wu, Laibin Zhang, Wei Liang, Jinqiu Hu 2012.A novel failure mode analysis model for gathering system based on Multilevel Flow Modeling and HAZOP, College of Mechanical and Transportation Engineering, China University of Petroleum, Beijing 102249, China Page 1-2.

[15] W. van Wassenhove, C. Twu and J. A. Feliu;Using a More Thermodynamically Adapted Model Improves Modelling of TEG Dehydration Units – Allows Better Operation and Efficiency; Aspentech.

[16] Campbell, J.M.: Gas conditioning and processing, vol. 2, CPS, 1982.

[17] McCabe-and-Smith,Unit Operations of Chemical Engineering ,7th edition,McGraw Hill Chemical Engineering Series

[18] Meyers.R.A, "Handbook of Petroleum Refining Processes", McGraw-Hill Professional, third edition, 2003.