A
Dissertation on

Design of Three-phase separator
And
Piping System for three-phase separation

By

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(M.Tech. Gas Engineering)

Harnessing Energy through Knowledge

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CERTIFICATE

This is to certify that **Mr. Maqusood Beig Tegginmani**, a student of **M.Tech in Gas Engineering**, bearing **R.No R-030204008**, has carried out a final semester project titled "**Design of Three Phase separator and piping system for three phase separation**" at Maritime Industrial Services Co. Inc. Sharjah UAE, in partial fulfillments of the requirements for the degree of Masters of Technology. This work is certified as bonafide.

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TO WHOM IT MAY CONCERN

This is to certify that **Maqusood Beig T.** was working with Maritime Industrial Services Co. Ltd. Inc. (MIS) from February 10th, 2006 till date as **Trainee Engineer.**

He has worked in our "Engineering Dept" with the "Process Section", During his tenure of training with MIS, it has been observed that he was sincere, hardworking & capable of handling the responsibilities assigned to him.

We wish him all success in his future endeavours..

Yours truly,

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Abstract

Design of Three-phase Separator and Piping system for three-phase separation

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Separation of oil and gas is a critical field processing operation. As the producing pressures have risen and lighter condensates are produced, efficient separation becomes more critical than ever. Basically separation means bifurcating a fluid mixture by the principles of

1. Gravity settling
2. Centrifugal force
3. Impingement
4. Electrostatic precipitation
5. Sonic precipitation
6. Filtration
7. Adhesive separation
8. Absorption
9. Thermal

Other factors that effect the separation are size and characteristics of the entering fluid, size and cost of the separator. The design of separator is based on the steady flow of fluids. Many books give the general design of separator, but this is not sufficient for a particular application. Design varies depending on the process requirements. The standard practices for the design of separator should be followed for designing for a particular process requirement. These practices are based on extensive research and experience. These practices guide the designers for selection and design for a particular application.

The design method presented in this report is based on standard practice followed by most of the engineering companies i.e. (Shells Shell's DEP Standards (31.22.05.11 and 31.22.05.12)- Gas/Liquid separators-type selection and design rules and Liquid/liquid and Gas/liquid /Liquid (Three phase) separators – type selection and design rules). Standard practices give the detailed guide for selection and design. The various types of separators presented here give the basic idea of the position of these separators in the process system.

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Maqusood Beig Tegginmani

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Chapter 1

Introduction.

1.1 Overview of the Gas-liquid separation

Hydrocarbon stream produced from a reservoir is a complex mixture of several hydrocarbons, intimately mixed with water in liquid and gaseous states. There are also solids with other contaminants present. The produced stream may be unstable with components undergoing rapid phase transformation as the stream is produced from a several hundred feet deep reservoir with a high temperature and pressure to surface conditions. It is important to remove any solids and contaminants and to separate the produced stream in to water, oil and gas, which are handled and transported separately. A gas and liquid separation operation involves the separation and stabilization of these phases into saleable products. Generally intermediate hydrocarbons in the liquid state fetch a higher price; therefore it is desirable to maximize liquid recovery.

Field processing of Natural gas includes:

- Gas and liquid separation operations to remove the free liquids, crude oil, hydrocarbon condensate and water and the entrained solids.
- Recovery of condensable hydrocarbon vapors; stage separator, or low temperature separation techniques are used.
- Further cleaning of the gas and oil streams after separations.
- Gas dehydration processing to remove from the gas condensable water vapor that may lead to the formation of hydrates under certain conditions.
- Processing the gas to remove other undesirable components such as hydrogen sulphide and carbon di oxide

Some of these processes are accomplished in field but in some cases, the gas goes to plant facility for further processing. Thus gas processing can be divided into field treatment and plant operations.

Field Treatment of Natural Gases.

Separation of well stream gas from free liquids is by far the most common of all field-processing operations and at the same time one of the most critical. A properly designed separator will provide a clean separation of free gases from the free hydrocarbon liquids. A well stream separator must perform the following:

- Cause a primary separation of the mostly liquid hydrocarbon from those that are mostly gas.
- Refine the primary separation by removing the most of the entrained liquid mist from the gas
- Further refine the separation by removing the entrained gas from the liquid; and
- Discharge the separated gas and liquid from the vessel and ensure that no re-entrainment of one into the other occurs.

If these Functions are accomplished, the basic separator design must:

1 Control and dissipate the energy of the well stream as it enters the separator.
2 Ensure that the gas and liquid velocities are low enough so that gravity segregation and vapor liquid equilibrium can occur.
3 Minimize the turbulence in the gas section of the separator and reduce velocity.
4 Control the accumulation of froths and foams in the vessel.
5 Eliminate re-entrainment of the separated gas and liquid.

6 Provide an outlet for gasses with suitable controls to maintain preset operating pressure.
7 Provide outlets for liquids, with suitable liquid level controls.
8 If necessary, provide clean out ports at points where solids may accumulate.
9 Provide relief for excessive pressures in case the gas or liquid outlets should be plugged and,
10 Provide equipment (Pressure gages, thermometers and liquid level gauges, glass assemblies) to check visually for proper operation.

The figure1.1 gives the brief outline for the field treatment of the natural gases.

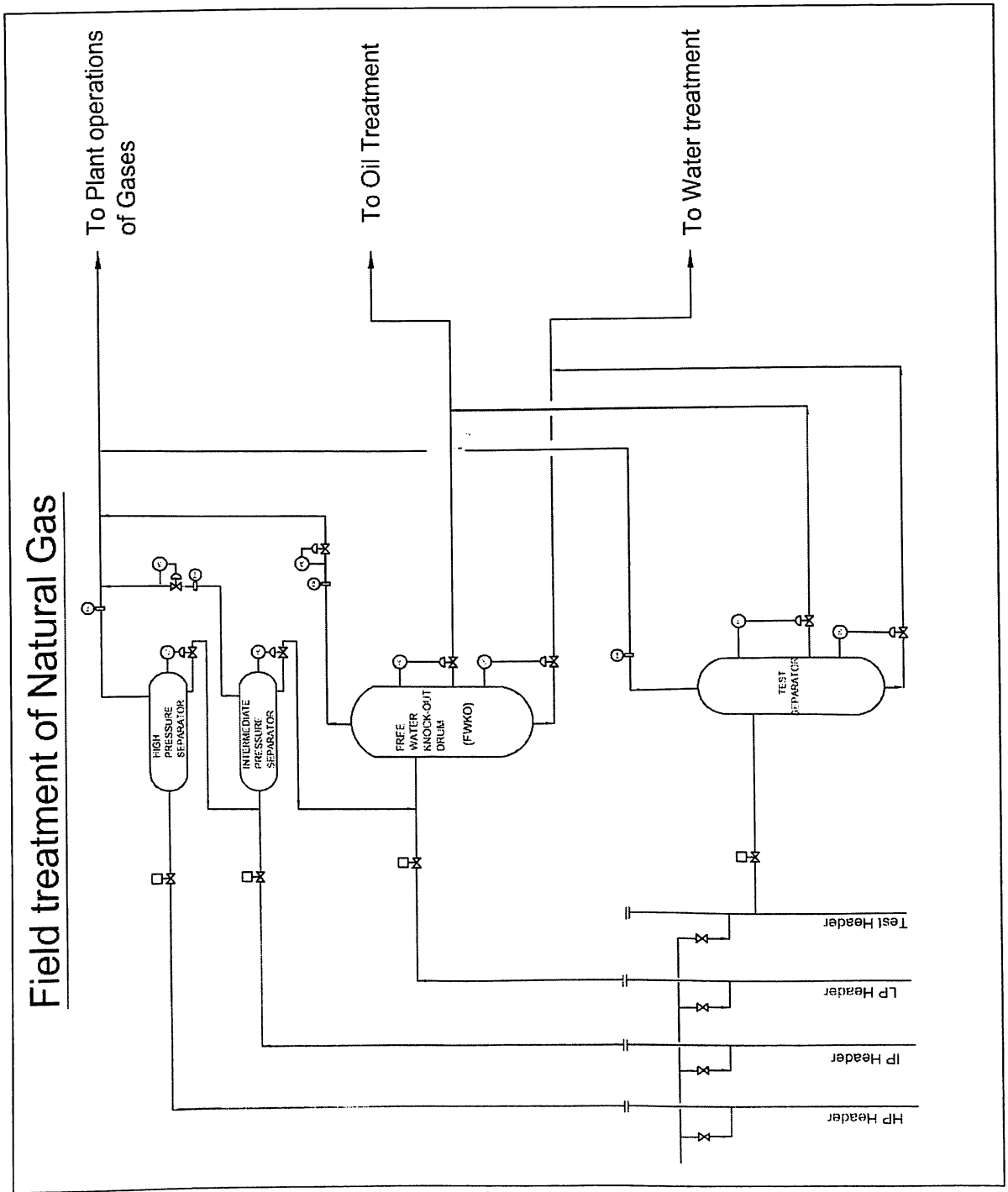


Fig1.1

1.2 Problem under consideration.

The gas supply for Qatar QG 3 & 4 LNG plant will be extracted form the north field, Offshore Qatar. The reservoir condition is 365 bars and 109°c. Gas will be produced through three normally unmanned installations without any processing facilities. Each installation will have 10 production wells with future expansion to 15 wells. The installations will have a phased development schedule with a two outermost installations and flowing into individual pipelines and onward to shore, the third platform will supply gas to both pipelines.

Please refer the fig 1.2 for process flow diagram

The following pressure vessels will be designed, for the project.

- Test separator:

The test separator is a three phase vertical separator. Required internals are shoepentoeter inlet device, plate pack, and demister. A detailed data sheet is provided in table 1.2.

General Data:
Location:
The pressure vessel will be located on an offshore, in the Qatar sector of the Arabian Gulf, located approximately 60 KM Offshore.

Environmental Data:
The pressure vessel will be designed to comply with the following environmental data.

Wind speed	34.7m/s
Max Air temperature	49 °c
Minimum Air Temperature	4 °c
Environment	Offshore saline
Humidity Maximum	74%

Front End Engineering (FEED) Data.

The separator inlet gas composition is given in table 1.1.
Conditions: 69.9 °c and 108 barg

Front End Engineering (FEED) Data: The table1.1 shows the data.

1.3 The need for this study.

The design of separators of vapor and liquids is essential to all processes. The design concepts of a simple separator can may be extended to several other processes such as fractionation towers, two-phase flow, slug catcher design, desalters. The purpose of this study is to review the principles governing the basic separation process.
Three phase separators handle gas plus two immiscible liquid phases. These two liquid phases be oil and water, glycol and oil, etc. The best potential applications of three-phase

separators occur where space is a major consideration. The design principles given here apply to both three-phase and two-phase separators.
The use of different standards in this study makes oneself aware of the design principles in standards like API 12J, DEP (31-22-05-11 & 12), API RP 14E and ASME B31.3 etc.

The understanding of the field treatment of gas is very critical before the development of the field because the wrong design of separator in field treatment leads to accidents. This study gives the brief understanding to the field treatment of the gases. The Control systems are very important to ensure liquid vapor separation.

The piping calculation done here gives the brief idea of how the material selection, fluid flow sizing and the pressure integrity are done for the piping system.

1.4 Objectives:

In the light of above mentioned needs the objectives of this project are:

1. Vapor-Liquid equilibrium calculations for primary separation.
2. Determination of Gas Oil ratio (GOR).
3. Determination of condensate gravity.
4. Determination of the gas gravity.
5. Design of the Separator.
    a. Sizing of the separator.
    b. Sizing of the nozzles.
    c. Design of internals of the separator.
    d. Conclusion
6. Material selection of the piping.
7. Sizing of piping system.
8. Pressure integrity Calculations (wall thickness of pipes).

1.5 An outline of this study:

The goal of this project is to design a three-phase separator that has to be installed in offshore conditions over a unmanned riser platform.

Chapter 2 gives the broad classification of different types of separator. This chapter gives the better understanding of the different types of separators. The applications of these separators in various scenarios are also highlighted.
Chapter 3 is selection of suitable separator, which is based on the design principles from the various standards. The process engineers apply this method in the design during the execution of the project.
Chapter 4 gives the brief idea for the design of separator. The principles presented here give the procedure for design of separator. The design varies for two-phase separator and three-phase separator. The flow chart presented here gives the scheme and sequence for the design of different separators.
Chapter 5 discusses the equilibrium calculations for a 3-stage separator. This discussion helps to find the GOR. These principles are utilized further in the calculations.
Chapter 6 discusses the control systems for the separator. The control system should be very efficient for a good separation system.

Chapter 7 discusses the piping system utilized in oil and gas gathering and separation systems. The understanding of this helps us to design a good separator.
Chapter 8 gives the process calculations for the design of the project problem. The sequence presented here is the actual sequence of design done in an industry.
Chapter 9 gives the conclusions and summary. The contributions and recommendations for further work are presented.

FEED Data:

Stream	**Unit**		**Molar Composition (%)**	**Z**
Total Stream			Methane	77.9952
Total Molar rate	KG-MOL/HR	7663.5	Ethane	4.5502
Total mass rate	TONNE/DAY	4124.75	Propane	1.7002
Total standard rate	MMSCFD	153.8524	Iso-butane	0
Temperature	C	69.9	N-butane	0.9277
Pressure	BARG	108	Iso-Pentane	0
Molecular Weight		22.4225	N-Pentane	0.5077
Total Vapor			N-Hexane	0.2852
Vapor Molar rate	KG-MOL/HR	7210.75	Heptanes +	2.2702
Vapor mass rate	KG/HR	145800	H2S	1.3652
Vapor rate (Actual)	M3/HR	1612.75	H2O	2.9877
Vapor molecular weight		20.215	Nitrogen	4.3152
Vapor density	KG/M3	90.38	CO2	3.0952
Vapor Cp	KJ/KG-C	2.635		
Vapor Viscosity	CP	0.02		
Vapor Z				
Hydrocarbon Liquid				
Liquid Molar rate	KG-MOL/HR	253.3725		
Liquid Mass rate	KG/HR	22480		
Liquid Rate (Actual)	M3/HR	35.1575		
Liquid Molecular weight		88.715		
Liquid Density	KG/M3	639.575		
Liquid Viscosity	CP	0.3275		
Water	KG-MOL/HR	198.795		
Water Molar rate	KG/HR	3586		
Water Mass rate	M3/HR	36725		
Water molecular weight		18.0375		
Water density	KG/M3	977		
Water Viscosity	CP	0.415		

Table1.1

Data sheet

					Units
Working Temperature Max./Normal/Min	104	80	22		c
Working Pressure Max./Normal/Min	125	115	110		barg
	@110 barg			@125 barg	
Condensate -Quantity	10897			15260	kg/hr
-Viscosity at working temperature	0.329			0.319	cP
-Density at working temperature.	628.5			625.7	kg/m3
Water -Quantity	4439			3929	kg/hr
-Viscosity at working temperature	0.261			0.366	cP
-Density at working temperature	949.5			969.1	kg/m3
Vapor -Quantity	164744			177568	kg/hr
-Molecular weight	20.91			20.7	
-Density at working temperature	82.91			103.6	kg/m3

Table1.2

Notes

1. Test separator is based on testing of a single well with a maximum combined gas and condensate flow of 150 mmscfd
2. The test separator is a three phase vertical separator.
   required internals are schoepentoeter inlet device, plate pack, and demister mat.
3. Plate pack to be designed to enable 150 micron droplets to be separated at maximum inlet flow rate
4. Condensate and water rates quoted include 1.2 Design margin.
5. Length of the plate pack is 0.30m with a height of 1.9m, based on 110 barg case, plate angle of 45 and plate spacing of 20mm.

Chapter 2

Classification of the separators

Classification of the separators:

2.1 Based on the separation fluids the separators are divided into

The incoming fluid in the separator may contain phases like gas and two immiscible liquids. Depending on the number of the phases, the separators and their types are divided into:

G/L Separators (Gas-liquid Separators) or two-phase separators
G/L/L separators (Gas-Liquid-Liquid Separators) or three phase separators

Two phase separator	Three Phase separators
1. Vertical Knock out Drums	1. Horizontal open two phase Settlers
2. Horizontal knock out Drums	2. Horizontal Two phase settler with plate pack
3. Vertical wire mesh demister	3. Coalescer fitted with a compressed coalescer mat
4. Horizontal Wire Mesh demister	4. Coalescer fitted with a prefabricated coalescer mat.
5. Vertical Inline separator with vane pack	5. Coalescer fitted with coalescer cartridges
6. Vertical two-stage separator with vane pack.	6. Horizontal open three phase settler with boot
7. Horizontal vane type demister	7. Horizontal open three phase settler with weir arrangement
8. Cyclones with tangential Inlet	8. Horizontal three phase settler with plate pack and weir arrangement
9. Cyclones with straight inlet and swirler	9. Vertical three-phase settler with plate pack.
10. Filter separator	

The following paragraphs will give the detailed information regarding each of these separators:

VERTICAL KNOCK-OUT DRUM

Application:

- Bulk separation of gas and liquid.

Characteristics:

- Unlimited turndown;
- High slug handling capacity;
- Liquid removal efficiency typically 90%;

Warning: poor removal efficiency of liquid from mist

- Very low pressure drop;
- Insensitive to fouling.

Recommended use:
- Vessels where internals have to be kept to a minimum (e.g. flare knock-out drums);
- Fouling service e.g. wax, sand, asphaltenes;
- Foaming service.

Non-recommended use:
- Where efficient demisting of gas is required.

Typical process applications:
- Vent and flare stack knockout drums;
- Production separator;
- Bulk separator (e.g. upstream of gas coolers);
- Flash vessel.

HORIZONTAL KNOCK-OUT DRUM

Application:
- Bulk separation of gas and liquid.

Characteristics:
- Can handle large liquid fractions;
- Unlimited turndown;
- Very high slug handling capacity;
- Liquid removal efficiency typically 90%

Warning: poor removal efficiency of liquid from mist;

- 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, asphaltenes;
- For foaming or very viscous liquids.

Non-recommended use:
- Where efficient demisting of gas is required.

Typical process applications:
- Vent and flare stack knockout drums;
- Production separator for low gas/oil ratio (GOR);
- Bulk separator;
- Slug catcher.

VERTICAL WIREMESH DEMISTER

Application:

- Demisting of gas.

Characteristics:

- High turndown ratio (factor 4);
- High slug handling capacity;
- Liquid removal efficiency > 98%;
- Sensitive to fouling;
- Low-pressure drop.

Recommended use:

- For demisting service with a moderate liquid load in form of droplets;
- Where slug-handling capacity may be required.

Non-recommended use:

- Fouling service (wax, asphaltenes, sand, hydrates)
- For viscous liquids where de-gassing requirement determines vessel diameter
- For compressor suction scrubbers unless precautions are taken to prevent the possibility of loose wire cuttings entering the compressor or of the demister mat becoming clogged and thereby increasing the suction pressure drop.

Typical process applications:

- Production/test separator - moderate GOR;
- Non-fouling;
- Inlet/outlet scrubbers for glycol contactors;
- Inlet scrubbers for gas export pipelines;
- For small diameter and/or low-pressure vessels, where extra costs of e.g. vane or SMS internals cannot be justified.

HORIZONTAL WIREMESH DEMISTER

Application:

- Demisting of gas where a high liquid handling capacity is required.

Characteristics:

- High turndown ratio (factor 4);
- Very high slug handling capacity;
- Liquid removal efficiency > 98%;
- Sensitive to fouling;
- Low-pressure drop.

Recommended use:

- Typically for demisting service with a high liquid load and a low GOR;
- Applied where slug-handling capacity may be required;
- For viscous liquids where liquid de-gassing requirement determines vessel diameter;
- In situations where head room is restricted;
- For foaming liquids.

Non-recommended use:
- Fouling service (wax, asphaltenes, sand).

Typical process applications:
- Production/test separator for low GOR
- Applications with height limitations.

VANE-TYPE DEMISTER (Both Vertical and Horizontal)
Application:
- Demisting of gas.

Characteristics:
- Liquid removal efficiency > 96%;
- Moderate turndown ratio (factor 3);
- Suitable for slightly fouling service (if without double-pocket vanes);
- Robust design;
- Sensitive to liquid slugs (in-line separator cannot handle slugs).

Recommended use:
- Typically for demisting service;
- In-line separator to be used only with relatively low flow parameter ($\phi_{feed} < 0.01$);
- Two-stage separator to be used if $\phi_{feed} >= 0.01$;
- Attractive for slightly fouling service (if without double-pocket vanes);
- May be used where demister mats may become plugged, i.e. waxy crudes.

Non-recommended use:
- Heavy fouling service (heavy wax, asphaltenes, sand, hydrates);
- For viscous liquids where de-gassing requirement determines vessel diameter;
- The in-line vertical flow vane pack separator shall not be used where liquid slugging may occur or where $\phi_{feed} >= 0.01$;
- If pressure exceeds 100 bar (abs), due to the consequent sharp decline in liquid removal efficiency.

Typical process applications:
- Compressor suction scrubbers - where vane packs are preferred to demister mats since their construction is more robust;
- Demisting vessels with slightly fouling service.

CYCLONE WITH TANGENTIAL INLET (CONVENTIONAL CYCLONE)
Application:
- Demisting of gas in fouling service

Characteristics:
- Liquid removal efficiency > 96%;
- Insensitive to fouling;
- Limited turndown ratio (factor 2);
- High-pressure drop.

Recommended use:
- Typically for use in a fouling (e.g. coke-formation) environment and where a high demisting efficiency is still required.

Non-recommended use:
- If high pressure drop cannot be tolerated.

Typical process application:
- In oil refineries: Thermal Gas oil Units (TGU); Visbreaker Units (VBU);
- In chemical plants: Thermoplastic Rubber Plants.

CYCLONE WITH STRAIGHT INLET AND SWIRLER

Application:
- Demisting of gas where a high gas handling capacity and a high liquid removal efficiency is required.

Characteristics:
- Very compact separator;
- High liquid removal efficiency (> 99%);
- Very high gas handling capacity (maximum allowable gas load factor =0.9 m/s);
- High turndown ratio (factor 7);
- High-pressure drop;
- Suitable for slightly fouling service (e.g. low sand loading);
- Slug handling capacity.

Recommended use:
- Where there is little plot space available (e.g. in offshore industry or in general for high-pressure conditions);
- As retrofits of existing vessels where capacity debottlenecking and/or improved separation efficiency are required.

Non-recommended use:
- If a low pressure drop is essential;
- If insufficient headroom is available.

Typical process application:
- Wellhead separators;
- Compressor suction and interstage scrubbers;
- Cold separators;
- Inlet separators to adsorption plants.

FILTER SEPARATOR

Application:
- After cleaning (liquid and solids) of already demisted gas when very high liquid removal efficiency is required.

Characteristics:
- Liquid removal efficiency > 99%;
- Very high-pressure drop;

- Sensitive to high liquid loading or slugs;
- Sensitive to fouling by sticky material.

Recommended use:
- Typically as a second-line gas/liquid separator to after-clean the gas stream exiting from the first-line gas/liquid separator.

Non-recommended use:
- Heavy fouling (sticky material) service;
- High liquid loading;
- Slugs.

Typical process application:
Last demisting stage of natural gas prior to dispatch for sale

HORIZONTAL OPEN THREE-PHASE SETTLER

Application:
Bulk separation of primary L/L dispersion and a relatively small gas flow rate.

Characteristics:
- Large liquid handling capacity
- Insensitive to fouling.

Recommended use:
General:
- Fouling service;
- Where the L/L separation is the controlling factor and only bulk L/L separation is required.

Settler with boot:
- Where the volume ratio of the heavy to light liquid phase in the feed is smaller than 0.2 and the de-oiling of the heavy phase is not important.

Settler with weir arrangement:
- Ill-defined L/L interface (use double weir arrangement with no IL control);
- De-oiling of the heavy phase required.

Not recommended:
- For efficient L/L separation

Typical process applications:
- Free Water Knock Out Drums

Overhead Accumulators
Ejector Effluent Separators

VERTICAL THREE-PHASE SETTLER WITH PLATE PACK

Application:

- Efficient L/L separation of a primary dispersion at a relatively high gas load;
- Efficient G/L separation provided appropriate separation internals (e.g. mist mat or SMS-internals) are installed.

Characteristics:

- Compact;
- High gas handling capacity can be achieved (depends on G/L separation internals).

Recommended use:

- If the gas load is high and L/L separation is readily accomplished (i.e. not high viscosity, primary dispersion).

Not recommended:

- if gas load is low;
- if L/L separation is difficult (i.e. secondary dispersion).

Typical process applications:

- Fractionator's overhead vessels in the work-up section of Long Residue Catalytic Cracking Units.

2.2 Based on Configuration, three types of separators are used for Gas-Liquid Separation they are

1 Vertical
2 Horizontal
3 Spherical

Relative Advantages of these Separators:

A Vertical separator can handle relatively large liquid slugs without the carryover into the gas outlet. It thus provides better surge control and is often used on low to intermediate gas-oil ratio (GOR) wells and wherever else large liquid slugs are expected. Vertical vessels can handle more sands. Liquid level control is not so critical in a vertical separator. The tendency of the liquid to re-vaporize is also minimized, because less surface area is available to the liquid for evaporation. It occupies less space, a particularly important advantage for operations on an offshore platform where floor area is at a premium.

Horizontal Separators have much greater gas-liquid interface area, permitting higher gas velocities. They can, therefore handle large volumes economically and efficiently. They are cheaper to fabricate and ship than vertical separator. They are also easier and cheaper to install and service. Horizontal separators minimize turbulence and foaming. For a given capacity, horizontal separators are smaller and cheaper than vertical separators. Horizontal separators are always used for high GOR wells, for foaming well streams, and for liquid-liquid separation.

Spherical separators are very inexpensive, cheaper than either the vertical or the horizontal separators. They are very compact, and offer better cleanout and bottom drain features than even the vertical type. Spherical separators are applicable to well streams with low to intermediate GOR's

Relative Disadvantages are:

Vertical vessels are more expensive to fabricate, and also more expensive to transport to location. A vertical separator for the same capacity is usually larger than a horizontal separator, since the upwards-flowing gas in the vertical separator opposes the falling droplets of the liquid.

In horizontal separators, the liquid level control is critical and the surge space is rather limited. They are much harder to clean, and are therefore not advisable to use where the well produces a lot of sand. They occupy a lot of space. Stacking several of these on top of each other for stage separation operations however can minimize the space requirements.

In spherical separators the liquid level control is critical. They have limited surge capacity and liquid settling section. Because of the limited internal space, it is difficult to use a spherical separator for three-phase (Gas-Oil-Water) separation.

Chapter 3

Selection of the separators

3.1 Selection of Two-phase separators or G/L Separators.

Selection Strategy:
To facilitate the choice of a separator type for a given application the following points are compared.

1. Gas handling capacity: -Max capacity (Gas load factor)
   -Turn down ratio (Ratio of Maximum to Minimum Flow
2. Liquid Removal Efficiency: -Overall
   -With respect to fine mist
   -With respect to the possible flooding above the maximal load factor (Flooding will cause a sharp decrease in efficiency)
3. Liquid handling capacity: - Slugs
   -Droplets (Over loading of the separation internal)
4. Fouling Tolerance: - Sand
   - Sticky Material
5. Pressure Drop:

1: Gas Handling Capacity:

The separator should be large enough to handle the gas flow rate under the most severe process conditions.

We have gas flow rate at operating conditions Q_G in m^3/s

Let $Q_{G\max}$ = the maximum envisaged gas flow rate.

This is found by including the Design Margin for the gas flow rate at operating conditions

And $Q^*_{G\max}$ = Maximum volumetric Gas load factor.

We have

$$Q^*_{G\max} = Q_{G\max}\sqrt{\frac{\rho_g}{\rho_l - \rho_g}}$$

Where $Q^*_{G\max} = A_{\min}\lambda_{\max} = \frac{\pi}{4}D^2 k\left(\frac{\rho_g}{\rho_l - \rho_g}\right)^{0.5}$ Where, $A_{\min}$ → Minimum area of Cross-section of Vessel

D → Internal diameter of the Vessels

$\lambda_{\max} = k.\left(\frac{\rho_l - \rho_G}{\rho_G}\right)^{0.5}$ → Which represents the Maximum gas velocity and k is called Separation coefficient.

The value of k depends on all factors that effect the separation other than density, vortex action, foaming, pulsating flow, liquid flowing in heads, presence of solids, degree of separation needed, separation length, varying gas liquid ratios and the like. It is not surprising that k varies widely in different applications.

Since the value of λ_{max} depends on k which in turn varies for different types of separators.
By knowing the value of $Q^*_{G\max}$ and λ_{max} we can find the minimum area, A_{min} for the gas flow in the separator.

For vertical Separator this A is the cross section area of the vessel but for horizontal vessel this area is the area of cross section of vessel above the highest liquid level.

2: Turn Down Ratio.

This is defined as the ratio of the maximum flow rate to the minimum flow rate through the separator. This ratio indicates the capacity of the separator for large variations in feed.

3: Liquid removal efficiency.

Efficiency of a G/L Separator is normally defined as the ratio of the liquid flow rate separated from the gas stream and the liquid flow rate in the feed entering the separator.

$$E_f = \frac{Q_l}{Q_g} x100 \ \%$$

Liquid removal efficiency for the overall removal of liquid and with respect to fine mist and also with respect to the possible flooding above the maximal load factor is accounted for selection.

4: Liquid Handling Capacity

The ability of the separator for handling slugs and droplets of liquid is also considered during selection.

5: Fouling tolerance:

The fouling of a separator is the choking of the separator due to presence of sand and sticky material present in the feed inlet. This is a important consideration during the selection procedure.

6: Pressure drop:

The pressure drop occurring in the separator is also considered for selection.

Table 3.1 gives the comparison of Performance characteristics of various separators for selection

Performance comparison of the various G/L Separators

	VKO	HKO	VW	HW	VV1	VV2	HV	SMS	SVS	SMSM	CT	CS	VRMC	FS
Gas handling														
Max capacity(λ_{max})	B	B	C	C	D	D	D	E	E	E	E	F	E	C
Turndown (Max/Min flow)	∞	∞	4	4	3	3	3	10	4	10	2	3	2	∞
Liquid Removal efficiency														
Overall, %	90	90	>98	>98	>96	>96	>96	>98	>96	>99	>96	>99	>93	>80
With respect to fine mist	A	A	E	E	C	C	C	E	D	F	B	D	B	F
Flooding above λ_{max} (Y/N)	N	N	Y	Y	*	*	*	N	N	N	N	N	Y	Y
Liquid handling capacity														
As slugs	D	E	D	E	A	D	E	D	D	D	D	D	D	-
As droplet($Q_{L,max}$)	D	D	D	D	B	C	C	D	D	D	D	D	B	B
Fouling Tolerance														
Sand	E	E	B	B	**	**	**	B	C	B	E	C	C	B
Sticky Material	E	E	A	A	**	**	**	A	C	A	E	C	C	A
Pressure drop	A	A	B	B	B	B	B	C	C	C	D	C	D	***

A = Very Low	D = High	∞ = Infinite
B = Low	E = Very High	
C = Moderate	F = Exceedingly High	

* = If double pocket vane pack: N; if single pocket or no pocket vane pack: Y

* * = If double pocket vane pack: A; if single pocket vane pack: B; if no pocket vane pack: C

*** = Depending on the degree of fouling, rating from C to F

VKO Vertical Knock out Drum	SVS Schoepentoeter-vane pack-swirl deck separator
HKO Horizontal knock out drum	SMSM Schoepentoeter-mistmat-swirldeck-mistmat separator
VW Vertical wiremesh demister	CT Cyclone with tangential Inlet(Conventional cyclone)
VV1 Horizontal wire mesh demister	CS Cyclone with straight inlet and swirler
VV2 Vertical two stage separator with vane pack	VRMC Vertical separator with reversed flow
HV Horizontal vane-type demister	Multicyclone bundle Multicyclone
SMS Shoepentoeter-Mistmat-Swirldeck Separator	FS Filter separator

Table3.1

3.2 Selection of a Three-phase separator.

Selection of a suitable LL- separator or GLL separator depends on the following factors such as:

1. Required separation efficiency
2. Required gas and liquid handling capacity.
3. Whether the LL separation or the GLL Separation is the controlling factor.
4. Required fouling factor.
5. Required fouling tolerance.

Definitions for LL Separations

- Type of Dispersion:

  A primary dispersion is one in which the majority of the dispersed droplets are larger than 30μm. A secondary dispersion is one in which the majority of the dispersed droplets are smaller than 30μm.

- Separation efficiency

In general drop size distribution of the dispersed phase in the feed is not known, it is impossible to quantify the separation efficiency in terms of "Cut of Diameter " this is the diameter of the smallest droplets to be removed with an efficiency of 100%

Primary Dispersion:

- Bulk Separation in open settlers; Droplets larger than 150μm removed
- Efficient Separation in Platepacks; Droplets larger than 50μm removed.
- Efficient separation in coalescer mat; Droplets larger than 30μm removed.

The table 3.2 guides the selection of the separator

Screening of L/L and G/L/L separators

√Recommended use X: Non-recommended use

	L/L separators					G/L/L separators			
Separator types	**L1**	**L2**	**C1**	**C2**	**C3**	**T1**	**T2**	**T3**	**T4**
L/L separation									
Primary dispersion	√	√		√		√	√	√	√
Bulk separation	√					√	√		
Efficient separation		√		√				√	√
Secondary dispersion	X	X	√	X	√	X	X	X	X
Ill-defined L/L interface	X	X					√	√	
G/L separation									
Bulk separation						√	√	√	
Efficient separation									√
Liquid handling controlled (in G/L/L separators)						√	√	√	X
Gas handling controlled (in G/L/L separators)								X	√
Fouling service	√		X	X	X	√	√		
High temperature			X	X	X				

Table 3.2

L1: Horizontal open two-phase settler
L2: Horizontal two-phase settler with plate pack
C1: Coalescer fitted with a compressed coalescer bed
C2: Coalescer fitted with a prefabricated coalescer mat
C3: Coalescer fitted with coalescer cartridges
T1: Horizontal open three-phase settler with boot
T2: Horizontal open three-phase settler with weir arrangement
T3: Horizontal three-phase settler with plate pack and weir arrangement
T4: Vertical three-phase settler with plate pack

Chapter 4

Separator Design Principles

4.1 Design of Two-phase separator

Following are the set of steps followed for a separator design.

(Step I)

Determining a max load factor λ_{max} for the selected separator

We have,

$\lambda_{max} = 0.07$ For **Vertical and Horizontal Knock out drum**

$\lambda_{max} = 0.25$ For **flare knock out drums**

$\lambda_{max} = 0.1$ If No feed Inlet device is used for **Horizontal and Vertical Knock out Drums**

$\lambda_{max} = 0.15$ If Shoepentoeter is used for **Horizontal and Vertical Knock out Drums**

For Wire Mesh Demister

$$\lambda_{max} = 0.105.f_{\mu}.f_{\phi}$$

Where $f_{\mu} = \left(\frac{0.001}{\mu_l}\right)^{0.04}$ If $\mu_l > 0.001$ Pa-s

$= 1$ If $\mu_l < 0.001$ Pa-s

& $f_{\phi} = \left(\frac{1}{1+10.\phi_{wm}}\right)$ Where $\phi_{wm} = 0.05.\frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$ for Schoepentoeter

$= 0.2.\frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$ For Half Open Pipe

For a Vane pack Demister

If Archimedes Number, $Ar = \left(\frac{\rho_l}{\mu_l}\right)\sqrt{\frac{\sigma^3}{g(\rho_l - \rho_g)}}$ >225

Then, $$\lambda_{max} = \frac{1.75\left(\frac{g.\sigma}{\rho_l - \rho_g}\right)^{0.24} * \left(\frac{\sigma}{\mu_l}\right)^{0.04}}{1+25\phi_v}$$

Where $\phi_v = \frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$ Flow parameter at van face $\phi_v \leq 0.01$

Else

$$\lambda_{max} = \left(\frac{0.14\left(\frac{\sigma}{\mu_l}\right)}{1+25\phi_v}\right) \text{ If } Ar \leq 225$$

For 2 Stage Separator with Horizontal Flow Vane Pack (With Shoepentoeter and Horizontal Vane type demister)

$$\lambda_{max} = 0.1 + 0.008\rho_g^{-0.14}\phi_{feed}^{-0.76}$$

4.1.1 Sizing of the Nozzles

(Step II)

This shall be based on the maximum flow rates, including the appropriate Design Margin.

Feed/Inlet Nozzle

For determining the internal diameter of the feed nozzle the momentum criteria shall be satisfied, which is given by $\rho_m v_m^2$

Where ρ_m → Mean Density of the Mixture of the feed pipe given by

v_m → Mean velocity of the mixture in inlet Nozzle

We have $$\rho_m = \frac{\rho_l Q_l + \rho_g Q_g}{Q_l + Q_g}$$

and $$v_m = \frac{Q_l + Q_g}{\pi / 4 * d_1^2}$$ Where d_1 → Diameter of the Inlet Pipe.

The momentum Criterion for Different Inlet conditions are:

For No inlet Device $\rho_m v_m^2 \leq 1400$

For Half Open Pipe $\rho_m v_m^2 \leq 2100$

For Schoepentoeter $\rho_m v_m^2 \leq 8000$

Gas Outlet Nozzle:

The diameter of the gas outlet nozzle, d_2 should normally be taken equal to that of the outlet pipe, but the following criterion shall be satisfied

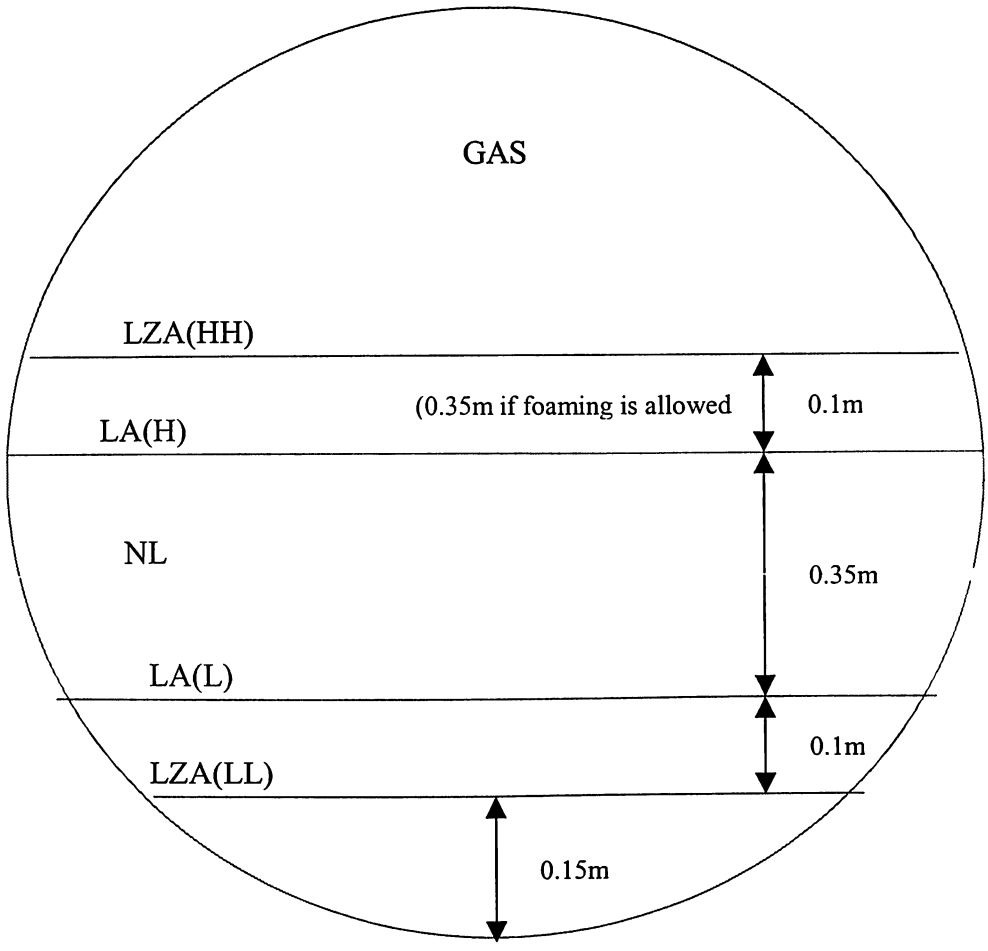
$$\rho_m v_m^2 \leq 3750$$

Liquid Outlet Nozzle:

The diameter of the liquid outlet nozzle d_3 shall be chosen such that the liquid velocity does not exceed 1m/s. The minimum diameter is 0.05 a vortex breaker is installed.

4.1.2 Level Control

(Step III)



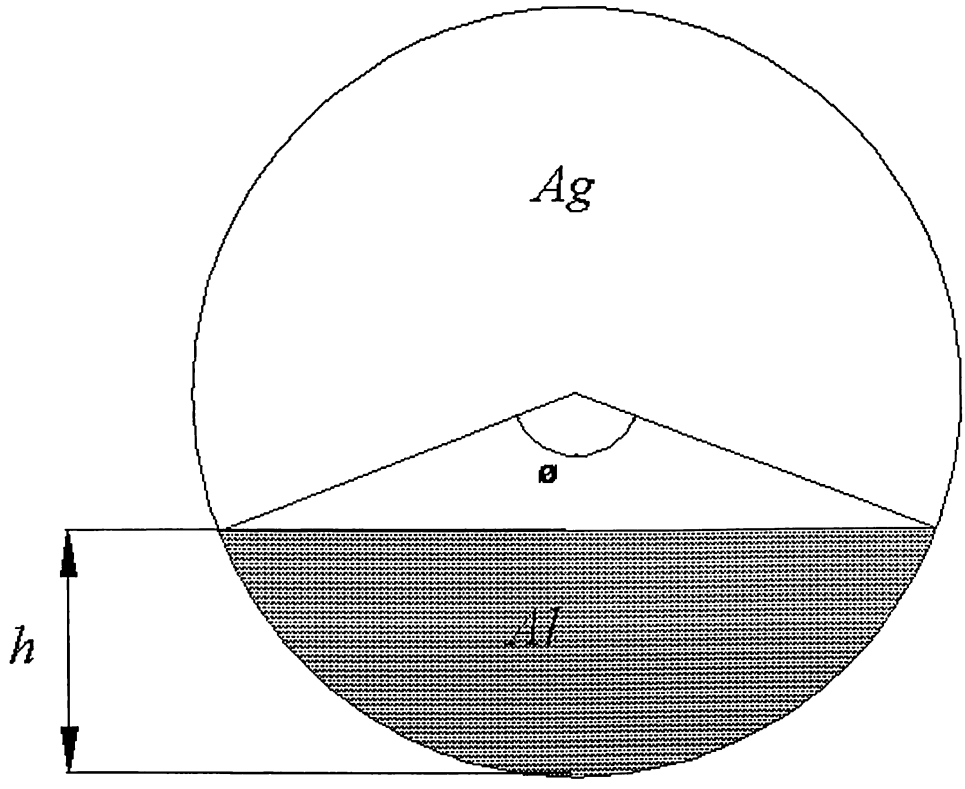
1. LZA (LL)-Low level trip; is 0.15 m above the bottom of the vessel i.e 0.15 from BTL (Bottom Tangent line) and 0.15m from bottom of the Horizontal Separator
2. LA(L) – Low level Prealarm; is either 0.1 m above LZA (LL) or if required, Located such that there is sufficient liquid hold time between the two levels for the operator intervention. Generally 1 minute is the control time for this level
3. LA(H) – High Level Pre-alarm; The minimum distance between LA(H) and LA(L) is 0.35. The distance between LA(L) & LA(H) shall be such that there is sufficient liquid hold up time for control purposes. This time is generally 3 minutes.
4. LZA(HH) – High level Trip; is either at least 0.1m above LA(H) or located such that there is sufficient liquid hold up time between the two levels for operator intervention

If the Liquid has the foaming tendency the distance between LA(H) and LZA(HH) has to be increased by further 0.25m

4.1.3 Vessel Geometrical Relationships

(Step IV)

The cross-section of a horizontal G/L separator with diameter D is shown in fig.



The positions of the liquid and gas phases have been indicated and have a cross-section area A_g, A_l.

The height of the G/L Interface is at h

For more general representation the area and heights are made dimensionless by dividing them by the vessel Cross-section and Vessel Diameter respectively

$$\text{i.e } A_l^* = \frac{A_l}{\pi/4 * D^2} \qquad A_g^* = \frac{A_g}{\pi/4 * D^2} \qquad h_l^* = \frac{h_l}{D}$$

A_l^* & A_g^* are now dimensionless chord areas with dimensionless chord heights h_l^*

We have the relationship as

$$A^* = \frac{0.5(\phi - \sin\phi)}{\pi} \qquad \text{and} \qquad h^* = 0.5(1 - \cos\phi/2)$$

To find A^* from h^*

Since $h^* = 0.5(1 - \cos\phi/2)$ find ϕ from this, then find A^* from $A^* = \frac{0.5(\phi - \sin\phi)}{\pi}$

To find h^* from A^*

Iterate ϕ to equate the following formula

$$\phi_{i+1} = \phi_i - \frac{(2\pi A - \phi_i + \sin\phi_i)}{\cos\phi_i - 1}$$

Once ϕ is found find h^* by $h^* = 0.5(1 - \cos\phi/2)$

4.1.4 Calculation of the Area between the control Levels

Foa given control levels find the individual areas and then algebraically add the areas to get the required areas between control levels.

Ex. $A_{h\_l} = A_h - A_l$
$A_{hh\_h} = A_{hh}$-A_h
$A_{l\_ll} = A_l$-A_{ll}

4.1.5 Calculation of the volumes between the control Levels

We have $\Delta V_{hd} = \frac{\alpha\pi D^3\{0.75(h_2^* - h_1^*) - (h_2^* - 0.5)^3 + (h_1^* - 0.5)^3\}}{6}$

Where h_1^* and h_2^* are the Dimensionless lower and upper boundary Control levels i.e h_1 and h_2

Where $h_1^* = \frac{h_1}{D}$ and $h_2^* = \frac{h_2}{D}$

And α =0.5 for Semi elliptical Vessel

4.1.6 Calculation of Tangent-to-Tangent Length of the vessel. (Step V)

$$L = \frac{(V_{slug} + Q_l t_{h\_l} - 2\Delta V_{hd,h\_l})}{A_{h\_l}}$$

Where V_{slug} is the volume of the anticipated slug (if any) to be accommodated

t_{h_l} is the required control time between the levels LA(L) and LA(H) and can be calculated from the relationship presented above

4.1.7 Calculate $\frac{L}{D}$ ratio

- If $\frac{L}{D} < 2.5$ Take $L = 2.5D$ and go to Step VII
- If $\frac{L}{D} \geq 2.5$ and $\frac{L}{D} \leq 6$ Go to Step VII
- If $\frac{L}{D} > 6$ Then got to Step VI

4.1.8 Iterate the Value of LZA (HH)

(Step VI)

Increase or Decrease LZA (HH)

Take new value of LZA (HH) and go to step III and proceed until the condition in Step V Satisfies

4.1.9 Check for Control Times

(Step VII)

Check If Specified Control Time is satisfied between the levels by.

$$t_{l\_ll} = \frac{\left(lA_{l\_ll} + 2\Delta V_{hd,l\_ll}\right)}{Q_l} \geq \text{Specified control Time (Usually 1 minute)}$$

For Non-Foaming System:

$$t_{hh\_h} = \frac{\left(lA_{hh\_h} + 2\Delta V_{hd,hh\_h}\right)}{Q_l} \geq \text{Specified Control Time}$$

For Foaming System:

$$t_{hh\_h} = \frac{\left(lA_{(hh-0.25)\_h} + 2\Delta V_{hd,(hh-0.25)\_h}\right)}{Q_l} \geq \text{Specified Control Time}$$

If the control times are met increase the width of the Associated control band by multiplying with the ratio of Specified and calculated control times and return to Step III

4.1.10 Degassing and Defoaming Criterion.

(Step VIII)

For degassing Criterion:

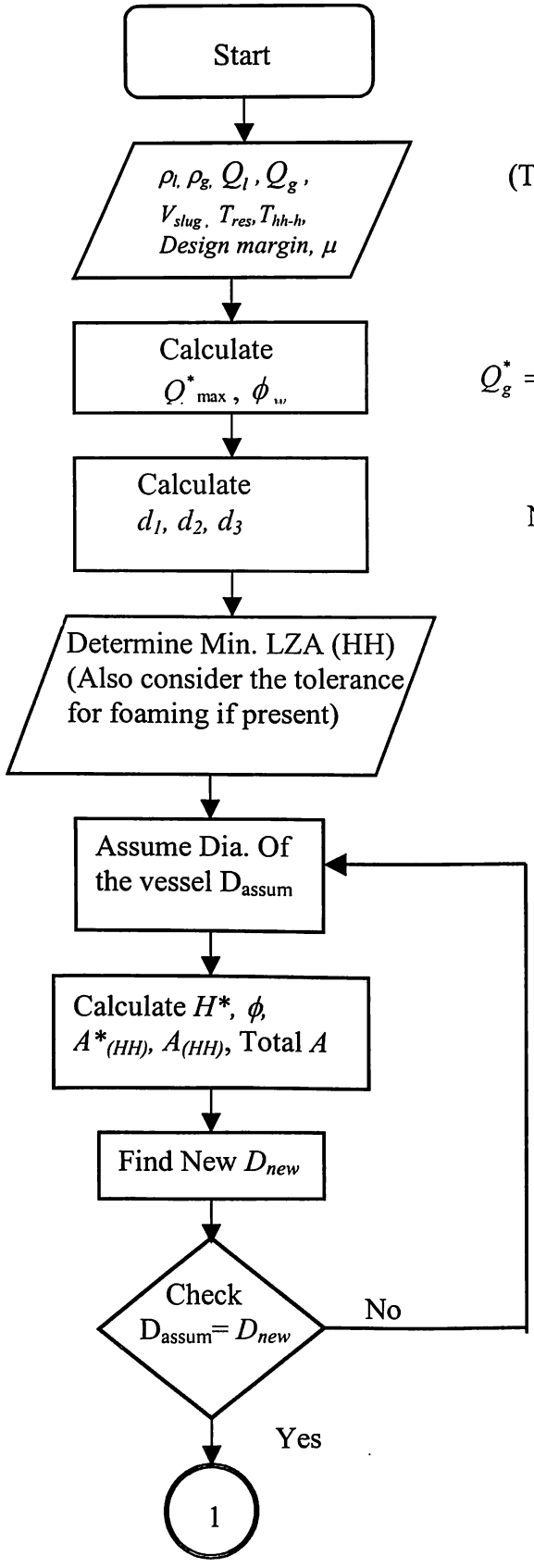
$$\text{Check } D > \frac{4.5x10^7 Q_{L\max}\mu_l}{(\rho_l - \rho_g)L}$$

For De-Foaming following Criterion should be satisfied.

$$D > 7000Q_{L\max}\left(\frac{\mu_l}{(\rho_l - \rho_g)}\right)^{0.27} * \frac{1}{L}$$

[A Detailed Flow chart is given next page]

4.2 The flow chart for the design of Horizontal Separator



(These values are given)

$$Q_g^* = Q_g.(\frac{\rho_G}{\rho_l - \rho_G})^{0.5} \quad \text{and} \quad \phi_w = \frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$$

Nozzle Sizing

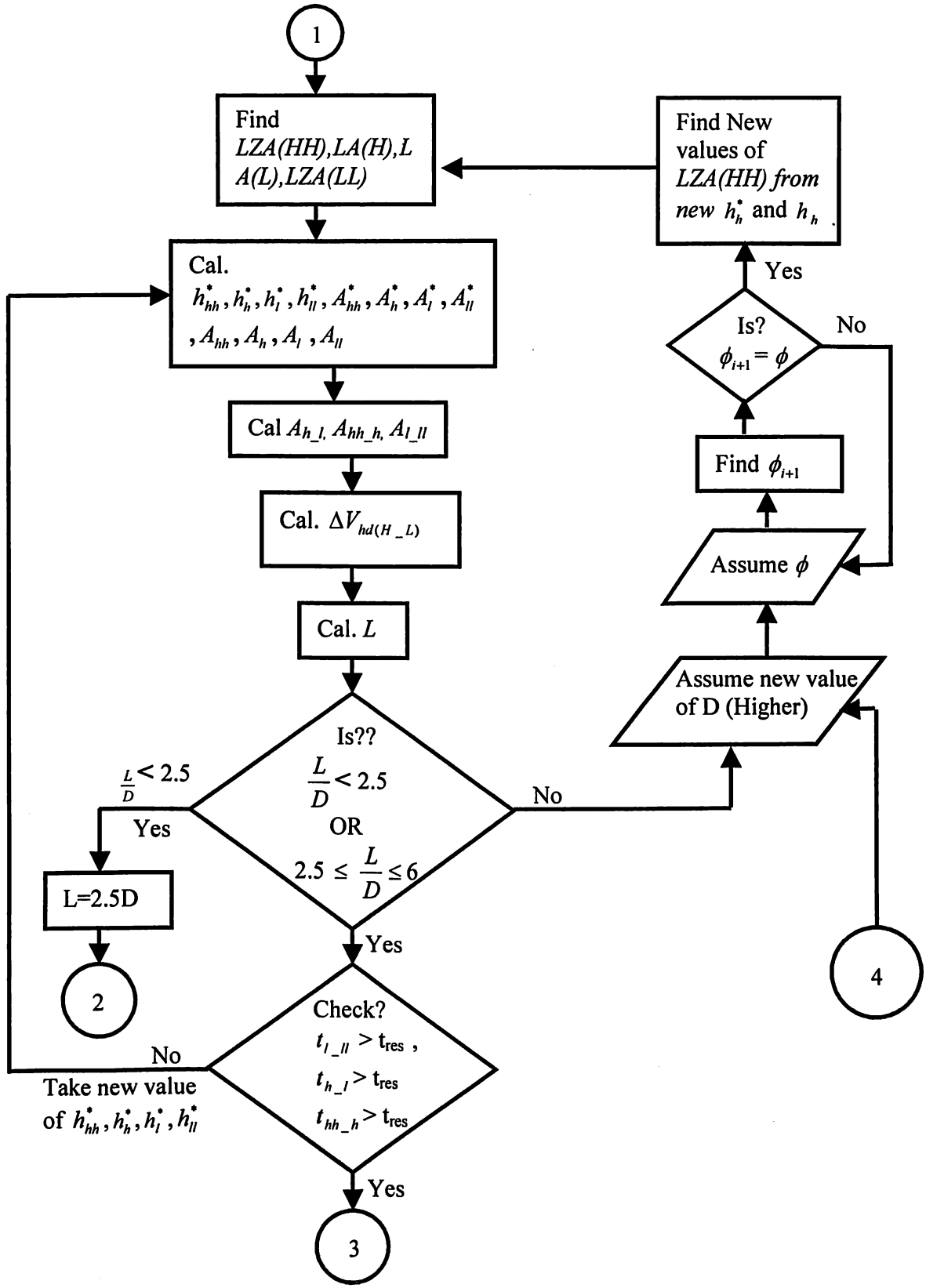
By considering the Level Control

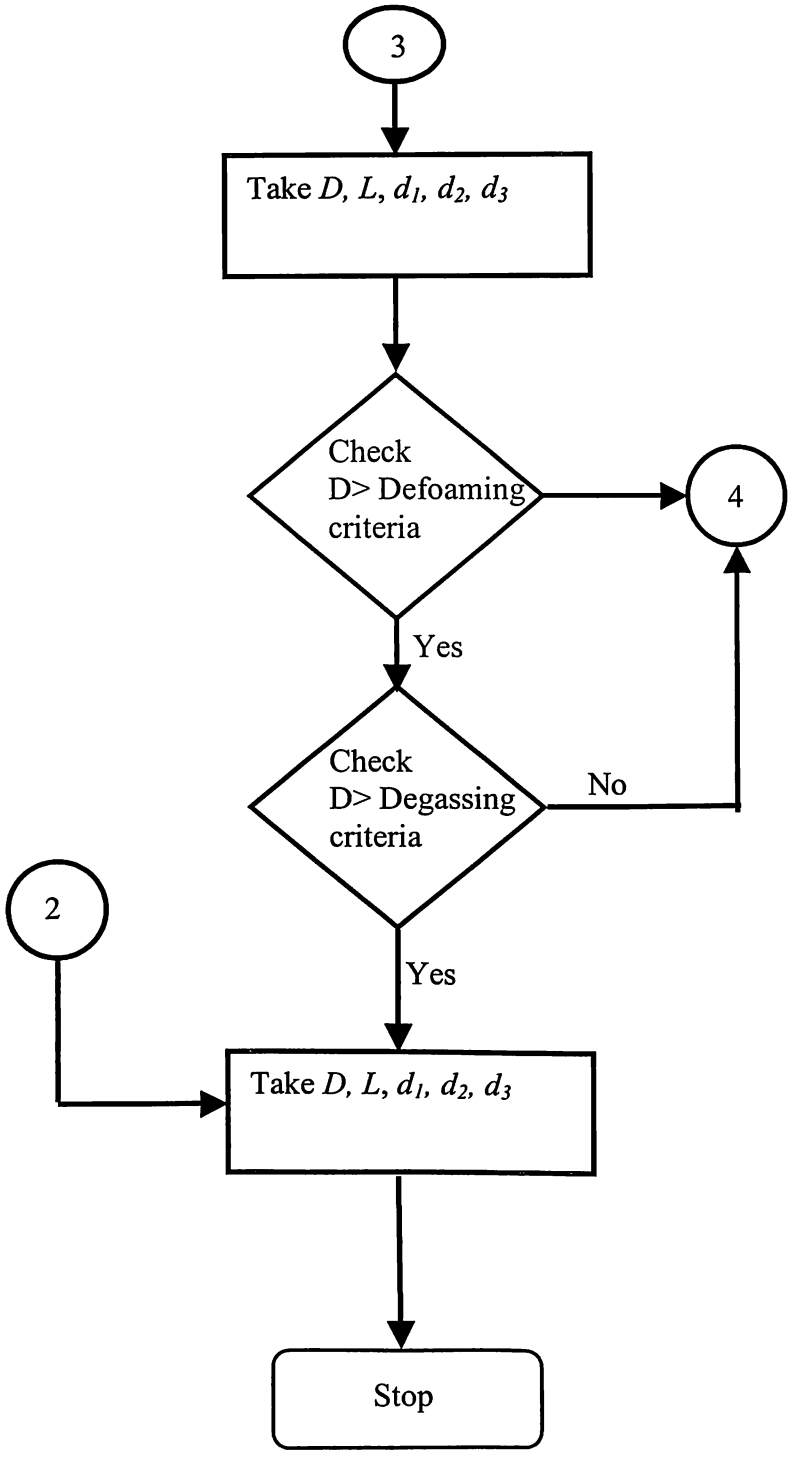
$$A_l^* = \frac{A_l}{\pi/4 * D^2}$$

$$A_g^* = \frac{A_g}{\pi/4 * D^2} \qquad h_l^* = \frac{h_l}{D}$$

$$A^* = \frac{0.5(\phi - \sin\phi)}{\pi}$$

$$h^* = 0.5(1 - \cos\phi/2)$$


1
Find LZA(HH),LA(H),LA(L),LZA(LL)
Cal. $h^*_{hh}, h^*_h, h^*_l, h^*_{ll}, A^*_{hh}, A^*_h, A^*_l, A^*_{ll}, A_{hh}, A_h, A_l, A_{ll}$
Cal $A_{h\_l}, A_{hh\_h}, A_{l\_ll}$
Cal. $\Delta V_{hd(H\_L)}$
Cal. L
Is?? $\frac{L}{D} < 2.5$ OR $2.5 \le \frac{L}{D} \le 6$
$\frac{L}{D} < 2.5$
Yes
L=2.5D
2
No
Take new value of $h^*_{hh}, h^*_h, h^*_l, h^*_{ll}$
Yes
Check? $t_{l\_ll} > t_{res}$, $t_{h\_l} > t_{res}$ $t_{hh\_h} > t_{res}$
Yes
3
No
Assume new value of D (Higher)
Assume ϕ
Find ϕ_{i+1}
Is? $\phi_{i+1} = \phi$
No
Yes
Find New values of LZA(HH) from new h^*_h and h_h
4


3
Take D, L, d1, d2, d3
Check
D> Defoaming
criteria
4
Yes
Check
D> Degassing
criteria
No
2
Yes
Take D, L, d1, d2, d3
Stop

4.3 Design of Three-Phase Separator

The design of the vessel of the three-phase separator is same as the two-phase separator, but some additional requirements are to be considered. The three-phase separator has plate packs, which needs to be designed first. The liquid level control is different here as compared to two-phase separators.
The detail design principles are presented here.

4.3.1 Design Of Plate packs

The function of the plate pack is to improve the efficiency of liquid / Liquid separation. Installation of Plate packs in settlers will also result in smaller vessels as compared to open settlers
The main Reasons for this increase in efficiency are:

1. The presence of plate pack results in a substantial reduction of the settling distance of the droplets.
2. If the platepack is properly sized, the flow between the plates is laminar.
   (By comparison, in an open settler the flow is nearly always turbulent which hampers the settling process.)

A plate pack is only effective for the separation of primary dispersions i.e. Droplets have to be larger then 30μm to satisfy the above criterion. If secondary dispersion (droplets <30μm) is present, separation can only be done by coalescers.
The platepack can be arranged as crossflow type, co-current type and countercurrent flow type.
Platepacks should be mounted in panels between the panels at least one gutter is present to transport the coalesced separated phase to its bulk phase.
Measures should be taken to prevent by passing of the platepack such as mounting a platepack in front of gutters and sealing the clearance between the plate pack and the vessel wall.

Design of the cross flow plate pack.
Sizing of the front area of the plate pack:
In cross flow plate pack with flat plates, flow is no longer fully laminar at a Reynolds number higher than 850.

We have $R_e = \dfrac{2d_{pp}Q_c\rho_c}{\mu_c Af_{net}}$

Where d_{pp} = Distance between two plates.

Q_c = Volumetric flow rate of the continuous phase

ρ_c =Density of the continuous phase

μ_c = Dynamic viscosity of the continuous phase.

But $Af_{net} = \dfrac{2d_{pp}Q_c\rho_c}{\mu_c R_e}$ i.e. $(Af_{net})_l = \dfrac{2d_{pp}Q_l\rho_l}{\mu_l R_e}$ and $(Af_{net})_h = \dfrac{2d_{pp}Q_h\rho_h}{\mu_h R_e}$

$$\Rightarrow Af_{Gross} = \frac{[(Af_{net})_l + (Af_{net})_h](t_{pp} + d_{pp})}{d_{pp}F_{loss}}$$

Where t_{pp} -> Plate thickness $\approx$1mm

F_{loss} -> Correction Factor to account for constructional elements, risers etc $\approx 0.9 \sim .95$

And

$$H_{pp} = \frac{Af_{gross}}{w_{pp}} + H_{contot} = H_l + H_h + H_{contot} \qquad \text{Where, } H_l, H_h \geq 0.3$$

and

$$H_{contot} = \sum H_{con} + H_{db} = (H_{con})_l + (H_{con})_h + H_{db} = \frac{(Q_l t_{con})_l}{\pi / 4D^2} + \frac{(Q_h t_{con})_h}{\pi / 4D^2} + H_{db}$$

Where w_{pp} =Width of the plate pack

H_{db} = Dispersion band width $\approx$0.2m

t_{con} = Specified Control time for Liquid-Liquid Control

Plate pack length, L_{pp}

$$L_{pp} = \frac{V_{pp} d_{pp}}{Vp_{set} \cos\theta} + ft_l L_{entry} \geq 0.3m$$

Where; V_{pp} -> Mean velocity of Total Liquid Flow. $= \frac{Q_c}{Af_{net}}$

Vp_{set} -> Settling velocity of smallest droplet,

If Diameter of droplet $d_p \geq 40\mu$m and the dispersed phase concentration is <1% and Re <= 1 then

$$Vp_{set} = \frac{|\rho_d - \rho_c| g d_p^2}{18\mu_c} - 0.05V_{cax}$$

V_{ax} =Axial Velocity of flow through plate pack

$L_{entry} = 0.02 d_{pp} R_e \approx 16 d_{pp}$

θ = angle of plate with the horizontal plane

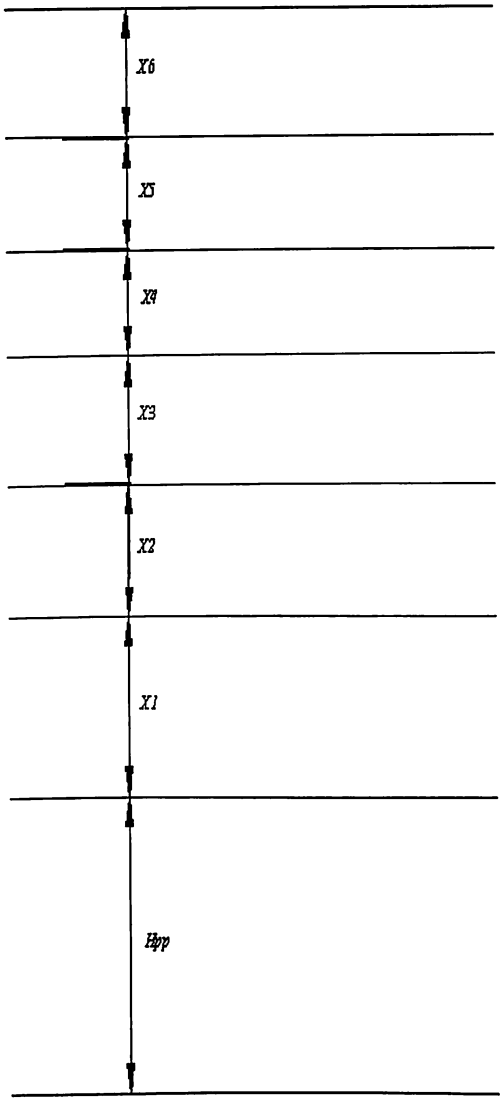
ft_l=0.5

Choice of Plate type, Plate spacing and plate angle:

For Clean service, -- $\theta = 45^0$

For Solids Presence, -- $\theta = 60^0$

4.3.2 Level Control For GLL Separators:



X6 is the distance between top of demisting internals and TTL X6 ≈0.15D ≥0.15m

X5 is the height of the demisting material
If Mistmat is there X5=0.1m

X4 is distance between the top of the schoepentoeter and the bottom of the demisting device, X4 = d_l ≥0.3m

X3 is height of Schoepentoeter. X3 = d_l + 0.02m
d_l is feed nozzle diameter.

X2 is the distance between LHZ(HH) & bottom of Schoepentoeter X2 = 0.05D

X1 is the sum of the control bands required for GL Level control.

$$H_{pp} = H_l + H_h + H_{contot}$$

$$H_l = \frac{(Af_{gross})_l}{w_{pp}} \qquad H_h = \frac{(Af_{gross})_h}{w_{pp}} \Bigg\} \geq 0.3\text{m}$$

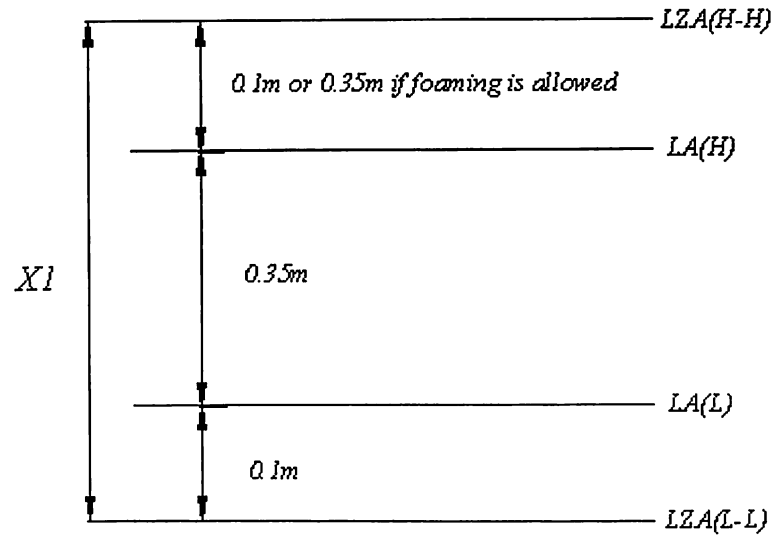
Af_{gross} → Gross frontal areas of the plate pack section associated with cleaning of the light and heavy phase.

$H_{contot} = \sum H_{con} + H_{db}$ H_{db} → is the sum of the heights of control bands required for LL control

$$H_{contot} = \frac{(Q_l t_{con})_l}{\pi/4D^2} + \frac{(Q_h t_{con})_h}{\pi/4D^2} + H_{db}$$

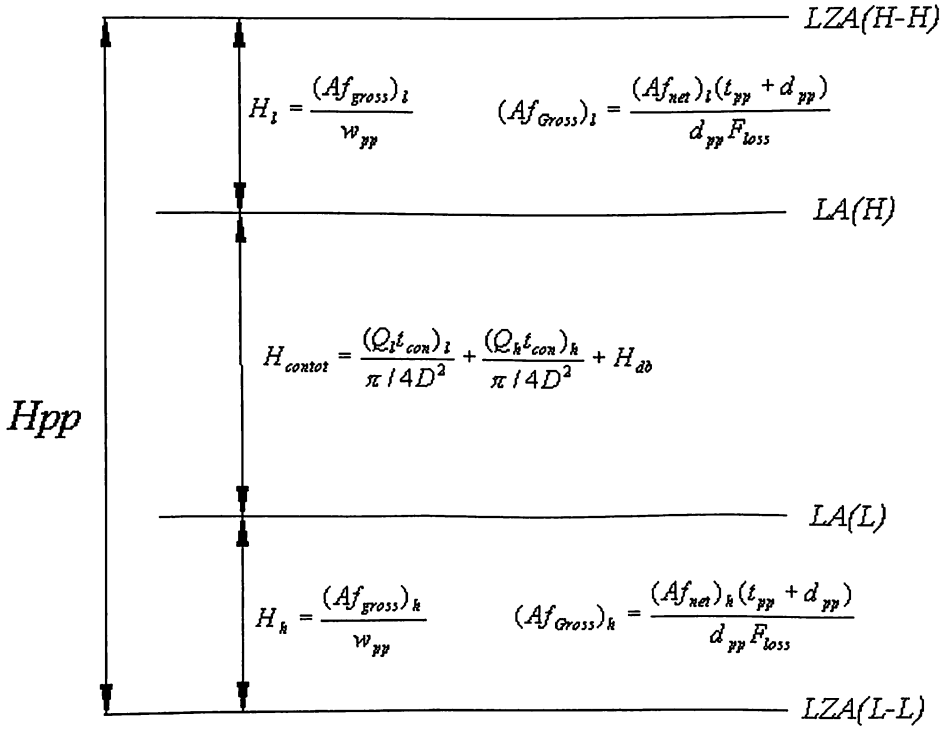
4.3.3 Finding the Sum of the control band for the GL Level Control i.e. X1

We have;



The values given here are for the minimum distance between the levels the final values are found after checking the control time for each level.

4.3.4 Finding the Sum of the control band for the L-L Level Control i.e. *Hpp*



4.3.5 Sizing of the Feed and outlet nozzles.

Feed/ inlet nozzle:

In LL Separator:

The Inner diameter of the feed nozzle should be equal to the inner diameter of the inner piping but shall also be sufficiently large, so that the liqud velocity does not exceed 1m/s.

To promote even distribution of the feed flow into the separator the feed nozzle should be equipped with a feed inlet device, such as an elbowed pipe or a slotted pipe.

In GLL Separator:

The inner diameter of the feed nozzle should be equal to the inner diameter of the inner piping but shall also be sufficiently large to satisfy the relevant criterion.

1. If no Inlet device is used:

$$\rho_m v_m^2 \leq 1000$$

Where ρ_m → Mean Density of the Mixture of the feed pipe given by

v_m → Mean velocity of the mixture in inlet Nozzle

We have $$\rho_m = \frac{\rho_l Q_l + \rho_g Q_g}{Q_l + Q_g}$$

and $$v_m = \frac{Q_l + Q_g}{\pi / 4 * d_1^2}$$ Where d_1 → Diameter of the Inlet Pipe.

2. If Half Open Pipe is used as inlet device:

$$\rho_m v_m^2 \leq 1500$$

3. If Schoepentoeter is used as inlet device:

$$\rho_m v_m^2 \leq 6000$$

Gas Outlet Nozzle:

The diameter of the gas outlet nozzle should normally be taken as equal to that of the outlet pipe, but also the following criterion shall be satisfied.

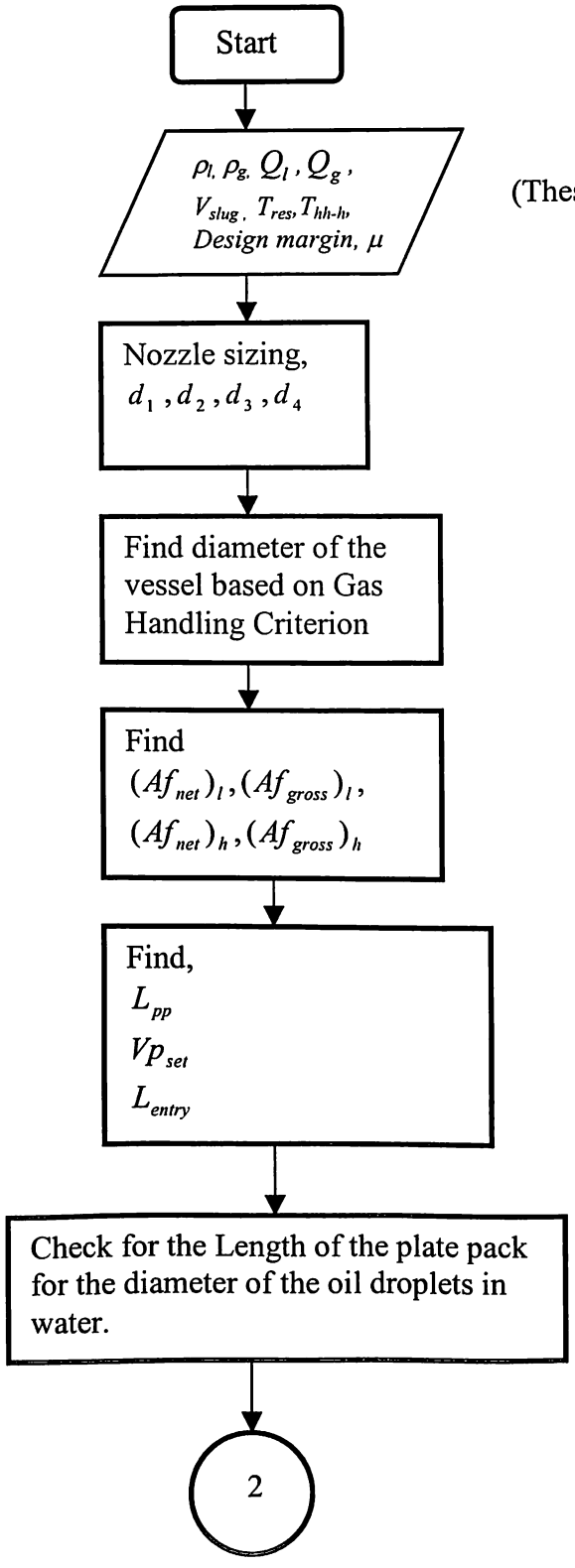
$$\rho v_m^2 \leq 3750$$

Liquid Outlet Nozzle:

The diameter of the liquid outlet nozzle shall be chosen such that the liquid velocity does not exceed 1 m/s. The minimum diameter is 50mm or 2"

4.4 Flow chart

Designing Vertical Three phase settler with Plate pack.



(These values are given)

$$Af_{net} = \frac{2d_{pp}Q_c\rho_c}{\mu_c R_e}$$

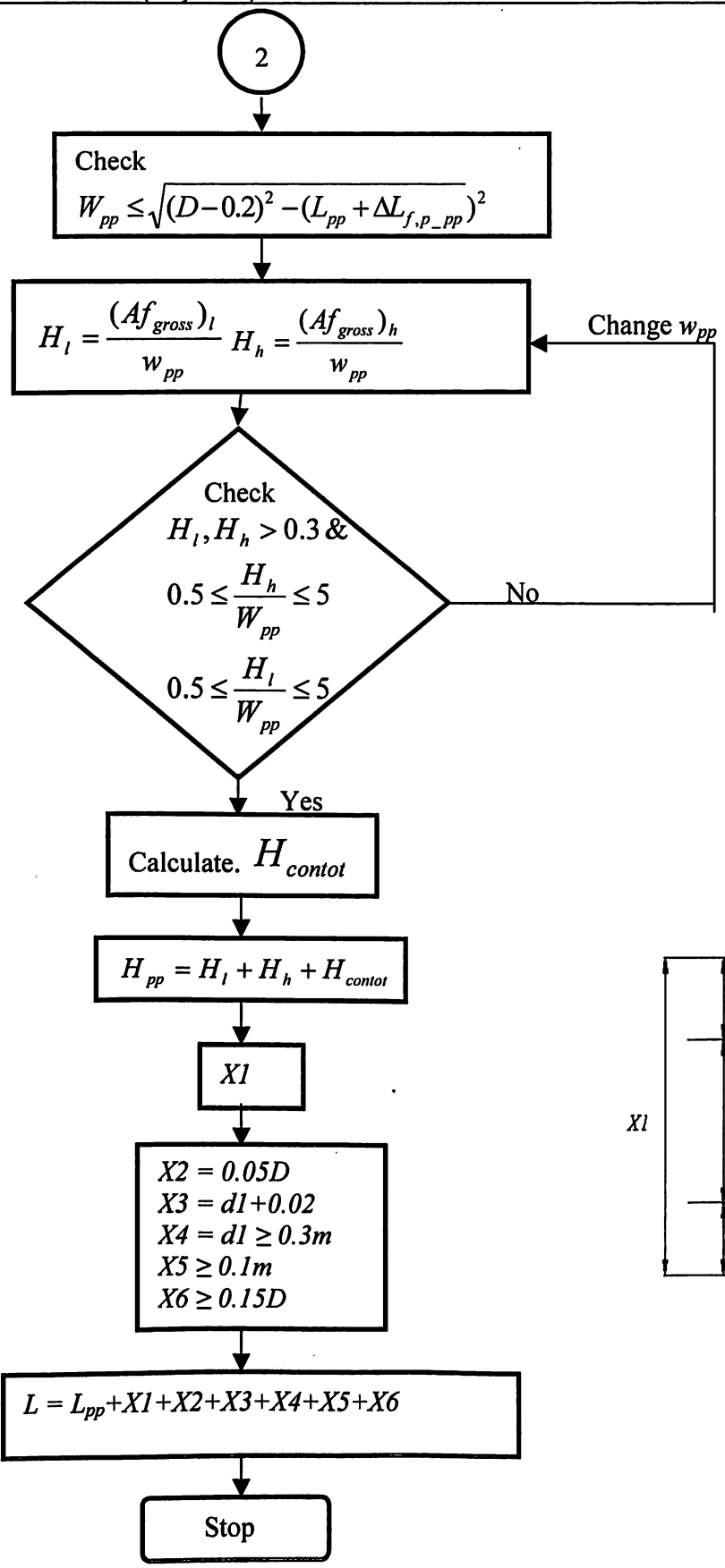
$$Af_{Gross} = \frac{[(Af_{net})_l + (Af_{net})_h](t_{pp} + d_{pp})}{d_{pp}F_{loss}}$$

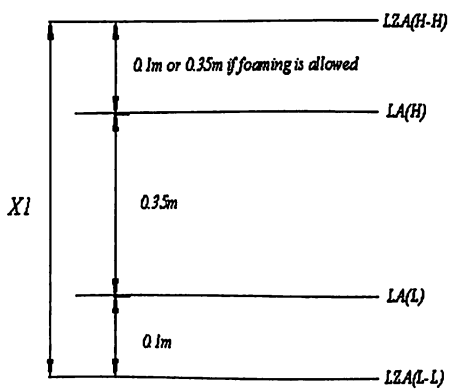
$$L_{pp} = \frac{V_{pp}d_{pp}}{Vp_{set}\cos\theta} + ft_l L_{entry} \geq 0.3m$$

$$Vp_{set} = \frac{|\rho_d - \rho_c|gd_p^2}{18\mu_c} - 0.05V_{cax}$$

$$L_{entry} = 0.02d_{pp}R_e \approx 16d_{pp}$$

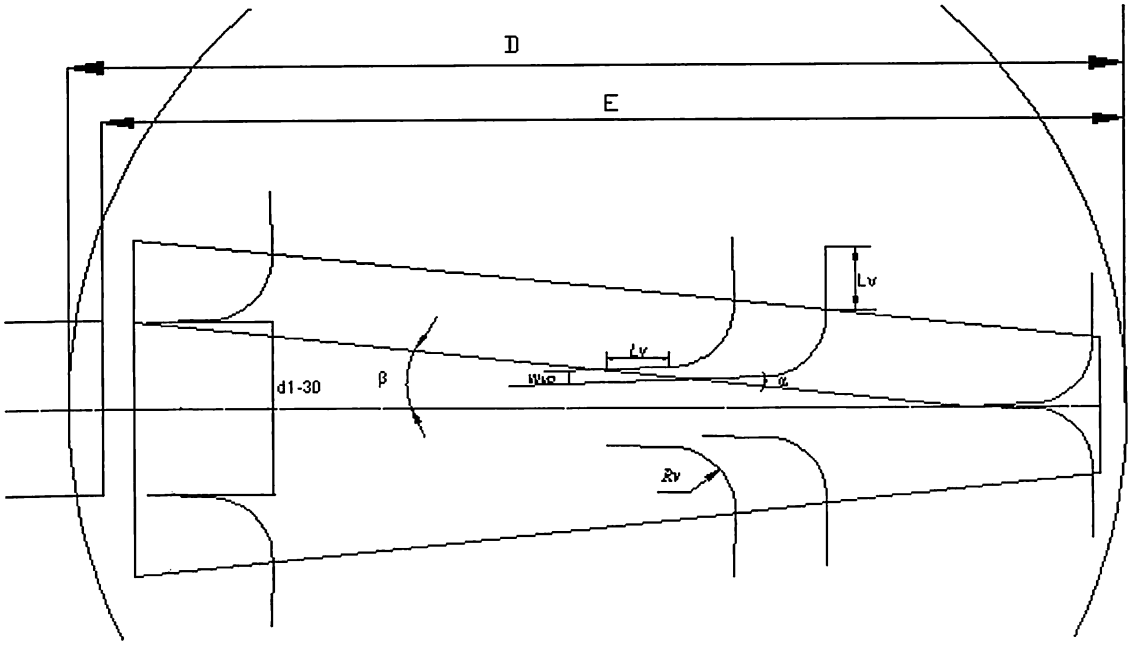
$$V_{pp} = \frac{Q_c}{Af_{net}} \qquad ft_l = 0.5$$


2
Check
$W_{pp} \le \sqrt{(D-0.2)^2 - (L_{pp} + \Delta L_{f,p\_pp})^2}$
$H_l = \frac{(Af_{gross})_l}{w_{pp}}$ $H_h = \frac{(Af_{gross})_h}{w_{pp}}$
Change w_{pp}
Check
$H_l, H_h > 0.3$ &
$0.5 \le \frac{H_h}{W_{pp}} \le 5$
$0.5 \le \frac{H_l}{W_{pp}} \le 5$
No
Yes
Calculate. H_{contot}
$H_{pp} = H_l + H_h + H_{contot}$
X1
X2 = 0.05D
X3 = d1+0.02
X4 = d1 ≥ 0.3m
X5 ≥ 0.1m
X6 ≥ 0.15D
L = Lpp+X1+X2+X3+X4+X5+X6
Stop


LLA(H-H)
0.1m or 0.35m if foaming is allowed
LA(H)
X1
0.35m
LA(L)
0.1m
LLA(L-L)

4.5 Design of Schoepentoeter

SCHEMATIC OUTLINE OF THE SCHOEPENTOETER



α = vane angle, angle made by straight part of vanes with center line
β = edge angle, angle made by edge of the row of vanes with center line
D = vessel inside diameter, mm
d_1 = inlet nozzle inner diameter, mm
E = available space, mm
L_v = length of straight part of vanes (normally 75, 100, 150 or 200 mm)
n_v = number of vanes per side
R_v = vane radius, mm (normally 50 or 100 mm)
t = vane material thickness, mm (normally 3 mm, but typically 5 mm for heavy duty, e.g slugs)
W_{vo} = width of vane entrance opening, mm

i) For a new vessel the inside diameter D and the inner diameter of the feed nozzle d_1 will be determined by process considerations.
For an existing vessel, D and d_1 will be known.
Schoepentoeters are not used in vessels of diameter less than 500 mm.
Schoepentoeters are only fitted on nozzles with $d_1 \geq 150$ mm.
If nozzle diameters $d_1 > 1/3$ D are encountered the Principal should be consulted.

ii) Evaluate the available space E.
For a vertical vessel, take $E = D - 50$ ($D \leq 2\,000$ mm)
$E = D - 100$ ($2\,000 < D \leq 4\,000$ mm)
$E = D - 200$ ($D > 4\,000$ mm)

In Schoepentoeter assembly type IV, D/2 instead of D shall be taken in the expression for E and each Schoepentoeter of the type IV configuration shall be designed with the general sizing rules given below.
If $E > 5d_1$ then the available space is excessive and a shortened Schoepentoeter (Types II to IV) should be used; in which case the required space, E, should be reassessed.

For a horizontal vessel, $3\ d_1 \leq E \leq 5\ d_1$
It should be noted that the **inner** (rather than the outer) diameter of the feed nozzle is taken in designing the Schoepentoeter.

iii) Evaluate $X = (E - 270)/(d_1 - 30)$ and select L_v and R_v from the table below:

X	E (mm)	L_v (mm)	R_v (mm)
≤ 2.5	≤ 550	75	50
≤ 2.5	> 550	100	50
$2.5 < X \leq 6$	> 550	100	100
$6 < X \leq 20$	> 550	150	100
$X > 20$	> 550	200	100
< 6	$> 6\ 000$	400	300 *

iv) Calculate number of vanes per side, n_v, from:
$n_v = (E - R_v - 70) / L_v$
Rounded down to the nearest integer.

v) Evaluate $\tan \beta = (d_1 - 30) / \{ 2(n_v-1)L_v \}$

vi) Choose $\alpha = 8$ degrees initially (maximum value)

vii) Evaluate W_{vo} from

$W_{vo} = L_v (\sin \alpha + \cos \alpha \tan \beta) - t$ mm

W_{vo} should be in the range of $W_{vo,\,min}$ to $W_{vo,\,max}$:

if R_v is 50 or 100 mm: $20 \leq W_{vo} \leq 30$ mm

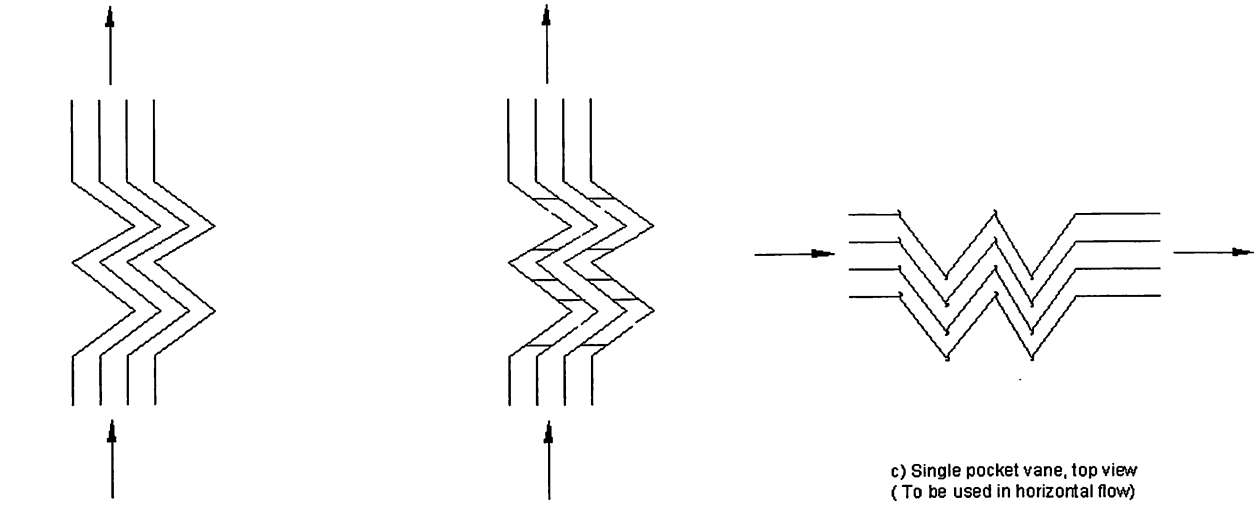
if R_v is 300 mm: $40 \leq W_{vo} \leq 80$ mm

viii) if R_v is 50 or 100 mm:
If $W_{vo} > 30$, reduce α in steps of 1 degree (minimum value 0°) until $W_{vo} \leq 30$mm
If $W_{vo} < 20$ mm, reduce the Schoepentoeter length, i.e. reduce n_v.
If $20 \leq W_{vo} \leq 30$ mm, the design is finished and n_v, R_v, L_v and α are selected.

ix) if R_v is 300 mm:
If $W_{vo} > 80$, reduce α in steps of 1 degree (minimum value 0°) until $W_{vo} \leq 80$mm
If $W_{vo} < 40$ mm, reduce the Schoepentoeter length, i.e. reduce n_v.
If $40 \leq W_{vo} \leq 80$ mm, the design is finished and n_v, R_v, L_v and α are selected.

4.6 Design of vane pack assembly:

The vane type design uses an intricate array of metal plates, called vanes. The vane type mist extractor is mounted such that the gas stream flows horizontally through the vanes. During this flow a change in direction is induced several times, resulting in a centrifugal action that aids the primary impingement separation mechanism in removing the finer liquid droplets entrained in the gas. The liquid droplets are forced into the liquid collection pockets entrained in the gas. The liquid droplets are forced into the liquid collection pockets, out of the gas flow path, and drain out by gravity. The pressure drop across a vane type mist eliminator is very small, it can handle solids in the flowing gas stream and can remove droplets down to about 40μm in size.
The vanes in vane packs can be either of no pocket (straight), single-pocket or double-pocket type.



a) Straight Vane

b) hollow vane side vane
(To be used in vertical flow)

c) Single pocket vane, top view
(To be used in horizontal flow)

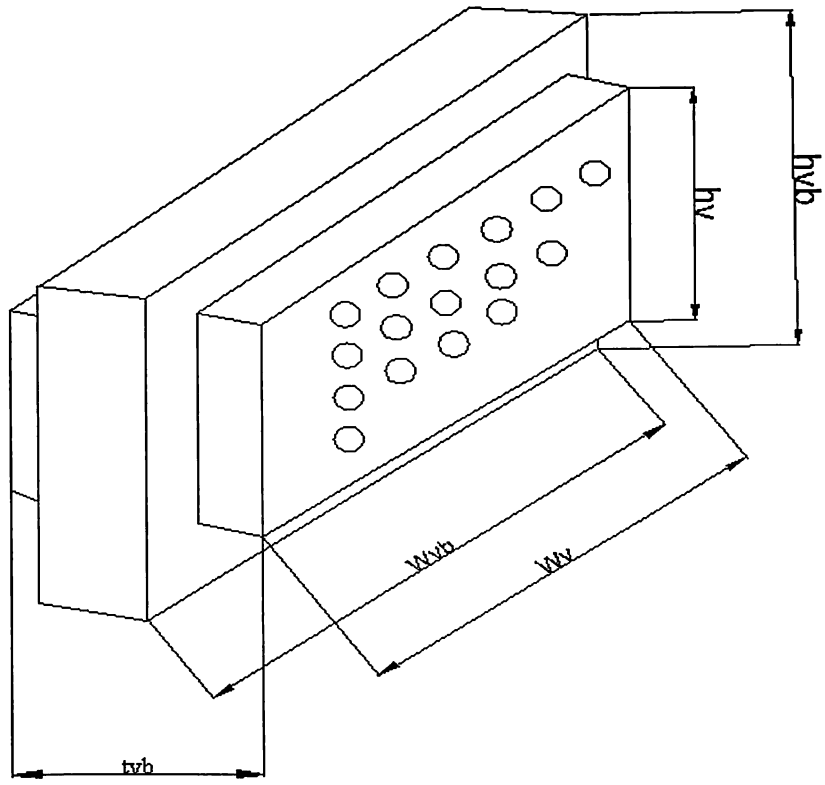
d) Double-Pocket Vane: Top view
(To be used in Horizontal flow)

Of the three types the straight vane has the lowest liquid separation efficiency. The double-Pocket vane type has the highest sensitivity to fouling but its efficiency above λ_{max} (Where liquid re-entrainment starts) Deteriorates to a lesser degree than with the other types because the separated liquid is better shielded from the gas flow.

In horizontally flowed through vane packs are equipped with either straight or double-pocket vanes. A vertically flowed through vane pack with double-Pocket vanes. A vertically flowed through vane pack with double-pocket vanes is only recommended, if the service is clean. Since it has a higher gas handling capacity than mistmats and requires little height(At most about 0,2) it is suitable retrofitting device to upgrade undersized wiremesh demister by installing it downstream of the mistmat.

Layout of the Vane pack

Inline separator with –Horizontal flow vane pack.



Initially take the height h_v as $h_v = \sqrt{1.5A_v}$ Where A_v = Vane pack face area $A_v = \frac{Q^*_{max}}{\lambda_{max}}$

If $\frac{Q_l\sqrt{\rho_l}}{Qg\sqrt{\rho_g}} < 0.01$ Or even if Not exactly known take as default $\frac{Q_l\sqrt{\rho_l}}{Qg\sqrt{\rho_g}} = 0.01$

h_v should be adjusted to fall within the following range

$$\boxed{0.3 \le h_v \le 1.5.}$$

The Vane pack width:

$$W_v = \frac{A_v}{h_v}$$

The width of the vane pack box:

$$W_{vb} = W_v + 0.1$$

The height of the vane pack, h_{vb} shall include a margin to obtain sufficient coverage of the vanes in order to prevent vapor by passing the demister. Also sufficient height shall be available to allow proper draining of the separated liquid.

Typically, $h_{vb} = h_v + 0.3$

Liquid shall be drained from the vane pack to the bottom compartment of the vessel via drain packs having a minimum diameter of 0.05. At least one pipe for each meter of vane pack width shall be used. The drain pipe shall extend at least 0.1m below LZA(LL) for sealing purposes.
The depth of the vane box, T_{vb} is dependent on the type of vane selected and is normally between 0.3 and 0.45m

Two Stage separators with horizontal flow vane pack:

$$W_v = \sqrt{(D-0.2)^2 - t_{vb}^2} - 0.1$$

The depth of the vane box, t_{vb} is dependent on the type of vane selected and ranges typically between 0.3and 0.45m
The corresponding vane height, h_v is calculated using the required vane pack face area,

A_v as determined with ϕ_v=0.01

$$h_v = \frac{A_v}{W_v}$$

The dimensions W_v and h_v should be adjusted to fulfill the condition

$$0.3 \le h_v \le 1.5$$

Further $W_{vb} = W_v + 0.1$ and

$$h_{vb} = h_v + 0.3$$

4.7 Design Margins for separator design:

To determine the highest envisaged gas and liquid load for vessel design, design margins (surge factors) are required.

In Exploration and production applications.

1: Offshore service:

Separator handling natural flowing production from

A: Its own platform.	1.2
B: Another Platform or well jacket in shallow water	1.3
C: Another platform or well in deep water	1.4

Separator handling gas lifted production from

A: Its own platform.	1.4
B: Another platform	1.5

2: Onshore Service:

Separator handling natural flowing production or gas plant inlet separator in:

A: Flat or low rolling country

B: hilly country

Separator handling gas lifted production in

A: Flat or low rolling country	1.4
B: Hilly country	1.5

In refineries and chemical plants the design margins ranges typically from 1.15 to 1.25

Chapter 5

Equilibrium Calculations for Three Stage Separators

5.1 Equilibrium Calculations for Three-Stage Separators.

The compositions and relative abundance of the separation products can be predicted either by laboratory tests, or in a fair approximation by theoretical considerations and using diagrams based on practical experience. In calculations stream and the discharged oil and gas do not vary in composition with time, that separation is of the flash type and that the system is in thermodynamic equilibrium at given temperature and pressure. The calculations are based on following three equations.

$$n = n_l + n_g$$

$$Z_i n = n_l X_i + n_g Y_i$$

$$K_i = \frac{Y_i}{X_i}$$

Where; $n \rightarrow$ Total number of moles in Liquid-gas system

$n_l \rightarrow$ Number of Moles in Liquid Phase

$n_g \rightarrow$ Number of moles in Gas phase

$Y_i \rightarrow$ Mole fraction of the i_{th} component of gas

$X_i \rightarrow$ Mole fraction of the i_{th} Component of Liquid

$K_i \rightarrow$ Equilibrium Ratio of the i_{th} component in the Liquid gas system

The total number of moles in the system equals the number of moles in the liquid and gas phase taken separately. The total number of moles of any component in the system equals the number of moles of that component taken separately in the liquid and gas phases.

$$\text{i.e. } \sum_{i=1}^{m} X_i = \sum_{i=1}^{m} Y_i = \sum_{i=1}^{m} Z_i = 1$$

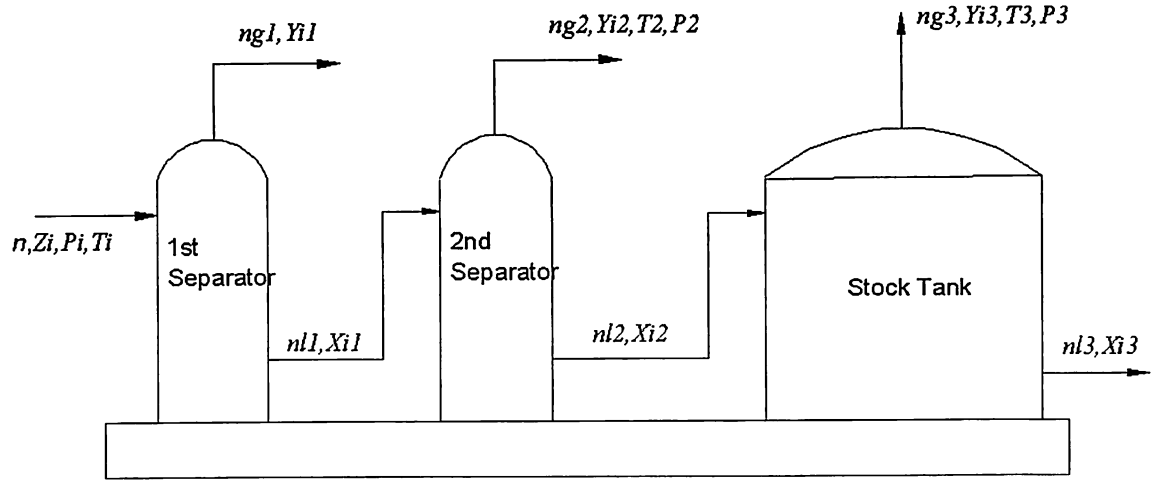
Dividing this equation by n on both sides

$$Z_i = \frac{n_l}{n} X_i + \frac{n_g}{n} Y_i$$

Let $\frac{n_l}{n} = \bar{n}_l$ and $\frac{n_g}{n} = \bar{n}_g$ $\qquad Z_i = \bar{n}_l X_i + \bar{n}_g Y_i$

$$X_i = \frac{Z_i}{1 + \bar{n}_g (K_i - 1)} \qquad \& \qquad Y_i = \frac{Z_i}{1 + \bar{n}_g (\frac{1}{K_i} - 1)}$$

i.e. $$\sum X_i = \sum \frac{Z_i}{1 + \bar{n}_g (K_i - 1)} = 1 \; \& \qquad \sum Y_i = \sum \frac{Z_i}{1 + \bar{n}_l (\frac{1}{K_i} - 1)} = 1$$



Three stage Separation

For 1st Separator we have;

$$\sum X_i 1 = \sum \frac{Z_i}{1 + \bar{n}_{g1}(K_{i1} - 1)} = 1 \qquad \& \qquad \sum Y_i 1 = \sum \frac{Z_i}{1 + \bar{n}_{l1}(\frac{1}{K_{i1}} - 1)} = 1$$

Where $\bar{n}_{l1} = \frac{n_{l1}}{n}$ and $\bar{n}_{g1} = \frac{n_{g1}}{n}$

For Second Separator we have;

$$\sum X_i 2 = \sum \frac{X_i 1}{1 + \bar{n}_{g2}(K_{i2} - 1)} = 1 \qquad \& \qquad \sum Y_i 2 = \sum \frac{X_i 1}{1 + \bar{n}_{l2}(\frac{1}{K_{i2}} - 1)} = 1$$

Where $\bar{n}_{l2} = \frac{n_{l2}}{n_{l1}}$ i.e. $n_{l2} = \bar{n}_{l2} . \bar{n}_{l1} . n$.

& $\bar{n}_{g2} = \frac{n_{g2}}{n_{g1}}$ i.e. $n_{g2} = \bar{n}_{g2} . \bar{n}_{g1} . n$

For Stock Tank we have;

$$\sum X_i 3 = \sum \frac{X_i 2}{1 + \bar{n}_{g3}(K_{i3} - 1)} = 1 \qquad \& \qquad \sum Y_i 3 = \sum \frac{X_i 2}{1 + \bar{n}_{l3}(\frac{1}{K_{i3}} - 1)} = 1$$

Where $\bar{n}_{l3} = \frac{n_{l3}}{n_{l2}}$ i.e. $n_{l3} = \bar{n}_{l3} \bar{n}_{l2} . \bar{n}_{l1} . n$.

& $\bar{n}_{g3} = \frac{n_{g3}}{n_{g2}}$ i.e. $n_{g3} = \bar{n}_{g3} \bar{n}_{g2} . \bar{n}_{g1} . n$

5.1.1 Gas to Oil Ratio (GOR)

GOR is defined as the ratio of the gas discharged from first 2 stages to the volume of liquid collecting in the 3rd stage (Stock Tank oil)

I.e. $$GOR = \frac{(n_{g1} + n_{g2})V_{mol}}{n_{l3}\frac{M_{l3}}{\rho_{l3}}}$$

Dividing numerator and denominator by n

i.e. $$GOR = \frac{\frac{(n_{g1} + n_{g2})V_{mol}}{n}}{\frac{n_{l3}\frac{M_{l3}}{\rho_{l3}}}{n}}$$ i.e. $$GOR = \frac{\frac{(n_{g1} + \bar{n}_{g2}.\bar{n}_{g1}.n)V_{mol}}{n}}{\frac{\bar{n}_{l3}\bar{n}_{l2}.\bar{n}_{l1}.n\frac{M_{l3}}{\rho_{l3}}}{n}}$$

Finally $$\boxed{GOR = \frac{(\bar{n}_{g1} + \bar{n}_{g2}.\bar{n}_{g1})V_{mol}}{\bar{n}_{l3}\bar{n}_{l2}.\bar{n}_{l1}\frac{M_{l3}}{\rho_{l3}}}}$$

Where,

V_{mol} is the molar volume of the gas molecules in the standard state. In $m^3/Kmoles$

M_{l3} is the molar mass of the liquid collecting in the third stage in $Kg/Kmoles$

ρ_{l3} is the density of the liquid collecting in the third stage in Kg/m^3

5.2 Gas-Liquid Equilibrium Calculations with equations of state.

The Peng-Robinson equation of state is:

$$p = \frac{RT}{V_m - b} - \frac{a_T}{V_m(V_m + b) + b(V_m - b)}$$

Where; V_m --Molar volume

a_T, b -> Individual Component coefficients

Mixing Rules are:

$$b = \sum_j Y_j b_j$$

and $$a_T = \sum_i \sum_j Y_i Y_j (a_{Ti} a_{Tj})^{1/2}(1 - \delta_{ij})$$ Where subscripts i and j refer to components

Also, $\delta_{ii} = \delta_{jj} = 0$ and $\delta_{ij} = \delta_{ij}$

The values of the coefficients for the individual components are calculated as

$$b_j = 0.07780\frac{RT_{cj}}{p_{cj}} \qquad \text{And} \qquad a_{Tj} = a_{cj}\alpha_j$$

Where;

$$a_{cj} = 0.45724\frac{R^2T_{cj}^2}{p_{cj}}$$

And

$$\alpha_j^{1/2} = 1 + (0.37464 + 1.54226s_j - 0.26992s_j^2)(1 - T_{rj}^{1/2})$$

The Peng-Robinson equation can be written as

$$z^3 - (1-B)z^2 + (A - 2B - 3B^2)z - (AB - B^2 - B^3) = 0$$

Where;

$$A = \frac{a_T p}{R^2T^2} \qquad \text{And} \qquad B = \frac{bp}{RT}$$

When the three roots of the equation are obtained, the lowest root is the z factor of the liquid. The highest root is the z-factor of the gas, and the middle root is discarded.

Combining the Peng-Robinson equation of state and equation for fugacity coefficient:

i.e. Fugacity coefficient may be calculated as

$$\ln\phi_j = \frac{1}{RT}\int_{\infty}^{V}[\frac{RT}{V} - (\frac{\partial p}{\partial n_j})_{T,V,n_i}]dV - \ln z$$

Therefore

$$\ln\phi_j = -\ln(z-B) + (z-1)B_j^{'} - \frac{A}{2^{1.5}B}(A_j^{'} - B_j^{'})\ln(\frac{z + (2^{1/2}+1)B}{z - (2^{1/2}-1)B})$$

Where;

$$B_j^{'} = \frac{b_j}{b} \qquad \text{And} \qquad A_j^{'} = \frac{1}{a_t}[2a_{Tj}^{1/2}\sum Y_i a_{Ti}^{1/2}(1-\delta_{ij})]$$

The procedure for calculating gas-liquid equilibria at a given temperature and pressure is as follows.

Values of a_{Tj} and b_j for each component of the mixture are obtained with the knowledge of the critical properties and acentric factors of the pure components.

A first trial set of K-factors is obtained. the Initial; values of K can be taken from this equation.

$$Kin_i = \frac{\exp(5.37(1+s_i)(1-\frac{1}{Tr_1})}{pr_i}$$

These are used in Gas-Liquid Equilibrium Calculation. For finding the values of liquid and gas composition, different values of fractional moles of gas and liquid are iterated. so that the following conditions are satisfied

$$\sum X_i = 1 \qquad \text{And} \qquad \sum Y_i = 1$$

The remaining equations are solved twice, once for liquid and once for Gas.

The values of a_T and b are found by the equations given above for the compositions found from initial Values of Equilibrium constant.. When the compositions of Liquid are used, the values are a_{TL} and b_L. When the composition of the gas is used, the values are a_{TG} and b_G. The values of Binary interaction coefficients, δ_{ij}, can be included in Equation if they are known. If unknown, the values of δ_{ij} can be set equal to zero. Values of A and B for liquid, A_L and B_L, are calculated using a_{TL} and b_L. Similarly A_G and B_G, are calculated from a_{TG} and b_G, for gas.

Also, $A_j^{'}$ and $B_j^{'}$ must be calculated for each component j. $B_{Lj}^{'}$ results when b_L is used and $B_{Gj}^{'}$ results when b_G is used.. Similarly $A_{Lj}^{'}$ and $A_{Gj}^{'}$ result from a_{TL} and a_{TG}.

The largest root of the z equation is z_g when A_G and B_G are used. Similarly for z_l.

The fugacity coefficient equation is solved for fugacity coefficients of the components of the Liquid, ϕ_{Lj}, using the values of A_L, B_L, $z_l, A_{Lj}^{'}$ and $B_{Lj}^{'}$. The values of ϕ_{Gj} result when corresponding gas coefficients and z-factor are used in that equation.

The Values of liquid Fugacity and Gas fugacity for each component are obtained from

$$fu_{Lj} = X_j p \phi_{Lj}$$

And

$$fu_{Gj} = Y_j p \phi_{gt}$$

Equilibrium is obtained and the calculation is complete when all

$$fu_{Gj} = fu_{Lj}$$

There are as many equations as there are components. All these equations cannot be satisfied simultaneously. Thus some sort of error function must be devised. One approach is

$$\varepsilon_j = fu_{Lj} - fu_{Gj}$$

Solution is obtained when $\varepsilon_j \leq 0.0001$

Another error function used in converging on a correct solution by a method of successive substitution involves K-factors. The K_j for the mixture are determined from fugacity coefficients with the following equation

$$K_j = \frac{\phi_{Lj}}{\phi_{Gj}} = \frac{\frac{fu_{Lj}}{X_j p}}{\frac{fu_{gj}}{Y_j p}} = \frac{Y_j}{X_j}$$

Then,

$$\varepsilon_j = \frac{(K_j^T - K_j^C)^2}{K_j^T K_j^C}$$

Where; K_j^C are the K-factors Just calculated

And K_j^T are the trial values of K-factors

Convergence on a correct solution is obtained when the sum of the error functions is less than some selected tolerance. If sum of the error functions is greater than the tolerance, the K_j^C are used as new trial values of K_j, and the process repeated.

An example of the calculation is shown in appendix

Chapter 6

Control Systems for Separators

6.1 Two-Phase Separation control

Control of pressure and level are basic to good separator operation. The choice of control modes, sensitivity characteristics, control hardware etc. depends on the purpose of the separator and the process modules immediately proceeding and following it.

As a general rule, the pressure of the separator must be held rather constant, independent of the operation of adjacent equipment. Usually this means a backpressure valve on the gas outlet. A deviation of 10% from the pressure set point can occur, although 6-7% is a good design number. If a high-pressure alarm is intermediate between normal operating pressure and high pressure shut down, there must be a broad enough pressure range allowed for the operator interaction to correct the problem.

The gas is often sent to a suction of boaster compressors. If this is reciprocating compressor or an axial compressor, a backpressure valve definitely is needed. No valve is absolutely required with a centrifugal compressor, since the suction pressure oftentimes can be kept rather constant with compressor controls.

If efficient vapor-liquid separation is the primary purpose of the vessel, the liquid level should be held relatively constant. The figure 6.1 shows a split range approach often used as long as the liquid level is steady valve A is operating. It is fully closed at pilot output pressure of 3 PSI and fully open at 9 PSI. When a large slug of liquid hits the separator, the level rises, the pilot output pressure rises towards 9 PSI, valve A is is fully open, but the liquid level continues to rise. At 9 PSI, valve B starts to open to relieve the surge. Once the slugging is over and the level is back within normal range, valve B closes and waits for next such upset.

This system provides sensitive routine level control plus the added capability for reliving surges not possible with a single valve system.

Suppose that the liquid is being pumped out the bottom of the separator. The figure 6.2 shows three different arrangements. In (A) the level control valve simply changes the pump backpressure to match pump output to level set point. A relief by pass circuit s provided to protect the pump from over pressure by the valve closing too much.

In (B) is provided for a rate controlled by pass valve. This is more expensive than (A) because the extra valves and controls needed. Both (A) and (B) are inefficient from a power utilization standpoint. Throttling and bypassing utilize power that serves no useful purposes. In some installations it is desirable to use a variable speed drive. The level controller than adjusts speed to maintain level.

As shown in (C), it is not possible to throttle a Positive displacement pump. A level will control by –pass rate or pump speed.

6.2 Three Phase Separation control

Figure 6.3 shows various ways of controlling the high-pressure separation of oil, gas and water. Shown are the use of interface controls, buckets and weirs. In interface control a displacement float reacts to the different density of the two adjacent liquid phases. Buckets are chambers within the vessel where one or more liquid phases are segregated. Weirs are used in segregation to eliminate the need for interface controls.

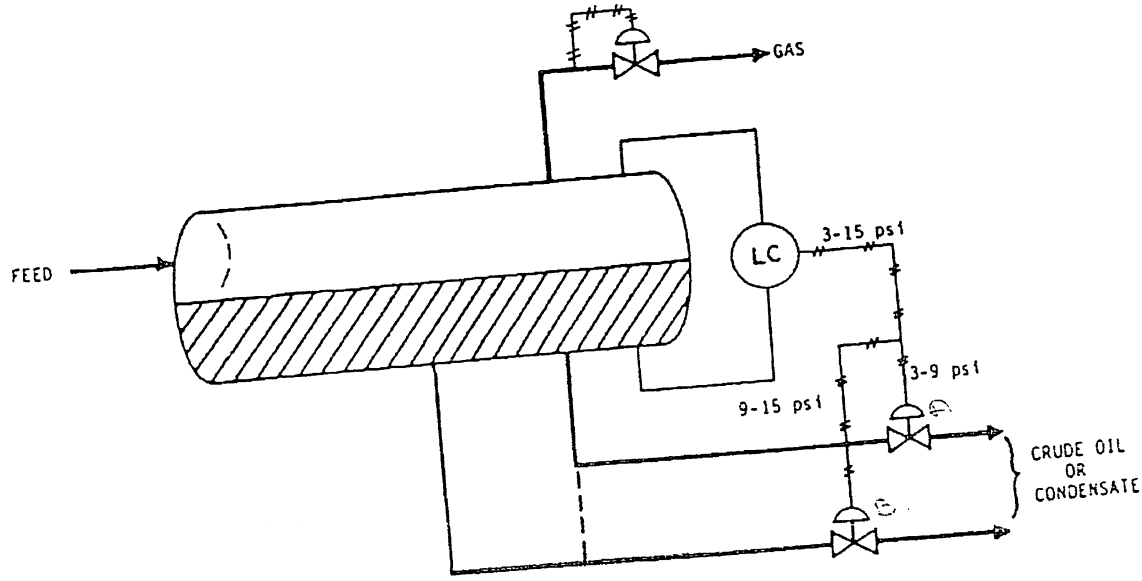


Fig.6.1
Example of a split-Range level control

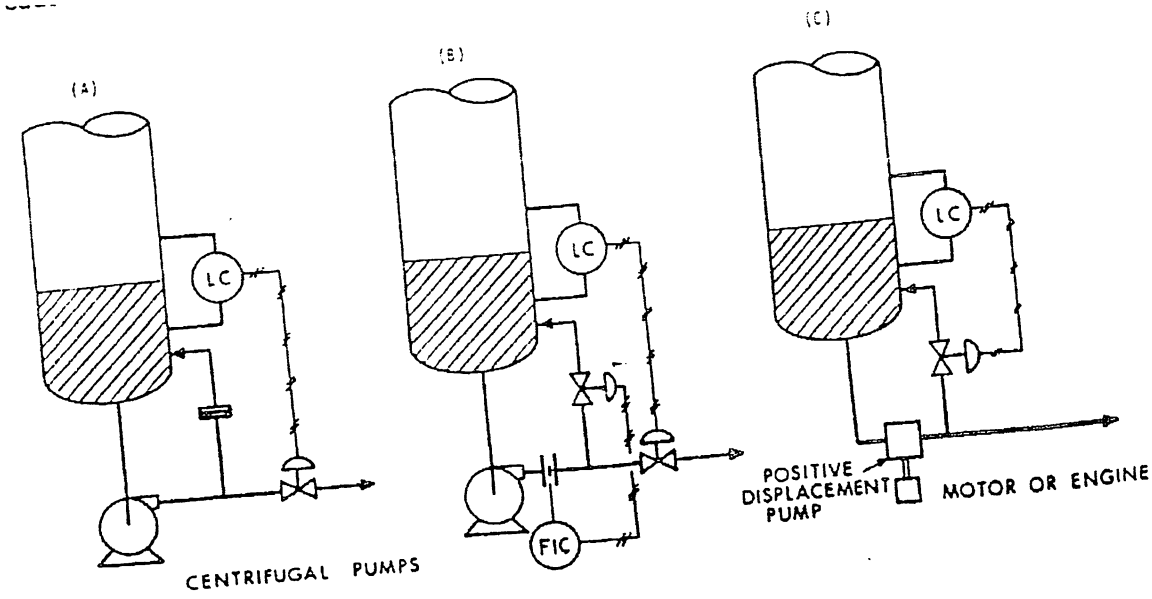
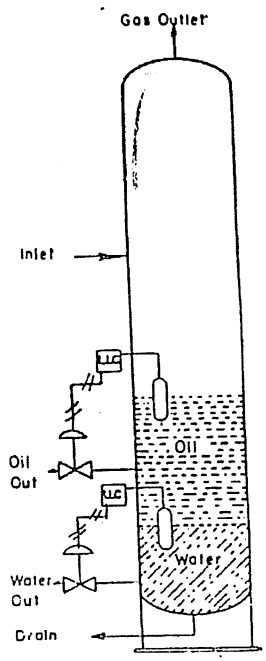
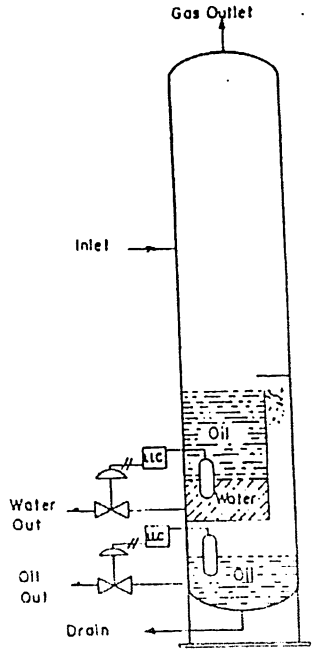


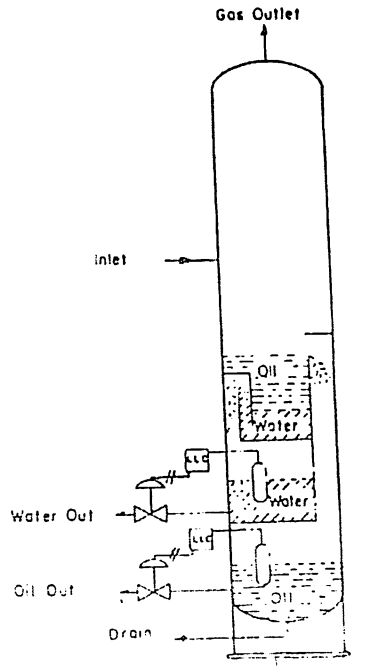
Fig.6.2
Example of Level control with pump.



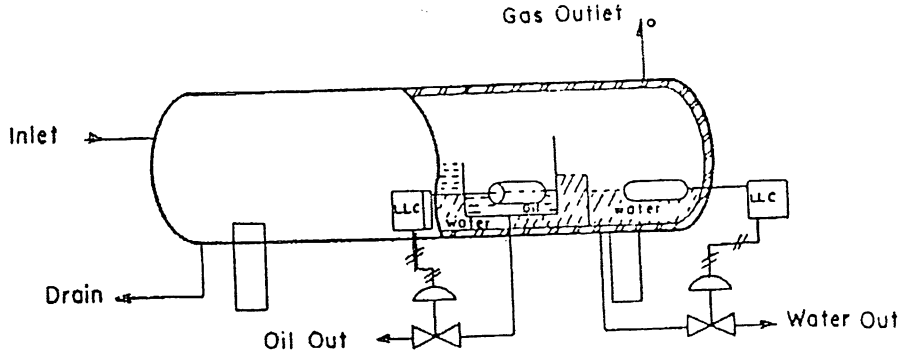
Method 1 Separation obtained through level control only.



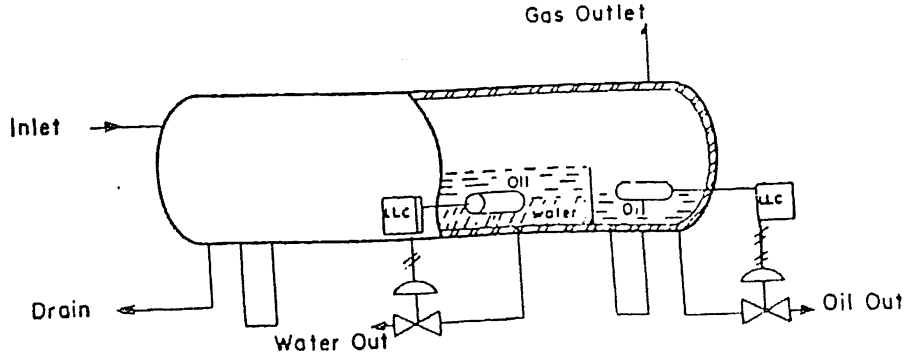
Method 2 Separation obtained by interfacial level control and weir.



Method 3 Weir used to separate oil and water to avoid use of interfacial controls.



Method 4 Operates in same manner as vessel in Method 3.



Method 5 Operates in same manner as vessel in Method 2.

Fig.6.3
Various methods for controlling Three-phase Separation

Chapter 7

Piping Systems for three-phase Separation.

7.1 Oil and gas Gathering and Separation systems

The well streams of wells producing crude oil and natural gas are conveyed to centers located on the oilfield. A system comprising piping, pipe fittings and central facilities permit us to separate the liquid from the gas, to measure the quantity of both, to adjust their properties so as to fall within sales contract and/or other specifications, and to transport them to the consumers or refineries.

Gathering and separating systems fall into three great groups. All three types of system start at the well and end at the storage tanks of the pipeline or in the intake of the pipeline driver pump.

The first group includes production systems of extremely high-capacity wells. Each well has its own facilities for separation and metering, possibly also for treatment. This setup is seldom economical

A more frequent type of system involves gathering and separating facilities permitting the common handling of several well streams.

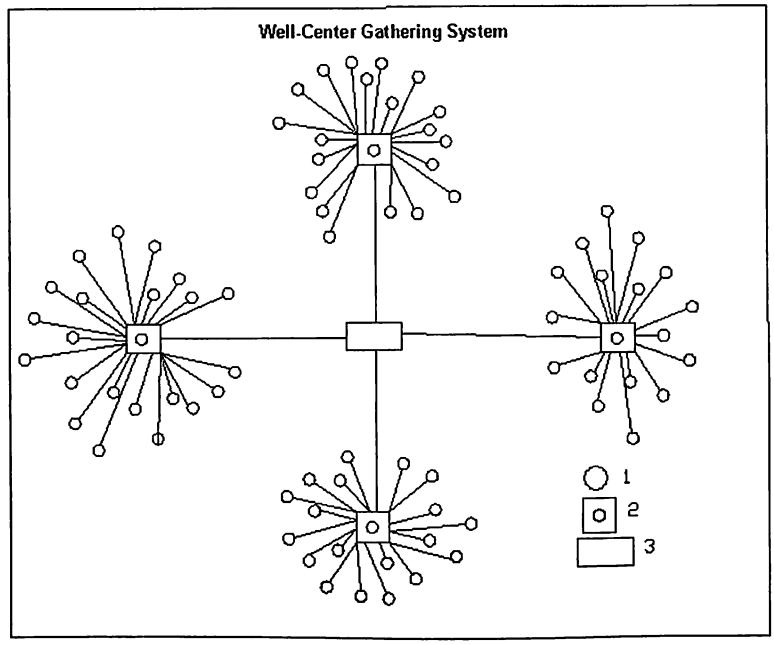


Fig.7.1

Fig.7.1 shows a so-called well-center gathering system.

Individual wells 1 are connected to well centers 2. Each well stream is transported to the well center in an individual flow line. At the well centers, the well streams at least of the wells selected for individual testing and metering are kept separate; their oil, gas and water rates are metered; then either the united well streams are transported as they are to the central gathering station 3 or the gas separated at each well center is introduced into the gas gathering line, whereas the liquid is forwarded to the central gathering station.

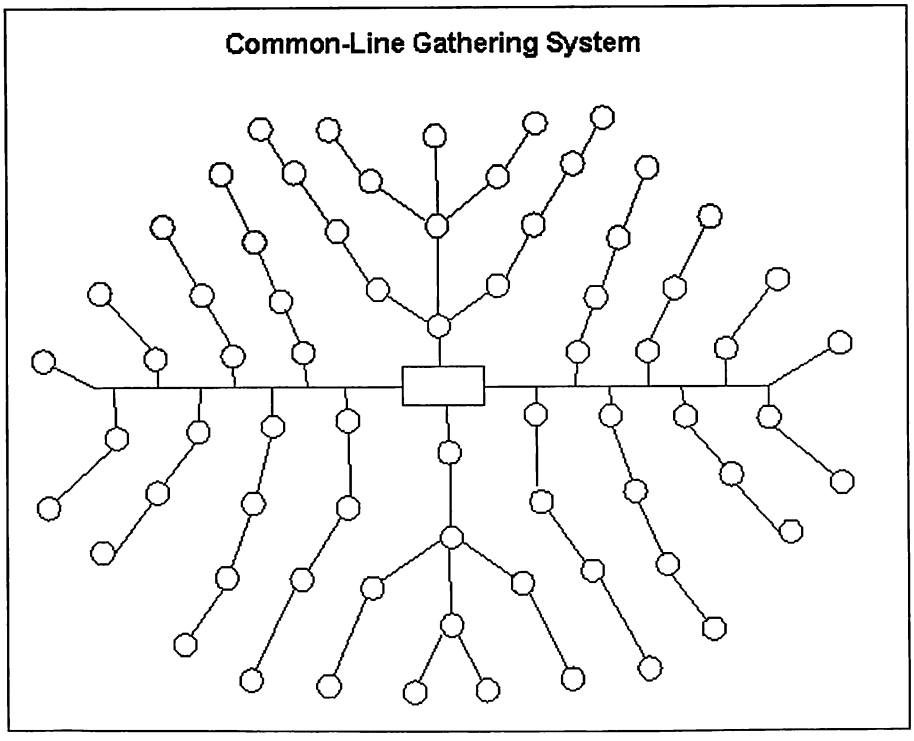


Fig 7.2

In third group several wells produce into a common flow line.(fig 7.2). Oil, gas and water production of individual wells are metered at intervals by means of portable well testers installed at the well sites.

All other treatment takes place at the central station.
Of the above named groups, the well-center system is the most widespread.

Viewpoints for designing gathering systems with well testing centers.

The principles to be observed in designing are as follows:

1. The wellhead pressures of the wells should be as low as feasible. The resulting main advantages are: longer flowing life; lower specific injection-gas consumption in gas lift wells; higher yield of wells produced with bottom hole pumps in the last phase of production.
2. The hydrocarbon loss of the system should be minimum.
3. The system should be easy to oversee as far as control and checking are concerned. This makes for disciplined production and permits fast intervention in times of operating troubles.
4. Metering the individual and common oil, gas and water production of the wells and testing for impurities in the liquid produced should be ensured to the necessary accuracy.
5. When determining oil storage volume, the off take rate to be expected is to taken into consideration, together with interruptions in off take due to breakdowns to be expected, as well as the settling time possibly required for the removal of water and impurities.
6. Expansion of facilities made necessary by bringing in of further well should require the least possible modifications to the existing installations; said modifications should be possible to achieve without disturbing the wells already in production.

7. Specific cost, referred to the volume unit of oil and /or gas produced, of installing and operating the system should be as low as possible
8. Safety prescriptions should be meticulously observed.

The realization of these principles may arise the following, partly contradictory, viewpoints

A. 1. Head loss in the flow line between wellhead and the separator should be as low as possible. Hence, sudden breaks in the flow line trace and sudden changes in cross-section should be avoided; flow line trace size, length and trace should be chosen with a view to attaining minimum head loss; if the crude is waxy and/or sandy, measures are to be taken to prevent the formation of the of bothersome deposits in the flow line and its fittings; in case of high viscosity or waxy crudes, reducing head loss may be achieved by heating the crude at the well head; heat-insulating the flow line, or injecting a friction reducing chemical; if water is readily separated from the well stream , it is recommended to install a water knockout next to each well.

2. Separator pressure should be as low as possible. In order to keep it so, it is usually recommended to install separator higher then storage tank level at the well centers so as to make oil flow by gravity from the separators into the tanks. The less the pressure needed to convey gas from the separator into the tanks. The less the pressure needed to convey gas from the separator through the gathering line to the compressor station, the better. This can be achieved primarily by the gathering –line network of low flow resistance (big size, small aggregate length, efficient liquid knock out or scrubbing) and by using low intake Pressure compressors

B. 1. Well streams from flowing wells of high wellhead pressure should be directed into a high-pressure separator; stage separation is recommended.
2. The tank system should be closed if possible.
3. Evaporation losses of open storage tanks
4. Oil and gas leaks should be kept minimum

C. 1. Particularly in manually operated and those with local automation, care should be taken to concentrate all the gathering and separation facilities at the well centers and the central gathering station. In remote controlled systems designs ensuring the fast supply of meaningful information should be starved at. Information should be supplied both to men working on the lease and to remote-control center.

D. 1. The number of separators enabling individuals wells to be tested at the well center should be sufficient to permit measuring the oil, gas and water production of each well at the intervals of 4 to 7 days
2. The number of common separator handling the well streams that are not separately metered should be composed of uniform units for each stage. The number of separators required is then determined by unit capacity
3. In on-lease gas metering, the accuracy of +/- 1-2 % of the orificemeter is adequate. The quantity of gas fed into a sales line or a transmission pipeline should be determined more accurately, if possible

4. If the liquid output is measured in the storage tanks at the well centers, then the number of so called test tanks metering the liquid production of individual wells should agree with the number of test separators. Common tanks metering the production of remaining wells should number at least two per center
5. For accounting within the lease it is usually sufficient to meter liquid at an accuracy of about 0.5 % . It is desirable to meter oil delivered outside the lease at an accuracy of at least 0.2%

E. 1. Optimum Storage-tank volume may differ widely depending on the nature of the gathering, separation, and transmission facilities. It is recommended in the usual case to have storage capacity for 2-3 days production

F. 1. Well centers are to be designed so that binging well in or off, and changing separator, tank or metering capacities might be achieved without effecting the equipment in operation. All the equipment and fittings of well centers within a lease should be uniformized; fitting should be transportable to the wells center ready for installation and it should be possible to install them without welding.

G. 1. The mechanical production equipment of wells should be chosen, and phased during the productive life of the wells, so as to minimize specific production cost over the entire life of the lease.
2. Gas from wells having a high well head pressure should be led at the lowest possible pressure loss into the compressor station. One and possible two compressor stages may thus be saved temporarily.
3 The number and location of well centers, as well as the location of the central gathering station, should be chosen so as to minimize total cost.
4 Temporary piping should be joined by couplings.
5. Well testing centers, usually hand operated early in life of the lease, should be devised so as to require the least possible modification when converting to automation. It is usually preferable to install an LACT (Lease Automatic Custody Transfer) system rather than a central gathering station early in lease life.

7.2 Piping system for three phase separation

The various piping and accessories used in piping for the three-phase separation are:

1. Well head accessory items:
   a. Sampling and injection connections:
      Connections may be desired near the wellhead for chemical injection and for obtaining samples.
   b. Chokes:
      Chokes are normally installed to control the flow from oil and gas wells. Chokes types include adjustable, positive and combination. The number and location of chokes depend on the amount of pressure drop taken, well fluid, flow rate and solids in the well streams. Usually if only one choke is used, it should be located near the wellhead. Additional chokes may be located near the manifold, entering low temperature separation units, in conjunction with line heaters etc.

2. Flow line and flow line accessories
   Flowline: Piping which carries well fluid from the wellhead to manifold or first process vessel.
   For designing flowlines consideration should be given to pressure, temperature, velocity, erosive effects on the pipe, etc. The various accessories over a flow line are:
   a. A flowlines pressure sensor is connected to sense the pressure in the flowline. These should be located to minimize the possibility of plugging and freezing. Also it is installed with an external and block valve.

   b. Flowline orifice fitting: A flowline orifice fitting may be desirable in gas well service for either a well monitoring aid or as a means of production allocation.

   c. Flowline heat exchanger: A flowline heat exchanger is used whenever the fluid needs to be heated to increase its flowability. Heavier crudes tends to be more viscous with the decreasing temperature A heat exchanger is added to provide heat to the liquid so as to remain it in liquid state.

   d. Flowline check valve: A flow line check valve should be installed to minimize back flow due to inadvertent switching of a low pressure well into a higher pressure system or in case of line rupture. Provisions should be made for blowdown of the flow line segment between wellhead and check valve to facilitate periodic testing at check valve.

   e. Flowline support: flowlines should be supported and secured to minimize vibration and to prevent whip. While designing flowline supports, it should be recognized that even though the well head may be fixed to a platform, there s a possibility of independent well head movement due to wave action, wind forces etc on the conductor.

3. Production manifolds:
   Manifold: An assembly of pipe, valves and fittings by which fluid from one to more sources is selectively directed to various process systems.
   The figure shown is a six header manifold. The actual number and function of headers depend upon the specific application.
   Manifold branch connections: These are made on the manifold for branching. The terminus of the manifold runs should be blind flanged to provide a fluid cushion area for possible future expansion.
   Manifold valve installations: There is a valve connection for a manifold; it should be arranged in such a way to provide easy access to each valve. For operational purposes and easy removal.

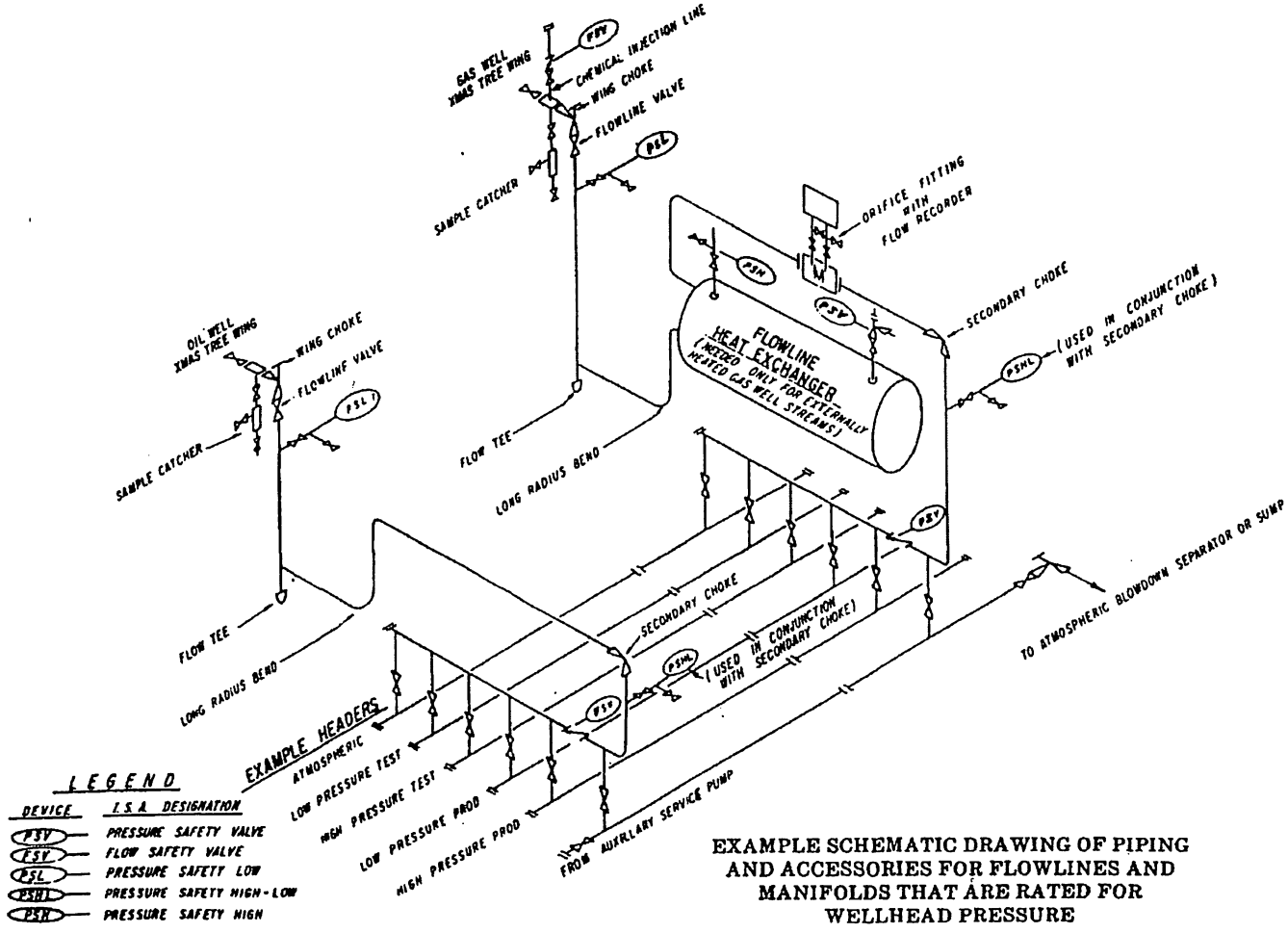
4. Process vessel piping:
   A typical 3-phase process vessel with standard accessories and many optional items is shown in figure. Different vessels are required for different functions in processing; however all of the flow streams to and from a vessel are generally handled in a similar manner

5. Utility systems:
    a. Pneumatic systems: Pneumatic systems are required to provide a dependable supply for pneumatically operated components.
    b. Firewater systems: Fire water systems are required to provide water for extinguishing fire.
    c. Potable water systems: required for drinking water.
    d. Sewage systems: A sewage systems are provided for the living quarters areas.

6. Heating fluid and glycol systems

7. Pressure relief and disposed systems.
    The commonly used safety relief devices are the conventional spring loaded relief valve, the balanced bellows spring loaded relief valve, the pilot operated relief valve, the pressure vacuum relief valve and the rupture disk
8. Drain systems.
    Drain systems should be designed to collect and dispose of contaminants from all sources. A good drain system prevents contaminants from spilling overboard; prevents the accumulation of flammable liquids on the deck or pans and promotes good house keeping practices.

    The figure 7.3 can referred for the above explanations



EXAMPLE SCHEMATIC DRAWING OF PIPING AND ACCESSORIES FOR FLOWLINES AND MANIFOLDS THAT ARE RATED FOR WELLHEAD PRESSURE

Fig7.3

Chapter 8

Detail Design

8.1 Vapor-Liquid Equilibrium calculations for Primary Separation

Given Conditions: Temperature, T = 69.9 °c = 617.49 °R
Pressure, P = 108 barg = 109.0135bara = 1581 Psia

Molar Composition (%)	Z	Equilibrium Constant at given Conditions, K	$\bar{n}_g$	X		Y
Methane	77.9952	2.534574	0.9784	0.308765		0.782606
Ethane	4.5502	1.052380	$\bar{n}_l$	0.043249		0.045515
Propane	1.7002	0.580597	0.0216	0.029165		0.016933
Iso-butane	0	0.321635		0		0
N-butane	0.9277	0.392457		0.023435		0.009197
Iso-Pentane	0	0.184549		0		0
N-Pentane	0.5077	0.212169		0.023442		0.004974
N-Hexane	0.2852	0.110019		0.024799		0.002728
Heptanes +	2.2702	0.064889		0.323730		0.021006
H2S	1.3652	0.709137		0.019207		0.013621
H2O	2.9877	0.172190		0.0168967		0.029094
Nitrogen	4.3152	4.543544		0.009539		0.043341
CO2	3.0952	1.175472		0.168967		0.030978
			$\sum X =$	1	$\sum Y =$	1

Where, X -> Molar composition of components in Liquid State.
Y - > Molar composition of components in Gaseous State.
K -> Equilibrium Constant (These are found by the method described in Appendix).

$\bar{n}_g$, $\bar{n}_l$ -> Fractional Moles of Gas and Liquid.

8.2 Determination of GOR at Standard Conditions:

We have;

$$GOR = \frac{n_g * V_{molG}}{n_l * \frac{M_l}{\rho_l}} = \frac{2138 \bar{n}_g * \rho_l}{\bar{n}_l * M_l} \frac{scf}{SCB}$$

Where, $V_{molG} = \frac{RT}{P}$ $\quad R = 10.73$

M_l = Molecular weight of Liquid (Assume M_l for Heptanes+ = 218)

ρ_l = Density of the Liquid

Components	Mole Fraction Liquid Phase, X	Molecular weights, M	$X.M$	Liquid Density of Propane + at 60of & 14.7 Psia ρ	X.M / ρ
Methane	0.308765	16.043	4.95352		
Ethane	0.043249	30.07	1.3005		
Propane	0.029165	44.097	1.28609	31.62	0.0406733
Iso-butane	0	58.124	0	35.12	0
N-butane	0.023435	58.124	1.36214	36.42	0.0374008
Iso-Pentane	0	72.151	0	38.96	0
N-Pentane	0.023442	72.151	1.69136	39.36	0.0429717
N-Hexane	0.024799	86.178	2.13713	41.4	0.0516215
Heptanes +	0.323730	218	70.5731	53.11	1.3288108
H2S	0.019207	34.076	0.6545		
H2O	0.0168967	18.015	0.30439		
Nitrogen	0.009539	28.013	0.26722		
CO2	0.168967	44.01	7.43624		
	1	$M_l = \sum X$ =	91.9662	$\sum \frac{X.M}{\rho} =$	1.50148

Density of the Liquid (Propane +) $= \frac{91.9662}{1.50148} = 61.25037.\frac{lb}{ft^3}$

$$GOR = \frac{2138 * 0.9784 * 61.25037}{0.0216 * 91.9662} = 64498.6823.\frac{scf}{STB}$$

$$\boxed{GOR = 64498.6823.\frac{scf}{STB}}$$

8.3 Determination of the gravity of the Liquid: γ_l

$$\gamma_l = \frac{\rho_l}{\rho_{water}} = \frac{61.25037}{62.37} = 0.98204$$

$$\gamma_l = \frac{141.5}{0.98204} - 131.5 = 12.5865^0 API$$

$$\boxed{\gamma_l = 12.5865^0 API}$$

8.4 Determination of the specific gravity of the Gas: γ_g

Components	Mole fraction In Gaseous phase, y	Molecular weights, M	$Y.M$
Methane	0.782606	16.043	12.5553
Ethane	0.045515	30.07	1.36864
Propane	0.016933	44.097	0.74669
Iso-butane	0	58.124	0
N-butane	0.009197	58.124	0.53457
Iso-Pentane	0	72.151	0
N-Pentane	0.004974	72.151	0.35888
N-Hexane	0.002728	86.178	0.23509
Heptanes +	0.021006	218	4.57931
H2S	0.013621	34.076	0.46415
H2O	0.029094	18.015	0.52413
Nitrogen	0.043341	28.013	1.21411
CO2	0.030978	44.01	1.36334
		$M_l = \sum Y.M =$	23.9443

We have; $$\gamma_g = \frac{\sum Y.M}{29} = \frac{23.9443}{29} = 0.8257$$

$$\boxed{\gamma_g = 0.8257}$$

8.5 Sizing of the Separator

The test separator is a vertical three-phase settler with inlet device as schoepentoeter, Mistmat, Vanepack, and Vortex Breaker for the oil and water outlets. The data provided in the datasheets has given for two conditions i.e. Maximum and minimum. The properties of vapor, condensate and water, vary with these conditions. Therefore the sizing is made for both the conditions.

i. For Working Conditions: Temperature, T = 22°c (Min)
Pressure, P = 110 barg (Min)

Condensate	Water	Vapor
$m_l = 10897 \frac{Kg}{hr}$	$m_h = 4439 \frac{Kg}{hr}$	$m_g = 164744 \frac{Kg}{hr}$
$\mu_l = 0.329x10^{-3} Pa-s$	$\mu_h = 0.261x10^{-3} Pa-s$	$M = 20.91$
$\rho_l = 628.5 \frac{Kg}{m^3}$	$\rho_h = 949.5 \frac{Kg}{m^3}$	$\rho_g = 82.91 \frac{Kg}{m^3}$
$Q_l = 4.8161x10^{-3} \frac{m^3}{s}$	$Q_h = 1.2986x10^{-3} \frac{m^3}{s}$	$Q_g = 0.55195 \frac{m^3}{s}$

We have;

$$Q_{G\max} = Q_g\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 0.21516\frac{m^3}{s} \qquad \phi_{wm} = \frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}} = 0.02402$$

$\lambda_{\max} = 0.105.f_\mu.f_\phi$ -> Because Mistmat is present

Where $f_\mu = \left(\frac{0.001}{\mu_l}\right)^{0.04}$ If > 0.001 Pa-s

$= 1$ If < 0.001 Pa-s

& $f_\phi = \left(\frac{1}{1+10.\phi_{wm}}\right)$ Where $\phi_{wm} = 0.05.\frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$ for Schoepentoeter

$\phi_{wm} = 1.2012x10^{-3}$

$\boxed{f_\mu = 1}$ And $\boxed{f_\phi = 0.9881}$

And

$\boxed{\lambda_{\max} = 0.10375m/s}$

We have; $Ag_{\min} = \frac{Q_{G\max}}{\lambda_{\max}} = 2.07375 = \frac{\pi D^2}{4}$ i.e.

$\boxed{D = 1.650m}$

Inlet nozzle sizing:

Since Shoepentoeter is used as a inlet device

We have;

$$\rho_m v_m^2 \leq 6000$$

Where ρ_m → Mean Density of the Mixture of the feed pipe given by

v_m → Mean velocity of the mixture in inlet Nozzle

We have $\rho_m = \frac{\rho_l Q_l + \rho_g Q_g}{Q_l + Q_g} = 88.9255$

and $v_m = \frac{Q_l + Q_g}{\pi/4 * d_1^2} = \frac{0.5580561}{\pi d_1{}^2/4} = 8.21414$ Where d_1 → Diameter of the Inlet Pipe.

i.e. $d_1 = 0.2941m \approx 12"$ $\boxed{d_1 = 12"}$

Gas Outlet nozzle sizing:

We have; $\rho\, v_m^2 \leq 3750$

i.e. $v_m = 6.4938m/s$ i.e.

$\boxed{d_2 = 14"}$

Condensate Outlet nozzle sizing:

We have; $v_m = 1m/s$ i.e. $v_m = \frac{Q_l}{\pi d_3^2/4}$

$$\boxed{d_3 = 4"}$$

Water Outlet nozzle sizing:

We have; $v_m = 1m/s$ i.e. $v_m = \frac{Q_h}{\pi d_4^2/4}$

$$\boxed{d_4 = 2"}$$

Length of the Vessel

Assume:

$d_{pp} = 20mm$

$\theta = 45^0$

$t_{pp} = 1mm = 0.001m$

$f_{loss} = 0.95$

$R_e = 850$

$d_p = 150\mu m$

$f_{tl} = 0.5$

$L_{entry} = 16 * d_{pp}$

$$Af_{Gross} = \frac{(Af_{net})_l (t_{pp} + d_{pp})}{d_{pp} F_{loss}}$$

For light Phase		For Heavy Phase	
$(Af_{net})_l = \frac{2d_{pp}Q_l\rho_l}{\mu_l R_e} =$	0.4329	$(Af_{net})_h = \frac{2d_{pp}Q_h\rho_h}{\mu_h R_e} =$	0.2231
$(Af_{Gross})_l = \frac{(Af_{net})_l (t_{pp} + d_{pp})}{d_{pp} F_{loss}} =$	0.4785	$(Af_{Gross})_h = \frac{(Af_{net})_h (t_{pp} + d_{pp})}{d_{pp} F_{loss}} =$	0.2457
$(V_{pp})_l = \frac{Q_l}{(Af_{net})_l} =$	0.011125	$(V_{pp})_h = \frac{Q_h}{(Af_{net})_h} =$	5.841e-3
$(Vp_{set})_l = \frac{\lvert\rho_h - \rho_l\rvert g d_p^2}{18\mu_l} =$	0.01196	$(Vp_{set})_h = \frac{\lvert\rho_h - \rho_l\rvert g d_p^2}{18\mu_h} =$	0.01508
$(L_{pp})_l = \frac{(V_{pp})_l d_{pp}}{(Vp_{set})_l \cos\theta} + ft_l L_{entry} =$	0.18630	$(L_{pp})_h = \frac{(V_{pp})_h d_{pp}}{(Vp_{set})_h \cos\theta} + ft_h L_{entry} =$	0.170955

The Value of $L_{pp} = 0.18630m$ (Maximum of the above value)

But $L_{pp} \geq 0.3m$ Therefore $\boxed{L_{pp} = 0.3m}$

We have the criterion for W_{pp} as $W_{pp} \leq \sqrt{(D-0.2)^2 - (L_{pp} + \Delta L_{f.p\_pp})^2}$

$$\boxed{W_{pp} = 1.37822m}$$

Also; $$H_l = \frac{(Af_{gross})_l}{w_{pp}} = 0.31517 > 0.3$$

$$H_h = \frac{(Af_{gross})_h}{w_{pp}} = 0.1618 < 0.3$$

Hence Reduce W_{pp}

Put $W_{pp} = 0.7$

Therefore $$H_l = \frac{(Af_{gross})_l}{w_{pp}} = 0.6835 > 0.3$$

$$H_h = \frac{(Af_{gross})_h}{w_{pp}} = 0.3515 > 0.3$$

Check;

$$0.5 \le \frac{H_l}{W_{pp}} \le 2 \text{ and } 0.5 \le \frac{H_h}{W_{pp}} \le 2$$

$\frac{H_l}{W_{pp}} = 1.3292 > 0.5$ And $\frac{H_h}{W_{pp}} = 0.6825 > 0.5$

Therefore the conditions are satisfied

i.e. The selected width of the plate pack $\boxed{W_{pp} = 0.7m}$

$\gamma_l = 12.5865^0 API$

& $\boxed{H_l = 0.6835m}$ and $\boxed{H_h = 0.3515m}$ for $\boxed{L_{pp} = 0.3m}$

Now we have; $H_{contot} = \sum H_{con} + H_{db}$

Assume $H_{db} = 0.2m$ and liquid retention time for control $t_{con} = 60s$

$$H_{contot} = \frac{(Q_l t_{con})_l}{\pi/4D^2} + \frac{(Q_h t_{con})_h}{\pi/4D^2} + H_{db} = 0.37158$$

$$H_{pp} = H_l + H_h + H_{contot} = 0.6835 + 0.37158 + 0.3515 = 1.41m$$

$$\boxed{H_{pp} = 1.41m}$$

Calculation of sum of control bands required for GL Control level:

$Q_l = 4.8161x10^{-3} \frac{m^3}{s}$

$Ag_{\min} = 2.07375m^2$

Let A be the length for the Levels LZA (LL) - LA (L)

The liquid Hold up time for this level $t_s = 60s$

i.e $$A = \frac{Q_l * t_s}{Ag_{\min}} = 0.1394m$$

Let B be the length for the levels LA (L) – LA (H)
The liquid Hold up time for this level t_s = 180s

i.e $$B = \frac{Q_l * t_s}{Ag_{\min}} = 0.418034m$$

Let C be the length for the levels LA (L) – LA (H)
The liquid Hold up time for this level t_s = 60s

i.e $$C = \frac{Q_l * t_s}{Ag_{\min}} + 0.25 = 0.3851m$$ (Foaming allowance of 0.25m)

Therefore

$$X_1 = A + B + C = 0.9467m$$

$$\boxed{X_1 = 0.9467m}$$

X_2 = Distance between the LZA (HH) and the bottom of the Schoepentoeter.

$$X_2 = 0.05D = 0.0825m$$

$$\boxed{X_2 = 0.0825m}$$

X_3 = Height of the Schoepentoeter.

$$X_3 = d_1 + 0.02m = 0.3248$$

$$\boxed{X_3 = 0.3248m}$$

X_4 = Distance between top of the Schoepentoeter to bottom of demisting device

$$X_4 = d_1 = 0.3048m$$

$$\boxed{X_4 = 0.3048m}$$

X_5 = Height of the Demisting device.

$$\boxed{X_5 = 0.1m}$$

X_6 = Distance between top of demisting device to Top tangent Length (TTL)

$$X_6 = 0.15d = 0.2475m$$

$$\boxed{X_6 = 0.2475m}$$

Total Length of the Separator,

$$\boxed{L = H_{pp} + X_1 + X_2 + X_3 + X_4 + X_6 + X_7}$$

$$\boxed{L = 3.6062m}$$

The tangent-to-tangent length of the separator, $L = 3.6062m$

8.6 For Working Conditions:

Temperature, T = 104°c (Min)
Pressure, P = 125 barg (Min)

Condensate	Water	Vapor
$m_l = 15260\frac{Kg}{hr}$	$m_l = 3926\frac{Kg}{hr}$	$m_l = 177568\frac{Kg}{hr}$
$\mu_l = 0.319x10^{-3}\,Pa-s$	$\mu_l = 0.366x10^{-3}\,Pa-s$	$M = 20.7$
$\rho_l = 625.7\frac{Kg}{m^3}$	$\rho_h = 969.1\frac{Kg}{m^3}$	$\rho_l = 103.6\frac{Kg}{m^3}$
$Q_l = 6.7746x10^{-3}\frac{m^3}{s}$	$Q_h = 1.1253x10^{-3}\frac{m^3}{s}$	$Q_g = 0.4761\frac{m^3}{s}$

We have;

$$Q_{G\max} = Q_g\sqrt{\frac{\rho_l - \rho_g}{\rho_g}} = 0.21208\frac{m^3}{s} \qquad \phi_{wm} = \frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}} = 0.035$$

$\lambda_{\max} = 0.105.f_\mu.f_\phi$ -> Because Mistmat is present

Where $f_\mu = \left(\frac{0.001}{\mu_l}\right)^{0.04}$ If > 0.001 Pa-s

$= 1$ If < 0.001 Pa-s

& $f_\phi = \left(\frac{1}{1+10.\phi_{wm}}\right)$ Where $\phi_{wm} = 0.05.\frac{Q_l}{Q_g}\sqrt{\frac{\rho_l}{\rho_g}}$ for Schoepentoeter

$\phi_{wm} = 1.748x10^{-3}$

$\boxed{f_\mu = 1}$ And $\boxed{f_\phi = 0.9828}$

And

$\boxed{\lambda_{\max} = 0.1032}$

We have; $Ag_{\min} = \frac{Q_{G\max}}{\lambda_{\max}} = 2.05512 = \frac{\pi D^2}{4}$ i.e.

$\boxed{D = 1.650m}$

Inlet nozzle sizing:

Since Shoepentoeter is used as a inlet device
We have;

$$\rho_m v_m^2 \le 6000$$

Where ρ_m → Mean Density of the Mixture of the feed pipe given by

v_m → Mean velocity of the mixture in inlet Nozzle

We have $\rho_m = \frac{\rho_l Q_l + \rho_g Q_g}{Q_l + Q_g} = 112.9434$

and $v_m = \frac{Q_l + Q_g}{\pi/4 * d_1^2} = \frac{0.5580561}{\pi d_1{}^2/4} = 7.2886 m/s$ Where d_1 → Diameter of the Inlet Pipe.

i.e. $d_1 = 0.2901m \approx 12"$ $\boxed{d_1 = 12"}$

Gas Outlet nozzle sizing:

We have; $\rho\, v_m^2 \leq 3750$

i.e. $v_m = 5.7621 m/s$ i.e.

$\boxed{d_2 = 14"}$

Condensate Outlet nozzle sizing:

We have; $v_m = 1 m/s$ i.e. $v_m = \frac{Q_l}{\pi d_3^2/4}$

$\boxed{d_3 = 4"}$

Water Outlet nozzle sizing:

We have; $v_m = 1 m/s$ i.e. $v_m = \frac{Q_h}{\pi d_4^2/4}$

$\boxed{d_4 = 2"}$

Length of the Vessel

Assume:

$d_{pp} = 20mm$

$\theta = 45^0$

$t_{pp} = 1mm = 0.001m$

$f_{loss} = 0.95$

$R_e = 850$

$d_p = 150\mu m$

$f_{tl} = 0.5$

$L_{entry} = 16 * d_{pp}$

$$Af_{Gross} = \frac{(Af_{net})_l (t_{pp} + d_{pp})}{d_{pp} F_{loss}}$$

For light Phase		For Heavy Phase	
$(Af_{net})_l = \frac{2d_{pp}Q_l\rho_l}{\mu_l R_e} =$	0.4329	$(Af_{net})_h = \frac{2d_{pp}Q_h\rho_h}{\mu_h R_e} =$	0.14022
$(Af_{Gross})_l = \frac{(Af_{net})_l(t_{pp}+d_{pp})}{d_{pp}F_{loss}} =$	0.4784	$(Af_{Gross})_h = \frac{(Af_{net})_h(t_{pp}+d_{pp})}{d_{pp}F_{loss}} =$	0.155
$(V_{pp})_l = \frac{Q_l}{(Af_{net})_l} =$	0.01565	$(V_{pp})_h = \frac{Q_h}{(Af_{net})_h} =$	8.02548e-3
$(Vp_{set})_l = \frac{\lvert\rho_h - \rho_l\rvert g d_p^2}{18\mu_l} =$	0.0132	$(Vp_{set})_h = \frac{\lvert\rho_h - \rho_l\rvert g d_p^2}{18\mu_h} =$	0.01150
$(L_{pp})_l = \frac{(V_{pp})_l d_{pp}}{(Vp_{set})_l \cos\theta} + ft_l L_{entry} =$	0.1935	$(L_{pp})_h = \frac{(V_{pp})_h d_{pp}}{(Vp_{set})_h \cos\theta} + ft_h L_{entry} =$	0.17973

The Value of $L_{pp} = 0.1935m$ (Maximum of the above value)

But $L_{pp} \geq 0.3m$ Therefore $\boxed{L_{pp} = 0.3m}$

We have the criterion for W_{pp} as $W_{pp} \leq \sqrt{(D-0.2)^2 - (L_{pp} + \Delta L_{f,p\_pp})^2}$

$$W_{pp} = 1.3784m$$

Choose $\boxed{W_{pp} = 1.2m}$

Also; $$H_l = \frac{(Af_{gross})_l}{w_{pp}} = 0.3986 > 0.3$$

$$H_h = \frac{(Af_{gross})_h}{w_{pp}} = 0.13 < 0.3 \text{ (Criterion not satisfied)}$$

Hence Reduce W_{pp}

Put $W_{pp} = 0.5$

Therefore $$H_l = \frac{(Af_{gross})_l}{w_{pp}} = 0.9568 > 0.3$$

$$H_h = \frac{(Af_{gross})_h}{w_{pp}} = 0.31 > 0.3$$

Check;

$$0.5 \leq \frac{H_l}{W_{pp}} \leq 2 \text{ and } \quad 0.5 \leq \frac{H_h}{W_{pp}} \leq 2$$

$$\frac{H_l}{W_{pp}} = 1.9136 > 0.5 \quad \text{and} \quad \frac{H_h}{W_{pp}} = 0.62 > 0.5$$

Therefore the conditions are satisfied

i.e. The selected width of the plate pack $\boxed{W_{pp} = 0.5m}$

& $\boxed{H_l = 0.9568m}$ and $\boxed{H_h = 0.31m}$ for $\boxed{L_{pp} = 0.3m}$

Now we have; $H_{contot} = \sum H_{con} + H_{db}$

Assume $H_{db} = 0.2m$ and liquid retention time for control $t_{con} = 60s$

$$H_{contot} = \frac{(Q_l t_{con})_l}{\pi/4D^2} + \frac{(Q_h t_{con})_h}{\pi/4D^2} + H_{db} = 0.4216$$

$$H_{pp} = H_l + H_h + H_{contot} = 0.9568 + 0.31 + 0.4216 = 1.6884m$$

$$\boxed{H_{pp} = 1.6884m}$$

Calculation of sum of control bands required for GL Control level:

$$Q_l = 6.7746x10^{-3}\frac{m^3}{s}$$

$$Ag_{\min} = 2.13824m^2$$

Let A be the length for the Levels LZA (LL) - LA (L)

The liquid Hold up time for this level $t_s = 60s$

i.e $$A = \frac{Q_l * t_s}{Ag_{\min}} = 0.19m > 0.1$$ (condition satisfied)

Let B be the length for the levels LA (L) – LA (H)

The liquid Hold up time for this level $t_s = 180s$

i.e $$B = \frac{Q_l * t_s}{Ag_{\min}} = 0.570m > 0.35m$$ (condition satisfied)

Let C be the length for the levels LA (L) – LA (H)

The liquid Hold up time for this level $t_s = 60s$

i.e $$C = \frac{Q_l * t_s}{Ag_{\min}} + 0.25 = 0.44m > 0.35$$ (Foaming allowance of 0.25m)

Therefore;

$$X_1 = A + B + C = 1.2m$$

$$\boxed{X_1 = 1.2m}$$

X_2 = Distance between the LZA (HH) and the bottom of the Schoepentoeter.

$$X_2 = 0.05D = 0.0825m$$

$$\boxed{X_2 = 0.0825m}$$

X_3 = Height of the Schoepentoeter.

$$X_3 = d_1 + 0.02m = 0.3248$$

$$\boxed{X_3 = 0.3248m}$$

X_4 = Distance between top of the Schoepentoeter to bottom of demisting device

$$X_4 = d_1 = 0.3048m$$

$$\boxed{X_4 = 0.3048m}$$

X_5 = Height of the Demisting device.

$$\boxed{X_5 = 0.1m}$$

X_6 = Distance between top of demisting device to Top tangent Length (TTL)

$$X_6 = 0.15d = 0.2475m$$

$$\boxed{X_6 = 0.2475m}$$

Total Length of the Separator,

$$\boxed{L = H_{pp} + X_1 + X_2 + X_3 + X_4 + X_6 + X_7}$$

$$\boxed{L = 3.9476m}$$

The tangent-to-tangent length of the separator, $L = 3.9476m$

8.7 Conclusion

The design for both the conditions show the dimensions of the vessel more or less the same. Therefore the final dimensions of the Separator vessel are:

Diameter of the Vessel, $\boxed{D = 1.650m}$

Tangent to tangent length of the vessel, $\boxed{L = 4.0m}$

The Ends of the Vessel are semi-elliptical.
The figure 8.1 shows the figure of the vessel.

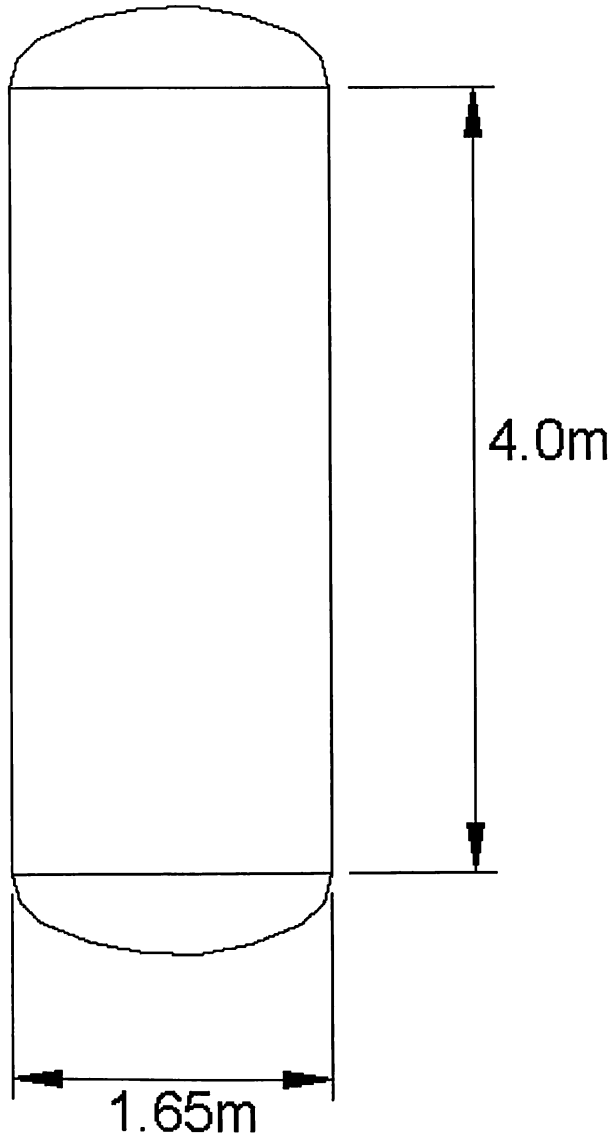


Fig 8.1

8.8 Design of Internals

i. Design of Schoepentoeter

a. We have available space in the vessels, E

$$E = D - 25 - \frac{D}{2}[1 - \cos(\sin^{-1}(\frac{d_1}{D}))]$$

$\boxed{D = 1.650m}$

and

$\boxed{d_1 = 0.2901m \approx 12"}$

Therefore; $E = 1610.8mm$

Since $E > 5d_1$ ie E to be type II or Type IV

$$E = 1524mm$$

b. Find L_v and R_v

$$\lambda = \frac{E-270}{d_1-30} = 4.88 \qquad \text{ie} \qquad 2.5 \le \lambda \le 6.0$$

$$\boxed{L_v = 100mm} \quad and \quad \boxed{R_v = 100mm}$$

c. Number of Vanes per side, n_v

$$n_v = \frac{E-R_v-25}{L_v} = 13.94 \approx 14$$

$$\boxed{n_v = 14}$$

d. Evaluate β

$$\text{We have, } Tan\beta = \frac{d-30_1}{2(n_v-1)L_v} = 0.1057$$

$$i.e. \qquad \boxed{\beta = 6.0332^0}$$

e. Let $\alpha = 8^0$

$$W_{vo} = L_v(\sin\alpha + \cos\alpha \tan\beta) - t$$

Choose t=3mm as for light duty (No slugs)

$$W_{vo} = 21.3835mm$$

ie. $W_{vo(\min)} = 20mm$ $W_{vo(\max)} = 30mm$

Figure 8.2 gives the schematic of the schoepentoeter

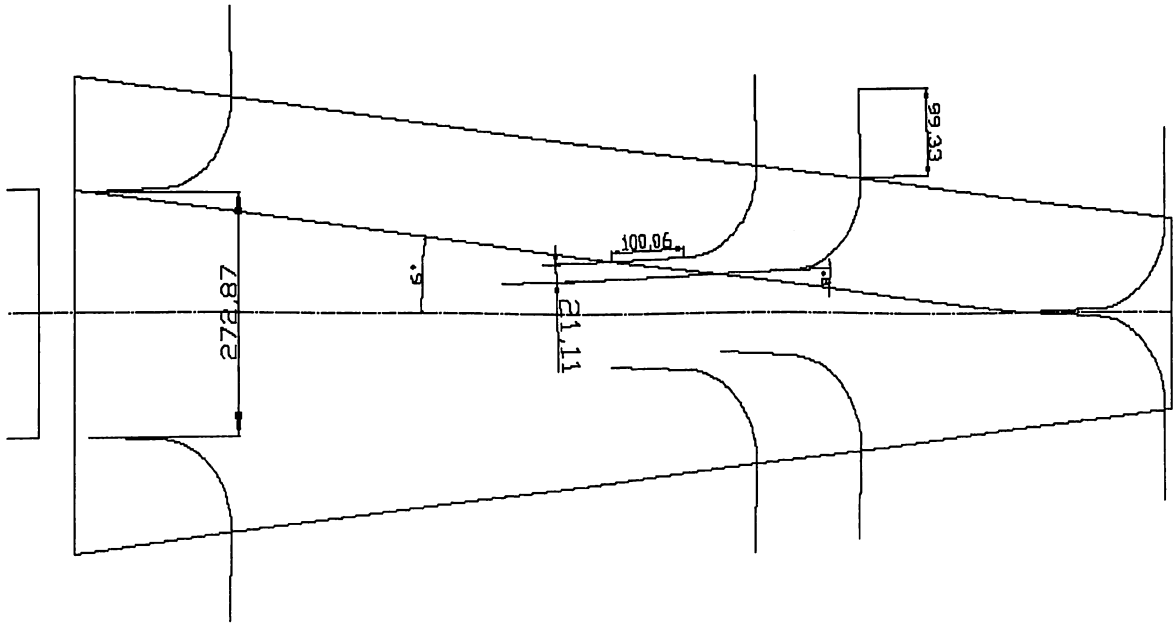


Fig 8.2

ii. Design Of the Mist mat

The demister matt shall be made of Knitted wire formed to give correct shape.
Demister mats are normally stainless steel.

a. The free volume of the mistmat = 97%
b. The wire thickness between 0.23mm to 0.28mm

The thickness of a horizontal mat in a vertical vessel is normally 0.1m
Mist mat shall be placed in between two girds having a free area of at least 97%. The mat shall be fastened in such a way that it cannot be compressed when being mounted.

iii. Design of Plate pack

As found from the calculations

a. The height of the plate pack, H_{pp} = 1.41m
b. Width of the plate pack, W_{pp} = 0.7m
c. Length of the plate pack, l_{pp} = 0.3m
d. Thickness of each plate = 1mm
e. Angle of the plate pack with horizontal, $\theta = 45^{\circ}$
f. Distance between two plates, d_{pp} = 20mm
g. The distance between the plate pack and perforated plate = 0.15m

The Final Diagram of the Vertical Three phase separator is shown in fig 8.3

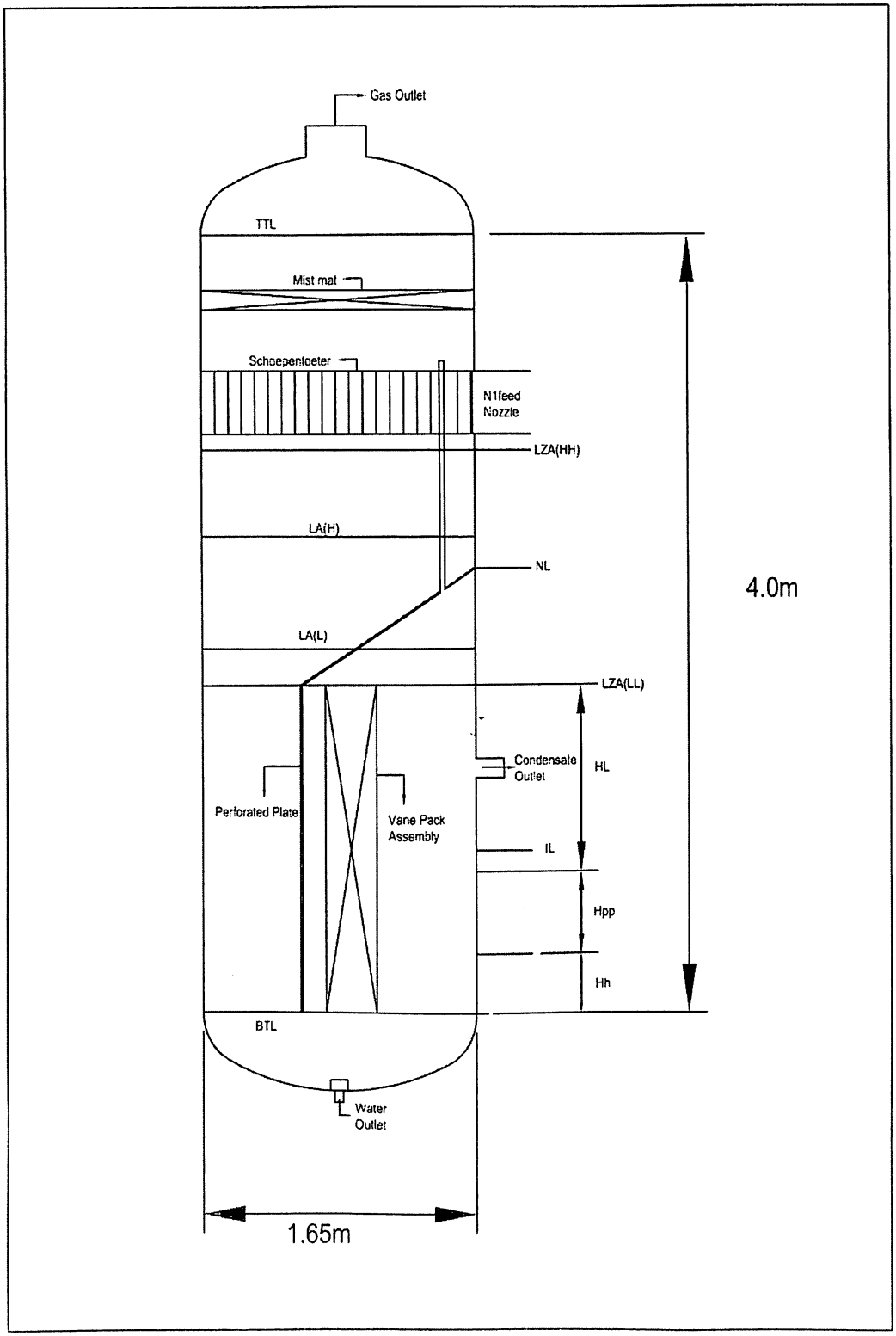

Gas Outlet
TTL
Mist mat
Schoepentoeter
N1feed
Nozzle
LZA(HH)
LA(H)
NL
LA(L)
LZA(LL)
Condensate
Outlet
HL
Perforated Plate
Vane Pack
Assembly
IL
Hpp
Hh
BTL
Water
Outlet
4.0m
1.65m

Fig 8.3

8.9 Material Selection for the piping

The material selection shall be optimized considering investment and operational cost, such that the life cycle costs (LCC) are minimized while providing acceptable safety and reliability.

The following key factors apply to material selection:

1. Primary consideration shall be given to materials with a good market availability and documented fabrication and service performance.
2. The number of different materials type shall be minimized considering cost, interchangeability and availability of relevant spare parts.
3. Design life
4. Operating conditions.
5. Expensive with the materials and corrosion protection methods from conditions with similar corrosivity
6. System availability requirement
7. Philosophy applied for maintenance and degree of system redundancy.
8. Weight reduction
9. Inspection and corrosion monitoring possibilities.
10. Effect of internal and external environment including compatibility of different materials.
11. Evaluation of failure probabilities, failure modes, criticalities and consequences. Attention shall be paid to any adverse effects material selection may have on a human health, environment, safety and material assets.
12. Environmental issues related to corrosion inhibition and other chemical treatments.
13. for main systems where material/ fabrication represent significant investment and operational costs and LCC analysis shall be basis for material selection.

Corrosivity evaluations in Hydrocarbon systems:

Evaluation of corrosivity shall a minimum include

- Co2 content
- H2S content
- Oxygen content and content of other oxidizing agents
- Acidicity, PH
- Haloginide concentration
- Velocity flow regime.

A gas system is defined wet when the relative humidity exceeds 50%

Oil and Gas processing:

When assessing corrosivity throughout a processing system, the partial pressure of CO2 and H2S in the gas phase in separators and scrubbers can be used as basis for evaluating corrosivity. To compensate for the fact that these gases are not at equilibrium in each separator, the following assumptions shall be made unless otherwise justified by experience.

- When the corrosivity in a separator and the liquid carrying piping down stream this separator is evaluated, the mean partial pressure of the foregoing separator and the actual separator shall be used.
- For gas scrubbers and piping downstreams separators carrying gas, the actual partial pressure can be used directly.

Pressure rating Maximum / minimum design temperature and size shall be taken into account and when selecting materials.

All components, which may contact oil well streams, shall be resistant against well treating, well simulating chemicals and other additives.

According to,
NORSOK Standard (Design Principals-Material selection M-OP-001 Rev.1. Dec. 1994)
By table 6.2 the material recommended for the following applications are:

Oil and gas Processing and Production
Piping and vessels -22CrDuplex, 6Mo, 316
Piping, Vessels for Produced water -316, 22CrDuplex, 6Mo, Titanium or GRP

Material Selection for piping for Project problem:

Since there is presence of H2S and CO2, the material selection can be either 22CrDuplex or 316 for Austenitic Stainless Steel.
22CrDuplex has more corrosion resistant properties than 316 stainless steel.
Therefore the material selected is 22CrDuplex

The properties of 22CrDuplex are given in the table below

Composition %							
Cr	Ni	C	S	Mn	Yeild strength Kip/in2(Mpa)	Tensile Strength Kip/in2(Mpa)	Elongation %
22-24	12-15	0.08	1.0	2.0	40(276)	95(655)	45

8.10 Sizing of the Piping

In determining the diameter of pipe used in separator piping both flow velocity and pressure drop should be considered, when determining line sizes, the maximum flow rate expected during the life of the facility should be considered rather than the initial flow rate. A surge factor of 20-50% to the anticipated normal flow rate, unless surge expectations have been more precisely determined by pulse pressure measurements in similar systems.

Determination of the nature of the phase in inlet for the project problem
The feed entering the piping system can be in the form of mist, stratified flow, slug flow etc. depending upon the flow rates and physical properties of the gas and liquid phases and on the feed pipe characteristics.

In Figure 8.4 and 8.5 Two flow maps are presented. The first flow map gives the two-phase flow regimes in a horizontal pipe and the second one is vertical pipe (up flow)

The gas and liquid Froude number is given by

$$Fr_g = v_g\sqrt{\frac{\rho_g}{(\rho_l-\rho_g)gd_1}} \qquad Fr_l = v_l\sqrt{\frac{\rho_l}{(\rho_l-\rho_g)gd_1}}$$

Where ρ_l=Mean density of the light and heavy phase

$$\rho_l = \frac{628\ .5 + 949\ .5}{2} = 789\ \frac{kg}{m^3}$$

& $\rho_g = 82.9\frac{kg}{m^3}$

$$v_g = \frac{Q_g}{\pi/4d_1^2} = 8.707 m/s \qquad v_l = \frac{Q_l}{\pi/4d_1^2} = 0.045005 m/s$$

Therefore;

$Fr_g = 1.75643$ $Fr_l = 0.028$

From the Fig8.4 and fig 8.5 we conclude that for Horizontal Flow the flow is annular dispersed flow and for vertical flow it is annular dispersed flow.

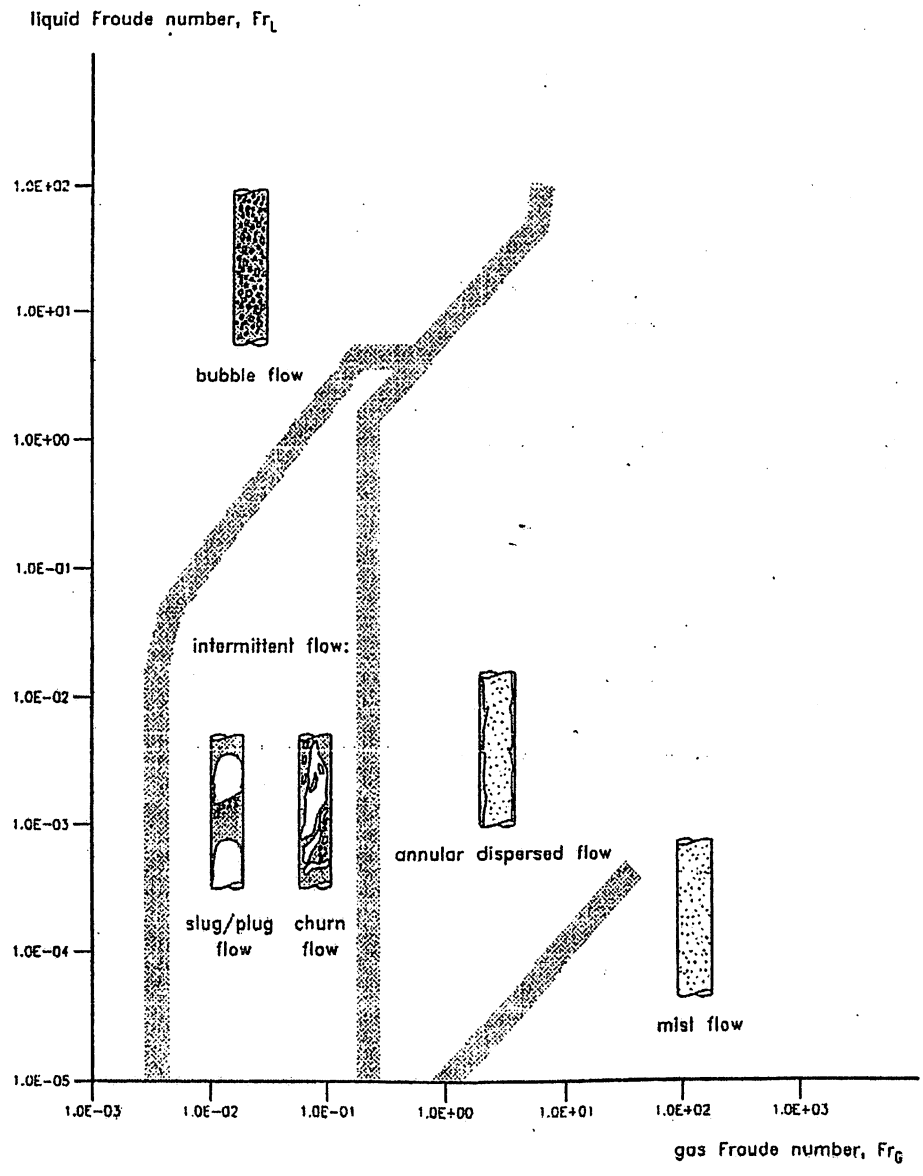


Fig 8.4
Two phase flow maps for vertical feed pipes

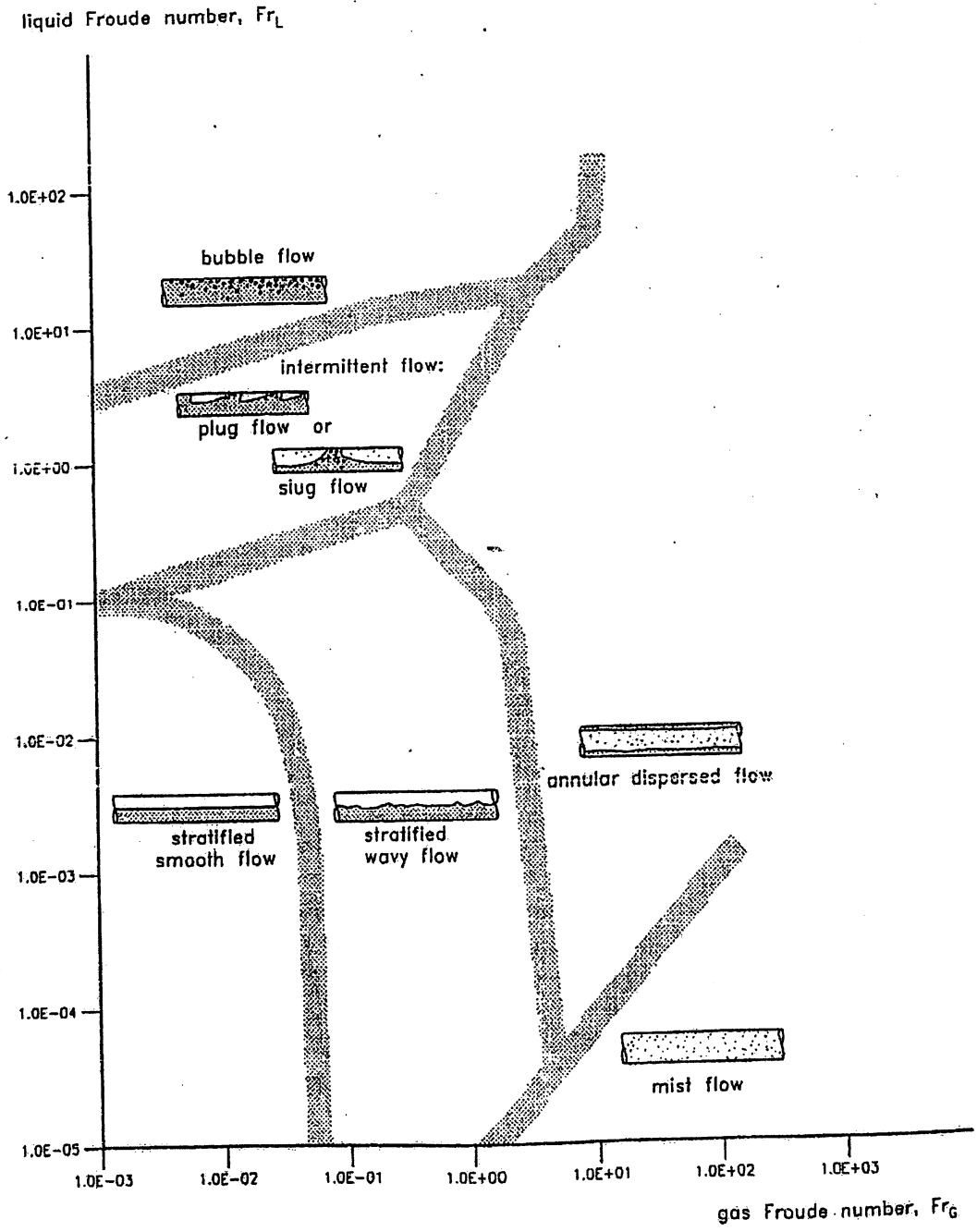


Fig 8.5
Two phase flow maps for Horizontal feed pipes

Sizing criteria for Gas/Liquid Two phase Inlet Pipe.

Erosional Velocity:

Flowlines, production manifolds, process headers and other lines transporting gas and liquid in two phase flow should be sized primarily on the basis of flow velocity. Experience has shown that loss of wall thickness occurs by a process of erosion/ corrosion. This process is accelerated by high fluid velocities, presence of sand, corrosive contaminants such as CO2 and H2S and fittings, which disturb the flow path such as elbows.

The velocity above which erosion may occur can be determined by the following empirical equation:

$$V_e = \frac{c}{\sqrt{\rho_m}}$$ Where V_e=Fluid erosional velocity ft/s

c = Empirical Constant

ρ_m =Gas/Liquid Mixture density at flowing temp. and pressure *lbs/ft3*

Industry experience to date indicates that for solids free fluids values of C=100 for continuous service and c=125 for intermittent service are conservative. For solids free fluids where corrosion is not anticipated or when corrosion is controlled by inhibition or by employing corrosion resistant alloys, values of C=150 to 200 have been used for continuous service; values up to 250 have been used successfully for intermittent service, if solids production is anticipated, fluid velocities should be significantly reduced. Different values of C may be used where specific application studies have shown them to be appropriate.

Where solids and or corrosive contaminants are present or where "c" values higher than 100 for continuous service are used, Periodic surveys to assess pipe wall thickness should be considered. The design of any piping system where solids are anticipated should consider the installation of sand probes, cushion flow tees and a minimum of three feet of straight piping downstream of choke outlets.

Density of the Gas/Liquid Fluid Mixture:

The density of the Gas/Liquid Mixture may be calculated using the following derived Eqn.:-

$$\rho_m = \frac{1240\gamma_l p + 2.7R\gamma_g p}{198.7p + RTz} \frac{lb}{ft^3}$$

Where, p = Operating Pressure, PSIA

γ_l = Liquid Specific Gravity(Water=1 use average gravity for hydrocarbon-water mixture) at standard conditions.

R= Gas/Liquid ratio $\frac{ft^3}{barrel}$ at standard conditions.

T = Operating Temperature R

γ_g = Gas specific gravity (Air=1) at std conditions.

z = gas compressibility factor, dimensionless.

Calculation of v_e and ρ_m for the project problem

We have; $p = 125$ barg $= 1827.671$ psia

$T = 104$ c $= 678.87$ R

$\gamma_l = \frac{0.639 + 0.977}{2} = 0.808$ (Taken from the feed data for flowline)

$R = 64498.6823$ (Found from separator calculations)

$Z = 0.86$

Therefore,

$$\rho_m = \frac{220456179.4}{38019307.82} = 5.7985 \frac{lb}{ft^3}$$

Assuming the value of $c = 100$

$$\text{Therefore, } v_e = \frac{c}{\sqrt{\rho_m}} = \frac{100}{\sqrt{5.7985}} = 41.5281 ft/s$$

Sizing of the Inlet Pipe

Calculation of Minimum cross sectional area required to avoid fluid erosion.

We have $$A = \frac{9.35 + \frac{zRT}{21.25p}}{v_e} \frac{in^2}{1000 Barrels/day}$$

Therefore;

$$A = \frac{9.35 + \frac{0.86 * 64498.6823 * 678.87}{21.25 * 1827.671}}{41.5281} = 23.5724 \frac{in^2}{1000 Barrels/day}$$

We have

$Q_l = 36.10 m^3/hr = 5.4494921$ *Thousand barrels/day*

Total Area,

$A = 23.5724 * 5.4494 = 128.45794 in^2$

But;

$$A = \frac{\pi d_1^2}{4} \quad \Rightarrow \quad d_1 = 12.789 in$$

Which is close to the value considered $d_1 = 12.0 in$

Therefore the value of inlet pipe is correct i.e. $d_1 = 12.0 in$

Sizing of the Condensate and water outlet pipes.
For condensate and water outlet pipes assume the single-phase liquid lines.

Sizing criteria for liquid lines:
The single-phase liquid lines should be sized primarily on the basis of flow velocity. For lines transporting liquids in single phase from one pressure vessel to another by pressure differential, the flow velocity should not exceed 1.5 feet/sec at maximum flow rates, to minimize flashing ahead of the control valve. If practical flow velocity should not be less than 3 ft/s in order to minimize the sand and other solid deposition.

Let us assume that the diameter of condensate outlet be equal to the outlet nozzle of separator.

i.e. $d_3 = 4$"

$$v_l = \frac{Q_l}{\pi/4 d_1^2} \Rightarrow \qquad v_l = \frac{42.48}{\pi 0.1016^2/4} = 1.45547 m/s = 4.7751640 ft/s$$

i.e. the criterion $3 < v_l < 5$ satisfies, therefore Our estimate of outlet condensate nozzle diameter is correct

i.e. $d_3 = 4$"

Let d_4 be the diameter of the water outlet pipe. As we have taken nozzle diameter for water outlet as 2", now let us assume that diameter of the piping be 2"
Check for the velocity criterion

$$\text{i.e. } v_l = \frac{1.2986 x 10^3}{\pi 0.0508^2/4} = 0.64070 m/s = 2.1020 ft/s$$

There is no sand production, the diameter of the water outlet piping can be taken as 2"

i.e $d_4 = 2$"

Sizing of the gas outlet piping

The 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 create noise problem if it exceeds 60ft/s. The design of any piping system where corrosion inhibition is expected to be utilized should consider the installation of additional wall 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.

Let us assume that the diameter of the gas outlet piping be equal to outlet gas nozzle.

i.e. $d_4 = 14" = 0.3356$ $\qquad A = \frac{\pi d_1^2}{4} = 0.09932 m^2$

$$Q_g = 0.55195 m^3/s$$

$$v_g = \frac{Qg}{A} = \frac{0.55195}{0.09932} = 5.5575 m/s = 18.2333 ft/s$$

i.e. v_g <60ft/s for gas flow in pipe

Therefore the diameter of the gas outlet pipe d_4 = *14*" can be taken.

8.11 Pipe wall Thickness Calculation

The pipe wall thickness required for a particular piping service is primarily a function of internal operating pressure and temperature. The standard under which pipe is manufactured permit a variation in wall thickness below nominal wall thickness. It is usually desirable to include a minimum corrosion/mechanical strength allowance of 0.05" for carbon steel piping.

The pressure design thickness required for a particular application may be calculated by the following equation. From ANSI B31.3

$$t = \frac{p_i D_o}{2(SE + p_i y)}$$

where,

t - Pressure design thickness, inches. Minimum wall thickness minus corrosion/Mechanical strength allowance or thread allowance.

p_i - Internal design Pressure, Psig

D_o - Pipe outside diameter, inches.

E –Longitudinal weld joint factor (See ANSI B31.3) = 1.00 for seamless = 0.85 for ERW

y - Temperature factor (0.4 for ferrous material at 900 F or below when t<D/6)

S - Allowable stress in accordance with ANSI B31.3 Psi

For the selected material 22CrDuplex the allowable stress in accordance with ANSI B31.3 is = 90 Kip/in2 = 620.5281Mpa

& E = 1 as seamless pipe is used.

y = 0.4 for temperature = 69.9 c from ASME B31.3

p_i = 109.0135 bar

Therefore,

For 12" Pipe $D_o = 0.3048m$

$$t = \frac{109.0135x10^5 * 0.3048}{2(620.5281x10^6 + 109.1035x10^5 x0.4)} = 2.6586x10^3 m = 2.65mm$$

Similarly

For 14" pipe $t = 3.10176mm$

For 4" Pipe $t = 0.8862mm$

For 2" Pipe $t = 0.4431mm$

This thickness is the minimum Thickness for the applied pressure.

But the standard available pipes have to be selected based on this minimum thickness.

The pipes selected from ASME B36.10M – 1996 Standard are

14" Pipe - Schedule 10 – Thickness 6.35mm
12" Pipe – Schedule 10 -std– Thickness 9.53mm
4" Pipe - Schedule 40-std-- Thickness 6.02mm
2" Pipe - Schedule 40 – std-Thickness 3.91mm

Fittings:
14" Elbow - Schedule 10 – Thickness 6.35mm
12" Elbow -std– Thickness 9.52mm
4" Elbow - std-- Thickness 6.02mm
2" Elbow – Std – Thickness 3.91mm

Pressure rating of the Inline components.

Determination of pressure rating for In-line components.

Following is the procedure for the pressure rating calculation:
1.The operating conditions like temperature & pressure of the system is known initially.
2. The Material of the piping system is also known
3. The ASTM specifications classifies the material under groups, determine the group.
4. Refer to ANSI pressure-temperature ratings for different classes.
5. Identify the class for the given pressure & temperature.

Pressure rating for components in project problem

The material of the in-line component is 22CrDuplex
The operating conditions are
Temperature - 35-80^{0}C
Pressure - 750 KPa

The design temperature and pressure:
Temperature - 100^{0}C = 212^{0}F
Pressure - 950 KPag = 137.78585 psig
= 152.48585 psia
Refer : ASME B16.5
For 22CrDuplex i.e. material group = 2.8
From pressure temperature rating table 2.2.8
For 50 – 100^{0}C, the pressure is 109.0135bar.

Therefore select 900 # class for minimum pressure up to 131bar

Therefore Rating chosen for In-Line components = 900 #
The dimensions of the Fittings, Flanges, valves etc are taken form the respective standards

Finally the Piping Layout of the three-phase Test Separator is shown in figure 8.6

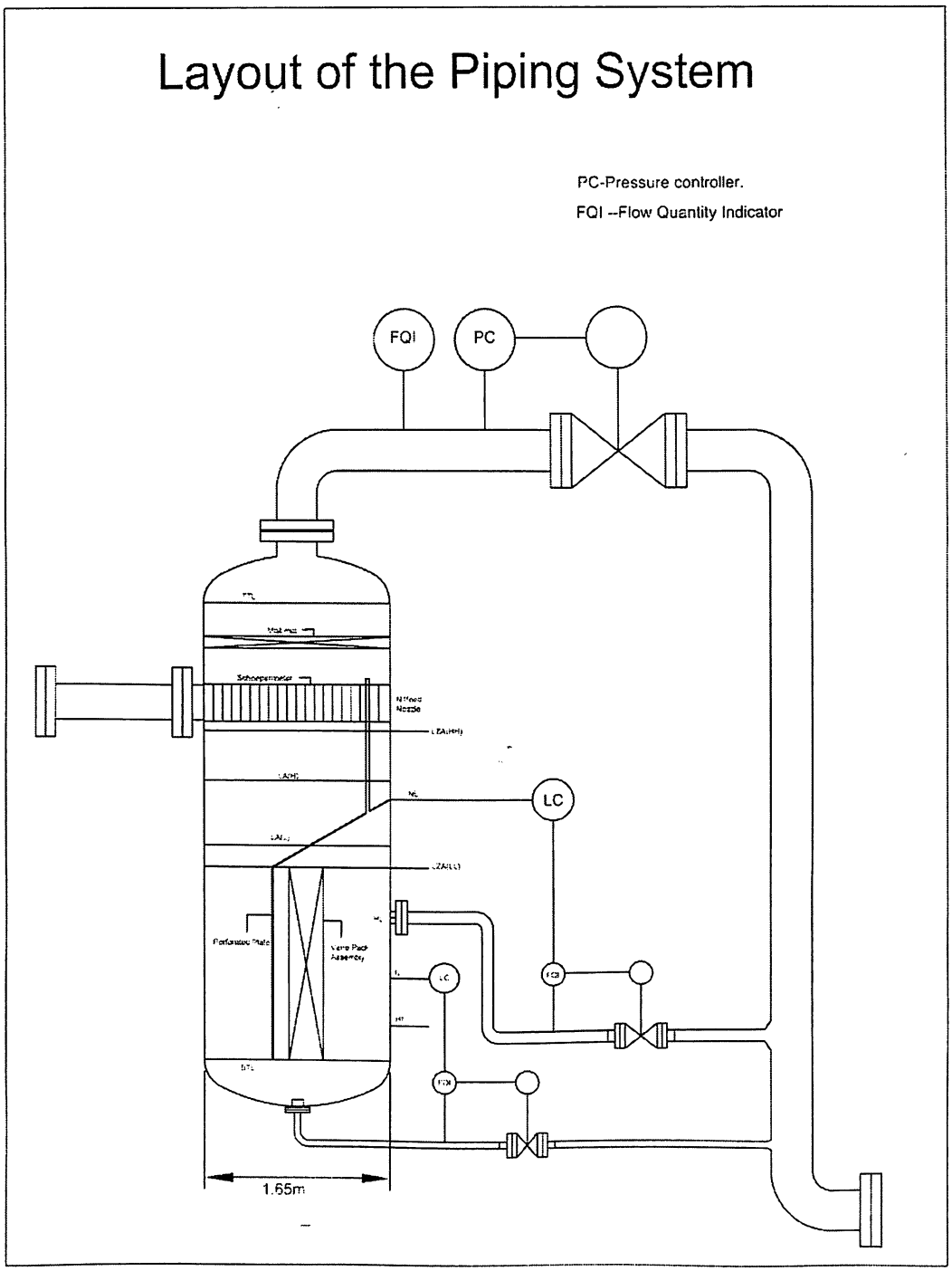

Layout of the Piping System
PC-Pressure controller.
FQI --Flow Quantity Indicator
FQI
PC
LC
1.65m

Fig 8.6

Chapter 9

Conclusions and Recommendations

9.1 Summary

Field processing of natural gas includes gas and liquid separation operations, recovery of condensable hydrocarbons, gas dehydration and plant processing. Plant processing includes desulphurization and further treatment for the recovery of natural gas liquids (NGL). The gas and liquid separation should cause a primary separation, refine and discharge the separated gas.

The project problem is a to design a three phase test separator for installation on offshore platform in Qatar. The details of the required output are summarized in chapter 1. The design principles should be adopted from the standard engineering practices. A detail datasheet and FEED data is also given. The composition of the gasses at inlet to the test separator is given. The outline presented here gives the sequence of operations performed during this project.

The classification done here is based on the standard practices. Various types of separators mentioned here give their application for given situations. The classification done here is based on the separation fluids and orientations. Each separator given here gives the information for their characteristics, recommended use, non-recommended use and typical process applications. Based on these parameters the suitability of the separator is judged.

The selection of the separator is based on the selection strategy described in Recommended practices. The strategy includes comparing different separators on each of these criteria. I.e. Gas handling capacity (Maximum capacity and turn down ratio), Liquid removal efficiency (Overall, with respect to fine mist, flooding above λ_{max}), liquid handling capacity (i.e. as slugs and as droplets), fouling tolerance (Sand and sticky material) and pressure drop. Based on these parameters a suitable separator is selected. For three-phase separator, type of dispersion and separation efficiency commands the suitability. The tables given in each of these chapters guide the designer to select the best separator.

The separation design principles presented here gives the sizing calculations for dimensions of vessel, liquid level control for G-L Separator, Design of internals like shoepentoeter, vane pack and Plate pack. The design margins given are applied for the input data depending upon the environment on which the separator is fitted.

The equilibrium calculations are made for the separator for finding the composition of liquid and gas phases from the output. The principles presented for equilibrium calculations for three-stage separation are used to find the Gas to Oil Ratio (GOR). The equilibrium constants applied for the equilibrium calculations should be accurate to get correct output results. A method of determining equilibrium constant by using equations of state is given. This method is utilized to find the equilibrium constants for the separation in our project.

A good control system is very much necessary for the correct separation of liquids and gasses. The liquid level control in G-L and G/L/L Separators is very critical for its performance. The various methods discussed here are applicable to most of the separators.

The piping system is discussed here is the most common piping system for the applied for field treatment of natural gases. The oil and gas gathering system described here gives the general idea of piping configuration. Here the piping and accessories described gives the knowledge of the layout of a piping system for three-phase separation.

The design of the project problem is presented in chapter 8 here the equilibrium calculations for primary separation gives the important data like GOR of well, specific gravity of gas, Gravity of Condensate and composition of the separated liquids and gases. Employing principles of standard practices does the sizing of separators vessel. Accordingly the nozzles are also sized. The design of schoepentoeter, mistmat and plate pack assembly is done. The final diagram of the designed separator is shown.

Considering the presence of H2S and CO2 in the incoming stream does material selection for the piping. This selection is based on a standard practice done in design engineering. The sizing of the piping is done by knowing the nature of the fluid flow. The pipe wall thickness and rating of the inline fittings and valves is also done by considering various piping standards.

9.2 Contributions and Conclusions

The following contributions were made in this study:

- A detailed summary of the actual design of G/L and G/L/L separator is presented in a short and compact manner
- The flow charts developed here gives the quick understanding of the sizing of vessel of separator, which is based on criterions like ratio of length to diameter, residence time, defoaming and degassing.
- The methods adopted here give the better understanding for a fresh graduate in actual work scenario in detail engineering.
- The design calculations helped to technically evaluate the dimensions of the vessel and ensure the output meets the desired quality from the separator.
- The method of material selection for gas piping presented here helps to understand the standard practice practiced in detail engineering
- The design methods presented here gives better understanding of the use of a suitable separator under different operating conditions.

9.3 Recommendations for further work

- The methods adopted here can be easily programmed using programming languages. The next recommended work after this project is to implement these methods over a computer, to give quick selection and design
- There are many other types of separators present in the actual field these are not mentioned here. Such separators should be explored and their design should be worked upon.
- The variation of the separator performance with respect to the operating pressure, temperatures, and flow rate needs to be studied.

- The Algorithms used in finding the Equilibrium constant calculations using C-Programming language should be worked upon, as the algorithms used here are inefficient.

- The stage separation in field treatment of gas needs to be studied and optimum pressures drops are to be found for these groups of separators.

- The material selection of the vessel is not done here this can be worked upon using standard engineering Practices.

- Mechanical Vessel design is also not done here this can be studied.

Nomenclature

Symbols

A	Area of Cross section or cord area
A^*	Dimensionless Area
Ar	Archimedes number
Af	Frontal Area for plate pack
d	Distance
d_3	Diameter of the liquid outlet nozzle.
d_2	Diameter of the gas outlet nozzle
d_1	Diameter of the inlet pipe.
D	Internal Diameter of vessel
E_f	Liquid Removal Efficiency
f	Correction factor
fu	Fugacity
h	Height
h^*	Dimensionless Height
k	Separation Coefficient
L	Tangent to Tangent Length
l	Length
$LZA(LL)$	Low level trip
$LA(L)$	Low level pre-alarm
$LA(H)$	High level Pre-alarm
$LZA(HH)$	High level Trip.
m	Mass flow rate
nl	Normal level
p	Pressure
p_r	Reduced Pressure
Q	Flow Rate or Capacity
R	Universal Gas constant
R_e	Reynolds number.
t	Time
t_{pp}	Thickness
T	Temperature
T_r	Reduced Temperature
v_m	Mean velocity of the mixture
Δv	Volume between control heights
v	Velocity
w	Width

Greek Symbols

Φ	Flow parameter
λ	Gas load factor
ρ	Density
μ	Viscosity
θ	Angle
ϵ	Error

Subscripts

g	Gas
max	Maximum
l	Light liquid
m	Mean value
pp	Plate pack
gross	Gross value
net	Net value
loss	loss correction factor
h	Heavy liquid
contot	Control band
db	Dispersion band
con	Control
pset	Settling
ax	Axial
entry	Entry

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Appendices

Appendix A

Calculations of Equilibrium Coefficients for the Project problem.

Given conditions: Temperature, T = 69.9 °c = 617.49 °R
Pressure, P = 108 barg = 109.0135bara = 1581 Psia

Molar Composition (%)	Z
Methane	77.9952
Ethane	4.5502
Propane	1.7002
Iso-butane	0
N-butane	0.9277
Iso-Pentane	0
N-Pentane	0.5077
N-Hexane	0.2852
Heptanes +	2.2702
H2S	1.3652
H2O	2.9877
Nitrogen	4.3152
CO2	3.0952

1. Calculate the coefficients of the components of the mixture.

$$\alpha_j^{1/2} = 1 + (0.37464 + 1.54226 s_j - 0.26992 s_j^2)(1 - T_{rj}^{1/2})$$

$$a_{cj} = 0.45724 \frac{R^2 T_{cj}^2}{p_{cj}}$$

$$a_{Tj} = a_{cj} \alpha_j$$

$$b_j = 0.07780 \frac{R T_{cj}}{p_{cj}}$$

Components	T_{cj} R	p_{cj}	w_c	α_j	a_{cj}	a_{Tj}	b_j
Methane	343.1	667.8	0.0115	0.7499	9283.261099	6961.876714	0.42898
Ethane	549.8	707.8	0.0908	0.9395	22490.776	21129.83811	0.64857
Propane	665.7	616.3	0.1454	1.0446	37867.83626	39555.15922	0.90188
Iso-butane	765.4	550.7	0.1928	1.1409	56023.13184	63917.58734	1.16047
N-butane	734.7	529.1	0.1756	1.1099	53726.41498	59632.91982	1.1594
Iso-Pentane	845.4	488.6	0.251	1.2322	77032.95688	94922.47099	1.44467
N-Pentane	828.8	490.4	0.2273	1.2072	73765.71688	89049.30792	1.41111
N-Hexane	913.4	436.8	0.2957	1.3146	100587.7118	132230.8459	1.74598
Heptanes +	972.5	396.8	0.3506	1.4022	125520.0491	176004.4294	2.04634
H2S	227.3	493	0.0355	0.5209	5518.963175	2874.716798	0.38496
H2O	547.6	1070.9	0.225	0.9131	14746.31492	13464.75496	0.42695
Nitrogen	672.4	1306	0.0949	1.0439	18231.30584	19031.08439	0.42988
CO2	1165.1	3203.6	0.321	1.524	22314.83416	34008.41882	0.30366

2. Select the trial values of K-factors and calculate trial compositions of equilibrium gas and liquid. Only the final trial, with K-factors as given below, is shown

$$\sum X_i = \sum \frac{Z_i}{1 + \bar{n}_g (K_i - 1)} = 1$$

This calculation requires trial and error; only the final trial with $\bar{n}_g = 0.9784$ is shown

Components	K	Z	X	Y
Methane	2.534574	77.9952	0.308765	0.782606
Ethane	1.052380	4.5502	0.043249	0.045515
Propane	0.580597	1.7002	0.029165	0.016933
Iso-butane	0.321635	0	0	0
N-butane	0.392457	0.9277	0.023435	0.009197
Iso-Pentane	0.184549	0	0	0
N-Pentane	0.212169	0.5077	0.023442	0.004974
N-Hexane	0.110019	0.2852	0.024799	0.002728
Heptanes +	0.064889	2.2702	0.323730	0.021006
H2S	0.709137	1.3652	0.019207	0.013621
H2O	0.172190	2.9877	0.0168967	0.029094
Nitrogen	4.543544	4.3152	0.009539	0.043341
CO2	1.175472	3.0952	0.168967	0.030978

3. Calculate the composition dependent coefficients necessary for z-factor calculations for both liquid and gas.

$$a_T = \sum_i \sum_j Y_i Y_j (a_{Ti} a_{Tj})^{1/2} (1 - \delta_{ij})$$

$$b = \sum_j Y_j b_j$$

$$A = \frac{a_T p}{R^2 T^2}$$

$$B = \frac{bp}{RT}$$

Phase	a_T	b	A	B
Liquid	54185.449219	1.027303	1.950604	0.245073
Gas	10400.047852	0.490543	0.374388	0.117024

4. Calculate the z-factors of liquid and gas.

$$z^3 - (1 - B)z^2 + (A - 2B - 3B^2)z - (AB - B^2 - B^3) = 0$$

i.e. $\boxed{z_L = 0.35431}$ and $\boxed{z_G = 0.8149}$

5. Calculate the composition dependent coefficients necessary for calculating fugacity coefficients for both liquid and gas.

$$A_j' = \frac{1}{a_t}[2a_{Tj}^{1/2} \sum Y_i a_{Ti}^{1/2} (1 - \delta_{ij})]$$

$$B_j' = \frac{b_j}{b}$$

Components	Liquid		Gas	
	$A_j^{'}$	$B_j^{'}$	$A_j^{'}$	$B_j^{'}$
Methane	0.716889	0.417577	1.636347	0.874495
Ethane	1.248926	0.631329	2.850758	1.322138
Propane	1.708797	0.877906	3.900442	1.838524
Iso-butane	2.172195	1.129627	4.958182	2.365681
N-butane	2.098127	1.128585	4.789115	2.363497
Iso-Pentane	2.647118	1.406276	6.04224	2.945042
N-Pentane	2.563917	1.373603	5.852313	2.876617
N-Hexane	3.124317	1.699574	7.131464	3.559272
Heptanes +	3.604546	1.991956	8.227617	4.171581
H2S	1.185279	0.418453	2.705478	0.876329
H2O	1.584462	0.295588	3.616642	0.619024
Nitrogen	0.460666	0.374726	1.051501	0.784757
CO2	0.996983	0.415601	2.275681	0.870357

6. Calculate the fugacity coefficients of the components of liquid and gas.

$$\ln\phi_j = -\ln(z-B) + (z-1)B_j^{'} - \frac{A}{2^{1.5}B}(A_j^{'} - B_j^{'})\ln\left(\frac{z+(2^{1/2}+1)B}{z-(2^{1/2}-1)B}\right)$$

Components	ϕ_{Lj}	ϕ_{Gj}
Methane	2.301335	0.907977
Ethane	0.614698	0.584102
Propane	0.237401	0.408892
Iso-butane	0.091932	0.285827
N-butane	0.120655	0.307436
Iso-Pentane	0.036820	0.199512
N-Pentane	0.045367	0.213825
N-Hexane	0.015389	0.139875
Heptanes +	0.006342	0.097736
H2S	5.225814	1.150163
H2O	0.808298	0.687637
Nitrogen	0.405194	0.571390
CO2	0.064	0.366503

7. Calculate the K-factors of the components and the error functions

$$K_j = \frac{\phi_{Lj}}{\phi_{Gj}}$$

$$\varepsilon_j = \frac{(K_j^T - K_j^C)^2}{K_j^T K_j^C}$$

Components	K	ε_j
Methane	2.534574	0
Ethane	1.052380	0
Propane	0.580597	0
Iso-butane	0.321635	0
N-butane	0.392457	0
Iso-Pentane	0.184549	0
N-Pentane	0.212169	0
N-Hexane	0.110019	0
Heptanes +	0.064889	0
H2S	0.709137	0
H2O	0.172190	0
Nitrogen	4.543544	0
CO2	1.175472	0
		0

Conclusion: The sum of the error functions is less than a tolerance of 0.001, so set of trial values of K-factors was correct and the calculated values of liquid and gas compositions are correct.

Appendix B

C-Program for the calculation of the equilibrium constant for a given composition of Gas

The following program in written in C-language.
The calculation of equilibrium constant requires repetitive calculation until convergence is reached for the correct value. The program is written in Turbo C.

```
#include<stdio.h>
#include<stdlib.h>
#include<conio.h>
#include<math.h>

int main(void)
{
float tc[14],pc[14],w[14],m[14],z[14],M[14],alpha[14],ac[14],at[14],b[14],tr[14],pr[14];
float t,p,k,kin[14],ng,sumx,x[14],sumy,y[14],zl,zg,eravg;
float anl[14],ang[14],bnl[14],bng[14],phl[14],phg[14],fl[14],fg[14],Kfin[14],err[14],er;
float A,B,C,D,E,F,G,H,atl,atg,bl,bg,Bl,Bg,al,ag,ngf,sumz;
float pl,ql,pg,qg,il,ig,r,ll,ww,signq,sinq,u,v,x1,x2,x3,max,phi,s;
float f,flll,vk,vkk,fll,e;
int i,j,l,c,ln;

clrscr();
tc[1] = 343.1;tc[2] = 549.8;tc[3] = 665.7;tc[4] = 765.4;tc[5] = 734.7;tc[6] = 845.4;tc[7] =
828.8;tc[8] = 913.4;tc[9] = 972.5;tc[10] = 227.3;tc[11] = 547.6;tc[12] = 672.4;tc[13] =
1165.1;
pc[1] = 667.8;pc[2] = 707.8;pc[3] = 616.3; pc[4] = 550.7; pc[5] = 529.1;pc[6] =
488.6;pc[7] = 490.4;pc[8] = 436.8;pc[9] = 396.8;pc[10] = 493.0;pc[11] = 1070.9;pc[12]
= 1306.0,pc[13] = 3203.6;
w[1] = 0.0115;w[2] = 0.0908;w[3] = 0.1454;w[4] = 0.1928;w[5] = 0.1756;w[6] =
0.2510;w[7] = 0.2273;w[8] = 0.2957;w[9] = 0.3506;w[10] = 0.0355;w[11] =
0.2250;w[12] = 0.0949;w[13] = 0.3210;
m[1] = 16.043;m[2] =30.070;m[3] = 44.097;m[4] = 58.124;m[5] = 58.124;m[6] =
72.151; m[7] = 72.151;m[8]=86.178;m[10] = 28.013; m[11] = 44.01; m[12] = 34.076;
m[13] = 18.015;
printf("\nGive the temperature R (OR if wanna exit press zero)\n");
scanf("%f",&t);
if(t==0)
{
printf("\nTemperature cant be zero");
goto q;
}
printf("Give the Pressure PSIA\n");
scanf("%f",&p);
```

```
if(p==0)
{
printf("\nPressure cant be zero");
goto q;
}
//printf("\ngive the Flow rate");
//scanf("%f",&f);
printf("plz give the Mol wgt of Hepatnes Plus \n  or \n otherwise to use the default heptane mol wgt press 0\n");
scanf("%f", &k);
if(k>=1)
m[9] = k;
else
m[9] = 100.205;
printf("Give the compositions in fraction\n");
for(i=1;i<=3;i++)
{
printf("\nc%d=", i);
scanf("%f",&z[i]);
}
j=4;
for(i=4;i<=6;i=i+2)
{
if(i>=5) j=i-1;
printf("\ni-c%d=",j);
scanf("%f",&z[i]);
printf("\nn-c%d=",j);
scanf("%f",&z[i+1]);
}
for(i=8;i<=9;i++)
{
printf("\nn-c%d=",i-2);
scanf("%f",&z[i]);
}
i=10;
printf("\nN2=");
scanf("%f",&z[i]);i++;
printf("\nCO2=");
scanf("%f",&z[i]);i++;
printf("\nH2S=");
scanf("%f",&z[i]);i++;
printf("\nH2O=");
scanf("%f",&z[i]);
sumz=0;
for( i=1;i<=13;i++)
sumz=sumz+z[i];
if(sumz>=1.0001||sumz<=-0.0001)
{
        printf("\n The sum of Compostion is not equal to 1");
```

```
goto q;
}

 i=1;
for( i=1;i<=13;i++)
{
tr[i]=t/tc[i];
pr[i]=p/pc[i];
kin[i]=exp(5.37*(1+w[i])*(1-1/tr[i]))/pr[i];
M[i]=0.37464+(1.54226*w[i])-(pow((0.26992*w[i]),2));
alpha[i]=pow((1+M[i]*(1-sqrt(tr[i]))),2);
ac[i]=pow((10.732*tc[i]),2)*0.45724/pc[i];
at[i]=alpha[i]*ac[i];
b[i]=0.07780*10.732*tc[i]/pc[i];
}

c=0;

do
{
ng=1;
er=0;
i=1;
l=0;
//////////////////Ng iterations
do
{
sumx=0;
for(i=1;i<=13;i++)
{
x[i]=z[i]/(1+ng*(kin[i] - 1.0));
sumx=sumx+x[i];
}

if(sumx<=1.001&&sumx>=0.999) break;
 ngf=ng;
ng=ng-0.0001;
l++;
}while((ngf-ng)<=0.001&&(ngf-ng)>=-0.001);

//////////////////////Ng iterations finished

i=1;
for(i=1;i<=13;i++)
{
y[i]=x[i]*kin[i];
}
i=1;
sumy=0;
```

```
for(i=1;i<=13;i++)
{
sumy=sumy+y[i];
}

i=1;
j=1;
atl=0;
atg=0;
bl=0;
bg=0;
for(i=1;i<=13;i++)
{
 for(j=1;j<=13;j++)
 {
        atl=atl+x[i]*x[j]*sqrt(at[i]*at[j]);
        atg=atg+y[i]*y[j]*sqrt(at[i]*at[j]);
 }
 bl=bl+x[i]*b[i];
 bg=bg+y[i]*b[i];
}
al=atl*p/(pow((10.732*t),2));
ag=atg*p/(pow((10.732*t),2));
Bl=bl*p/(10.732*t);
Bg=bg*p/(10.732*t);
A=1;
B=1-Bl;
B=-1*B;
C=al-(2*Bl)-(3*pow(Bl,2));
D=(al*Bl)-(pow(Bl,2))-(pow(Bl,3));
D=-1*D;
E=1;
F=1-Bg;
F=-1*F;
G=ag-(2*Bg)-(3*pow(Bg,2));
H=(ag*Bg)-(pow(Bg,2))-(pow(Bg,3));
H=-1*H;
////////////////////////////////////////////////////
////////////roots
pl=(C/3.0)-pow(B/3.0,2);
ql=B/3.0*(pow(B/3.0,2)-(C/2.0))+D/2.0;
pg=(G/3.0)-pow(F/3.0,2);
qg=F/3.0*(pow(F/3.0,2)-(G/2.0))+H/2.0;

il= (pow(pl,3))+(pow(ql,2));
ig= (pow(pg,3))+(pow(qg,2));
```

```
if(il>0.001)
{
        r=sqrt(il);
        ll=-1*ql+r;
        ww=(-1*ql)-r;
        if(ll>0)signq=1;
        if(ll<0)signq=-1;
        if(ww>0) sinq=1;
        if(ww<0) sinq=-1;
        if(ll<0)ll=-1*ll;
        if(ww<0)ww=-1*ww;
        u=pow(ll,0.3333333)*signq;
        v=pow(ww,0.3333333)*sinq;
        x1=u+v-(B/3.0);
        zl=x1;
}

if(il<=0.001&&il>=-0.001)
{
        k=-1*ql;
        if(k>0)signq=1;
        if(k<0)signq=-1;
        x1=2*sqrt(-1*pl)*signq-B/3.0;
        x2=-1*(x1/2.0)-(B/3.0);
        x3=x2;
        if(x1>x2)
        {
                if(x1>x3)       max=x1;
                else max=x3;
        }
        else
        {
                if(x2>x3) max=x2;
                else max=x3;
        }
        zl=max;
}
if(il<-0.001)
{
        phi=acos(-1*ql/(sqrt(-1*pow(pl,3))));
        s=2*sqrt(-1*pl);
        x1=(s*cos(phi/3.0))-(B/3.0);
        x2=(s*cos((phi/3.0)+(2.0*3.142/3.0))-(B/3.0));
        x3=(s*cos((phi/3.0)+(4.0*3.142/3.0))-(B/3.0));
        if(x1>x2)
        {
                if(x1>x3)       max=x1;
                else max=x3;
        }
        else
```

```
        {
                if(x2>x3) max=x2;
                else max=x3;
        }
        zl=max;
}

///////////////

if(ig>0.001)
{

        r=sqrt(ig);
        ll=-1*qg+r;
        ww=-1*qg-r;
        if(ll>0)signq=1;
        if(ll<0)signq=-1;
        if(ww>0) sinq=1;
        if(ww<0) sinq=-1;
        if(ll<0)ll=-1*ll;
        if(ww<0)ww=-1*ww;
        u=pow(ll,0.3333333)*signq;
        v=pow(ww,0.3333333)*sinq;
        x1=u+v-(F/3.0);
        zg=x1;
}

if(ig<=0.001&&ig>=-0.001)
{
        k=-1*qg;
        if(k>0)signq=1;
        if(k<0)signq=-1;
        x1=2*sqrt(-1*pg)*signq-F/3.0;
        x2=-1*(x1/2.0)-(F/3.0);
        x3=x2;
        if(x1>x2)
        {
                if(x1>x3)       max=x1;
                else max=x3;
        }
        else
        {
                if(x2>x3) max=x2;
                else max=x3;
        }
        zg=max;
}


if(ig<-0.001)
```

```
{
        phi=acos(-1*qg/(sqrt(-1*pow(pg,3))));
        s=2*sqrt(-1*pg);
        x1=(s*cos(phi/3.0))-(F/3.0);
        x2=(s*cos((phi/3.0)+(2.0*3.142/3.0))-(F/3.0));
        x3=(s*cos((phi/3.0)+(4.0*3.142/3.0))-(F/3.0));
        if(x1>x2)
        {
                if(x1>x3)       max=x1;
                else max=x3;
        }
        else
        {
                if(x2>x3) max=x2;
                else max=x3;
        }
        zg=max;
}
sumx=0;
sumy=0;
i=1 ;
for( i=1;i<=13;i++)
{
sumx=sumx+x[i]*sqrt(at[i]);
sumy=sumy+y[i]*sqrt(at[i]);
}
i=1;
for( i=1;i<=13;i++)
{
anl[i]= 2*sqrt(at[i])*sumx/atl;
ang[i]= 2*sqrt(at[i])*sumy/atg;
bnl[i]=b[i]/bl;
bng[i]=b[i]/bg;
}

i=1;
for( i=1;i<=13;i++)
{
fl[i]=-log(zl-Bl)+((zl-1)*bnl[i])-al*(anl[i]-bnl[i])*log((zl+(sqrt(2)+1)*Bl)/(zl-(sqrt(2)-
1)*Bl))/(Bl*pow(2,1.5));
fg[i]=-log(zg-Bg)+((zg-1)*bng[i])-ag*(ang[i]-bng[i])*log((zg+(sqrt(2)+1)*Bg)/(zg-
(sqrt(2)-1)*Bg))/(Bg*pow(2,1.5));
phl[i]= exp(fl[i]);
phg[i]= exp(fg[i]);
Kfin[i]=phl[i]/phg[i];
}

i=1;
for( i=1;i<=13;i++)
err[i]=(Kfin[i]-kin[i])/(Kfin[i]*kin[i]);
```

```
for( i=1;i<=13;i++)
er=er+err[i];
eravg=er/13.0;


if(eravg<=0.00001&&eravg>=-0.00001)break;
for( i=1;i<=13;i++)
kin[i]=Kfin[i];
c++;
}while(c<25);

printf("\n The number of Iterations =%d",c);
 for( i=1;i<=13;i++)
 printf("\nKfin=%f     X[%d]=%f     Y[%d]=%f ",Kfin[i],i,x[i],i,y[i]);
printf("\nng=%f                    nl=%f",ng,1-ng);
q:
getch();
return 0;
}
```