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Performance of Internals in Three-Phase Tank Separators

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PERFORMANCE OF INTERNALS IN THREE-PHASE SEPARATORS

DOMINIC GEORGE AKPAN

A MASTER OF SCIENCE THESIS WORK DONE IN PARTIAL FULFILMENT FOR THE AWARD OF MASTERS DEGREE (MSC) IN PETROLEUM ENGINEERING

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Preface

This thesis is the final work of my Masters degree at the Norwegian University of Science & Technology, NTNU. The work weighs 30 points which amounts to a full semester work. It was carried out at the faculty of Engineering, and Science Technology-Department of Petroleum Engineering and Applied Geophysics during Spring 2013.

The main objective of the thesis is how internals increase the capacity and efficiency of separation in separators.

My warm appreciation and thank go to my supervisor, Professor Jon Steinar Gudmundsson for the interesting topic he gave, his availability despite the short time to guide me and the useful corrections he offered.

At the same time I thank all the Staff and lecturers at NTNU especially; Professor Pal Skalle, Tone Sane, Dr. Uduak Mme, and Rita Kuma for their kindness, concern and care.

Lastly I thank the management and Staff of University of Uyo, Nigeria especially Dr. Francis Udoh, my co-supervisor Professor Dulu Appah and all my friends and colleagues. God bless you all.

Abstract

This thesis work focused partly on the sizing of a three phase vertical and horizontal separators without internals and with internals (mesh pad) at different pressures and the performance of internals to increase the capacity and efficiency of separation. The stage separation of oil, gas and water was carried out with a series of three separators operating at consecutively reduced pressures of 80bar, 15bar and 2bar. The physical separation of these three phases occurs at three different stages (steps) and the feed fluid to the separator was a volatile oil composition as depicted in Table 1-1. The purpose of the stage separation was to obtain maximum recovery of liquid hydrocarbon from the feed fluid coming to the separators and to provide maximum stabilization of both gas and liquid effluents.

The work uses model as proposed by Monnery and Svrcek (1994) as a basic design for vertical and horizontal separators to obtain the diameters and lengths of these separators at different pressures. The diameters for the vertical separators were assumed 2m for separators without mist eliminator and 2.15m with mist eliminator. For the vertical separator without mist eliminator, the heights were obtained as 3.63m, 3.32m, and 3.16m with height/diameter ratios of 1.8, 1.66 and 1.58. For the vertical separator with mist eliminator, the heights were obtained as 3.48m, 3.50m, and 3.37m with height/diameter ratios of 1.60, 1.63 and 1.57.

The diameters for the horizontal separators were calculated as 1.42m, 1.54m, 1.48m for separators without and with mist eliminator at those pressures. For the horizontal separators without mist eliminator, the lengths were obtained as 6.0m, 6.0m, and 5.0m with height/diameter ratios of 4.2, 3.9 and 2.9. For the horizontal separator with mist eliminator, the lengths were obtained as 7.0m, 7.0m, and 6.0m with height/diameter ratios of 5, 4 and 3.

The properties of the fluid were determined using an engineering simulation software-HYSYS. The wire mesh pad was size in the horizontal separator at 80bar, the design velocity was obtained as 0.27m/s and a cross section area of 0.19m^2. The separation capture efficiency of this pad was 60% of a 0.09m thickness element of the mesh for removal of 5μm droplets size and 79% for removal of 10μm droplets. For a 0.15m thickness of the mesh element, the separation capture efficiency was 95% for the removal of 5μm droplet and 99% for the removal 10μm droplets. Foamy crude, paraffin, sand, liquid carry over and emulsion were found as operating problems affecting separation.

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Nomenclature

Symbols	Definitions and Units
A	Area of vessel, m^2
A_D	Downcomer cross-sectional area, m^2
A_G	Gas cross sectional area, m^2
A_{LL}	Cross section area of light liquid, m^2
A_{HL}	Cross section area of heavy liquid, m^2
A_L	Area of baffle, m^2
A_{LLL}	Cross section area of normal liquid level, m^2
A_{NLL}	Cross section area of light liquid above bottom of vessel, m
A_T	Total cross sectional area, m^2
C_D	Drag coefficient
D	Vessel diameter, m
D_I	Vertical vessel internal diameter, m
d_D	Droplet diameter, μm, m
D_N	Nozzle diameter, m
d_W	Wire diameter, mm, m
ε	Separation capture efficiency, %
E	Impaction efficiency fraction

F	Frictional factor
F_B	Force of buoyancy, N
F_D	Drag force, N
F_G	Force of gravity, N
F_{GA}	Fractional gas phase cross sectional area
g	Acceleration due to gravity, m/s^2
H_{BN}	Liquid height from above baffle to feed nozzle, m
H_D	Disengagement height, m
H_G	Gas space height, m
H_H	Hold up height, m
H_L	Height from liquid phase, m
H_{HL}	Height of heavy liquid, m
H_{LL}	Height of light liquid, m
H_{LLL}	Low liquid level in light liquid compartment, m
H_{NLL}	Normal liquid level, m
H_R	Height from light liquid nozzle to baffle, m
H_S	Surge height, m
H_T	Total height of vertical vessel, m

H_w	Weir height, m
k_d	Demister capacity factor
k_s	terminal settling velocity constant
L	Vessel length, m
M	Mass of particle, kg
m_G	Mass flow rate of gas, kg/s
m_{LL}	Mass flow rate of light liquid, kg/s
m_{LL}	Mass flow rate of heavy liquid, kg/s
New	Weber Number
P	Pressure, bara, kPa
Q	Volume flow rate, m^3/s, Sm^3/s
q_G	Gas (vapour) volume flow rate, m^3/s
q_{LL}	Light liquid volume flow rate, m^3/s
q_{HL}	Heavy liquid volume flow rate, m^3/s,
R_e	Reynolds Number
SO	Corrected pad specific surface area, m^2
t_S	Surge time, sec, min
T	Temperature, °C, K
t_R	Residence time, retention time, sec, min

TSV	Terminal settling velocity, m/s
U	Velocity of particle, m/s
u_c	Continuous phase velocity, m/s
u_G	Gas velocity, m/s
u_{GA}	Actual vapour velocity, m/s
u_D	Velocity of droplet, m/s
u_{LH}	Settling velocity of light liquid out of heavy liquid phase, m/s
u_{HL}	Settling velocity of heavy liquid out of light liquid phase, m/s
V	Volume, m^3
V_H	Hold up volume, m^3
V_S	Surge volume, m^3
W_D	Downcomer chord width, m
Z	Compressibility factor
σ	Interfacial tension, N/m
ρ_{HL}	Density of heavy liquid, kg/m^3
ρ_{LL}	Density of light liquid, kg/m^3
ρ_G	Density of vapour, kg/m^3
ρ_c	Density of continuous phase, kg/m^3

μ	Viscosity, Pa.s
μ_{em}	Emulsion viscosity, Pa.s
μ_G	Gas viscosity, Pa.s
μ_{LL}	Viscosity of light liquid, Pa.s
μ_{HL}	Viscosity of heavy liquid, Pa.s
$\emptyset$	Volumetric ratio of inner phase to outer phase
θ	Phase dispersion co-efficient
$\Delta\gamma$	Specific gravity difference

1 Introduction

Separators in oilfield terminology designate a pressure vessel used for separating well fluids produced from oil & gas wells into gaseous and liquid components. The goal for ideal separator selection and design is to separate the well stream into liquid-free gas and gas free-liquid. Ideally, gas and liquid reach a state of equilibrium at the existing conditions of pressure and temperature within the vessel. Separators work on the principle that these components have different densities, which allows them to stratify when moving slowly with gas on top, water on the bottom and oil in the middle. Any solids such as sand will also settle in the bottom of the separator. These separating vessels are normally used on a producing lease or platform near the wellhead, manifold, or tank battery to separate fluids produced from oil and gas wells into oil, gas and water.

Most separators are two-phase in design; separating the gas and total liquids, three-phase separators mainly needed in processing to separate gas, oil or other liquid hydrocarbons and free water. Most platforms have a series of production separators; starting with a high-pressure (HP) separators which separates the gas from the liquids. Liquids are then piped to a medium pressure (MP) separator which remove more gas and then pass the liquid to a low pressure (LP) separator that removes even more gas and then separates water from the oil (Arnold and Stewart 1999).

Separators are built in various designs such as horizontal, vertical or spherical. Vertical separators are commonly used where the gas to oil ratio is high and where horizontal space is considered a barrier. Horizontal separators are installed when volume of total fluid is available and large amount of dissolved gas in it. Proper design of separator is important in order to obtain satisfactory separation efficiency, and at the same time minimize size and weight. The processing will mainly be phase separation of oil, gas, water and perhaps sand normally performed in one to four stages where pressure is successively reduced for each stage as illustrated in Figure 1-1 below.

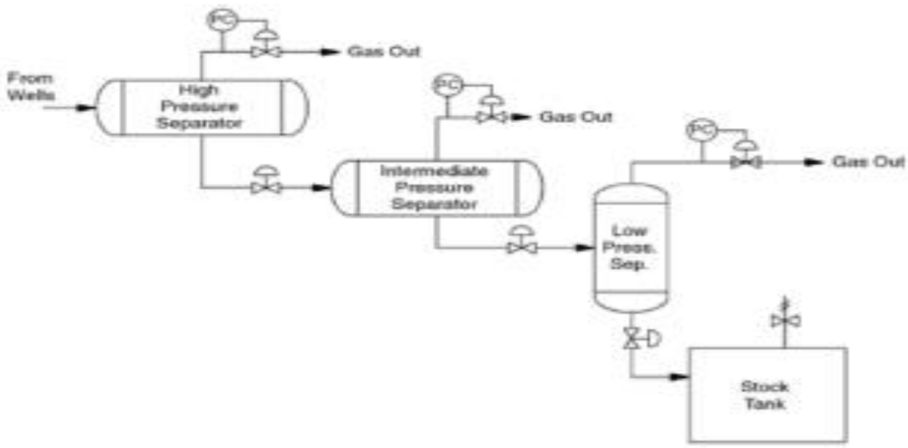


Figure 1-1: Schematic of a three-stage separation system (Ken and Maurice 2008)

This thesis is a continuation of my project work on “Droplets in Production Tubing and Separators” the focus of this work has been on the performance of internals in three-phase tanks separators and emphasis put on horizontal separators, separating oil, gas and water. The sizing of separators is divided into two parts; the first part without internals (only original internals) and the second part with the proposed internals. Internals are the various types of equipment put inside separators to increase the capacity and efficiency of separation. Internals are found in the separator inlet zone such as inlet cyclones, gravity/coalescing zones such as mesh pads, plate packs and the gas outlet zone such as the mist extractors which include wire mesh and vane type plates. The proposed internals for this work are wire mesh pads, vane type plates and cyclones but emphasis on sizing a wire mesh mist extractor at the gas outlet zone and calculating its separation capture efficiency.

In the previous work a two-phase tank separator was used separating gas and water using a produced gas well fluid (Guo & Ghalambor 2005).

In this work; Hysys is used to determine the properties of the fluid using the fluid composition of volatile oil in Table 1-1 in the appendix obtained from a typical molar composition of petroleum reservoir fluid (Pedersen et al 1989). Since a droplet size in separation is important, diameter of droplet sizes are assumed in sizing the wire mesh. A three-phase tank separator is used and the processing is phase separation separating oil, gas, water. The separation is in two stages; gas – liquid and liquid – liquid separations.

The liquid carryover in gas-liquid separation depends not only on the vessel configuration and operating conditions but also on droplet breakup and coalescence processes due to vessel internals. Normally, the separation efficiency of liquid - liquid separator is specified in terms of the "cut-off diameter". This is the diameter of the smallest droplets removed with an efficiency of 100%.

Because of the sensitivity of droplet size to the gravity settling process, a variety of vessel internals including (inlet cyclone, perforated baffles, vane packs, mesh pads, spiral flow demister) have been developed to enhance droplet coalescence and reduce liquid carryover involved in the separator.

As part of this work; separation in three stages is assumed (at 80 bara, 15 bara and 2 bara) and separators at these pressures are sized without internals and with internals based on the combination of API recommended practice and NORSOK standard which is one of the purposes of this work. Based on Sounder-Brown equation (Sounders and Brown 2008), and API SPEC 12J (API specification 12J 2008) sizes a gas-liquid separator using maximum allowable gas velocity, at which the minimum droplet can settle out of a moving gas stream. To prevent re-entrainment of liquid droplets from the interface, a simplified Kelvin-Helmholtz criterion is practically used to estimate a critical interface velocity. If the actual velocity under a given operating condition exceeds the critical interface velocity, it is then assumed that the liquid droplet will be entrained into the gas stream from the interface and eventually leads to higher liquid carryover at gas outlet.

The physical and chemical effects that make separation more difficult including foaming, emulsion and phase inversion are addressed.

2 Theory and literature review

2.1 Separators

The three phase separator works on the principle that the three components have different densities, which allows them to stratify when moving slowly with gas on top, water on the bottom and oil in the middle. Any solids such as sand will also settle in the bottom of the separator. These separating vessels are normally used on a producing lease or platform near the wellhead, manifold, or tank battery to separate fluids produced from oil and gas wells into oil, gas and water.

Separators are often classified by their geometrical configuration; vertical, horizontal, and vertical (Saeid et al 2006) and their function; two-phase and three-phase separators. Separators are two-phase if they separate gas from the total liquid stream and three-phase if they also separate the liquid stream into its crude oil and water-rich phases (Ken and Maurice 1998).

Additionally, separators can be categorized according to their operating pressure; high, medium and low. Low-pressure units handle pressure of about 0.7 to 12 bar (70 to 1200 kPa). Medium-pressure separators operate from about 15 to 48 bar (1500 to 4800 kPa). High-pressure units handle pressure of about 65 to 103 bar (6500 to 10300 kPa). In other words, they may be classified by applications (test, production, low temperature, metering, and stage separators) and by principles (gravity settling, centrifugation and coalescing).

For this work, only three-phase separators (vertical and horizontal) are used for design. The principle of gravity settling, centrifugation and coalescing are involved in separators. In the gravity settling section, gravitational forces control separation, and the efficiency of the gas-liquid separation is increases by lowering the gas velocity. Because of the large vessel size required to achieve settling, gravity separators are rarely designed to remove droplets smaller than 250 μm (Taravera, 1990). Also, residence time in the vessel is an important criterion for better separation.

In centrifugal separators, centrifugal forces act on droplet at forces several times greater than gravity as it enters a cylindrical separator. Generally, centrifugal separators are used for removing droplets greater than 100 μm in diameter, and a properly sized centrifugal separator can have a reasonable removal efficiency of droplet sizes as low as 10 μm. They are also extremely useful for gas streams with high particulate loading (Talavera, 1990).

Very small droplets such as fog or mist cannot be separated practically by gravity. However, they can be coalesced to form larger droplets that will separate out. Coalescing devices in separators force gas to follow a tortuous path. The momentum of the droplets causes them to collide with other droplets or with the coalescing device, forming larger droplets. These can then separate out of the gas phase due to the influence of gravity. Wire mesh screens, Vane elements, and Filter cartridges are typical examples of coalescing devices.

2.1.1 Horizontal separators

Horizontal separators are almost always used for high GOR wells, for foaming well streams, and for liquid-liquid separation (Beggs, 1984).

They are available for two-phase and three-phase operations. They vary in size (in diameter and in seam to seam). Figure 2-1 is a typical scheme of a three-phase horizontal separator. The fluid enters the separator and hits an inlet diverter. This sudden change in momentum generates the initial bulk separator of liquid and gas. In most designs, the inlet diverter contains a downcomer that directs the liquid flow below the oil-water interface.

This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil-water interface. This process called 'water-washing' promotes the coalescence of water droplets that are entrained in the oil continuous phase. The inlet diverter assures that little gas is carried with the liquid, and the water-wash assures that the liquid does not fall on top of the gas-oil or oil-water interface, mixing the liquid retained in the vessel and making control of the oil-water interface difficult. The liquid-collecting section of the vessel provides sufficient time so that the oil and emulsion form a layer or oil-pad at the top. The free water settles to the bottom. The produced water flows from a nozzle in the vessel located upstream of the oil weir. An interface level controller sends a signal to the water dump valve, thus allowing the correct amount of water to leave the vessel so that the oil-water interface is maintained at the design height.

The gas flows horizontally and outs through a mist extractor (normally known as a demisting device) to a pressure control valve that maintains constant vessel pressure. The level of gas-oil interface can vary from half

(50%) the diameter to 75% of the diameter depending on the relative importance of liquid-gas separation and what purpose the separator has.

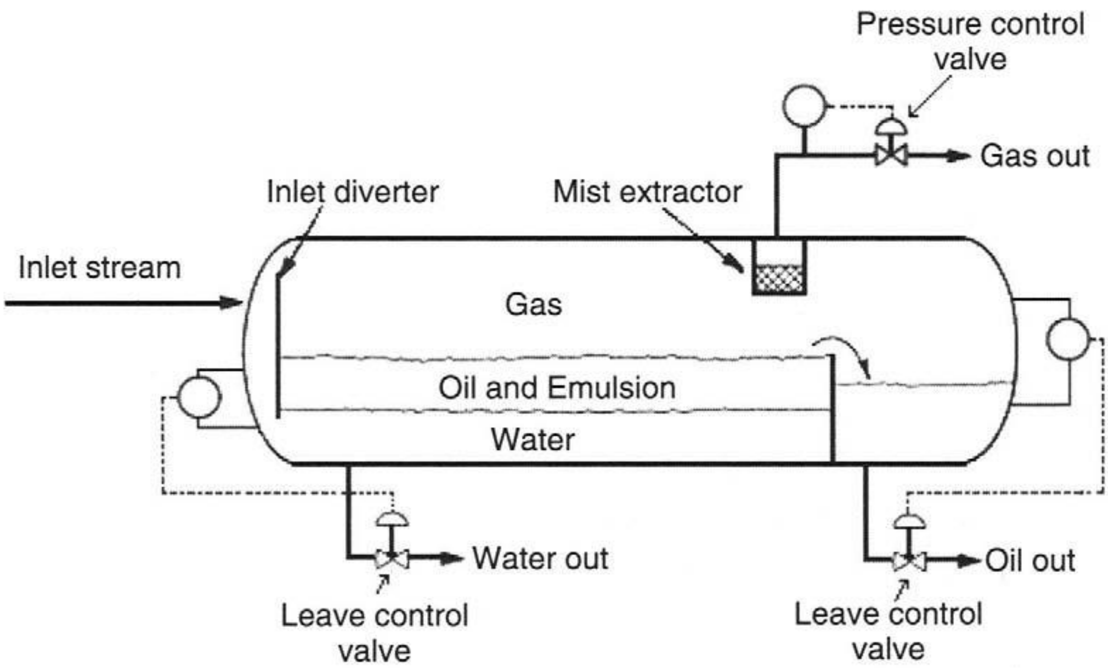


Figure 2-1: Typical scheme of a horizontal three-phase separator (Saeid et al 2006)

In addition, as illustrated in Figure 2-2 below, four major functional zones can be generally identified in the horizontal three-phase separator (ken and Maurice 2008). The section between the inlet nozzle and first baffle can be considered as the primary separation zone, which is desired to separate the bulk liquid from the gas stream. Downstream from the primary separation zone is the gravity settling zone, which is used for the entrained droplets to settle from the wet gas stream. This section normally occupies a large portion of the vessel volume through which the gas moves at a relatively low velocity. Following the gravity settling zone is a droplet coalescing zone, which could be parallel plates, vane packs, mesh pads, and spiral flow demisters. This zone helps to remove very small droplets based on impingement and inertial separation principles. In some designs, the droplets coalescing zone, and the gravity settling zone work sequentially. In other design, however, they are integrated together. The last section is the liquid collection zone may have a certain amount of surge volume over a minimum liquid level necessary for control system to function properly. However, the most common configuration is half full.

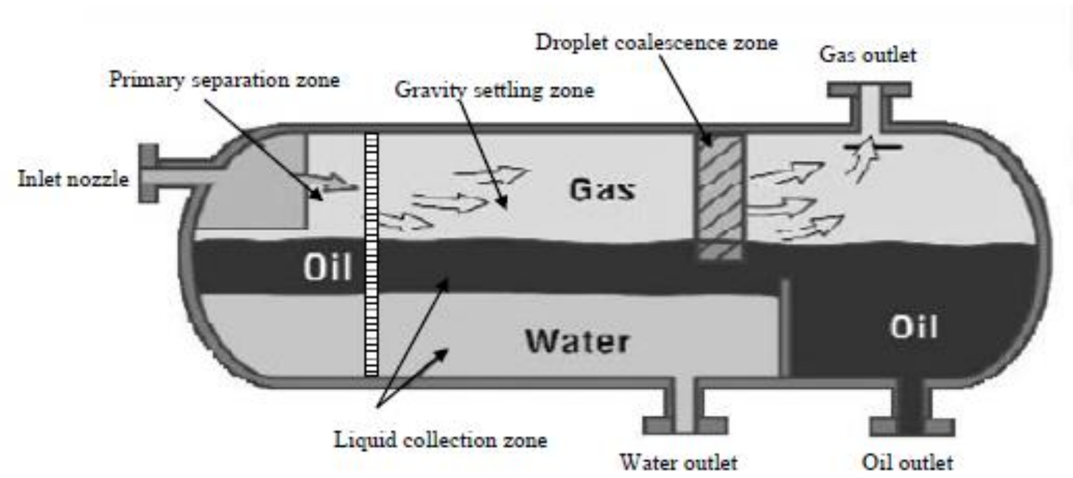


Figure 2-2: Major functioning zones in a horizontal three-phase separator (ken and Maurice 2008)

2.1.2 Vertical separators

A vertical separator can handle relatively large liquid slugs without 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 and more sands are expected.

They are available for two-phase and three-phase operations. They also vary in size (in diameter and height). Figure 2-3 is a typical scheme of a three-phase vertical separator. The flow enters the vessel through the side as in the horizontal separator and the inlet diverter separates the bulk of the gas. The gas moves upward, usually passing through a mist extractor to remove suspended mist, and then the dry gas flows out. A downcomer is required to transmit the liquid collected through the oil-gas interface so as not to disturb the oil-skimming action taking place. As illustrated by Powers et al (1990), vertical separators should be constructed such that the flow stream enters near the top and passes through a gas-liquid separating chamber even though they are not competitive alternatives unlike the horizontal separators.

A chimney is needed to equalize gas pressure between the lower section and the gas section. The spreader or downcomer outlet is located at the oil-water interface. From this point as the oil raises any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. Similarly, the water flows downward and oil droplets trapped in the water phase tend to raise countercurrent to the water flow. The horizontal separators have separation acting tangentially to flow, whereas vertical separators have separation acting parallel to flow. In the vertical separator, level control is not critical, where the liquid level can fluctuate several inches without affecting operating efficiency (GPSA, 1998). However, it can affect the pressure drop for the downcomer pipe (from the demister), therefore affecting demisting device drainage.

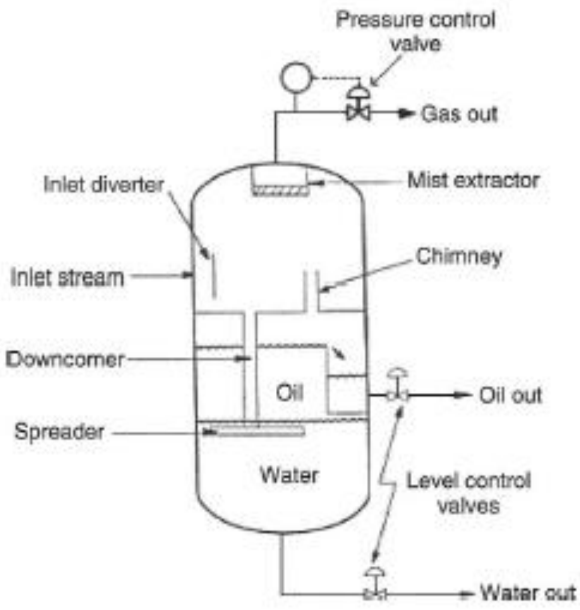


Figure 2-3: A typical scheme of a vertical three-phase separator (Saeid et al 2006)

2.1.3 Separator selection

There are no simple rules for separator selection. Sometimes, both configurations (vertical and horizontal) should be evaluated to decide which is more economical.

The relative merits and common applications of vertical and horizontal separators are summarized by Manning and Thompson (1995) as follow:

Horizontal Separators are used most commonly in the following conditions;

- Large volumes of gas and/or liquids.
- High-to-medium gas/oil (GOR) streams.
- Foaming crude
- Three-phase separation

Advantages of these separators are:

- Require smaller diameter for similar gas capacity as compared to vertical vessels.
- No counter-flow (gas flow does not oppose drainage of mist extractor).
- Large liquid surface area for foam dispersion generally reduces turbulence.
- Larger surge volume capacity

Vertical Separators are used in the following conditions;

- Small flow rates of gas and/or liquids
- Very high GOR streams or when the total volumes are low.
- Plot space is limited.
- Ease of level control is desired

Advantages of these separators are as follow:

- Have good bottom-drain and clean-out facilities.
- Can handle more sand, mud, paraffin, and wax without plugging.
- Fewer tendencies for entrainment.
- Has full diameter for gas flow at top and oil flow at bottom.
- Occupies smaller plot area.

2. 2 Vessel internals

Vessel internals are essential to enhance droplet coalescence processes in separators. Generally, gas-liquid separators without any enhancement internals can only remove liquid entrainment with sizes above 100 micron. By adding efficient internals, the corresponding droplet size can be reduced to 5-10 microns (Yaojun and John 2009). This indicates that the gas-liquid separation efficiency can be enhanced considerably by properly designed vessel internals. It is for this reason that varieties of vessel internals have been developed which include; inlet devices, perforated baffles, mesh pads, vane packs, and spiral flow demisters. The details of the various internals below are based on the SPE paper of Yaojun Lu and John Green of FMC Technologies Inc and and a research paper of Saeid Rahimi.

2.2.1 Inlet devices

A number of different inlet devices are available, with different working mechanisms. Their performances differ from each other, both in efficiency and complexity. The inlet devices have large impact on the overall separator efficiency. Inlet devices perform the following functions below;

- **Separate bulk liquids**

One of the main functions of the inlet device is to improve the primary separation of liquid from the gas. Any bulk liquids separated at the inlet device will decrease the separation load on the rest of the separator and thus improve the efficiency. Good bulk separation will also make the separator operation less sensitive to changes in the feed stream. When mist extractors (mesh or vane pads) are utilized to enhance the liquid droplet separation, the amount of liquid in gas in the face of mist extractor (liquid loading) adversely affects the performance of the mist extractor. Therefore using an appropriate inlet device plays a major role in achieving required separation.

- **Ensure good gas and liquid distribution**

A properly sized inlet device should reduce the feed stream momentum and ensure the distribution of the gas and liquid(s) phases entering the vessel separation compartment, in order to optimize the separation efficiency. Mal-distribution of liquid can lead to a large spread in residence times, decreasing the separation efficiency. Also a gas mal-distribution at the entrance of the mist extractor or cyclone deck can locally overload the demister and cause severe carryover.

- **Prevent re-entrainment and shattering**

Re-entrainment of liquid droplets can be caused by blowing gas down or across the liquid surface at very high velocities. This phenomenon often occurs when vessels with deflector baffles or half pipes are operated at the higher gas flow rates than what they were designed for. Liquid shattering inside the inlet device can also happen in a vessel with no inlet device or with a deflector baffle when the feed stream's liquid smashes into the plate and is broken up in extremely small droplets. This can create smaller droplets than were present in the feed stream, making the separation in the rest of the separator even harder. Selecting a proper inlet device and following common design guidelines for setting the distance between the bottom of the inlet device and highest liquid level inside the vessel should minimize this problem.

- **Facilitate de-foaming**

If the feed stream has a tendency to foam, an inlet device that prevents or even breaks down foam can significantly improve the separation efficiency of the vessel, reduce the size of the vessel and the use of chemicals.

Common types of inlet devices include:

- Diverter plate
- Half pipe
- Inlet Vane distributor
- Inlet Cyclone
- Slotted tee distributor
- Tangential inlet with annular ring
- Deflector baffle

2.2.1.1 Diverter plate

A diverter or baffle plate can be a flat plate, dish, cone that induce a rapid change in flow direction and velocity, causing separation of the two phases. Because the higher-density liquid possesses more energy than the gas at the same velocity it does not change direction as rapidly. The gas will flow around the diverter while the liquid strikes the diverter and falls down in the liquid section of the vessel. The design of such devices is relatively simple, it mainly needs to withstand the forces acting on it, but the geometry can vary according to fluid conditions. It can be used for flows with little gas load and little tendency for foaming. Figure 2-4 shows two examples of diverter plates, horizontal separator (left) and vertical separator (right). In addition to relatively poor bulk separation, problems of liquids droplets becoming shattered may occur. This creates small droplets which are more difficult to separate.

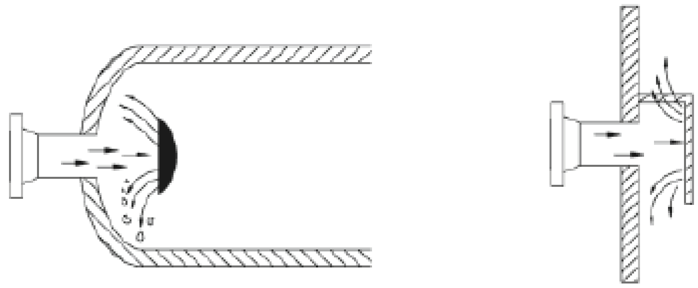
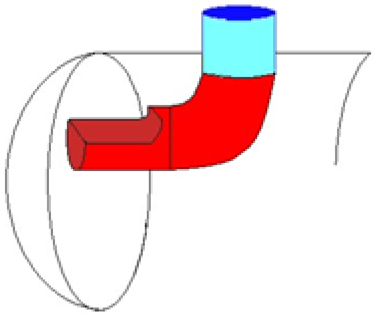


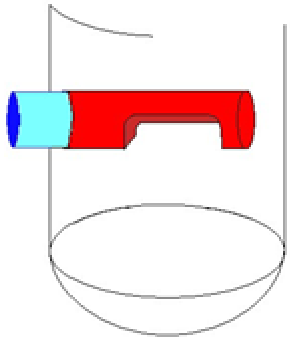
Figure 2-4: Showing examples of diverter plates (ken and Maurice 2008)

2.2.1.2 Half pipe

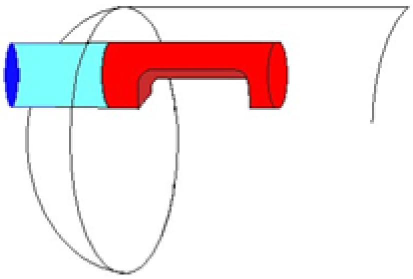
A half pipe inlet is a horizontally oriented cylinder where the bottom half is removed lengthwise. It has a simple design, but sends both gas and liquid downward into the separator and some gas may be entrained in to the liquid. Half open pipes are the modified versions of 90° elbow devices, suitable for both vertical and horizontal separators, with slightly improved bulk liquid removal and reasonable gas distribution. In this type, a piece of pipe with a length up to three times the inlet nozzle diameter is welded to the inlet 90° elbow. In horizontal vessels, the last section of the half open pipe should be horizontal; pointing opposite to the flow direction in the vessel and with its opening directed upward. In vertical vessels, the last section is closed and its opening is directed downward. The same configuration is used when the half open pipe is used for a horizontal vessel with a side nozzle.



(a) Horizontal vessel-top entry



(b) Vertical vessel

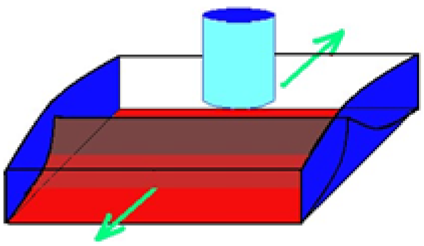


(c) Horizontal vessel-side entry

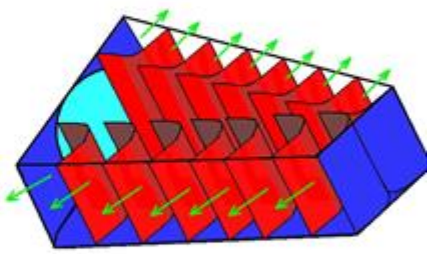
Figure 2-5: Showing half open pipes installation configuration in horizontal and vertical vessels (Saeid 2013)

2.2.1.3 Inlet vane distributor

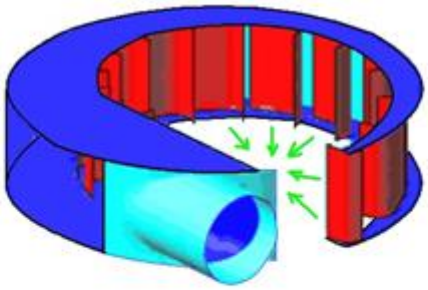
The simplest form of the inlet vane distributor is the dual vane inlet device as shown in Figure 2-6 which offers a reasonable flow distribution with low shear and pressure drop. In horizontal vessels, it is suited for top entry only. The benefits of this device compared with simpler deflectors such as deflector plates include reduced agitation and hence improved phase operational performance, more stable level control, and reduced foaming. For liquid slugging applications, usually where there is a long incoming flow line, this device provides excellent mechanical strength. The dual vane works by smoothly dividing the incoming flow into two segments using curved vanes to suit the overall geometry of the inlet nozzle. The gas phase readily separates and disperses along the vessel, whilst the liquid phase velocity is reduced and the flow directed to the vessel walls where it further disperses and falls into the bulk liquid layer at relatively low velocity. For services where there is a high gas flow relative to the liquid flow, the multi-vane inlet device provides excellent vapour distribution allowing a reduced height to the mass transfer or mist eliminator internals. The inlet vane distributors work by smoothly dividing the incoming flow into various segments using an array of curved vanes to suit the overall geometry of the inlet nozzle and distributor length. To achieve this effect the vanes start with a wide spacing and gradually reduce the gap, giving the unit its characteristic tapering shape. It can be installed in both vertical and horizontal (top and side entry) three phase separators.



(a) Dual vane



(b) Multi vane



(c) Multi vane (for vertical vessel only)

Figure 2-6: Showing the different types of inlet vane devices (Saeid 2013)

2.2.1.4 Inlet cyclone

The inlet cyclonic device is used in horizontal and some vertical separators where there is a requirement for high momentum dissipation, foam reduction and high capacity. They work on the principle of enhanced gravity separation by accelerating any incoming stream to a high gravity force, which particularly helps foam to break down into separate liquid and gas phases. Unlike most inlet devices that are positioned in the gas phase, the inlet cyclone is partly submerged in the liquid phase. The liquid phases are also separated centrifugally through the perimeter of the cyclone tubes and fall down in to the bulk liquid layers, whilst the gas form s a central vortex core and escapes through a top outlet hole into the gas space. The mixing elements on top of the cyclone outlet section usually provide a proper distribution of the cleaned gas to downstream devices. The device has a high pressure drop associated with it. The designs of the inlet cyclones have evolved over the past decades from short single (conventional cyclones) or dual cyclones into multi-cyclone arrangements as shown in Figure 2-7. The main characteristic of the cyclone inlet device is its high flow capacity, meaning that more throughput is possible through any given size separator.

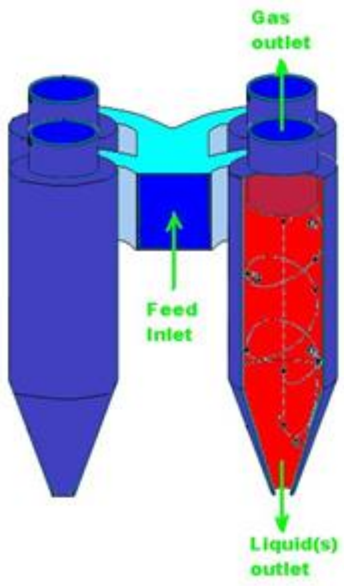


Figure 2-7: Showing multi cyclone inlet device (Saeid 2013)

2.2.1.5 Slotted tee distributor

The slotted T-shaped distributor consists of a vertical pipe extended inside the vessel to bring the distributor to the right elevation and a slotted pipe with large holes or rectangular slots (perpendicular to the inlet pipe) ensuring a reduced feed stream velocity and minimized flow turbulence. As shown in Figure 2-8, it can be used in both vertical and horizontal (top entry only) separators. The openings of the slots are usually 120° (±60°) and towards the dish end and liquid interface in horizontal and vertical vessels, respectively.

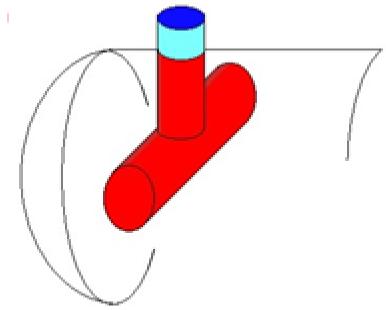


Figure 2-8: Showing tee distributor (Saeid 2013)

2.2.1.6 Tangential inlet with annular ring

Tangential inlet devices have been exclusively developed for vertical vessels. The feed flow radially enters the vessel and accelerates passing through the inlet device, the cyclonic action of the inlet device helps the liquid droplets flow on the inner wall of the vessel and the stripped gas to flow through the central section of the inlet device (annular ring) to the gas outlet nozzle. The two options with regards to the inlet nozzle arrangements are shown in Figure 2-9. The round entry type generates higher centrifugal force and slightly better separation efficiency. However, it is not recommended for pressures higher than 5.0 bar (500kPa) due to its construction difficulties at high pressures. Furthermore, both types can have a circular or rectangular inlet nozzle. A larger cross sectional area can be provided when a rectangular (with height larger than the width) nozzle is used.

(a) Round entry (b) Straight entry

Figure 2-9: Tangential inlet entry arrangement (Saeid 2013)

2.2.1.7 Deflector baffle

Deflector baffles are historically one of the most common types of inlet devices in oil and gas industries before the advent of inlet devices with higher separation efficiency become so popular. This device simply uses a baffle plate in front of the inlet nozzle to change the direction of the inlet stream and separate the bulk of the liquid from the gas. Figure 2-10 shows an horizontal baffle designed so that settled fluid flows to the inlet end of the drum and down of the drum wall to the bottom.

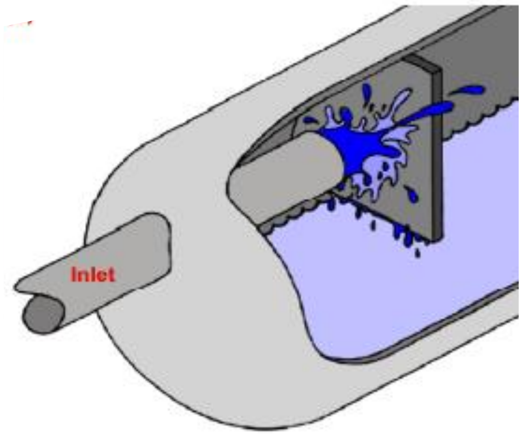


Figure 2-10: Showing a deflector baffle (Alvin & Ronaldo 2009)

2.2.1.8 Comparison of performances of different inlet devices

In order to make a proper selection of the inlet device to use, one needs to know how different types of inlet devices perform in similar conditions. When designing the inlet device, conservative liquid carry-over from upstream equipment (e.g. separators) shall be assumed. A typical number for an upstream separator, with moderate carry over, is 0.15 m^3 liquid/MSm^3 gas. Different inlet devices exist. Operating outside their design point will have detrimental effect to the overall performance. A poor inlet separation will cause liquid overloading of the demisting section and result in carry-over. A good inlet device shall reduce the inlet momentum, separate bulk liquid with minimum creation or shattering of droplets, and create good vapour distribution. Table 2-1 below evaluates the different functions of the inlet devices.

Table 2-1: Comparison of performances of different inlet devices (NORSOK standard, P-100, Nov 2001)

Inlet device functions	**None**	**Inlet vane**	**Cyclone**	**Half pipe**	**Baffle**
Momentum reduction	Poor	Good	Good	Good	Good
Bulk separation	Good	Good	Good	Average	Poor

Prevent re-entrainment	Good	Good	Average	Average	Average
Prevent liquid shatter	Good	Good	Good	Average	Poor
Ensure good gas distribution	Poor	Good	Average/ poor	Poor	Poor

Table 2-1 shows a comparison between the different inlet devices. Inlet cyclone and vane type are ranked to have the best performances. It would always be necessary to install an inlet device such as vane type distributors or cyclones. Based upon the table above, the inlet vane arrangement is usually used. Inlet cyclones may provide high inlet separation, but the design is critical and the operating envelope is more limited than for the inlet vane. However, it should also be noted that the weakness of the inlet device can be compensated if proper engineering practices are taken into consideration.

2.2.2 Mist extractors

Mist extractors otherwise called demisters are a commonly used internal devices to eliminate mist (very small disperse droplets) from gaseous streams. They are used in oil and gas industry as internal devices to gravitational separators in primary oil processing unit, in order to minimize carryover by affluent gas stream. The gas drag force causes small liquid particles to follow the gas stream. Mist extractors must therefore somehow intervene the natural balance between gravitational and the drag forces. This can be accomplished by reducing the gas velocity (hence reduce drag), introduce additional forces by use of cyclones or increase gravitational forces by boosting the droplet size (impingement). The selection of mist extractor is based on evaluation of:

- Droplet sizes that must be removed.
- Tolerated pressure drop.
- Presence of solids and the probability or risk of plugging because of this.
- Liquid handling in the separator.

The rate of droplets following the gas stream is governed by simple laws of fluid mechanics. As gas flows upward, two opposing forces are acting on a liquid droplet namely a gravitational force (accelerates the droplet down) and a drag force (slows down the droplet's rate of fall). An increase in gas velocity will increase the drag and when the drag force equals the gravitational force the droplet will settle at a constant velocity called the terminal velocity. Further increase in the gas velocity causes the droplet to move upwards and then follow the gas stream out of the separator.

Mist extractors' operations are usually based on a design velocity and depend on the demister type and the manufacturing company. The designed velocity is given by;

$$u_d = k_d \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \qquad (2\text{-}1)$$

Where

u_d = design maximum velocity, m/s

ρ_L = the density of liquid, kg/m^3

ρ_G = the density of gas (vapour), kg/m^3

k_d = demister capacity factor, m/s and depends upon the demister type.

It is in other words the k_d-factor that determines the operating gas velocity, where a too low factor can cause the droplets to remain in the gas streamlines and pass through the device uncollected while a too high value can cause re-entrainment because of droplet breakup.

Some functions of mist extractors include

- Collect/capture drops
- Remove drops
- Avoid maintenance problems
- Keep cost as low as possible

In order for a mist eliminator to work properly, it must be designed to collect and capture the droplets present in the system. Therefore, it is imperative to define the size of the droplets present. After the droplets have been captured, the mist eliminator must be able to remove the droplets from the system by draining effectively. Through the initial selection process, the most appropriate mist eliminator media must be selected so that liquid hold-up in the pad does not become an issue after the mist eliminator is installed. When designing a mist eliminator, engineers are also responsible for defining the environment the mist eliminator will operate in so that maintenance requirements are minimized. The mist eliminator must keep the operating costs of the system as low as possible and match the budget of the end user.

Common types of mist extractors are:

- Wire mesh pads
- Vane packs
- Demisting cyclones

2.2.2.1 Wire mesh pad

The mesh pad demister captures small droplets with high efficiency. The mesh can be of metal (wire mesh) or plastic material, or a combination. Typical minimum droplet removal size is:

Metal mesh: 10 micron

Plastic/fibre: 3-5 micron

The mesh depth is typically from 100-300 mm with typical pressure drop of 0.1-3.0 millibar. High liquid load/flooding will increase the pressure drop significantly. For efficient operation the demister k_d value must generally be below 0.1 m/s.

The most common impingement type mist extractor are the wire mesh type, as shown in Figure 2-11, where a large surface area is obtained by knitting wire together to a pad. The mesh pad is mounted close to the gas outlet of the separator. As the gas flows through, the inertia of the entrained droplets make them contact the wire surfaces and coalesce. Because of the dense structure of the pad it is best suited for low viscosity, non-congealing liquids with no solids present. Otherwise it may get clogged. At k_d -values above 0.1m/s the mesh will be flooded. This causes loss of separation efficiency, and the element will then act as an agglomerator, coalescing small droplets into larger. In this service the mesh can act as a conditioner for a secondary demisting element such as vanes or cyclones, as larger droplets will separate more easily in cyclones or vanes.

At elevated pressures and in critical services, the k_d -factor value must be multiply by the following adjustment factor.

Table 2-2: Scrubber conditions and adjustment factors (GPSA engineering data book, 1998, vol. 2)

Scrubber condition	Adjustment factor
1 bar pressure	1.00
20 bar pressure	0.90
40 bar pressure	0.80
80 bar pressure	0.75

Also as shown in Figure 2-12, there are one-layered and multi layered mesh pads and are usually constructed from wire of diameter ranging from 0.1 to 0.28mm, typical void fraction from 0.95 to 0.99, and thickness from 100 to 300mm. During operation, gas stream carrying entrained droplet passes through the mesh pad. The gas moves freely from the mesh pad while the entrained droplets are forced to the wire surface and coalesced due to the inertial effect. Drops formed in the mesh pad ultimately drain and drop out of the mesh pad. It is evident that a well designed and operated mesh pad can effectively removed droplets larger than 3-5 microns from the gas stream, and corresponding pressure drops is less than 0.25kPa. In addition, mesh pad can be operated between 30-110% of the design capacity, thus exhibits excellent turndown behaviour.

Figure 2-11: Wire mesh extractor for vertical separator (NATCO, 2009)

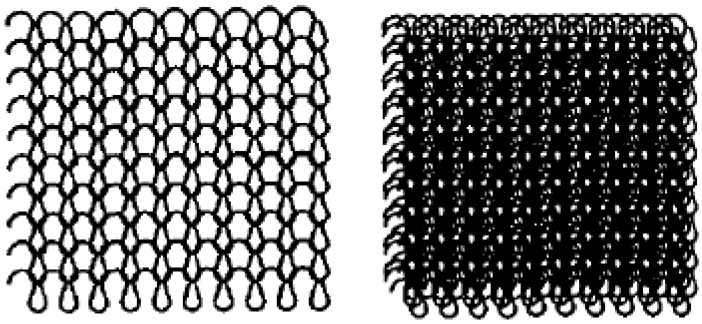


Figure 2-12: Typical configuration of mesh pads

2.2.2.2 Vane packs

Vane mist eliminators also known as plate types consist of closely spaced corrugated plates that force mist-laden gas to follow serpentine paths. These devices are generally not efficient for mist droplets smaller than about 20 microns, but they are sturdier than mesh pads and impose less pressure drop. Vane arrays can be mounted horizontally or vertically. They are preferred in applications involving high vapour velocities, low available pressure drop, viscous or foaming liquids, lodging or caking of solids, slugs of liquid or violent upsets. Like mesh pads, vane units are usually round or rectangular. The operating principle for a vane pack is that the feed stream passes through parallel vane plates and is forced to change direction several times. The droplets impinge and collect at the surface of the plates and create a liquid film which is drained through slits into a liquid sump and then further to the liquid compartment of the vessel. Figure 2-13 (a & b) show a vane pack design by Koch-Otto York for a horizontal and vertical gas flows. Here the collected liquid (green arrow) is guided into separate channels which move the liquid away from the gas. Because the liquid is isolated from the gas the chance for re-entrainment of liquid into the gas again is reduced. This is called a double pocket design. Simpler single pocket designs are also common, but here is the liquid drained with the gas flowing by, increasing the chance of re-entrainment of liquid. Hence, gas velocities can be much higher for double pockets (Koch-Glitsch 2007).

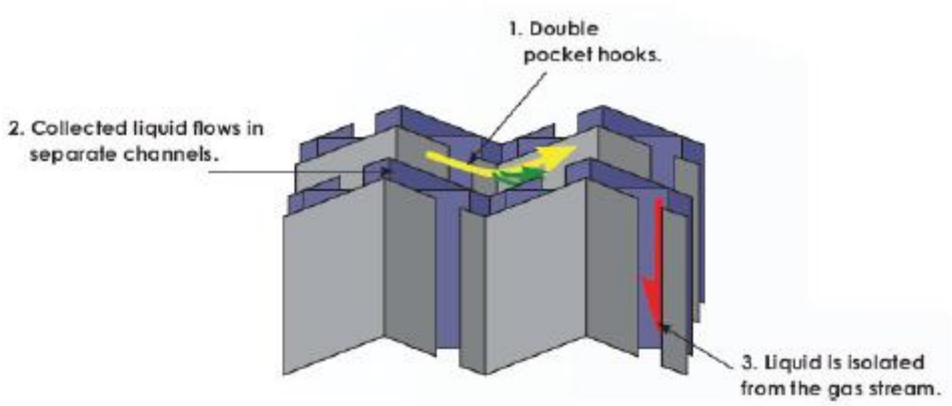


Figure 2-13: (a) Horizontal gas flow in a vane pack (Koch-Glitsch 2007)

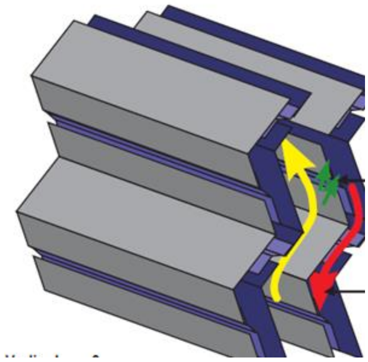


Figure 2-13: (b) Vertical gas flow in a vane pack (Koch-Glitsch 2007)

Also as shown in Figure 2-14, it illustrates the vane pack with strategically designed slots or pocket hooks, which allows the coalesced liquid on the blade surface be collected and directed into the internal channels shielded from the gas flow. Once droplets gets into these channels, the collected liquid is directed to drain and lead to a liquid sump in the separator vessel. Since the liquid is isolated from the gas stream and less subject to re-entrainment, the gas velocities can go significantly high both in horizontal and vertical applications. The space between the two adjacent blades ranges from 5 to 75mm with a total depth in the flow direction of 150 to 300mm. By passing the wet gas through the vane pack, the mist droplet undergoes changes in momentum, causing impingement and coalescence on the vane blades. The coalescence droplets then drain down along the vane surfaces. It is reported that the convectional vane pack can separate droplets larger than 40 micron, while the vane pack with pocket design can remove droplets down to less than 15 micron (Yaojun and John2009). In most cases, the drop through the vane pack ranges from 0.1 to 1.5 kPa. Double pocket vanes have higher efficiency and larger capacity as compared to single pocket vanes. Many different vane designs exist and a general performance factor can not be given, but table 2-3 below gives some conservative values.

Table 2-3: Sizing factors for demisting vane elements (NORSOK standard, P-100, Nov 2001)

	Vertical gas flow	Horizontal gas flow
ρv_s^2 (kg/ms^2)	20 – 30	30 – 45
K-value (m/s)	0.12 – 0.15	0.20 – 0.25

Where v_s = superficial velocity, m/s

Vanes give lower pressure drop than demisting cyclones and this may be an advantage at lower pressures such as in re-compressor scrubber applications (1 to 20 bar). At higher pressures, careful design is required to limit re-entrainment problems, but vanes have been used at pressures above 100 bar.

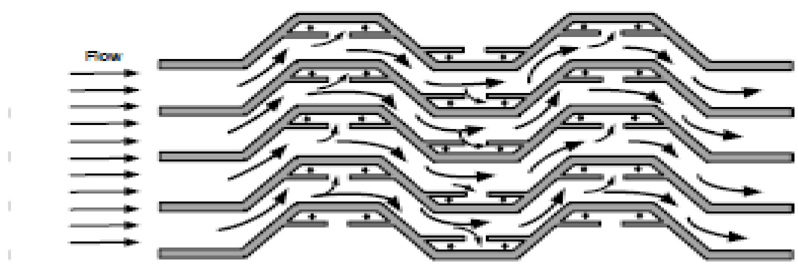


Figure 2-14: Showing vane pack with pockets (Yaojun 2009)

2.2.2.3 Demisting cyclones

Demisting cyclones are sometime axial type cyclones, but some other designs are also based on multi-cyclones with tangential entry. For axial cyclones, typical minimum droplet removal size is 5-10 microns depending on swirl velocity. The typical pressure drop is 20-100 bar. The relatively high pressure drop requires a high drainage head, and is one of the critical parameters in cyclone design. The drainage is normally internal, into the vessel bottom, but should be routed externally in case of insufficient drainage height. The total differential pressure over the demisting section, measured in liquid height, shall not be more than 50% of the available drainage height related to LAHH (Level alarm high high). A cyclone based scrubber should usually have a mesh upstream the cyclones. The mesh will act as a demister at low gas rates, and as an agglomerator at high gas rates. The performance curves of the mesh and cyclones shall overlap to assure good demisting in the whole operating range. The liquid handling capacity of the cyclones may limit the gas capacity and performance. Sufficient separation in the inlet-mesh section is required to stay below the liquid capacity limits of the cyclones.

In the cyclonic demisting device multiple cyclone tubes are mounted on a deck or into housing. Cyclone demisters can handle high gas capacities combined with efficient droplet removal, and are more efficient than mist extractors and vanes and less susceptible to clogging. Figure 2-15 shows a principle sketch of a cyclone mist extractor. Gas and mist enters the cyclone and goes through a swirl element. This induces high centrifugal forces causing the liquid droplets to move outwards and coalesce to a liquid film on the cylinder wall. The liquid is purged through slits in the wall together with some gas into a chamber where the phases are separated. The purge gas, with some remaining mist is led to a low pressure zone of the cyclone where the remaining entrainment is removed. The main gas flow is discharged at the top of the cyclone while the liquid is drained at the bottom.

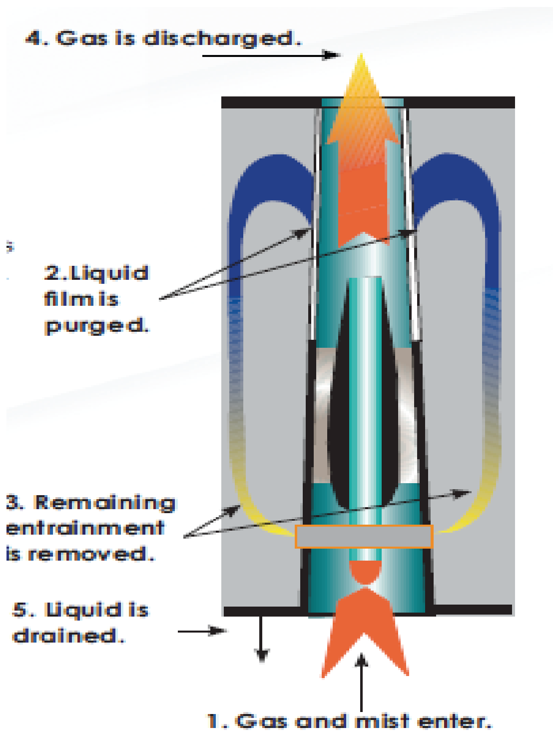


Figure 2-15: Cyclone mist extractor (Koch Glitsch 2007)

2.2.2.4 Perforated baffles

The perforated baffle can be just a single plate with uniformly distributed holes as shown in Figure 2-16. Some advance baffles are constructed with steel as double plate with varied hole-size and pattern or combination of the full and partial baffles property spaced. As the gas stream approaches the baffle surface, flow is force to change direction and spread along the baffle surface. Due to the presence of perforated baffle, additional pressure drop is created, kinetic energy of the gas stream is dissipating, and flow across the baffle is re-distributed accordingly. Since the gravity setting process is closely related to the flow distribution, the perforated baffles are commonly used to manage the flow condition and further to control re-entrainment of droplet from the gas-liquid interface. In conjunction with the inlet devices, the perforated baffle are frequently utilized to establish a primary separation section, where the momentum is reduced prior to entering the gravity setting zone where the condition are optimized for setting separation.

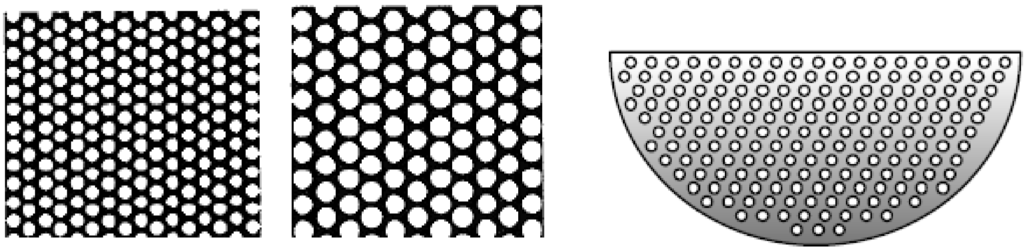


Figure 2-16: Typical configuration of perforated baffles (Yaojun 2009)

Common applications include; calming the inlet zone in horizontal

Separators, liquid flow redistribution in long vessels, surge suppression in vessels, gas distribution upstream or downstream of mist eliminators.

The flow distribution characteristic of perforated baffles is well established, but modern design tools such as CFD enable today's designers to tailor the baffle design to achieve optimum distribution by adjusting the hole size, percentage of the open area, number of baffle and their overlap as illustrated below;

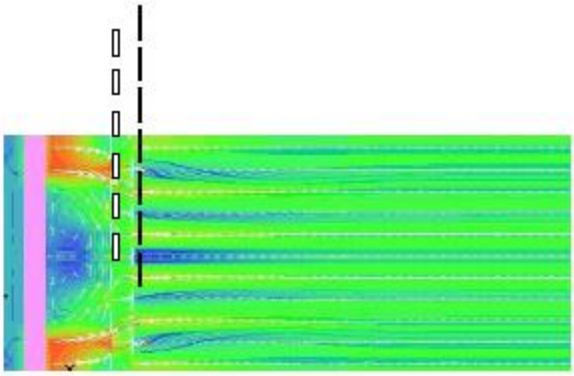


Figure 2-17: Showing CFD enabled baffle design (HAT International)

2.2.2.5 Spiral flow demister

Spiral flow demisters consist of multiple cyclone tubes mounted into housing. Figure 2-18 shows only one of the cyclone tubes. As can be seen, the gas stream enters the cyclonic inlet and flows through the spiral flow element that imparts a high centrifugal force. The droplets are then flung outward and are coalesce into a liquid film on the inner wall of the wall of the cyclone tube. The liquid film is purge out of the cyclonic unit through slit in the wall, along with a small portion of gas flow, into an outer chamber where most of the gas and liquid separate. The gas along with some remaining mist is educed back into a low pressure zone of the cyclone unit and the remaining entrainment is removed. The demisted gas is then discharge from the top and separation liquid is drained from the bottom. Depending on nominal diameter of the spiral element, droplets of 25 micron and above can be effectively separated, and the corresponding pressure drop range from 2.5 to 7.5kPa. An important advantage of the spiral demister is the high gas handling capacity combined with excellent droplets removal efficiency even at elevated pressure. Its downside, however, is sensitivity to flow change. Therefore, the spiral flow demister is more suitable for application where the flow fluctuation is not very significant.

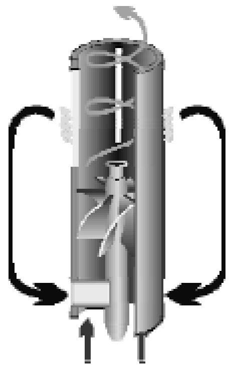


Figure 2-18: Showing spiral-flow demister (Yaojun 2009)

3 Determination of fluid properties using Hysys

Hysys, an engineering simulation software developed by Aspen Technologies Incorporated was used in this work to determine the properties of the mixed fluid (volatile oil and water) using the fluid composition of volatile oil in Table 1-1 in the appendix obtained from a typical molar composition of petroleum reservoir fluid (Pedersen 1989). Hysys was started by first click on the start menu and select it among the various programs in the computer. It already had a database of about 1500 components making it easy to select the specific needed components with reference to Table 1-1. After specifying the components list, the composition of the fluid was directly put into the program and normalized to give a total mole fraction of one. Peng-Robinson equation of state was selected as the fluid package. The fluid package contains information about the physical and flash properties of components. It determines the relation between each component and how they react together. The Peng-Robinson fluid package is the preferred fluid package for hydrocarbon mixtures. It is recommended for oil, gas, water and petrochemicals because it calculates with a high degree of accuracy the properties of two-phase and three-phase systems. The simulation environment was entered and material streams added.

The volatile oil was assumed to be the fluid at the bottom of the well was mixed with water using a mixer and the properties of the mixed fluids were obtained. The volatile oil was assumed to have a temperature of 90°C (363 K), a pressure of 300 bara (30000 kPa) and a mass flow of 36,000 kg/hr (10 kg/s). The water was set at the same temperature and pressure of the volatile oil with a mass flow of 1300 kg/hr (0.36 kg/s).

The mixed fluid (volatile oil and water) was considered as the fluid from the reservoir which flows through a well of 2500 m with five different pipe segments of 500 m each representing the well to the surface through the tubing. The pipes had an outer diameter of 4.50" (0.114 m) and an inner diameter of 4.02" (0.102 m). The flow was distributed and assumed to be a mist flow throughout the different segments of the pipe.

The fluid at the surface passed through the controlled valve-1 to the first-stage separator. Three separation units were considered. The drop in pressure caused flash vapourisation. Crude oil from the first stage flows to the second stage and then to the third stage. There is a pressure-reducing valve at the input of each separator vessel and the pressures used were 80, 15, and 2 bar for first, second and third stages separation respectively. Tables 3-1, 3-2 and 3-3, are the fluid properties in first-stage (high pressure), second-stage (medium pressure) and third-stage (low pressure) respectively. For convenient and ease calculations, some of the fluid properties are summarised in Table 3-4 (Appendix A) for gas, oil, and water in both field and SI units.

3.1 Stage and phase separation

Stage separation of oil and gas is carried out by a series of separators whose pressures gradually decrease. The fluid is discharged from a high pressure separator to the next low pressure separator. The purpose of stage separation is to get the highest amount of hydrocarbon liquid from the well fluid, and provide the highest stability of the streams of liquids and gas.

To achieve good separation between gas and liquid phases and maximizing hydrocarbon liquid recovery, it is necessary to use several separation stages at decreasing pressures in which the well stream is passed through two or more separators arranged in series. The operating pressures are sequentially reduced, hence the highest pressure is found at the first separator and the lowest pressure at the final separator. In practice, the number of stages normally ranges between two and four, which depends on the gas-oil ratio (GOR) and the well stream pressure, where two-stage separation is usually used for low GOR and low well stream pressure, three-stage separation is used for medium to high GOR and intermediate inlet pressure, and four-stage separation is used for high GOR and a high pressure well stream. The main objective of stage separation is to provide maximum stabilization to the resultant phases (gas and liquid) leaving the final separator, which means that the considerable amount of gas or liquid will not evolve from the final liquid and gas phases, respectively. The quantities of gas and liquid recovered at a given pressure are determined by equilibrium flash calculation using an appropriate equation of state (EOS). This helps optimize the value of pressure that is set for each separator. The pressures are often staged so that the ratio of the pressure in each stage is constant. Therefore, if the pressure in the first separator (which is normally fixed by specification or economics) and the pressure in the final separator (which will be near the atmospheric pressure) are known, the pressure in each stage can be determined. In this work, pressures at the three different stages are given as 80, 15 and 2 bar. The different stages of separation are completed using the principles of; momentum, gravity settling, and coalescing. Momentum force is utilized by changing the direction of flow and is usually employed for bulk separation of the fluid phases. The gravitational force is utilized by reducing velocity so the liquid droplets can settle out in the space provided. Gravity settling is the main force that accomplishes the separation, which means the heaviest fluid settles to the bottom and the lightest fluid rises to the top. However, very small droplets such as mist cannot be separated practically by gravity. These droplets coalesced to form larger droplets and settle by gravity.

3.1.1 Gas-liquid separation

Gas liquid separation is often based on the principle of gravity settling, when liquid droplets suspended in rising gas vapours settle down at the bottom of the separation vessel and are eventually taken out through the bottom. Gas stream separated from liquid is taken out from the top of the separation vessel.

Gas-liquid separation is usually accomplished in three stages. The first stage; primary separation, uses an inlet diverter to cause the largest droplets to impinge by momentum and then drop by gravity. The next stage; secondary separation, is gravity separation of smaller droplets as the vapour flows through the disengagement area. Gravity separation can be aided by utilizing distribution baffles that create an even velocity distribution in the fluid, thus allowing enhanced separation. The final stage; is mist elimination, where the smallest droplets are coalesced on an impingement device, such as a mist pad or vane pack, followed by gravity settling of the larger formed droplets. In the liquid-liquid separation, the volume must be sufficiently large to allow sufficient time for the dispersed-phase drops to reach the liquid-liquid interface and coalescence. Thus, the residence time has two components. These are the time required for the droplets to reach the interface and the time required for the droplets to coalesce.

The separation of liquid droplets from vapour phase can be explained with the help of equation for terminal velocity of liquid droplets.

In the gravity settling section of a separator, liquid droplets are removed using the force of gravity. Liquid droplets will settle out of a gas phase if the gravitational force acting on the droplet is greater than the drag force of the gas flowing around the droplet. These forces can be described mathematically using the terminal or free settling velocity. Figure 3-2 (a and b) show the forces on liquid droplet in gas stream and gravity settling theory where small droplets are entrained in the gas vapour, droplet of critical sizes stay in suspension and large droplet settled.

A liquid drop in a gas stream will be carried upward if the gas velocity is higher than the terminal settling velocity (TSV). According to Gudmundsson (2000), the TSV of a liquid droplet in a gas stream is given by the equation;

$$u_D = \sqrt{\frac{4gd_D}{3\mathrm{C_D}}}\ \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \qquad (3\text{-}1)$$

Where $\sqrt{\frac{4gd_D}{3\mathrm{C_D}}} = k_s$ (3-2)

$$u_D = k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \qquad (3\text{-}3)$$

Also;

The drag coefficient is a function of the Reynolds number for liquid droplet which is expressed as;

$$R_e = \frac{\rho_G u_G d_D}{\mu_G} \qquad (3\text{-}4)$$

The gaseous mass flow rate is given as;

$$m_G = u_G \, \rho_G \, A_G = \rho_G \, A_G \, k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \qquad (3\text{-}5)$$

And the gaseous volume flow rate is given as;

$$q_G = u_G \, A_G = A_G \, k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \qquad (3\text{-}6)$$

At standard conditions, the gaseous volume flow rate is given as

$$q_{Gsc} = A_G \, k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \left(\frac{P}{P_{sc}}\right)\left(\frac{T_{sc}}{T}\right)\frac{1}{Z} \qquad (3\text{-}7)$$

$0.05 \, m/s < k_s < 0.11 \, m/s$ (API 12J SPEC for vertical separators.)

Most vertical separators are sized based on equation (3-3) which have been developed using the terminal settling velocity equation and the drag coefficient expressed as a function of Reynolds number. The drag coefficient has been found to be a function of the shape of the particle and the Reynolds number of the flowing gas. For the purpose of this equation particle shape is considered to be a solid, rigid sphere.

For smaller horizontal vessels (length less than 3m), equation (3-3) given above can be used for sizing. For horizontal vessels larger than 3m equations (3-8 & 3-9) given below have to be followed.

$$u_G = k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \left(\frac{L}{6}\right)^{0.58} \qquad \text{(For NORSOK standard)} \tag{3-8}$$

$$u_G = k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \left(\frac{L}{3.05}\right)^{0.56} \qquad \text{(For API 12J SPEC.)} \tag{3-9}$$

$$m_G = u_G \rho_G A_G = \rho_G \, A_G \, k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \left(\frac{L}{6}\right)^{0.58} \tag{3-10}$$

$$q_G = A_G \, k_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \left(\frac{L}{6}\right)^{0.58} \tag{3-11}$$

0.12 m/s < k_s < 0.15 m/s (API 12J SPEC)

k_s = 0.137 m/s (NORSOK standard)

Where;

A_G = area of the gas particle, m^2

m_G = mass flow rate of gas, kg/s

C_D = drag co-efficient

d_D = droplet diameter, m, µm

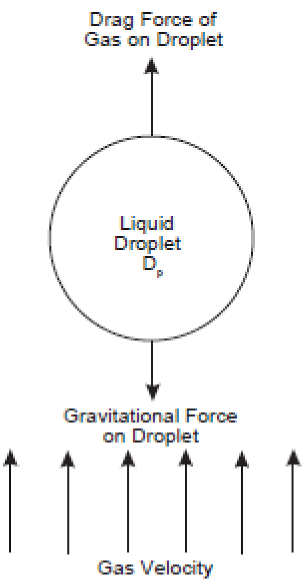
q_G = volume flow rate of gas, m^3/s

q_{Gsc} = volume flow rate of gas at standard conditions, Sm^3/s

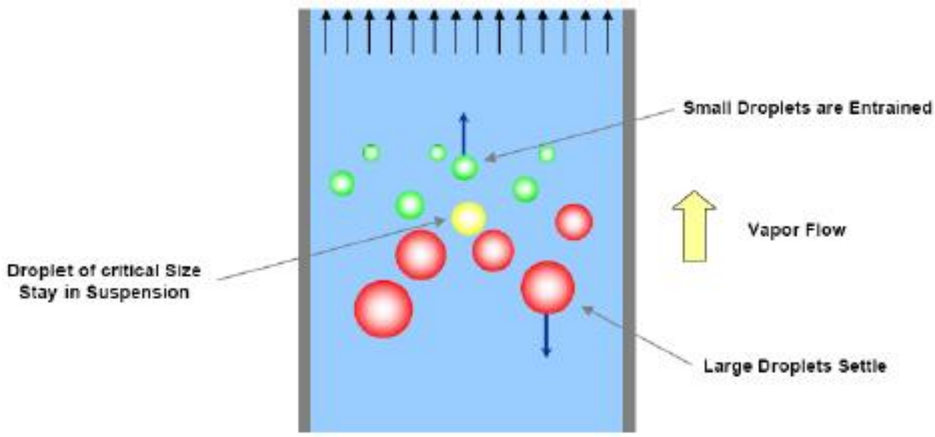
L = effective length of separator, m

R_e = Reynolds's number

μ_G = Gas viscosity, Pa.s



(a) Forces (drag and gravitation) on liquid droplet in gas stream



(b) Showing gravity settling theory

Figure 3-1: 2001 ASHRAE Meeting in Cincinnati, Nestle USA

3.1.2 Liquid-liquid (oil-water) separation

Two immiscible liquids can be separated using the difference between densities of the two phases. Similar to gas-liquid separation principles, liquid-liquid separation is also governed by settling of heavier phase droplets or rise of the droplets of lighter liquid phase. In the liquid-liquid separation, the volume must be sufficiently large to allow sufficient time for the dispersed-phase drops to reach the liquid-liquid interface and coalescence. Thus, the residence time has two components. These are the time required for the droplets to reach the interface and the time required for the droplets to coalesce. The separation of liquids in this ways is also governed by the equation of the terminal settling velocity of the droplets.

For a liquid droplet in another phase Reynolds number is expressed and for low Reynolds numbers (less than 2), the drag coefficient (C_D) has a linear relationship with Reynolds number (R_e). Then the terminal settling velocity equation can then be reduced to the Stokes law.

The difference between two liquid densities is often low and viscosities are high, resulting in low terminal velocities. Hence, for liquid-liquid separation, the residence time required for the separation is much higher than often required for gas-liquid separation. Thus for high degrees of separation, liquid-liquid separation requires a big size.

According to Stewart et al (1998), the oil-water separation is governed by Stoke's law for terminal velocity of spheres in a liquid medium. The terminal velocity of the continuous phase is defined by;

$$u_D = \frac{1}{18} g d_D^2 \frac{(\rho_L - \rho_G)}{\mu_G} \qquad (3\text{-}12)$$

Where g = gravitational constant, 9.8m/s^2

From equation 3-12 above, the terminal velocity is a function of an emulsion (oil-water) viscosity that takes into account an oil-rich or a water-rich system. The viscosity of an emulsion as given by Taylor is given by;

$$\mu_{em} = \mu_c \left[1 + 2.5\emptyset \left(\frac{p + {}^{2}/_{5}}{p+1}\right)\right] \qquad (3\text{-}13)$$

Where $p = \frac{\mu_{inner\ phase}}{\mu_{outer\ phase}}$ (3-14)

As production goes on, inversion from oil-dominant to water-dominant emulsion takes place and is estimated by;

$$\theta = \left(\frac{q_{LL}}{q_{HL}}\right) \left(\frac{\rho_{LL}\mu_{HL}}{\rho_{HL}\mu_{LL}}\right)^{0.3} \qquad (3\text{-}15)$$

Where;

μ_{em} = emulsion viscosity, Pa.s

μ_c = continuous phase viscosity, Pa.s

$\emptyset$ = volumetric ratio of inner phase to outer phase

θ = phase dispersion coefficient

q_{LL} = flow rate of light liquid (light phase), m^3/s

q_{HL} = flow rate of heavy liquid (heavy phase), m^3/s

ρ_{LL} = density of light liquid, kg/m^3

ρ_{HL} = density of heavy liquid, kg/m^3

μ_{HL} = viscosity of heavy liquid, Pa.s

μ_{LL} = viscosity of light liquid, Pa.s

Table 3-5 below summarized the types of emulsion based on the phase dispersion coefficient, θ

Table 3-5: Emulsion type resulting from phase dispersion coefficient (Boukadi et al 2012)

Phase dispersion coefficient, θ	Result
< 0.3	Light phase always dispersed

0.3 – 0.5	Light phase probably dispersed
0.5 – 2.0	Phase inversion possible
2.0 – 3.3	Heavy phase probably dispersed
> 3.3	Heavy phase always dispersed

For practical purposes, phase dispersion of 0.5 is used as an inversion point. The emulsion viscosity obtained from the above procedure can only be used to calculate the minimum capacity of the separator. There is no limit on the size of the separator as viscosity does not directly influence the capacity of a separator. For this purpose, a new retention time is used that is calculated using the formula below to yield a more direct correlation to the size.

$$t_{R(future)} = \frac{\mu_{future}}{\mu_{base}} t_{R(base)} \qquad (3\text{-}16)$$

Where t_R = retention time, sec

Figure 3-2 below illustrates the new methodology of sizing separators.

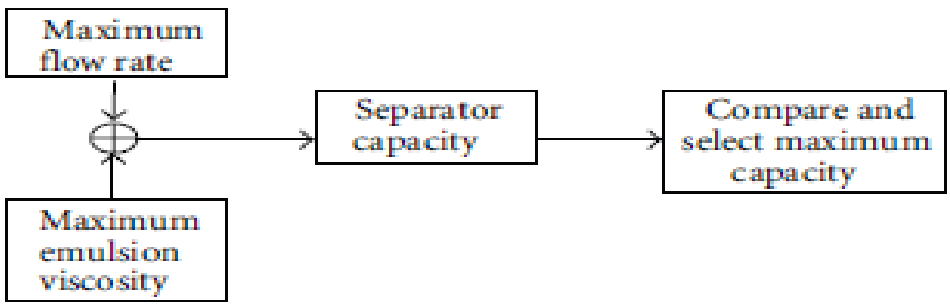


Figure 3-2: Showing separator sizing methodology (Boukadi et al 2012)

4 Separator sizing

Based on engineering design guidelines, separator sizing must satisfy several criteria for good operation during the lifetime of the producing field. These include;

- Providing sufficient time to allow the immiscible gas, oil, and water phases to separate by gravity.
- Providing sufficient time to allow for the coalescence and breaking of emulsion droplets at the oil–water interface.
- Providing sufficient volume in the gas space to accommodate rises in the liquid level that result from the surge in the liquid flow rate.
- Providing for the removal of solids that settle to the bottom of the separator.
- Allowing for variation in the flow rates of gas, oil, and water into the separator without adversely affecting separation efficiency.

Gas-liquids separators may be sized for horizontal or vertical operation, but Younger (1955) found that for seven separators in use, with L/D varying from 1.7 to 3.6, all were installed vertically. This is consistent with the rule given by Branan (1994) that if L/D > 5, a horizontal separator should be used. Scheiman (1963) recommends that the settling length should be to 0.75D or a minimum of 12in (0.305 m) whereas Gerunda (1981) specifies a length equal to the diameter or a minimum of 3ft (0.914 m). Also, to prevent flooding the inlet nozzle, Scheiman (1963) allows a minimum of 6in (0.152 m) from the bottom of the nozzle to the liquid surface or a minimum of 12in (0.305 m) from the center line of the nozzle to the liquid surface. Branan (1994) recommends using 12in (0.305 m) plus half of the inlet nozzle outside diameter or 18in (0.4570 m) minimum. Gerunda (1981) specifies a length equal to 0.5 D or 2 ft (0.610 m) minimum.

Scheiman (1963) recommends a surge time in the range of 2 to 5 min, whereas Younger (1955) recommends 3 to 5 min. There is a minimum liquid height required to prevent a vortex from forming. The design of the separator will have to include a vortex breaker. The minimum liquid level should cover the vortex breaker plus an additional liquid height. Experiments conducted by Patterson (1969) showed that the lower liquid level varies slightly with the liquid velocity in the outlet nozzle. For a velocity of 7ft/s (2.13 m/s) in the outlet piping of a tank, with no vortex breaker, a vortex forms at a liquid level of about 5in (0.127 m). The flow should be turbulent to break up any vortex. Thus, Gerunda's (1981) recommendation, allowing a 2ft (0.610 m) minimum liquid level, should suffice. The thickness of the mist eliminator must be specified, which must be thick enough to trap most of the liquid droplets rising with the vapour. The thickness of the eliminator is usually 6in (0.152 m). An additional 12in (0.305 m) above the eliminator is added to obtain uniform flow distribution across the eliminator. If the eliminator is too close to the outlet nozzle, a large part of the flow will be directed to the center of the eliminator, reducing its efficiency. The total length of the separator can be calculated by summing up the dimensions. According to Branan (1994), if L/D is greater than 5, use a horizontal separator. Also, Branan states that if $L/D < 3$, increase L in order that $L/D > 3$, even if the liquid surge volume is increased. Increasing the surge volume is in the right direction.

4.1 Factors affecting separation

Characteristics of the flow stream greatly affect the design and operation of a separator. The following factors must be determined before separator design;

- Gas and liquid flow rates (minimum, average, and peak)
- Operating and design pressures and temperatures
- Surging or slugging tendencies of the feed streams
- Physical properties of the fluids such as density viscosity and compressibility
- Designed degree of separation (e.g., removing 100% of particles greater than 10 microns)
- Presence of impurities (paraffin, sand, scale, etc.)
- Foaming tendencies of the crude oil
- Corrosive tendencies of the liquids or gas.

4.1.1 Sizing considerations

The following must be considered in designing separator vessel based on engineering design guideline.

- The volumes of the dished heads are negligible as compared with the volume of the cylinder.
- Unless specifically stated the length/diameter (L/D) is considered to be acceptable when it is in the range 1.5 to 6.0. There is not a great change in cost over this ranger and other factors such as foundations, plant layout, and symmetry are significant.
- For a vertical separator, the gas flows through the entire cross section of the upper part of the vessel. The feed enters the separator just above the vapor-liquid interface, which should be at least 2ft (0.61m)

from the bottom and at least 4ft (1.22m) from the top of vessel. The interface does not have to be at the center of the vessel.

- For a horizontal separator, the interface does not have to be at the centerline of the vessel. In some cases, a smaller-diameter vessel may be obtained by making the interface location off-center and a design variable. The feed enters at the end of separator just above the vapor-liquid interface, which should be at least 10in (0.25m) from the bottom and at least 16in (0.41m) from the top of the vessel.

4.2 Sizing procedures

This section addressed the basics of three-phase separator (vertical and horizontal) design and provides step-by-step procedures for three-phase gas/liquid/liquid separator design.

In the separator design, it is also worthwhile to clarify two definitions;

holdup and surge times. Holdup is the time it takes to reduce the liquid level from normal (NLL) to Low (LLL) while maintaining a normal outlet flow without feed makeup. Surge time is the time it takes for the liquid level to rise from normal (NLL) to high (HLL) while maintaining a normal feed without any outlet flow. Holdup time (t_H) is based on the stream facilities, whereas surge time (t_S) is usually based on requirements to accumulate liquid as a result of upstream or downstream variations or upset. Table 4-1 shows typical values of holdup time and surge time (Svrcek and Monnery, 1994).

Table 4-1: Typical values of holdup (t_H) and surge (t_S) times (Monnery and Svrcek 1994)

Service	t_H, min	t_S , min
A. Unit feed drum	10	5
B. Separator		
• Feed to column	5	3
• Feed to other drum or tankage with pump or through exchange	5	2
• Without pump	2	1
• Fee to fire heater	10	3

Also, the separator k_s factors based on York Demister and Gas Processors Suppliers' Association is obtained from table 4-2 below.

Table 4–2 Separator k_s- factors *(Monnery and Svrcek, 1994)*

Vendor: Otto H. York Company Inc.
With mist eliminator
$1 \le P \le 15$ $k_s = 0.1821 + 0.0029P + 0.0461 \mathrm{In} P$
$15 \le P \le 40$ $k_s = 0.35$
$40 \le P \le 5{,}500$ $k_s = 0.430 - 0.023 \mathrm{In} P$
Where P is in Psia

Gas Processing Suppliers' Association
$0 \le P \le 1500$ $\quad k_s = 0.35 - 0.0001(P - 100)$
For most vapours under vacuum, $k_s = 0.20$
For glycol and amine solutions, multiply k_s by 0.6 to 0.8
For vertical vessel without demister, divide k_s by 2
For compressor suction scrubber, mole sieve scrubbers, and expander inlet separators, multiply k_s by 0.7 to 0.8 where P is in psig.

Theoretically, equation (3-2) can be used to obtain k_s for separators without mist extractors or typically one-half (½) of that used for vessels with mist extractors.

Separators can be any length, but the ratio of seam-to-seam length to the diameter of the vessel, L/D is usually in the range of 2:1 to 4:1 or in the range of 1.5 to 6.0. Table 4-3 below shows the L/D ratio guidelines as proposed by Monnery and Svrcek 1994.

Table 4–3 L/D ratio guidelines (Monnery and Svrcek, 1994)

Vessel operating pressure, psig	L/D
$0 < P \le 250$ (18 bar) 250 (18 bar) $< P < 500$ (35 bar) $P > 500$ (35 bar)	1.5 – 3.0 3.0 – 4.0 4.0 – 6.0

4.2.1 Vertical separator design procedure

For a three-phase vertical separator, the total height can be broken into different sections, as shown in Figure 4-1. The separator height is then calculated by adding the height of these sections. If a mist eliminator pad is used, additional height is added. The design is based on the methodology of a basic design of three phase vertical separator (Monnery and Svrcek 1994).

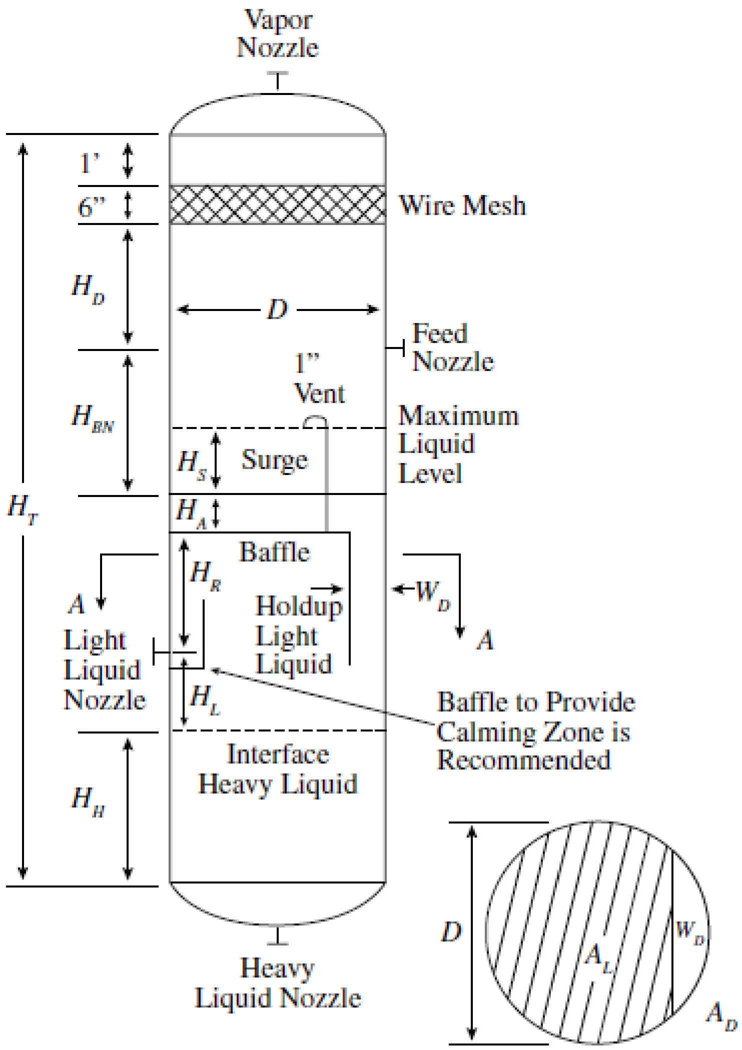


Figure 4-1: Basic design of three phase vertical separator (Monnery and Svrcek 1994)

The calculations of diameter and height based on the methodology of Monnery and Svrcek (1994) are detailed as follow;

1. Calculating the terminal settling velocity (TSV) of droplets using Equation 3-3

$$u_D = k_S \sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$$

And setting $u_G = \frac{2}{3} u_D$ (Tamagna 2012) (4-1)

2. Calculating the vapour (gas) volumetric flow rate, q_G

$$q_G = \frac{m_G}{\rho_G} \quad (4\text{-}2)$$

3. Calculating the vessel internal diameter, D_I:

$$D_I = \left(\frac{4q_G}{\pi u_G}\right)^{1/2} \quad (4\text{-}3)$$

If there is a mist eliminator; D_I + 3-6 inch (0.08 – 0.15m)

If there is no mist eliminator, $D = D_I$.

4. Calculating the settling velocity of the heavy liquid out of the light liquid. (the maximum is 10in./min (0.0042m/s)

$$u_{HL} = \frac{k_s(\rho_{HL} - \rho_{LL})}{\mu_{LL}} \qquad (4\text{-}4)$$

5. Similarly, calculating the rising velocity of the light liquid out of the heavy liquid phase

$$u_{LL} = \frac{k_s(\rho_{HL} - \rho_{LL})}{\mu_{HL}} \qquad (4\text{-}5)$$

6. Calculating the light and heavy liquid volumetric flow rates, q_{LL} and q_{HL}

$$q_{LL} = \frac{m_{LL}}{\rho_{LL}} \qquad (4\text{-}6)$$

$$q_{HL} = \frac{m_{HL}}{\rho_{HL}} \qquad (4\text{-}7)$$

7. Calculating the settling times for the heavy liquid droplets to settle through a distance, H_L (minimum 1ft, 0.3042m) and for the light liquid droplets to rise through a distance, H_H (minimum 1ft, 0.3042m).

$$t_s, H_L = \frac{H_L}{u_{HL}} \tag{4-8}$$

$$t_s, L_L = \frac{H_H}{u_{LL}} \tag{4-9}$$

8. Calculating the area of a baffle plate (if any); A_L, which is the settling area for the light liquid

$$A_L = A - A_D \tag{4-10}$$

Where A is vertical vessel cross-sectional area, and A_D is downcomer cross-sectional area given as;

$$A = \frac{\pi D^2}{4} \tag{4-11}$$

$$A_D = \left(\frac{q_{HL+} q_{LL}}{G}\right) \tag{4-12}$$

Where the baffle liquid loads (G) be obtained from Figure 4-1 below;

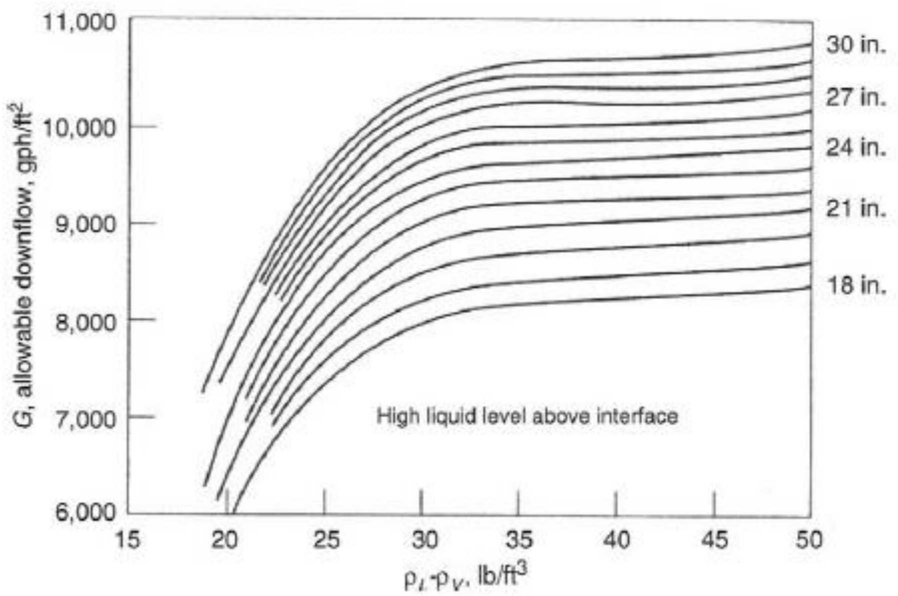


Figure 4-2: Determining the downcomer allowable flow (Monnery and Svrcek 1994)

Or using equation 4-13 below;

$$\frac{A_D}{A} = \left\{-4.755930\ \times 10^{-5}\ + 0.174875\left(\frac{W_D}{D}\right) + 5.668973\left(\frac{W_D}{D}\right)^2 - 4.916411\left(\frac{W_D}{D}\right)^3 - 0.145348\left(\frac{W_D}{D}\right)^4\right\} / \left\{1.0 + 3.924091\left(\frac{W_D}{D}\right) - 6.358805\left(\frac{W_D}{D}\right)^2 + 4.018448\left(\frac{W_D}{D}\right)^3 - 1.801705\left(\frac{W_D}{D}\right)^4\right\} \qquad (4\text{-}13)$$

Where D is the vessel diameter and the downcomer chord width (W_D) is assumed 4in (0.1016m).

9. Calculating the residence time (t_R) of each phase based on the volumes occupied by the light and heavy phases as;

$$t_{R,LL} = \frac{H_L\, A_L}{q_{LL}} \qquad (4\text{-}14)$$

$$t_{R,HL} = \frac{H_H\, A_H}{q_{HL}} \qquad (4\text{-}15)$$

Where $A_H = A$

If $t_{R,LL} < t_{HL}$ or $t_{R,HL} < t_{LH}$ this implies that liquid separation is controlling, the diameter needs to be increasing and procedure repeated from step 7.

10. Calculating the height of the light liquid above the outlet (holdup height) H_R based on the required holdup time (t_H) as

$$H_R = \frac{q_{LL}\, t_H}{A_L} \qquad (4\text{-}16)$$

Check this value with that assumed in step 7 to ensure that the assumed value is reasonable.

11. If surge is not specified, calculating the surge height (H_S) based on surge time (t_S)

$$H_S = \left(\frac{t_s(q_{LL} + q_{HL})}{A}\right) \qquad (4\text{-}17)$$

12. Calculating the vessel total height (H_T) as

$$H_T = H_H + H_L + H_R + H_A + H_{BN} + H_D \qquad (4\text{-}18)$$

Where H_A is liquid level above baffle, which is 6 in. (0.15m) minimum, and H_{BN} is liquid height from above baffle to feed nozzle.

$H_{BN} = \frac{1}{2} D_N +$ greater of 2 ft (0.61m) or $H_s + 0.5$ ft (0.15m)

H_D = 0.5D or a minimum of; 36in (0.91m) + $\frac{1}{2} D_N$ (without mist eliminator)

Or 24 in (0.61m) + $\frac{1}{2} D_N$ (with mist eliminator)

Where the nozzle diameter (D_N) is calculated using the following criterion:

$$D_N \geq \left[\frac{4q_M}{\pi 60/\sqrt{\rho}_M}\right]^{0.5} \qquad (4\text{-}19)$$

q_M and ρ_M are inlet mixture volume flow rate and density of mixture respectively. H_D is disengagement height.

If a mist eliminator pad is used, additional height is added as shown in Figure 4-1.

Using the design procedure outlined above and the fluid properties in Table 3-4, the design calculations and results are shown in Table 4-4 (Appendix A) at 80, 15 and 2bar for the vertical separators.

The following assumptions are made;

- No mist eliminator
- Setting $u_G = \frac{2}{3} u_D$ (Tamagna 2012)
- Vessel is half filled
- Vessel internal diameter$D_I = D$ (vessel diameter) = 2m if less that.
- Height from liquid interface to light liquid nozzle, H_L = 1ft (0.3042m)
- Downcomer width (W_D) = 1in (0.1016m)
- Holdup time (t_H) = 5min (300s) and surge time (t_S) = 3min (180s)
- Liquid level above baffle (H_A) = 6in (0.15m)
- Disengagement height (H_D) = 0.5D

The ratio of the total height to the diameter of the separators is in the range of 1.5 to 6.0, which is the acceptable range.

With mist eliminator, the design calculations and results are shown in Table 4-5 (Appendix A) at 80, 15 and 2bar.

The following assumptions are made;

- With mist eliminator
- Setting $u_G = \frac{2}{3} u_D$ (Tamagna 2012)
- $D_I = D$ = 2m + 6in (0.15m) .
- Height from liquid interface to light liquid nozzle, H_L = 1ft (0.3042m)

- Downcomer width (W_D) = 1in (0.1016m)
- Holdup time (t_H) = 5min (300s) and surge time (t_S) = 3min (180s)
- Liquid level above baffle (H_A) = 6in (0.15m)
- Disengagement height (H_D) = 0.5D

The ratio of the total height to the diameter of the separators is in the range of 1.5 to 6.0, which is the acceptable range for 80bar. Additional height (0.3042m) is added to the separators at 15 and 2 bara.

4.2.2 Horizontal separators design procedure

For a three-phase horizontal separator, the horizontal design procedures incorporate optimizing the diameter and the length. Wall thickness, surface area and approximate vessel weight are obtained from Table 4-6 below.

Table 4-6: Wall thickness, surface area and approximate vessel weight (Monnery & Svrcek 1994)

Components	Wall thickness	Surface area

Shell	$\frac{PD}{2SE-1.2P}+W_C$	πDL
2:1 Eliptical heads	$\frac{PD}{2SE-0.2P}+W_C$	$1.09D^2$
Hemispherical heads	$\frac{PD}{4SE-0.4P}+W_C$	$1.571D^2$
Dished heads	$\frac{PD}{4SE-0.4P}+W_C$	$0.84D^2$

Approximate vessel weight is given by;

$$W\left(\frac{490Ib}{ft^3}\right)\left(\frac{t\ in}{12}\right)(A_{shell}+2A_{head}) \qquad (4\text{-}20)$$

Selection of the horizontal separator heads is based on Table 4-7 below.

Table 4-7: Selection of horizontal separator separator heads (Monnery & Svrcek 1994)

Conditions	Typical heads used
D < 15ft (4.57 bar) and P < 100psig (7.9bar)	Dished with knuckle radius = 0.6D
D < 15ft and P > 100psig	2:1 Eliptical
D > 15ft regardless of pressure	Hemispherical

Figure 4-2 is a basic design of horizontal three-phase separator with weir.

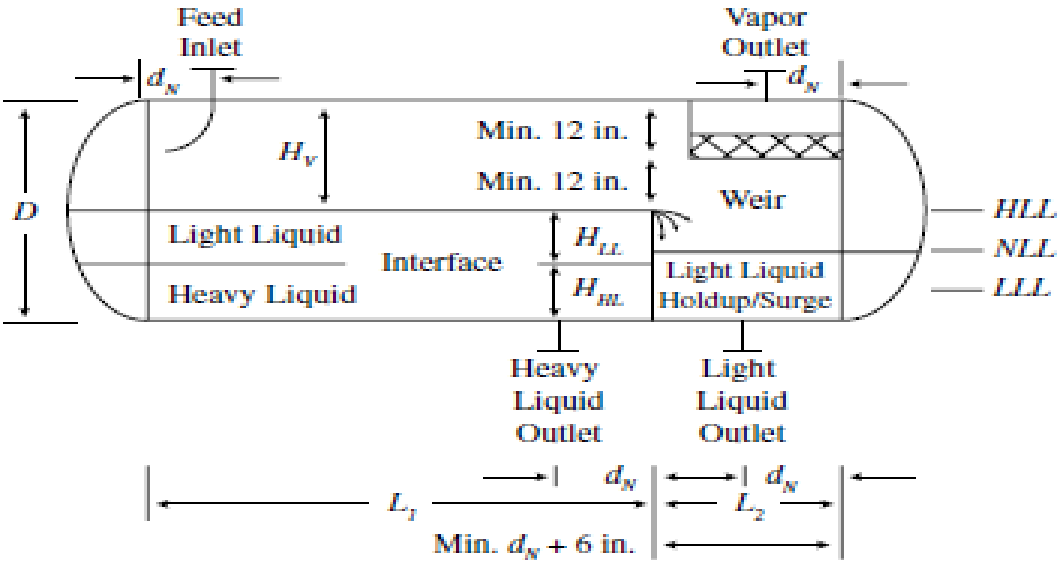


Figure 4-3: Basic design of three phase horizontal separator with weir (Monnery and Svrcek 1994)

The horizontal design procedures with weir are detailed as follow:

1. Calculating the vapour volumetric flow rate

$q_G = \frac{m_G}{\rho_G}$ from equation 4-2

2. Calculating the light and heavy liquid volumetric flow rates, q_{LL} and q_{HL}

$q_{LL} = \frac{m_{LL}}{\rho_{LL}}$ from equation 4-6

And $q_{HL} = \frac{m_{HL}}{\rho_{HL}}$ from equation 4-7

3. Calculating the terminal settling velocity using equation 3-3

$u_D = k_s\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$

And setting $u_G = \frac{2}{3}u_D$ (Tamagna 2012) from equation 4-1 for a conservative design and k_s from Table 4-2.

4. Selecting hold up and surge times from Table 4-1, and calculate the holup and surge volumes, V_H *and* V_S.

$$V_H = t_H\, q_{LL} \qquad (4\text{-}21)$$

$$V_S = t_S\, q_{HL} \qquad (4\text{-}22)$$

5. Obtaining L/D from Table 4-3 and initially calculate the diameter according to

$$D = \left(\frac{4(V_H + V_S)}{0.5\pi(L/D)}\right)^{1/3} \qquad (4\text{-}23)$$

Then calculating the total cross-sectional area, using $A_T = \frac{\pi D^2}{4}$

6. Setting the vapor space height, H_G, to the largest of 0.2D or 2 ft (1 ft if there is no mist eliminator).
   Using $\frac{H_G}{D}$ in equation 4-13 (replacing with $\frac{W_D}{w}$) obtain $\frac{A_G}{A}$ and calculate A_G
7. Calculating the low liquid level in the light liquid component in the vessel

$$H_{LLL} = 0.5D + 7 \qquad (4\text{-}24)$$

If $D \leq 4.0$ft (1.2m), then H_{LLL} = 9in (0.23m)

8. Calculating the cross-sectional area of the light liquid above the bottom of the vessel, A_{LLL} Using $\frac{H_{LLL}}{D}$ (instead of $\frac{W_D}{W}$) in equation 4-13.
9. Calculating the weir height

$H_w = D - H_G$ (4-25)

If H_w < 2ft, increase D and repeat the calculation from step 6.

10. Calculating the minimum length, L_2 to accommodate the liquid hold/surge.

$$L_2 = \frac{V_H + V_S}{A_T - A_G - A_{LLL}} \quad (4\text{-}26)$$

Round to the nearest $\frac{1}{2}$ ft. The minimum for for $L_2 = D_N + 12in$ (0.3m)

11. Setting the interface at the height $H_W/2$ obtaining the heights of the heavy and light liquids H_{HL} and H_{LL}

12. For the liquid settling compartment, calculating the cross-sectional area, A_{HL} of the heavy liquid using H_{HL}/D (instead of $\frac{W_D}{W}$) in equation 4-13 and the cross-sectional area of the light liquid from; $A_{LL} = A_T - A_V - A_{HL}$ (4-27)

13. Calculating the settling velocities of the heavy liquid out of the light liquid phase u_{HL}, and the light out of the heavy phase u_{LH}, using equations 4-4 and 4-5.

14. Calculating the settling times of the heavy liquid out of the light liquid phase and the light liquid out of the heavy liquid phase;

$$t_s, H_L = \frac{H_{LL}}{u_{HL}} \quad (4\text{-}28)$$

$t_s, L_L = \frac{H_{HL}}{u_{LL}}$ (4-29)

15. Calculating the minimum length, L_1 to facilitate liquid-liquid separation as the larger of;

$$L_1 = \max\left(\frac{t_s, L_L\, q_{HL}}{A_{HL}}\ or\ \frac{t_s, H_L q_{LL}}{A_{LL}}\right) \quad (4\text{-}30)$$

Round to the nearest $\frac{1}{2}$ft (0.15m)

16. Finding L

$L = L_1 + L_2$ (4-31)

17. Calculating the liquid dropout time, t_{DL}, using the following equation:

$$t_{DL} = \frac{H_G}{u_G} \quad (4\text{-}32)$$

18. Calculating the actual vapor velocity, u_{GA}, as

$$u_{GA} = \frac{q_G}{A_G} \quad (4\text{-}33)$$

19. Calculating the minimum length required for vapor/liquid separation,

$$L_{MIN} = u_{GA}t_{DL} \quad (4\text{-}34)$$

20. If $L < L_{MIN,}$ then set $L = L_{MIN}$ (vapor/liquid separation controls).

If $L \ll L_{MIN}$ increase $H_{G,}$ and recalculate A_G

If $L > L_{MIN}$, the design in acceptable for vapor/liquid separation.

If $L \gg L_{MIN,}$ (liquid separation and holdup controls). L can only be reduced and L_{MIN} increase if H_G is reduced.

H_G may only be reducing if greater than the minimum specified in step 6. With reduced$H_{G,}$ recalculate A_G and repeat from step 10.

21. Calculating L/D, If L/D $\ll$ 1.5, then decrease D (unless it is already at a minimum) and repeat from step 6.

If L/D $\gg$ 6.0, then increase D and repeat from step 5

22. Calculating the thickness of the shell and heads according to table 4-6.

23. Calculating the surface area of the shell and head according to table 4-6

24. Calculating the approximate vessel weight according to equation 4-20

25. Increase or decrease the diameter by 6 in. increment and repeat the calculations until L/D ranges from 1.5 – 6.0

26. With the optimum vessel size (minimum weight), calculate normal and high liquid levels:

$$H_{HLL} = D - H_G \quad (4\text{-}35)$$

$$A_{NLL} = A_{NLV} + V_H/L_2 \quad (4\text{-}36)$$

Obtain H_{NLL} with replacing A_{NLL}/A_T with $\frac{W_D}{W}$ in equation 4-13.

Using the design procedure outlined above and the fluid properties in Table 3-4, the design calculations and results are shown in Table 4-8 (Appendix A) at 80, 15 and 2bar for the horizontal separators.

The following assumptions are made;

- No mist eliminator
- k_s = 0.0305m/s (Silla 2003) without mist eliminator
- Setting $u_G = \frac{2}{3} u_D$ (Tamagna 2012)
- Vessel is half filled
- Holdup time (t_H) = 10min (600s) and surge time (t_S) = 5min (300s)
- Gas disengagement area height (H_G) = 1ft (0.3042m)
- L/D = 4 for $P >$ 30bar
- L/D = 3 for $P <$ 30bar

The following assumptions are made for the design of horizontal separator with mist eliminator;

- k_s = 0.137m/s (typical NORSOK standard)
- Setting $u_G = \frac{2}{3} u_D$ (Tamagna 2012)
- Holdup time (t_H) = 10min (600s) and surge time (t_S) = 5min (300s)
- Gas disengagement area height (H_G) = 2ft (0.6084m)
- L/D = 4 for $P >$ 30bar
- L/D = 3 for $P <$ 30bar

The design calculations and results are shown in Table 4-9 (Appendix A) at 80, 15 and 2bar for the horizontal separators.

5 Droplet settling theory

In gravity settling, the dispersed phase drops/bubbles will settle at a velocity determined by equating the gravity force on the drop/bubble with the drag force caused by its motion relative to the continuous phase. In horizontal vessels, a simple ballistic model can be used to determine a relationship between vessel length and diameter. In the vertical vessels, the settling theory results in a relation for the vessel diameter.

For horizontal separators; droplet settling theory using a ballistic model results in the relationship for liquid drops in gas phase as shown below;

$$\frac{L_{eff} D_I^2 F_{GA}}{H_G} = \frac{TZq_G}{P}\left[\left(\frac{\rho_G}{\rho_L - \rho_G}\right)\frac{C_D}{d_D}\right]^{\frac{1}{2}} \qquad (5\text{-}1)$$

Where

L_{eff} = effective length of vessel where separation occurs, m

F_{GA} = fractional gas phase cross sectional area

Z = gas compressibility

d_D = droplet diameter, m

C_D = drag coefficient

D_I = vessel internal diameter, m

For bubbles or liquid drops in liquid phase is given by;

$$\frac{L_{eff} D_I^2 F_{CA}}{H_C} = q_C\left[\left(\frac{\rho_C}{\rho_d - \rho_C}\right)\frac{C_D}{d_D}\right]^{\frac{1}{2}} \qquad (5\text{-}2)$$

Where

F_{CA} = fractional continuous phase cross sectional area

H_C = continuous liquid phase space height, m

ρ_C = continuous liquid phase density, kg/m^3

ρ_D = dispersed liquid phase density, kg/m^3

q_C = continuous liquid phase flow rate, m^3/s

For low Reynolds number flow, the continuous liquid phase space height H_C can be obtained by;

$$H_C = \frac{t_{RC}(\Delta\gamma)d_D^2}{\mu_C} \tag{5-3}$$

Where

t_{RC} = continuous phase retention time, s

μ_C = continuous phase viscosity, Pa.s

$\Delta\gamma$ = specific gravity difference (heavy/light) of continuous and dispersed phases

For vertical vessels; droplet settling theory using a ballistic model results in the relationship for liquid drops in gas phase as shown below;

$$D_I^2 = \frac{TZq_G}{P}\left[\left(\frac{\rho_G}{\rho_L-\rho_G}\right)\frac{C_D}{d_D}\right]^{\frac{1}{2}} \tag{5-4}$$

For bubbles or liquid drops in liquid phase is given by;

$$D_I^2 = q_C \left[\left(\frac{\rho_C}{\rho_d - \rho_C}\right)\frac{C_D}{d_D}\right]^{\frac{1}{2}} \qquad (5\text{-}5)$$

Assuming a low Reynolds number flow, the vertical vessel internal diameter is obtained as;

$$D_I^2 = \frac{q_C \mu_C}{(\Delta\gamma) d_D^2} \qquad (5\text{-}6)$$

5.1 Retention time of liquid phase

For horizontal vessels, the retention time of the liquid phase is obtained from the relationship of vessel diameter and length given by;

$$D_I^2 L_{eff} = \frac{t_{RL}\, q_{LL} + t_{RH}\, q_{HL}}{F_{VA}} \qquad (5\text{-}7)$$

Where

t_{RL} = retention time of light liquid, s

t_{RH} = retention time of heavy liquid, s

F_{VA}= fraction of vessel cross section area filled by liquid

Similarly for vertical vessels, the retention time of the liquid phase is obtained from the relationship of vessel diameter and liquid pad heights given by;

$$D_I^2 (H_O + H_W) = t_{RL}\, q_{LL} + t_{RH}\, q_{HL} \qquad (5\text{-}8)$$

Where

H_O =oil (light liquid) pad height, m

H_W = water (heavy liquid) pad height, m

6 Internal sizing

Separation of a gas-liquid stream in a vessel is assisted by process internals which provide additional separation beyond that of gravity.

As mentioned previously, many types of demisters and other internals are limited by a maximum velocity given by equation 2-1 and depend upon their types and the manufacturer specifications. Therefore sizing internals

Droplets are removed from a vapor stream through a series of three stages; collision and adherence to a target, coalescence into larger droplets, and drainage from the impingement element. Knowing the size distributions is important because empirical evidence shows that the target size is important in the first step of removal and must be in the order of magnitude as the particles to be removed. These steps are shown schematically in Figure 6-1 for mist elimination using wire mesh mist elimination.

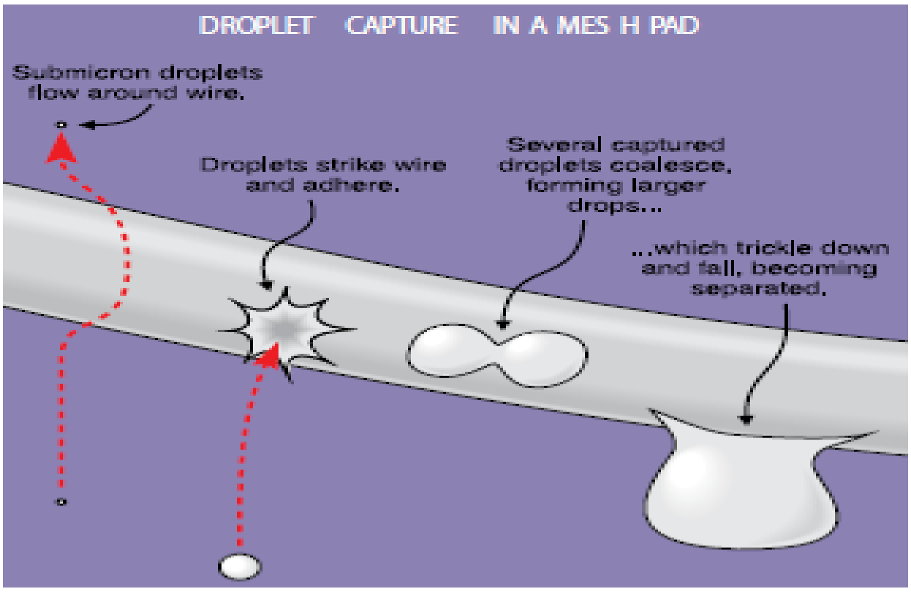


Figure 6-1: Mist elimination using wire mesh mist extractor (AMACS process tower internals)

This work is limited to sizing wire mesh pad internal and calculating its efficiency.

6.1 Sizing wire mesh pads

The sizing procedures are detailed below.

- Determination of the optimum design gas velocity. The Souders-Brown equation/equation 2-1 is used to determine this velocity based on the physical properties of the liquid droplets and carrying vapour.

The recommended value of k_d varies and depends upon several factors such as liquid viscosity, surface tension, liquid loading, and operating pressure. Each manufacturer has its own recommended values. For general sizing, a k_d value of 0.1 m/s can be used as a guideline.

- Obtaining the capacity factor using Table 6-1 below.

This is influenced by type and style of mesh or vane targets used, and the geometry of the targets (vertical or horizontal relative to the vapour flow).

Table 6-1 Standard Sounder's Brown coefficient (k-factor) for mesh and plate packs (AMACS process tower internals, Houston)

Pad arrangement	*k, m/s*
1. Horizontal style 4CA pad 2. Style 4CA mistermesh pad 3. Horizontal plate pack 4. Vertical plate pack	0.107 0.128 0.152

	0.198

- Determination of the cross-sectional area.

After selecting the appropriate capacity factor and calculating the ideal vapour velocity, the cross-sectional area of mist eliminator is readily determined by dividing the volumetric flow rate by the velocity.

- Predicting the efficiency of the mesh pad.

Having established the design velocity for the application, the efficiency of the mesh pad for droplet of a particular size can be predicted by calculating the inertial parameter *k* as follows;

$$k = \frac{(\rho_L - \rho_G) u_d d_D^2}{9 \mu d_w} \qquad (6\text{-}1)$$

Where

d_w = wire diameter or thickness, m

u_d = design velocity of the wire mesh, m/s

Using this calculated *k* value with Figure 6-2 below to find the corresponding value of the impaction efficiency fraction, *E.*

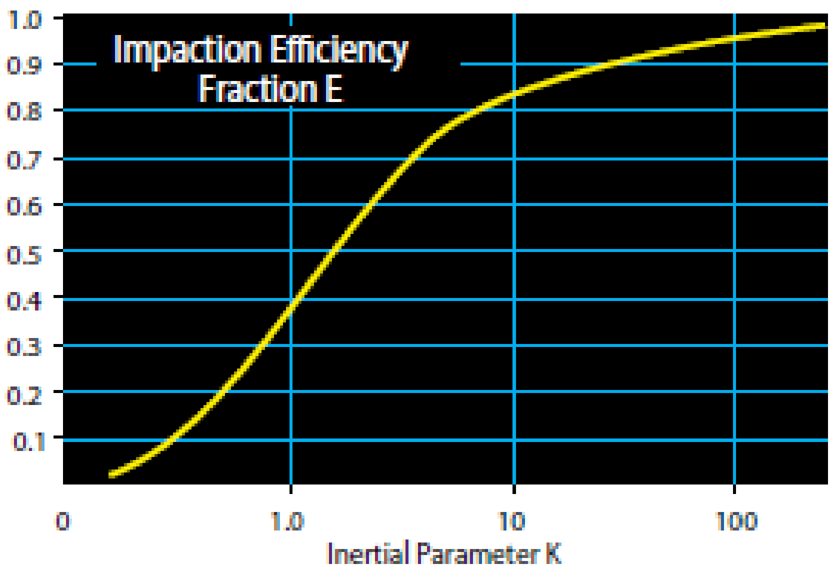


Figure 6-2: Determining impact efficiency fraction, E using inertial parameter, k (AMACS process tower internals, Houston)

From Table 6-2 below, we can find the specific surface area, S for the mesh style of interest and determine SO, the area of the area of the mist eliminator perpendicular to vapour flow and with a correction factor of 0.67 to remove that portion of the knitted wire not perpendicular to the gas flow.

Table 6-2: Wire and plastic mesh styles (AMACS process tower internals, Houston)

Mesh Style	Density lbs/ft³	Diameter D, inches	Surface, S, ft²/ft³	Percent Voids, ε
Metal mesh				
7CA	5.0	0.011	45	99.0
5CA	7.0	0.011	65	98.6
4CA	9.0	0.011	85	98.2
4BA	12.0	0.011	115	97.6
3BF	7.2	0.006	120	98.6
3BA	12.0	0.006	200	97.6
Plastic mesh				
8P	4.0	0.011	130	92.0
8K	4.0	0.011	160	96.3
8T	4.0	0.011	130	97.0

Mesh Style	Density lbs/ft³	Diameter D, inches	Surface, S, ft²/ft³	Percent Voids, ε
Metal mesh				
8D	9	0.0008	615	99.0
8TMW 11	12	0.0008	1170	99.0
6BE	9	0.000036	3725	99.0
Plastic mesh				
8PP	3	0.001	480	99.0
8TT	5	0.0008	1240	99.0

Table 6-2 shows a few of the more common mesh styles available, together with mesh density and void fraction, and most importantly, the diameter and specific surface area (i.e. the target density) of filaments used. It is the amount of targets per unit volume which influences removal efficiency, not the density of mesh (the greater the number of targets the greater the probability of a successful collision).

$$SO = 0.67\frac{1}{\pi}t_w S \quad (6\text{-}2)$$

- Calculating the capture efficiency as given in equation 6-3 below;

$$\varepsilon\ (\%)\ =\ 100\ -\left(\frac{100}{e^{ESO}}\right) \qquad (6\text{-}3)$$

Where

ε = the capture efficiency, %

SO = corrected pad specific surface area

E = impaction efficiency fraction

Using the procedures and guidelines above, the design velocity is obtained using equation 2-1 as;

u_d = $107\sqrt{\frac{598-69}{69}}$ = 0.27m/s (0.89ft/s) for horizontal style 4CA mesh pad.

The cross sectional area is obtained as;

A = 0.05/0.27 = 0.19m^2

Calculating the removal efficiency at 5μm droplet size and obtaining the value of k using equation 6-1

$$k\ = \frac{(598-69)0.27(5*10^{-6})^2}{9*1.56*10^{-5}\ 2.79*10^{-4}} = 0.095\text{m/s}\ (0.3\text{ft/s})$$

From Figure 6-2, the corresponding impaction efficiency fraction $E \sim 0.12$

Applying 0.3ft (0.09m) and 0.5ft (0.15m) of the thick element of the mesh

SO = 0.67*$\frac{1}{3.142}$*0.3*85+36 = 7.7

ε =100-$\frac{100}{e^{0.12*7.7}}$ = 60.3%

Also for 0.5ft thick;

SO = 0.67*$\frac{1}{3.142}$*0.5*85+36 = 12.9

ε =100-$\frac{100}{e^{0.12*12.9}}$ = 79%

Calculating the removal efficiency at 10µm droplet size;

$k \ = \frac{(598-69)0.27(10*10^{-6})^2}{9*1.56*10^{-5}\ 2.79*10^{-4}}$ = 0.36m/s (1.18ft/s)

From Figure 6-2, the corresponding impaction efficiency fraction $E \sim 0.40$

Applying 0.3ft (0.09m) and 0.5ft (0.15m) of the thick element of the mesh

SO = 0.67*$\frac{1}{3.142}$*0.3*85+36 = 7.7

ε =100-$\frac{100}{e^{0.40*7.7}}$ = 95.4%

Also for 0.5ft thick;

$SO = 0.67*\frac{1}{3.142}*0.5*85+36 = 12.9$

$\varepsilon = 100 - \frac{100}{e^{0.4*12.9}} = 99.4\%$

The plot of separation (capture) efficiency versus the droplet diameter in micron is shown in figure 6-3 below;

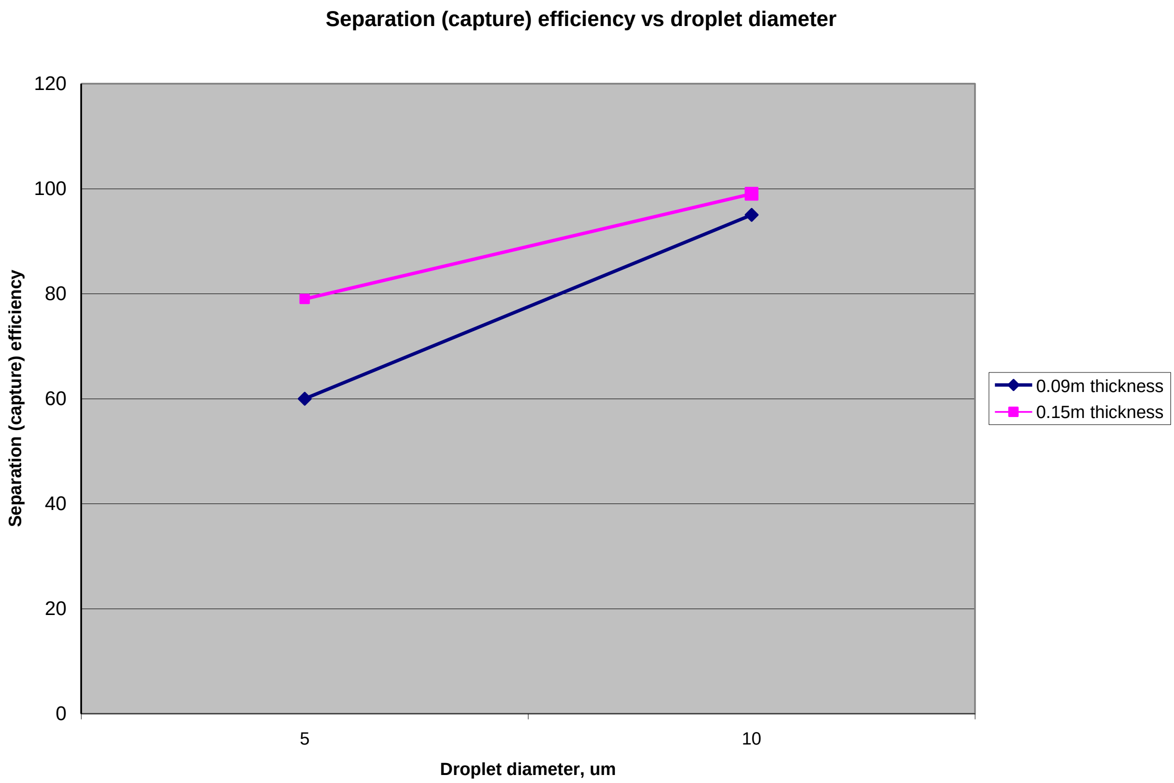


Figure 6-3: Separation (capture) efficiency versus droplet diameter, μm

The efficiency of a separator is defined here as the fraction (or percentage) of the liquid entering the vessel that is separated off.

The design settling velocity of the mesh pad versus the droplet diameter is shown in Figure 6-4 below.

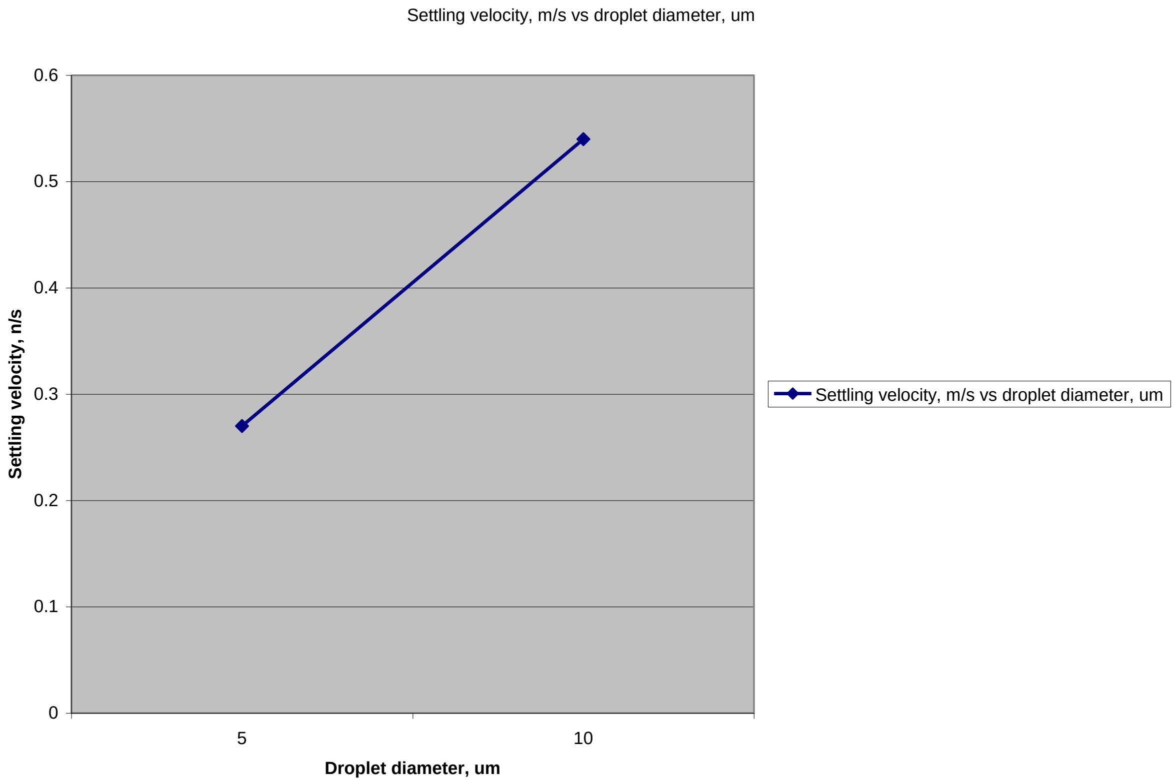


Figure 6-4: Design settling velocity of the mesh and the droplet diameter at 80bar for the horizontal separation

Also the separating efficiency of a vane mist extractor depends on:

- The number of vanes in the element
- Distance between the vanes
- Angle of the vanes and
- Size of liquid particles

6.2 Predicting pressure drop

Operating pressure loss across the pad within the above design range is normally less than 0.5 kPa depending upon mesh density, pad thickness, liquid loading and vapour rate. An approximate pressure drop can be estimated from equation 6-4 below;

$$\Delta P_{wet}(\text{kPa}) = C(\rho_L - \rho_G)k^2 t_w \qquad (6\text{-}4)$$

Where

ΔP_{wet} = Wet pressure drop, (kPa). Dry pressure drop is about half of the wet figure.

C = 0.20 for a typical style mesh demister

t_w = is the pad thickness, m

The overall pressure drop is the sum of the head loss incurred as the gas travels through the mesh, as well as that due to the resistance to captured liquids. Liquid accumulates as a pool in the bottom of the mist eliminator.

7 Operating problems

Problems occasionally occurred in the operation of separators (Arnold and Stewart, 1998) are as follows;

- Foamy crude
- Paraffin
- Sand
- Liquid carryover and gas blowby
- Emulsions

7.1 Foamy crude

The major cause of foam in crude oil is the appearance of impurities, other than water, that are impractical to remove before the stream reaches the separator. Foam presents no problem within a separator if the internal design assures adequate time or sufficient coalescing surface for the foam to "break." Problem of foaming in a separating vessel is as follows;

- Mechanical control of liquid level is aggravated because any control device must deal with three liquid phases, an emulsion is the third phase, and instead of two-phases.
- Foam has a large volume-to-weight ratio. Therefore, it can occupy much of the vessel space that would otherwise be available in the liquid-collecting or gravity-settling sections.
- In an uncontrolled foam bank, it becomes impossible to remove separated gas or degassed oil from the vessel without entraining some of the foamy material in either the liquid or the gas outlets.

Essentially as the foam is dispersed, it creates very small liquid droplets, which carry over. The amount of foam is dependent on the pressure drop to which the inlet liquid is subjected, as well as the characteristics of the liquid at separator conditions. In some cases, the effect of temperature may be significant. Foam will often be effective in increasing the capacity of a given separator.

Foam can be reduced by;

- Using a defoaming pack
- Using defoaming chemicals, and
- Utilizing heat to break it down.

7.2 Paraffin

Coalescing plates in the liquid section and mesh pad mist extractors in the gas section are particularly prone to clogging by accumulations of paraffin waxes. Hand holes, and nozzles should be provided to allow steam, solvent, or other types of cleaning of the separator internals.

Also, the bulk temperature of the liquid should always be kept above the cloud point of the crude oil to prevent paraffin wax formation in the separators.

7.3 Sand

Sand is often troublesome in separators by causing cutout of valve trim, plugging of separator internals, and accumulation in the bottom of the separator, thus leading to level control problems. Traditionally, sand has only been removed once it has collected in the main production separators.

However, removal of sand upstream of these separators reduces sand problems to a minimum, giving substantial operational benefits.

To meet these needs, the Mozley Wellspin desander has been developed to remove sand effectively in simple, compact systems based on solid/liquid hydrocyclones, which remove the sand before it enters the separator. Sand problems may be solved by using a filter or desanding cyclone before the separator. However, filters will quickly block in sandy service and are not often used.

7.4 Liquid carryover and gas blowby

Liquid carryover occurs when free liquid escapes with the gas phase and can indicate high liquid level, thus causing damage to vessel internals, foam, improper design, plugged liquid outlets, or a flow rate that exceeds the design rate of the vessel. Gas blowby occurs when free gas escapes with the liquid phase and can be an indication of low liquid level, vortexing, or level control failure.

7.5 Emulsion

Emulsions are often troublesome in the operation of three-phase separators. Over a period of time an accumulation of emulsified materials and/or other impurities usually will form at the interface of the water and oil phases. In addition to adverse effects on the liquid level control, this accumulation will also decrease the effective oil or water retention time in the separator, with a resultant decrease in water-oil separation efficiency. The addition of chemicals and/or heat often minimizes this difficulty. Also; lowering the settling time needed for oil-water separation by either the application of heat in the liquid section of the separator, or the addition of demulsifying chemicals.

8 Discussions

Separating vessels in oil and gas processing service are of two kinds; those substantially without internals and those with internals. The main functions of the first kind, called drums or tanks, are intermediate period for storage or to provide a phase separation by settling. Their sizes may be established by definite process calculations or by general rules based on experience. The second category comprises the shells of equipment whose housing can be designed and constructed largely independently of whatever internals are necessary.

8.1 Separating vessel without internals (no mist extractor)

The separators without internals were designed for different pressures (80, 15 2bar). For an oil and gas separator to accomplish its primary functions, pressure must be maintained in the separator so that the liquid and gas can be discharged into their respective processing or gathering systems. Pressure was maintained in the separator by use of a gas backpressure valve on each separator. For the vertical separators without internals, the terminal settling velocities of the droplets for the three separators were obtained as 0.102m/s, 0.35m/s, and 0.79m/s for 80, 15 and 2 bar respectively. The settling velocities of the heavy liquid out of the light liquid were 0.0006m/s, 0.00025m/s, 0.00019m/s and that of the rising velocities of the light liquid out of the heavy liquid phase were 0.00041m/s, 0.00022m/s, and 0.00018m/s. The surge times for the high light liquid were 8.45, 20 and 27 minutes while that of the heavy liquid were 12.4, 23 and 28 minutes. The retention time for the light liquid (oil) was approximate 2 minutes for the three separators while that of the heavy liquid (water) were 46, 80 and 106 minutes. These were greater than the surge time and the calculation was proceeded to obtain the total height as 3.63m, 3.32m and 3.16m. The height/diameter ratios were 1.82, 1.66, and 1.58 which were acceptable in the range of 1.5 to 6.0.

For the horizontal separators without internals, the terminal settling velocities of the droplets for the three separators were obtained as 0.085m/s, 0.206m/s, and 0.439m/s for 80, 15 and 2 bar respectively. The settling velocities of the heavy liquid out of the light liquid were 0.00049m/s, 0.000149m/s, 0.000103m/s and that of the rising velocities of the light liquid out of the heavy liquid phase were 0.00034m/s, 0.000127m/s, and 0.00010m/s which were less than maximum of 0.0042m/s. The surge times for the high light liquid were 19, 69 and 95 minutes while that of the heavy liquid were 27, 81 and 98 minutes. The total length was obtained as 6m, 6m and 5m. The length/diameter ratios were 4.2, 3.9, and 2.9 which were acceptable in the range of 1.5 to 6.0.

8. 2 Separating vessel with internals (mist extractor)

The separators with internals were also designed for different pressures (80, 15 2bar). Difference in density of the liquid and gaseous hydrocarbons accomplished acceptable separation in the oil and gas separation. However, it is necessary to use mechanical devices commonly referred to as "mist extractors" to remove liquid mist from the gas before it is discharged from the separator.

For the vertical separators with internals, the terminal settling velocities of the droplets for the three separators were obtained as 0.205m/s, 0.70m/s, and 1.57m/s for 80, 15 and 2 bar respectively. The settling velocities of the heavy liquid out of the light liquid were 0.00012m/s, 0.00050m/s, 0.00037m/s and that of the rising velocities of the light liquid out of the heavy liquid phase were 0.00082m/s, 0.00043m/s, and 0.00035m/s. The surge times for the high light liquid were 6.2, 11.8 and 14.5 minutes while that of the heavy liquid were 4.2, 10 and 13.7 minutes. The retention times for the light liquid (oil) was 1.6, 2, and 2.26 minutes for the three separators while that of the heavy liquid (water) were 52, 91 and 122 minutes. These were greater than the surge time and the calculation was proceeded to obtain the total height as 3.48m, 3.21m and 3.07m. The height/diameter ratios were 1.6, 1.49, and 1.42. Additional heights of 1ft (0.3042) were added to the second and third stage separators.

For the horizontal separators with internals, the terminal settling velocities of the droplets for the three separators were obtained as 0.379m/s, 0.928m/s, and 1.97m/s for 80, 15 and 2 bar respectively. The settling velocities of the heavy liquid out of the light liquid were 0.0015m/s, 0.00057m/s, 0.00044m/s and that of the rising velocities of the light liquid out of the heavy liquid phase were 0.0015m/s, 0.00057m/s, and 0.00044m/s which were less than maximum of 0.0042m/s. The terminal settling velocity is inversely proportional to the viscosity of the continuous phase. Therefore the bigger the viscosity of the continuous phase is, the more difficult would be to settle droplets out of the continuous phase. The surge times for the high light liquid were 3, 11.7 and 16 minutes while that of the heavy liquid were 4.5, 13.7 and 16.7 minutes. The total length was obtained as 7m, 7m and 5m. The length/diameter ratios were 5, 4, and 3 which were acceptable in the range of 1.5 to 6.0. It is observed that the terminal settling velocity in separators with internals are higher than that with no internals.

The effect of the mist eliminator is to increase the maximum allowable velocity and therefore to reduce the drum diameter.

8.3 The wire mesh pads

Wire-meshs or Meshpads are used to separate liquid from gas phase. Wire Mesh mist extractors are made by knitting wires into tightly packed layers which are stacked to achieve the thickness needed. They are installed horizontally in vessels with gas stream flowing vertically upwards through the pad. Meshpads can be installed in both vertical and horizontal vessels.

Meshpads removes liquid droplets by impingement of droplets onto the wire by coalescing them into larger droplets. These larger and heavier droplets will subsequently disengage and drop to the bottom leaving drier gas moving out of the vessel.

In this work, the designed velocity for horizontal style 4CA mesh pad was obtained as 0.27m/s with a cross sectional area of 0.19m^2.

The separation efficiency of 0.09m thickness of the thick element of the mesh was 60% for the removal of 5μm droplet size and 95% for the removal of 10 μm droplet size. Also the separation efficiency of 0.15m thickness of the thick element of the mesh was 79% for the removal of 5μm droplet size and 99% for the removal of 10 μm droplet sizes as depicts in Figure 6-3.

This shows that for the effective removal of 5μm droplet size and above, a 0.15m thick element of the mesh should be used.

9 Conclusion

The most important gas/liquid separations that take place in oil and gas processing fields operation are; gas-liquid and liquid separations. The conditions under which the separations have to take place and requirements are to be fulfilled. The present available separator types have to be sized with or without internals and evaluated with respect to the suitability to fulfill the separation requirements and perhaps in stages.

The number of stages in stage separation is actually determined by the form and the quantity of the liquids offered to the separator and the maximal amount of liquid quantity permitted in the outlet of the separator.

The effectiveness of stage separation resulted in the maximum stabilization of the resultant phases; gas, oil and water leaving the separator. The terminal settling velocities of the droplets increases in the separators as the pressure decreases in the different stages of the separators.

In the vertical three-phase separators with mist extractor (mesh pad), the retention time which is the effective time available for each phase droplets to be separated from the other phase of the heavy liquid (oil) is more than 100% higher than that of the vertical separator without mist extractor. Also the surge time which is the time the vessel can accommodate inlet flow rate if outgoing flow rate cuts off was greater in in vertical vessels without mist extractor than in the vessel with mesh.

In the horizontal three-phase separators with mist extractor (mesh pad), the surge time was less than that in horizontal vessels without mesh. The values are depicted in Tables 4-4, 4-5, 4-8 and 4-9 above.

The designed length of the three phase vertical separators without mesh was in the range of 1.5 to 6.0 at the different pressures showing that the design standard was acceptable. For vertical separators with mesh, only the separator at 80bar was in the acceptable range. Additional height of 1ft (0.3042) was added to the design height of the separators at 15 and 2 bar.

For the horizontal separators without mesh, the length/diameter ratios were obtained as 4.2, 3.9 and 2.9 for the separators at 80, 15 and 2bar respectively.

Also, for the horizontal separators without mesh, the length/diameter ratios were obtained as 5, 4 and 3 for the separators at 80, 15 and 2bar respectively.

The designed settling velocity for horizontal style 4CA mesh pad used in the horizontal separator at 80bar was 0.27m/s with a cross sectional area of $0.19m^2$. The separation capture efficiency of 0.09m thickness of the thick element of the mesh was 60% for the removal of 5µm droplet size and 95% for the removal of 10 µm droplet size.

Also the separation efficiency of 0.15m thickness of the thick element of the mesh was 79% for the removal of 5µm droplet size and 99% for the removal of 10 µm droplet sizes as depicts in Figure 6-3.

This shows that for the effective removal of 5μm droplet size and above, a 0.15m thick element of the mesh should be used.

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Appendix

Appendix A: Showing Tables

Table 1-1: Typical molar composition of petroleum reservoir fluid (Pederson 1989)

Typical Molar Compositions of Petroleum Reservoir Fluids

Component	Gas	Gas Condensate	Volatile Oil	Black Oil
N_2	0.3	0.71	1.67	0.67
CO_2	1.1	8.65	2.18	2.11
C_1	90.0	70.86	60.51	34.93
C_2	4.9	8.53	7.52	7.00
C_3	1.9	4.95	4.74	7.82
C_4 $(i+n)$	1.1	2.00	4.12	5.48
C_5 $(i+n)$	0.4	0.81	2.97	3.80
C_6 $(i+n)$	6+: 0.3	0.46	1.99	3.04
C_7		0.61	2.45	4.39
C_8		0.71	2.41	4.71
C_9		0.39	1.69	3.21
C_{10}		0.28	1.42	1.79
C_{11}		0.20	1.02	1.72
C_{12}		0.15	12+: 5.31	1.74
C_{13}		0.11		1.74
C_{14}		0.10		1.35
C_{15}		0.07		1.34
C_{16}		0.05		1.06
C_{17}		17+: 0.37		1.02
C_{18}				1.00
C_{19}				0.90
C_{20}				20+: 9.18

Table 3-1: Table 3-1: Fluid properties at first stage separation (80 bara) using Hysys

Name	Feed Fluid	Oil 1	Gas 1	Water 1
Vapour Fraction	0.62263	0.00000	1.00000	0.00000
Temperature [C]	77.214	77.214	77.214	77.214
Pressure [kPa]	8000.0	8000.0	8000.0	8000.0
Actual Vol. Flow [m3/h]	222.22	39.302	181.68	1.2467
Mass Enthalpy [kJ/kg]	-3284.5	-2260.7	-4010.4	-15604
Mass Entropy [kJ/kg-C]	3.9663	2.2483	7.1956	3.6717
Molecular Weight	39.938	82.306	21.676	18.024
Molar Density [kgmole/m3]	4.2027	7.2606	3.2008	53.808
Mass Density [kg/m3]	167.85	597.59	69.380	969.81
Std. Liquid Mass Density [k	<empty>	664.84	<empty>	1014.9
Molar Heat Capacity [kJ/kg	102.56	203.14	56.022	78.146
Mass Heat Capacity [kJ/kg	2.5680	2.4681	2.5845	4.3358
Thermal Conductivity [W/m	<empty>	9.1313e-002	4.3873e-002	0.66784
Viscosity [cP]	<empty>	0.23008	1.5536e-002	0.36366
Surface Tension [dyne/cm]	<empty>	9.7604	<empty>	62.827
Specific Heat [kJ/kgmole-C	102.56	203.14	56.022	78.146
Z Factor	<empty>	0.37824	0.85801	5.1038e-002
Vap. Frac. (molar basis)	0.62263	0.00000	1.0000	0.00000
Vap. Frac. (mass basis)	0.33793	0.00000	1.0000	0.00000
Vap. Frac. (Volume Basis)	0.47829	0.00000	1.0000	0.00000
Molar Volume [m3/kgmole]	0.23794	0.13773	0.31243	1.8585e-002
Act. Gas Flow [ACT_m3/h]	<empty>	<empty>	181.68	<empty>
Act. Liq. Flow [m3/s]	<empty>	1.0917e-002	<empty>	3.4632e-004
Std. Liq. Vol. Flow [m3/h]	80.213	35.326	<empty>	1.1913
Std. Gas Flow [STD_m3/h]	22082	6747.0	13749	1586.2
Watson K	14.186	12.963	17.109	8.5110
Kinematic Viscosity [cSt]	<empty>	0.38501	0.22393	0.37498
Cp/Cv	1.0882	1.2383	1.4524	1.1656
Lower Heating Value [kJ/k	1.7241e+006	3.6811e+006	9.6261e+005	3.4536
Mass Lower Heating Value	43168	44725	44409	0.19161
Liquid Fraction	0.37737	1.0000	0.00000	1.0000
Partial Pressure of CO2 [kP	205.98	0.00000	205.98	0.00000
Avg. Liq. Density [kgmole/r	12.949	7.8361	16.857	55.362

Table 3-2: Fluid properties at second-stage separation (15 bar)

Name	Oil 2	Oil 3	Gas 2	Water
Vapour Fraction	0.32348	0.00000	1.00000	0.0000
Temperature [C]	69.681	69.681	69.681	69.68
Pressure [kPa]	1500.0	1500.0	1500.0	1500.
Actual Vol. Flow [m3/h]	198.32	32.070	166.25	0.0000
Mass Enthalpy [kJ/kg]	-2260.7	-2107.0	-3610.0	-1564
Mass Entropy [kJ/kg-C]	2.3089	1.8136	6.6568	3.583
Molecular Weight	82.306	109.22	26.022	18.02
Molar Density [kgmole/m3]	1.4389	6.0196	0.55522	45.64
Mass Density [kg/m3]	118.43	657.44	14.448	822.6
Std. Liquid Mass Density [k	<empty>	706.64	<empty>	1014.
Molar Heat Capacity [kJ/kg	191.03	256.13	54.885	78.16
Mass Heat Capacity [kJ/kg	2.3210	2.3452	2.1092	4.337
Thermal Conductivity [W/m	<empty>	0.10447	3.3828e-002	0.6620
Viscosity [cP]	<empty>	0.34342	1.2751e-002	0.4022
Surface Tension [dyne/cm]	<empty>	14.359	<empty>	64.20
Specific Heat [kJ/kgmole-C	191.03	256.13	54.885	78.16
Z Factor	<empty>	8.7422e-002	0.94780	1.1529e-00
Vap. Frac. (molar basis)	0.32348	0.00000	1.0000	0.0000
Vap. Frac. (mass basis)	0.10227	0.00000	1.0000	0.0000
Vap. Frac. (Volume Basis)	0.16508	0.00000	1.0000	0.0000
Molar Volume [m3/kgmole]	0.69499	0.16612	1.8011	2.1908e-00
Act. Gas Flow [ACT_m3/h]	<empty>	<empty>	166.25	<empty
Act. Liq. Flow [m3/s]	8.9084e-003	8.9084e-003	<empty>	0.0000
Std. Liq. Vol. Flow [m3/h]	35.326	29.837	<empty>	0.0000
Std. Gas Flow [STD_m3/h]	6747.0	4564.5	2182.5	0.0000
Watson K	12.963	12.760	16.401	8.527
Kinematic Viscosity [cSt]	<empty>	0.52236	0.88257	0.4889
Cp/Cv	1.0455	1.0335	1.2357	1.167
Lower Heating Value [kJ/k	3.6811e+006	4.8884e+006	1.1562e+006	1.008
Mass Lower Heating Value	44725	44758	44431	5.5981e-00
Liquid Fraction	0.67652	1.0000	0.00000	1.000
Partial Pressure of CO2 [kP	48.587	0.00000	48.587	0.0000
Avg. Liq. Density [kgmole/r	7.8361	6.3495	15.355	55.37

Table 3-3: Fluid properties at the third-stage separation (2 bar)

Name	Oil 4	Oil 5	Gas 3	Water
Vapour Fraction	0.16576	0.00000	1.00000	0.0000
Temperature [C]	60.957	60.957	60.957	60.95
Pressure [kPa]	200.00	200.00	200.00	200.0
Actual Vol. Flow [m3/h]	463.14	28.926	434.21	0.0000
Mass Enthalpy [kJ/kg]	-2107.0	-2075.0	-2552.4	-1568
Mass Entropy [kJ/kg-C]	1.8315	1.6619	4.1900	3.473
Molecular Weight	109.22	122.13	44.208	18.01
Molar Density [kgmole/m3]	0.41683	5.5676	7.3697e-002	45.94
Mass Density [kg/m3]	45.525	680.00	3.2580	827.9
Std. Liquid Mass Density [k	698.98	718.38	508.18	1014.
Molar Heat Capacity [kJ/kg	245.61	277.83	83.465	78.01
Mass Heat Capacity [kJ/kg	2.2488	2.2747	1.8880	4.329
Thermal Conductivity [W/m	<empty>	0.11344	2.3485e-002	0.6544
Viscosity [cP]	<empty>	0.43514	9.9740e-003	0.4564
Surface Tension [dyne/cm]	<empty>	17.130	<empty>	65.77
Specific Heat [kJ/kgmole-C	245.61	277.83	83.465	78.01
Z Factor	<empty>	1.2931e-002	0.97694	1.5669e-00
Vap. Frac. (molar basis)	0.16576	0.00000	1.0000	0.0000
Vap. Frac. (mass basis)	6.7095e-002	0.00000	1.0000	0.0000
Vap. Frac. (Volume Basis)	9.0888e-002	0.00000	1.0000	0.0000
Molar Volume [m3/kgmole]	2.3991	0.17961	13.569	2.1764e-00
Act. Gas Flow [ACT_m3/h]	<empty>	<empty>	434.21	<empty
Act. Liq. Flow [m3/s]	8.0350e-003	8.0350e-003	<empty>	0.0000
Std. Liq. Vol. Flow [m3/h]	30.164	27.380	2.7837	0.0000
Std. Gas Flow [STD_m3/h]	4564.5	3807.9	756.63	0.0000
Watson K	12.760	12.699	14.460	8.527
Kinematic Viscosity [cSt]	<empty>	0.63991	3.0614	0.5512
Cp/Cv	1.0350	1.0308	1.1218	1.164
Lower Heating Value [kJ/k	4.8884e+006	5.4641e+006	1.9911e+006	0.2162
Mass Lower Heating Value	44758	44738	45039	1.2003e-00
Liquid Fraction	0.83424	1.0000	0.00000	1.000
Partial Pressure of CO2 [kP	4.5731	0.00000	4.5731	0.0000
Avg. Liq. Density [kgmole/r	6.3495	5.8266	11.580	55.38

Table 3-4: Summary of some fluid properties at first-stage (80 bar), second-stage (15 bar) and third-stage (2 bar) separation using Hysys

Parameters	80 bar	15 bar	2 bar
σ_G	-	-	-
σ_O	9.76 dyne/cm = 0.00976 N/m	14.36 dyne/cm = 0.01436 N/m	17.13 dyne/cm = 0.01713 N/m
σ_W	62.83 dyne/cm = 0.06283 N/m	64.20 dyne/cm = 0.06420 N/m	65.78 dyne/cm = 0.06578 N/m

ρ_G ρ_O ρ_W	69.38 kg/m^3 (4.33 Ib/ft^3) 597.59 kg/m^3 (37.30 Ib/ft^3) 969.81 kg/m^3 (60.54 Ib/ft^3)	14.45 kg/m^3 (0.90 Ib/ft^3) 657.44 kg/m^3 (41.04 Ib/ft^3) 822.60 kg/m^3 (51.35 Ib/ft^3)	3.26 kg/m^3 (0.20 Ib/ft^3) 680.0 kg/m^3 (42.43 Ib/ft^3) 827.95 kg/m^3 (51.68 Ib/ft^3)
P	80 bar = 8000 kPa (1161 psia)	15 bar = 1500 kPa (217 psia)	2 bar = 200 kPa (29 psia)
T	77˚C = 350K (626˚R)	70˚C = 343K (613˚R)	61˚C = 334K (598˚R)
z_G z_O z_W	0.86 0.38 0.051	0.95 0.087 0.012	0.98 0.013 0.0016

μ_G μ_O μ_W	1.56E-02cp = 1.56-05Pas 2.30E-01cp = 2.30-04Pas 3.36E-01cp = 3.36-04Pas	1.30E-02cp = 1.30E-05Pas 3.40E-0cp = 3.40-04Pas 4.00E-01cp = 4.00-04Pas	1.0E-02cp = 1.0E-05Pas 4.40E-01cp = 4.40-04Pas 4.60E-01cp = 4.60-04Pas
q_G q_O q_W	181.68m^3/h = 0.05m^3/s 39.30m^3/h = 0.011m^3/s 1.25m^3/h = 0.00035m^3/s	166.25m^3/h = 0.046m^3/s 32.07m^3/h = 0.009m^3/s 0.72m^3/h = 0.00020m^3/s	434.21m^3/h = 0.12m^3/s 28.93m^3/h = 0.008m^3/s 0.54m^3/h = 0.00015m^3/s

Table 4-4: Showing design calculations and results for a three phase vertical separator with no mist eliminator using fluid property in Table 3-4

Input and notes	Calculations and results at 80 bar	Calculations and results at 15 bar	Calculations and results at 2 bar

From Hysys			
ρ_G	69 kg/m^3	14 kg/m^3	3.26 kg/m^3
ρ_O (ρ_{LL})	598 kg/m^3	657 kg/m^3	680 kg/m^3
ρ_W (ρ_{HL})	970 kg/m^3	823 kg/m^3	828 kg/m^3
μ_G	1.56-05Pas	1.30-05Pas	1.00-05Pas
μ_O (μ_{LL})	2.30-04Pas	3.40-04Pas	4.40-04Pas
μ_W (μ_{HL})	3.36-04Pas	4.00-04Pas	4.6-04Pas
q_G	0.05m^3/s	0.046m^3/s	0.012m^3/s
q_O (q_{LL})	0.011m^3/s	0.009m^3/s	0.0008m^3/s
q_W (q_{HL})	0.000035m^3/s	0.0002m^3/s	0.00015m^3/s
From GPSA value in Table 4-2 k_s = 0.3 – 0.001(P-100) *P in Psig*	k_s = 0.3 – 0.001(1160 -100)= 0.244ft/s (0.074m/s)	k_s = 0.3–0.001(217-100)=0.338ft/s (0.103m/s)	k_s = 0.3–0.001(29-100)=0.357ft/s (0.109m/s)
k_s/2 (no mist eliminator)	0.074/2 =0.037m/s	0.103/2 = 0.052m/s	0.109/2 =0.055m/s

$u_D = k_s\sqrt{\frac{\rho_L-\rho_G}{\rho_G}}$	$0.037\sqrt{\frac{598-69}{69}}$ = 0.102m/s	$0.052\sqrt{\frac{657-14}{14}}$ = 0.35m/s	$0.055\sqrt{\frac{680-3}{3.2}}$ = 0.75m/s
$u_G = \frac{2}{3}u_D$	$\frac{2}{3}(0.102) =$ 0.068m/s	$\frac{2}{3}(0.35) = 0.023\text{m/s}$	$\frac{2}{3}(0.75) = 0.49\text{m/s}$
$D_I = \left(\frac{4q_G}{\pi u_G}\right)^{1/2}$ Where $D_I = D$	$\left(\frac{4*0.05}{3.142*0.068}\right)^{\frac{1}{2}}$ = 0.97m Using D_I = 2.0m	$\left(\frac{4*0.046}{3.142*0.23}\right)^{\frac{1}{2}}$ = 0.51m Using D_I =2.0m	$\left(\frac{4*0.12}{3.142*0.49}\right)^{\frac{1}{2}}$ = 0.56m Using D_I =2.0m
$u_{HL} =$ $\frac{k_s(\rho_{HL}-\rho_{LL})}{\mu_{LL}}$	$\frac{0.037(970-598)}{2.30}10^{-4}$ = 0.0006m/s	$\frac{0.052(823-657)}{3.4}10^{-4}$ 0.00025m/s	$\frac{0.055(828-680)}{4.4}10^{-4}$ = 0.00019m/s
$u_{LL} =$ $\frac{k_s(\rho_{HL}-\rho_{LL})}{\mu_{HL}}$	$\frac{0.037(970-598)}{3.36}10^{-4}$ = 0.0004m/s	$\frac{0.052(823-657)}{4.0}10^{-4}$ = 0.00022m/s	$\frac{0.055(828-680)}{4.6}10^{-4}$ = 0.00018m/s
$t_s, H_L = \frac{H_L}{u_{HL}}$	$\frac{0.3042}{0.0006}$ = 507s (8.45min)	$\frac{0.3042}{0.00025}$ = 1217s	$\frac{0.3042}{0.00019}$ = 1601s

		(20min)	(27min)
$t_s, L_L = \frac{H_H}{u_{LL}}$	$\frac{0.3042}{0.0004}$ = 742s (12.4min)	$\frac{0.3042}{0.00022}$ = 1383s (23min)	$\frac{0.3042}{0.00018}$ = 1690s (28min)
$A = \frac{\pi D^2}{4}$	$\frac{3.142*2^2}{4}$ = 3.14m^2	$\frac{3.142*2^2}{4}$ = 3.14m^2	$\frac{3.142*2^2}{4}$ = 3.14m^2
$\frac{A_D}{A}$ (Using equation 4-13)	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2}$, $\frac{A_D}{A}$ = 0.0194	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2}$, $\frac{A_D}{A}$ = 0.0194	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2}$, $\frac{A_D}{A}$ = 0.0194
$A_D = \frac{A_D}{A}(A)$	0.0194(3.14) = 0.06m^2	0.0194(3.14) = 0.06m^2	0.0194(3.14) = 0.06m^2
$A_L = A - A_D$	3.14 – 0.06= 3.08m^2	3.14 – 0.06= 3.08m^2	3.14 – 0.06= 3.08m^2
$t_{R,LL} = \frac{H_L\, A_L}{q_{LL}}$	$\frac{0.3042*3.08}{0.011}$ = 85s (1.4min)	$\frac{0.3042*3.08}{0.009}$ = 104s (1.7min)	$\frac{0.3042*3.08}{0.008}$ = 117s (1.95min)
$t_{R,HL} = \frac{H_H\, A_H}{q_{HL}}$ where	$\frac{0.3042*3.14}{0.00035}$ = 2731s (46min) since $t_{R,HL} > t_s, L_L$,	$\frac{0.3042*3.14}{0.00020}$ = 4779s (80min)	$\frac{0.3042*3.14}{0.00015}$ = 6372s (106min)

$A_H = A$	proceeding to the next step	since $t_{R,HL} > t_S, L_L$, proceeding to the next step	since $t_{R,HL} > t_S, L_L$, proceeding to the next step
$H_R = \frac{q_{LL}\, t_H}{A_L}$	$\frac{0.011*300}{3.08}$ = 1.07m	$\frac{0.009*300}{3.08}$ = 0.88m	$\frac{0.008*300}{3.08}$ = 0.78m
H_S = $\left(\frac{t_s(q_{LL} + q_{HL})}{A}\right)$	$\frac{180(0.011 + 0.00035)}{3.14}$ = 0.65m	$\frac{180(0.009 + 0.0002)}{3.14}$ = 0.53m	$\frac{180(0.008 + 0.00015)}{3.14}$ = 0.47m
$H_{BN} = H_S$ +0.15m	0.65 + 0.15 = 0.8m	0.53 + 0.15 = 0.68m	0.47 + 0.15 = 0.62m
$H_D = 0.5D$	0.5(2)=1.0m	0.5(2)=1.0m	0.5(2)=1.0m
H_T = $H_H + H_L + H_R + H_A + H_{BN} + H_D$	0.3042 + 0.3042 + 1.07 + 0.15 + 0.8 + 1.0 = 3.6m	0.3042 + 0.3042 + 0.88 + 0.15 + 0.68 + 1.0 = 3.3m	0.3042 + 0.3042 + 0.78 + 0.15 + 0.62 +1.0 = 3.2m
Checking;	3.6/2 = 1.80		

$H_T/_D$		3.3/2 = 1.65	3.2/2 = 1.60

Table 4-5: Showing design calculations and results for a three phase vertical separator with mist eliminator using fluid property in Table 3-4

Input and notes	Calculations and results at 80 bar	Calculations and results at 15 bar	Calculations and results at 2 bar
$k_s$from Table 4-5 above	0.074m/s	0.103m/s	0.109m/s
$u_D = k_s\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$	$0.074\sqrt{\frac{598-69}{69}}$ = 0.205m/s	$0.103\sqrt{\frac{657-14}{14}}$ = 0.70m/s	$0.109\sqrt{\frac{680-3}{3.26}}$ = 1.57m/s
$u_G = \frac{2}{3}u_D$	$\frac{2}{3}(0.205)$= 0.14m/s	$\frac{2}{3}(0.70)$ = 0.47m/s	$\frac{2}{3}(1.57)$ = 1.05m/s
$D_I = D$+0.15m	2.0m + 0.15m =2.15m	2.0m + 0.15m =2.15m	2.0m + 0.15m =2.15m
$u_{HL} = \frac{k_s(\rho_{HL} - \rho_{LL})}{\mu_{LL}}$	$\frac{0.074(970-598)}{2.30}10^{-4}$	$\frac{0.103(823-657)}{3.4}10^{-4}$	$\frac{0.109(828-680)}{4.4}10^{-4}$

	= 0.0012m/s	0.0005m/s	= 0.00037m/s
u_{LL} = $\frac{k_s(\rho_{HL}-\rho_{LL})}{\mu_{HL}}$	$\frac{0.074(970-598)}{3.36}10^{-4}$ = 0.00082m/s	$\frac{0.103(823-657)}{4.0}10^{-4}$ = 0.00043m/s	$\frac{0.109(828-680)}{4.6}10^{-4}$ = 0.00035m/s
$t_s, H_L = \frac{H_L}{u_{HL}}$	$\frac{0.3042}{0.0012}$ = 254s (4min)	$\frac{0.3042}{0.0005}$ = 608s (10min)	$\frac{0.3042}{0.00037}$ = 822s (13.7min)
$t_s, L_L = \frac{H_H}{u_{LL}}$	$\frac{0.3042}{0.00082}$ = 371s (6.2min)	$\frac{0.3042}{0.00043}$ = 707s (11.8min)	$\frac{0.3042}{0.00035}$ = 869s (14.5min)
$A = \frac{\pi D^2}{4}$	$\frac{3.142*2.15^2}{4}$ = 3.6m^2	$\frac{3.142*2.15^2}{4}$ = 3.6m^2	$\frac{3.142*2.15^2}{4}$ = 3.6m^2
$\frac{A_D}{A}$ (Using equation 4-13)	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2.15}$, $\frac{A_D}{A}$ = 0.00829	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2.15}$, $\frac{A_D}{A}$ = 0.00829	Replacing $\frac{W_D}{D}$ with $\frac{0.1061}{2.15}$, $\frac{A_D}{A}$ = 0.00829
$A_D = \frac{A_D}{A}(A)$	0.000829(3.6) = 0.03m^2	0.000829(3.6) = 0.03m^2	0.000829(3.6) = 0.03m^2
$A_L = A - A_D$			

	3.6 – 0.03= 3.57m²	3.6 – 0.03= 3.57m²	3.6 – 0.03= 3.57m²
$t_{R,LL} = \frac{H_L\, A_L}{q_{LL}}$	$\frac{0.3042*3.57}{0.011}$ = 99s (1.6min)	$\frac{0.3042*3.57}{0.009}$ = 121s (2min)	$\frac{0.3042*3.57}{0.008}$ = 136s (2.3min)
$t_{R,HL} = \frac{H_H\, A_H}{q_{HL}}$ where $A_H = A$	$\frac{0.3042*3.6}{0.00035}$ = 3152s (52min) since $t_{R,HL} > t_s, L_L$, proceeding to the next step	$\frac{0.3042*3.6}{0.00020}$ = 5476s (91min) since $t_{R,HL} > t_s, L_L$, proceeding to the next step	$\frac{0.3042*3.6}{0.00015}$ = 7301s (122min) since $t_{R,HL} > t_s, L_L$, proceeding to the next step
$H_R = \frac{q_{LL}\, t_H}{A_L}$	$\frac{0.011*300}{3.57}$ = 0.92m	$\frac{0.009*300}{3.57}$ = 0.76m	$\frac{0.008*300}{3.57}$ = 0.67m
$H_S = \left(\frac{t_s(q_{LL} + q_{HL})}{A}\right)$	$\frac{180(0.011 + 0.00035)}{3.6}$ = 0.57m	$\frac{180(0.009 + 0.0002)}{3.6}$ = 0.46m	$\frac{180(0.008 + 0.00015)}{3.6}$ = 0.41m
$H_{BN} =$ H_S +0.15m	0.57 + 0.15 = 0.72m	0.46 + 0.15 = 0.61m	0.41 + 0.15 = 0.56m
$H_D = 0.5D$ $H_T =$	0.5(2.15) = 1.08	0.5(2.15) = 1.08	0.5(2.15) = 1.08

$H_H + H_L +$ $H_R + H_A +$ $H_{BN} + H_D$ Checking; H_T/D	0.3042 + 0.3042 + 0.92 + 0.15 + 0.72 + 1.08 = 3.48m 3.48/2.15 = 1.62	0.3042 + 0.3042 + 0.76 + 0.15 + 0.61 + 1.08 = 3.21m 3.21/2.15 = 1.49 Adding 1ft (0.3042m) to H_T 3.21+0.3042 = 3.5m Check; 3.5/2.15 =1.63m	0.3042 + 0.3042 + 0.67 + 0.15 + 0.56 +1.08 = 3.07m 3.07/2.15 = 1.43 Adding 1ft (0.3042m) to H_T 3.07+0.3042=3.37m Check; 3.37/2.15 =1.57m

Table 4-8: Design calculations and results for a three phase horizontal separator with no mist eliminator using fluid property in Table 3-4

Input and notes	Calculations and results at 80 bar	Calculations and results at 15 bar	Calculations and results at 2 bar

$u_D = k_s\sqrt{\frac{\rho_L-\rho_G}{\rho_G}}$	$0.0305\sqrt{\frac{598-69}{69}}$ = 0.085m/s	$0.0305\sqrt{\frac{657-14}{14}}$ = 0.206m/s	$0.0305\sqrt{\frac{680-3.26}{3.26}}$ = 0.439m/s
$u_G = \frac{2}{3}u_D$	$\frac{2}{3}(0.085)=$ 0.057m/s	$\frac{2}{3}(0.206) =$ 0.138m/s	$\frac{2}{3}(0.439) =$ 0.294m/s
$V_H = t_H\, q_{LL}$	10*60*0.011=6.6m^3	10*60*0.009=5.4m^3	10*60*0.008=4.8m^3
$V_S = t_S\, q_{HL}$	5*60*0.00035 =0.105m^3	5*60*0.00020 =0.060m^3	5*60*0.00015 =0.045m^3
$D =$ $\left(\frac{4(V_H+V_S)}{0.5\pi(^L/_D)}\right)^{1/3}$	$\left(\frac{4(6.6+0.105)}{0.5*3.142*4}\right)^{\frac{1}{3}}$ = 1.42m	$\left(\frac{4(5.4+0.06)}{0.5*3.142*3}\right)^{\frac{1}{3}}$ = 1.54m	$\left(\frac{4(4.8+0.45)}{0.5*3.142*3}\right)^{\frac{1}{3}}$ = 1.48m
$A_T = \frac{\pi D^2}{4}$	$\frac{3.142*1.42^2}{4}$ = 1.58m^2	$\frac{3.142*1.54^2}{4}$ = 1.86m^2	$\frac{3.142*1.48^2}{4}$ = 1.72m^2
$\frac{A_G}{A_T}$ (Using	Replacing $\frac{W_D}{D}$ with	Replacing $\frac{W_D}{D}$ with	Replacing $\frac{W_D}{D}$ with

equation 4-13)	$\frac{H_G}{D}(\frac{0.3042}{1.42})$, $\frac{A_G}{A_T} = 0.158$	$\frac{H_G}{D}(\frac{0.3042}{1.54})$, $\frac{A_G}{A_T} = 0.139$	$\frac{H_G}{D}(\frac{0.3042}{1.48})$, $\frac{A_G}{A_T} = 0.147$
$A_G = \frac{A_G}{A_T}(A_T)$	0.158(1.58) = 0.250m^2	0.139(1.86) = 0.259m^2	0.147(1.72) = 0.252m^2
$H_{LLL} = 0.5D$ +7 (in)	9.3in (0.24m)	9.5in (0.24m)	9.42in (0.24m)
$\frac{A_{LLL}}{A_T}$ (Using equation 4-13)	Replacing $\frac{W_D}{D}$ with $(\frac{0.24}{1.42})$, $\frac{A_{LLL}}{A_T} = 0.0083$	Replacing $\frac{W_D}{D}$ with $\frac{H_{LLL}}{D}(\frac{0.24}{1.54})$, $\frac{A_{LLL}}{A_T} = 0.075$	Replacing $\frac{W_D}{D}$ with $\frac{H_{LLL}}{D}(\frac{0.24}{1.48})$, $\frac{A_{LLL}}{A_T} = 0.078$
$A_{LLL} = \frac{A_{LLL}}{A_T}(A_T)$	0.0083(1.58) = 0.013m^2	0.075(1.86) = 0.14m^2	0.078(1.72) = 0.13m^2
$H_w = D - H_G$	1.42 – 0.3042 = 1.12m	1.54 – 0.3042 = 1.24m	1.48 – 0.3042 = 1.18m
$L_2 = \frac{V_H + V_S}{A_T - A_G - A_{LLL}}$	$\frac{6.6 + 0.15}{1.58 - 0.25 - 0.013}$ = 5.09m ~ 5m	$\frac{5.4 + 0.060}{1.86 - 0.259 - 0.14}$ = 3.7m ~ 4m	$\frac{4.8 + 0.45}{1.72 - 0.252 - 0.13}$ = 3.62m ~ 4m
$H_{HL} = H_{LL} = H_W/_2$			

Table 4-9: Design calculations and results for a three phase horizontal separator with mist eliminator using fluid property in Table 3-4

Input and notes	Calculations and results at 80 bar	Calculations and results at 15 bar	Calculations and results at 2 bar
$u_D = k_s\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$	$0.137\sqrt{\frac{598-69}{69}}$ $= 0.379 m/s$	$0.137\sqrt{\frac{657-14}{14}}$ $= 0.928 m/s$	$0.137\sqrt{\frac{680-3.26}{3.26}}$ $= 1.97 m/s$
$u_G = \frac{2}{3} u_D$	$\frac{2}{3}(0.379)=$ $0.253 m/s$	$\frac{2}{3}(0.928) =$ $0.619 m/s$	$\frac{2}{3}(1.97) = 1.31 m/s$
$V_H = t_H\, q_{LL}$	10*60*0.011=6.6m^3	10*60*0.009=5.4m^3	10*60*0.008=4.8m^3
$V_S = t_S\, q_{HL}$	5*60*0.00035 =0.105m^3	5*60*0.00020 =0.060m^3	5*60*0.00015 =0.045m^3
$D =$			

$\left(\frac{4(V_H+V_S)}{0.5\pi(L/D)}\right)^{1/3}$	$\left(\frac{4(6.6+0.105)}{0.5*3.142*4}\right)^{\frac{1}{3}}$ = 1.42m	$\left(\frac{4(5.4+0.06)}{0.5*3.142*3}\right)^{\frac{1}{3}}$ = 1.54m	$\left(\frac{4(4.8+0.45)}{0.5*3.142*3}\right)^{\frac{1}{3}}$ = 1.48m
$A_T = \frac{\pi D^2}{4}$	$\frac{3.142*1.42^2}{4}$ = 1.58m²	$\frac{3.142*1.54^2}{4}$ = 1.86m²	$\frac{3.142*1.48^2}{4}$ = 1.72m²
$\frac{A_G}{A_T}$ (Using equation 4-13)	Replacing $\frac{W_D}{D}$ with $\frac{H_G}{D}(\frac{0.6048}{1.42})$ $\frac{A_G}{A_T} = 0.357$	Replacing $\frac{W_D}{D}$ with $\frac{H_G}{D}(\frac{0.6048}{1.54})$ $\frac{A_G}{A_T} = 0.473$	Replacing $\frac{W_D}{D}$ with $\frac{H_G}{D}(\frac{0.6048}{1.48})$ $\frac{A_G}{A_T} = 0.315$
$A_G = \frac{A_G}{A_T}(A_T)$	0.357(1.58) = 0.564m²	0.473(1.86) = 0.88m²	0.315(1.72) = 0.54m²
$H_{LLL} = 0.5D + 7$ (in)	9.3in (0.24m)	9.5in (0.24m)	9.42in (0.24m)
$\frac{A_{LLL}}{A_T}$ (Using equation 4-	Replacing $\frac{W_D}{D}$ with $(\frac{0.24}{1.42})$,	Replacing $\frac{W_D}{D}$ with $\frac{H_{LLL}}{D}(\frac{0.24}{1.54})$,	Replacing $\frac{W_D}{D}$ with $\frac{H_{LLL}}{D}(\frac{0.24}{1.48})$,

13)	$\frac{A_{LLL}}{A_T}$ = 0.0083	$\frac{A_{LLL}}{A_T}$ = 0.075	$\frac{A_{LLL}}{A_T}$ = 0.078
$A_{LLL} = \frac{A_{LLL}}{A_T}(A_T)$	0.0083(1.58) = 0.013m^2	0.075(1.86) = 0.14m^2	0.078(1.72) = 0.13m^2
$H_w = D - H_G$	1.42 – 0.6084 = 0.81m	1.54 – 0.6084 = 0.93m	1.48 – 0.6084 = 0.87m
$L_2 = \frac{V_H + V_S}{A_T - A_G - A_{LLL}}$	$\frac{6.6 + 0.105}{1.58 - 0.56 - 0.013}$ = 6.68m ~ 7m	$\frac{5.4 + 0.060}{1.86 - 0.88 - 0.14}$ = 6.5m ~ 7m	$\frac{4.8 + 0.45}{1.72 - 0.54 - 0.13}$ = 4.6m ~ 5m
$H_{HL} = H_{LL} = H_W/_2$	0.81/2 = 0.41m	0.93/2 = 0.47m	0.87/2 = 0.44m
$\frac{A_{HL}}{A_T}$ (Using equation 4-13)	Replacing $\frac{W_D}{D}$ with $(\frac{0.41}{1.58})$	Replacing $\frac{W_D}{D}$ with $(\frac{0.47}{1.86})$	Replacing $\frac{W_D}{D}$ with $(\frac{0.44}{1.72})$
$A_{HL} = \frac{A_{HL}}{A_T}(A_T)$	$\frac{A_{HL}}{A_T}$ = 0.205	$\frac{A_{HL}}{A_T}$ = 0.199	$\frac{A_{HL}}{A_T}$ = 0.249

$u_{HL} = \frac{k_s(\rho_{HL}-\rho_{LL})}{\mu_{LL}}$	$0.205(1.58) =$ $0.32m^2$	$0.199(1.86) =$ $0.370m^2$	$0.249(1.72) =$ $0.428m^2$
$u_{LL} = \frac{k_s(\rho_{HL}-\rho_{LL})}{\mu_{HL}}$	$\frac{0.137(970-598)}{2.30}10^{-4}$ $= 0.0022m/s$	$\frac{0.137(823-657)}{3.40}10^{-4}$ $0.00067m/s$	$\frac{0.137(828-680)}{4.4}10^{-4}$ $= 0.00046m/s$
$t_s, H_L = \frac{H_{LL}}{u_{HL}}$	$\frac{0.137(970-598)}{3.36}10^{-4}$ $= 0.0015m/s$	$\frac{0.137(823-657)}{4.0}10^{-4}$ $= 0.00057m/s$	$\frac{0.137(828-680)}{4.6}10^{-4}$ $= 0.00044m/s$
$t_s, L_L = \frac{H_{HL}}{u_{LL}}$	$\frac{0.41}{0.0022} = 186s$ (3min)	$\frac{0.47}{0.00067} = 701s$ (11.7min)	$\frac{0.44}{0.00046} = 957s$ (16min)
$L_1 = \frac{t_{S,LL}q_{HL}}{A_{HL}}$	$\frac{0.41}{0.0015} = 273s$ (4.5min)	$\frac{0.47}{0.00057} = 825s$ (13.7min)	$\frac{0.44}{0.00046} = 957s$ (16.7min)
$L = L_2 + L_1$	$\frac{273 * 0.00035}{0.32}$ $= 0.30m$	$\frac{825 * 0.00020}{0.370}$ $= 0.45m$	$\frac{1000 * 0.00015}{0.428}$ $= 0.35m$

$t_{DL}=\frac{H_G}{u_G}$	7+0.30 = 7.3m~7m	7+0.45= 7.45m~7m	5+0.35= 5.35m~5m
$u_{GA}=\frac{q_G}{A_G}$	$\frac{0.6084}{0.253}$ = 2.4s	$\frac{0.6084}{0.619}$ = 0.98s	$\frac{0.6084}{1.31}$ = 0.46s
$L_{MIN}=u_{GA}t_{DL}$	$\frac{0.05}{0.564}$ = 0.089m/s	$\frac{0.046}{0.88}$ = 0.052m/s	$\frac{0.012}{0.54}$ = 0.22m/s
L/D	0.089*2.4 = 0.2m $L > L_{MIN}$ (acceptable for vapour-liquid separation)	0.052*0.98 = 0.05m	0.22*0.46 = 0.10m
Checking; L/D	7/1.42= 4.9	7/1.54= 4.5	5/1.72= 2.9
From AMSE (1986) E=0.85, S=17,500psi (1207bar), W_C= 0.0625in	Acceptable range of 1.5 to 6.0	Acceptable range of 1.5 to 6.0	Acceptable range of 1.5 to 6.0

(0.00159m) $t_{SH} = \frac{PD}{2SE - 1.2P} + W_C$ $t_{HD} = \frac{PD}{2SE - 0.2P} + W_C$ $A_{SH} = \pi DL$ $A_{HD} = 1.09D^2$ W (Using equation 4-20)	P = 80+8(10%) = 88bar $\frac{88*1.42}{2*1207*0.85-1.2*88}$ +0.00159 = 0.0653m Using 2:1 elliptical head $\frac{88*1.42}{2*1207*0.85-0.2*88}$ +0.00159 = 0.063m 3.142*1.42*7 = 31m^2 (334ft^2) 1.09*(1.42)^2 = 2.2m^2(ft^2)	P = 15+3(30psig) = 18bar $\frac{18*1.54}{2*1207*0.85-1.2*18}$ +0.00159 = 0.0152m Using 2:1 elliptical head $\frac{18*1.54}{2*1207*0.85-0.2*18}$ +0.00159 = 0.0029m 3.142*1.54*7 = 31m^2(334ft^3) 1.09*(1.54)^2 =	P = 2+3(30psig) = 5bar $\frac{5*1.48}{2*1207*0.85-1.2*5}$ +0.00159 = 0.0036m Using dished head $\frac{0.885*5*1.48}{1207*0.85-0.1*5}$ +0.00159 = 0.00634m 3.142*1.48*5 = 23m^2(247.6ft^2)

$H_{HLL} = D - H_G$ $A_{NLL} = A_{LLL} + \frac{V_H}{L_2}$ $\frac{H_{NLL}}{A_T}$ (Using equation 4-13) $H_{NLL} = \frac{A_{NLL}}{A_T}(A_T)$	490*$\frac{2.57}{12}$(334+2*24) = 40025Ib (18115kg) 1.42-0.6084 = 0.81m 0.013+$\frac{6.6}{7}$ = 0.956m^2 Replacing $\frac{W_D}{D}$ with ($\frac{0.956}{1.42}$) $\frac{H_{NLL}}{A_T}$ = 0.715 0.715(1.58) = 1.13m	2.6m^2(28ft^2) 490*$\frac{5.92}{12}$(334+2*28) = 88950Ib (40358kg) 1.54 -0.6084 = 0.93m 0.14+$\frac{5.4}{7}$ = 0.91m^2 Replacing $\frac{W_D}{D}$ with ($\frac{0.91}{1.86}$) $\frac{H_{NLL}}{A_T}$ = 0.719 0.719(1.86) = 1.34m	0.842*(1.48)^2 = 1.8m^2(19.4ft^2) 490*$\frac{0.25}{12}$(248+2*19) = 2919Ib (1324kg) 1.48-0.6084 = 0.87m 0.13+$\frac{4.8}{6}$ = 0.93m^2 Replacing $\frac{W_D}{D}$ with ($\frac{0.93}{1.72}$) $\frac{H_{NLL}}{A_T}$ = 0.798

			0.798(1.72) = 1.37m

Appendix B : Development of terminal settling velocity (TSV) of droplet

The force of gravity F_G, buoyancy F_B and drag F_D on the droplet may be determined from the following equations:

$F_G = \rho_G V[N]$ (B-1)

$= \rho_L g V_d$ (B-2)

The force of buoyancy of the drop is

$F_B = \rho_G g V_d$ (B-3)

The drag force on the drop is

$F_D = \frac{1}{2} C_D A_D \rho_G u_D{}^2$ (B-4)

Where

F_D = drag force, lb_f, N

C_D = drag coefficient

A_d = cross-sectional area of the droplet, ft^2, m^2

ρ = density of the continuous phase, lb/ft^3, kg/m^3

u_D = terminal (settling velocity) of liquid droplet, ft/s, m/s

g = gravitational constant, 32.2 $lb_m ft/lb_f\, s^2$, m/s^2).

When the liquid drop acquires a steady speed defined as the settling speed, the drag force is just the same as gravity. This state could be described as;

$F_D = F_G - F_B$ (B-5)

That is;

$\frac{1}{2} C_D A_D \rho_G u_D{}^2 = gV_d(\rho_L - \rho_G)$ (B-6)

Where V_d is the volume of the drop given as;

$V_d = \frac{\pi D^3}{6}$ (B-7)

And $A = \frac{\pi D^2}{4}$ (B-8)

Therefore the terminal settling velocity (TSV) of a liquid droplet in a gas stream is given by;

$$u_D = \sqrt{\frac{4gd_D}{3\mathrm{C_D}}}\ \sqrt{\frac{\rho_L-\rho_G}{\rho_G}} \qquad \text{(B-9)}$$

Where $\sqrt{\frac{4gd_D}{3\mathrm{C_D}}} = k_s$ (B-10)

In practical situation in gas - liquid separation, TSV can be written as;

$$u_D = k_s\sqrt{\frac{\rho_L-\rho_G}{\rho_G}} \qquad \text{(B11)}$$

Where k_s is the separation constant; ranging between 0.05 to 0.11 as recommended by API.