Technical University of Crete | School of Mineral Resources Engineering

Basic Design of Oil Process Train in Upstream Facilities

Klothakis Andreas

Thesis for the MSc Program in Petroleum Engineering

Submission Date: October 2016

Supervisor: Dr. Dimitrios Marinakis

Supervisor Professor: Prof. Nikos Varotsis

Technical University of Crete
School of Mineral Resources Engineering

Abstract

The term "surface or production facilities" for an oilfield includes the necessary equipment for processing the production fluids from the wells into commodities. The core part of this process is the separation of the wellhead fluids into three distinct fluid phases namely, oil, water and gas. The separation train may consist of a single primary separation vessel or multi-stage separation vessels, each one operating at different conditions. The performance of the separation process is defined by the percentage of oil that is extracted from the wellhead fluids, as well as from the properties of the produced oil and gas with respect to the market specifications (RVP, BSW, H2S, water content, etc.). As the production of an oilfield changes with time, production wells are closed while new wells start producing, the separation conditions should adapt in order to achieve maximum performance at any given time. There are often cases in which surface separation facilities impose constraints on the production processes or they interfere with the total oil production output. The best way to avoid such constraints is through simulation of the upstream production facilities, including sensitivity and performance analysis.

This study examines the design and performance of a three-stage separation process at steady state conditions. The input flow to this process is the composite product of three wellhead fluids, each one with different hydrocarbon properties and watercuts. Using AspenTech's HYSYS simulation software, together with literature data and equipment standards, we manage to simulate the basic process flow diagram and test its performance over different numbers of separation stages for a predefined set of inflow conditions and flow rates. Temperature and pressure conditions at each separation stage were optimized according to the produced oil volumetric flow rate, under different production schemes – all wells producing simultaneously, one out of three wells shut in, start/end of production life for each well, etc.

Equipment specification and sizing is also a part of this study. Separator vessel sizing and internal configuration were prerequisites in order to monitor and control water carry over between the separation stages and the water content of the final stock tank oil. A short review on recent literature data for carry over estimation and technical information about separator internals are provided. HYSYS correlation tools were again employed to perform the water carry over calculations, to define the droplet size distribution in each separator and to design the internal configuration of each vessel. Calculation of the gas compressor sizes and specifications were included – as input data - in the optimization process.

Acknowledgments

This thesis concludes my Master of Science degree in Petroleum Engineering at the School of Mineral Resources Engineering, Technical University of Crete.

First of all, I owe my deepest gratitude to my supervisor Dr Marinakis Dimitrios. His continuous guidance throughout the course of the thesis, especially his technical input and his comments and suggestions were of crucial help in framing the report. Above all, I would like to thank him for his support to my personal problem and his encouragement which led me to the completion of this thesis.

A special word of appreciation is offered to my supervisor Professor Varotsis Nikos for providing me with this opportunity to work on this project and his guiding instructions through it. In addition, I appreciate the helpful examination of my work by Professor Kalogerakis Nikolas.

Moreover, I would like to acknowledge the support of Petroleum and Mechanical Engineering Department of Eastern Macedonia and Thrace Institute of Technology, which kindly provide me with access to ASPENTECH's HYSYS simulation software package. Without their support, it would have been impossible for me to successfully complete this study.

Finally, I am extremely grateful to the director of this Msc course, Professor Pasadakis Nikos, not only for his careful reading of my thesis but also for his respect to my personal problem and his significant help and support during my Msc studies.

Contents

1. Introduction

This MSc thesis simulates the oil and gas separation process of an oil production site with a common inlet from three oil-reservoir fields. Performance analysis and optimization are subsequently conducted to the simulation model with respect to possible changes in the inflow fluids.

A short introduction to surface facilities simulation and the motivation for this project is provided in the subsequent section of this Chapter. The formulation of the production problem of this study is described in subsection 1.2 of this Chapter, while the simulation software and the characteristics of the software's tools which were used in the simulation are presented in subsection 1.3.

In Chapter 2, a brief outline of the basic theoretical background of upstream production processes is presented.

In Chapter 3, the main input and assumptions of the simulation are presented. Multistage separation stage scenarios are examined and categorized based on their volumetric performance of oil production. Process conditions for gas and oil treatment are then simulated. According to them, detail design of first separator is performed.

In chapter 4, carry over calculation is inserted for real separation simulation. Literature and software information about the effect are provided. Droplet distribution and carry over is estimated in each separator. Internals design and separator sizing is completed.

In chapter 5, separated gas compression system is introduced. Compression equipment is designed.

Finally, in chapter 6, the study focuses on a detailed performance analysis of the system in order to indicate the sensitivity and the dependence of the process parameters (conditions) at the oil production performance.

1.1 Introduction to surface facilities simulation and motivation for this study

Simulation models have been extensively used in the downstream refining and chemical industry since the 1980s, reducing the engineering and operating costs, maximizing the production throughput, improving the reliability and increasing the profit margins. Despite the benefits of simulation in the oil downstream sector, the leverage from the use of this technology is not fully exploited in the oil upstream and midstream sectors.

The term "surface or production facilities" is used in the context of this study in order to describe the equipment and processes required to accomplish the separation and refinement of the oil wellhead fluids into oil, water and gas products, which are suitable for disposal to the market (oil, gas) and to the environment (water). All the auxiliary equipment, utility, metering, storage and export systems, is also included in the aforementioned term. The combination of the wellhead, the flow lines, bulk headers, valves and fittings needed to collect and transport the raw produced fluid to the production platform is referred as the gathering system.

The gathered fluids must be processed to standard commercial products in order to increase their value. Initially, wellhead fluids must be separated into the three main fluid phases namely, oil, water, and natural gas. This is a task for the separation process system. Then each fluid stream should be refined: The separated gas must be compressed and treated in order to meet the sales' contract specification. The oil and the water-in-oil emulsion from the separators must be treated to remove suspended water and other contaminants. The water from the separators must be treated to remove small quantities of produced oil and solid particles from it.

Traditionally, exploration and production (E&P) companies have focused on flourishing through the discovery and development of untapped reserves in order to increase their production rate. Today, the emphasis is more on maximizing the production from the tapped oil reserves. Reservoir modeling and Enhanced Oil Recovery are among the key techniques to achieve this goal.

Although the increase of the oil production from the reservoir is the main target for the oil companies, the improvement of upstream separation and production facilities should also be taken into account. There are often cases in which surface facilities impose constraints on the production processes and interfere with the oil production output. The best way to avoid such constraints is through simulation of the upstream production facilities, including sensitivity and performance analyses.

Owners and operators are increasingly using simulation models to support and optimize performance. Examples of such applications include steady-state and dynamic models leading to operation decisions, performance and equipment monitoring, offline and real-time optimization for improving planning models and for the assessment of the equipment lifecycle.

Process models become source for determining the design parameters for the plant. The use of them in conjunction with the real-time data gives the operator actionable information for development and planning the behavior of the infrastructure of production in order to face current and future needs, bearing in mind the operational and quality requirements.

Simulation models are used during the early stages of design to analyze the facilities performance under a variety of possible production scenarios as well as to predict the performance during shut in and start up stages. The results may indicate important process and equipment modifications for reducing the cost and operational risks in emergency situations. Besides, simulation may facilitate the training of the personnel to better understand of the interaction between the petroleum wells and the facilities on the platform.

1.2 Problem formulation

Based on literature data a production problem was formulated in order to create a stand-alone and steady state separation process simulation model. This model was used to analyze the performance of the equipment and optimize the oil production.

Stand-alone models reflect simulation problems with predefined inflow data as boundary conditions. Production decisions based exclusively on stand-alone model results may fail to fully exploit the potential value of the asset (from the reservoir to the market). On the other hand, an integrated model combines both reservoir, well, network surface, and process facilities simulations. Integrated model optimization offers more economic benefits (e.g., produces better economic outcomes), but it requires far more computational and time resources, including the risk of higher complexity results, in comparison to the stand-alone optimization. Due to the higher complexity of integrated models and the time constraints, it was decided to use only steady-state model in this study.

The combined effluent from three oil reservoirs was selected as inflow conditions to the upstream production facility. Details of the production fluids and of the flow rates for each reservoir were combined with crude oil product specifications for market, in order to define the boundary conditions of an efficient oil separation process train. The main specification was a final product Reid Vapor Pressure 10-12 psia and tank oil water content 0.5% (v/v). The design included mass/energy balance calculations, vessel details (sizing, fluid rates and conditions), compressor details, heating/cooling requirements. Internal configurations of any separation / treating vessel were also included in the study. Aspen HYSYS Process simulation software was used to conduct all the necessary calculations and form the model. The fluid and production characteristics from each reservoir are provided in Tables 1.1 and 1.2:

Component	Oil 1	Oil 2	Oil 3
Nitrogen	0,57	0,34	1,67
Carbon Dioxide	2,46	0,02	2,18
Methane	36,37	34,62	60,51
Ethane	3,47	4,11	7,52
Propane	4,05	1,01	4,74
i-Butane	0,59	0,76	0
n-Butane	1,34	0,49	4,12
i-Pentane	0,74	0,43	0
n-Pentane	0,83	0,21	2,97

Hexanes	1,62	1,16	1,38
Heptanes plus	47,96	56,85	14,91
Total	100	100	100
C_{7+} molecular weight	329	274	181
C_{7+} specific gravity	0,9594	0,92	0,799
Live oil molecular weight	171,4		46.69
Stock tank oil API gravity	19	23,6	47
Asphaltene content in stock tank oil, wt%	16,8		
Reservoir temperature, °F	212	203	246
Saturation pressure, psia	2950	2810	4677
GOR (SCF/STB)		300	2,909
initial oil FVF (RB/STB)		1.16	2.704

Table 1.1: Oil characteristics

	Estimated production BBL/D	**State 1**	**State 2**	**State 3**	**State 4**	**State 5**
Oil 1	Oil	3000	2800	2300	2000	1800
	Water	100	180	320	560	720
	Watercut	0,032258	0,060403	0,122137	0,21875	0,285714
Oil 2	Oil	6000	5700	5300	4800	4200
	Water	10	400	1000	1500	2000
	Watercut	0,001664	0,065574	0,15873	0,238095	0,322581
Oil 3	Oil	5000	4800	4500	4100	3600
	Water	0	100	400	800	1200
	Watercut	0	0,020408	0,081633	0,163265	0,25
Total	Watercut	0,007796	0,048641	0,124457	0,207849	0,289941

Table1.2: Production data

The scope of this thesis is to provide:

- Review of the latest literature on surface process facilities design
- Better comprehension of simulation, performance analysis and optimization procedure of upstream production facilities
- Better understanding of the interactions between the petroleum wells and the facilities as well as better understanding of the interaction between the production processes variables

- Familiarization with process simulation software for upstream facilities modeling.

1.3 Software Overview

Aspen HYSYS, developed by Aspen Technology, Inc., was the main software tool used to simulate the process train. Aspen HYSYS is a comprehensive process modeling tool used by the world's leading oil and gas producers, refineries, and engineering companies for process simulation and process optimization in design and operations. It offers the capability:

- To solve Mass and Energy Balance using in-built Mathematical Models.
- To obtain the Flow Rates, Compositions and Thermo Physical Properties of process streams at its various operating conditions.
- To Predict Phase behavior of Fluids.

Simulations are also performed to detect abnormal conditions like:

- Formation of hydrates by hydrocarbons due to fall in pressure & temperature.
- Fall in temperature below hydrocarbon or water- dew point.
- Flashing of liquids across control valves or drain valves, etc.
- Condensation of vapors due to cooling.

Especially for steady state simulation it offers:

- Process design (to determine the process conditions required to produce the desired product)
- Process equipment design (to size the equipment required to produce the desired product)
- Process design optimization (to determine the optimum configuration of equipment and maximize energy recovery)
- Process optimization (to determine changes to the current operating conditions that can either reduce operating costs or Increase production).

HYSYS offers a very user friendly interaction environment (fig1.1). Flowsheet can be graphically designed with streams and "blocks" (vessels). Each stream can be defined by its composition and a wide variety of chemical and physical properties. Blocks include various options depending on the type of the vessel. Simplified and detailed approaches to simulate each type of vessel for both steady and transient conditions are provided. In addition, there are theoretical blocks which perform logical operations.

Besides, it offers a wide variety of ways to edit results and data graphically or in charts. Data can be easily transferred to text editors and spreadsheets in numerical or plot form.

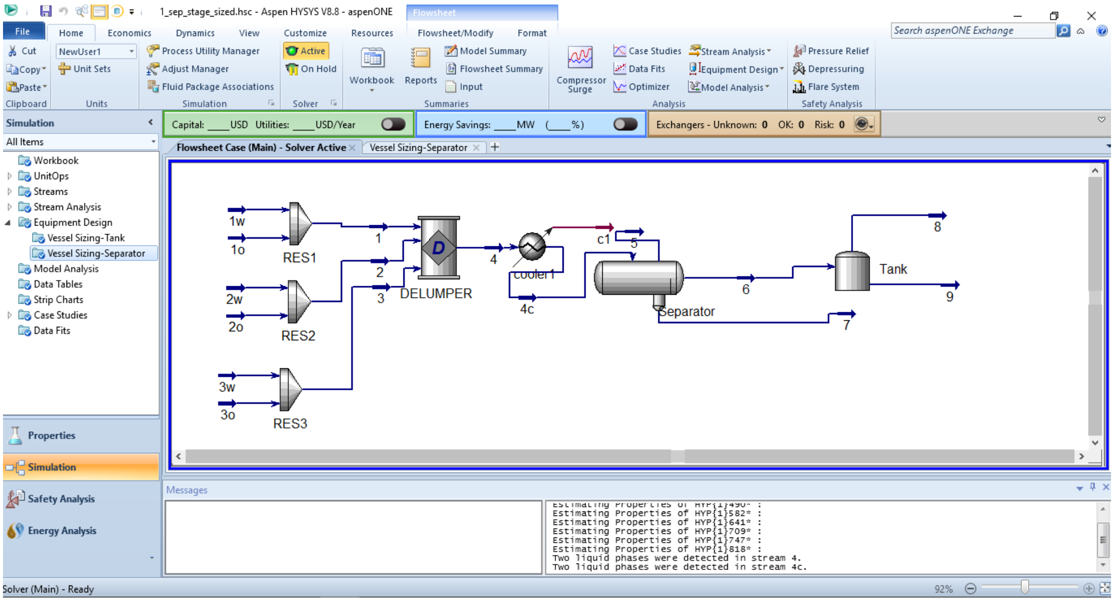

Figure1.1: Aspen HYSYS interface

2. Theory

2.1 Introduction to production facilities

Production facilities include all the treatment processes of the produced fluids from the well-head up to the point of sale. The treatment aims to separate the well stream into the three basic fluid phases (oil, gas and water) and to process each of the phase stream either into marketable products (i.e. to sales specification) or to a form suitable for disposal in the environment.

The primary separation process is conducted in closed steel drums which are called separators. The process is based on the density difference between the three fluids. There are various types of separators with different configurations and advantages for each case.

The separated gas must be compressed to the gas facilities pipeline pressure and be subjected to further treatment for sale. Gas treatment mainly includes removal of water vapor (dehydration) and heavier hydrocarbons. Additional equipment may be needed to “sweeten” the gas namely removing other contaminants such as carbon dioxide,hydrogen sulphide etc.

The separated oil is subjected to further refinement: Very light hydrocarbons must be removed from the oil in order volatility (vapor pressure) to be within market specification range. This process is called stabilization of crude oil. Although oil and water can be effectively separated by gravity in most of the cases, the removal of the smallest water droplets which may occur in the oil in the form of emulsion is difficult. Heat treatment is usually required to break the emulsion and achieve the required high degree of separation. Suspended solids can also be present in the oil so additional processes may be required to reduce the composition of the basic sediment to acceptable values, typically less than 0.5%.

The separated water goes through additional treatment process to remove dispersed oil and suspended or dissolved solids so that it becomes suitable for reuse (water injection) or for disposal. An overview of the production facilities is provided in Figure 2.1.

Except from the aforementioned separation and treatment processes, the upstream facilities include the equipment for accurate measurement and sampling of the crude oil. Important auxiliary systems such as fuel treatment, power generation, control rooms, accommodation facilities, emergency and safety equipment are also included in production facilities.

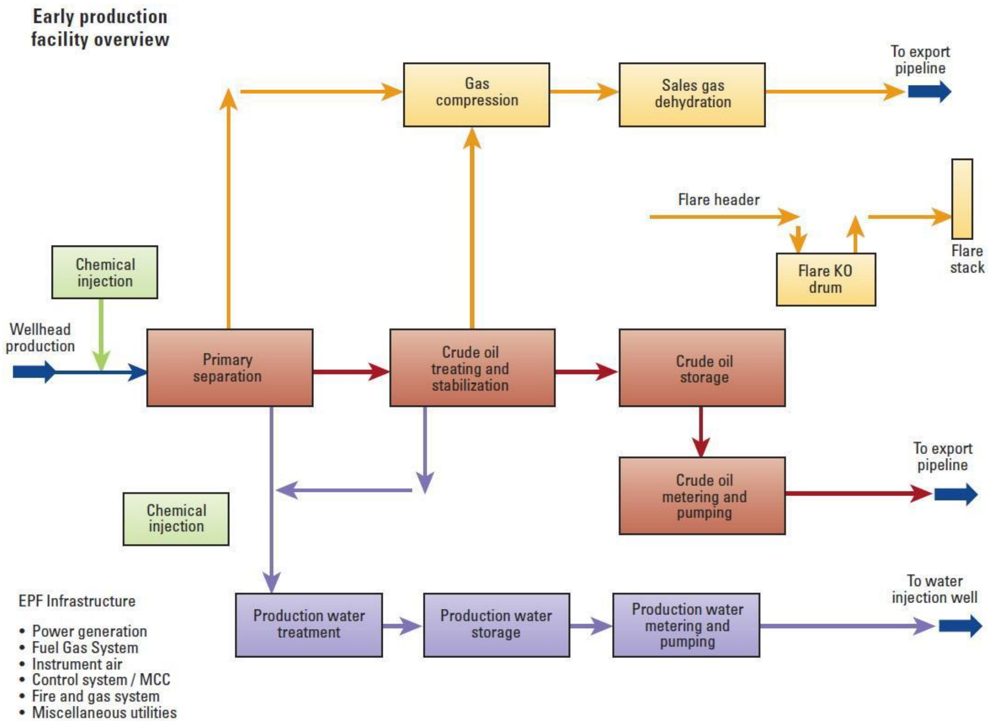

Figure 2.1: Simplified Processing scheme for an oil facility.

2.2 Separators

Wellhead fluids are complex mixtures of gas and liquid hydrocarbons, brine and suspended solids. The fluid sustains pressure and temperature reduction from reservoir to wellhead so that part of liquid is vaporized while part of each phase is encaged to another continuous phase.

The separation of each phase is necessary before the latter is subjected to any further processing and treatment. Improper separation can be the cause of severe problems to the downstream equipment: Centrifugal pumps cannot handle gas bubbles in a liquid flow, while gas compressors and dehydration equipment treat only gases that are cleared from any liquid droplets. Moreover, product specifications impose constraints to basic sediment, water, condensable and volatile hydrocarbon content. Last but not least, accurate oil or gas measurement can be significantly impaired by the presence of a second phase.

Thus, pressure vessels called i.e. separators have been employed to accomplish three primary tasks:

a) Primary phase separation between gas, liquid and water
b) Refinement of each of the separated phases by removing any dispersed second phase

c) Discharge of each refined phase in a way that prevents any re-entrainment of another phase.

The separation is achieved by applying physical forces: a) centrifugal forces - the fluid mixture is subjected to a whirling motion so that the heavier liquid is concentrated to the perimeter while the gas is concentrated to the center of the spiral b) gravity separation -the denser phase moves downwards while the light phase escapes upwards c)inertia forces -when the direction of a fluid stream is abruptly changed, the contained phases are separated due to different inertia d) surface tension (adhesion)– the fluid is forced into contact with a metal surface where the oil phase is preferably adhered to due to the difference in surface properties with respect to the water phase. The contact can be achieved through impingement, coalescence or filtering.

A separator can be divided into four main parts:

1) The inlet diverter section: The bulk of the wellhead liquid is separated from the gas by the abrupt change in the direction of flow and by deceleration.
2) The liquid settling section: It provides the necessary time - retention time - for the gas bubbles which are dispersed in the liquid to be removed and for the liquid mixture to equilibrate. The liquid storage capacity serves also as attenuator for irregular liquid flows (slugs).
3) The gravity settling section: Droplets remaining in the gas stream even after the inlet diverter are separated by gravity and coalesce to the gas-liquid interface.
4) The mist extractor section: Device which removes the droplets remaining in gas after the gravity settling section. The device may use a combination of filtering, coalescence, impingement or flow direction change.

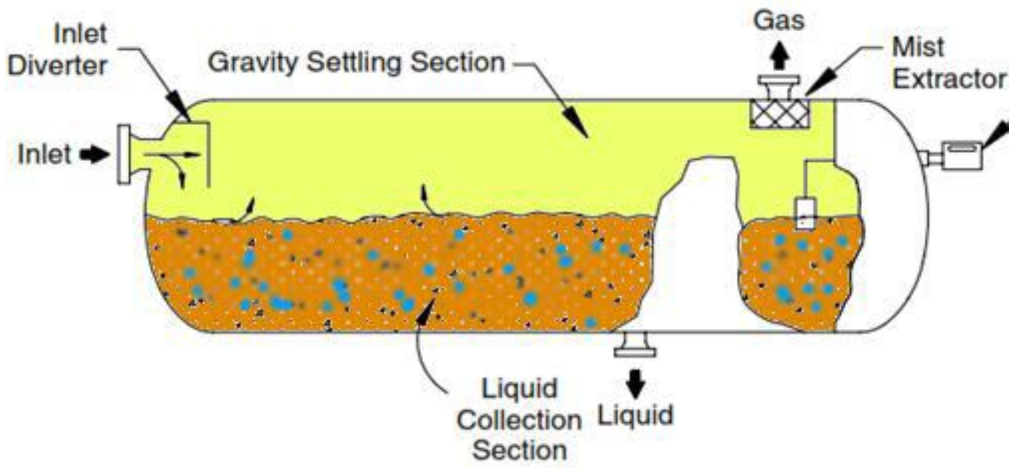

Figure 2.2: Main functional section of separators

Apart from the primary tasks a separator must also effectively address several secondary tasks and operational problems which may be encountered during transient process conditions:

- Optimum pressure maintenance: pressure must be maintained in the separator within certain limits both for separation performance issues, as well as for the fluids to be discharged into downstream operations without requiring any additional mechanical assistance. Pressure is maintained either by a set of gas backpressure regulator valves, each one installed on each separator or by a single master backpressure valve that simultaneously controls the pressure on all separators.
- Level control: To maintain pressure on a separator, a liquid seal must be effected in the lower portion of the vessel. This liquid seal prevents loss of gas with the oil and requires the use of a liquid-level controller and a valve. There is a wide variety of mechanical devices for level control. The amount of time a liquid stays in a vessel, the retention time, assures that equilibrium between the liquid and gas has been reached at separator pressure. The retention time in a separator is determined by dividing the liquid volume inside the vessel by the liquid flow rate.
- Foaming: Gas bubbles may concentrate on a thin oil film when the pressure is reduced or due to high viscosity, surface tension and impurities (CO_2, completion fluids) acting as foaming agents in oil. Foaming greatly reduces the capacity of oil and gas separators because it requires substantial longer retention time to break and to separate. Foaming can also interfere with the level control and the phase discharge equipment. The most common methods to reduce and break the foam in oil include settling, agitation (baffling-defoaming plates), heat, chemicals, and centrifugal forces.
- Parrafins: Paraffin deposition in oil and gas separators reduces their efficiency and blocks the mist extractor and discharge outlets of the fluids. Steam, chemical solvents or special coating are used to prevent paraffin agglomeration and deposition.
- Solids: Various solids such as sand, silt, salt, wax etc. can be transferred to the vessel with the produced fluids and cause plugging and erosion. Conical bottoms, water and steam injection (sand jets and drains) can remove sand and other deposited solids. Salts may be dissolved by injection of fresh water in the oil and subsequent separation of the aqueous phase and drainage. Scale and asphaltenes can be controlled be chemical inhibitors.
- Surging Flow:The influx of the separator may not always be at an even flow rate. The separation system must be capable to regulate abrupt high feed rates which can cause disturbance to level control. The level control system and the storage capacity need to be adequately designed to address such issues. Wave breakers, perforated baffles or plates that are placed perpendicular to the flow and vortex breakers can be used to reduce fluid waves and stabilize the flow.

- <u>Emulsion:</u> Both oil-in-water and water-in-oil emulsions have adverse effect on the separation efficiency and level control of the separator. Momentum breakers and corrugated plate settling packs are usually installed in the separator to promote emulsion breaking and oil droplet coalescence. The oil from the separator is routed to a heater-treater or a gun-barrel tank, where the remaining water-in-oil emulsion breaks with heating and with long retention time. Additionally, de-emulsifying chemicals may be used to enhance the emulsion destabilization effect and to reduce the required retention time.

2.2.1 Separator Types

There are several criteria to categorize separators into different types. Primary classification is by their function: Two- or three-phase operation. In a two-phase unit, gas is separated from the liquids with the gas and liquids being discharged separately. In a three-phase separator, well fluids are separated into gas, oil, and water, with each fluid being discharged separately. The choice between each type of separator depends on the expected fluid and production characteristics of the wells.

A special case of a two phase separator is the free water knockout drum(FWKO). Free water is defined as the water which is produced with the oil and which will settle out within five minutes retention time in a vessel. FWKO is a vessel that separates the free water which is discharged from the bottom of the vessel, while gas and oil are both discharged from the upper part of the vessel without any additional separation. FWKOs are usually operated as packed vessels

In some cases it is preferable to remove excess water from the well fluids before they undergo any pressure reduction, which occurs by flowing through chokes and valves. Water removal may prevent downstream problems such as corrosion, hydrates formation, and tight emulsions. If high water cut is expected then excess-water removal at an early stage can reduce the number and the size of the required separators. For production wells with similar characteristics, it may be possible to use a common FWKO, otherwise ,i.e. if the wells fluid pressures differ significantly, it is preferable to use early two phase separators adapted to each well pressure and finally to remove free water with a FWKO upstream treatment.

Separators can be also classified by their shape and external configuration. There are two types for cylindrical shaped separators, horizontal and vertical. Their pros and cons are summarized in Table 2.1

	Horizontal	Vertical
Advantages	1) Higher GOR. They perform well even at much higher gas velocities than vertical separators (no countercurrent flow)	1) Easier to clean and can handle large quantities of sand and solids
	2) Longer residence time for liquid-liquid separation thus accomplishing higher efficiency and stabilization of the fluids	2) Occupies smaller surface area
	3) Cheaper than the vertical separator. Easier to ship and to assemble	3) Better surge control (liquid level changes do not affect gas capacity)
	4) Requires less piping or field connections. Several separators may be stacked, minimizing space requirements	4) Liquid level control is not critical (smaller horizontal dimensions)
	5) Reduces turbulence and reduces foaming (thus, it can handle foaming crudes)	5) Limited reentrance of liquids into of liquid into the gas phase due to the relatively greater vertical distance between liquid level and the gas outlet
Disadvantages	1) Greater space requirements	1) Longer-diameter compared to the one of a horizontal separator for the same load
	2) Accurate liquid level control is more critical	2) Expensive construction and commision
	3) Surge space is somewhat limited	3) Difficult and more expensive to ship (transport)
	4) Difficult to clean (hence a bad choice in any sand-producing area)	4) Difficult to reach and service top-mounted instruments and safety devices
Ideal Use	High GOR crudes, foaming crudes, preferably liquid-liquid separation. Good for a diverse range of situations.	Low to intermediate GORs and where relatively large slugs of liquid are expected

Table 2.1: Comparison of horizontal and vertical separators

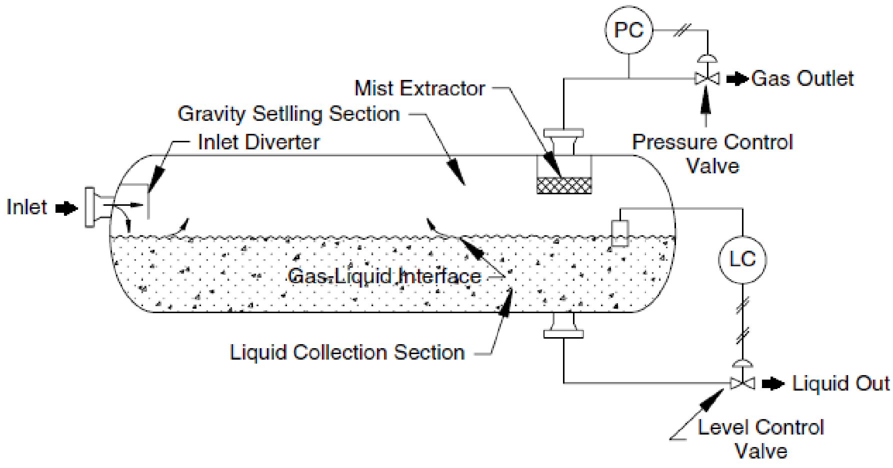

Figure 2.3: Horizontal separator schematic (Arnold K.,1986)

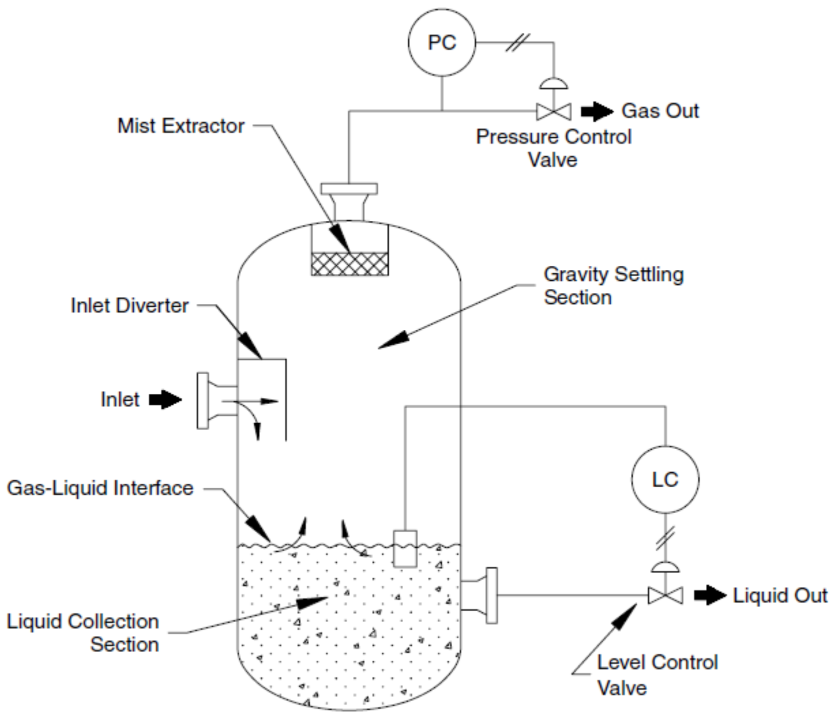

Figure 2.4: Vertical separator schematic (Arnold K., 1986)

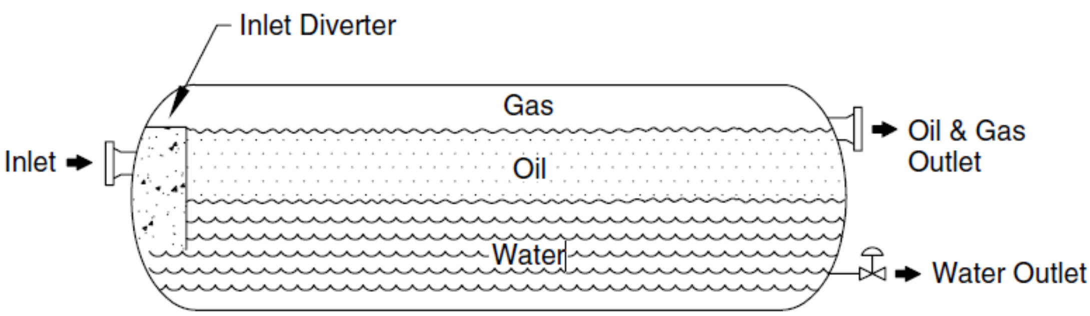

Figure 2.5: Schematic of horizontal FWKO (Arnold K., 1986)

Another separator type is the spherical. Spherical separators were originally considered to combine the advantages of both horizontal and vertical separators. In practice, however, these separators actually perform worse than both types and they are difficult to size and to operate. They have limited liquid surge capability and they exhibit construction difficulties, therefore they are rarely used in oil field facilities. Other separator types are the centrifugal and the venturi separators which share some advantages such as efficiency, low-cost, no moving parts and low maintenance requirements. However, their main drawbacks are the high sensitivity to flow rate fluctuations and the high pressure drop which lead to rare use of these devices.

Several variations of the two basic types of separator, horizontal and vertical, are installed in specific flow cases:

- Double barrel horizontal separators (Figure 2.6): They are used in cases of high gas flow rates and large liquid slugs. They consist of two cylinders-barrels. The flow stream is inserted in the upper barrel where the gas is routed to the mist extractor while the liquid is led through baffles and pipes to the lower barrel. In this manner the liquid accumulation is separated from the gas stream so that there is no chance of liquid re-entrainment in the gas stream due to bursts of high velocity gas. Due to their cost, they are used only in cases where the liquid flow rate is extremely low relative to the gas flow rate.
- Horizontal separator with a “Boot” (Figure 2.7): It is a single-barrel separator with a liquid “boot”, the outlet of which is used only when there are very low liquid flow rates so the “boot” serves as a liquid-liquid separator. The small amounts of liquid flow into the bottom boot which provides the liquid collection section while the main part of the separator is almost completely occupied by the gas. Hence, the liquid handling capability is limited but the gas is processed in a more efficient way. In the case of three phase separators with boot, water and oil separation is better separation, as long as the boot volume is sufficient large and the water flow rate is small.
- Vertical separator with external cone: These are vertical cylindrical separators with an external cone for better handling of fluids with high solid content.

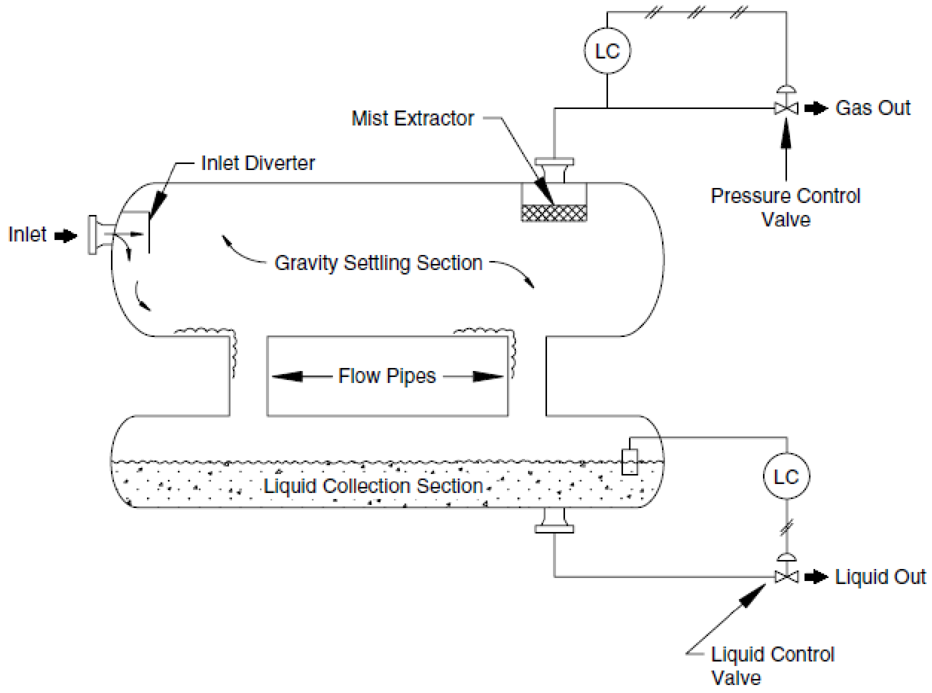

Figure 2.6: Schematic of double barrel horizontal separator (Arnold K., 1986)

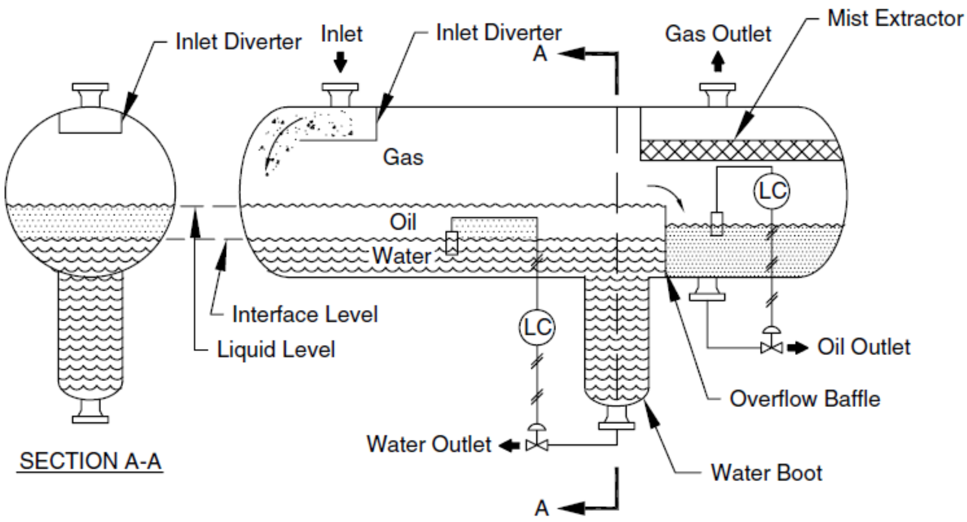

Figure 2.7: Schematic of three phase horizontal separator with a boot (Arnold K., 1986)

2.2.2 Separator Design

Several factors affect separator's efficiency and should be taken into account during the process design:

- Fluid properties, mainly density differences between the phases
- Operating pressure and temperature
- Liquids and gas flow rates
- Required degree of separation

- Surging, foaming and emulsion tendency
- Impurities composition
- Corrosion tendency

The flow rates of vapor and liquids, as well as some of their physical properties, are not always definite during the design period of an upstream facilities' project, so assumptions and forecasts should be utilized. An essential piece of information is the amount and composition of the inlet feed to the separator. Based on this information operating conditions can be simulated for the amount and composition of the produced vapor and liquid streams, assuming that equilibrium is attained between these exit streams (Chilingarian.G, 1987).

Equilibrium is a theoretical condition which describes an operating system that has reached a "steady-state" where the vapor is condensing to liquid at the same rate as the liquid is boiling to vapor. In most production systems, true equilibrium is never actually reached; however, vapors and liquids move through the systems low enough that a "pseudo" or "quasi" equilibrium is assumed. This assumption simplifies the process calculations (Arnold K., 1986).

Under the assumption of equilibrium conditions, flash calculations can be applied to define the optimal operating conditions (separator's pressure and temperature). This allows the determination of the gas and liquids outflows as well as the size of the separator. Important properties for the outlet products are then calculated, such as the oil formation volume factor, the producing gas-oil ratio and the API gravity of the stock tank oil.

The purposes of the separation design are to maximize the oil volumetric production from the well fluids and to minimize the compressors horsepower. Stage separation is usually the option to fulfil the aforementioned requirements. Stage separation of oil and gas is carried out with a series of separators operating at subsequently reduced pressures. Liquid is discharged from the higher-pressure separator into the lower-pressure separator as it is depicted in Figure 2.8.

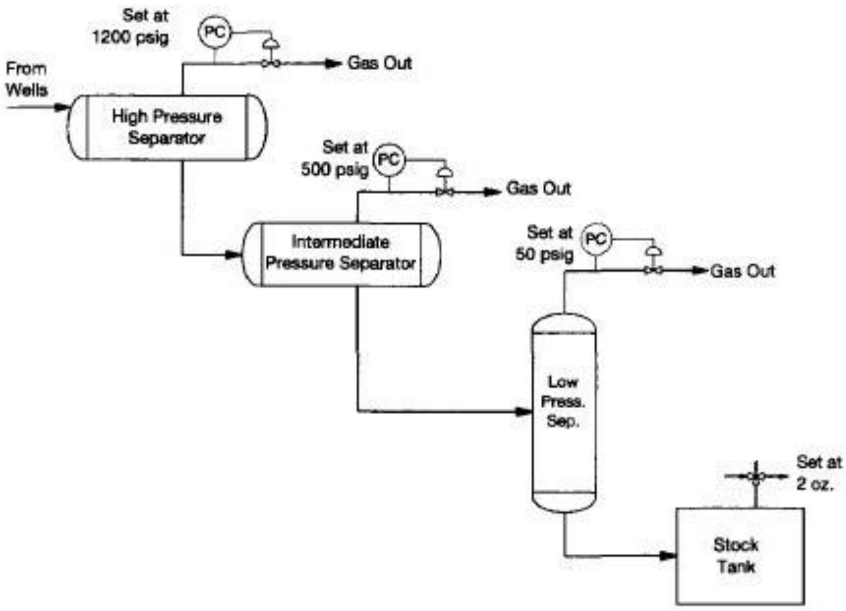

Figure 2.8: Stage separation (Arnold K., 1986)

When the pressure of the separator increases, the gas/oil ratio of the separator decreases, and more of the light hydrocarbons enter the stock tank liquid. At the end of the process the oil will be discharged to the tank at atmospheric pressure and so some of the dissolved light hydrocarbons will evaporate from the liquid phase and increase the gas/oil ratio in the stock tank. At the optimal design conditions, the sum of the gas from the separator and the stock tank reaches a minimum value and the light hydrocarbon content of the stock tank oil maximizes. The pressure at this minimum is referred as the optimized pressure of the separator. The optimum value of pressure is also the one that produces the maximum liquid yield (by minimizing the oil volume factor Bo) and maximizes the API gravity of the product. Thus, the smaller the value of GOR and Bo, the larger the volumetric liquid yield and API value.

This phenomenon can be calculated by flash equilibrium method and be explained qualitatively considering partial pressures. If the pressure in the vessel is high, the partial pressure for the component will be relatively high and the molecules of that component will tend toward the liquid phase. However, many of these molecules are the lighter hydrocarbons (methane, ethane, and propane), which have a strong tendency to flash to the gas state at stock-tank conditions (atmospheric pressure). The present quantity of these molecules in the tank leads to low partial pressure of the intermediate range hydrocarbons (butanes, pentane, and heptane) whose flashing behavior at tank conditions is very sensitive to partial pressure modifications. Hence, keeping lighter molecules in the stock tank results in a small amount of them stocked as liquid, but many more of the intermediate-range molecules are vaporized and escape. Therefore, beyond some optimum point there is actually a decrease in stock-tank liquid by separator operating pressure augment. The above phenomenon is repeated during every stage in a multi-stage separation process, providing more light molecules in the stabilized liquid.

A very large number of stages is neither economical nor practical. Besides, the increase in liquid yield between subsequent stages reduces fast as the number of stages increases. Thus, for each facility there is an optimum number of stages which may also change during the production progress of an oil field.

The optimum number of stages for the separation can be determined either by field testing or by equilibrium calculations based on laboratory tests of the well fluid. Flash calculations can be made for different number of separation stages and determine the optimum number for each specific field case. Equilibrium calculations should also conducted for different separator conditions and favor the ones that will maximize oil production.

Other factors that should be taken into consideration are:

- physical and chemical characteristics of the well fluid
- flowing wellhead pressure and temperature.
- operating pressures of available gas-gathering systems
- facilities for transporting oil and aqueous liquids.

Multi-stage separation is often the preferable choice for processing fluids from wells that produce under different pressures, as it is in the case of different reservoirs, or different production stages. In this case, wells with lower well pressure would be rerouted to a low pressure separator. By using a combination of high and low pressure separators, high-pressure wells can continue to flow at sales pressure requiring no compression, while those with lower tubing pressures can flow into whichever system minimizes compression(Arnold K., 1986).

Usually it is most economical to use three to four stages of separation (including the stock tank as a separation stage) for the hydrocarbon mixture. Five or six stages may payout under favorable conditions, as it is - for example -when the incoming wellhead fluid is at very high pressure. The two-stage separation is commonly used for low-gravity oils, low gas/oil ratios, and low flow pressures. On the other hand, the three-stage separation is used for intermediate-gravity oils, intermediate to high gas/oil ratios and intermediate wellhead flow pressures. Finally, the four-stage, separation is applicable for high-gravity oils, high gas/oil ratios, and high flowing pressures. Four-stage separation is also used when a high-pressure gas is required from the sales contract of for pressure maintenance reasons (Bradley H., 1987).

Compression of the separated gas to sales pipeline pressure is always required. So the produced gases from each separation stage are gathered and compressed by gas compressors. The allowable compression ratios and the power requirements of the compressors will determine the pressure ratios between the different stages of the separation. Usually a number of gas compressors are used instead of a single one. In most cases the gas from each separation stage is compressed to the operating pressure of the one from the previous stage.

Minimizing the compression power requirements while maximizing the liquid recovery, will dictate that the gas from each oil separation stage is compressed to the operating pressure of the previous stage.

The primary separator pressure must be low enough to allow effective choke operation and thus control of well behavior, but it should not be higher than the sales gas pressure. There are guidelines in literature such as the ones depicted in Table2.2. A minimum pressure for the lowest-pressure stage would be in the range of 25 to 50psig. This pressure is probably needed to allow the oil to flow to a treater or tank and the water to be dumped to the water treating system. Increasing the low-pressure separator may decrease the compression horsepower required for the separated gas but it will also add backpressure to wells, thus restricting their flow, and result in more gas vented from the stock tank to the atmosphere. Usually, operating pressures between 50 and 100 psig are optimum (Arnold K., 1986).

Table 2-2
Stage Separation Guidelines

Initial Separator Pressure		
(kPa)	(PSIG)	Number of Stages[1]
170–860	25–125	1
860–2100	125–300	1–2
2100–3400	300–500	2
3400–4800	500–700	2–3[2]

[1] Does not include stock tank.
[2] At flow rates exceeding 650 m^3/hr (100,000 BPD), more stages may be justified.

Table 2.2: Initial separator pressure guidelines (Arnold K., 1986)

A simplified analysis of a multi-stage separation system involves the determination of the operating pressure for each stage (Campbell J.M., 1976):

$$R = \sqrt[n]{\frac{p_1}{p_n}} \quad (2.1)$$

Where: R=pressure ratio; p_1=pressure in stage 1 (high-pressure end),psia; p=stocktank pressure, psia; and n=number of stages minus 1. This implies assumption of equal pressure ratios between the stages, which has been found to be the optimum operating condition for maximizing liquid recovery. The maximum ratio per stage will normally be in the range of 3.6 to 4.0 in order to minimize inter-stage temperatures. The pressure at any intermediate stage can then be determined using the following equation:

$$p_r = \frac{p_{r-1}}{R} \ (2.2)$$

Where: p_r = pressure at stage *r ,* psia; and *R* = pressure ratio.

The above equations will give a first approximation that can be used when no other information is available. Equilibrium calculations should be subsequently applied to accurately determine the optimal pressure conditions for each stage.

2.2.3 Separator sizing

The separator vessel sizing requires understanding of droplet settling theory. In gravity settling section, the droplets are removed from a continuous phase due to gravity forces. The gravity force applied on a droplet is equalized by the drag force due to the continuous gas phase. The drag force on a droplet can be calculated from the following equation:

$$F_D = C_D A d\rho \frac{{V_t}^2}{2g} \ (2.3)$$

Where: F_D=drag force, lb_f, C_D=drag coefficient, A=cross-sectional area of the droplet, ft^2, ρ=density difference between phases, lb/ft^3 , Vt=terminal (settling velocity) of the droplet, ft/s, g =gravitational constant, 32.2 lbm_ft/lb_fs^2

Assuming laminar flow, $C_D = \frac{24}{Re}$, according to Stokes' law, where Re = Reynolds number, which is dimensionless. The buoyant force applied to droplet due to gravity is determined by equation:

$$F_B = (\rho_l - \rho_g)\frac{\pi {D_m}^3}{6} \ (2.4)$$

Where: Dm=droplet diameter, ft. Equaling the above two forces the following equation can be produced to determine the settling velocity of droplet:

$$V_t = 0.0119\sqrt{(\frac{\rho_l - \rho_g}{\rho_g})\frac{d_m}{C_D}} \ (2.5)$$

Where: ρ_l=density of liquid, lb/ft^3, ρ_g=density of the gas at the temperature and pressure in the separator, lb/ft^3, dm=droplet diameter. Laminar flow does not always govern in production facilities so that C_D determination is more complex:

$$C_D = \frac{24}{Re} + \frac{3}{\sqrt{Re}} + 0.34 \ (2.6)$$

Equations (2.5) and (2.6) can be solved by an iterative process to determine the settling velocity:

1. Start with C_D=0.34 and calculate Vt with eq.2.5
2. Calculate $Re = 0.0049 \frac{\rho_g d_m V_t}{\mu}$
3. Calculate C_D with eq. 2.6 and new Re.
4. Recalculate Vt with eq. 2.5
5. Return to step 2 and iterate until it converges to the same value

Using the preceding settling theory, the gas capacity constraint of a separator can be computed. This constraint implies that droplets have enough time to settle before gas carry them out of the separator. In horizontal vessels, a simple ballistic model (fig.2.9) can be used to determine a relationship between the vessel length and the diameter. When the gas retention time is equal to the minimum droplet settling time, the minimum droplet size and the liquid load of the vessel can be defined. A commonly used set of specifications is the 100 μm minimum droplet size and the 50% of the vessel full of liquid. Then the produced relation is:

$$dLeff = 420\left(\frac{TZQ_g}{P}\right)\sqrt{\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}} \quad (2.7)$$

Where: d=vessel internal diameter in inches, Leff=effective length of the vessel where the separation occurs-in ft, T=operating temperature in R, Qg=gas flow rate in MMscfd, P=operating pressure in psia, Z=gas compressibility, C_D=drag coefficient, dm=liquid droplet to be separated, micron, g=density of gas, lb/ft3,l=density of liquid, lb/ft^3

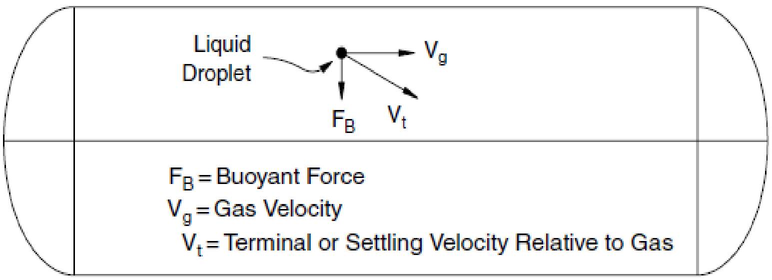

Figure 2.9: Horizontal separator model (Arnold K., 1986)

For vertical vessels (fig.2.9), droplet settling theory using a ballistic model results in the relationship for liquid drops in gas phase as shown below:

$$d^2 = 5040\left(\frac{TZQ_g}{P}\right)\sqrt{\left(\frac{\rho_g}{\rho_l - \rho_g}\right)\frac{C_D}{d_m}} \text{ (2.8)}$$

Separators must be sized to provide adequate liquid retention time so that the liquid can reach phase equilibrium with the gas. Hence, the liquid capacity constraint is defined. For a two phase vessel 50%full of liquid, with a given liquid flow rate and retention time, the following equation can be used to determine vessel size:

Vertical Vessel: $d^2h = \frac{t_r Ql}{0.12}$ (2.9)

Horizontal Vessel: $d^2Leff = \frac{t_r Ql}{0.7}$ (2.10)

Where t_r=desired retention time for the liquid, min, Ql=liquid flow rate, bpd

In three phase separators, a further analysis has to be made for oil-water and water-oil interphase levels. Stokes' law is used in these cases because laminar flow regime can be assumed. As this study uses only three phase horizontal separators, the analysis will be limited to this case. Liquid retention time constraints can be used to develop the following equation, which determines the acceptable combinations of d and Leff for half-full horizontal three phase vessels.

$$d^2Leff = 1.42[(Q_w)(t_r)_W + (Q_o)(t_r)_o] \text{ (2.11)}$$

Where, Qw=water flow rate, BPD, tr_w= water retention time, min, Qo=oil flow rate, BPD, tro= oil retention time, min.

The settling velocity of water droplets through oil can be calculated using Stokes' law and eq.2.5. By using the calculated velocity and the specified oil phase retention time, the settling distance (oil pad thickness)of the minimum water droplet may be determined:

$$h_o = \frac{0.00128((t_r)_o(\Delta SG){d_m}^2}{\mu} \text{ (2.12)}$$

The maximum oil pad thickness constraint establishes the minimum diameter in accordance with the following procedure:

1. Define droplet sizes and retention times and compute (h_o)max with eq.2.12.
2. Calculate the fraction of the vessel cross-sectional area occupied by the water phase. This is given by $\frac{A_w}{A} = 0.5\frac{Q_{w(t_r)_W}}{(Q_w)(t_r)_W + (Q_o)(t_r)_o}$ (2.13)

3. From Figure 2.10, determine the coefficient β.

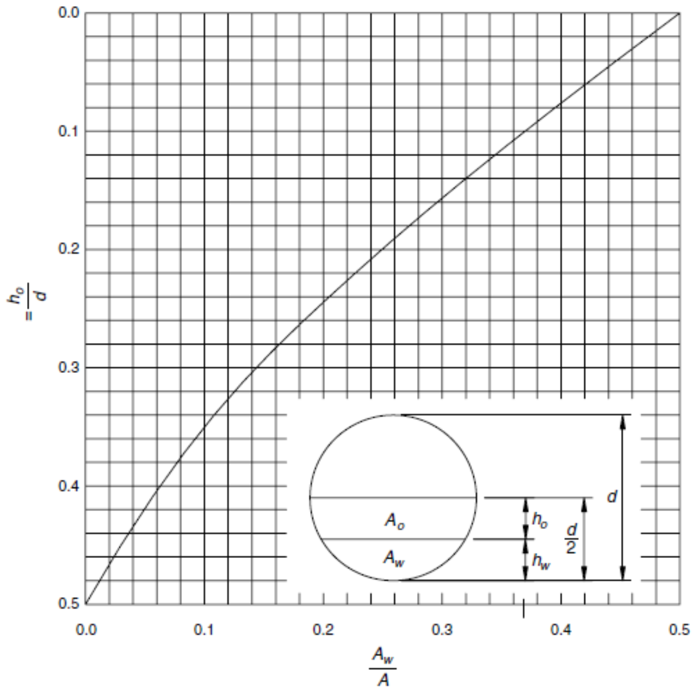

Figure 2.10: Coefficient "β" for a cylinder half filled with liquid**.** (Arnold K., 1986)

4. Calculate dmax from: $d_{max} = \frac{(h_o)max}{\beta}$ (2.14) where
5. Any combination of d and Leff that satisfies gas capacity constrain and eq. 2.11 and 2.12 will meet the necessary criteria.

As far as oil droplets in the water phase are concerned, they rise at a terminal velocity defined by Stokes' law. As with water droplets in oil, the velocity and retention time may be used to determine a minimum vessel diameter from eqs. 2.5 and 2.12.

<u>Droplet Size</u>

All of the aforementioned calculations are based on droplet size distribution. As it is difficult to predict the droplet size that must be settled out in order to reach a predefined liquid content, the following guidelines according to Arnold K.(1986) are utilized.

The purpose of the gravity settling section of the vessel for the gas phase is to condition the gas for final polishing by the mist extractor. Field experience indicates that if the droplets which are larger than 140-micron are removed in this section, the mist extractor will perform well in removing the smaller droplets which are between 10- and 140-micron in diameter. So the default value for gas capacity design equations in this section is the removal of all droplets that are larger than 140 micron in diameter. In some cases this will result in a very conservative solution, thus the same techniques can be easily implemented for any value of minimum droplet size.

There are special cases where the separator is designed to remove only very small quantities of liquid that could condense due to temperature or pressure changes in a stream of gas that has already passed through a separator and a mist extractor. These separators, named as “gas scrubbers,” could be designed for removal of droplets on the order of 500 microns without fear of flooding their mist extractors. Compressor suction scrubbers are examples of vessels to which this might apply.

In the case of water droplets suspended in the oil phase, there are even less available laboratory or field data. In practice, acceptable results may be obtained when sizing the oil pad so that water droplets of 500 microns or larger can settle out. If the 500 μm water droplet is removed effectively, then the downstream equipment can treat a water-in-oil emulsion product that contains less than5-10% of water. In heavy crude oil systems it is difficult to remove water droplets, so the design can account for 1,000-micron water droplets and the produced w/o emulsion may contain as much as 20% to 30% water.

The effectiveness of droplet separation by gravity in a continuous liquid phase depends strongly on the viscosity of the continuous phase. When the oil viscosity is 5 to 20 times greater than that of the water, droplet separation is significantly reduced. Sizing for oil droplet removal from the water phase is normally unnecessary except from the rare case of exceptionally viscous aqueous phase (e.g. glycol/water mixtures).In such case, a minimum oil droplet size of 200micron can be used as a specification.

In most cases, the maximum vessel diameter which is determined from a200-micron oil droplet rising through the water phase, is greater than the one calculated for 500-micron water droplets falling through the oil phase. Therefore, the maximum vessel diameter is usually determined from the water droplet settling through the oil phase

<u>Retention time</u>

Retention time refers to the average time for which the molecules of a phase remain in the vessel, assuming a plug flow. The retention time is thus calculated by the volume of the liquid storage in the vessel divided by the respective liquid flow rate.

A certain amount of time is required for each fluid phase to reach equilibrium or/and to formulate a dispersed phase of significant size in case of saturated conditions. Generally, the retention time increases as the oil specific gravity or viscosity increases(Tables 2.3 and 2.4).

Retention Time for Two-Phase Separators

°API Gravity	Retention Time (Minutes)
35+	0.5 to 1
30	2
25	3
20+	4+

1. If foam exists, increase above retention times by a factor of 2 to 4.
2. If high CO_2 exists, use a minimum of 5-minute retention time.

Table 2.3: Oil retention time for 2-phase separators(Arnold K., 1986)

Oil Retention Time

°API Gravity	Minutes
Condensate	2–5
Light crude oil (30°–40°)	5–7.5
Intermediate crude oil (20°–30°)	7.5–10
Heavy crude oil (less than 20°)	10+

Note: If an emulsion exists in inlet stream, increase above retention times by a factor of 2 to 4.

Table 2.4: Oil retention time for 3-phase separators (Arnold K., 1986)

Similarly, a certain amount of water retention time is required to ensure that most of the large oil droplets suspended in the water will coalesce and rise up to the oil–water interface. If laboratory field information is not available, a water retention time of 10 minutes is recommended for design purposes. Custom retention times for the maximum oil rate and for the maximum water rate should be calculated, when laboratory data indicate so.

<u>Seam to seam Length</u>

The effective length for horizontal separators can be determined by equation 2.7, 2.10 and 2.11. The final seam-to-seam length is also determined by the design of the internal equipment of the vessel. At the inlet section a certain inlet length is required to distribute the flow evenly near the inlet diverter. Part of the vessel length is also reserved for the mist extractor. The length of the vessel that is between the inlet diverter and the mist extractor is called effective length. As a vessel's diameter increases, more length is required to evenly distribute the gas flow. Based on these concepts coupled with field experience, the following formulas (2.15 a and b) have been defined for three phase horizontal separators (Arnold K., 1986):

$$L_{ss} = \frac{4}{3L_{eff}} \; (when\ liquid\ capacity\ constraint\ governs) \;\; (2.15a)$$

$$L_{ss} = L_{eff} + \frac{d}{12}(when\ gas\ capacity\ constraint\ governs)\ (2.15b)$$

Similarly to a horizontal separator, in vertical vessels the design of the internal equipment strongly affects the seam-to-seam length(L_{ss}). As a rule of thumb, L_{ss} may be estimated from the diameter and liquid height according to the following formulas(Arnold K., 1986):

$$L_{ss} = \frac{h + 76}{12}\ diameter < 36in\ \ (2.16a)$$
$$L_{ss} = \frac{h + d + 40}{12}\ diameter > 36in\ (2.16b)$$

Where h = height of liquid level, in., d = vessel ID, in.

Slenderness Ratio

The calculation procedure will result in several combinations of L_{eff} and d. The smaller the diameter of a vessel, the less it weights and consequently the cost is lower. There is a minimum limit however, after which as the diameter decreases, it increases the possibility of poor separation due to high velocity gases, creation of waves and carryover of liquids in the gas phase.

Empirical correlation indicates that the ratio of the seam-to seam length divided by the outside diameter should be between 3 and5 for three phase horizontal vessels. This ratio is referred as the "slenderness ratio" (SR) of the vessel. Similarly, the slenderness ratio for two phase vertical separators is usually kept below 3-4in order to keep the liquid collection section to manageable height(Arnold K., 1986).

Sizing procedures used in this study

For two phase vertical separator sizing:

1. Specify the design conditions, namely the maximum and minimum flow rates, operating pressures and temperatures, droplet sizes, densities, viscosities, retention time etc.
2. The minimum required vessel diameter (d) is determined by eq.2.8.
3. For the selected d, the height of the vessel (h) is determined by Eq.2.9.
4. From d and h, the seam-to-seam length is estimated using eq. (2.16a-b). The larger value of Lss is selected for safety reasons.
5. Check if the slenderness ratio is between 3 and 4.

For two phase horizontal separator sizing:

1. Specify the design conditions, namely the maximum and minimum flow rates, operating pressure and temperature, droplet size, densities, viscosities, retention time etc.
2. Prepare a table with calculated values of Leff for selected values of d with eq. 2.7. Calculate Lss using eq.2.15b.
3. For the each d, calculate the Leff using eq.2.10 for the liquid capacity constraint and Lss using eq.2.15a.
4. For each d, the larger Leff should be selected.
5. Calculate the slenderness ratio for each d. Select the combination of d and Lss that has a slenderness ratio between 3 and 4.

For three phase horizontal separator sizing:

1. Specify the design conditions, namely the maximum and minimum flow rates, operating pressure and temperature, droplet size, densities, viscosities, retention time etc.
2. Calculate max (h_{oil}) with eq. 2.12
3. Calculate d_{max} from the procedure previously described, using eq.2.13-2.14 and fig.2.10
4. Calculate combinations of $d<d_{max}$ and L_{eff} that comply with the gas capacity constraint (eq. 2.7).
5. Calculate combinations of $d<d_{max}$ and L_{eff} that comply with the oil and water retention time constraints (eq.2.11).
6. Estimate the seam-to-seam length (eq.2.15a,b).
7. Select a reasonable slenderness ratio of 3 to 5.

As final choice, it is always more economical to select between standard vessel sizes. Standard vessels outside diameters are usually multiples of 6 in. The width of steel sheets for the shells is usually 10 ft, thus it is common practice to specify Lss in multiples of five.

2.3 Gas Compression Train

The gas which is liberated from the production fluids must be delivered to the export pipeline at the specified pressure. The initial gas pressure is defined by the operating pressure of each separation stage in a multi-stage separation process. Gas boosters are used to increase the pressure of the gas from a separation stage to the pressure of the immediately preceding separation stage as this is proved to be the most economically efficient method.

As gas is compressed to higher pressures its temperature also increases considerably. In order to minimize the compressor's power and maximize the liquid recovery, a set of inter-stage coolers are used, as it is depicted in fig.2.11. Cooling increases the gas density and reduces the volume of the gas resulting in more economical compressor requirements, both in size (piston volume) and in capacity. Sea water is used as a cooling medium in off-shore sites and air coolers in onshore ones. The temperature of the cooled gas is commonly set to 100F.

When the gas is cooled down, some part of the heavy hydrocarbon content will condense. Therefore a scrubber must be placed immediately after heat exchangers to remove the liquid. Compressors are susceptible to damage by liquid droplets; hence the presence of the liquid knockout vessels (scrubbers) prior to each compressor is vital. The scrubber is a large vertical separator where the gas enters near the bottom and is discharged at the top of the cylinder. The reduced gas speed and a vane pack in the cylinder lets the oil droplets fall to the bottom. This condensate, the oil droplets, is then pumped back to the inlet of the separator (liquid recovery).

Worldwide accident records indicate that compressors are the single most hazardous piece of equipment in the process. Compressor operating conditions are typically not well known when the compressor is installed, but even when they are, they are liable to change greatly as production wells come on and off. Thus, expect from the scrubbers, compressors are also equipped with safety valves and surge control systems.

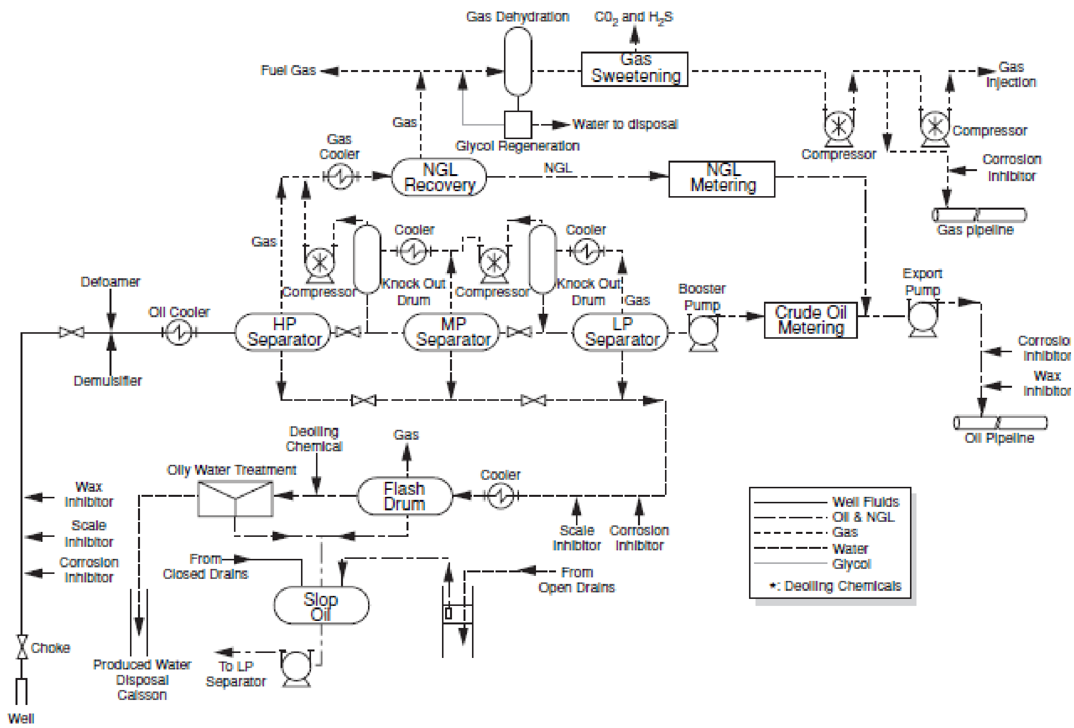

Fig.2.11: 3-stage compression scheme

Especially for the vapor handling of the stock tank gas, a vapor recovery unit (VRU) is used. The most usual type of vapor recovery unit is the compression system. It is usually electrically driven and all components are skid mounted. Some of these systems use a vane-type rotary compressor and they lubricate the seal system of the vanes by injecting refined oil onto the compressor walls. The sealing oil absorbs the condensed hydrocarbons from the compressed vapors and returns them back to storage. The applications for this type of system are (1) compression of the rich stock tank vapors and route condensates to a gasoline plant and (2) recovery of condensates from the rich stock-tank vapors. Condensates can also be recovered from the compressed vapor-recovery unit (VRU) by a vapor cooling system. Air-cooled or water-cooled heat exchanger, coupled with a separator, can be installed downstream of the hot compressed VRU or between the stages of a multiple-staged compression system. Alternatively, a mechanical refrigeration unit may be installed downstream of the VRU for a higher-yield liquid hydrocarbon recovery.

The prime mover or compressor's engine type depends on the compressor's location and the power requirements. Gas turbines, diesel engines and electric motors are frequently employed.

Centrifugal and positive displacement reciprocating compressors are both commonly used in oil field applications. Reciprocating compressors compress the gas with a piston moving linearly in a cylinder. Thereby, the outlet gas flow is not steady and a buffer tank is usually employed to reduce vibrations. Reciprocating compressors are particularly attractive for low horsepower(<2,000 hp), high-ratio applications, although they are available in sizes up to approximately 10,000 hp. They have higher fuel efficiencies than centrifugal, and much higher turndown capabilities. Centrifugal compressors use high-speed rotating wheels to create a gas velocity that is converted into pressure by stators. Centrifugal compressors are particularly well suited for high horsepower(>4,000 hp) or for low-ratio (<2,5) in the 1,000-hp and greater sizes. They are lighter, less expensive, take up less space and they tend to have higher availability and lower maintenance costs than the reciprocating compressors. Their overall fuel efficiency can be increased if the heat from the high-temperature exhaust is used in the process.

The compressed gas is subsequently treated in the dehydration and the gas sweetening units (fig.2.11), which are not included in the scope of the thesis. Removing traces of the water vapors from the gas is necessary because it prevents hydrates from forming when the gas is cooled in the transmission and distribution systems and it also prevents the formation of water condensate, which can be the root cause for corrosion problems. Dehydration also increases marginally the line capacity. There are various dehydration methods:

The usual pressure for the gas export pipeline is 120 bar=1740psi. If the gas is re-injected to the reservoir for pressure maintenance then the final gas discharge pressure must be even larger than 148 bar.

3. Separation Design and Simulation

Separators constitute "the heart" of a production process, so both design and simulation should begin from optimizing separators performance and efficiency. This is achieved by testing different separation stages and configurations.

3.1 Properties Specification

The first step in HYSYS simulation is to define the component properties, namely to set up the properties of the basic elements, pure and pseudo components, and select the computation method which will be used in the simulations. Firstly, the number and type of components should be defined. The component list which is used in this study is the following:

<u>Component List</u>

Methane	Pure Component
Ethane	Pure Component
Propane	Pure Component
i-Butane	Pure Component
n-Butane	Pure Component
n-Pentane	Pure Component
i-Pentane	Pure Component
n-Hexane	Pure Component
Hydrogen	Pure Component
H2O	Pure Component
Nitrogen	Pure Component
CO	Pure Component
CO2	Pure Component
H2S	Pure Component
C7+(1)*	User Defined Hypothetical Component
C7+(2)*	User Defined Hypothetical Component
C7+(3)*	User Defined Hypothetical Component

The heavy ends of the three samples shall be defined as User Defined Hypothetical Component. It is not necessary to supply all of the component properties. By supplying only molecular weight and ideal liquid density HYSYS will estimate the critical properties as well as all the variables used to calculate vapor pressure. Specifically:

C7+(1)*: Molecular weight is given in the table 1.2 equal to 329. Liquid density can be calculated: $\mathrm{SG} = 0.9594 \leftrightarrow \frac{\rho}{\rho_w} = 0.9594 \leftrightarrow \frac{\rho}{62.4} = 0.9594 \leftrightarrow \rho = 59.86 lb/ft^3$ The estimated base and critical properties are:

Base Properties		*Critical Properties*	
Molecular Weight	*329.0*	*Temperature (F)*	1112,89
Normal Boiling Pt (F)	746,81	*Pressure (psia)*	251,25
Std Liq Density (lb/ft3)	*57.43*	*Volume (ft3/lbmole)*	15,25
		Acentricity	0,77

C7+(2)*: Molecular weight is given in the table 1.2 equal to 284. Liquid density can be calculated: $\mathrm{SG} = 0.92 \leftrightarrow \frac{\rho}{\rho_w} = 0.92 \leftrightarrow \frac{\rho}{62.4} = 0.92 \leftrightarrow \rho = 57.43 lb/ft^3$ The estimated base and critical properties are:

Base Properties		*Critical Properties*	
Molecular Weight	*274.0*	*Temperature (F)*	*1006*
Normal Boiling Pt (F)	*640.8*	*Pressure (psia)*	*272.6*
Std Liq Density (lb/ft3)	*57.43*	*Volume (ft3/lbmole)*	*13.51*
		Acentricity	*0.6616*

C7+(3)*: Molecular weight is given in the table 1.2 equal to 181. Liquid density can be calculated: $\mathrm{SG} = 0.799 \leftrightarrow \frac{\rho}{\rho_w} = 0.799 \leftrightarrow \frac{\rho}{62.4} = 0.799 \leftrightarrow \rho = 49.86 lb/ft^3$ The estimated base and critical properties are:

Base Properties		*Critical Properties*	
Molecular Weight	*181.0*	*Temperature (F)*	*773.5*
Normal Boiling Pt (F)	*443.4*	*Pressure (psia)*	*284.1*
Std Liq Density (lb/ft3)	*49.86*	*Volume (ft3/lbmole)*	*11.33*
		Acentricity	*0.5185*

In the various equations of state (EOS), the Binary Interaction Parameter (BIP) is used to analyze the extent of non-ideality in a binary mixture (Jaubert and Privat, 2010). The BIP values of Nitrogen and CO2 towards the pseudo components are the standard HYSYS values.

Property Packages

The next step is to use the HYSYS Fluid Packages viewer to select the specialized sets of parameters and calculation methods (property packages) you want to use with the component list. The combination of the component list and the property package comprises the basic HYSYS fluid package. It is crucial for the simulation and model optimization to choose an equation of state which describes well all the input mixtures.

Peng-Robinson approach was considered sufficiently accurate for the scope of this thesis. The Peng-Robinson (PR) model is ideal for VLE (Vapor Liquid Equilibrium) calculations as well as calculating liquid densities for hydrocarbon systems. The PR property package rigorously solves any single-, two-, or three-phase system with a high degree of efficiency and reliability and is applicable over a wide range of conditions:

Temperature Range > -271°C or -456°F

Pressure Range < 100,000 kPa or 15,000 psia

The Peng-Robinson generates all required equilibrium and thermodynamic properties of the phases in the systems. Equations of state, such as the Peng Robinson model, were developed originally to deal with hydrocarbon gas systems. Although they have proven to be very reliable in predicting properties of most hydrocarbon based fluids over a large range of operating conditions, their application has been limited to primarily non-polar or slightly polar components. Polar or non-ideal chemical systems have been traditionally handled using dual model approaches. For non-library or hydrocarbon pseudo-component, HC-HC interaction parameters are generated automatically by HYSYS. Key components such as He, H_2, N_2, CO_2, H_2S, H_2O, CH_3OH, EG, DEG, and TEG require case specific interaction parameters.

Formulations used in HYSYS for the PR equations of state are the following:

$$P = \frac{RT}{V-b} - \frac{a}{V(V+b)+b(V-b)}$$

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

where:

$$A = \frac{aP}{(RT)^2}$$

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

$$b = \sum_{i=1}^{N} x_i \left(0.077796 \frac{RT_{ci}}{P_{ci}}\right)$$

$$a = \sum_{i=1}^{N} \sum_{j=1}^{N} x_i x_j \left[\left(0.457235 \frac{(RT_{ci})^2}{P_{ci}}\right) \alpha_i \left(0.457235 \frac{(RT_{cj})^2}{P_{cj}}\right) \alpha_j\right]^{0.5} \left(1 - k_{ij}\right)$$

$$\alpha_i^{0.5} = 1 + m_i\left(1 - T_{ri}^{0.5}\right)$$

$$m_i = 0.37464 + 1.54226\omega_i - 0.26992\omega_i^2$$

P = Pressure

T = Temperature

$_c$ = indicates the variable at the critical point

$_r$ = indicates the variable at the reduced point

Note: When the component's acentric factor is greater than 0.49, HYSYS uses the following correction term form:

$$m_i = 0.379642 + 1.48503\omega_i - 0.164423\omega_i^2 + 0.016666\omega_i^3$$

For the Peng-Robinson Equation of State, the enthalpy and entropy departure calculations use the following relations:

$$\frac{H - H^{ID}}{RT} = Z - 1 - \frac{1}{2^{1.5} bRT}\left[a - T\frac{da}{dt}\right] \ln\left(\frac{V + (2^{0.5} + 1)b}{V - (2^{0.5} - 1)b}\right)$$

$$\frac{S - S_{\circ}^{ID}}{R} = \ln(Z - B) - \ln\frac{P}{P^{\circ}} - \frac{A}{2^{1.5} bRT}\left[\frac{T}{a}\frac{da}{dt}\right] \ln\left(\frac{V + (2^{0.5} + 1)b}{V - (2^{0.5} - 1)b}\right)$$

Where:

H^{ID} = Ideal Gas Enthalpy used by HYSYS

$$a = \sum_{i=1}^{n}\sum_{j=1}^{n} x_i x_j (a_i a_j)^{0.5}\left(1 - k_{ij}\right)$$

$$b_i = 0.077796\frac{RT_{ci}}{P_{ci}}$$

$$a_i = 0.457235\frac{(RT_{ci})^2}{P_{ci}}\alpha_i$$

$$\sqrt{\alpha_i} = 1 + m_i\left(1 - T_{ri}^{0.5}\right)$$

$$m_i = 0.37464 + 1.54226\omega_i - 0.26992\omega_i^2$$

R = Ideal Gas constant

H = Enthalpy

S = Entropy

^{ID} = indicates Ideal Gas

° = indicates reference state

Fluid Package:

A fluid package is a combination of a component list and a collection of tasks or industry-specific property-derivation methods named as property package. The Peng-Robinson method which was used includes the following default property calculation methods.

Enthalpy Calculation Method: Lee Kesler method is used for calculating enthalpy. The combined Property Package, employs the appropriate equation of state (PR) for vapor-liquid equilibrium calculations and the Lee-Kesler equation for the calculation of enthalpies and entropies. This method yields comparable results to the Aspen HYSYS standard equations of state and has identical ranges of applicability. Lee-Kesler enthalpies may be slightly more accurate for heavy hydrocarbon systems, but the method requires more computer resources because a separate model must be solved. As this model includes heavy hydrocarbons Lee-Kesler proves to be a reasonable choice.

Density Calculation Method: COSTALD combined with "The Smooth Liquid Density option" is selected so that Aspen HYSYS interpolates the liquid densities from Tr=0.95 to Tr=1.0, giving a smooth transition. It is recommended for petroleum and hydrocarbon liquid mixtures at low and moderate pressure. The densities differ if the option is not selected. COSTALD provides better results for liquid densities and smoothing near Tr=1 where there is sometimes a risk for discontinuities. COSTALD model includes both temperature and pressure dependence. The critical temperature and pressure of hydrogen is modified as a function of temperature. This feature produces better results for simulation systems containing hydrogen.

Viscosity Calculation Method: Aspen HYSYS default method was selected which provides an estimate of the apparent liquid viscosity of an immiscible hydrocarbon liquid-aqueous mixture using only the viscosity and the volume fraction of the hydrocarbon phase.

Root Searching Methods: Newton-Rhapson is chosen which can quickly find the real root for both liquid and vapor phases when calculating bubble pressure at low temperature, or dew temperature at low pressure (down to 1e-5 Pa).

3.2 Multistage Separation Simulation

3.2.1 Model Assumptions

After the specification of components and the fluid packages, the study can proceed to the main part of simulation. The following assumptions were considered as an acceptable simplification of the real case:

- Steady State
- Negligible tube pressure drop. Separator and heater pressure drops are equal to 1 bar
- Negligible vessel and tube heat losses
- Ideal Separation. In ideal separators, complete/perfect separation between the gas and liquid phases is assumed. In real separators, separation is not perfect: liquid can entrain the gas phase and each liquid phase may include entrained gas or entrained droplets of the other liquid phase. The use of vessel internals - mesh pads, vane packs, weirs, etc. - reduce the carry-over of entrained liquids or gases. This phenomenon will be thoroughly analyzed in the following chapter. Ideal separator was used only for preliminary estimations of separator's performance and size.

3.2.2 Inlet Arrangement

The process flow diagram was formulated and depicted in Figure 3.1. The diagram includes the following units:

<u>MixersOil1, Oil2, Oil3</u>

Three oils are chosen to be simulated as three distinct streams, while the watercut for each oil stream is simulated as a separate stream. This configuration was decided as it offers higher flexibility for production flow management. Consequently, three mixers are required to combine the HC and the water streams from each production forming respectively the three oil-water streams. The initial production rate of each oil-water stream is defined by the input data of the case. The fluids are assumed to reach the inlet of the separation process with the same pressure and temperature which is close to ambient conditions (60F), due to pipeline cooling.

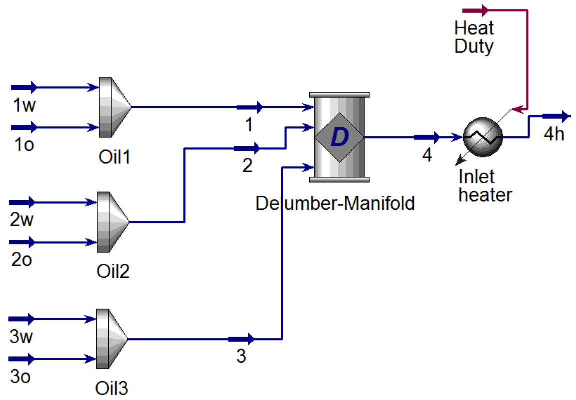

Figure 3.1: *Inlet Arrangement Process Flow Diagram*

<u>Delumber-Manifold</u>

A delumbing unit was used as a collection manifold of the three oil-water fluids. The aim was to have more descriptive illustration of heavy ends properties for higher model accuracy. In practice, though, minor differences were noticed at final results whether using a simple mixer, a delumber or a lumper unit. According to Kazemi, (2011) the way that heavy end is simulated, affects significantly the results.

Peng-Robinson EOS property package and T-P flash calculation method were used. HYSYS Oil char was the delumbing method set, which defined a hypothetical group based on temperature ranges specified automatically. Methane to hexane components were lumbed while the rest were delumbed to eight hypothetical cuts and they are summarized below:

name	Pc (atm)	T(K)	Accentric Factor	Mw
HYP366	22,44	644,20	0,433	136,6
HYP434	20,80	686,09	0,4869	160,5
HYP490	18,63	714,53	0,5523	188
HYP582	14,80	753,42	0,6879	203,6
HYP641	19,39	818,15	0,6496	263,1
HYP709	15,94	843,15	0,7608	294
HYP747	15,10	861,48	0,8104	327,5
HYP818	13,03	891,48	0,9212	381,8

Table 3.1: Delumbing heavy ends results

Heater

A heater was installed upstream of the primary separator, in order to control the separator inlet temperature. As it is analyzed in Chapter 5, the inlet temperature affects significantly the oil production rate. As well fluids reach the separation facility their temperature can be close to the ambient one due to pipeline heat losses, which is more intense in the case of offshore production pipelines.

At ambient temperatures, it may become difficult to separate the phases due to the high viscosity of the liquid. If the temperature at the final separation stage is not controlled, the vapor pressure of the stabilized product might be too high. Even without taking into consideration the cooling due to the adiabatic flash of the separation process. The produced STO might have true vapor pressure higher than 1 atm at 30°C, and Reid vapor pressure well above the specification. So, in such a case the inlet temperature must increase.

The separation of high viscosity/low API gravity oils (less than 25 API), water and sand is also enhanced by heating the separator in a temperature range from 100F to 160F. Elevated temperatures permits lower retention times and therefore reduces the size of the separation equipment. The size of the first stage separator is usually specified by the required water content of its oil effluent that is routed to the second separator.

Heating is also used as a precaution measure for wax and foam formation and accumulation. Viscous crude oil (>53cP) is more likely to foam when the API gravity is less than 40° API and the operating temperature is lower than 160°F.The wax appearance temperature is estimated within the range of 75F and 100F while wax disappearance temperature is estimated within the range of 120F and 130F.

Besides, separators should be operated above the hydrate-formation temperatures, otherwise hydrates may form in the vessel with a risk of plugging. Even if the plug is only partial, it reduces the capacity of the separator and can be the cause of activation for safety valves, rupture disks. Using the HYSYS integrated tool for hydrate formation estimation with the selected as default the method of Ng & Robinson, type II hydrates will form at pressures of 500-1000psi if the temperature is between 55 to 60F. Safety guidelines indicate that a margin of at least 15F should exist to ensure that hydrates will not form, so the fluid temperature must be preferably higher than 75F.

Taking into account the aforementioned guidelines and the estimation of an almost 25 API as stock tank oil product, it was decided to increase the inlet temperature to 130 F in order to avoid or at least alleviate the aforementioned problems.

The composition of the combined oil at the inlet section of each separation stage in a multi-stage separation scheme is estimated to be the following:

	stage1	stage2	stage3	stage4	stage5
Methane	0,464764	0,356293	0,240905	0,168634	0,464764
Ethane	5,58E-02	4,28E-02	2,90E-02	2,03E-02	5,58E-02
Propane	3,33E-02	2,55E-02	1,71E-02	1,20E-02	3,33E-02
i-Butane	2,96E-03	2,25E-03	1,48E-03	1,03E-03	2,96E-03
n-Butane	2,48E-02	1,91E-02	1,30E-02	9,10E-03	2,48E-02
i-Pentane	2,25E-03	1,70E-03	1,10E-03	7,54E-04	2,25E-03
n-Pentane	1,73E-02	1,33E-02	9,05E-03	6,35E-03	1,73E-02
Hydrogen	0	0	0	0	0
H2O	5,60E-02	0,277747	0,514342	0,660762	5,60E-02
Nitrogen	1,05E-02	8,05E-03	5,47E-03	3,84E-03	1,05E-02
CO	0	0	0	0	0
CO2	1,49E-02	1,14E-02	7,61E-03	5,31E-03	1,49E-02
H2S	0	0	0	0	0
n-Hexane	1,27E-02	9,75E-03	6,53E-03	4,56E-03	1,27E-02
HYP{1}366*	2,13E-02	1,64E-02	1,12E-02	7,92E-03	2,13E-02
HYP{1}434*	2,77E-02	2,13E-02	1,46E-02	1,03E-02	2,77E-02
HYP{1}490*	2,84E-02	2,12E-02	1,33E-02	1,11E-02	2,84E-02
HYP{1}582*	3,11E-03	2,35E-03	1,53E-03	1,11E-03	3,11E-03
HYP{1}641*	9,69E-02	7,46E-02	5,20E-02	3,46E-02	9,69E-02
HYP{1}709*	4,00E-02	3,04E-02	2,04E-02	1,42E-02	4,00E-02
HYP{1}747*	7,45E-02	5,62E-02	3,54E-02	2,41E-02	7,45E-02
HYP{1}818*	1,30E-02	9,72E-03	5,85E-03	3,93E-03	1,30E-02

Table 3.2: Separator inlet fluid composition

3.2.3 Optimal number of separation stages and separation pressures optimization

The optimal number of separation stages is defined by production optimization analysis under different production schemes. The optimization was based on the volumetric oil production performance. Together with the number of stages, the optimum pressure of each stage is defined.

Three phase horizontal separators are chosen at each separation stage, due to the intermediate gas flow and the comparatively high liquid volume apart from their other advantages which described in theory chapter .The majority of the three phase separation units is horizontal type. The vertical type is used when there is a large volume of gas and relatively small total volume of liquids (<10-20% by weight). Separators are designed in the simulation as half-full (50% liquid level) which is the dominant case in practice because it maximizes the liquid settling area. Beside the three phase separators, free water knockout separators were also examined as high pressure separators, but the result was poor in respect to stabilization, possibly due to the low watercut of the oil wells.

HYSYS considers tank properties similarly to two phase and three phase separator. The input data are same to the separator's input: half-full, zero vapor outlet pressure drop and outlet pressure=vessel pressure=atmospheric pressure.

Single, two and three stage separation scenarios were examined. As it was analyzed in the theory chapter, the optimal separation operation conditions should be defined in order to maximize production. For this reason, the case study and optimizer tool of HYSYS software were used.

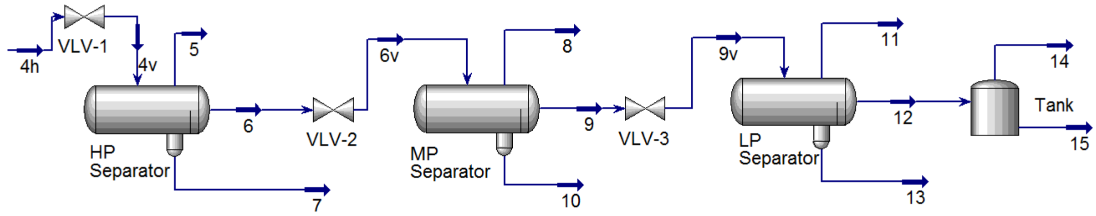

Figure 3.2: *Three stage separation train*

The case study tool facilitates the monitoring of key variables at steady state, as well as how the latter shift when the process conditions change to a new set of values. From the list of all possible variables in a case, some are designated as independent variables, while the rest as automatically designated as dependent ones. For each independent variable, the user can specify a lower and an upper bound, as well as a step size for change. HYSYS will change the value of each independent variable at a step of one variable at a time. With each change, all of the dependent variables are calculated and the new state is defined. HYSYS displays the number of states that are calculated as you define the bounds and the step sizes of change for the independent variables. Once the case study analysis is completed, the user can monitor the states in table format or view the results in plots. The independent variables for study were selected to be the outlet control valves' set pressure.

The optimizer tool enables the analysis of multi-variable steady state optimization problems. It manipulates a given set of independent variables in a predefined range of values, in order to minimize (or maximize) a user-defined objective function, constructed from any number of process variables. Additional equality or inequality constraint functions can be imposed. The outlet control valves' pressure was chosen as independent variables in order to maximize the tank's volumetric oil rate. The range for each variable was specified according to Chapter 2 guidelines for separator's pressure estimation, while the initial value was set according to eq. (2.1) and (2.2) for a pressure ratio of 3.8.

Internally, the Optimizer multiplies the objective function by minus one for maximization problems because all methods refer to minimization. The optimization HYSYS Mixed method was chosen. This method takes advantage of the global convergence characteristics of the BOX method and the efficiency of the Sequential Quadratic Programming method. The minimization starts with the BOX method using a very loose convergence tolerance (50 times the desired tolerance).After convergence, the SQP method is used to locate the final solution using the user defined tolerance.

The BOX Method is loosely based on the "Complex" method of BOX1; the Downhill Simplex algorithm of Press et al (1992) and the BOX algorithm of Kuester and Mize(1973). The BOX method is a sequential search technique which solves problems with non-linear objective functions, subject to non-linear inequality constraints. No derivatives are required. The BOX method is not very efficient in terms of the required number of function evaluations. It generally requires a large number of iterations to converge on the solution. However, if applicable, this method can be very robust.

Procedure:

1. Given a feasible starting point, the program generates an original "complex" of n+1 points around the centre of the feasible region (where n is the number of variables).
2. The objective function is evaluated at each point. The point having the highest function value is replaced by a point obtained by extrapolating through the face of the complex across from the high point (reflection).
3. If the new point is successful in reducing the objective function, HYSYS tries an additional extrapolation. Otherwise, if the new point is worse than the second highest point, HYSYS does a one-dimensional contraction.
4. If a point persists in giving high values, all points are contracted around the lowest point. The new point must satisfy both the variable bounds and the inequality constraints. If it violated the bounds, it is brought to the bound. If it violated the constraints, the point is moved progressively towards the centroid of the remaining points until the constraints are satisfied.
5. Steps #2 through #4 are repeated until convergence.

The Sequential Quadratic Programming (SQP) Method handles inequality and equality constraints. SQP is considered by many to be the most efficient method for minimization with general linear and non-linear constraints, provided a reasonable initial point is used and the number of primary variables is small. The implemented procedure is based entirely on the Harwell subroutines VF13 and VE17. The program follows closely the algorithm of Powell (1978). It minimizes a quadratic approximation of the Lagrangian function subjected to linear approximations of the constraints. The second derivative matrix of the Lagrangian function is estimated automatically. A line search procedure utilizing the "watchdog" technique (Chamberlain and Powell, 1982) is used to force convergence.

The optimization results and pressure ranges used for each scenario are following:

1) Single separation stage

Manipulation variable range	Low Bound	High Bound	initial value
Pressure (Psia)	25	300	250

Optimum operation conditions per production state	state 1	state 2	state 3	state 4	state 5
Separator - Vessel Pressure [psia]	194	185	204	214	207
Tank Liq Vol Flow [barrel/day]	**10073,3**	**9545,4**	**8627,3**	**7749,2**	**6825,7**
Tank Liquid Density (API)	27	27	27	27	27
Tank Liquid Reid Vapour Pressure [psia]	11	11	11	11	11

2) Two separation stages

Manipulation variable range	Low Bound	High Bound	Initial Value
HP separator	125	500	212
LP separator	25	100	56

Optimum operation conditions per production state	state 1	state 2	state 3	state 4	state 5
HP Separator - Vessel Pressure [psia]	398	375	398	449	485
LP Separator - Vessel Pressure [psia]	58	54	58	58	58
Tank Liq Vol Flow [barrel/day]	**10151,2**	**9621,4**	**8699,5**	**7816,7**	**6886,4**
Tank Liquid Density (API)	27,3	27,4	27,8	27,9	27,8
Tank Liquid Reid Vapour Pressure [psia]	10	10	10	10	10

3) Three separation stages

Manipulation variable range	Low Bound	High Bound	Initial Value
HP separator	300	900	806
MP separator	115	450	212
LP separator	25	100	55

Optimum operation conditions per production state	state 1	state 2	state 3	state 4	state 5
HP Separator - Vessel Pressure [psia]	511	571	577	593	511
MP Separator - Vessel Pressure [psia]	115	138	107	103	115
LP Separator - Vessel Pressure [psia]	33	32	29	29	33
Tank Liq Vol Flow [barrel/day]	**10181,0**	**9648,9**	**8727,4**	**7842,5**	**6908,6**
Tank Liquid Density (API)	27,5	27,6	27,9	28,1	28,0
Tank Liquid Reid Vapour Pressure [psia]	10	10	10	10	10

Table 3.3: Multistage separation optimization results

The results of Tables 3.1 to 3.3 were also verified by the CMG's WinProp software. Separator tests were carried in WinProp using the same fluid composition data at the inlet of the primary separator followed by minimization of the oil formation volume factor.

According to HYSYS results, the two-stage separation process increased oil production at 71 barrels per day on average, while the three-stage separation process supplemented the two-stage process by 27 barrels per day on average. Furthermore, due to better stabilization of the final product the RVP pressure was reduced by 1psia with respect to the two-stage and the API gravity increased (28 API) possibly leading to a better market price. Any further separation stage, i.e. in addition to the three-stage separation, would only exhibit marginal increase in the oil production and oil properties so it was excluded from the analysis.

The selection choice between the two and the three stages is not straightforward. A more thorough on the economic feasibility study, which will compare the additional operational and installation cost of the third separator with the possible additional profit must be performed. The production increase due to the third stage can be considered as unprofitable. However, from a process perspective, the third stage can be also used as safety margin in cases of possible debottlenecking procedures, vapor liquid recovery and heat reuse, so the profit can be substantially higher than that originally anticipated. Moreover, the three stage scheme offers higher process flexibility with respect to the handling of water carryover handling, although this is not expected to be important as the watercut remains relatively low (except of the end of production cycle). Conclusively, for this study the three stage separation train will be chosen, although a more detailed investigation is needed to reduce the benefit uncertainty. The average optimal pressures for high, medium and low pressure separators are respectively defined to 550, 115 and 30 psia.

3.3 High Pressure Separator Sizing

Proper separator sizing is possibly more of art than science. Three primary factors should be considered in separator sizing: 1) vapor capacity, 2) liquid capacity, and 3) operability. The vapor capacity will determine the cross-sectional area which is necessary for the gravitational forces to separate the liquid from the vapor. The liquid capacity is typically set by determining the volume required to provide adequate residence time for "de-gasing" the liquid or for allowing the immiscible liquid phases to separate. Operability issues include the separator's ability to deal with possible solids, unsteady flow/liquid slugs, turndown, etc. Finally, the optimal design will usually result in an slenderness ratio that satisfies these requirements in a vessel at a reasonable cost. These factors often result in an iterative calculation approach (GPSA,EDB, 2014).

In section 2.2.3, sizing methods based on droplet settling theory were presented. These methods were included in Excel spreadsheets in order to size the three phase separation train. In addition to this, some technical guidelines (Monneray W., 2004 and Lake,2015) were adopted to define the exact separator type, lower and higher liquid limits and nozzle sizing for each stage.

In order for the design basis of the separator sizing to be set, the following factors should be determined:

- Maximum Gas and liquid flow rates. As it was mentioned in section 2.2.3, methods are quite conservative so peak values have to be used.
- Operating and design pressures and temperatures
- Physical properties of the fluids, such as density and viscosity
- Expected separation performance (e.g., removing 100% of particles greater than 10 microns)

The ideal separation model that was used in 3.2.3 to define the number of stages, had zero degree of carry over. Hence, water was totally removed in HP separator, while MP and LP separators operated in practice as two phase separators with zero water flow (except from the liquefied water vapors in each stage). This is an unrealistic assumption but it is easy to solve for studying oil stabilization. For the high pressure separator sizing, there are sufficient data to formulate the design basis. However, middle and lower pressure separator sizing requires a more realistic calculation of the carry over in their input streams. This is a complex subject that is discussed in the next chapter, preceding to the HP separator sizing. Simulation was carried out for three different cases in order to define maximum flow rates and decide a safety margin for extreme conditions:

a. Normal operation condition, temperature=130F, optimal pressures calculated by the method described in section 3.2.3
b. Highest liquid rate, separator high pressure limit of 650psia, inlet low temperature limit of 100F
c. Highest gas rate, separator low pressure limits of 450psia, high temperature limit of 180F

A safety factor of 10% with respect to the flow rates was determined to account for the adverse conditions. Density, viscosity, pressure and temperature average values were selected. Droplet size and retention times were set according to the guidelines of section 2.2.3.The design basis for each separator is provided in table 3.5.

HP Separator Design Basis			
Qg (MMscfd)	7,57	(SG)w	0,98
Qo (BOPD)	12188	μo (cp)	2,49
Qw (BWPD)	3408	μw (cp)	0,52

Po (psia)	555	dm,l (microns)	100
To (F)	130	dm,w (microns)	500
Sg	0,69	dm,o (microns)	200
APIo	32	(tr)o (min)	7,5
(SG)o	0,74	(tr)w (min)	10

Table 3.4: Separators Design Basis

The next step was to define dmax and to examine combinations of allowable d and Leff that satisfy the gas capacity constraint. The results of very low Leff indicate that gas capacity is not the limiting factor in this study (Table 3.6-3.8):

HP separator d_{max} calculation and Gas capacity constraint			
Maximum oil pad thickness calculation (eq. 2.12)			
(h_o) max (inch)	233,50		
The fraction of the vessel cross-sectional area of water phase (eq. 2.13)			
Aw/A	0,14	coef.β (fig. 2.10)	0,30
Maximum inner diameter (eq. 2.14)			
dmax (inch)	778,34		
`			
C_D Calculation (Iterative process section 2.3.3)			
ρl (lb/ft^3)	46,29	Z	0,88
ρg (lb/ft^3)	1,99	μ gaz (cp)	0,013
C_{D1} (initial value)	0,34		
	V (ft/s)	Re	C_D
Iteration1	0,96	295,57	0,60
Iteration2	0,73	223,30	0,65
Iteration3	0,70	214,06	0,66
Iteration4	0,69	212,60	0,66
Iteration5	0,69	212,36	0,66
Diameter vs. Length for Gas Capacity Constraint (eq. 2.7)			
d (inch)		Leff (ft)	
60		0,85	
72		0,71	
84		0,61	
96		0,53	

Table 3.5: HP separator d_{max} calculation and gas capacity constraint

Since gas is not the limiting factor, the study proceeds to the retention time constraint (eq.2.11), using Lss=3/4Leff. Results are exhibited in the following table.

HP separator Retention Time Constraint			
do (in.)	**Leff (in.)**	**Lss (ft)**	**SR (12Lss/do)**
60,00	49,5	66,0	13,2
72,00	34,4	45,8	7,6
78,00	29,3	39,1	6,0

84,00	25,3	33,7	4,8
96,00	19,3	25,8	3,2
108,0	15,3	20,4	2,3

Table 3.6: Diameter vs. Length for Retention Time Constraint and Slenderness ratio

The final choice will be conducted from the total number of possible combinations that comply with the retention and gas capacity constraints, taking into consideration a reasonable size for separator vessel and a slenderness ratio between 3 and 5. The lowest possible diameter is then chosen for each separator in order to decrease the vessel's weight, while its seam-to-seam length is selected to be equal to the nearest multiple of 5 for a standard steel plate size. Slenderness ratio is then recalculated for the increased Lss in order the ratio to be in the range of 3-5. The results are summarized in table 3.11:

Final size selection	**do (in.)**	**Lss (ft)**	**SR (12Lss/do)**
HP separator dimensions	84	35	5

Table 3.7: HP Separator final dimensions

There are different sub types of horizontal three-phase separators. The liquid separation section may include a boot or a weir and it is usually configured so as to provide adequate interface level control. A boot is normally specified when the volume of water is less than 15-20% of the total weight of the liquid (Monneray W., 2004), otherwise a weir is used when the water volume is substantial. The weight of the water in the total liquid fraction was calculated by HYSYS and it was found lower than 20%, with the exception of production state 5 which includes the highest watercut. Thus, a vessel design with a boot as it is in fig.3.3 is suggested.

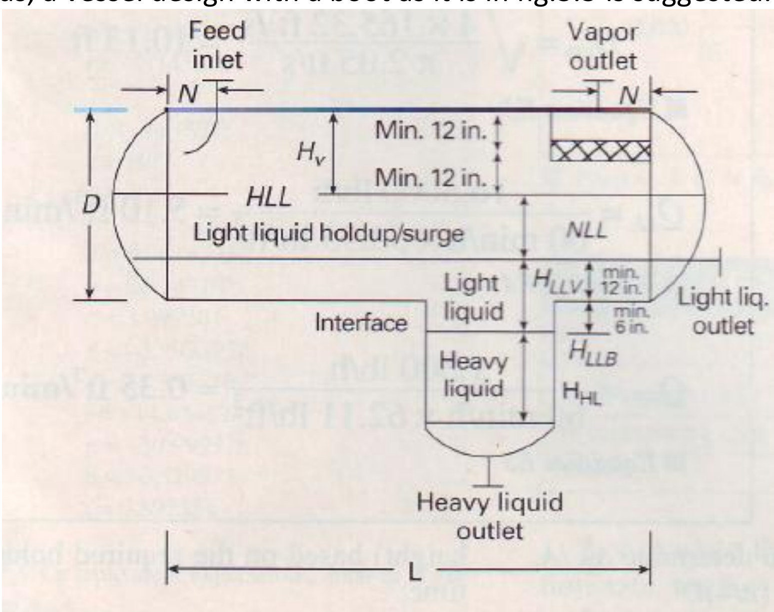

Figure 3.3: Basic design of a three phase separator with a "boot" (Monneray W. & Svrcek W., 2004)

Normal light liquid level is set to d/2 as the sizing procedure considered half full vessel. The exhibited heights (fig3.3) have to be calculated according to Monneray W. & Svrcek W. (2004) to complete separator sizing. At this point, holdup and surge time must be defined. Holdup is defined as the time it takes to reduce the light liquid level from normal (NLL) to empty (LLL) while maintaining a normal outlet flow without feed makeup. Surge time is defined as the time it takes for the light liquid level to rise from normal (NLL) to maximum (HLL) while maintaining a normal feed without any outlet flow. Holdup time is related to the reserve required to maintain good control and safe operation of downstream facilities. Surge time is usually related to requirements to accumulate light liquid as a result of upstream or downstream variations or upsets, for example, slugs. There are guidelines as table 3.12 in literature for holdup and surge time specification. Then, the respective surge and holdup volumes (Vs and Vh) can be calculated by multiplying time with oil flow.

Service	Holdup time, min (NLL — LLL)	Surge time, min (NLL — HLL)
A. Unit feed drum	10	5
B. Separators		
1. Feed to column	5	3
2. Feed to other drum or tankage		
a. With pump or through exchanger	5	2
b. Without pump	2	1
3. Feed to fired heater	10	3
C. Reflux or product accumulator		
1. Reflux only	3	2
2. Reflux and product	3+	2+
(Based on reflux (3 min) plus appropriate holdup time of overhead product (as per B 1–3))		
D. Column bottoms		
1. Feed to another column	5	2
2. Feed to other drum or tankage		
a) With pump or through exchanger	5	2
b) Without pump	2	1
3. Feed to fired reboiler	5–8	2–4
(Based on reboiler vapor expressed as liquid (3 min) plus appropriate holdup time for the bottom product(as per D 1, 2))		
E. Compressor suction/interstage scrubber		
3 min between HLL (high liquid alarm) and high level shutdown		
10 min from bottom tangent line to high liquid alarm		
F. Fuel gas knock-out drum		
20 ft. slug in the incoming fuel gas line between NLL and high level shutdown		

Table 3.8: Holdup and surge time guidelines (Monneray W. & Svrcek W., 2004)

The highest and light liquid levels have to be specified so that vessel can accommodate holdup and surge volumes, namely $Voil_{max}$>Vs+Vh which can be transformed to the following formula:

$$Lss(A_t - A_v - A_{LLV}) > Vs + Vh \ (3.1)$$

Where A_t, A_v, A_{LLV} are cross sectional areas of vessel, vapor and lowest light liquid level which can be calculated with correlation in table 3.13 if H_V and H_{LLV} are specified.

$y = \frac{a + cx + ex^2 + gx^3 + ix^4}{1 + bx + dx^2 + fx^3 + hx^4}$	
H/D to A/A_t	
Y=A/A_t	
X=H/D	
a	-4,8E-05
b	3,924091
c	0,174875
d	-6,35881
e	5,668973
f	4,018448
g	-4,91641
h	-1,80171
i	-0,14535

Table 3.9: Cylindrical height and area conversion correlation (Monneray W., 2004)

The most conservative approach is adopted to maximize available light liquid level by minimizing H_v and H_{LLV}. H_v defines the lowest vapor space. It's minimum value is determined from the following formula:

$$if\ 0.2d > 24in, then\ Hv = 0.2d\ else\ Hv = 24in \ \ (3.2)$$
$$where\ d\ vessel\ diameter, in$$

The heights specification and holdup and surge criterion examination is presented to the next table:

Holdup Time (min)	2
Surge Time (min)	1
Vh (bbl)	16,93
Vs (bbl)	8,46
H_{LLV} (inch)	12
H_{LLB} (inch)	6
Hv (inch)	24
H_{LLV}/D	0,14
A_{LLV}/A_t	0,09
Hv/D	0,29
Av/D	0,24
Voilmax (bbl)	71,88
Vh+Vs (bbl)	25,39

Table 3.10: Heights specification and holdup and surge examination

It can be concluded that Vh+Vs<Voilmax for all cases proving that our design is conservative and holdup and surge volumes can be easily accommodated. However, if an exchanger or pump was mounted downstream of a separator (holdup 5min and surge 2min), HP separator capacity is only sufficient while the rest two separators need resizing.

The boot also needs to be sized. Initially, the height of the heavy liquid must be set (H_{HL}). Then, the rising velocity of the light liquid out of the heavy liquid phase, U_{LH}, has to be calculated with eq.2.5 and Stokes law. A real velocity U_P is set equal to 0.75 U_{LH} in order to add conservatism. Hence, bottle diameter can be calculated by:

$$D_b = \sqrt{\frac{4x12Q_{HL}}{\pi U_P}} \quad (3.3)$$

The settling time of the light liquid out of the heavy liquid phase, namely $t_{LH} = 12\frac{H_{HL}}{U_{LH}}$ (3.4), must be compared to the residence time of the heavy liquid in boot: $\theta_{HL} = \frac{\pi {D_b}^2 H_{HL}}{4Q_{HL}}$ (3.5). If $\Theta_{HL} < T_{HL}$ the boot diameter must be increased else it is kept as it is. The results of boot sizing with the above procedure are presented in table 3.15.

	HP separator
H_{HL} (inch)	18
U_{HV} (ft/s)	0,03
Dboot (inch)	11,67
T_{HL} (min)	0,75
Θ_{HL} (min)	5,03

Table 3.11: Boot sizing

Nozzles were sized setting limits either to velocity or the product of density multiplied by velocity^2 (momentum) based on literature guidelines. According to Monneray W. & Svrcek W., the fluid velocity at the inlet nozzle of separator must be equal or lower than $60/\rho_m$, where ρ_m inlet fluid density. Hence:

$$V_M = \frac{Q_M}{\frac{\pi}{4}{d_n}^2} \leq \frac{60}{\sqrt{\rho_m}} \leftrightarrow d_n \geq \sqrt{\frac{4Q_M}{60\pi/\sqrt{\rho_m}}} \quad (3.6)$$

Where d_n nozzle diameter, ft, Q_M flow rate, ft^3/s, ρ_m= density, lb/ft^3, V_M velocity, ft^3/s

From table 3.16, $\rho_m {V_m}^2$ equals to 2500 ft/s in the gas outlet nozzle so that:

$$\rho_m {V_M}^2 = 2500 \leftrightarrow \rho_m (\frac{Q_M}{\frac{\pi}{4}{d_n}^2})^2 = 2500 \leftrightarrow d_n = 2\sqrt{\frac{Q_M\sqrt{\rho_m}}{50\pi}} \quad (3.7)$$

Where d_n nozzle diameter, Q_M flow rate, ft^3/s, ρ_m= density, lb/ft^3, V_M velocity, ft^3/s

Similarly, in liquid outlet nozzles velocity equals to a value in the range of 3-6ft/s. Assuming V_M=3ft/s:

$$V_M = 3 \leftrightarrow \frac{Q_M}{\frac{\pi}{4}{d_n}^2} = 3 \leftrightarrow d_n = 2\sqrt{\frac{Q_M}{3\pi}} \quad (3.8)$$

Inlet nozzle:	$\rho_m {V_m}^2$, lbm/ft-sec^2
No inlet device:	600–700
Half-open pipe inlet:	1,000
Vane inlet spreader:	4,000–6,000
Inlet cyclones:	6,000–10,000
Gas outlet nozzle:	2,500
Liquid outlet nozzles:	velocity 3–6 ft/sec

Table 3.12: Guidelines for nozzle sizing *(*Lake, 2015)

Nozzle position of feed and gas outlet are defined from distance N in fig.3.3 where N= dn/2+6in. Remembering that mass flow=volume flow/density and using HYSYS output results for flow and density, the following nozzles are calculated:

nozzles sizing	HP separator
inlet ρ (lb/ft^3)	13,35
inlet mass flow (lb/hr)	160542,19
inlet Dn_{min} (inch)	6,11
gas outlet ρ (lb/ft^3)	2,03
gas outlet vol flow (MMscfd)	7,57
gas outlet Dn (inch)	21,41
oil outlet ρ (lb/ft^3)	38,89
oil outlet mass flow (lb/hr)	126618,18
oil Dn (inch)	7,44
water ρ (lb/ft^3)	61,62
water outlet mass flow (lb/hr)	16992,21
water Dn (inch)	2,16
N inlet (inch)	9,05
N gas outlet (inch)	16,71

Table 3.13: Nozzles sizing results

4. Carry Over Calculation

In section 3.2, ideal separation was assumed, namely complete separation of each phase, in order to decide the number of separation stages. In addition to this, in separators sizing technical guidelines were used to alleviate liquid entrainment in gas (slenderness ratio restrictions) and the inlet fluid to HP separator was heated expecting more efficient liquid-liquid separation.

There is a wide range of separator sizing methods. One of them was used in the previous chapter. There are several weaknesses associated with most of these methods, including:

- Quantification of feed flow steadiness
- Entrainment/droplet size distribution quantification
- Velocity profile/distribution quantification
- Separator component performance quantification

The question is how separator performance and carry over effect can be quantified. In this chapter a literature review about this subject is introduced based on the series of articles of Bothamley M. (2013), while HYSYS tools are applied to calculate carry over and design internals in order to increase separation efficiency.

4.1 Literature Review about Carry Over Estimation and Internals Design

Carry over study can be divided per separator's major parts. These parts were defined in Chapter 2 as inlet diverter, liquid settling, gravity settling (primary gas separation) and mist extractor section (secondary gas separation).

<u>Inlet section</u>

In the inlet section, the following questions have to be considered:

- What is the flow pattern at the separator's inlet?
- How much of the dispersed phase (liquid droplets) is present in entrained form?
- What are the sizes of the droplets?
- What inlet diverter must be chosen and how efficient it is?

Unfortunately, quantification of these values is difficult and typically requires simplifying assumptions. The validity of these assumptions has a significant bearing on equipment selection, sizing, and performance prediction. In figure 4.1 various possible inlet flow patterns are presented. There are various literature map and guidelines to predict the prevalent flow pattern in a pipe related to the fluid velocity. What is important to notice is that liquid entrainment begins when droplets are sufficiently small and increases rapidly as the transition to annular flow is approached. Increasing gas velocity is going to strongly correlate with the increasing entrainment and with decreasing droplet sizes, both of which will negatively affect separation. Taking into account these observations, sizing the inlet feed pipe for stratified/wave flow is desirable, if possible. Specifically, inlet nozzle sizing (section 3.3) and feed pipe length must be determined according to technical guidelines.

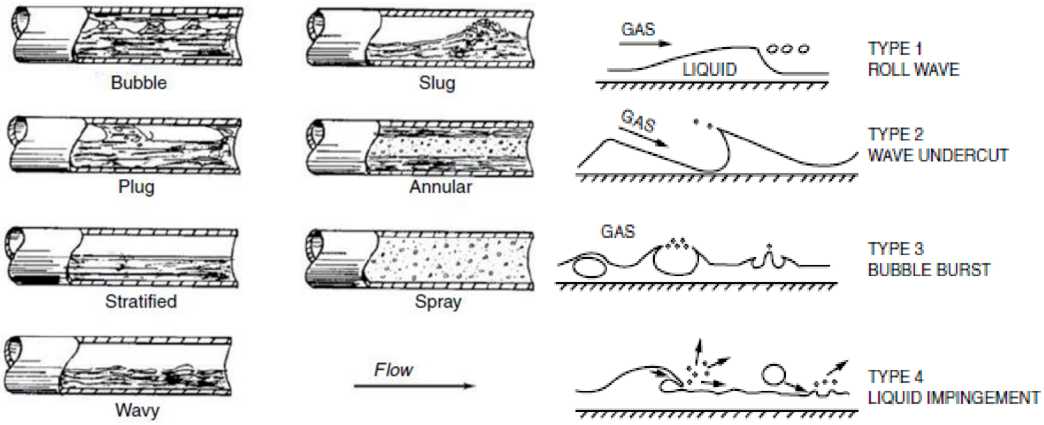

Figure 4.1: Flow patterns and liquid entrainment mechanisms (Bothamley M., 2013)

Liquid entrainment mechanism (fig.4.1) is strongly dependent on liquid phase Reynolds number (inlet momentum) and droplet size (fig.4.2). Droplet size decreases with increasing gas velocity, increasing gas density and decreasing liquid surface tension. Several sophisticated correlations can be found in literature which include these variables and produce droplet size distribution (e.g. Kataoka, 1983), and entrainment fraction functions (e.g. Ishii, 1975). Most of them are referred to annular flow and definitely are not robust enough to serve all cases and rigorous calculations. They can only be used for estimation purposes. Experimental data using various droplet size measurements methods are more reliable procedures.

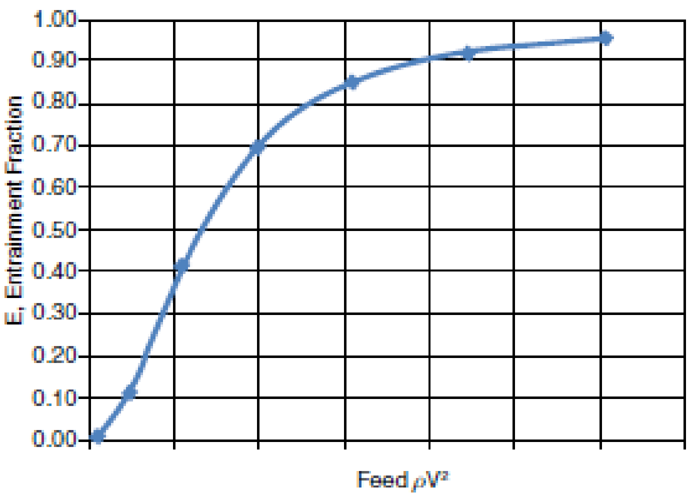

Figure 4.2: Liquid entrainment behavior example

Inlet devices are installed to improve separation performance by:

- Separating bulk liquid from the gas
- Reducing the inlet momentum and ensuring good gas and liquid distribution
- Minimization of droplet creation or shattering
- Facilitating de-foaming

The above actions reduce liquid entrainment. There are various types of inlet devices (fig.4.3) which share different performance characteristics (table 4.1):

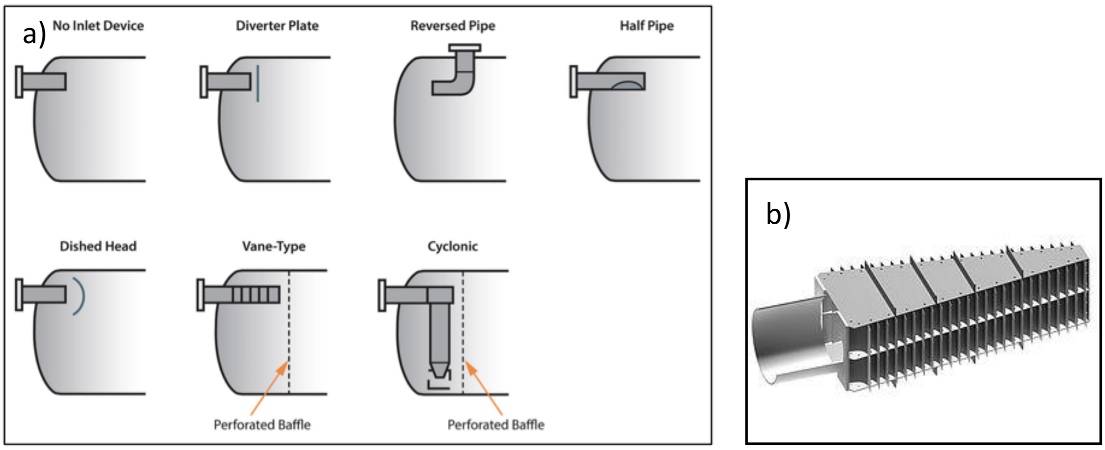

Figure 4.3: a) Various inlet devices b) Vane-Type Inlet Diverter

Function	No inlet device	Diverter plate	Half-pipe	Van-type	Cyclonic
Momentum reduction	Poor	Average	Good	Good	Good
Bulk liquid separation	Poor	Poor	Average	Good	Good
Prevent re-entrain-ment	Poor	Poor	Average	Good	Average-Good
Minimize droplet shattering	Poor	Poor	Average	Good	Good
Defoam	Poor	Poor	Poor	Aver-age	Good
Low pres-sure drop	Good	Good	Good	Good	Average
Ensure good gas distri-bution	Poor	Poor	Poor	Good	Average

Table 4.1: Inlet devices performance guidelines (NORSOK standard, 2001)

As it can be noticed vane type devices have the highest overall performance in various functions. This is the reason why it is widely used being the most popular choice. However, other devices such as half pipes and diverter plates can be selected due to their low cost and their relative good performance in less critical installations. Cyclonic are very efficient but too sensitive to surge so they need special design.

Inlet devices are typically selected and sized based on the inlet momentum ($\rho_m V^2_m$) of the separator feed stream. The intent is to reduce the energy/velocity of the feed fluids to provide conditions favorable for phase separation. Limits on $\rho_m V^2_m$ are suggested such as table 3.16 which used for nozzle sizing.

Bothamley (2013) used mostly anecdotal information obtained from various sources, mainly technical literature available in the public domain to quantify inlet device efficiency in relation to feed momentum producing fig.4.4.

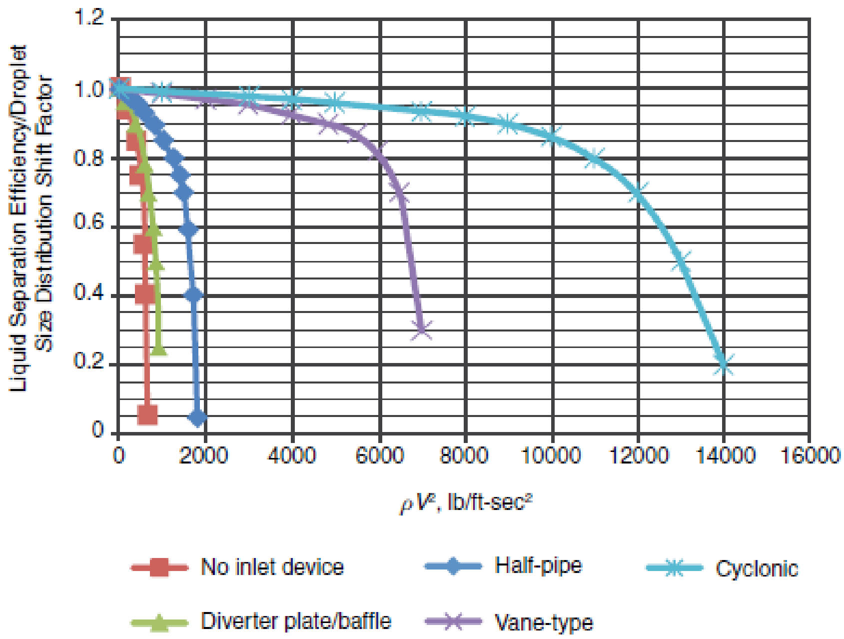

Figure 4.4: Inlet device liquid separation efficiency and effect on droplet sizes (Bothamley M., 2013).

The amount of unseparated liquid as predicted by Fig. 4.4 is assumed to be in the form of entrained droplets immediately downstream of the inlet device (at the entry to the gas gravity separation section). The fall off in separation performance with increasing feed ρV2 reflects a larger fraction of the feed liquid in entrained droplet form, breakup of bulk liquid into droplet form, and smaller droplet sizes associated with the higher inlet velocities. As far as droplet size distribution of unseparated liquid downstream of inlet device is concerned, there are limited literature data. Hence, a simplified approach is to define a droplet size distribution shift factor (fig.4.4) which should be a reasonable reflection of the droplet shattering/shearing effect of the different inlet devices.

Another primary performance attribute of inlet device that must be approached is the downstream flow distribution improvement. Inlet internals may straighten out the velocity profile of the inserted continuous phases and promote droplet coalescence or foam collapse to improve separation performance and/or reduce separator size. In real conditions, actual velocity profile rarely agrees with the ideal plug flow calculated by volumetric flow rate (V=Q/A). Only experimental tracer surveys and computational fluid dynamics (CFD) can accurately simulate this effect. Another simplified approach is fig.4.5.

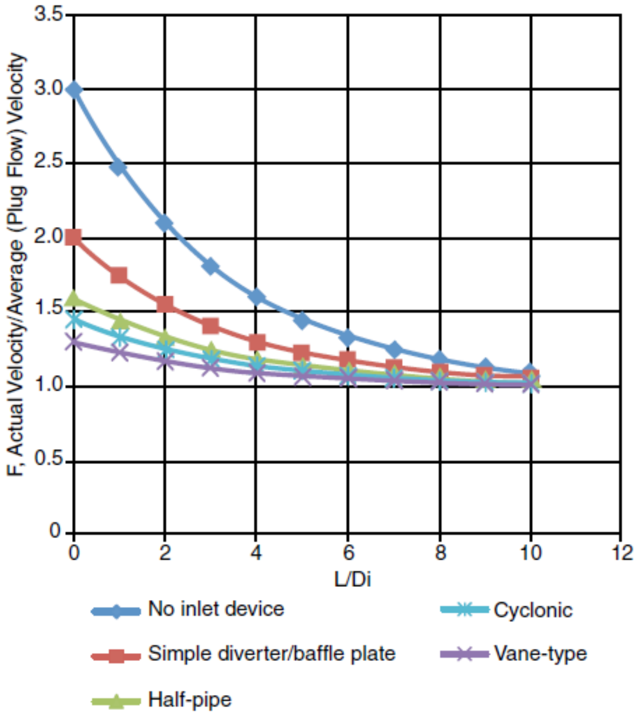

Figure 4.5: Effect of inlet device on downstream gas and liquid velocity profiles (Bothamley M., 2013)

Fig.4.5 offers an estimation of the quality of the flow distribution immediately after the inlet device (L/Di=0) and of the flow profile with distance downstream of the inlet device. The quality of the flow distribution is characterized by the factor F, the actual average velocity/ideal plug flow velocity. F values greater than 1.0 imply unused cross-sectional flow area. Use of this factor will allow estimation of the effective actual velocity, which can then be used in the droplet settling calculations for the gas (and liquid) gravity sections. The calculated effective actual velocity for the gas gravity section will be the entrance velocity to the mist extractor section (Bothamley M., 2013).

<u>Gas gravity section</u>

The entrained liquid load not removed by the inlet device is reduced in gas gravity separation section. In addition, this section straightens gas velocity profile. Droplet settling theory that has already been analyzed in Chapter 2 and used for separators sizing, is the approach that can more accurately quantify this section performance.

The droplet settling method aims at removing a target liquid droplet size and all droplets larger than the target size, after some assumptions to simplify the calculations. With the droplet size distribution and effective actual gas velocity profile determined by one of the methods described in the inlet section the droplet settling calculations are relatively straightforward. Even this approach is only semi-quantitative. The selection of an appropriate target droplet size is an inherently ill-defined procedure, and provides little indication as to the amount of entrainment remaining.

Another vague assumption is the determination of the effective release point of the droplets, namely the exact initial position of a droplet to settle for a horizontal vessel. The relatively chaotic gas flow patterns expected to be exiting from the inlet device cannot warranty any specification of the position/height of the release point. The most conservative approach would be that all droplets are released at the top (inlet) of the gas gravity separation section and a more optimistic approach that the release of entrained droplets occurs uniformly over the vertical height of the gas gravity separation section. A release point closer to the gas/liquid interface results in lower entrainment as droplets can be easier removed.

An additional uncertainty introduced to droplet settling theory application is liquid re-entrainment prediction. There are technical guidelines to avoid liquid re-entrainment such as slenderness ratio limits. Besides, there are complex correlations in literature to estimate re-entrainment fraction similarly to those used to predict entrainment (Viles, 1993).

<u>Liquid settling section</u>

The functions that take place in liquid settling section are water-oil separation and liquid degassing. Droplet settling theory is again applied for liquid droplets. Stokes law can be assumed to simplify calculations. As it has been referred oil droplet removal from water phase is much easier than water droplets from oil phase due to viscosity difference. Thus, water carry over in oil is more important than oil carry over in water. However, there is the same uncertainty of droplet distribution and the selection of a standard droplet size removal. The real challenge is to predict droplet size distribution in the emulsion where experimental analysis and field data are needed for reliable determination.

There is no much information about liquid degassing estimation in literature but some liquid residence time limits can be specified. As with liquid droplets in the gas phase, the difficulty is estimating how much gas is entrained in the liquid and the size distribution of the bubbles. The entrained gas is minor in comparison to the total gas flow so that it does not affect separation performance significantly.

Mist Extraction Section

The mist extractor is the final gas cleaning device in a conventional separator. The selection and design to a large degree, determine the amount of liquid carryover remaining in the gas phase. The most common types include wire mesh pads ("mesh pads"), vane-type (vane "packs") and axial flow demisting cyclones (fig.4.6).

These droplets are typically less than 150–500 micron in size and usually much smaller. It is generally not economic to separate these droplets by gravity alone by making the separator larger. The different types of mist extractors use principles other than simple droplet settling by gravity to achieve efficient removal of small droplets.

Mesh pads are the most common types of mist extractors used in vertical separator applications. The primary separation mechanism is liquid impingement onto the wires, followed by coalescence into droplets large enough to disengage from the mesh pad (fig.4.7). Mesh pads are not recommended for dirty or fouling service as they tend to plug easily. They capture small droplets with high efficiency and are preferred to low gas capacities. Their cost is the lowest among the other devices while they cause relatively low pressure drop.

Vane packs, like mesh pads, capture droplets primarily by inertial impaction. The vane bend angles force the gas to change direction while the higher density liquid droplets tend to travel in a straightline path, and impact the surface of the vane where they are collected and removed from the gas flow. They can handle higher gas capacity, solids and they are especially suited for low pressure applications. They are though more expensive than mesh pads and less efficient for small droplets.

Cyclonic mist extractors use centrifugal force to separate solids and liquid droplets from the gas phase based on density difference. Very high G forces are achieved which allows for efficient removal of small droplet sizes. These units can be more efficient than either wire-mesh or vanes and are the least susceptible to plugging. Besides, they are better for high pressure and high gas capacity applications. However, they are not in common use in production operations because their removal efficiencies are sensitive to small changes in flow. In addition, they require relatively large pressure drops to create the centrifugal force.

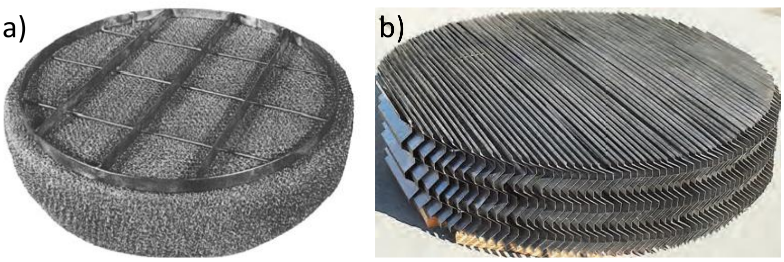

Figure 4.6:a)Mesh pad demister b) Vane type demister

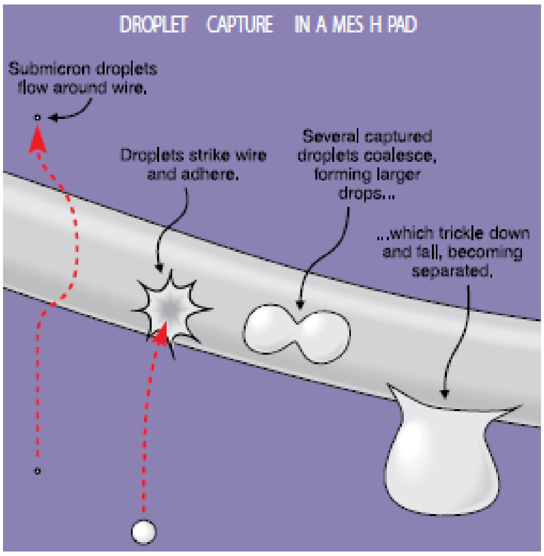

Figure 4.6: Mist elimination using wire mesh mist extractor

Analysis will focus to mesh pad performance because it is the most widely used mist eliminator. There are two main aspects to mesh pad separation performance:

- Droplet removal efficiency: The ability to remove smaller droplets will correspond to less droplet penetration through the mist extractor and, therefore, less carryover of liquid into the separated gas. The required removal efficiency and tolerable amount of carryover is determined by the sensitivity of the downstream process or equipment to the liquid content of the gas.
- Gas handling capacity: It is usually determined by the maximum allowable velocity before re-entrainment becomes excessive. This velocity can be determined by a load or sizing factor, Ks, as utilized in the Souders and Brown equation given below:

$$V_{max} = K_s \sqrt{\frac{\rho_l - \rho_g}{\rho_g}} \quad (4.1)$$

  Where Vmax maximum velocity (ft/s) , Ks load factor (ft/s) and ρl,ρg liquid and gas phase density (lb/ft^3)
- Liquid handling. This refers to the amount of entrained liquid load ($gal/min/ft^2$) that the mist extractor can handle before separation performance is substantially degraded (usually because of re-entrainment).

It is in other words the Ks -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. Efficiency and capacity are normally inversely related, i.e. as droplet removal efficiency increases, allowable gas throughput decreases.

This is an oversimplification. Droplet removal efficiency is typically given by the manufacturer as a curve or tables (table 4.2) showing % removal as a function of droplet size at design flow and a nominal liquid loading. Each mist extractor type has its own Ks factor. Ideally, there should be correlations/equations that quantify Ks values for mist extractors as a function of the mist extractor design/construction details and the in-situ fluid properties and flow conditions. Although some information is available about this, further work is required. Most mist extractor capacity information is based on low pressure air/water tests, which do not scale up well to real world conditions.

The recommended value of Ks varies and depends upon several factors such as liquid viscosity, surface tension, liquid loading, and operating pressure. Each manufacturer has its own recommended values.

A methodology that can be used to quantify droplet capture efficiency of a mesh type mist extractor based on inertial impaction is the following:

1) Select the appropriate load factor (table 4.2&4.3) and calculate the ideal vapor velocity with eq. (4.1). Then, the cross-sectional area of mist eliminator is readily determined by dividing the volumetric flow rate by the velocity.
2) Calculate the inertial impaction parameter Stk (Stokes number) as follows:

$$Stk = \frac{\left(\rho_l - \rho_g\right)\left(d_p{}^2\right)V_g}{9\mu_g d_w} \quad (4.2)$$

Where d_w=wire diameter or thickness, V_g=design velocity of the wire mesh, μ_g=gas viscosity, d_p=target droplet size and ρ phase density

3) Calculate the singlewire removal efficiency from Fig. (Langmuir and Blodgett, 1946) assuming that the motion of the droplets relative to the gas phase is governed by Stokes'law.

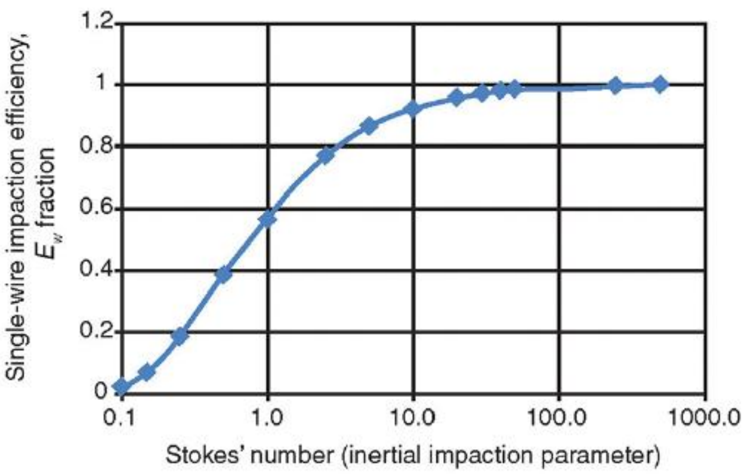

Figure 4.7: Single wire droplet capture efficiency (Langmuir and Blodgett, 1946)

4) Find the specific surface area, S for the mesh style of interest (tables 4.2 & 4.3) and determine SO, the area of the area of the mist eliminator perpendicular to vapor flow with a correction factor of 0.67 to remove that portion of the knitted wire not perpendicular to the gas flow.

Description	Density, kg/m³, (lb/ft³)	Voidage, %	Specific surface area, m^2/m_3, (ft^2/ft^3)	K_s, m/s, (ft/sec)	Separable droplet size, 90% removal, microns	Liquid Load Before Capacity Deteriorates, L/min/m² (gal/min/ft²)
"Standard" mesh pad	144 (9)	98.5	279 (85)	0.107 (0.35)	5	31.5 (0.75)
"High capacity" mesh pad	80 (5)	99.0	148 (45)	0.12 (0.4)	8 – 10	63 (1.5)
"High efficiency" co-knit mesh pad	192 (12)	96.2	365 (12 000)	0.07 (0.22)	2 – 3	21 (0.5)

Table 4.2: Mesh pad characteristics and performance parameters (Snow-Mcgregor, 2014)

Density	Surface Area	%Voids	Wire Dia	Mesh Styles			
Lb/Ft³	Ft²/Ft³		inch	Koch	Otto York	ACS	Description
12.0	115	97.6	.011	4120		4BA	
10.8	110	97.7	.011	4210	421		All Around, Heavy Duty
10.0	163	94.0	.006	3710	371		Liq-Liq Coalescer, Fog
9.0	86	98.2	.011	4310	431	4CA	Standard, good all around
8.0	140	98.4	.006	3260	326	3BF	Super High Eff, fine mist
7.3	65	98.5	.011	6440	644		High Eff - anti fouling
7.0	65	98.6	.011	5310	531	5CA	Economy Performance
5.0	48	99.0	.011	9310	931	7CA	High Thruput
20.0	450	96.0		5520		X200	
27.0	610	94.6		5540	333	X100	

Table 4.3: Mesh Styles (www. oilgasprocessing.com)

Table 4-3 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).

$$SO = 0.67\ \frac{t_w}{\pi} S \qquad (4.3)$$

Where t_w mesh pad thickness.

5) Convert the singlewire capture efficiency into the meshpad removal efficiency using the following equation based on the work by Carpenter and Othmer (1955): $E_{pad} = 1 - e^{-SO*Stk}$ (4.4)

The Ks capacity factor for mesh pads is often derated for higher pressure operation, (table 4.3). All factors being equal, this is normally due to the reduction in surface tension of the liquid phase that occurs with increasing pressure. Mesh pads normally operate efficiently over a range of 30–110% of the design gas rate. Among other things, Ks is also a function of the amount of entrained liquid reaching the mesh pad. As would be expected, Ks decreases with increasing inlet liquid loading. However, per the earlier discussion on droplet size distributions, it is difficult to predict what the inlet liquid loading reaching the mist extractor will be for a given separator application.

Pressure, kPa	Pressure, psia	K_s factor, % of design value
100	14.7	100
500	73	94
1 000	145	90
2 000	290	85
4 000	580	80
8 000	1 160	75

Table 4.4: Ks factor duration due to pressure

High liquid load/flooding will increase the pressure drop significantly. For efficient operation the demister Ks value must generally be below 0.1 m/s. For typical mesh pad designs, liquid loads greater than 1gpm/ft^2 are considered high and will require deration of the standard Ks factor, to prevent excessive entrainment carryover. The mesh can be of metal or plastic material, or a combination. Typical minimum droplet removal size is:

Metal mesh: 10 micron

Plastic/fibre: 3-5 micron

Wire mesh pads are normally installed horizontally with gas flow vertically upwards through the pad. Most installations will use a 6-inch thick pad with 9-12 lb/ft^3 bulk density with typical pressure drop of 0.1-3.0 millibar. Minimum recommended pad thickness is 4 inches. They are usually constructed from wires of diameter ranging from 0.10 to 0.28 mm, with a typical void volume fraction of 0.95 to 0.99.

4.2 HYSYS Carry Over Calculation Tools

The previous section introduced some literature information about real separation simulation and specifically about carry-over calculation. It can be inferred the complexity of this subject and the need for more research about it. Only simplifications and assumptions can be applied to produce correlations and technical guidelines for carry-over estimation while more of them require field and experimental data for reliable results. Based on them, inlet and mist elimination devices are designed. Computational fluid dynamics can possibly enrich existing knowledge in the future.

HYSYS offers two sets of correlations to calculate carry-over, “Horizontal Vessel” and “ProSeparator”. Both of them require separator dimensions, nozzle setup and pressure drop specification. For droplet size distribution, a modified Rossin-Rammler distribution method is applied by the software. These methods will be analyzed separately below based on Rojek & Tiwary Hysys Guide for real separators (2004):

<u>Rossin-Rammler droplet size distribution</u>

In order to properly analyze the particle size distribution in a system, the engineer must characterize these particles by collecting particle size measurements. Where such data is not available the engineer can resort to known typical values for the system in question. As a last resort the default values provided in HYSYS can be used but great care should then be taken in interpreting the results.

According to Karabelas (1978), droplet size spectra generated in turbulent pipe flow of dilute liquid /liquid dispersions can be well represented by a Rosin-Rammler type of equation with constant parameters and only one variable, the maximum drop diameter dmax or d_{95}. The latter can be predicted independently as a function of flow conditions and liquid physical properties (table 4.5). For engineering calculations in particular, the drop size spectrum can be estimated by using the Rosin-Rammler equation with a constant slope n=2.5 and the computed d_{95}.

$$\frac{d_{95}}{D} = 4.0\ We^{-0.6}$$

where $We = D\rho_c\overline{U^2}/\sigma$ is a dimensionless Weber number based on pipe diameter and mean velocity.

Table 4.5: d_{95} calculation (Karabelas, 1978)

Based on Karabelas’ work, HYSYS uses a modified Rossin-Rammler function to estimate droplet size distribution. Rossin-Rammler distributions are defined by:

$$F = \exp(-d/dm)^{x} \quad (4.5)$$

Where:

- F = fraction of droplets larger than d
- dm is related to d_{95}. It is the commonly occurring diameter (peak of the histogram / frequency curve), as compared to other different measures of central tendency such as mean or median diameters.
- x = RR index
- d_{95} = 95% of droplets are smaller than this diameter for the specified dispersion
- RR Index = exponent used in the RR equation (also known as the “spread parameter”). Spread is a measure of degree of deviation from the central tendency; its value is characteristic of the substance / system being considered

Another way of expressing this is:

$$\ln(F) = (-d/dm)x$$

or

$$\ln(\ln(1/F)) = B + x\ln(d) \quad (4.6)$$

Where: B = constant = ln (1/dm)

Therefore plotting ln(ln(1/F) against ln(d) can be used to calculate the R-R parameters. HYSYS correlations ask the user for d_{95} data rather than dm. If required dm can be relatively simply calculated from d_{95} as follows:

$$dm = d95 / (-\ln(1 - 0.95))^{(1/x)} \quad (4.7)$$

It must be emphasized that the use of continuous size distribution functions to represent experimental data is almost always a compromise, since measured data rarely fit the models exactly. However, distribution functions have the advantage that they enable the comparison of a large amount of data using a few basic parameters. An important feature is the ability to represent the size distribution in cumulative form as a straight line by means of a scaling that is constructed so as to linearise the cumulative size distribution function.

<u>Horizontal Vessel</u>

The Horizontal Vessel correlations were developed for a horizontal three-phase separator. For the inlet calculations the Horizontal Vessel correlations calculate the six types of dispersions in the feed according to an assumed efficiency of a user-defined inlet device, and user-defined dispersion fractions termed “Inlet Hold up”. The droplet distribution of the dispersed phase is then calculated using user-supplied Rosin-Rammler parameters. The droplet d95 of the liquid-liquid dispersions is not specified but calculated using the inlet droplet d95 and the densities of the two liquid phases.

The primary gas-liquid separation in the gravity section and the liquid-liquid separation in the settling section are calculated from the droplet settling theory. A droplet is carried over if the vertical distance traveled during its residence in the vessel is less than the vertical distance required to rejoin its bulk phase. This effectively applies to horizontal vessels.

The settling velocities are calculated using the GPSA correlations (table 4.5) for all dispersions, except for the water in oil dispersion for which the settling velocity is calculated by the method of Barnea and Mizrahi. A user defined liquid phase inversion point is used in the calculation of the appropriate liquid phase viscosities (i.e. water-in-oil and oil-in-water). The inversion point is the water fraction at which the system changes in behaviour from a water-in-oil emulsion to an oil-in water emulsion. In many cases it is observed to occur at 50-70% water; however there is no reliable means of determining the actual point and it must usually be determined experimentally. A residence time correction factor can also be applied.

Settling Law	Reynolds Number (Re_p)	Terminal Velocity Equation
Stokes' Law	<2	$V_t = \frac{g d_p^2 (\rho_l - \rho_g)}{18 \mu_g}$ ....(4)
Intermediate Law	2–500	$V_t = \frac{0.1529 g^{0.714} d_p^{1.142} (\rho_l - \rho_g)^{0.714}}{\rho_g^{0.286} \mu_g^{0.428}}$ ....(5)
Newton's Law	500–200,000	$V_{max} = K_s \sqrt{\frac{\rho_l - \rho_g}{\rho_g}}$ ....(6)
Note: For calculations involving separation of gas bubbles from liquid, the gas viscosity (μ_g) in the equations is replaced with the liquid viscosity.		

Table 4.6: Settling velocity equations for settling law regions (Gas Processors Suppliers Association, EDB, 2014)

The secondary gas liquid separation (mist eliminator) calculations for the gas phase are defined by a user-defined critical droplet size. The gas loading factor of the device is used to calculate its size.

<u>ProSeparator</u>

ProSeparator correlations are based on several SPE papers (see reference section) and proprietary research from BP Exploration and Operating Co. They have been successfully applied in many BP operations offering benefits in debottlenecking applications and separator configurations assessment.

Their main advantage is that they include droplet size distribution calculation in contrast to conventional methods of sizing which set critical droplet sizes. ProSeparator correlations are rigorous but are limited to calculating liquid carry over into gas. There are no calculations of liquid-liquid separation or gas entrainment in the liquid phases (they are set to zero). Light liquid and heavy liquid entrainments are calculated separately and the total carry over is the sum of the separate light and heavy liquid carry over calculations.

Minimum and maximum droplet diameters are calculated based on inlet flow conditions (inlet gas flow rate and gas/liquid phase physical properties such as density and surface tension) and inlet pipe size. Maximum droplet size is determined with ProSeparator using empirical correlations (table 4.6). Accurate physical properties of the fluids (particularly surface tension) are very important to this calculation.

$$\frac{D_{p(\max)}}{d_{pipe}} = k \cdot \frac{R_e^{0.1}}{W_e^{0.6}} \cdot \left[\frac{\rho_g}{\rho_l}\right]^{0.6} + 4.0 \cdot \frac{G_{le}}{\rho_l \cdot U_g} \qquad (3)$$

Where :-

$$R_e = \frac{\rho_g \cdot U_g \cdot d_{pipe}}{\mu_g} \qquad (4)$$

$$W_e = \frac{\rho_g \cdot U_g^2 \cdot d_{pipe}}{\sigma} \qquad (5)$$

The value of k and the last two terms of Formula (3) have been simplified/revised to match operational data.

Table 4.7: Inlet pipework maximum stable particle size and distribution formula (Baker M. & Dick D.R., 1991)

The droplet distributions of light and heavy liquids in the inlet gas are then calculated using a Rossin-Rammler type distribution. It must be noted that ProSeparator effectively calculates its own Rossin-Rammler parameters (droplet diameters), fitting them to match the predetermined minimum and maximum droplet sizes and does not require the user to specify any of these parameters. The only user input in the inlet calculations is the ability to limit the amount of phase dispersion calculated.

Primary separation is based on critical droplet size; however, the critical droplet size is not user-specified but calculated based on the gas velocity through the vessel. It is determined from the terminal velocity of the droplets as calculated from the inlet gas velocity, vessel dimensions, and fluid properties (liquid & gas density, gas viscosity).

Secondary separations accomplished using exit devices are calculated by device specific correlations. Demister characteristics are required to be input. In this study mesh pad device was selected which uses Carpenter and Othmer approach described in the previous section (eq.4.4).

4.3 HP Separator Carry Over Calculation – Internals Design

High pressure separator has been sized in section 3.3 and its design basis has been set so that HYSYS software tools can be applied to estimate carry over in the first separation stage. These results will prove the adequacy of sizing procedure while they can be used for the design of the internals and of the next separation devices. Literature generally indicates the following upper carry over limits for separators:

- free gas of up to 2% v/v in separated oil
- 1000ppm oil in water
- 0.1 gal/ MMscf/D liquid in gas= $1,33*10^{-8}$ %v/v
- 30-5% v/v water in oil in high pressure, 5% v/v for medium pressure and 0.5% v/v for low pressure separators

Operational problems associated with too much liquid in the gas are far more prevalent than those caused by too much gas in the liquid. Especially, liquid carry over in gas can have destructive effect on compressors. Besides, as it was analyzed in droplet settling theory, separation of water droplets from oil phase is much more difficult than those of oil droplets from water and liquid droplets from gas phase due to the relatively much higher viscosity of oil. Final product specifications always define a maximum of 0.5%v/v water in oil product. Hence, this analysis will focus on liquid in gas and water in oil entrainment. The other carry over limits are less crucial and usually are easily satisfied with a conservative separator design as in this study.

"Proseparator" method of HYSYS was selected to simulate gas/liquid separation in gas gravity section and mist elimination device. "Horizontal vessel" method was selected to simulate water/oil separation in liquid settling section. With the absence of experimental data, inversion point was set to 60% but it is not expected to affect calculation due to the low water content of the emulsion. Inlet nozzle sizing of section 3.3 refers to an inlet momentum of 3600 (lb/(ft*sec^2) . Selecting the most advantageous inlet diverter type, vane type, an efficiency of 97% is estimated from fig.4.4. Separator dimensions and nozzles (diameter and position) are inserted according to section 3.3. Pressure drop is calculated using "horizontal vessel" method. Inlet droplet d95 diameter for Rossin –Rammler distribution was set to the conservative value of 0,0315 in.(800 micron).

The most vulnerable production states for each carry over case have to be chosen for examination. Higher water flows lead to more water droplets entrained in oil so that the production state with the highest watercut (state 5) is the worst scenario for water in oil entrainment. If separator performance satisfies carry over specifications in this state, it will readily satisfy the requirements in the earlier states, too. Considering the most conservative scenario that the water phase is totally dispersed in the oil phase in the inlet due to the lack of emulsion field data, the calculated water in oil product carry over is 0.00443%v/v, lower than the final product specification. The produced feed and product droplet distribution are shown in fig.4.8. It can be noticed that all droplets bigger than 0.0094 in (240 micron) have settled. These results reveal that sizing procedure led to a sufficient but conservative separator size. Furthermore, it proves that water in oil specification can be easily met by the first separator due to the low water cut ratios of production so that lower pressure separator can be replaced by two phase separators.

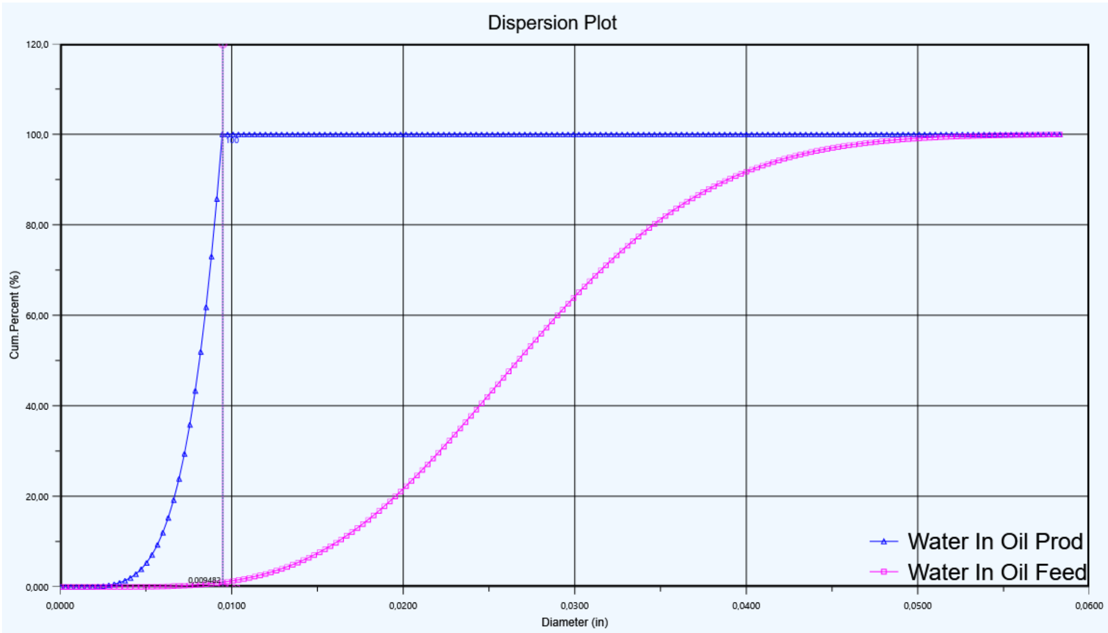

Figure 4.8: Water in oil droplet distribution

As far as liquid in gas carry over is concerned, it is expected to be maximized to the earlier production states where oil flows are higher. Production state 2 was found to have the highest liquid entrainment in gas without using demister and specifically a value of 1,72E-08 %v/v liquid in gas. Applying a standard mesh pad demister Otto York 431 from fig. 4.3 with 6 in. thickness the resultant carry over was reduced to 9,17E-09 %v/v meeting specifications. Liquid droplet distributions are presented in table.4.8. It can be inferred that mesh pad partially removes droplets larger than 4.5 μm while the very small observed droplet sizes prove that gas capacity constraint does not govern. Although sizing consideration was the removal of 100 micron, liquid capacity constraint led to a longer separator removing smaller droplets.

Droplet Distribution	Oil in Gas		Water in Gas	
	Drop. Diam. (micron)	Cum. Per. (%)	Drop. Diam. (micron)	Cum. Per. (%)
Mesh Pad Demister Inlet	0,23	6,52E-07	0,23	1,25E-06
	2,25	0,12	2,25	0,22
	4,50	2,20	4,50	4,20
	6,75	12,74	6,75	24,32
	9,00	45,97	9,00	87,77
	11,25	100,00	11,25	100,00

Mesh Pad Demister Outlet	0,23	1,21E-06	0,23	2,14E-06
	2,25	0,19	2,25	0,34
	4,50	3,12	4,50	5,51
	6,75	15,94	6,75	28,14
	9,00	50,92	9,00	89,81
	11,25	100,00	11,25	100,00

Table 4.8: Liquid droplet distribution in gas

4.4 MP and LP Separators Design

It was deduced from the previous section that medium and lower pressure separators have to be two phase separators. As in chapter 3, MP and LP separators were set to ideal separation mode receiving the output of the previously calculated high pressure separator. Simulation was carried out to define the design basis for sizing them. The sizing method described in section 2.2.3 was used in combination with technical guidelines (Svrcek W., 2003 and Lake, 2015) to design the separators. The design bases which were formed are described in the following table:

Design Bases:	**MP separator**	**LP separator**
Qg (MMscfd)	0,22583	0,968
Qo (BOPD)	11464	11352
Po (psia)	125	45
To (F)	120	120
Sg	0,820	1,010
API^{o}	30	28
(SG)o	0,74	0,74
$d_{m,l}$ (microns)	140	140
$(tr)_o$ (min)	2	2

Table 4.9: MP and LP separators design bases

Horizontal two phase design was selected due to the low gas-oil ratio. The iterative procedure for C_D calculation which was described in section 2.3.3 was applied. Its results are given in the next table:

MP separator C_D calculation			
ρl (lb/ft3)	46,18	Z	0,97
ρg (lb/ft3)	0,49	μ gaz (cp)	0,012
C_{D1} (initial value)		0,34	
	V (ft/s)	Re	C_D
Iteration1	2,33	272,52	0,61
Iteration2	1,74	203,49	0,67
Iteration3	1,66	194,39	0,68
Iteration4	1,65	192,89	0,68
Iteration5	1,64	192,64	0,68

LP separator C_D calculation			
ρl (lb/ft3)	46,18	Z	0,99
ρg (lb/ft3)	0,21	μ gaz (cp)	0,011
C_{D1} (initial value)		0,34	
	V (ft/s)	Re	C_D
Iteration1	3,54	196,55	0,68
Iteration2	2,51	139,38	0,77
Iteration3	2,36	130,92	0,79
Iteration4	2,33	129,31	0,79

Table 4.10: MP and LP separators C_D calculation

Effective lengths for both liquid and gas capacity constraint were calculated (table 4.11). The low values of gas capacity constraint indicate that liquid capacity constraint governs. According to guidelines slenderness ratios between 3 and 4 have to be selected. For both separators, calculations were concluded to a diameter of 54 in. and length of 15ft.

MP separator Diameter vs Length calculation per capacity constraint				
do (in.)	Gas Leff (ft)	Liquid Leff (ft.)	Lss (ft)	SR (12Lss/do)
42,00	0,07	18,6	24,8	7,1
48,0	0,06	14,2	19,0	4,7
54,00	0,06	11,2	15,0	3,3
60,0	0,05	9,1	12,1	2,4
LP separator Diameter vs Length calculation per capacity constraint				
do (in.)	Gas Leff (ft)	Liquid Leff (ft.)	Lss (ft)	SR (12Lss/do)
42,00	0,63	18,4	24,5	7,0
48,0	0,55	14,1	18,8	4,7
54,00	0,49	11,1	14,8	3,3
60,0	0,44	9,0	12,0	2,4

Table 4.11: Separator length calculation for both liquid and gas capacity constraint

Nozzles were sized according to the procedures of section 3.3 while their position was defined according to fig.4.9 (Svrcek W., 2003). Table 4.12 includes the results.

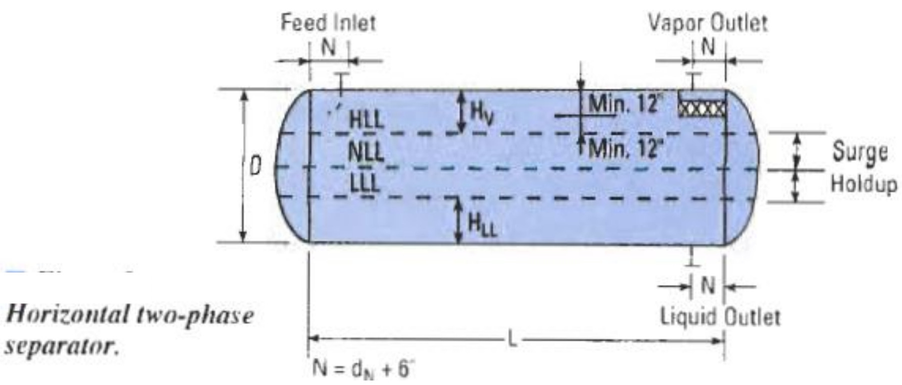

Figure 4.9: Horizontal two phase separator scheme (Svrcek W., 2003)

MP and LP separator nozzles design		
inlet ρmin (lb/ft^3)	12,76	9,92
inlet mass flow (lb/hr)	137894,83	96544,48
inlet Dn (inch)	5,73	5,10
gas outlet ρmin (lb/ft^3)	0,43	0,14
gas outlet vol flow (MMscfd)	1,15	0,81
gas outlet Dnmin (inch)	5,68	3,61
liq outlet ρmin (lb/ft^3)	23,02	24,17
liq outlet mass flow (lb/hr)	119658,04	112339,76
liq Dnmin (inch)	9,40	8,89
N inlet (inch)	12	11
N gas outlet (inch)	12	10
N oil outlet (inch)	15	15

Table 4.12: MP and LP separator nozzles design

Sizing results and nozzles set up were inserted to HYSYS to estimate liquid carry over in gas. "Proseparator" method of HYSYS was selected to simulate gas/liquid separation in gas gravity section and mist elimination device. Similarly to HP separator, selecting the most advantageous inlet diverter type, vane type, an efficiency of 97% is estimated from fig.4.4. The production state with the maximum liquid carry-over in gas has to be examined for demister design. This state was found to be production state 1. Calculations proved that MP separator requires a higher efficiency mesh-pad demister of 6 in. thickness (Otto York 326, fig. 4.3) while LP separator a standard mesh pad of 8in. thickness (Otto York 431, fig. 4.3) to satisfy liquid in gas ratio specification (1.33* 10^-8 v/v). Their effect to droplet distribution can be noticed in the following table.

Droplet Distribution		Demister Inlet	Demister Exit
	Drop. Diam. (micron)	Cum. Per. (%)	Cum. Per. (%)
MP separator	2	0,05	0,51
	5	0,85	6,13
	7	4,90	23,36
	10	17,67	53,13
	12	48,77	84,28
	15	100,00	100,00
LP Separator	2	0,02	0,18
	5	0,37	2,71
	7	2,17	12,74
	10	7,82	34,92
	12	21,59	64,29
	15	50,04	87,67
	17	100,00	100,00

Table 4.13: MP and LP Liquid droplet distribution in gas

5. Gas Compression Train

In this chapter the procedure used to simulate the compression train in HYSYS will be described. The new operation conditions will be calculated. Finally, all compression equipment will be sized according to literature.

5.1 Gas Compression Train Set Up in HYSYS

The HYSYS flowsheet which was designed for gas compression is presented in fig.5.1. The simulation of gas compression train in HYSYS requires the addition of two theoretical blocks, the Recycle and Set block apart from mixers, compressors, scrubber and coolers.

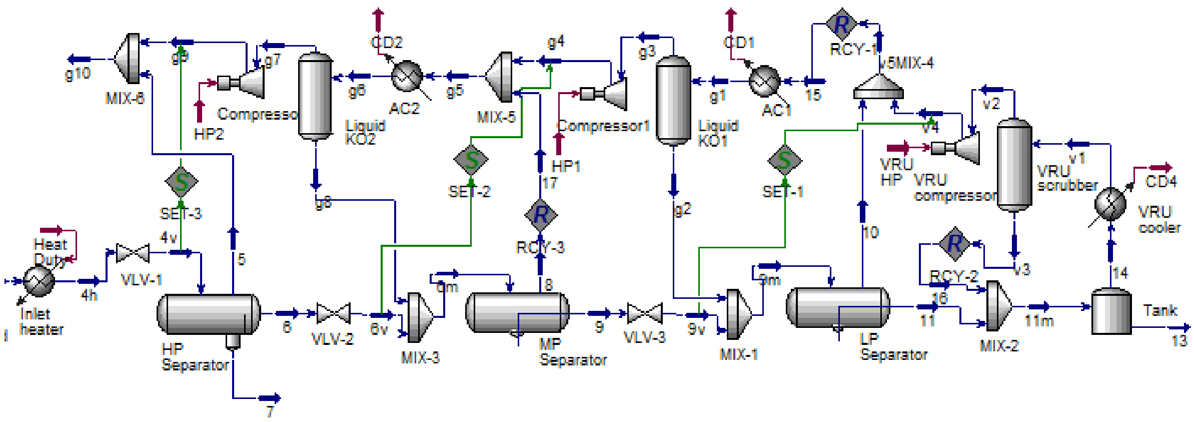

Figure 5.1: Separated Gas Compression Train in HYSYS

Recycle

Material recycles, where downstream material mixes with upstream material, require a “Recycle” operation. The “Recycle” installs a theoretical block in the process stream. The stream conditions can be transferred forward between the inlet and outlet streams of this block. In terms of the solution, there are assumed values (outlet streams) and calculated values for each of the variables in the inlet streams.

The following steps take place during the convergence process:

1. HYSYS uses the assumed values and solves the flowsheet around the recycle.
2. HYSYS then compares the assumed values in the attached streams to the calculated values in the opposite stream.
3. Based on the difference between the assumed and calculated values, HYSYS generates new values to overwrite the previous assumed values.
4. The calculation process repeats until the calculated values match the assumed values within specified tolerances.

The tool of “Recycle Adviser” ensures that flowsheet contain the minimum number of recycles at their optimal locations to minimize calculation time. The Adviser picks the best location for recycles and assigns the best calculation order for optimized convergence. Thus, three recycle loops were defined: tank - vapor recovery unit (VRU) scrubber, LP separator – liquid knock-out1, MP separator-liquid knockout-2.

Assumed values were estimated by running models with unconnected inlet streams from “Recycle” blocks. The calculated values of these streams were set as initial values for the outlets of recycle blocks.

Set

The “Set” is an operation used to set the value of a specific process variable in relation to another process variable. It was applied to equal the pressure of the outlet of pressure control valve of each separation stage to the respective discharge pressure of compressor. As it was explained in theory chapter, gas has to be compressed to the operating pressure of the preceding separation stage. Pressure drop in separator, scrubber and inter-stage cooler were neglected in order the recycle loop to be able to converge.

Compressor (VRU compressor,Compressor1,Compressor2)

All compressors were set as reciprocating with adiabatic efficiency of 82%. Vapor Recovery Unit (VRU) compressor is actually a rotary vane type compressor. HYSYS does not include such an option so that VRU was simulated as reciprocating, too. More information about compressor design will be introduced in section 5.2.

Coolers(VRU cooler,AC1,AC2)

Inter-stage coolers were initially simulated by simple theoretical cooler blocks with specified standard temperature outlet of 95F. In reality, heat exchanger or air coolers are used upstream of reciprocating compressors. More information about heat exchanger design will be introduced in section 5.2.

Scrubbers (VRU scrubber, Liquid KO1 and KO2)

Scrubbers or liquid knockout vessels are simulated as vertical two phase separators due to the expected high gas to oil ratios. They will be sized in section 5.2.

Duty streams

Compressors, heaters and coolers require specification of inlet energy streams to represent the duty of the device.

Mixers

Mixers were used to collect liquid recycle streams and direct them to the respective separator with the main inlet streams.

5.2 Gas Compression Train Design

The optimal operation conditions with liquid recovery from gas per production state were calculated. The same optimization method used in chapter 3 was applied. The separators and internals sizing results from the previous chapters will be used. Scrubbers are assumed to completely separate gas and liquid phase (ideal separation). The calculated gas properties will be used for compression equipment design. The production results of the simulation are summarized in the next table:

Separation Train Perfomance with liquid recovery from gas					
Optimum operation conditions per production state	state 1	state 2	state 3	state 4	state 5
HP Separator – Vessel Pressure [psia]	600	610	610	550	600
MP Separator – Vessel Pressure [psia]	115	269	282	115	115
LP Separator – Vessel Pressure [psia]	30	47	30	62	62
Tank Liq Vol Flow [barrel/day]	10171	9639	8727	7862	6943
Tank Liquid Density (API)	27,46	27,49	27,81	27,88	27,77
Tank Liquid Reid Vapour Pressure [psia]	10,0	9,8	9,7	9,5	9,4

Table 5.1: Oil production with liquid recovery from gas

Equipment design will focus on the production state with maximum gas and oil flow rates. It was figured out that this state is number 1. Gas stream properties are introduced in table 5.2:

Name	**5**	**8**	**10**	**14**	**g1**	**g2**	**g3**	**g4**
Temperature [F]	127,4	125,33	122,34	121,33	81,58	81,58	81,58	216,90
Pressure [psia]	600	115	30	14,7	30	30	30	115
Molar Flow [lbmole/hr]	797,6	106,83	35,24	9,49	44,74	2,10	42,64	42,64
Mass Flow [lb/hr]	15744	2617,6	1257,7	416,89	1674,9	38,474	1636	1636
Name	**g5**	**g6**	**g7**	**g8**	**g9**	**g10**	**v1**	**v2**
Temperature [F]	159,9	95	95	95	301,50	163,58	101	101
Pressure [psia]	115	115	115	115	600	600	14,7	14,7
Molar Flow [lbmole/hr]	149,47	149,47	147,83	1,64	147,83	945,44	9,49	9,49
Mass Flow [lb/hr]	4254	4254	4223	30,55	4223	19968	417	417
Name	**v3**	**v4**	**v5**	**15**	**16**	**17**		
Temperature [F]	101,33	159,73	131,57	131,58	101,33	125,33		
Pressure [psia]	14,7	30	30	30	14,7	115		
Molar Flow [lbmole/hr]	1,66E-05	9,49	44,73	44,74	1,66E-05	106,83		
Mass Flow [lb/hr]	2,94E-03	416,88	1674,90	1674,91	2,94E-03	2617,59		

Table 5.2: Gas stream resulted conditions according to fig.5.1

5.2.1Compressor inter-stage cooler

Aerial coolers were selected for compressor inter-stage coolers. Cooling in the vapor recovery unit of the tank is achieved by a more complex way explained in section 2.3. Therefore, its simulation was simplified by a theoretical cooler achieving a 20F temperature decrease. Aerial coolers are mechanically simple and flexible, and they eliminate the nuisance and cost of a cold source. In warm climates, aerial coolers may not be capable of providing as low a temperature as shell-and-tube exchangers, which use a cool medium. In aerial coolers the tube bundle is on the discharge or suction side of a fan, depending on whether the fan is blowing air across the tubes ("forced draft") or sucking air through them ("induced draft"). This type of exchanger can be used to cool a hot fluid to something near ambient temperature.

Louvers, fan variable speed drives, blade pitch or recirculation of process fluid. As the process flow rate and heat duties change, and as the temperature of the air changes from season to season and night to day, some adjustment must be made to assure adequate cooling while assuring that the process fluid is not over cooled. Too cool gas temperature could lead to hydrates forming and developing ice plugs in the cooler.

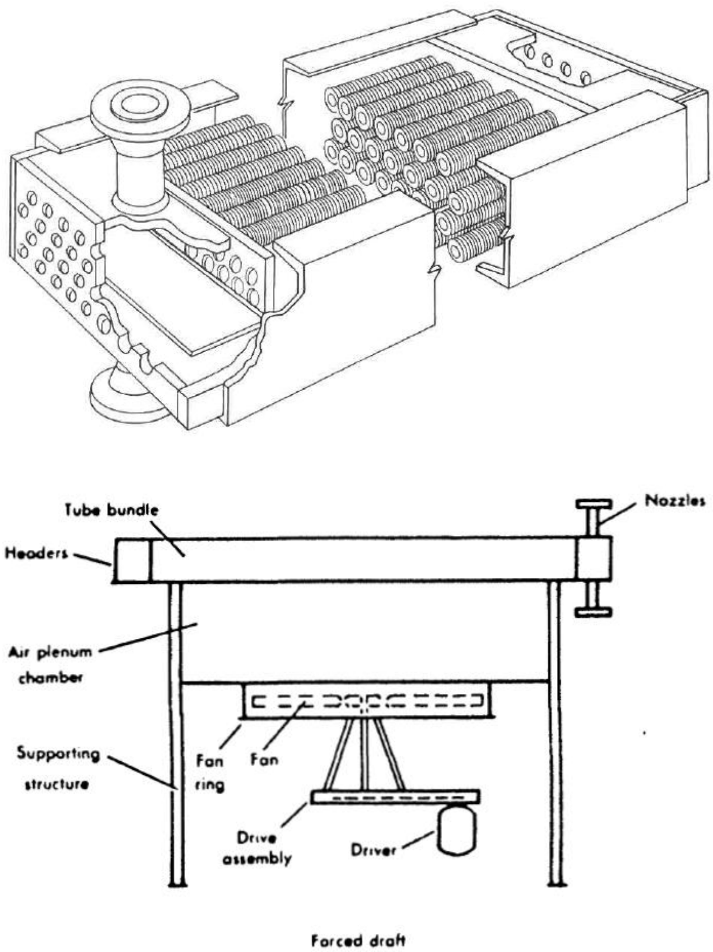

Fig5.2: Aerial cooler (Arnold K.,vol.2, 1986)

For an air-cooler design a first estimation of heat-transfer coefficient and air outlet temperature shall be made using table 5.3. The tubes have fins on them since air is non-fouling and it has very low heat transfer efficiency. The fins increase efficiency by effectively adding surface area to the outside surface of the tubes. U_b should be used when the outside surface area of the bare tube (neglecting fins) is used in the heat ing fins) is used in the heat transfer equation. U_x is used when the extended surface area including fins is used for the area term in the general heat transfer equation. From the method described in table 5.3, U_x=2.4 (fintube 1/2in. by 9) and $T_{airoutlet}$= 78 F were estimated for the first air cooler (AC1) while for the second U_x=2.4 and $T_{airoutlet}$=80 F.

Typical Overall Heat-Transfer Coefficients for Air Coolers

Service	Fintube ½ in. by 9 U_b	Fintube ½ in. by 9 U_x	Fintube ⅝ in. by 10 U_b	Fintube ⅝ in. by 10 U_x
Water & Water Solutions				
Engine jacket water (r_f = .001)	110	7.5	130	6.1
Process water (r_f = .002)	95	6.5	110	5.2
50-50 ethyl glycol-water (r_f = .001)	90	6.2	105	4.9
50-50 ethyl glycol-water (r_f = .002)	80	5.5	95	4.4
Hydrocarbon Liquid Coolers				
Viscosity C_p				
0.2	85	5.9	100	4.7
0.5	75	5.2	90	4.2
1.0	65	4.5	75	3.5
2.5	45	3.1	55	2.6
4.0	30	2.1	35	1.6
6.0	20	1.4	25	1.2
10.0	10	0.7	13	0.6
Hydrocarbon Gas Coolers				
Temperature, °F				
50	30	2.1	35	1.6
100	35	2.4	40	1.9
300	45	3.1	55	2.6
500	55	3.8	65	3.0
750	65	4.5	75	3.5
1000	75	5.2	90	4.2

$$\Delta T_a = \left(\frac{U_x + 1}{10}\right)\left(\frac{T_1 + T_2}{2} - t_1\right)$$

where ΔT_a = air temperature rise, °F

U_x = overall heat transfer coefficient based on extended area, Btu/hr-ft^2-°F

T_1 = process fluid inlet temperature, °F

T_2 = process fluid outlet temperature, °F

t_i = ambient air temperature, °F

Table 5.3: Typical heat transfer coefficient for Air coolers and air outlet temperature estimation (Arnold K.,vol.2, 1986)

HYSYS defines Air Cooler duty, q, in terms of the overall heat transfer coefficient, the area available for heat exchange, and the log mean temperature difference:

$$q = -UA(LMTD)Ft \qquad (5.1)$$

where:

UA = (overall heat transfer coefficient) x (surface area available for heat transfer)

ΔT_{LM} = log mean temperature difference = LMTD

$$LMTD = \frac{\Delta T_A - \Delta T_B}{\ln\left(\frac{\Delta T_A}{\Delta T_B}\right)} = \frac{\Delta T_A - \Delta T_B}{\ln \Delta T_A - \ln \Delta T_B} \qquad (5.2)$$

We assume that a generic heat exchanger has two ends (which we call "A" and "B") at which the hot and cold streams enter or exit on either side; ΔT_A is the temperature difference between the two streams at end A, and ΔT_B is the temperature difference between the two streams at end B. Using data from table 5.2 $LMTD_{AC1}$=43.58 F and $LMTD_{AC2}$= 44.33F.

Ft = correction factor which is calculated from the geometry and configuration of the Air Cooler. For an one pass three row air cooler (fig.5.3) according to GPSA, EDP,9th edition and the calculated LMTD Ft=0.6 and 0.85 for AC1 and AC2 respectively.

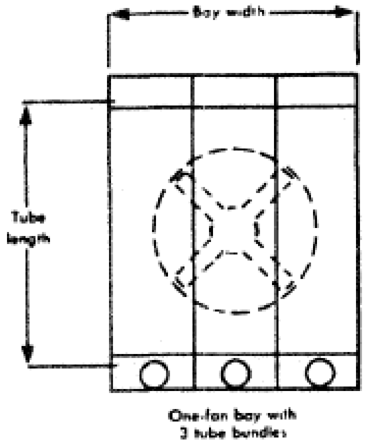

Figure 5.3: Three row one pass air cooler

UA has to be calculated and inserted in HYSYS. From table 5.4, it can be inferred that:

$$UA = U_x (A_{bf})(APSF) \qquad (5.3)$$

Where

1. APSF: tube expanded area per square foot of bundle=68.4 for fintube 1/2in. by 9, tube pitch 2in. and 3rows cooler table (5.4)
2. A_{bf} :Bundle face area (ft^2) calculated by the correlation of table 5.4 for AC1 17 ft^2 and for AC2 26 ft^2

External Area of Fin Tube Per Ft² of Bundle Surface Area (APSF) for 1-in. OD Tubes				
	½ in. Height by 9 fins/in.		⅝ in. Height by 10 fins/in.	
Tube Pitch	2 in. Δ	2¼ in. Δ	2¼ in. Δ	2½ in. Δ
3 rows	68.4	60.6	89.1	80.4
4 rows	91.2	80.8	118.8	107.2
5 rows	114.0	101.0	148.5	134.0
6 rows	136.8	121.2	178.2	160.8

$$A = \frac{q}{U_x\,(LMTD)\,(APSF)}$$

where A = required area of bundle face, ft^2
q = heat duty, Btu/hr
U_x = overall heat transfer coefficient, Btu/hr-ft^2-°F
LMTD = log mean temperature difference correctec
APSF = tube expanded area per square foot of bundle face

Table 5.4: Bundle face area and tube expanded area per square foot of bundle face calculation (Arnold K.,vol.2, 1986)

Consequently, the resultants UA for HYSYS calculation are for AC1: UA=2911.7 BTU/F/h and for AC2 UA=4278.5 BTU/F/h.

5.2.2 Scrubbers-Liquid Knock Out Vessels

Liquid knockout vessels will be sized according to the procedure of section 2.2.3 for vertical two phase separators. Scrubbers due to their high gas-oil ratio are designed in vertical arrangement with a higher liquid droplet specification (500micron). As it was explained in section 2.3 vapor recovery unit (VRU) removes liquid from gas with a more complex way than a simple scrubber. Hence, vapor liquid recovery will be simulated by a simple theoretical ideal two phase separator. The design bases for liquid knock out vessel 1 and vessel 2 (fig.5.1), based on the production of state 1 with the maximum gas and liquid content and a flow safety factor of 10%. Sizing spreadsheets results are the following:

Design Bases	Liquid KO1	Liquid KO2
Qg (MMscfd)	0,43	1,46
Ql (BOPD)	2,07	3,21
Po (psia)	26,9	115
To (F)	81,58	95
$d_{m,l}$ (microns)	500	500

Table 5.6: Design Basis for Liquid knockout vessels 1 and 2

Liquid KO1			
C_D Calculation (Iterative process section 2.3.3)			
ρl (lb/ft^3)	48,86	Z	0,95
ρg (lb/ft^3)	0,17	μ gaz (cp)	0,010
C_{D1} (initial value)	0,34		
	V (ft/s)	Re	C_D
Iteration1	7,62	1413,26	0,44
Iteration2	6,72	1246,89	0,44
Minimum Diameter Calculation			
dmin (eq.2.8)		8,63	
Liquid KO2			
C_D Calculation (Iterative process section 2.3.3)			
ρl (lb/ft^3)	61,14	Z	0,98
ρg (lb/ft^3)	0,58	μ gaz (cp)	0,010
C_{D1} (initial value)	0,34		
	V (ft/s)	Re	C_D
Iteration1	4,66	2872,13	0,40
Iteration2	4,28	2633,74	0,41
Iteration3	4,26	2623,27	0,41
Minimum Diameter Calculation			
dmin (eq.2.8)		9,87	

Table 5.7: C_D and minimum diameter calculation

Liquid KO1 Diameter vs Length				
tr (min)	do (in.)	h(in) (eq.2.9)	Lss (ft) (eq.2,16a)	SR (12Lss/do)
2	18,0	0,46	6,8	4,5
2	24,0	0,46	6,8	3,4
2	30,00	0,46	6,8	2,7
3	12,0	0,70	7,0	7,0
3	18,0	0,70	7,0	4,7
3	24,0	0,70	7,0	3,5
3	30,00	0,70	7,0	2,8
Liquid KO2 Diameter vs Length				
tr (min)	do (in.)	h(in) (eq.2.9)	Lss (ft) (eq.2,16a)	SR (12Lss/do)
2	18,0	0,55	6,9	4,6
2	24,0	0,55	6,9	3,4
2	30,00	0,55	6,9	2,8
3	12,0	0,82	7,2	7,2
3	18,0	0,82	7,2	4,8
3	24,0	0,82	7,2	3,6
3	30,00	0,82	7,2	2,9

Table 5.8: Liquid Knockout Diameter vs Length - Constraints Calculations

A slenderness ratio between 3 and 4 had to be selected. It was preferred the higher retention time, so for Liquid KO1 was selected d=24in, lss=7ft and SR=3.5 while for KO2 d=24in, lss=7.5ft and SR=3.8.

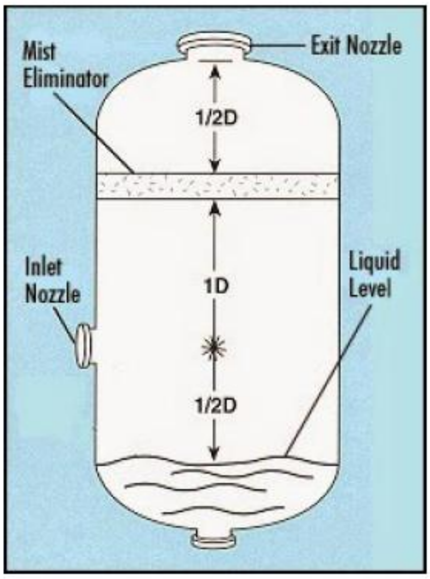

Figure 5.4: Vertical two phase separator

Nozzles were sized by the same method described in section 3.3 and their position was set according fig.5.4. Sizing results were introduced in HYSYS to estimate carryover and "Proseparator" method was applied. It was figured out that there is need for demister device to control liquid quantity entrained in gas because liquid was inserted in the compressor.

nozzles sizing	Liquid KO1	Liquid KO2
inlet ρ_{min} (lb/ft^3)	0,18	0,59
inlet mass flow (lb/hr)	1674,73	4228,71
inlet Dn (inch)	1,84	2,17
gas outlet ρ_{min} (lb/ft^3)	0,17	0,58
gas outlet vol flow (MMscfd)	0,39	1,33
gas outlet Dnmin (inch)	2,64	6,56
liquid outlet ρ_{min} (lb/ft^3)	62,32	61,14
liquid outlet mass flow (lb/hr)	27,46	41,71
liquid Dnmin (inch)	0,09	0,11

Table 5.8: Liquid Knock-Out nozzles sizing

The calculated carry over before and after the installation of a 6in thick standard mesh pad (Otto York 431, table 4.3) are presented in the following table:

Gas Carry Over	Liquid KO1		Liquid KO2	
	no demister	mesh pad	no demister	mesh pad
oil in gas (v/v)	4,72E-08	1,01E-11	1,01E-06	1,86E-10
water in gas (v/v)	3,09E-06	4,73E-10	3,25E-07	5,49E-11

Table 5.9: Liquid carry-over in gas for scrubbers

5.2.3 Compressors

Compressor which is used to compress separated gas, it is normally called a "flash gas compressor." Flash gas compressors are normally characterized by low throughput rate and high differential pressure. The use of large compressors is probably more prevalent in oil field facilities than in gas field facilities. Oil wells often require low surface pressure and the gas that flashes off the oil in the separator must be compressed in a flash gas compressor.

Vapors from tanks and other atmospheric equipment may be recovered in a "vapor recovery compressor" (VRU). Vapor recovery compressors have very low suction pressure (0 to 8 ounces gauge) and typically have low flow rates. They normally discharge into the suction of a flash gas compressor.

Vane rotary compressors are used extensively as vapor recovery compressors and vacuum pumps. The vanes slide into and out of the slots as the shaft rotates and the volume contained between two adjacent vanes and the wall of the compressor cylinder decreases. The more vanes the compressor has, the smaller the pressure differential across the vanes. Thus, high-ratio vane compressors tend to have more vanes than low-ratio compressors. VRU scrubber in Fig.5.1 is such type of compressor.

On the other hand, both centrifugal and reciprocating compressors are commonly used as flash gas compressor. Some selection criteria were referred in section 2.3. The type, the number of stages of compression, and the horsepower are required to be defined. In order to do this the volume of gas, suction and discharge pressure, suction temperature, and gas specific gravity must be known. Typical selection criteria are summarized in table 5.10 and figure 5.6.

The analysis of multistage separation has already led to the determination of the suction and discharge conditions (table 5.2). The optimal pressure conditions of production state 1 which offers the highest gas flow rates were used to define compressors' operation. According to the introduction of 2.3 and table 5.10 a vane type must be used for vapor recovery and reciprocating compressor is the most suitable choice for flash gas compressors. . There is no much literature about vane type; neither HYSYS offers tools to simulate it. Therefore, analysis will focus on reciprocating compressors and VRU compressor was replaced in the model by a reciprocating type of 82% adiabatic efficiency.

Gas flow rates are less than 2 MMscfd and 115 ACFM (actual cubic feet per minute) as it can be seen in table 5.11. Pressure ratio are higher than 3.5. Thus, table 5.10 and figure 5.5 suggest the use of high speed reciprocating compressors because they offer high pressure ratio to low throughput rates.

Example Compressor Type Selections

Service	Flow Rate MMscfd	R	n	Approx. bhp	Most Likely	Selection Alternate
Booster	100	2.0	1	4,400	Centrifugal	Integral (onshore only)
	10	2.0	1	440	High Speed	
Gas Lift	5	2.7	3	980	High Speed	
	20	2.7	3	3,920	Centrifugal	Integral (onshore only)
	100	2.7	3	19,602	Centrifugal	
Flash Gas	2	2.0	1	88	Screw	High Speed
	2	2.0	2	190	High Speed	Screw
	4	2.0	2	380	High Speed	
Vapor Recovery	0.1	4.0	1	9	Vane	Screw
	1.0	3.0	2	143	Screw	Vane
	2.0	3.0	2	286	High Speed	Screw

Table 5.10: Compressor type selection (Arnold K.,vol.2, 1986)

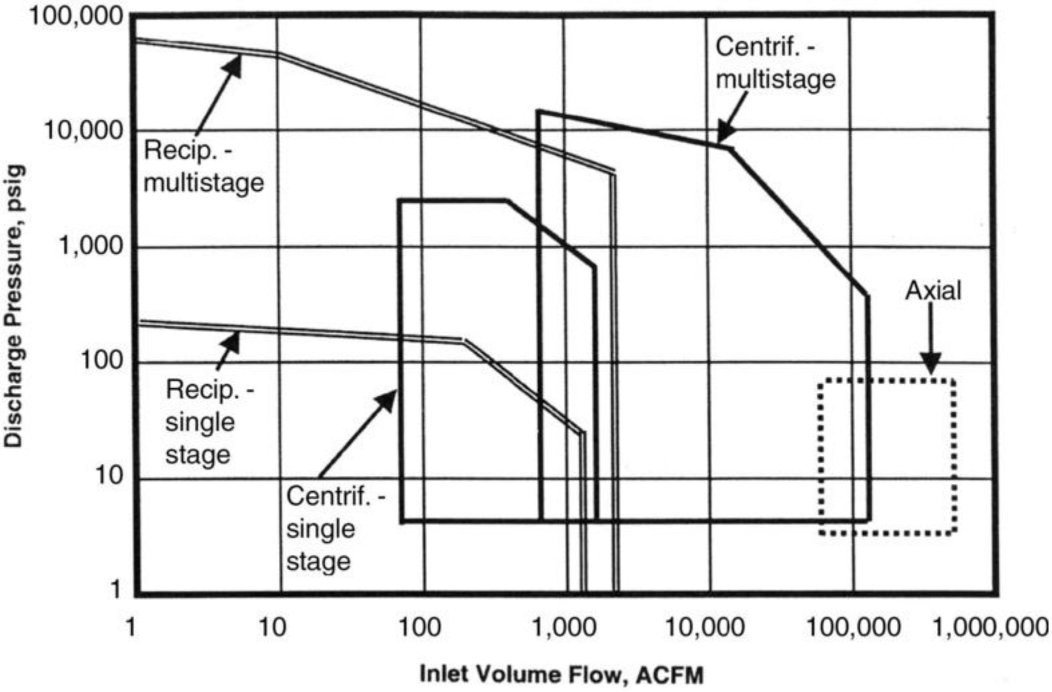

Figure 5.5: Compressor selection. Areas indicate regions of best performance (courtesy of Dresser-Rand, Petrowiki, 2015)

Table 5.10 and figure 5.5 give also indications for the number of stages of each compressor (n). The number of stages usually coincides with the number of cylinders or throws. However, there are two stage reciprocating compressors with a set of two cylinders working in parallel for each stage. Another rule is that for overall compression ratio >5, multistage reciprocating compressors have to be used but for ratio between 3 and 5 single stage must be selected.

	Compressor 2	Compressor 1
Ts (F)	95	81,58
Psuction (psia)	115	30
Pdischarge (psia)	600	115
Pressure Ratio	5,217	3,83
gas flow(MMscfd)	1,27	0,41
actual flow (ACFM)	114,3	43,18
stages	2	1
bhp per cf (Table 5.13)	116	70
break horse power	147	29

Table 5.11: Compressor 1 and 2 nominal break horsepower estimation

After deciding the type and the number of stages, an estimation of the nominal break horsepower must be carried out to define the compressor. The total horsepower for the compressor is the sum of the horsepower required for each stage and an allowance for inter-stage pressure losses. It is estimated that there is a 3% loss of pressure in going through the cooler, scrubbers, piping, etc., between the actual discharge of the cylinder and the actual suction of the next cylinder. Nevertheless, this pressure drop was neglected in the model in order to succeed recycle loop convergence. The brake horsepower per stage can be determined from the thermodynamic correlation of table 5.12 or from technical charts as 5.13.

$$BHP = 0.0857\,[Z_{av}]\left[\frac{(Q_g)(T_s)}{E}\right]\left[\frac{k\eta}{k-1}\right]\left[\left(\frac{P_d}{P_s}\right)^{\frac{k-1}{k\eta}} - 1\right]$$

where BHP = brake horsepower per stage
Q_g = volume of gas, MMscfd
T_s = suction temperature, °R
Z_s = suction compressibility factor
Z_D = discharge compressibility factor
E = efficiency
high-speed reciprocating units — use 0.82
low-speed reciprocating units — use 0.85
centrifugal units — use 0.72
η = polytropic efficiency
k = ratio of gas specific heats, C_p/C_v
P_s = suction pressure of stage, psia
P_d = discharge pressure of stage, psia
$Z_{av} = (Z_s + Z_D)/2$

Table 5.12: Compressor horsepower per stage (Arnold K.,vol.2, 1986)

Compressor Horsepower Selection Chart

BRAKE HORSEPOWER PER MILLION CUBIC FEET*
DISCHARGE PRESSURE (PSIG)

SUCTION PRESSURE PSIG	25	50	75	100	125	150	175	200	250	300	350	400	450	500	550	600	650	700	750	800	850	900	950	1000	1050	1100	1150	1200	1250	1300	1350	1400	1450	1500	Stages
0	65	99	128	144	156	168	178	187	203	218	233	233	241	248	254	260	266	272	277	282	286	291	295	299	303	307	311	315	312	318	318	321	324	328	4
10	35	63	85	104	121	131	140	149	163	175	186	196	205	214	223	231	228	233	237	242	245	250	253	257	260	264	267	270	267	270	273	276	279	282	
20		43	62	78	92	106	118	126	139	151	160	170	178	186	193	199	206	212	218	225	231	226	229	232	236	239	242	245	242	245	248	251	254	257	3
30		29	47	62	74	85	96	107	123	133	143	152	159	167	173	179	185	191	196	201	206	211	216	221	226	230	224	227	224	227	230	233	236	239	
40			36	50	61	72	81	90	107	121	130	138	145	152	158	164	170	175	180	185	190	194	198	202	206	210	214	218	214	218	222	226	230	234	
50			26	41	52	61	70	78	93	106	119	127	134	141	147	153	158	163	168	173	177	181	185	189	193	196	200	203	200	203	206	209	212	215	2
60				32	44	53	61	69	83	95	108	118	125	131	137	143	148	153	158	162	166	170	174	178	182	185	188	192	188	192	196	200	204	208	
70				25	37	46	54	61	74	86	97	109	117	123	129	135	140	145	149	153	157	161	165	169	172	176	179	182	179	182	185	188	191	194	
80					30	40	47	54	67	78	89	98	109	117	122	127	132	137	142	146	150	153	157	161	164	167	171	174	171	174	177	180	183	186	
90					24	34	42	49	61	72	81	91	100	109	116	121	126	131	135	139	143	147	150	154	157	160	163	166	163	166	169	172	175	178	
100						28	37	44	55	66	75	84	92	100	109	116	120	125	129	133	137	141	144	148	151	154	157	160	157	160	163	166	169	172	
125							25	32	44	54	63	71	78	85	92	99	106	113	117	121	124	128	131	134	137	140	143	146	143	146	149	152	155	158	
150								22	35	45	53	60	67	74	80	86	92	98	103	110	114	118	121	124	127	130	133	135	133	135	137	139	141	143	
175									27	37	45	52	57	60	71	76	82	87	92	97	102	107	112	115	118	121	123	126	123	126	129	132	135	138	
200										30	38	45	52	58	63	68	73	78	83	88	92	96	101	105	110	113	116	119	116	119	122	125	128	131	
250											26	33	40	46	51	56	60	65	69	73	77	81	85	88	92	95	99	102	99	102	105	108	111	114	
300												23	30	36	41	46	50	54	58	62	66	69	73	76	79	83	86	89	86	89	92	95	98	101	
350													21	27	33	38	42	46	50	53	57	60	63	67	70	73	75	78	75	78	81	84	87	90	1
400															25	30	35	39	43	46	50	53	56	59	60	64	67	70	67	70	73	76	79	82	
450																23	28	32	36	40	43	46	49	52	55	58	60	63	60	63	66	69	72	75	
500																	22	26	30	34	38	41	44	46	49	52	54	57	54	57	60	63	66	69	
550																		20	25	20	32	36	39	41	44	46	49	51	49	51	53	55	57	59	
600																				23	27	30	34	37	39	42	44	46	44	46	48	50	52	54	
650																					22	26	29	32	35	38	40	42	40	42	44	46	48	50	
700																						22	25	28	30	33	36	38	36	38	40	42	44	46	
750																							20	24	27	29	32	34	32	34	36	38	40	42	

* Note: MMSCFD measured at 14.7 pisa, and 60F, suction temp = 100F, Natural Gas K=1.26, SG=.65
Example: Given = P1 =70 Psig, P2=650 Psig, MMSCFD = 8.0 Result: 140 Bhp/ MMSCFD = 8 X 140 = 1200 Total BHP.

Table 5.13: Reciprocating Compressor Horsepower Selection (GE Power, 2016)

If the nominal break horsepower is specified, high speed reciprocating compressor size can be selected from technical charts as table 5.14. For example, a frame size M with 1 cylinder (throw), piston rod diameter of 1.125 in. and stroke of 3 in. would be suitable for compressor 2 which has the lowest power demand. Compressor 1 requires a higher frame size (H) with two cylinders and the same piston rod diameter and stroke due to its higher power needs.

API 11P High Speed Reciprocating Compressors

Balanced opposed high speed compressors of rugged design for heavy duty service in oil and gas field applications.

Frame Size	No. of Throws	Max. Power Capability (*) hp (kW)	Piston Rod Diameter in. (mm)	Stroke in. (mm)	Max. Speed (rpm)
M	1-2	120 (90)	1.125 (28.6)	3 (76.2)	1800
H	1-2-4	400 (300)	1.125 (28.6)	3 (76.2)	1800
A	2-4	800 (600)	1.375 (34.9)	3.5 (88.9)	1800
B	2-4	1600 (1200)	1.75-2.0 (44.5-50.8)	3.5-4.5-5.0 (88.9-114.3-127.0)	1200-1500-1800
DS	2-4	2400 (1800)	1.5-2.0 (38.1-50.8)	4.25-5.0-6.0 (108-127.0-152.4)	1200-1500-1600
ES	2-4-6	7200 (5400)	2.25 (57.2)	5.0-6.0-7.0 (127.0-152.4-177.8)	1000-1200-1500
FS	2-4-6	7200 (5400)	2.5 (63.5)	5.0-6.0-7.0 (127.0-152.4-177.8)	1000-1200-1500
SHMB	2-4	5900 (4400)	2.5 (63.5)	6 (152.4)	1200
SHM	2-4-6	8800 (6600)	2.5 (63.5)	6 (152.4)	1200

Table 5.14: Typical High Speed Reciprocating Compressors sizes (GE Power, 2016)

Typically, high-speed compressors run at a speed of 900 to 1200 rpm in contrast to the slow speed units with speeds of 200 to 600 rpm. The most common driver for a high-speed compressor is a natural gas driven engine. The major characteristics of high-speed reciprocating compressors are:

Size:

• Numerous sizes from 50 hp to 3000 hp.

• 2, 4, or 6 compressor cylinders are common.

Advantages:

- Can be skid mounted.
- Self-contained for easy installation and easily moved.
- Low cost compared to low-speed reciprocating units.
- Easily piped for multistage compression.
- Size suitable for field gathering offshore and onshore.
- Flexible capacity limits.
- Low initial cost.

Disadvantages:

- High-speed engines are not as fuel efficient as low speed (7,500 to 9,000 Btu/bhp-hr).
- Medium range compressor efficiency (higher than centrifugal; lower than low-speed).
- Short life compared to low-speed.
- Higher maintenance cost than low-speed or centrifugal.

Figure 5.6 shows a high-speed engine-driven compressor package:

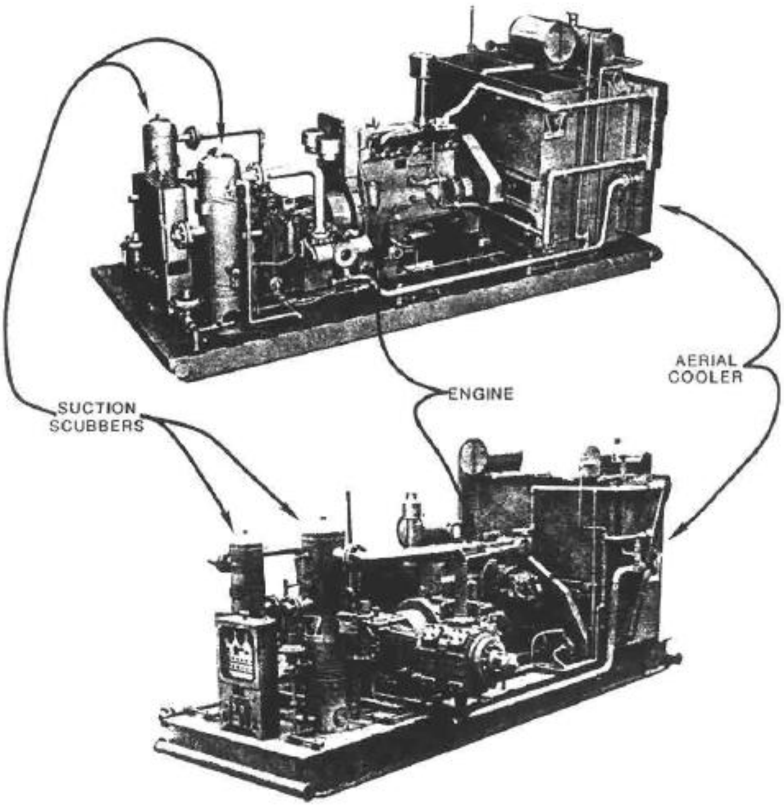

Figure 5.6: High Speed Reciprocating Compressors (Arnold K.,vol.2, 1986)

It is desirable to limit discharge temperatures to below 250°F to 275°F to ensure adequate packing life for reciprocating compressors and to avoid lube oil degradation. At temperatures above 300°F eventual lube oil degradation is likely, and if oxygen is present ignition is even possible. Under no circumstances should the discharge temperature be allowed to exceed 350°F. The discharge temperature can be lowered by cooling the suction gas and reducing the value of P_d/P_s, that is, by adding more stages of compression. It can be seen from results of table 5.2 that the above constraint is satisfied except of HP compressor whose discharge temperature is slightly greater than 300°F. With a more conservative cooling by air coolers, there would be no violation of discharge temperature constraint.

6. Performance analysis

In this chapter the effect of heating inlet fluid will be studied to simulate the relation of temperature with production facilities performance. In addition to this, some scenarios of shutting down wells will be tested to estimate separation train behavior under different input conditions.

6.1 Heating Effect

Different inlet fluid temperatures in the range of 100-160 F will be applied to estimate separation and production as long as compression equipment performance. As it was analyzed in section 3.2.3, this range is suggested for handling heavy oils by literature technical guidelines. Higher temperatures would lead to excess vaporization reducing product significantly. On the other hand, lower temperatures could trigger flow assurance problems such as hydrates, paraffins and foam. HYSYS provides the discussed flash equilibrium calculations and carry over correlation methods to estimate production facilities behavior against temperature changes. In fact, these are not enough. Experimental field data are required to simulate water-oil emulsion. However, valuable conclusions can be gathered in respect of separation train performance to different inlet fluid temperatures even with those tools. It was selected to study the two limits of production, the beginning of production (state 1) where oil and gas flow rates are maximum and its end with the highest water-cut. Pressures were adjusted to the optimal values of each state. The study will emphasize on carry over ratios, oil production flow, oil product RVP and compression power needs.

<u>Beginning of production</u>

More gas is liberated by the increase of inlet temperature providing less and heavier oil product (fig.6.1).

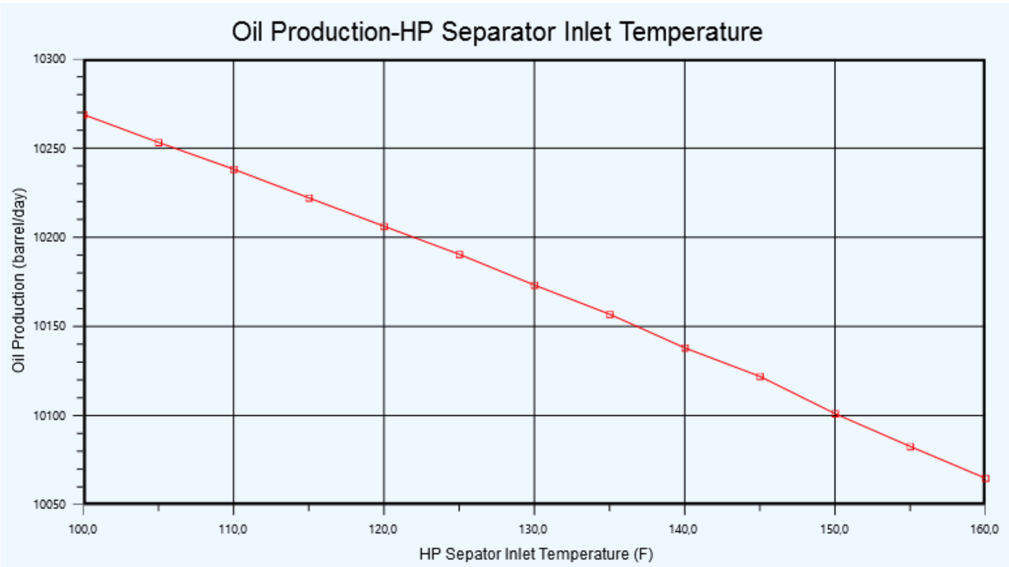

Figure 6.1: Oil production in relation to HP separator inlet fluid temperature

RVP is steadily decreasing with the increase of inlet temperature leading to a more stabilized product which though it is of higher quality than final product specification indicate (RVP 10-12 psia). This situation is not profitable for the producer considering the additional decrease of product flow (fig.6.2).

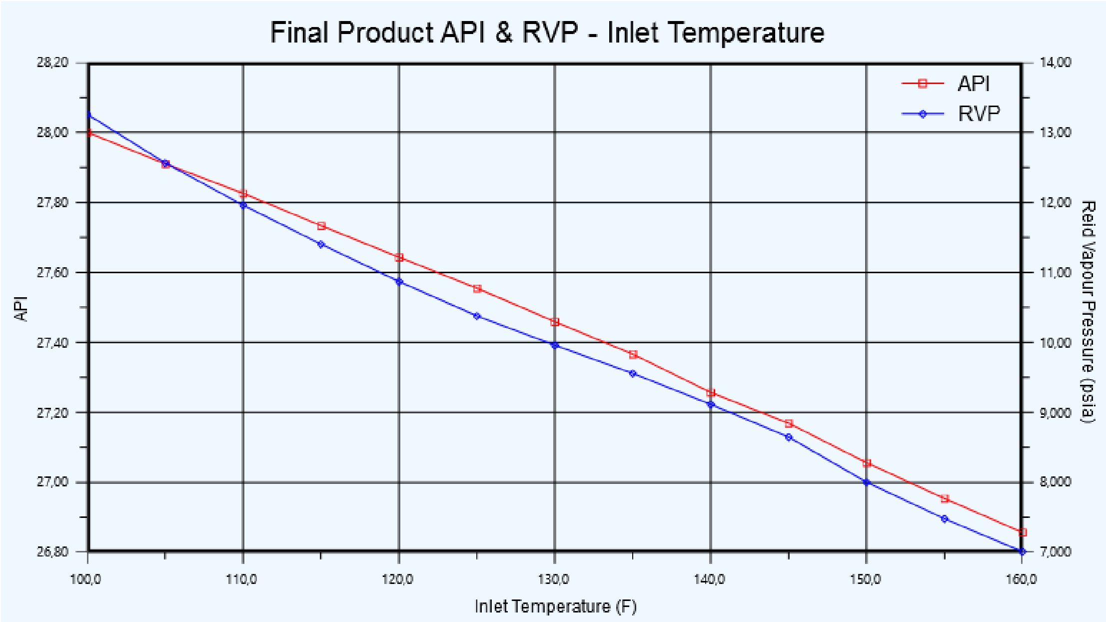

Figure 6.2: Final Product API density and Reid Vapor Pressure in relation to HP separator inlet fluid temperature

In table 5.12, it can be noticed that compressor break horse power depends on gas flow, suction temperature and other coefficients such as compressibility, c_p, c_v factors which are affected by temperature. Specifically, reciprocating compressors performance is more sensitive to gas flow changes than temperature changes. Figure 6.3 depicts the actual gas flows which are exerted from the separators. LP and HP separators are characterized by increasing gas flows in relation to inlet temperature. The increase of temperature favors vaporization of lighter hydrocarbons. Therefore, more volatile components are removed in HP separator so that the content of MP separator becomes more stable (lower gas liberation). For this reason, actual gas flow from MP separator is reduced.

Accordingly, compressors horsepower are affected (fig.6.4). Compressor 2 receives gas from MP separator resulting in lower power consumption. On the other hand, Compressor 1 and VRU compressor handle the increasing gas flows from tank and LP separator so that their energy demand is increased. All in all, total compression power consumption is governed by compressor 2 which has the highest pressure ratio and break horsepower. Hence, total power demand for compression is reduced by heating the fluid inserted in the separation train.

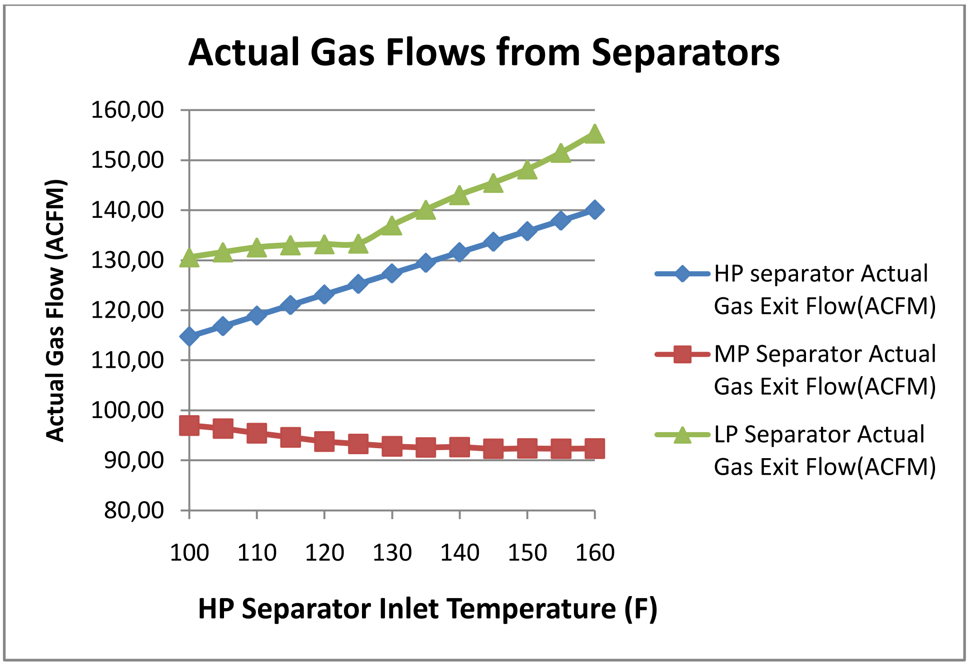

Figure 6.3: Actual Gas Flows from Separators in relation to HP separator inlet fluid temperature

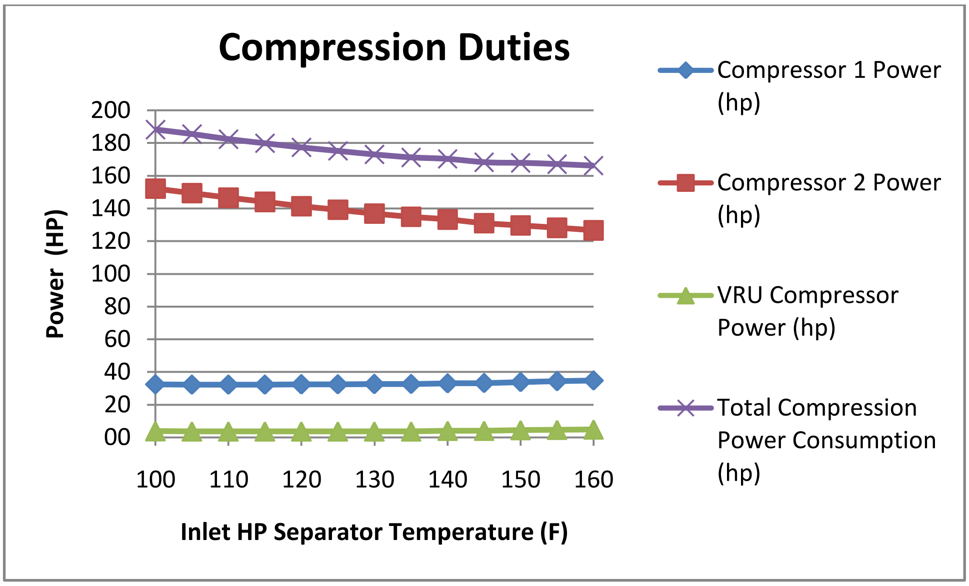

Figure 6.4: Compressors break horsepower in relation to HP separator inlet fluid temperature

As far as water carry-over is concerned, heating helps the removal of water droplets from oil in the HP separator by decreasing continuous phase viscosity and providing lower water in oil fraction (v/v). On the contrary, liquid entrainment in gas is increased due to the increase of the total gas flow and the operation temperature which favor smaller droplet size formation and entrainment (fig.6.5).

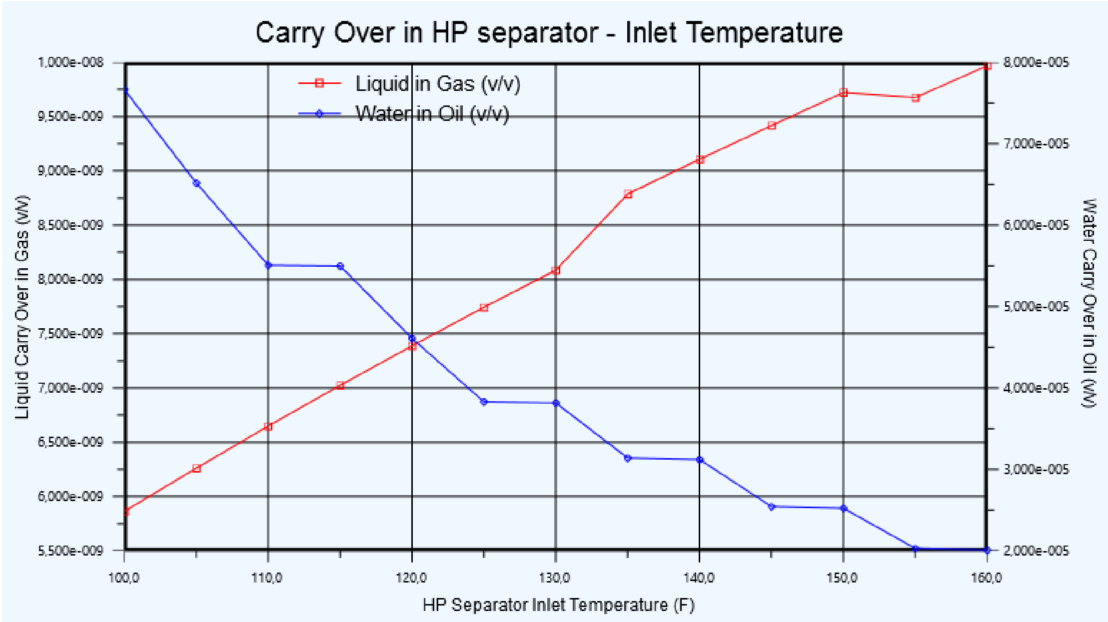

Figure 6.5: Carry over ratios in HP separator in relation to inlet temperature

However, water content is increased in tank due to the higher quantity of liquefied vapors from gas liquid recovery (fig. 6.6). It seems that until a temperature of 125 F, more water is separated from oil as liquid than vaporized. This explains the diminishing progress of water content. Then, liquid recovery from water vapors prevails so that water content is rapidly increased and a re-examination of water in oil restriction is suggested.

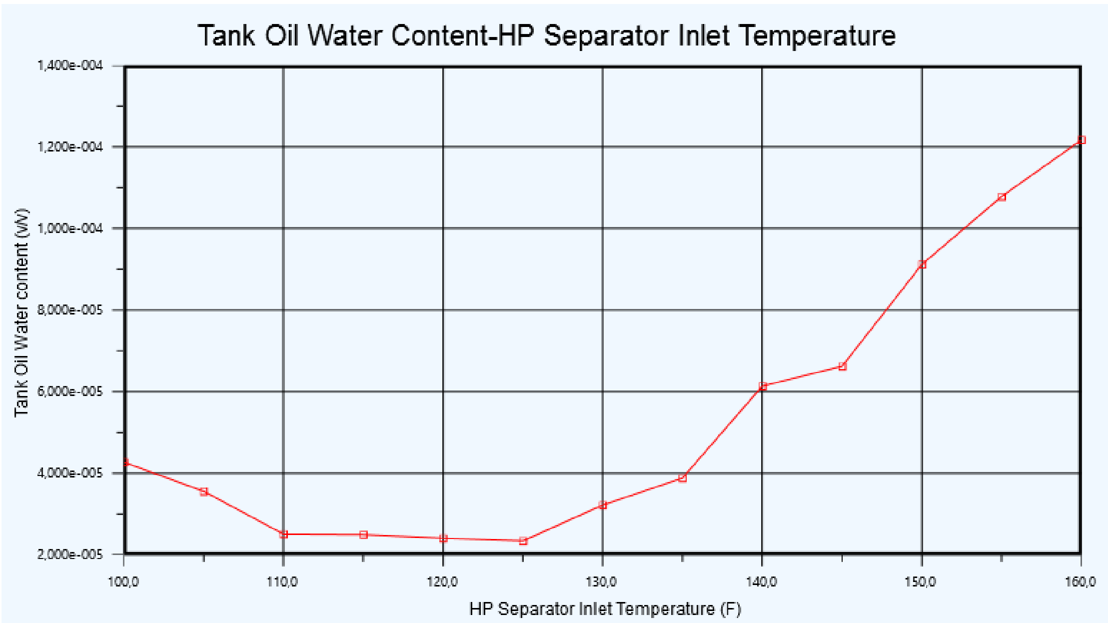

Figure 6.6: Water in tank oil in relation to HP separator inlet temperature

End of production

Process train during the end of production proved to have similar performance to the production start stage against inlet temperature changes. Figures 6.7 to 6.11 verify this behavior. This can be explained by the relatively low water cut even in the end of production, so that the resultant low water carry over does not have a significant effect on separation performance.

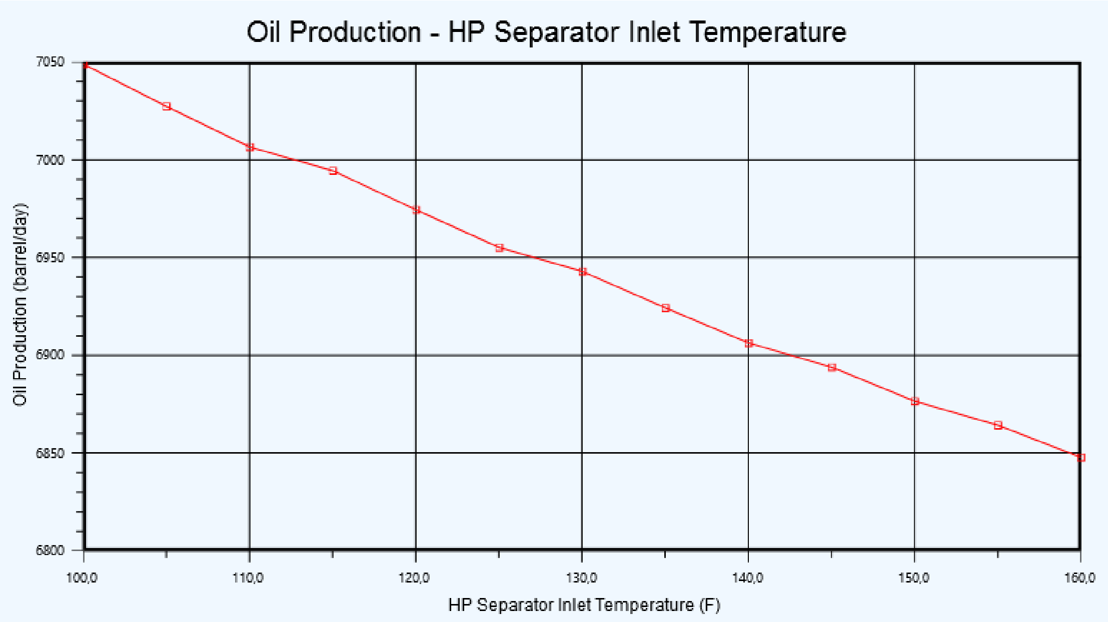

Figure 6.7: Oil production in relation to HP separator inlet fluid temperature

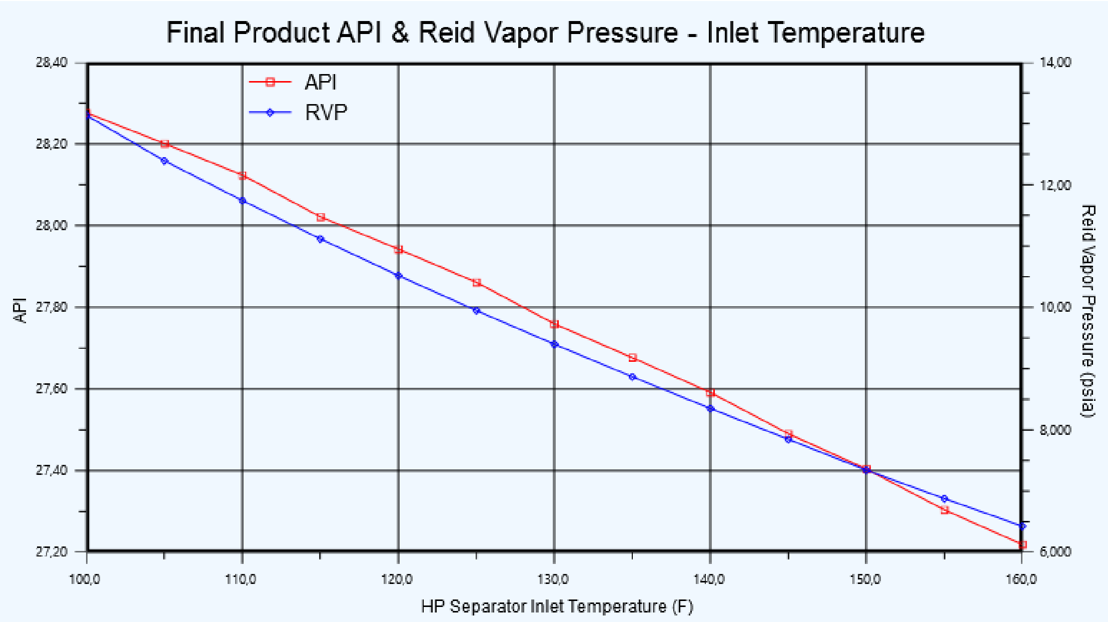

Figure 6.8: Final Product API density and Reid Vapor Pressure in relation to HP separator inlet fluid temperature

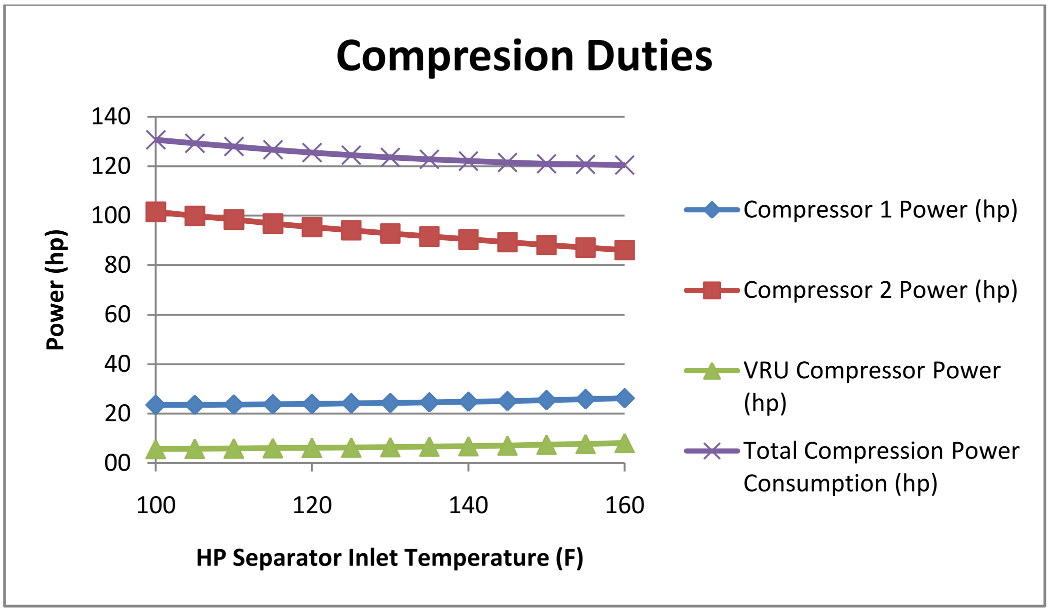

Figure 6.9: Actual Gas Flows from Separators in relation to HP separator inlet fluid temperature

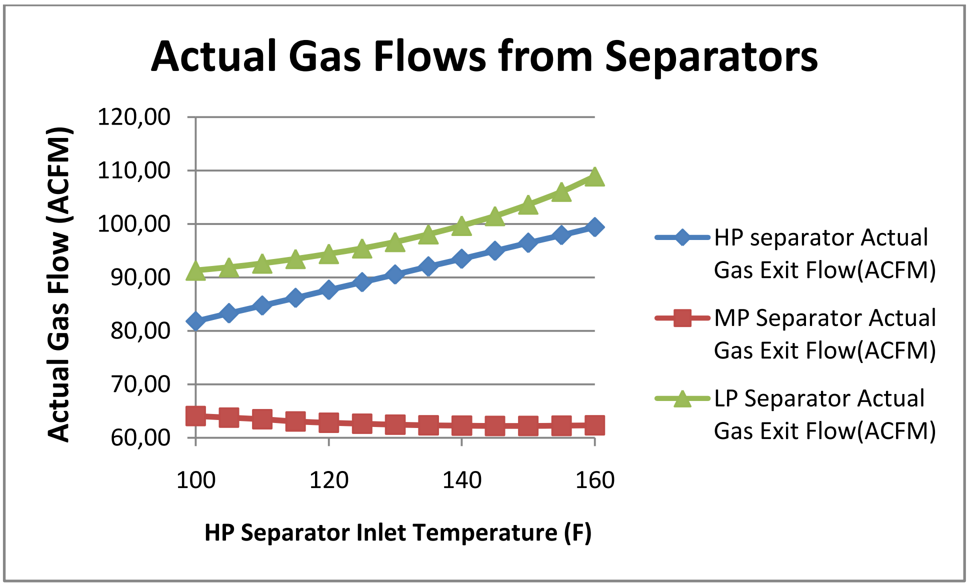

Figure 6.10: Compressors break horsepower in relation to HP separator inlet fluid temperature

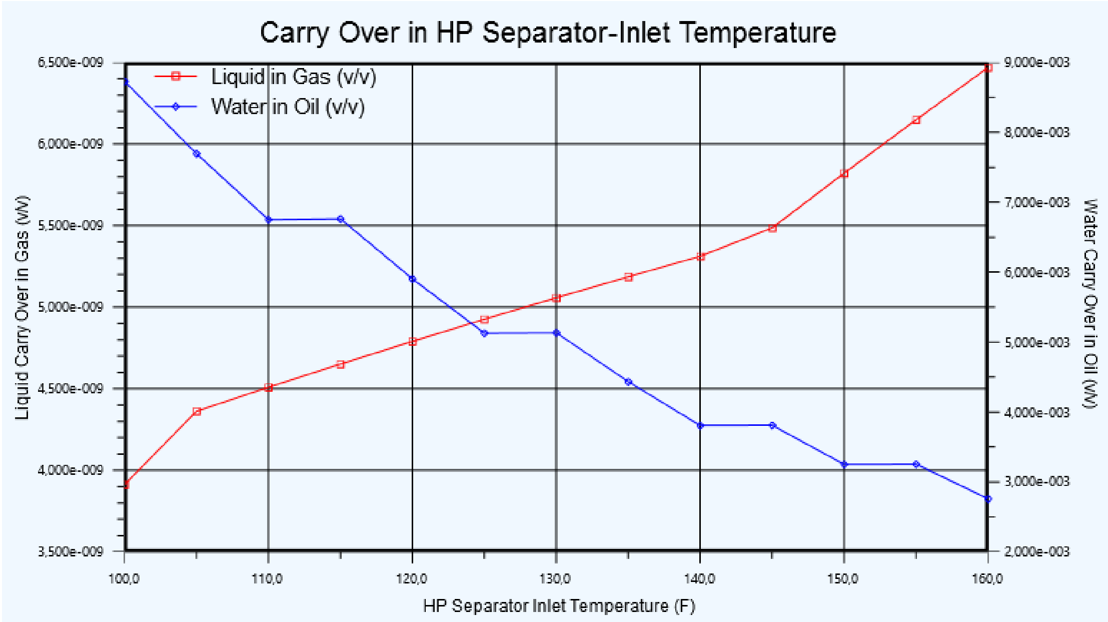

Figure 6.11: Carry over ratios in HP separator in relation to inlet temperature

The only noticeable difference in comparison to the beginning of production refers to tank oil water content. Fig.6.12 depicts a steady diminishing progress of water content. This indicates that water from vapor liquid recovery is significantly less than the entrained water in oil phase. As a result, the improving effect of heating to liquid-liquid separation prevails and final water content is reduced.

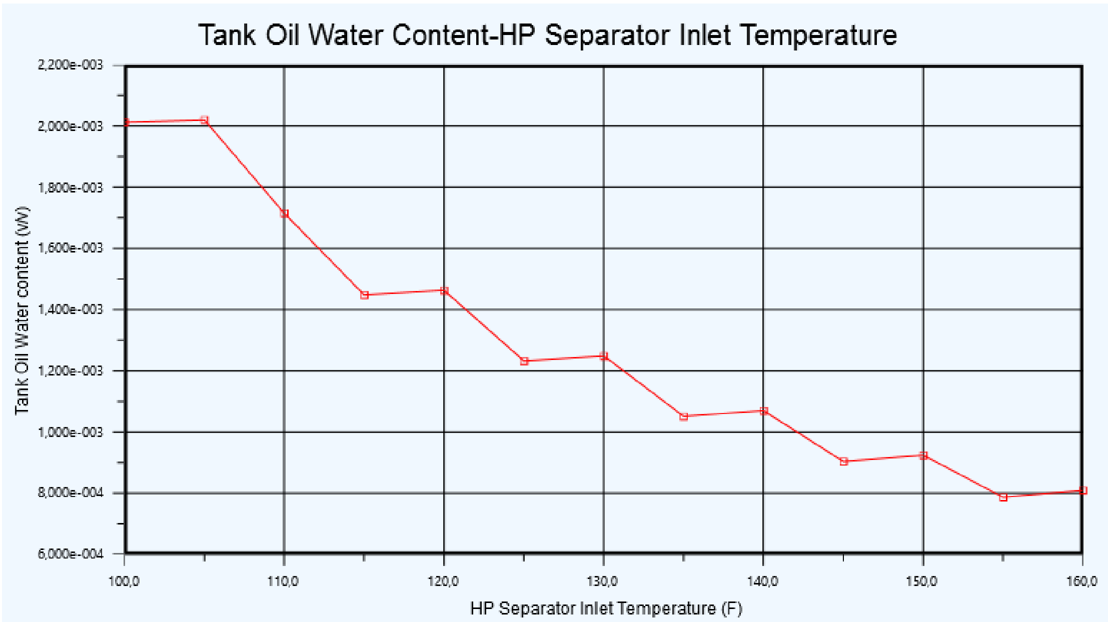

Figure 6.12: Water in tank oil in relation to HP separator inlet temperature

6.2 Wells Shut-Down scenarios

Scenarios with different shut-down wells will be examined to study the resultant product characteristics and process train behavior. During production a well may have to be shut down due to various reasons such as a kick, flow assurances, well efficiency or even oil market demand may force to reduce productive oil wells. The well 1 with the heaviest oil and the well 3 with the most volatile oil will be closed to investigate how upstream process train will perform due to the new inlet data, namely to a more volatile and a heavier inlet fluid respectively. Their performance will be investigated both in existing and optimized pressure conditions according to the optimization method described in chapter 3.

Well 1 Shut down

By closing well 1 a lighter hydrocarbon mixture is inserted so that both RVP and API values are increased while oil production is decreased (table 6.1). Optimization of new pressure conditions only altered high pressure while medium and low pressure remained nearly stable at their lower limits. Lower high pressures were calculated leading to less hydrocarbon vaporization from high pressure separator. This is an indication that three stage separation is extravagant and that possibly a two stage separation would be preferable. Also, it may imply that the ideal liquid recovery from vapor system and the conservative design of separators led to a too stable system. Another fact is that, API density and RVP were not increased greatly due to the relative low flow ratio of well 1 to total well flows. As a result, fluid mixture behavior did not change rapidly, although it became lighter.

	Well1 Shut Down				All Wells Open	
	Production State 1		Production State 5		state 1	state5
Pressure Conditions:	Existing	Optimized	Existing	Optimized	Optimized	Optimized
HP Separator Inlet Temperature [F]	130,00	130,00	130,00	130,00	130,00	130,00
HP Separator - Vessel Pressure [psia]	599,76	539,07	599,78	590,3	599,62	599,68
MP Separator - Vessel Pressure [psia]	115,00	115,00	115,00	115,00	115,00	115,00
LP Separator - Vessel Pressure [psia]	26,90	25,00	26,90	25,00	26,90	26,90
Oil Production [barrel/day]	7600,4	7601,18	5362,72	5364,8	10172,3	6905,72
API	31,12	31,13	31,10	31,10	27,46	27,87
Reid Vapour Pressure [psia]	9,96	9,98	9,52	9,52	9,96	9,41

Table 6.1: Well1 Shut Down scenario production conditions

Compressors operating under the same pressure conditions after the shutdown of well 1, consume less power as they handle lower flows (table 6.2). After the adjustment of calculated optimal conditions, compressor 2 consumes even lower power as it has to provide lower pressure ratio. In contrary, the consumption of compressor 1 is increased because of the higher volumes of gas inserted from LP separator. The relatively low pressures of HP and MP separators have enclosed volatile hydrocarbon components to the stream more of which are liberated in the low pressure separator increasing its gas product.

	Well1 Shut Down				All Well Open	
	Production State 1		Production State 5		state 1	state5
Pressure Conditions:	Existing	Optimized	Existing	Optimized	Optimized	Optimized
Compressor1 - Power [hp]	26,83	43,20	20,64	30,65	32,52	24,28
Compressor2 - Power [hp]	110,77	93,18	78,39	72	136,88	92,78
VRU compressor - Power [hp]	3,12	1,84	2,14	2.08	3,70	6,46

Table 6.2: Well1 Shut Down scenario compressors performance

<u>Well 3 Shut down</u>

Well 3 shut down scenario provides a heavier hydrocarbon mixture and as it was expected API density decreases (table 6.3). RVP is increased proving that existing operation conditions have enclosed sufficient volatile components to the new mixture. The optimization procedure similarly to the previous scenario did not cause significant changes to the lower pressures. However, it provided lower high pressure operation. This fact proves that the new heavier fluid can be compressed more encaging more lighter components than the fluid of the previous scenario without losing product in the lower pressure separation stages due to vaporization.

Compression power needs are generally reduced due to the lower pressure ratios and flow rates (table 6.4). Fluid is significantly less volatile than the one of the previous scenario so that gas flow production from the lower pressure separators is low and the respective power requirements from the middle and low compression stages are decreased.

A crucial point to be noticed is that water and oil separation becomes more difficult for the new heavier mixture due to the higher oil phase viscosity. Although oil retention time increases due to the lower flow rate, this fact cannot totally compensate for the increased viscosity. Table 6.5 depicts the high water in oil carry over values. Especially, in the end of production (state 5) with the maximum water-cut the specification for 0.5% v/v water in oil is not satisfied. Consequently, a different separator design, heat treatment or an electrostatic treater are possible additional solutions to reduce water in oil content.

	Well3 Shut Down				All Well Open	
	Production State 1		Production State 5		state 1	state5
Pressure Conditions:	Existing	Optimized	Existing	Optimized	Optimized	Optimized
HP Separator Inlet Temperature [F]	130,00	130,00	130,00	130,00	130,00	130,00
HP Separator - Vessel Pressure [psia]	599,89	466,53	599,90	412,92	599,62	599,68
MP Separator - Vessel Pressure [psia]	115,00	115,00	115,00	115,00	115,00	115,00
LP Separator - Vessel Pressure [psia]	26,90	26,90	26,90	26,90	26,90	26,90
Oil Production [barrel/day]	7872	7872,69	5349,1	5371,65	10172,29	6905,72
API	21,37	21,37	21,33	21,28	27,46	27,87
Reid Vapour Pressure [psia]	10,43	10,45	10,39	10,38	9,96	9,41

Table 6.3: Well3 Shut Down scenario production conditions

	Well3 Shut Down				All Well Open	
	Production State 1		Production State 5		state 1	state5
Pressure Conditions:	Existing	Optimized	Existing	Optimized	Optimized	Optimized
Compressor1 - Power [hp]	14,32	25,95	9,46	12,93	32,52	24,28
Compressor2 - Power [hp]	71,20	47,26	46,40	24,72	136,88	92,78
VRU compressor - Power [hp]	1,47	3,10	0,97	1,80	3,70	6,46

Table 6.4: Well3 Shut Down scenario compressors performance

	Well3 Shut Down				All Well Open	
	Production State 1		Production State 5		state 1	state5
Pressure Conditions:	Existing	Optimized	Existing	Optimized	Optimized	Optimized
HP Separator - Water in Oil Carry Over (v/v)	2,76E-04	2,74E-04	1,47E-02	1,80E-02	3,82E-05	9,63E-04
Water content in Tank Oil (v/v)	3,43E-04	3,55E-04	1,84E-02	2,25E-02	3,22E-05	1,25E-03

Table 6.5: Well3 Shut Down scenario water in oil carry over

7. Conclusions

Petroleum companies are looking to optimize recovery from both their existing and new assets. Surface facilities simulation offers them the ability to improve upstream equipment design at low cost, reliably and safely while to extract even greater value from their capital investments. Performance analysis with simulation models can provide higher levels of efficiency and agility, with greater transparency and accuracy in decision making.

In the frame of this field, a model for a three phase separation train in service of a three oil fields bundle was built. Composition and properties data of the well fluids were introduced to HYSYS process simulator. Inlet fluid conditions to the process train were determined and flash equilibrium calculations were applied to simulate different ideal multi-stage separation scenarios. Additionally, numerical optimization methods were used to define optimum operational pressure conditions for each case. Three-stage separation was selected as the possible scheme to maximize production and increase process train flexibility. The high pressure separator was analytically designed according to these conditions using both technical guidelines and droplet settling theory.

Carry over calculation was introduced for a realistic equipment design and simulation. Based on recent literature guidelines and the correlation tools that HYSYS offers, carry over estimation was carried out for high pressure separator. Droplet distribution calculations were included. It was figured out that a three phase separator with a boot in combination with a standard mesh pad demister can meet final product water in oil specification. Then, lower and medium pressure separators were designed as two phase separators and their internals were analytically sized.

The model was upgraded by adding flash gas compression train and taking advantage of HYSYS capabilities. Gas auxiliary compression equipment was designed based on the new optimal operation conditions. High speed reciprocating compressors were chosen due to their high pressure ratios and low throughputs.

Finally, a performance analysis was performed to define the effect of heating inlet fluid to final product properties, compression power needs and separation efficiency. Moreover, two well shut down scenarios were set to investigate the reaction of the designed equipment to different inlet fluids and possible modifications to be suggested.

This relatively simplified and straight forward desk study proved that the combination of technical guidelines with process simulation software application can produce a conservative but reliable design of upstream facilities in addition to an inclusive model for process variables analysis and optimization.

8. References

[1] Adewumi M., *Design and Optimization of Separators, Phase Relations in Reservoir Engineering,* The Pennsylvania State University, 2016, Retrieved from: https://www.e-education.psu.edu/png520/m20_p3.html

[2] Akpan D. G. ,Performance of Internals in Three-Phase Tank Separator, MSc in Petroleum Engineering Thesis, Norwegian University of Science Department of Petroleum Engineering Science and Technology and Applied Geophysics, 2009

[3] Arnold K. and Stewart M., Surface Production Operations, Vol. 1-2 , Third Edition, Design of Oil-Handling Systems and Facilities, Houston, Texas: Gulf Publishing Co., 1986.

[4] Aspen HYSYS Help Guide

[5] Baker M. & Dick D., *Proseparator – a novel separator/scrubber design program,* BP Engineering, SPE21506, 1991

[6] Bandopadhyay M., Design optimization of heat exchangers in topside systems for offshore oil and gas processing, MSc Thesis, Norwegian University of Science and Technology, Department of Energy and Process Engineering ,2014

[7] Bianco Romano, Treatment plants for oil production, Vol.I / Exploration, Production and Transport, Encyclopedia of Hydrocarbons

[8] Bothamley M., Gas/Liquids Separators—Part 1-3: Quantifying Separation Performance, John M. Campbell/ PetroSkills, SPE Oil and Gas Facilities, 2013

[9] Bradley H., Gipson F., Petroleum Engineering Handbook, Society of Petroleum Engineers, 1987

[10] Campbell, J.M., Gas Conditioning and Processing**.** Campbell Petroleum Series, Norman, Okla, 1976

[11] Carpenter C.L., Othmer D.F., *Entrainment removal by a wire mesh separators,* AIChE J., 1 (December) (1955), pp. 549–557 , 1955

[12] Chamberlain R.M. and Powell M.J.D., Lemarechal C. and Pedersen H. C., *The Watchdog Technique for Forcing Convergence in Algorithms for Constrained Optimization*, Mathematical Programming Study. Springer Berlin Heidelberg. 16: 1-17., 1982

[13] Chilingarian G., Robertson J. and Kumar S., Surface Operations in Petroleum Productlon, , I, Developments in Petroleum Science Volume 19, Part A, 1987

[14] Chugh A., *How to Design Entrainment Separation Vessels / Gravity Deacanter,* 2014 Retrieved from: http://www.pipingguide.net/2014/12/how-to-design-entrainment-separation.html

[15] *Crude Oil Processing on Offshore Facilities, 2016,* Retrieved from: http://www.piping-engineering.com/crude-oil-processing-offshore-facilities.html

[16] Gaidhani H., Debottlenecking existing offshore production facilities to safely extend their operational capacities, SBM offshore, Oil and Gas facilities, 2014

[17] Gas Processors Suppliers Association, Engineering Data Book, Vol.I &II, FPS version, 12thedition, 2014

[18] General Electric Company, *Reciprocating Compressors,* GE Oil& Gas brochure,2005

[19] Ishii, M. and Grolmes, *Inception Criteria for Droplet Entrainment in Two-Phase Concurrent Film Flow*, AIChE J., 1975.

[20] Jentz R., *Separator Sizing, Oil and Gas Processing*, 2016, Retrieved from: http://www.oilgasprocessing.com/Separator%20Sizing.htm

[21] Karabelas A. *Drop Size Spectra Generated in Turbulent Pipe Flow of Dilute Liquid-Liquid Suspensions,* AIChE Journal, vol. 24, No. 2, pages 170-181, 1978

[22] Kataoka, I., Ishii, M., and Mishima, *Generation and Size Distribution of Droplets in Gas-Liquid Annular Two-Phase Flow*, ASME Journal of Fluids Engineering, 1983.

[23] Kazemi Mahsa, Optimization of Oil and Gas Multi Stage Separators Pressure to Increase Stock Tank Oil, Thermodynamic Laboratory, Faculty of Chemical Engineering, Tarbiat Modanes University, Tehran (Iran),2011

[24] Kuester, J. L., and Mize, J. H., *Optimization techniques with Fortran*, McGraw-Hill, New York., (1973)

[25] Kylling O. W., Optimizing separator pressure in a multistage crude oil production plant, MSc in Engineering Cybernetics Thesis, Norwegian University of Science and Technology, Department of Engineering Cybernetics, 2009

[26] Lake L., *PEH: Oil and Gas Separators*, 2015, Retrieved from: http://petrowiki.org/PEH%3AOil_and_Gas_Separators#Separator_Components

[27] Langmuir & Blodgett, *A mathematical investigation of water droplet trajectories*, U.S. Army Air Force Tech. Rep., 5418 ,1946

[28] Lawson P. & Little L, *Separator Design and Operation: Tools for Transferring "Best Practise",* BP Exploration and Operating CO. Ltd, SPE36647, 1996

[29] Monneray W. & Svrcek W. , Successfully Specify Three-Phase Separators, Multiphase Flow p.29-40, Chemical Engineering Progress, 2004

[30] Moshfeghian M., *Gas-Liquid Separators Sizing Parameter,* John M. Campbell/PetroSkills, 2014

[31] NORSOK standard, *Process systems*,P-100, 2001

[32] *Offshore Processing Options vary widely,* Oil and Gas Journal, *2004,* Retrieved from: http://www.ogj.com/articles/print/volume-102/issue-45/drilling-production/offshore-processing-options-vary-widely.html

[33] Pan, L. and Hanratty, *Correlation of Entrainment for Annular Flow in Horizontal Pipes*, International Journal of Multiphase Flow, 2002

[34] Powell M.J.D., *A Fast Algorithm for Non-Linearly Constrained Optimization Calculations*, Numerical Analysis. Proceedings of the Biennial Conference, Dundee. Springer Berlin Heidelberg. 630: 144-157., 1978

[35] Press, W. H.; Flannery, B. P.; Teukolsky, S. A.; and Vetterling, W. T., "*Downhill Simplex Method in Multidimensions" and "Linear Programming and the Simplex Method.*, §10.4 and 10.8 in *Numerical Recipes in FORTRAN: The Art of Scientific Computing, 2nd ed.* Cambridge, England: Cambridge University Press, pp. 402-406 and 423-436, 1992

[36] Rahmawati S. D., Integrated Field Modeling and Optimization, PhD Thesis, Norwegian University of Science and Technology, Department of Engineering Cybernetics, 2012

[37] Rojek L. & Tiwary D., *Real Separator Guide,* Aspen HYSYS ,2004

[38] SnowMcGregor K., Troubleshooting Gas Liquid Separators – Removal of Liquids from the Gas, John M. Campbell/ PetroSkills, 2014

[39] Svrcek W. & Monneray W., *Design of Two-Phase Separators within the Right Limits*, Fluids/Solids Handling p.53-60, Chemical Engineering Progress,2003

[40] Viles, J. C., *Predicting Liquid Re-entrainment in Horizontal Separators* (SPE 25474), Production Operations Symposium, 1993.