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Automated Process Design in Oil and Gas Field Development

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MASTER THESIS

for

Student John Swatson

Spring 2018

Automated process design in oil and gas field development

Automatiskert prosessdeisgn for olje og gass feltutvikling

Background and objective

A large part of worlds remaining oil and gas resources is found in harsh environments such as deep water and arctic conditions. The development of such oil and gas fields requires advanced process solutions for hydrate control, separation, dew point control and transport solutions. Furthermore Development of subsea processing technologies and concepts will be central in future field development.

The topics of this project will be to evaluate new processing ideas including process design applicable for subsea and offshore processing of natural gas. Simulation tools such as HYSYS will be used for process simulation. Necessary process design tools for equipment such as heat-exchangers, separators, compressors etc. should be developed.

Process design typically involves a number of persons in a company. The focus of this work will be to simplify the design process. An integrated design tool using HYSYS and Excel shall be developed.

The following tasks are to be considered:

1. Review of process equipment and design methods related to offshore gas processing
2. Development of automated process design methods
3. Process design for a field development case study
4. Suggestion of process design for case study

Within 14 days of receiving the written text on the master thesis, the candidate shall submit a research plan for his project to the department.

When the thesis is evaluated, emphasis is put on processing of the results, and that they are presented in tabular and/or graphic form in a clear manner, and that they are analyzed carefully.

The thesis should be formulated as a research report with summary both in English and Norwegian, conclusion, literature references, table of contents etc. During the preparation of the text, the candidate should make an effort to produce a well-structured and easily readable report. In order to ease the evaluation of the thesis, it is important that the cross-references are correct. In the making of the report, strong emphasis should be placed on both a thorough discussion of the results and an orderly presentation.

The candidate is requested to initiate and keep close contact with his/her academic supervisor(s) throughout the working period. The candidate must follow the rules and regulations of NTNU as well as passive directions given by the Department of Energy and Process Engineering.

Risk assessment of the candidate's work shall be carried out according to the department's procedures. The risk assessment must be documented and included as part of the final report. Events related to the candidate's work adversely affecting the health, safety or security, must be documented and included as part of the final report. If the documentation on risk assessment represents a large number of pages, the full version is to be submitted electronically to the supervisor and an excerpt is included in the report.

Pursuant to "Regulations concerning the supplementary provisions to the technology study program/Master of Science" at NTNU §20, the Department reserves the permission to utilize all the results and data for teaching and research purposes as well as in future publications.

The final report is to be submitted digitally in DAIM. An executive summary of the thesis including title, student's name, supervisor's name, year, department name, and NTNU's logo and name, shall be submitted to the department as a separate pdf file. Based on an agreement with the supervisor, the final report and other material and documents may be given to the supervisor in digital format.

☐ Work to be done in lab (Water power lab, Fluids engineering lab, Thermal engineering lab)
☐ Field work

Department of Energy and Process Engineering, 15. January 2018

Even Solbraa

Even Solbraa
Academic Supervisor

Research Advisor:

Abstract

Oil and Gas Processing plants require specialised equipment to effectively treat the hydrocarbons produced from the field. The processing equipment employed occupy a large amount of space and contribute a significant amount of weight to the platform which add to the cost of offshore structures. The design of an offshore oil and gas field incorporates both technical and economic factors that must be considered throughout the project life. The development concept, design and selection of process equipment, energy consumption, carbon footprint, commodity prices, tax regime and profitability are some of the factors that are critically investigated at each stage of project development. These indicators inform the decision criteria which underpin the feasibility of an oil and gas field development.

This master thesis presents an integrated automated model/tool that encompasses the technical and economic factors that can simplify the decision process. As a starting point, a hypothetical base case given a gas well composition and well parameters is used in this research. An offshore gas processing plant is modelled using ASPEN HYSYS in parallel with Microsoft Excel which was used to create equipment sizing calculators for each gas processing equipment. With such models, the impact to process design or to the entire project based on changes to technical and economic factors can be investigated. Different equations of state are also utilised to equally examine the influence on equipment design. The results from the base case showed that utilising different thermodynamic models can give up to ~ 3.5% difference in equipment weight and ~1.8% difference in footprint.

The calculator developed was taken a step further to incorporate automation. Automation of the sizing calculator was performed using Aspen Simulation Workbook to link MS Excel to Aspen HYSYS as well as visual basic codes to create the functionality that allows for investigating the process design based on changing parameters. The calculator/tool also presents an analytical model that gives results of design indicators including equipment footprint/weight, energy consumption, carbon footprint and cashflow (Net Present Value) depending on the development concept. As a myriad of technical and economic factors can impact an oil and gas field development, the thesis focusses on three hypothetical production profiles. The results of the analyses using the automated tool showed that producing at a high rate and quickly does not necessarily give the optimum results and/or high profitability. Also, with the "winning scenario" changing the thermodynamic model for the process simulation from Soave Redlich Kwong to Peng Robinson gave a significant relative difference of approximately 3.5% in equipment weight amounting to 22 tons and 5% in NPV which amounted to USD $ 12 million.

The research goes further to build up on the three scenarios and shows methods to determine the optimum production profile with the objective of maximising NPV. A trend was shown where increasing the flowrate (plateau production) increases the profitability of the project; however, beyond the optimum flowrate the capital expenditure of the project increases and the profitability of the project declines. The optimum flowrate of 8MMsm^3/d was determined.

Essentially, the master thesis has presented an automated tool capable of examining gas processing project indicators for field development. It gives a preliminary design of gas processing equipment and provides the functionality of analysing the effect of different thermodynamic models to the design. Furthermore, it enables investigative analysis into changing parameters during the production lifecycle.

Acknowledgement

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Trondheim, 5 June 2018

John Clement Swatson

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List of Abbreviations

ANSI	-	American National Standards Institute
APEA	-	Aspen Process Economic Analyser
API	-	American Petroleum Institute
ASME	-	Association Society of Mechanical Engineers
CAPEX	-	Capital expenditure
CO_2	-	Carbon Dioxide
CPA	-	Cubic Plus Association
Bara	-	Absolute pressure in bars
Barg	-	Gauge pressure in bars
EoS	-	Equations of State
HHC	-	Heavy hydrocarbon
H_2S	-	Hydrogen sulphide
ID	-	Internal diameter
MEG	-	Monoethylene glycol
MMscmd	-	Million standard cubic metre per day
NOK	-	Norwegian Kroner
OD	-	Outside diameter
OPEX	-	Operating expenditure
ppm	-	Part per million
PR	-	Peng Robinson
SRK	-	Soave Redlich Kwong
TEMA	-	Tubular Exchanger Manufacturers Association
TST	-	Twu-Sim-Tassone
USD	-	United States Dollar
NCS	-	Norwegian Continental Shelf
NPV	-	Net Present Value
YRS	-	Years

Nomenclature

Symbol		***Description***
A	-	Area
b	-	Correction factor for volume
C	-	Piping empirical Factor
C_d	-	Drag co-efficient
C_p	-	Specific Heat Capacity
C_uN_i	-	Copper Nickel
d_m	-	Particle diameter
D	-	Pipe Inside Diameter
f	-	Friction factor (dimensionless)
f	-	Polytropic Correction Factor
F	-	Cross-section area for gas flow
F	-	Correction Factor for Countercurrent Heat Exchanger
F_D	-	Drag force
F_G	-	Gravitational Force
G	-	Gas Gravity
g	-	Acceleration due to gravity
GC	-	Gas Chromatography
G_p	-	Cumulative Gas Production
h	-	Film transfer co-efficient
H	-	Height
H_p	-	Polytropic Head
H_T	-	Total Head
K	-	Equilibrium constant
K_s	-	Sizing Constant
L	-	Pipe Segment Length
$LMTD$	-	Logarithmic Temperature Difference
L_s	-	Seam-to-seam length
MW	-	Molecular Weight
n	-	Polytropic Exponent
N_b	-	Number of Baffles
P_b	-	Base Pressure
P_c	-	Pseudocritical pressure
PR	-	Pitch Ratio
P_t	-	Tube Pitch
PVT	-	Pressure Volume Temperature
P_1	-	Upstream Pressure
P_2	-	Downstream Pressure
Q	-	Flow rate at standard conditions (m^3/day)
q_a	-	Actual flow rate
q_g	-	Gas flowrate
q_{pp}	-	Production Potential flowrate
R	-	Universal Gas Constant
Re	-	Reynolds Number
s	-	Elevation factor
S	-	Allowable Stress
SR	-	Slenderness ratio

t	-	Time
t_w	-	Wall Thickness
T_b	-	Base Temperature
TBP	-	True-Boiling Point Distillation
T_c	-	Pseudocritical temperature
T_f	-	Average Gas flowing Temperature
T_r	-	Pseudoreduced Temperature
U	-	Heat Transfer co-efficient
u	-	Velocity
v	-	Velocity
V_m	-	Specific volume of mass
W_b	-	Weight of empty vessel shell
W_I	-	Weight of Internals
W_N	-	Weight of Nozzles
W_P	-	Weight of Piping
W_v	-	Weight of empty vessel
w_i	-	Weight fraction
X_i	-	Mol Fraction molecule i
y_i	-	Mol fraction
z	-	Gas Compressibility Factor at flowing temperature
z_i	-	Mol fraction

α	-	Correction Factor (degree of attraction)
ρ	-	Density
ω	-	Accentric Factor
μ	-	Viscosity
γ	-	Specific gravity
$\dot{m}$	-	Mass flow
κ	-	Isentropic exponent
η_p	-	Polytropic Efficiency
v	-	Specific Volume
ε	-	Pipe Roughness

1 Introduction

Gas processing plants, be it unmanned platforms or processing facilities, require specialised equipment to effectively treat the hydrocarbons produced from the field. The well-stream may consist of crude oil, gas, condensates, water and various contaminants. The objective for treating the gas is to;

- Ensure flow of the hydrocarbons hence transportability to the end user or process delivery system. This pertains to flow assurance to ensure the gas flows from one point to another without pushing the limits of the conduit in which it is transported, e.g. with respect to pressure rating of pipelines and vessels. Flow assurance challenges could include hydrate formation, scale formation and wax formation in gas condensate systems
- Protect and afford long life of process equipment such as compressors and consumer equipment.
- Meet quality specifications for sale.

With the advancement of unmanned offshore processing techniques, more innovative methods are being created thereby shifting the processing techniques to be performed on the seabed or offshore. Unmanned offshore natural gas platforms require a critical look at process design methods and the related preliminary equipment design. Offshore platform design face limitations with regards to weight and footprint. Such factors largely impact capital expenditure (CAPEX) and subsequently operating expenditure (OPEX) for the processing operations.

This master thesis titled "*Automated Process Design in Oil and Gas Field Development"* focusses on two main aspects with respect to natural gas processing on offshore unmanned platforms. Firstly, design and selection of gas processing equipment and secondly, automation of the offshore gas plant based on a case study using thermodynamic analysis. The stages of the natural gas encompassing processes such as condensate stabilisation, drying of natural gas and export processes are looked at critically to determine the effect of thermodynamic analyses on process parameters as well as the effect of these thermodynamic models on the design of equipment. The choice and sizing of equipment has in turn a major impact on the space requirements and weight limitations of the offshore structure as processing equipment takes a significant amount of space on offshore platforms. Automation of gas processing is performed to analyse factors such as the impact of varying parameters such as production flowrates on equipment weight and in turn project profitability.

The thesis takes into account a model gas composition scenario as given in *Table 1.1* and *Table 1.2* within the research work. ASPEN HYSYS simulation software was used in designing the offshore process and evaluating the equipment specifications.

Table 1.1: Process parameters and specifications (Solbraa, 2016)

Designations and Units	**Specification**
Well operating pressure [bara]	180
Well operating temperature [°C]	80
Platform inlet pressure [bara]	90
Platform inlet temperature [°C]	5
Platform outlet pressure [bara]	200
Platform outlet temperature [°C]	15
Flowrate at well [MMSm^3/d]	5
Sea water temperature [°C]	5
Water dew point specification [°C at 80bara]	-10
Hydrocarbon cricondebar Specification [bara]	90
Condensate/ oil specification [bara /°C]	1 / 20
Export pressure [bara]	200

Table 1.2: Well stream composition

Component	**Mole %**
Nitrogen	0.56
Carbon Dioxide	2.02
Methane	81.77
Ethane	7.77
Propane	3.91
i-Butane	0.56
n-Butane	0.90
i-Pentane	0.25
n-Pentane	0.24
n-Hexane	0.50
n-Heptane	0.30
n-Octane	0.20
n-Nonane	0.12
n-Decane	0.91

The master thesis focuses on four main tasks under the research topic; with each chapter detailing the work process, concepts and the build-up to the plant design and automation methods used.

The first section, Chapter 2, discusses the thermodynamic models. It touches on different thermodynamic models used predominantly in oil and gas processing. The chapter summarises the history and build-up of different Equations of State (EoS) and highlights their limitations and applications. The equations of state touched on are Redlich-Kwong, Soave-Redlich-Kwong (SRK) and Peng Robinson (PR). A further look into Pressure-Volume-Temperature (PVT) and fluid characterisation is detailed where the properties of reservoir fluids and mixing rules are discussed in relation to the thermodynamic models.

The second section, Chapter 3, presents a breakdown of the theory and fundamentals for sizing of gas processing equipment specifically with respect to *Separators*, *Heat Exchangers*, *Compressors*, *Pumps* and *Pipelines*. The theoretical design of the equipment incorporates industry standards such as API/ASME standards for design of the separators and pipeline as well as TEMA standards for design of heat exchangers. Based on theory, an equipment calculator was developed in *MS Excel* detailing simple design methods to efficiently size the gas processing equipment so as to investigate the impact of thermodynamic models on weight and footprint.

The third section, Chapter 4, simulates an offshore gas processing plant from a saturated gas stream based on the well composition and well parameter case scenario under *Table 1.1*. The simulation tool used in this thesis is ASPEN HYSYS v9.0. The offshore gas processing simulation stages are broken down into *Saturation of Gas Stream*, *Condensate Stabilisation*, *Hydrocarbon Dew Point Control* and *Export compression.*

The fourth section, Chapter 5, discusses the developed calculator and highlights examples for developing the sizing models for each equipment in the related Appendices. It further on utilises the theoretical sizing model to compare the equipment sizing in the HYSYS simulation based on different thermodynamic models and its impact on sizing parameters, weight and footprint of offshore equipment.

The fifth section, Chapter 0, gives an overview of the methodology used in automating the calculators developed in parallel with HYSYS. It shows a step-by-step approach in linking the two models i.e. *HYSYS* and *Equipment Calculator*. This involves setting up the required parameters to perform scenario analysis based on changes during the life cycle of the processing plant. It also outlines the visual basic code and programming involved in setting up the functionality to record data.

The sixth section, Chapter 7, outlines the analysis performed for the plant and process life cycle by examining three scenario production profiles. The analysis covers project indicators that determine the feasibility of the project in its entirety. The indicators captured are limited to *equipment sizing* and *weight* which translates to equipment cost, *carbon footprint* in relation to carbon intensity and emissions and *cash flow analysis* with respect to project revenues and costs.

The seventh section, Chapter 8, presents an overall summary and discussion of the results of the automated tool developed. It touches on how the calculator could be used as a tool for preliminary design models as well as an economic model for plant design. Different case scenarios are presented and a suggested case scenario for further rigorous study is presented.

The final section, Chapter 9, presents information on further research into the master thesis. Following the work from the master thesis this could be used as a tool and expanded to incorporate a more in-depth model covering reservoir to sales.

2 Thermodynamic Models

This chapter summarises the various thermodynamic models utilised within gas processing. It highlights the development and the history for both classical and more modern thermodynamic models. The master thesis herein analyses the gas process design utilising the Soave-Redlich-Kwong (SRK) and Peng-Robinson (PR) models in Aspen HYSYS. The effects of the models on the design of equipment sizing are highlighted and presented in *Chapter 5.*

The details of equation of state presented in this chapter are based on previous master thesis conducted in fall 2016, from experimental data in scientific articles and various references indicated herein (Whitson, Brule, & Society of Petroleum Engineers of AIME., 2000).

2.1 Equations of State

Over 100 equations of state have been developed in an attempt to improve on the ideal gas equation of state. British Chemist Robert Boyle performed experiments that supported the relation that gas volume varied inversely with pressure. This was the building block for further equations of state. Further on, Italian scientist, Amedeo Avogadro investigated the equation formulated by Boyle and the effects of molecules in a given volume and formulated what is currently being utilised and termed as the ideal gas law (*equation 2.1*)

$$PV_m = RT \tag{2.1}$$

where P represents pressure, V_m is volume, R is the gas constant and T is temperature.

An improvement over the ideal gas equation of state based on elementary molecular arguments was suggested by Johannes D. van der Waals, who noted that gas molecules actually occupy more than the negligibly small volume presumed by the ideal gas model and also exert long-range attractive forces on one another. In 1949, the equations and Van der Waals studies were modified by Redlich and Kwong which was further on utilised as the basis for both Peng-Robinson (PR, 1976) and Soave-Redlich-Kwong (SRK, 1972). PR and SRK derived the correlation factor for the attraction of molecules and temperature in gases. SRK and PR have become the most used equations of state for the development of models such as Cubic Plus Association (CPA) and the Twu-Sim-Tassone equation (TST).

2.1.1 Van der Waals Equation

The ideal gas is a hypothetical gas, whose molecules do not attract or repel one another, and their volume is negligible compared to a gas container. Real gases can approach the ideality only at low temperatures and pressures (<5atm). The repulsive forces of gas molecules tend to increase with the increasing temperature. With the increasing pressure, density of gas also increases, the molecules are closer to one another, and the intermolecular forces become significant to affect the motion of the molecules. In addition, the volume of real gas molecules also becomes a significant fraction of the total volume, thus causing deviations from the ideal gas behaviour.

Van der Waals equation was an improvement of the ideal gas laws incorporating correction to the volume of gas molecules and their interactions.

$$\left(P + \frac{\alpha}{V_m^2}\right)(V_m - b) = RT \tag{2.2}$$

The term $V_m - b$ refers to the "free volume", namely the free space where molecules can move around. The parameter α is an expression of the degree of attraction of gas molecules to each other. The parameter b is linked with the volume of the gas molecules and their repulsive forces. Both constants are unique for each gas molecule and are independent of pressure and temperature. External pressure P and attraction between molecules $\alpha/{V_m}^2$ act in the same direction, pushing molecules together. At equilibrium, this pressure is balanced by the thermal pressure $RT/(V_m - b)$, which is holding the molecules apart. Hence *equation 2.2* can be re-arranged to *equation 2.3*;

$$P = \frac{RT}{V_m - b} - \frac{\alpha}{V_m^2} \tag{2.3}$$

Van der Waals equation at middle pressures reasonably describes the behaviour of real gases but presents inconsistencies higher pressures, where repulsive forces prevail over attractive ones. The constants in the equation and critical parameters of a given gas are given by (Hurai, Huraiová, Slobodník, & Thomas, 2015):

$$P_c = \frac{\alpha}{27b^2}, T_c = \frac{8\alpha}{27bR}, V_c = 3b \tag{2.4}$$

Where P_c, T_c and $, V_c$ are the critical pressure, temperature and volume respectively. For a single component the critical pressure can be explained as the pressure above which liquid and vapour cannot coexist, regardless of temperature. Similarly, the critical temperature is the temperature above which a gas-liquid mixture cannot coexist, regardless of the pressure. In a multicomponent system, however, the two-phase region can extend beyond the systems critical point.

2.1.2 Redlich and Kwong

Van der Waals equation was modified by Redlich Kwong (MRK) in 1949 to improve the ability of the equation to reproduce fluid parameters at higher temperatures and pressures. MRK modifies the second term of *equation 2.2;*

$$\left(P + \frac{\alpha}{V_m(V_m + b)(T)^{0.5}}\right)(V_m - b) = RT \tag{2.5}$$

This allows MRK to be utilised for pure gases and their mixtures as well as for H_2O-CO_2 and NaCl fluids.

2.1.3 Soave-Redlich-Kwong (SRK)

The Soave-Redlich-Kwong equation of state (SRK) is developed from the Redlich-Kwong (MRK EoS) where modifications to the correction factor are given by *equation 2.6*;

$$P = \frac{R\,T}{V_m - b} - \frac{\alpha}{V_m(V_m - b)} \tag{2.6}$$

SRK uses the same equation as MRK; however Soave made some adjustments to the α factor.

$$\alpha = 0.42748\ \frac{R^2\, T_c^2}{P_c}\,[f(T)]^2 \tag{2.7}$$

The adjustment to the equation incorporates the function of the reduced temperature T_r and the accentric factor ω. Given by *equations 2.8* and *2.9*.

$$f(T) = 1 + k\left(1 - \frac{T}{T_c}\right) \tag{2.8}$$

$$k = 0.480 + 1.574\omega - 0.176\omega^2 \tag{2.9}$$

The accentric factor accounts for molecules without a spherical form. Molecules with a spherical form have an accentric factor equal to zero. The accentric factor was introduced by in 1955 by K. S Pitzer and is given by *equation 2.10*.

$$\omega = -\log_{10}\left(\frac{P^{sat}}{P_c}\right)_{T_r=0.7} - 1 \tag{2.10}$$

The volume correction factor b, was not changed in the analysis made by Soave and was maintained as in *equation 2.11*

$$b = 0.08664\ \frac{R\, T_c}{P_c} \tag{2.11}$$

The SRK equation of state presented a marked impact on calculation of hydrocarbons and represents one of the biggest advancements upon which cubic equations are built. (Robinson, Peng, & Chung, 1985)

2.1.4 Peng-Robinson

The Peng Robinson equation of state focusses on the natural hydrocarbon gas and petroleum systems. This similar to SRK equation of state except for a slightly better performance of the PR EoS around the critical point, making this EoS better suited for gas/condensate systems. The PR EoS has the following form:

$$\left(P + \frac{\alpha}{V_m(V_m + b) + b(V_m - b)}\right)(V_m - b) = RT \tag{2.12}$$

Peng Robinson conserved the temperature dependency of the attractive term and the acentric factor introduced by Soave. In addition, they presented different fitting parameters to describe this dependency and the coefficients. The correction factors are obtained as in the SRK equation, with a few changes. The SRK predicts a compressibility factor of 0.333 while PR predicts a value of 0.307. (Robinson et al., 1985).

$$\alpha = 0.45724 \frac{R^2 T_c^2}{P_c} [f(T)]^2 \tag{2.13}$$

$$b = 0.0778 \frac{R T_c}{P_c} \tag{2.14}$$

Changes to the function for the acentric factor k, can the correction factor α be calculated the same way as for SRK (equation 2.8)

$$f(T) = 1 + k\left(1 - \frac{T}{T_c}\right) \tag{2.15}$$

$$k = 0.37464 + 1.5422\omega - 0.26992\omega^2 \tag{2.16}$$

2.1.5 Cubic-Plus Association (CPA)

More modern equations of state have been developed based on the earlier equations of state. These models take into the hydrogen bonding interactions in ionic liquid systems. For example ionic liquid systems containing nitrogen, oxygen and fluorine can also form hydrogen bonding with other solvents like water and alcohols. Hence it is more accurate to account for the interactions into these models. The CPA EoS was proposed by Kontogeorgis et al. in 1996. Further on, later versions derived from the Peng Robinson EoS included an association term based on the stick-shield method.

The model is a combination of the regular cubic EoS and the association factor. The compressibility factor z is expressed as;

$$z = z_{cubic} + z_{assoc} \tag{2.17}$$

where the z_{cubic} represents the physical contribution and z_{assoc} represents the association contribution. This gives z_{cubic} as;

$$z_{cubic} = \frac{V_m}{V_m - V_b} - \frac{aV_m}{RT\,[V_m(V_m + b) + b\,(V_m - b)]} \tag{2.18}$$

where a and b are characteristic parameters based on the mixing rules highlighted under *Chapter 2.2.2*. The associated contribution to compressibility factor is given as;

$$z_{assoc} = \sum_i x_i \left(\frac{1}{X_i} - \frac{1}{2}\right) \rho_0 \left(\frac{\partial X_i}{\partial \rho_0}\right) \tag{2.19}$$

where X_i represents the mole fraction of molecule i not bonded, x_i is the mole fraction of component i, and $\rho_0$is the total molecule number density. (Ma et al., 2011)

2.2 PVT and Fluid Characterisation

Natural gas is composed primarily of low-molecular weight alkanes; methane through butane, carbon dioxide, hydrogen sulphide, nitrogen in some cases lesser quantities of helium, hydrogen, CO and carbonyl sulfide. The temperature and pressure gradients on a formation may cause reservoir-fluid properties to vary as a function of depth referred to as "compositional grading". (Whitson et al., 2000).

It is important to understand the composition of petroleum reservoir fluids at the onset as this aids in defining the value of the end product for market as well as the subsequent field development solution, which in turn encompasses wells and flowline design, processing equipment, pipeline transport systems and offloading systems.

This section explains the fluid characterisation methods employed. As the simulation ASPEN HYSYS was used; the section explains the theory and fundamentals of phase behavior and EoS employed by the software in characterizing the fluid composition.

2.2.1 Properties of Reservoir Fluids

Hydrocarbon with seven or more carbon atoms are called C_{7+} components. Petroleum reservoir fluids may contain hydrocarbons as heavy as C_{200}. A particular C_{7+} component falls under the following component classes also referred to as Paraffins-Napthenes-Aromatics (PNA) distribution;

Paraffins or Alkanes: These are carbon atoms that are connected by single bonds. Paraffins are divided into normal paraffins (*n-paraffins*) and iso-paraffins (*i-paraffins*). Paraffinic compounds consist of hydrocarbon segments of the type C, CH, CH_2, or CH_3.

Naphthenes or *Cycloalkanes*: These are similar to paraffins but contain one more cyclic structure. The segments in the ring structures are connected by single bonds. e.g. *Cyclohexane* and *methyl cyclopentane.*

Aromatics: Aromatics are similar to alkanes in that they contain one or more cyclic structures but have the carbon atoms connected by aromatic double bonds. e.g *Benzene.*

Due to different components, reservoir fluids cover a wide range of component properties for e.g. boiling points as depicted under *Appendix A.*

The pure component vapour pressures and critical points are essential in calculations of component and mixture properties. The pure component vapour pressures are experimentally determined by measuring the corresponding values of temperature (T) and pressure (P) at which the substance undergoes a transition from liquid to gas. *Figure 2.1* shows the vapour pressure for methane and benzene as pure components and as a mixture. The critical points, *CP*, shown are different for the pure components and mixture signifying the difference in phase behavior.

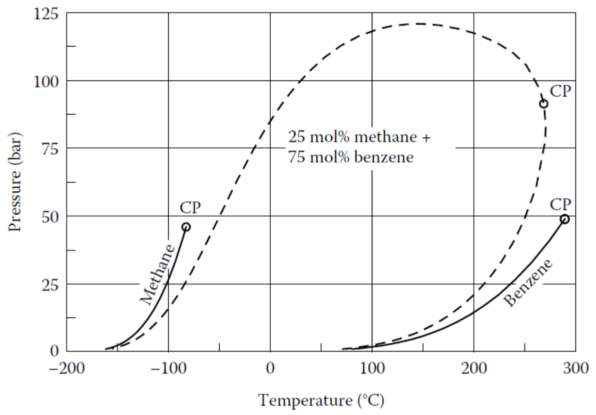


Figure 2.1: Vapour pressure curves for methane and benzene (full drawn line). Phase envelope (dashed line)

Some phase-behaviour applications require the use of an Equation of State (EoS - r*eference Chapter 2.1*) to predict the properties of petroleum reservoir fluids. The critical properties, acentric factor, molecular weight and Binary Interaction Parameters (BIP) are required for EoS calculations. The challenge is the chemical separation to identify the properties of many hundreds and thousands of components in reservoir fluids especially for compounds heavier than C_{20}.
The characterisation of C_{7+} fractions are done by;

1. Splitting the C_{7+} fractions into a number of fractions with known molar compositions.
2. Identifying the properties; molecular weight, specific gravity and boiling point of each fraction.
3. Estimating the critical properties and acentric factor of each C_{7+} fraction and key BIP for the specific EoS used.

For complete C_{7+} characterisation into discrete fractions the *True-Boiling Point (TBP)* distillation method provides necessary data as highlighted above. *Gas Chromatography (GC)* is a less-expensive, time-saving option to the TBP distillation method which only quantifies C_{7+} mass fractions and does not provide analysis of properties such as specific gravity, molecular weight and boiling point (Whitson et al., 2000). *Appendix A* and *Table 2.1* show examples of the results of true boiling point distillation.

Table 2.1: Experimental TBP results for a North Sea condensate (Whitson, C.H and Brule, M. R. 2000 Phase behaviour, Richardson, TX: Henry L. Doherty Memorial Fund of AIME, Society of Petroleum Engineers)

Fraction	Upper T_{bi} (°F)	Average T_{bi}* (°F)	m_i (g)	γ_i**	M_i (g/mol)	V_i (cm^3)	n_i (mol)	w_i (%)	x_{Vi} %	x_i %	Σw_i %	Σx_{Vi} %	K_w
C_7	208.4	194.0	90.2	0.7283	96	123.9	0.940	4.35	4.80	7.80	4.35	4.80	11.92
C_8	258.8	235.4	214.6	0.7459	110	287.7	1.951	10.35	11.15	16.19	14.70	15.95	11.88
C_9	303.8	282.2	225.3	0.7658	122	294.2	1.847	10.87	11.40	15.33	25.57	27.35	11.82
C_{10}	347.0	325.4	199.3	0.7711	137	258.5	1.455	9.61	10.02	12.07	35.18	37.37	11.96
C_{11}	381.2	363.2	128.8	0.7830	151	164.5	0.853	6.21	6.37	7.08	41.40	43.74	11.97
C_{12}	420.8	401.1	136.8	0.7909	161	173.0	0.850	6.60	6.70	7.05	48.00	50.44	12.03
C_{13}	455.0	438.8	123.8	0.8047	181	153.8	0.684	5.97	5.96	5.68	53.97	56.41	11.99
C_{14}	492.8	474.8	120.5	0.8221	193	146.6	0.624	5.81	5.68	5.18	59.78	62.09	11.89
C_{15}	523.4	509.0	101.6	0.8236	212	123.4	0.479	4.90	4.78	3.98	64.68	66.87	12.01
C_{16}	550.4	537.8	74.1	0.8278	230	89.5	0.322	3.57	3.47	2.67	68.26	70.33	12.07
C_{17}	579.2	564.8	76.8	0.8290	245	92.6	0.313	3.70	3.59	2.60	71.96	73.92	12.16
C_{18}	604.4	591.8	58.2	0.8378	259	69.5	0.225	2.81	2.69	1.87	74.77	76.62	12.14
C_{19}	629.6	617.0	50.2	0.8466	266	59.3	0.189	2.42	2.30	1.57	77.19	78.91	12.11
C_{20}	653.0	642.2	45.3	0.8536	280	53.1	0.162	2.19	2.06	1.34	79.37	80.97	12.10
C_{21+}			427.6	0.8708	370	491.1	1.156	20.63	19.03	9.59	100.00	100.00	
Sum			2,073.1			2,580.5	12.049	100.00	100.00	100.00			
Average				0.8034	172								11.98

Reflux ratio = 1 : 5; reflux cycle = 18 seconds; distillation at atmospheric pressure = 201.2 to 347°F; distillation at 100 mm Hg = 347 to 471.2°F; and distillation at 10 mm Hg = 471.2 to 653°F.

$V_i = m_i/\gamma_i/0.9991$; $n_i = m_i/M_i$; $w_i = 100 \times m_i/2073.1$; $x_{Vi} = 100 \times Vi/2580.5$; $x_i = 100 \times n_i/12.049$; $\Sigma w_i = \Sigma w_i$; $\Sigma x_{Vi} = \Sigma x_{Vi}$; and $K_w = (T_{bi}+460)^{1/3}/\gamma_i$.

*Average taken at midvolume point.

**Water = 1.

An important factor, the acentric factor, ω, proposed by Kenneth Pitzer (1955) is a measure of the non-sphericity (centricity) of molecules or the measure of the curvature of the pure component vapour pressure curve. The acentric factor of n-paraffins increases with carbon number. That is, methane has an acentric factor of 0.008, ethane 0.098 and propane 0.152. *Figure* 2.2 gives a representation of different acentric factors of components with same critical point.

As defined by Pitzer (1955), ω ,is given by

$$\omega = -1 - log_{10}\left(\frac{P^{sat}}{P_c}\right)_{T=0.7T_c} \tag{2.20}$$

where P^{sat} stands for saturation pressure or vapour pressure and is given by *equation 2.21* which when plotted against the reciprocal of the reduced temperature, T_r ,(given by *equation 2.22*) for most pure substances gives an approximate straight line. By definition, ω, is zero for noble gases; argon, krypton and xenon and very close to zero for other spherical molecules. *Figure* 2.3 shows the logarithm plot of the reduced component vapour pressure against the reciprocal of the reduced temperature, T_r.

$$P^{sat} = \frac{P^{sat}}{P_c} \tag{2.21}$$

$$T_r = \frac{T}{T_c} \tag{2.22}$$

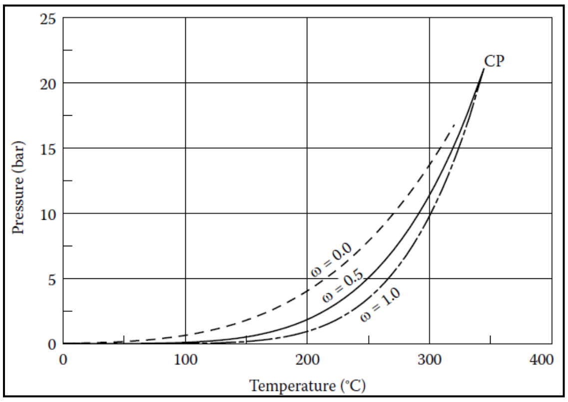


Figure 2.2: Vapour pressure curves of component with same critical point as nC_{10} and different acentric factors (Pedersen, Christensen, & Shaikh, 2015)

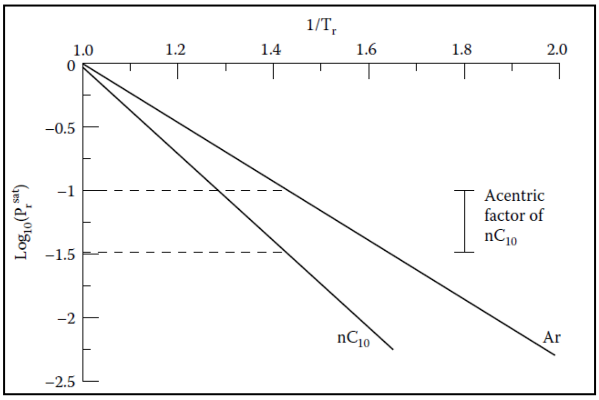


Figure 2.3: Acentric factor of nC_{10} from vapor pressure curves of Ar and nC_{10}. (Pedersen, Christensen, & Shaikh, 2015)

As petroleum reservoir fluids are multicomponent mixtures, the phase behaviour of the fluid (e.g. natural gas) must be characterised incorporating the vapour pressure curves of the components in a *Phase Envelope* as illustrated in *Figure 2.4.*

Petroleum reservoir fluids are divided into;

- Natural Gas mixtures
- Gas Condensate mixtures
- Near-critical mixtures or volatile oils
- Black oils
- Heavy oils

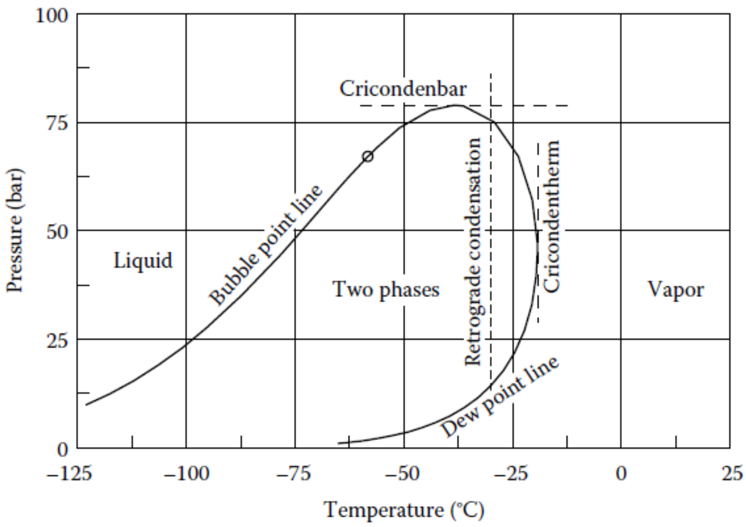


Figure 2.4: Phase envelope of natural gas (Pedersen, Christensen, & Shaikh, 2015)

Appendix B illustrates examples of each type of reservoir fluid. The classifications of the fluids are distinguished by the position of critical temperature of the mixture relative to the reservoir temperature. During production the reservoir temperature T_{res} remains fairly constant however the pressure decreases with production. The phase behaviour of the different fluid types differs with production.

As depicted in *Figure 2.5*, for a natural gas, there would be no impact of the number of phases as the gas remains in a single phase at all pressures. For a gas condensate, the pressure reduction will result in a second liquid phase below the dew point.

Near-critical mixtures have their critical temperatures close to the reservoir temperatures. For near-critical mixtures, a reduction pressure will also result in a second gas phase at the bubble point branch. This mixture is classified as a volatile oil. In the case where the reservoir temperature is slightly higher, indicated in *Figure 2.5* by T'_{res}, the pressure reduction will introduce a second liquid phase at the dew point line and resulting in a mixture classified as *gas condensate* mixture. The compositions and properties of the gas and liquid phases within the phase envelope are similar. (Pedersen, Christensen, & Shaikh, 2015)

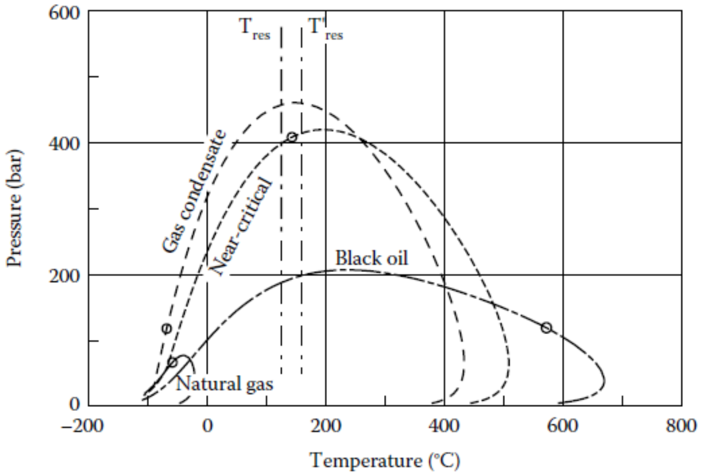


Figure 2.5: Phase envelope of various types of reservoir fluids (Pedersen, Christensen, & Shaikh, 2015)

2.2.2 Mixing Rules

Reservoir fluids contains mixtures of hundreds of components. The components are characterised based on mole, weight, and volume fractions. For a mixture with N components, $i = 1, \dots, N$, hence the overall mole fractions are given by

$$z_i = \frac{n_i}{\sum_{j=1}^{N} n_j} = \frac{m_i/M_i}{\sum_{j=1}^{N} m_j/M_j} \tag{2.23}$$

Given n=moles, m= mass, M = molecular weight/Molar mass; where the sum of z_i equals 1.0. Compositions for oil are denoted by x_i and gas compositions by y_i.
Mass fractions are given by *equation 2.24*; where w_i equals 1.0.

$$w_i = \frac{m_i}{\sum_{j=1}^{N} m_j} = \frac{n_i M_i}{\sum_{j=1}^{N} n_j/M_j} \tag{2.24}$$

The volume fractions based on component densities at standard conditions

$$x_{vi} = \frac{m_i/\rho_i}{\sum_{j=1}^{N} m_j/\rho_j} = \frac{n_i\, M_i/\rho_i}{\sum_{j=1}^{N} n_j M_j/\rho_j} = \frac{x_i\, M_i/\rho_i}{\sum_{j=1}^{N} x_j M_j/\rho_j} = \frac{x_i\, M_i/\gamma_i}{\sum_{j=1}^{N} x_j M_j/\gamma_j} \tag{2.25}$$

Some equations of state may give good approximations at low pressures and high temperatures, however at low temperatures and high pressures the impact of intermolecular interaction on gas behaviour increases. It is pertinent to account for the effect of these interactions on the relationship between pressure, volume and temperature. One such method in defining component fractions by average properties of mixtures is the use of Kay's mixing rule which uses the mole-fraction average given by *equation 2.26*. The mixing rule is acceptable for molecular weight, pseudo-critical temperature and acentric factor.

$$\ddot{\theta} = \sum_{i=1}^{N} z_i \theta_i \tag{2.26}$$

For a more generalised linear mixing rule;

$$\ddot{\theta} = \frac{\sum_{i=1}^{N} \Phi_i \theta_i}{\sum_{i=1}^{N} \Phi_i} \tag{2.27}$$

where Φ_i may represent one of the following: $\Phi_i = z_i$ (mole fraction) , $\Phi_i = w_i$ (weight fraction), $\Phi_i = x_{vi}$ (volume fraction).

With respect to each EoS (highlighted in *Chapter 2.1*) the "attraction" parameter, α, and "repulsion" parameter, b, needs to be expressed in a form to account for multi component systems. For a vapour phase with composition, y_i, the parameters are given by:

$$\alpha = \sum_{i=1}^{N} \sum_{j=1}^{N} y_i y_j A_{ij} \tag{2.28}$$

$$b = \sum_{i=1}^{N} y_i B_i \tag{2.29}$$

$$A_{ij} = \left(1 - k_{ij}\right)\sqrt{A_i A_j} \tag{2.30}$$

where k_{ij} represents binary interaction parameters given $k_{ii} = 0$, $k_{ij} = k_{ij}$. Also $k_{ij} = 0$, for most hydrocarbon-hydrocarbon pairs, with the exception of pairs of C_1 and C_{7+}. For Non-hydrocarbon-hydrocarbon pairs $k_{ij} \approx 0.1$ to 0.15 for nitrogen-HC pairs and CO_2-HC pairs. (Whitson et al., 2000)

The Aspen HYSYS simulation used for this research gives the *BIP* under a chosen fluid package and presents interaction parameters for each component pair as shown in *Figure 2.6*.

Equation of State Interaction Parameters

	Nitrogen	CO2	Methane	Ethane	Propane	i-Butane	n-Butane	i-Pentane	n-Pentane	n-Hexane	n-Heptane	n-Octane	n-Nonane
Nitrogen	---	-0,01710	0,03120	0,03190	0,08860	0,13150	0,05970	0,09300	0,09360	0,16500	0,07999	0,07999	0,07999
CO2	-0,01710	---	0,09560	0,14010	0,13680	0,13680	0,14120	0,12970	0,13470	0,14200	0,10920	0,13500	0,13500
Methane	0,03120	0,09560	---	0,00224	0,00683	0,01311	0,01230	0,01763	0,01793	0,02347	0,02886	0,03416	0,03893
Ethane	0,03190	0,14010	0,00224	---	0,00126	0,00457	0,00410	0,00741	0,00761	0,01141	0,01532	0,01932	0,02302
Propane	0,08860	0,13680	0,00683	0,00126	---	0,00104	0,00082	0,00258	0,00270	0,00514	0,00789	0,01085	0,01370
i-Butane	0,13150	0,13680	0,01311	0,00457	0,00104	---	0,00001	0,00035	0,00039	0,00157	0,00322	0,00521	0,00725
n-Butane	0,05970	0,14120	0,01230	0,00410	0,00082	0,00001	---	0,00050	0,00055	0,00187	0,00365	0,00575	0,00788
i-Pentane	0,09300	0,12970	0,01763	0,00741	0,00258	0,00035	0,00050	---	0,00000	0,00044	0,00146	0,00288	0,00445
n-Pentane	0,09360	0,13470	0,01793	0,00761	0,00270	0,00039	0,00055	0,00000	---	0,00039	0,00137	0,00276	0,00430
n-Hexane	0,16500	0,14200	0,02347	0,01141	0,00514	0,00157	0,00187	0,00044	0,00039	---	0,00030	0,00107	0,00210
n-Heptane	0,07999	0,10920	0,02886	0,01532	0,00789	0,00322	0,00365	0,00146	0,00137	0,00030	---	0,00024	0,00082
n-Octane	0,07999	0,13500	0,03416	0,01932	0,01085	0,00521	0,00575	0,00288	0,00276	0,00107	0,00024	---	0,00017
n-Nonane	0,07999	0,13500	0,03893	0,02302	0,01370	0,00725	0,00788	0,00445	0,00430	0,00210	0,00082	0,00017	---
n-Decane	0,12790	0,13390	0,04361	0,02673	0,01663	0,00945	0,01016	0,00621	0,00603	0,00335	0,00166	0,00064	0,00015
H2O	-0,67660	-0,11585	0,50000	0,50000	0,48190	0,51800	0,51800	0,48000	0,48000	0,51090	0,48000	0,48000	0,48000

Figure 2.6: Interaction Parameters for Fluid Components (ASPEN HYSYS)

2.2.3 K-Value Correlation

K*-value* is defined as the ratio of equilibrium gas composition, y_i, to the equilibrium liquid composition x_i. K_i is function of pressure, temperature and overall composition. K -values are estimated by empirical correlations or by satisfying equal-fugacity constraint with an EoS. Empirical correlations of K-value are useful in applications involving;

- Multi-stage surface operations (e.g. multistage flash separation)
- Compositional reservoir material balance
- Checking the consistency of separator-oil and gas compositions

There are several methods utilised in the correlation of K-values but are all based on two limiting conditions to describe the pressure dependence of K-values i.e. at low and high pressure.

At *low pressures,* below ~6 bara *Raoult's* and *Daltons* Law for ideal solutions provide a simplified approach for predicting equilibrium ratios as given in *equation 2.31*:

$$K_i = \frac{P_{vi}(T)}{P} \tag{2.31}$$

where P_{vi} is the component vapour pressure at the system temperature. *Equation 2.31* is limited as the temperature must be less than the component critical temperature and behaves as an ideal gas. Based on this, K-value is independent of overall composition.

Raoult's law states that the partial pressure, P_i , of a component in a multicomponent system is the product of its mole fraction in the liquid phase, x_i, and the vapour pressure of the component, P_{vi}, given as *equation 2.32*

Dalton's Law states that the partial pressure, P_i , of a component is the product of its mole fraction in the gas phase, y_i, and the total pressure of the system, P, given as *equation 2.33*. Combining Raoult and Dalton's Laws gives the correlation in *equation 2.31*.

$$P_i = x_i \, P_{vi} \tag{2.32}$$

$$P_i = y_i P \tag{2.33}$$

At high pressure, the K-value of all components in a mixture tend to converge to unity at the same pressure termed the *Convergence Pressure*. For binary mixtures this is the actual mixture critical pressure; however for multi-component mixtures, the convergence pressure is a non-physical condition unless the system temperature equals the mixture critical temperature. This is due to the fact that mixtures become single phase at bubble point or dew-point pressure before reaching the convergence pressure. The log-log plot of K_i against pressure represents how the ideal gas and convergence pressure conditions define the K-value behaviour at limiting conditions.

With respect to lighter components (where $T > T_{ci}$), K-values decrease monotonically toward the convergence pressure whereas for heavier components where (where $T < T_{ci}$), K-values initially decrease as a function of pressure at low pressures, passing through unity when system pressure equals the vapor pressure of a particular component, reaching a minimum, and finally increasing toward unity at the convergence pressure.

For reservoir fluids, the pressure at which K-values reach a minimum is usually >1,000 psia, indicating that Kvalues are more or less independent of convergence pressure/composition at pressures < 1000psia. (Whitson et al., 2000)
3

3 Review and Design of Equipment

This chapter reviews design methods in sizing gas processing equipment. It takes an in-depth look at fundamental theoretical procedures in sizing 2-phase and 3-phase separators, shell and tube heat exchangers, centrifugal compressors, sea water pumps and pipeline systems. The fundamental concepts highlighted are used to develop a tool; *Equipment Sizing Calculators* in MS Excel, for each processing equipment.

The calculator gives a summarised output of the equipment covering footprint (length, width and height) and mechanical design (thickness and weight). The design methods incorporate global standards and manufacturer specifications to give as close to accurate standard designs as possible.

3.1 Separation Train

The separation train focusses on the initial bulk removal process upon receiving the wet gas from the field. The separation equipment could be a two-phase separator – to allow for the separation of gas and oil/water, 3-phase separator – to allow for the separation of oil, gas and water. Inclusive in the system are flow control valves to decrease pressure to the required pressure level.

Within this master thesis, two methods are considered herein in performing sizing calculations for the separators. These are with respect to;

i. API Specification 12 J standards (based on two major references *Gas Conditioning and Processing* from Campbell, John; Maddox Robert and *Separator Sizing of Two-phase and Three-phase separators* by Monnery, Wayne and Svrcek, William. (Campbell, 1999 #2) and (Svrcek & Monnery, 1993)
ii. Fundamental theory highlighted in *Petroleum and Gas Field Processing* by Abdel-Aal, H.K ; M. Aggour and M. A Fahim which gives a detailed explanation of the theory. (Abdel-Aal, Aggour, & Fahim, 2003)

3.1.1 Two-Phase Separator

The two-phase separator is used to separate gas from oil in oil fields, or gas from oil/water for gas fields. The hydrocarbon mixtures to be separated contain essentially three main groups of hydrocarbon. (Abdel-Aal et al., 2003)

1. Light group, which consists of CH_4 (methane) and C_2H_6 (ethane)
2. Intermediate group, which consists of two subgroups; propane/butane (C_3H_8/C_4H_{10}) group and the pentane/hexane group (C_5H_{12}/ C_6H_{14})
3. Heavy group, which is the bulk of crude oil and identified as C_7H_{16}

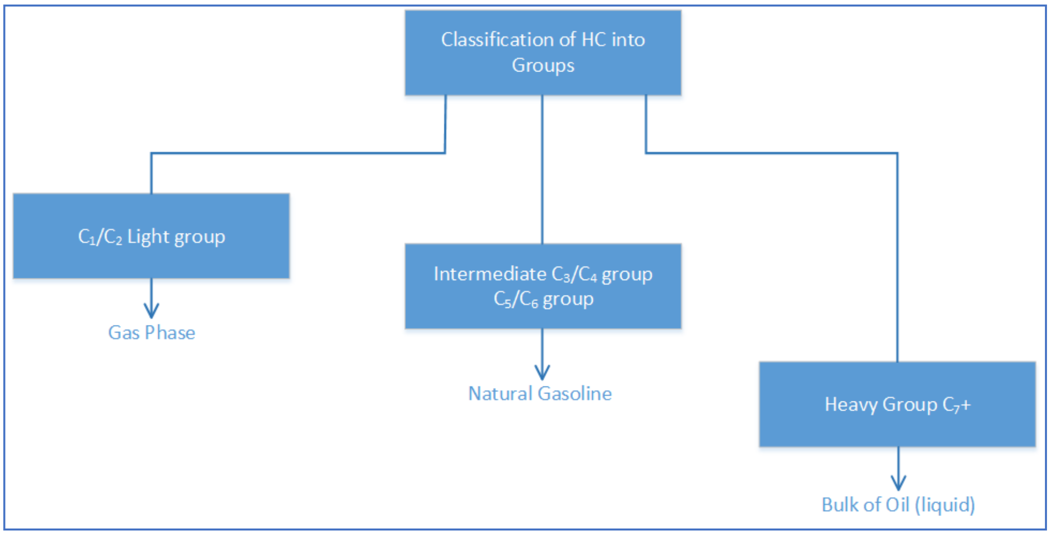


Figure 3.1: Classification of hydrocarbons in wellhead fluids (Abdel-Aal et al., 2003)

Within the separation process, the objective is to

- Separate the light gases mostly C_1 and C_2 gases from oil
- Maximize the recovery of heavy components of the intermediate group in the crude oil
- Save the heavy group components in liquid product.

Separation methods can be broadly classified into two main methods (Abdel-Aal et al., 2003);

1. Conventional Methods
2. Modified methods: this involves
   a. Including vapour recompression unit to the conventional method
   b. Replacing the conventional method by a stabiliser and a recompression unit

For the purpose of this thesis, focus is placed on the conventional method of separation. The conventional separator is the first vessel through which the gas from the wellstream flows. For some special cases there are heaters, water knock out drums upstream of the separator. The conventional separator is designed to achieve the following;

- Decrease in the flow velocity and optimum retention time allowing for the separation of gas and liquid by gravity
- Operation above the hydrate point of the flowing gas.

The choice of the configuration of a separator is based on a number of reasons. *Figure 3.2* gives the classification based on the application and operating conditions. The vertical separator occupies less ground area and is easier to clean. The horizontal separator can handle foaming crude oil better and is claimed to be more economical for handling large gas volumes. The spherical separator is easier to install and is more compact and adaptable for portable use.

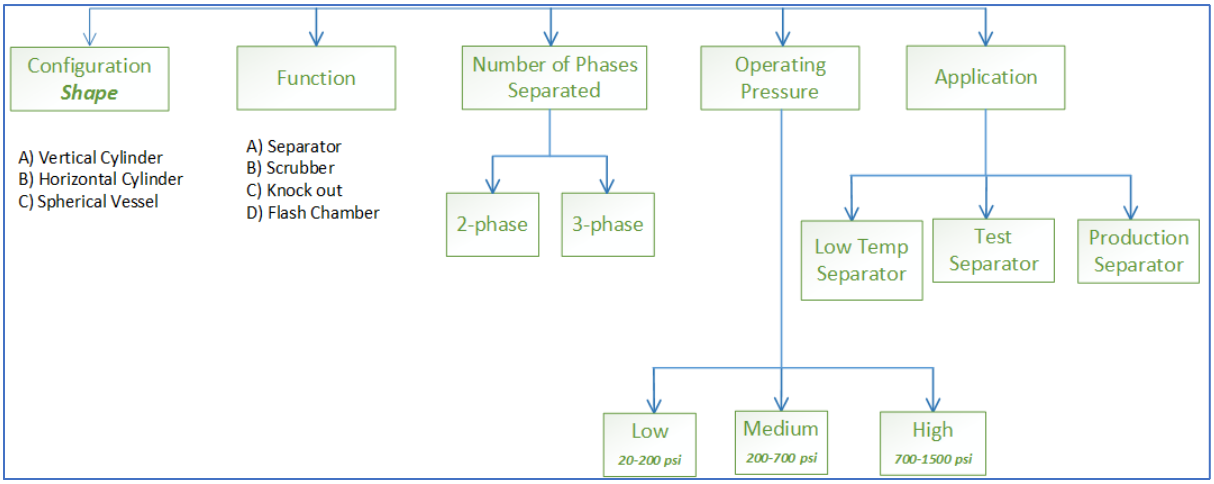


Figure 3.2: Separator classification (Abdel-Aal et al., 2003)

3.1.1.1 *Components of a Separator*

Gas-oil separators are equipped with control equipment for fluid level and pressure control as well as internal components to allow for the separation process. The control equipment include;

Liquid Level Controller – this is used to maintain the level of the fluid within the separator at a predetermined liquid height. This is achieved via a float and an automatic diaphragm motor valve on the oil outlet. The signal causes the valve to either open or shut, allowing or preventing more fluid into the separator hence maintaining the liquid level.

Pressure Control – the pressure control valve (PCV) is an automatic backpressure valve located on the outlet gas stream. The valve is set at a prescribed pressure that opens and closes automatically allowing more or less gas to flow out of the separator to maintain a fixed pressure inside the separator.

Pressure Relief Valve – this is a safety equipment used to prevent overpressure within the separator. It is set to a design pressure and vents pressure when the design limit is exceeded. The internal components of the separator include;

Mist Extractor – this section of the separator removes liquid mist or very fine liquid droplets from a gas stream via impingement, flow direction/velocity change, centrifugal force, filters or coalescing packs.

Several types of mist extractors are installed in separators. The types available are;

- *Wire-Mesh Mist Extractor* – these are made from stainless steel wire which are wrapped into tightly packed cylinder.
- *Vane- Mist Extractor* – This extractor type is made up of a series of closely spaced parallel and corrugated plates. It works on the principle that as the flow of fluid changes direction between the plates they impinge on the surface of the plates, thereby coalescing and falling down to the liquid section below.
- *Centrifugal Mist -Extractor*- this type of extractor utilises the principle of centrifugal force to separate the liquid droplets from the gas.

Within this thesis, the wire-mesh and vane type mist extractor are considered in the design as these are the most common design of internals in separators.

Inlet Diverters – This could be in the shape of a flat plate, spherical dish, a cone or centrifugal type. The inlet diverter causes the first bulk separation of liquid and gas. This occurs due to rapid change in velocity of the flowing gas stream and also separation due to difference in fluid densities.

Wave Breakers – Predominantly used in horizontal separators. The wave breakers are vertical baffles installed perpendicular to the flow direction to prevent unsteady fluctuations in the liquid level which would otherwise negatively affect the performance of the liquid level controller.

Defoaming Plates – Foam created in the separator occupies large space that could otherwise be available for the separation process. This causes a reduction in the separator efficiency. Also foam could affect the operation of the liquid level controller.
Vortex Breaker – This is utilised in the liquid exit of the separator to prevent vortices that could entrain gas in the liquid (*gas blowby*).

Sand Jets and Drains – Produced fluids from the wells may contain formation sand which could settle and accumulate at the bottom of the separator. Vertical separators are preferred in this case. In the case of horizontal separators, sand jets and drains may be installed where produced water is injected through the jets to fluidize the accumulated sand and is removed through the drains.

3.1.1.2 Design and Sizing of 2-phase Separators

The most important design factors in the design of separators are length and diameter and this depends on the fluid flow rates and operating conditions. The design factors considered are based on basic theories and assumptions to obtain as close to accurate design parameters for the gas model flow scenario. For the design factors, the following assumptions are made (Abdel-Aal et al., 2003):

1. No oil foaming takes place during the gas-oil separation otherwise retention time has to be drastically increased as foam occupies a large space in the separator and reduces the efficiency of the separation)
2. The cloud point of the oil and the hydrate point of the gas are below the operating temperature.
3. The smallest separable liquid drops are spherical ones having a diameter of 100 µm.
4. Liquid carryover with the separated gas does not exceed 13l/MMsm^3

The sizing and design criteria for vertical and horizontal separators differ and are based on gas or liquid sizing constraints. To understand the constraints it is required to consider the relative motion existing between particles during the separation process.

3.1.1.3 Theory behind Droplet Separation – 2-Phase Separators

Relative motion exists between the liquid particles and the surrounding fluid which is the gas particles. The liquid droplet has a greater density than the gas and tends to move vertically downward under gravitational or buoyant force, F_g. The gas conversely exerts a drag force, F_d, on the liquid droplet in the opposite direction. The liquid droplet will accelerate until the frictional resistance of the fluid drag force, F_d, approaches and balances gravitational force, F_g. After which the liquid droplet continues to fall at a constant terminal or settling velocity. (Abdel-Aal et al., 2003)

The drag force is proportional to the droplet surface area and perpendicular to the direction of gas flow, and its kinetic energy per unit volume:

$$F_d = C_d \frac{\pi}{4} d^2 \frac{\rho_g u^2}{2} \tag{3.34}$$

whereas F_g *is given by*

$$F_g = \frac{\pi}{6} d^3 \left(\rho_o - \rho_g\right) g \tag{3.35}$$

where C_d is the drag co-efficient and, d is the diameter of the oil droplet, u is the settling velocity of the oil droplet, ρ_o , and ρ_g are densities and g is the gravitational acceleration. The settling terminal velocity, u , is reached when $F_d = F_g$. Equating *equations 3.34* and *3.35* gives the droplet settling velocity as:

$$u^2 = \frac{8}{6} g \frac{\rho_o - \rho_g}{\rho_g} \left(\frac{d}{C_d}\right)$$

The droplet diameter expressed in microns is given as 1µm and acceleration due to gravity as 9.81m/s^2 the above equation gives:

$$u_s = 3.617 \times 10^{-3} \left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right) \frac{d_m}{C_d}\right]^{1/2} \quad m/s \tag{3.36}$$

3.1.1.4 Separator Gas Capacity

To evaluate the sizing of a separator the gas capacity of the separator is determined. The volumetric flow rate of the gas processed by the separator is related to the cross-sectional area and the maximum allowable gas velocity.

$$Q_g = \frac{A_g u}{35.313} \quad m^3/s \tag{3.37}$$

The above *equation 3.37* is given in m^3/s, however in standard pressure and temperature reported in million standard cubic metres per day MMscmd is given by

$$Q_g = 0.0865 \left(\frac{P}{TZ}\right) A_g u \quad MMscmd \tag{3.38}$$

which gives the gas velocity as

$$u = 0.09967\, Q_g \left(\frac{TZ}{P}\right) \left(\frac{1}{A_g}\right) \frac{m}{s} \tag{3.39}$$

3.1.1.5 Separator Liquid Capacity

The *liquid capacity* of the separator is given by the volume occupied by the liquid and the retention or residence time, t. This is given as below where 1ft^3/min = 0.0283168m^3/min = 257 bbl/day and V_o , is the volume of the separator occupied by oil and Q_o, the oil capacity of the separator.

$$Q_o\left(\frac{bbl}{day}\right) = 257\ \frac{V_o}{t} \tag{3.40}$$

3.1.1.6 Vertical Separator Sizing

The sizing of a vertical separator is determined by the gas capacity constraint.

3.1.1.7 Gas Capacity Constraint

For vertical separators, the upward average gas velocity should not exceed the terminal velocity of the smallest liquid droplet to be separated . This is given by equating *equations 3.36* and *3.39* which results

$$0.09967Q_g\left(\frac{TZ}{P}\right)\left(\frac{1}{A_g}\right) = 3.617\ \times\ 10^{-3}\ \left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \tag{3.41}$$

Substituting

$$A_g = \frac{\pi}{4}\ D^2$$

where D represents the internal diameter of the separator in metres and solving for D;

$$D^2 = 35.085\ Q_g\ \left(\frac{TZ}{P}\right)\left[\frac{\rho_g}{(\rho_o-\rho_g)}\ \frac{C_d}{d_m}\right]^{1/2}\ \mathrm{m}^2 \tag{3.42}$$

Equation 3.42 gives the *minimum acceptable diameter* of the separator. Larger diameters result in lower gas velocities hence better separation of the oil droplets from the gas. Smaller separator diameters give higher gas velocities hence causing liquid droplets to be carried over with the gas.

The drag co-efficient, C_d, is determined from the equation below which is related to the Reynolds number; this is given empirically for a spherical particle with Re in the range of 0.2< Re <2× 10^3. (Abdel-Aal et al., 2003)

$$C_d = 0.34 + \frac{3}{Re^{0.5}} + \frac{24}{Re} \tag{3.43}$$

where the Reynolds number is given by

$$Re = 0.0049\ \frac{\rho_g d_m u}{\mu_g} \tag{3.44}$$

The settling velocity is dependent on C_d and this is found by an iterative procedure.

a. Assume a value of C_d,
b. Calculate the velocity, u, from *equation 3.36*
c. Calculate Re from *equation 3.44*
d. Calculate C_d, from *equation 3.43* and compare to the assumed value
e. If no match is obtained, use the calculated value of C_d, and repeats steps b-d until convergence is obtained.

3.1.1.8 Liquid Capacity Constraint

Liquid within the separator has to be retained within the separator for a specific retention time, *t* for the separation process. The volume of the separator occupied by the oil, V_o, is obtained by the cross-sectional area by the height of the oil column, H (in.). *Equation 3.40* is therefore rewritten as

$$Q_o = 30.644 \left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2\left(\frac{H}{12}\right)\left(\frac{1}{t}\right)\left(\frac{m^3}{day}\right) \quad (3.45)$$

or

$$D^2H = 1.40355\ x\ 10^{-4}\ Q_o t \quad \mathrm{m}^3 \quad (3.46)$$

3.1.1.9 Sizing Procedure of Vertical Separator

The sizing procedure based on the above theory can be used to find the size of the vertical separator (diameter and seam-to-seam length or height).(Abdel-Aal et al., 2003)

1. *Equation 3.42* is used to determine the minimum allowable vessel diameter
2. For diameters larger than the minimum, *equation 3.46* is used to determine combinations of *D* and *H*.
3. The seam-to-seam length, L_s, for each combination of *D* and *H* is determined using one of the following expressions as appropriate:

   For D < 0.91m

   $$L_s = \frac{H+76}{39.3696} \quad \mathrm{m} \quad (3.47)$$

   For D > 0.91m

   $$L_s = \frac{H+D+40}{39.3696} \quad \mathrm{m} \quad (3.48)$$

4. For each combination of *D* and L_s, the *slenderness ratio*, *SR*, defined as the ratio of length to diameter is determined. Separators with SR between 3 and 4 are commonly selected.

3.1.1.10 Horizontal Gas – Liquid Separator Sizing

Similar to the vertical separator the size of the horizontal separator is determined by the gas and liquid capacity. For the horizontal separator the gas capacity constraint yields a relationship between the diameter and effective length of the separator. This along with a similar relationship derived from the liquid capacity constraint are used in determining the size of the separator. In reality, either the gas capacity constraint or the liquid capacity constraint governs the design and only one of the two constraints equations is used in determining the size. (Abdel-Aal et al., 2003)

The derivations below assume the gas and liquid phases each occupy 50% of the effective separator volume.

3.1.1.11 Gas Capacity Constraint

The average gas flowing velocity within the separator, u_g, is is obtained by dividing the volumetric flow rate, Q_g, by one-half of the separator cross-sectional area, A; that is,

$$u_g = \frac{Q_g}{0.5[(\pi/4)D^2]} \tag{3.49}$$

For Q_g, given in MMscmd the velocity is given as

$$u_g = 36.576\frac{Q_g}{D^2}\left(\frac{TZ}{P}\right) \quad \text{m/s} \tag{3.50}$$

The gas travels horizontally along the effective length of the separator, L (m), in a time, t_g, that is given by;

$$t_g = \frac{L}{u_g} \text{ s} \tag{3.51}$$

This time must, at least, be equal to the time it takes the smallest oil droplet, to be removed from the gas, to travel a distance of $(D/2)$ to reach the gas–oil interface. This settling time, t_s, is obtained by dividing the distance $(D/2)$ by the settling velocity (in *equation 3.36*);

$$t_s = \left(\frac{D}{2\,x12}\right)\left\{0.01186\left[\left(\frac{\rho_o-\rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2}\right\}^{-1} \text{ s} \tag{3.52}$$

Equating 3.51 and *3.52*, substituting u_g from *equation 3.50* and solving for the product LD, gives;

$$LD = 326.71\left(\frac{Q_gTZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o-\rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \text{ m cm} \tag{3.53}$$

Equation 3.53 provides a relationship between the vessel diameter and effective length that satisfies the gas capacity constraint. Any combination of D and L satisfying *equation 3.53* ensures that all oil droplets having diameter d_m and larger will settle out of the gas flowing at a rate of Q_g (in MMscmd) into the separator that is operating at a given pressure and temperature.

3.1.1.12 Liquid Capacity Constraint

The separator must have sufficient volume to retain liquid for a specified time. For a half full separator, the volume occupied by the liquid is given by

$$V_o = 0.01415 \left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2 L \ \text{m}^3 \tag{3.54}$$

Substituting *equation 3.40*, the below is obtained

$$D^2 L = 0.04044\ Q_o t \ \text{m}^3 \tag{3.55}$$

The above equation provides another relationship between D and L that satisfies the liquid capacity (retention) time constraint.

3.1.1.13 Sizing Procedure of Horizontal Separator

Based on operating conditions of pressure, temperature, gas and liquid flow rates, gas and liquid properties and oil retention time) the size of (diameter and seam-to-seam length) of a horizontal separator can be determined as follows:

1. Assume various values for the separator diameter, *D*.
2. For each assumed value of *D*, determine the effective length, L_g, that satisfies the gas capacity constraint from *equation 3.53* and calculate the seam-to-seam length, Ls, from

$$L_s = \frac{1}{3.2808}\left(L_g + \frac{D}{12}\right) \text{m} \tag{3.56}$$

3. For each assumed value of *D*, determine the effective length, L_o, that satisfies the liquid capacity constraint from *equation 3.55* and calculate the seam-to-seam length, Ls, from

$$L_s = \frac{4}{9.8424} L_o \ \text{m} \tag{3.57}$$

4. For each value of *D* used, compare the values of L_g and $L_o$to determine whether the gas capacity constraint or the oil capacity constraint governs the design of the separator. The larger required length governs the design.
5. Select reasonable combinations of *D* and *Ls* such that the slenderness ratio SR is in the range of 3–5. Following that the cost and availability would then determine the final selection. The cost and availability criteria for separator selection is not covered in this thesis.

For the API determination, the allowable gas velocity, v, is determined from the *Souders Brown* equation. This gives the apparent velocity in the space open to gas flow.

$$v = K_s \left[\frac{\rho_l - \rho_g}{\rho_g}\right]^{0.5} \tag{3.58}$$

where K_s depends on all factors that affect separation other than density – vortex action, foaming, pulsating flow, liquid flowing in heads, presence of solids, degree of separation needed, separation length, varying gas-liquid ratios etc. $K_s$values are determined from experience/field data and are dictated per API 12J standard. Examples are given in *Table 3.1* and *Appendix C.5.*

Table 3.1: Ks Values based on API 12J (Campbell & Maddox, 1999)

Type	**Height or Length**	**Typical Ks range**	
	[m (ft)]	*[ft/sec]*	*[m/sec]*
Vertical	1.52 (5)	0.12 to 0.24	0.037 - 0.073
Vertical	3.05 (10)	0.18 to 0.35	0.055 - 0.107
Horizontal	3.05 (10)	0.40 to 0.50	0.122 - 0.152
	Other lengths	$(L/10)^{0.56}$	$(L/3.05)^{0.56}$
Spherical	ALL	0.061 - 0.107	0.20 - 0.35

To determine the volume flow rate based on the cross-sectional area, F, of the separator, the actual gas flow is given by

$$q_a = (\pi/4)(d^2)(v)(F) \tag{3.59}$$

From *equation 3.59* the diameter of the separator can be determined.

Vessel Length
Knowing the liquid actual volume flow rate, q_l , the fraction of cross-section area occupied by liquid, F_l, and the residence time required for separation, t, the effective length or seam-to-seam length of the separator can be ascertained

$$L_e = \sqrt{\frac{4tq_l}{\pi D^2 F_l}} \tag{3.60}$$

From *equation 3.60* the actual length can be determined from the effective length and diameter.

$$L = L_e + D \tag{3.61}$$

Based on the fluid properties, *Appendix C* shows an example of the separator calculator developed for the separation process. *Chapter 5* gives a brief on the spreadsheet set up and its practical use.

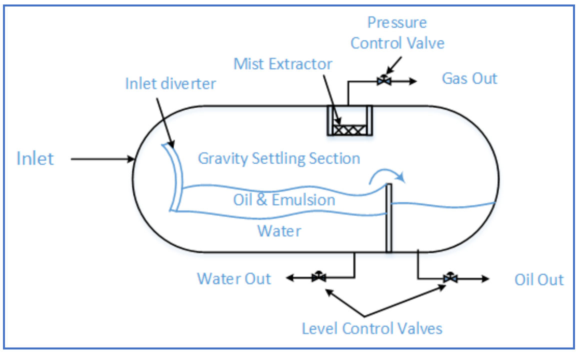
3.1.2 Three-Phase Separators

In most production operations, the produced fluid stream comprises three-phases: oil, gas and water. Water produced with the oil exists partly as free water and partly as water-in-oil emulsion. Free water produced with the oil will settle and separate from the oil by gravity. The emulsified water requires various methods of treatment including heat treatment, chemical treatment, electrostatic treatment, or a combination of these treatments in addition to gravity settling.

It is therefore advantageous to first separate the free water from the oil to minimize the treatment costs of the emulsion. Gas is mostly present along with the water and oil. If the volume of gas is small relative to the liquid, the separation of the water from oil will govern the design of the vessel. However, when the volume of the gas to be separated from the liquid is large then either the gas capacity requirement or the water–oil separation constraints govern the vessel design. (Abdel-Aal et al., 2003)

Three-phase separators are either horizontal or vertical. The design of three-phase separators differ from two-phase in that the design must incorporate separation and level control of two liquids.

Figure 3.3 and *Figure 3.4* highlight two main designs of horizontal three-phase separators and the vertical three-phase separator.



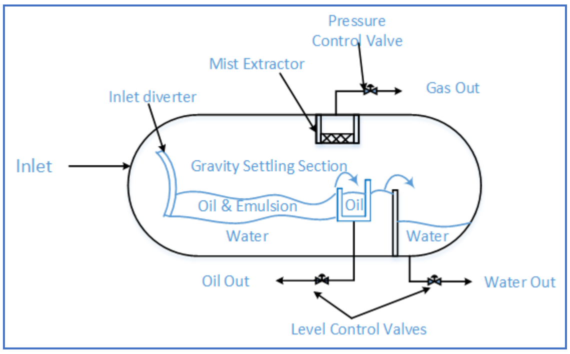


Figure 3.3 Three-phase horizontal separator - weir type (left) and bucket and weir type (right) (Abdel-Aal et al., 2003)

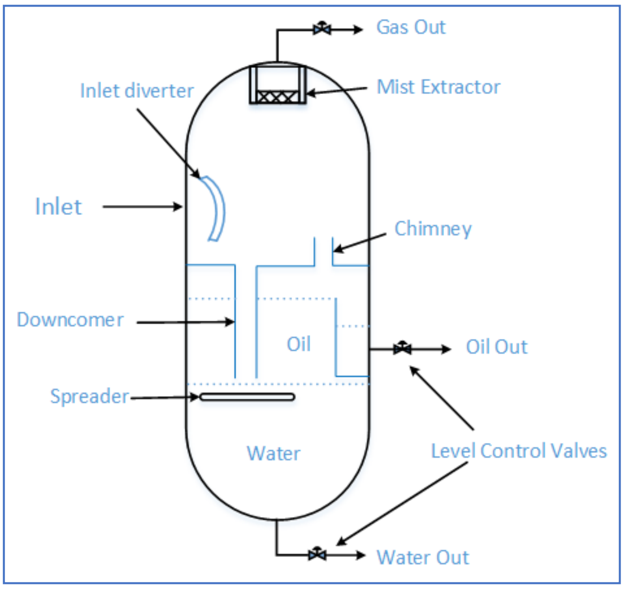


Figure 3.4: Three-phase vertical separator (Abdel-Aal et al., 2003)

The operation of the separator is similar to that of the two-phase separator. The produced fluid stream enters the separator and hits the inlet diverter, where the initial bulk separation of the gas and liquid takes place due to the change in momentum and difference in fluid densities.

The gas flows horizontally through the gravity settling section (the top part of the separator) where the entrained liquid droplets, down to a certain minimum size (normally 100 µm), are separated by gravity. The gas then flows through the mist extractor, where smaller entrained liquid droplets are separated, and out of the separator through the pressure control valve, which controls the operating pressure of the separator and maintains it at a constant value. The bulk of liquid, separated at the inlet diverter, flows downward, normally through a downcomer that directs the flow below the oil–water interface. (Abdel-Aal et al., 2003)

3.1.2.1 Theory behind Droplet Separation – 3-Phase Separators

For three-phase separator design the settling and separation of the oil droplets from water and of the water droplets from the oil must be considered in addition to the retention time constraint. This is unlike the two-phase separator where liquid retention time constraint is the only criterion used for determining the liquid capacity of two-phase separators.

With respect to separation of oil droplets from water, or water droplets from oil, a relative motion exists between the droplet and the surrounding continuous phase. An oil droplet, being smaller in density than the water, moves vertically upward under the gravitational or buoyant force, F_g. The continuous phase (water), on the other hand, exerts a drag force, F_d, on the oil droplet in the opposite direction. The oil droplet will accelerate until the fractional resistance of the fluid drag force, F_d, approaches and balances F_g after which the oil droplet reaches constant velocity or settling or terminal velocity.

Conversely, water droplets, are higher in density than the oil, hence move vertically downward under the gravitational or buoyant force, F_g. The continuous phase , on the other hand, exerts a drag force, F_d, on the water droplet in the opposite direction. The water droplet will accelerate until the frictional resistance of the fluid drag force, F_d, approaches and balances F_g; thereafter, the water droplet continues to rise at a constant velocity or settling or terminal velocity.

$$F_d = C_d \frac{\pi}{4} d^2 \frac{\rho_c u^2}{2} \tag{3.62}$$

whereas F_g is given by

$$F_g = \frac{\pi}{6} d^3 (\Delta\rho) \tag{3.63}$$

where d represents the diameter of the droplet, u is the settling velocity of the droplet (m/s), ρ_c ,is the density of the continuous phase (kg/m^3) , g is the gravitational acceleration (m/s) and C_d is the drag co-efficient. For low Reynolds number, Re, drag co-efficient is given by

$$C_d = \frac{24}{Re} = \frac{24\,\mu' g}{\rho d u} \tag{3.64}$$

where μ is the viscosity of the continuous phase (kg-s/m^2)
Substituting *equation 3.64* into *3.62* yields

$$F_d = 3\pi\mu' d u \tag{3.65}$$

The terminal velocity, u_s, is reached when $F_d = F_g$. therefore equating 3.63 and 3.65 gives

$$u_s = \frac{(\Delta\rho) d^2}{18\mu'} \tag{3.66}$$

The typical units for droplet diameter are in micrometers and viscosity in centipoise. Representing the diameter by d_m and viscosity by μ in the equation becomes

$$u = 8.729 \times 10^{-9} \frac{(\Delta\rho) d_m^2}{\mu} \text{ m/s} \quad \text{or} \tag{3.67}$$

$$u = 5.447 \times 10^{-7} \frac{(\Delta\gamma) d_m^2}{\mu} \text{ m/s} \tag{3.68}$$

where $\Delta\gamma = \gamma_w - \gamma_o$, which is the specific gravity of oil and water respectively.

From *Equation 3.67* and/or *3.68* the droplet settling velocity is inversely proportional to the viscosity of the continuous phase. Oil viscosity is several magnitudes higher than the water viscosity. Therefore, the settling velocity of water droplets in oil is much smaller than the settling velocity of oil droplets in water. The time needed for a droplet to settle out of one continuous phase and reach the interface between the two phases depends on the settling velocity and the distance travelled by the droplet. In operations where the thickness of the oil pad is larger than the thickness of the water layer, water droplets would travel a longer distance to reach the water–oil interface than that travelled by the oil droplets. This, combined with the much slower settling velocity of the water droplets, makes the time needed for separation of water from oil longer than the time needed for separation of oil from water. Hence, the separation of the water droplets from the continuous oil phase is always taken as the design criterion for three-phase separators.

The minimum size of the water droplet or the minimum size of the oil droplet that must be removed from the continuous phase (either oil or water) depends on the operating conditions and fluid properties. Data for this can be obtained from simulations of field data or offset fields. In the absence of such data the minimum water droplet size to be removed from the oil is taken as 500µm.

The required liquid volumes within the separator is determined by the retention time. The oil phase needs to be retained within the separator for a period of time that is sufficient for the oil to reach equilibrium and liberates the dissolved gas. The retention time should also be sufficient for appreciable coalescence of the water droplets suspended in the oil to promote effective settling and separation. Similarly, the water phase needs to be retained within the separator for a period of time that is sufficient for coalescence of the suspended oil droplets. This data can be obtained from laboratory test; however in the absence of such data it is common practice to use a retention time of 10 minutes for both oil and water. (Abdel-Aal et al., 2003)

3.1.2.2 Water Droplet settling constraint

Similar to the two-phase separator sizing criteria, the three-phase separator requires consideration of the gas capacity constraint, liquid retention time constraint as well as a the settling of water droplets in oil which gives the maximum diameter of the separator.

The additional constraint in the design of three-phase horizontal separators is that the oil retention time should be sufficient for the water droplets of certain minimum size to settle out of the oil. A conservative assumption is to take the water droplet to travel from the top of the oil pad. Hence the water droplet would have to travel a distance equal to the oil pad thickness, H_o, at a velocity (determined from *equation 3.68*). This gives;

$$t_{wd} = \left(\frac{1}{60}\right)\frac{(H_o/12)}{1.787\times 10^{-6}(\Delta\gamma)d_m^2/\mu_o} \text{ min} \tag{3.69}$$

To obtain the maximum allowable oil pad thickness and *equating 3.69* to the oil retention time, t_o, this gives:

$$H_{o,max} = \frac{3.2512 \times 10^{-3}(\Delta\gamma)d_m^2}{\mu_o} \text{ cm} \tag{3.70}$$

d_m is assumed to be 500µm in the absence of laboratory data.

For a separator half full of liquid the relation is

$$\frac{A}{A_w} = \left(\frac{1}{\pi}\right)\left[cos^{-1}\left(\frac{2H_o}{D}\right) - \left(\frac{2H_o}{D}\right)\left(1 - \frac{4H^2}{D^2}\right)^{-0.5}\right] \tag{3.71}$$

where A_w and A are the cross-sectional area of the separator occupied by water and the total cross-sectional area of the separator, respectively.

For a separator half full of liquid the total cross-sectional area of the separator, A, is equal to twice the area occupied by the liquid, which is equal to the area occupied by water, A_w, and the area occupied by oil, A_o, given as $A = 2\ (A_o + A_w)$. This gives the below relation

$$\frac{A}{A_w} = 0.5\ \frac{Q_w t_w}{Q_o t_o + Q_w t_w} \tag{3.72}$$

Upon determination of ratio A/A_w; the ratio of the oil pad height to the diameter H_o/D can be derived from equation 3.71. Knowing H_{omax}, and H_o/D the maximum diameter of the separator is obtained

$$D_{max} = \frac{H_{omax}}{H_o/D} \tag{3.73}$$

This gives the upper limit of the separator diameter. Different equations could be derived as opposed to the assumption made of 50% occupied by the different phases.

Gas Capacity Constraint

As with the two-phase separator the gas capacity constraint also holds with the three-phase separator. This gives a relationship between the separator diameter and the effective length where d_m is normally taken as 100µm;

$$LD = 326.71\ \left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o - \rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \text{m cm} \tag{3.74}$$

With diameters less than the maximum diameter from the water droplet settling constraint, *equation 3.74* is used to determine possible diameter and length combinations that meet the *gas capacity constraint*.

Retention Time Constraint

The space occupied by the oil and water should allow for sufficient retention time for separation. Since half of the liquid phase (both oil and water) occupy half of the separator volume; the diameter and effective length is given by

$$V_l = 0.0141584 \left(\frac{\pi}{4}\right)\left(\frac{D}{12}\right)^2 L \quad \text{m}^3$$

where 1 barrel = 0.15898 m^3 ; this gives:

$$V_l = 7.2959 \times 10^{-5} D^2 L \quad \text{m}^3 \tag{3.75}$$

The volume of separator occupied by oil, V_o, is the product of the oil flow rate, Q_o, and the oil retention time, t_o. For Q_o in cubic meters per day and t_o in minutes, gives

$$V_o = 1.104 \times 10^{-4}\, t_o Q_o \ \text{m}^3 \tag{3.76}$$

Similarly, the volume of the separator occupied by water, is the product of the water flow rate and the water retention time.

$$V_w = 1.104 \times 10^{-4}\, t_w Q_w \ \text{m}^3 \tag{3.77}$$

Since $V_l = V_o + V_w$; this gives

$$D^2 L = 2.8101\,(Q_o t_o + Q_w t_w) \quad \text{cm}^2\,\text{m} \tag{3.78}$$

Selecting diameters smaller than the maximum diameter determined from *equation 3.73;* with combinations of diameter and length are obtained to satisfy the retention time constraint.

3.1.2.3 Sizing Procedure for three-phase horizontal separator

The procedure for determining the diameter and length of a three-phase horizontal separator can therefore be summarised as:

1. Determine the value of A/A_w from *equation 3.72*.
2. From *equation 3.71* determine the value of H_o/D for the calculated A/A_w
3. Determine the maximum oil pad thickness, H_{omax} from *equation 3.70* assuming d_m equal to 500μm
4. Determine D_{max} from *equation 3.73*
5. For diameters smaller than D_{max}, determine the combinations of D and L that satisfy the gas capacity constraint from *equation 3.74*, substituting 100 μm for d_m .
6. For diameters smaller than D_{max}, determine the combinations of D and L that satisfy the retention time constraint from *equation 3.78.*
7. Compare the results obtained in steps 5 and 6 and determine whether the gas capacity or retention time (liquid capacity) governs the separator design.
8. If the gas capacity governs the design, determine the seam-to-seam length of the separator, L_s, from

$$L_s = \frac{1}{3.2808}\left(L + \frac{D}{12}\right) \quad (3.79)$$

If the liquid retention time (liquid capacity) governs the design, determine L_s from

$$L_s = 4\frac{L}{9.8424} \quad (3.80)$$

9. A reasonable diameter and length with a slenderness ratio (L_s/D) in the range of 3–5 is recommended. In some cases, the slenderness ratio might be different from the range of 3–5. In such cases, especially when the slenderness ratio is larger than 5, internal baffles can be installed to act as wave breakers in order to stabilize the gas–liquid interface.

For the stepwise sizing procedure using the API method, by using predetermined K_s values, (Svrcek & Monnery, 1993) reference *Chapter 5. Appendix C.9* presents for this method, L/D ratio guidelines in determining the optimum design.

3.1.2.4 Sizing Equations for Vertical Separators

Sizing of a vertical three-phase separator is done in a similar manner to sizing vertical two-phase separators where the gas capacity constraint is used to determine the minimum diameter of the vessel and the liquid retention time constraint is used to determine the height of the vessel. For three-phase separators, however, a third constraint is added. This is the requirement to settle water droplets of a certain minimum size out of the oil pad. This results in a second value for the minimum diameter of the separator. Therefore, in selecting the diameter of the vessel, the larger of the minimum diameters determined from the gas capacity constraint and water settling constraint is considered as the minimum acceptable vessel diameter.

3.1.2.5 Water Droplets Settling Constraint

The condition for the settling and separation of water droplets from the oil is established by equating the average upward velocity of the oil phase, u_o, to the downward settling velocity of the water droplets of a given size, u_w. The average velocity of the oil is obtained by dividing the oil flow rate by the cross-sectional area of flow;

$$u_o = 3.627 \times 10^{-3} \frac{Q_o}{D^2} \text{ m/s} \quad (3.81)$$

And the water droplet settling velocity:

$$u_w = 5.446 \times 10^{-7} \frac{(\Delta\gamma)d_m^2}{\mu_o} \text{ m/s} \quad (3.82)$$

For water droplets to settle out of the oil, u_w must be larger than u_o. Equating u_w to $u_o$would result, therefore, in determining the minimum diameter of the separator, D_{min}, that satisfies the water settling constraint. This results from *equation 3.81* and *3.82* gives;

$$D_{min}^2 = 4.3135 \frac{Q_o\mu_o}{(\Delta\gamma)\, d_m^2} \text{ m}^2 \quad (3.83)$$

Diameters larger than the D_{min} will yield a lower average oil velocity and ensure water separation.

3.1.2.6 Gas Capacity Constraint

The gas capacity constraint for a vertical separator yields an expression for the minimum vessel diameter

$$D_{min}^2 = 3.263\ Q_g \left(\frac{TZ}{P}\right)\left(\frac{\rho_g}{\rho_o - \rho_g}\frac{C_d}{d_m}\right)^{1/2} \text{m}^2 \qquad (3.84)$$

Diameters larger than the D_{min} will yield a lower gas velocity and ensure separation of liquid droplets of diameters equal to and larger than d_m out of the gas.

3.1.2.7 Liquid Retention Time (Capacity Constraint)

The separator volume must be sufficient to afford retention time to allow separation of entrained water droplets from the oil, separation of the entrained oil droplets from the water, and for the oil to reach equilibrium with the gas. Retention times are ideally determined from laboratory tests and range from 3-30 minutes depending on fluid properties and operating conditions.

The calculator assumes a retention time of 10 minutes to be used for both oil and water.

The volume (given in m^3) of each phase within the separator is given by;

$$V_o = 0.0283\left(\frac{1}{12}\right)^3\left(\frac{\pi}{4}\right)D^2 H_o \qquad (3.85)$$

and

$$V_w = 0.0283\left(\frac{1}{12}\right)^3\left(\frac{\pi}{4}\right)D^2 H_w \qquad (3.86)$$

Hence,

$$V_o + V_w = 1.286 \times 10^{-5}\ D^2(H_o + H_w) \qquad (3.87)$$

The volume (in m^3) is also calculated from the volumetric flow rate and the retention time (in minutes)

$$V_o = Q_o \frac{0.1589}{24x60}\frac{m^3}{min} \times t_o$$

$$V_w = Q_w \frac{0.1589}{24x60}\frac{m^3}{min} \times t_w$$

Then

$$V_o + V_w = 1.1035\ x\ 10^{-4}\ (Q_o t_o + Q_w t_w) \qquad (3.88)$$

Equating *3.87* and *3.88*; we obtain

$$D^2(H_o + H_w) = 8.576\ (Q_o t_o + Q_w t_w)\ \mathrm{m}^3 \quad (3.89)$$

3.1.2.8 Sizing Procedure for a three-phase vertical separator

The procedure for determining the diameter and seam-to-seam length of a three-phase vertical separator can therefore be summarised as:

1. Determine the minimum diameter that satisfies the water droplets settling constraint from *equation 3.83*.
2. Determine the minimum diameter that satisfies the gas capacity constraint from *equation 3.84.*
3. The larger of the two minimum diameters determined in steps 1 and 2 is then considered as the minimum allowable vessel diameter.
4. For various values of diameter larger than the minimum allowable vessel diameter, use *equation 3.89* to determine combinations of diameters and liquid heights.
5. For each combination, determine the seam-to-seam length (in metres) from the following:

   For $D > 0.914$ m

$$L_s = \frac{1}{39.3696}\ (H_o + H_w + D + 40) \quad (3.90)$$

   For $D < 0.914$ m

$$L_s = \frac{1}{39.3696}\ (H_o + H_w + 76) \quad (3.91)$$

It was assumed in the research a length-to-diameter ratio of 1.5 to 6.0 for the 2-phase horizontal and 3-phase horizontal separators. This was assumed as the optimum target for the equipment design.

3.1.3 Mechanical Design (Wall thickness and Weight)

The total weight of each separator assembly includes the weight of the empty vessel, the weight of the internals, and the skid weight. In addition, the associated piping also contributes to the weight of the equipment or unit and must be taken into consideration.

The weight of an empty vessel, W_b, (mass per unit length, given in kg/m, including heads is) is given by;

$$W_b = 3.47\ dt \quad (3.92)$$

where, d, is the internal diameter (cm) and, t, is the wall thickness (cm).

The wall thickness is estimated according to ASME standards. This is classified under Division 1 or Division 2. The wall thickness is a function of the diameter, d, operating pressure P and the maximum allowable stress, SE . The wall thickness, t (in cm), is calculated based on *equation 3.93*. The maximum allowable stress, S depends on the material and grade and the division code for the application. The reference standards and allowable stresses for determining the wall thickness is given under *Appendix G* and *Appendix H.*

$$t = 2.54\, x \frac{Pd}{2SE - 0.2P} \tag{3.93}$$

The weight of the empty vessel, W_v, is the sum of the weight of the internals, W_I, the weight of the external nozzles, W_N, and L is the seam-to-seam length of the separator. The internal and nozzle weights are determined from correlations given under *Appendix C.6.*

$$W_v = W_b L + W_I + W_N \quad \text{kg/m} \tag{3.94}$$

For skidded equipment the following factors are used for the weight of the piping, W_p, weight of skid steel, W_s, weight of electrical & instrument and the weight of the total skid, W_{skid}, is given by *equations 3.95*

$$\begin{gathered} W_p = 40\,\%\, W_v \\ W_s = 10\,\%\, W_v \\ W_E = 80\,\%\, W_v \\ W_{skid} = W_v + W_P + W_E + W_s \end{gathered} \tag{3.95}$$

3.1.4 Equipment Footprint

The footprint of the separators are calculated based on assumptions of preliminary estimates of the skid dimensions. These differ for horizontal vessels and vertical vessels and give an initial approximation of the space occupied by the process equipment.

Table 3.2: Skidded equipment footprint relations

	Horizontal Vessels	**Vertical Vessels**
Skid Width	I.D. x 2	I.D. x 2
Skid Length	Seam-to-Seam length x 1.5	I.D. x 2.5
Skid Height	I.D. x 2 +1 meter	Seam to Seam length x 1.5

3.2 Heat Exchanger

Heat exchangers are fundamental in the gas processing system. They are utilised to optimise the processing system in terms of energy utilisation and area considerations which in effect significantly impact cost.

This work focuses on the sizing criteria for the design of heat exchangers based on the model scenario and considers simple guidelines and rules of thumb for heat exchanger selection. The heat exchangers looked at within the research are *Shell and Tube Heat Exchangers*. For this master thesis, heat exchangers used in offshore gas processing focus on gas cooling by sea water. The function of the heat exchanger is to provide the medium for cooling and separation of the heavy hydrocarbon components. This process is done to ensure the cricondenbar and dew point requirements are met for rich gas transport.

3.2.1 Heat Exchanger Design

Many factors are considered in the design and selection of heat exchangers. These would include basic process-design variables and other factors such as temperature strains, thickness of tubes and shell, types of baffles, tube pitch, and standard tube lengths.
The design and manufacture of heat exchangers is given by standards provided by the Tubular Exchanger Manufacturers Association (TEMA). These standards identify heat exchanger size and type by designated numbers and letters.(TEMA, 1988)

3.2.1.1 Shell and Tube Heat Exchanger

For the purpose of this master thesis, for simplicity, consistency and for accurate comparison, all heat exchangers are assumed to be of the *single-pass shell and tube type*. In the design of heat exchangers, the amount of heat transfer must be determined and is given by the below equations for heat balance with no phase change of the gas and sea water.

$$Q = \dot{m}C_p(T_1 - T_2) \tag{3.96}$$

From *equation 3.96*, the duty of the heat exchanger could be determined from a heat balance given by *equation 3.97*. This assumes no phase change in any of the fluids.

$$Q = \dot{m}C_{pc}\left(T_{c,o} - T_{c,i}\right) = \dot{m}C_{p,h}\left(T_{h,i} - T_{h,o}\right) \tag{3.97}$$

given;

Q - Heat Transfer, W
$\dot{m}$ - Mass flowrate, kg/s
C_p - Heat Capacity of the cold or hot streams, J/kg-K
T - Temperature of inlet or outllet hot stream or cold stream, K

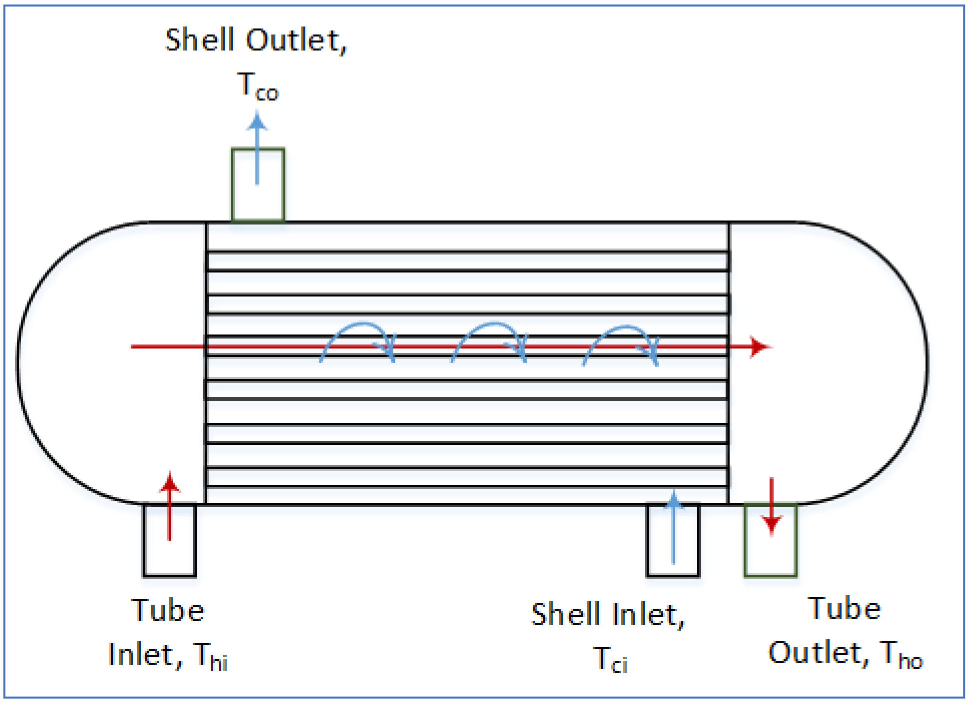


Figure 3.5: Shell and tube heat exchanger (Counter-current flow)

Equation 3.98 refers to the heat transfer utilising the overall heat transfer coefficients, U, total surface area, A, and Logarithmic Mean Temperature Difference, $LMTD$, for single pass design. The heat transfer area generally refers to the effective outside bare surface area of the tubes, and the overall heat transfer co-efficient must also be based on this area.

$$Q = UA(LMTD).F \tag{3.98}$$

The local temperature difference between the hot stream and the cooling stream (sea water) will not have a constant value throughout a heat exchanger, and so an effective average value must be used in the rate equation. The appropriate average depends on the configuration of the exchanger. For simple counter-current and co-current exchangers the Log Mean Temperature Difference (LMTD) applies as represented in *Figure 3.6*; where GTTD refers to Greatest Terminal Temperature Difference and LTTD refers to Least Terminal Temperature Difference.

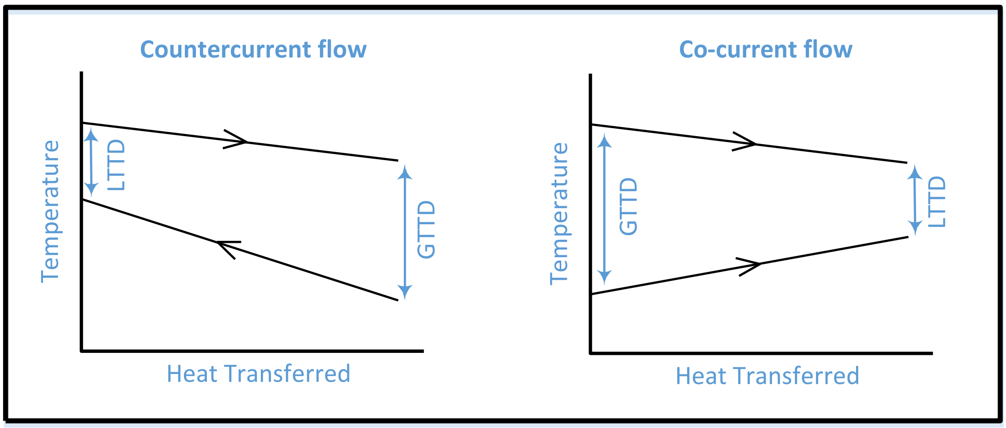


Figure 3.6: Counter-current flow and co-current flow

The factor F, refers to correction factor with exchanger configurations with flow passages being either partially countercurrent or co-current. The magnitude of the factor depends on exchanger configuration and stream temperatures as given in *Appendix D.1* (Gas Processors Suppliers Association (U.S.), 2012)

The logarithmic temperature difference is obtained from *equation 3.99*

$$LMTD = \frac{\Delta T_1 - \Delta T_2}{ln\frac{\Delta T_1}{\Delta T_2}} \tag{3.99}$$

Shell and tube exchanger nomenclature are characterised by the front end, shell type and rear end head type as depicted in *Figure 3.7*. Within HYSYS provides a default *AEL* configuration for the shell and tube heat exchanger.

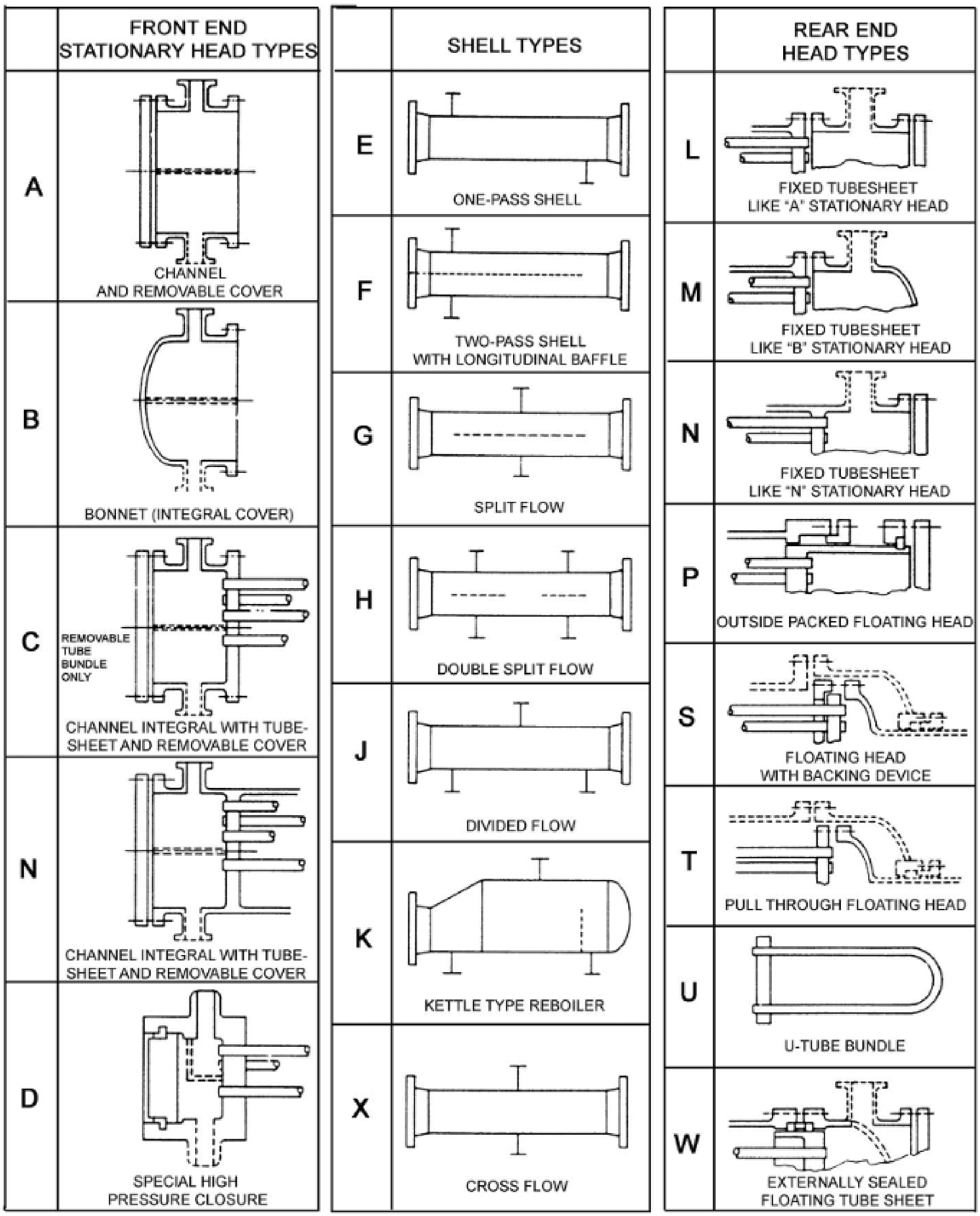


Figure 3.7: Shell and tube exchanger nomenclature courtesy of TEMA - (TEMA, 1988)

In sizing the heat exchanger, the heat transfer area is required which is derived knowing;

- Film heat transfer co-efficient
- Fouling factor
- LMTD (based on hot and cold stream temperatures)
- Duty of the heat exchanger
- Overall heat transfer co-efficient

Heat is transferred from a hot fluid to a cold fluid through the tube walls by the flowing process: *convection to the hot fluid wall*, *conduction through the wall* and *subsequent convection from the wall to the cold fluid*. Over a period with the heat transfer process; there is the formation and accumulation of scale and rust, deposits from the fluid, chemical reaction products between the fluid and wall material, and/or biological growth. This fouling has a low thermal conductivity and can increase the thermal resistance to heat flow from the hot fluid to the cold fluid. This thermal resistance of individual fluids is taken into account by a *fouling factor*, $R_f = 1/h$ (with units m^2.K/W); where h is the film transfer co-efficient.

For the purpose of the thesis and to obtain an approximate sizing of the heat exchanger, some pre-design criteria have been selected based on Heat Exchanger Design Handbook (Hewitt, 2002). The overall heat transfer co-efficient for unfinned tubular heat exchangers is found by utilising the empirical factors given by *equation 3.100*. (Shah & Sekuliâc, 2003)

$$\frac{1}{U} = \left(\frac{1}{h_{gas}} + R_{f,gas}\right)\frac{d_o}{d_i} + \frac{d_o ln(d_o/d_i)}{k_w} + \left(\frac{1}{h_{seawater}} + R_{f,seawater}\right) \quad (3.100)$$

where;

h	- film transfer co-efficient
R_f	- fouling resistance of gas (gas in tube, seawater on the shell side)
d_o , d_i	- outer and inner diameters respectively of the tube.

With reference to the Heat Exchanger Design Handbook (reference *Appendix D.3*) the parameters below are assumed. (Hewitt, 2002)

For seawater, the factors for calculating the overall heat transfer coefficient are given as;

h	5000 – 7500 W/m^{2}K
R_f	$10^{-4} to\ 2.5 \times 10^{-4}$ m^{2}K/W

Within the calculator, $h_{seawater}$ is assumed to be 5000 W/m^{2}K , $R_f - 1.5 \times 10^{-4}$ m^{2}K/W and the gas parameters given as:

h	250 – 400 W/m^{2}K (1MPa) 500 – 800 W/m^{2}K (10MPa)
R_f	$0 - 10^{-4}$ m^{2}K/W

Within the calculator h_{gas} is assumed to be 500 W/m^{2}K , $R_f - 10^{-4}$ m^{2}K/W.
Based on the assumed factors, the overall heat transfer co-efficient, U, is given as;

$$\frac{1}{U} = \left(\frac{1}{500} + 10^{-4}\right) \times 1.0 + 0 + \left(\frac{1}{5000} + 1.5 \times 10^{-4}\right) \quad (3.101)$$

$$U = 408.16 \sim 400\ W/m^2K \quad (3.102)$$

The area of the heat exchanger can be determined from *equation 3.98* knowing the heat transfer, overall heat transfer co-efficient and corrected LMTD.

Tube Side Parameters

The tube side specification can be determined from TEMA standards (reference *Appendix D.2)* where the outside parameters and the thickness of the tube are indicated and a standard length can be selected. (Gas Processors Suppliers Association (U.S.), 2012). The number of tube passes depending on the heat exchanger configuration is also specified (This has been assumed to be a single pass for the purpose of comparison within this master thesis)

The total length of tubes is determined knowing the total heat transfer area and the area of one standard tube. The diameter of the tube bundle is determined knowing the tube pattern; be it triangular or square. The *triangular tube configuration* is assumed as in *Figure 3.8*; where P_t is the tube pitch and d is the diameter of the tube. The pitch ratio given by *equation 3.103* has been assumed as **1.25**; this is normally the recommended ratio unless process requirements dictate otherwise. (Sinnott, Coulson, & Richardson, 2005)

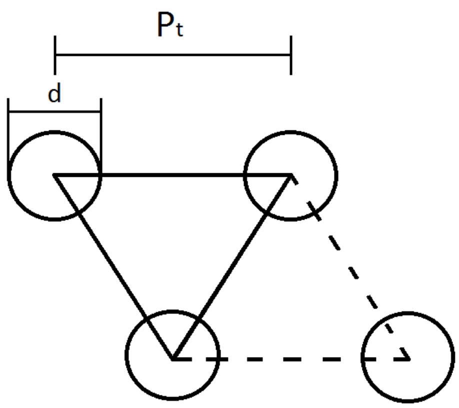


Figure 3.8: Triangular tube bundle configuration

$$\frac{P_t}{D} = 1.25 \tag{3.103}$$

For the triangular pitch, the diameter of the whole tube bundle is found from

$$D_{tight} = 2\left(\frac{N_T Area_{tube}}{\pi}\right)^{0.5} \tag{3.104}$$

Where

$$Area_{tube,\ triangular} = 2\,(PRd_o)^2\frac{\sqrt{3}}{4} \tag{3.105}$$

The corrected area may be calculated from *equation 3.106* for a tube pass greater than 1 (where n_p represents the number of tube passes in the shell) the cross sectional area can be added to account for the pass partition by multiplying the tube diameter by D_{tight}.

$$A_{corrected} = D_{tight}\, d_o(n_p - 1) + (N_T Area_{tube}) \tag{3.106}$$

Shell Side Parameters

The shell side minimum diameter is related to the number of tubes, tube passes, tube diameter, tube pitch, tube pitch layout (as indicated above under Tube design parameters) and tube omissions to allow space for impingement baffles or to decrease the number of tubes in the baffle windows. This shell side minimum diameter is given by *equation 3.107* where two tube diameters are added to the corrected area for tube passes.

$$D_{s,min} = 2\left(\frac{A_{corrected}}{\pi}\right)^{0.5} + 2d_o \tag{3.107}$$

Baffle Spacing

Heat exchangers are designed with baffles to divert the flow across the bundle to obtain a higher heat transfer co-efficient and also to give the tubes structural rigidity, preventing tube vibration and sagging. The number of baffles for a heat exchanger must be determined as these add to the weight of the heat exchanger. *Figure 3.9* depicts the baffle spacing and cut window in relation to shell.

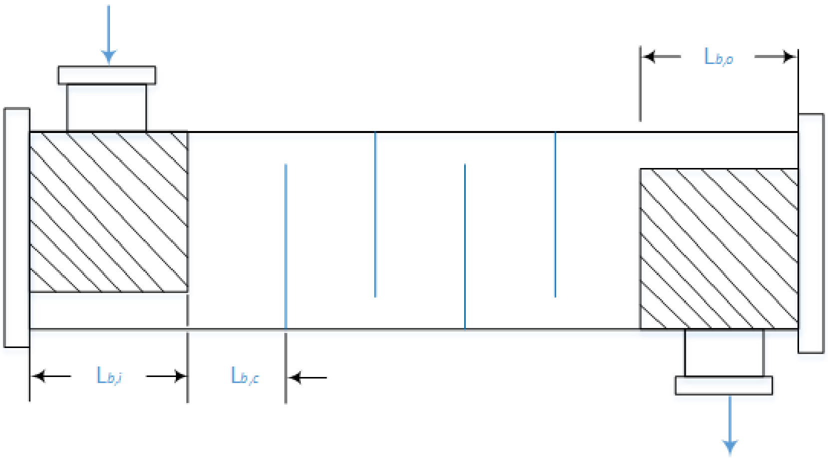


Figure 3.9 Heat exchanger baffle spacing- showing cut windows and entrance and exit sections

The number of baffles, N_b, within the heat exchanger is determined from *equation 3.108* knowing the length of the shell L, the central baffle spacing $L_{b,c}$ and the baffle spacings in the inlet and outlet regions $L_{b,i}$ and $L_{b,o}$ respectively. (Shah & Sekuliâc, 2003)

$$N_b = \frac{L - L_{b,i} - L_{b,o}}{L_{b,c}} + 1 \tag{3.108}$$

3.2.2 Mechanical Design (Wall thickness and Weight) and Footprint

The weight and footprint of the shell and tube heat exchanger is calculated based on assumptions of the empty shell weight as is done with the separator. The weight of the internals includes the weight of the tubes, the weight of the baffles as well as the nozzles.
The weight of the empty shell is given from *equation 3.109* as;

$$W_{empty\ shell} = Shell\ Volume\ \times\ Density\ of\ Steel \tag{3.109}$$

The total baffle weight is given as;

$$W_{baffle\ weight} = N_b\ (1 - Baffle\ Window\ height)\ \times \pi\ \frac{(Baffle\ Clearance\ \times ID_{shell})^2}{4} \times wall\ thickness\ \times Density\ of\ Steel \tag{3.110}$$

In addition to the empty vessel weight, the tube weights are estimated from tube weight per meter given in *Appendix D.2*, the length and total number of tubes. This from *equation 3.111* gives the total weight of the tubes.

$$W_{tubes} = Tube\ Weight\ per\ meter\ x\ Tube\ Length\ x\ N\ total\ tubes \tag{3.111}$$

For skidded equipment as is the case with separators, *equations 3.95* are used for the weight of the piping, W_p, weight of skid steel, W_s, weight of electrical & instrument and the weight of the total skid, W_{skid}.

3.2.3 Equipment Footprint

The footprint of the shell and tube heat exchanger skid is determined in the same manner as the separators i.e. from correlations highlighted in *Table 3.2*. This gives a preliminary estimation of the designed heat exchanger.

	Horizontal Vessels
Skid Width	I.D. x 2
Skid Length	Seam to Seam length x 1.5
Skid Height	I.D. x 2 +1 meter

3.3 Compressor

Within the offshore processing platform, compression of the gas is performed to transport the fluid in the gaseous phase to reach rich gas transport specification. The most commonly used compressors in offshore platforms are centrifugal compressors. They offer a high power to weight ratio and are manufactured in three configurations: *overhung impeller*, *horizontally split* or *vertically split* (barrel type).

Overhung impellers are commonly used in single stage service where the impeller is usually open, backward-bladed. Horizontally split cases are used in applications of high volume and lower pressure where the casings are split horizontally at the mid-section and repair and inspections are performed by removing the top half. The vertically split or barrel type compressors are used in high pressure and low volume applications and are maintained by removing the compressor barrel from the end of the compressor. More space is required in this case to facilitate removal, however, can be repaired more quickly with a spare barrel than horizontally split compressors. (Campbell & Maddox, 1999).

Compressor design is manufacturer specific and performance is based on in-house design techniques in improving the efficiency. The compressor is characterised by performance parameters specifically "*head*" and the theoretical head may be calculated using the isentropic or polytropic approach. The head is the amount of work per unit mass. In determining the size of the compressor which is manufacturer specific; the head requirement must be determined as well as the efficiency and the power.

This project focuses on centrifugal compressors as these are commonly used in offshore gas processing. *Figure 3.10* shows the coverage of centrifugal compressors in specific range of applications based on discharge pressure and inlet flow. For the purpose of sizing the compressor, data from a specific manufacturer (Elliott Group) has been obtained to obtain an approximate sizing. This is captured under *Appendix E.1*.

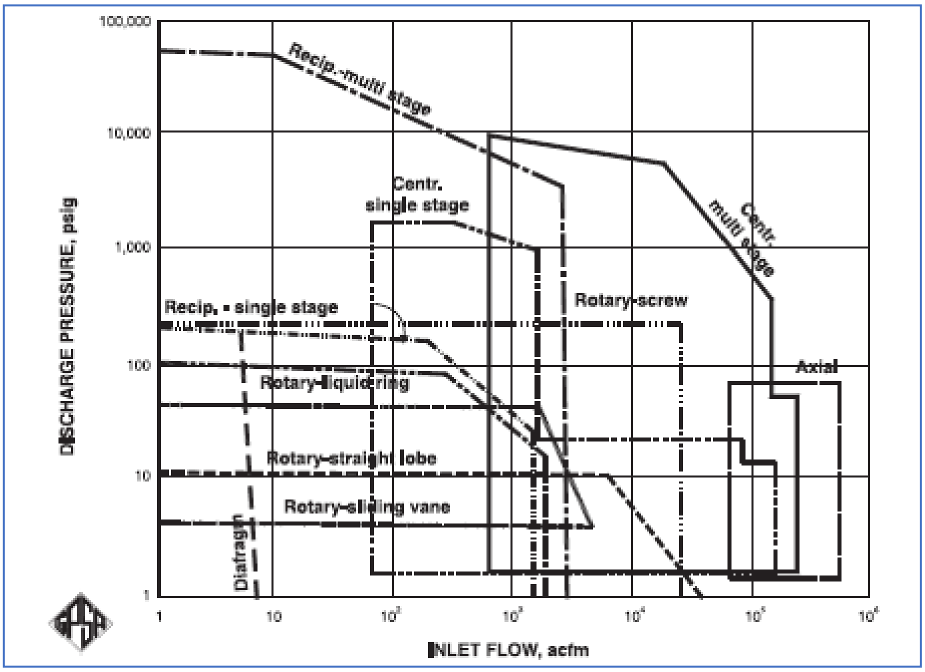


Figure 3.10: Compressor Coverage Chart (Gas Processors Suppliers Association (U.S.), 2012)

3.3.1 *Performance calculations*

The actual compression process follows a compression path given by the compressor efficiency, either isentropic or polytropic. *Figure 3.11* shows the isentropic compression process (1-2s) and polytropic compression process (1-2) which relates to infinite small isentropic compression steps along the actual compression path given by the compressor efficiency.

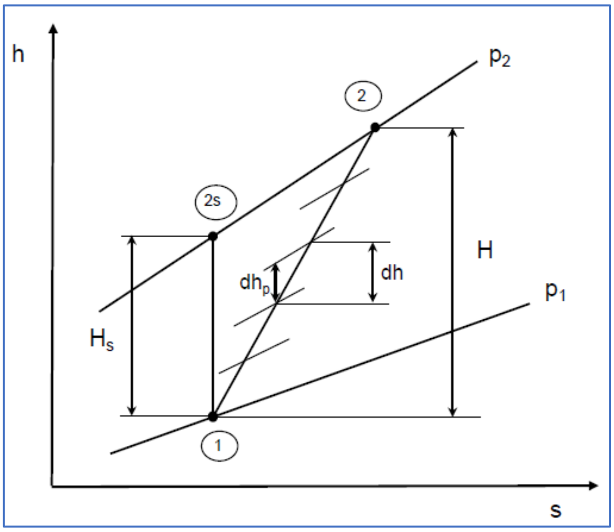


Figure 3.11: Compression process (Bakken, 2017)

The generalised polytropic process, at constant efficiency (polytropic) is defined as;

$$pv^n = constant \tag{3.112}$$

where

$$\frac{n-1}{n} = \frac{\kappa - 1}{\kappa \eta_p} \tag{3.113}$$

Given;

n - Polytropic exponent
κ - Isentropic exponent
η_p - Polytropic efficiency
p - Pressure
v - Volume

For real gas behaviour, the proper performance calculation has to distinguish between polytropic temperature exponent (n_T) and polytropic volume exponent (n_V) which takes into account the real gas behaviour when utilising the basic pressure-temperature and pressure-volume relationships.(Bakken, 2017)

The polytropic exponent varies through the compression process. This makes an exact analytical solution of the polytropic head, H_p, challenging. The exponent is assumed constant when solving for the polytropic head equation. This gives an approximate solution of the integral.

$$H_p = \int_1^2 vdp \approx \frac{n_v}{n_v - 1} [p_2 v_2 - p_2 v_2] \tag{3.114}$$

From the real gas equation and *equation 3.112,* the polytropic head becomes

$$H_p \approx f \frac{n_v}{n_v - 1} \frac{Z_1 R_o T_1}{MW} \left[\left(\frac{p_2}{p_1} \right)^{\frac{n_v - 1}{n_v}} - 1 \right] \tag{3.115}$$

where Z is the compressibility factor, f takes into account the change in polytropic volume exponent n_v along the compression path and is given by equation;

$$f = \frac{h_{2s} - h_1}{\frac{\kappa_v}{\kappa_v - 1} [p_2 v_{2s} - p_1 v_1]} \tag{3.116}$$

At given suction and discharge conditions the polytropic volume exponent is given by *equation 3.117.*

$$n_v = \frac{ln \left(\frac{p_2}{p_1} \right)}{ln \left(\frac{v_1}{v_2} \right)} \tag{3.117}$$

The polytropic efficiency is given by *equation 3.118*. The polytropic efficiency is normally used by vendors when quoting compressor performance as this is essentially independent of compression ratio and gas composition and is determined from compressor tests. (Campbell & Maddox, 1999)

$$\begin{aligned} \eta_p &= f \frac{n_v}{n_v - 1} \frac{(p_2 v_2 - p_1 v_1)}{(h_2 - h_1)} \\ &= f \frac{n_v}{n_v - 1} \frac{Z_1 R_o T_1}{MW (h_2 - h_1)} \left[\left(\frac{p_2}{p_1} \right)^{\frac{n_v - 1}{n_v}} - 1 \right] \end{aligned} \tag{3.118}$$

3.3.2 Total Compressor Head and Power

The total compressor head is derived from the polytropic head and the polytropic efficiency of the compressor. Based on experience from evaluation of compressor performance and from predictions from compressor vendors and process simulation systems; large deviations in isentropic and polytropic exponents as well as polytropic head and efficiency are obtained. This is largely due to different equations of state used in the performance analysis.
The total head, H, is calculated by the polytropic head given by *equation 3.119.*

$$H = \frac{H_p}{\eta_p} \tag{3.119}$$

The amount of power, P_{fluid}, required to compress a fluid, excluding the mechanical and friction losses is given by *equation 3.120*

$$P_{fluid} = \dot{m}\, H = \rho_1 Q_1 H \tag{3.120}$$

where

ρ_1 - Density at suction , kg/m^3

Q_1 - Flowrate at suction, m^3/s

H - Total Head, m

3.3.3 Mechanical Design (Wall thickness and Weight) and Footprint

The compressor weight and footprint was determined from the manufacturers catalog corresponding to the calculated power requirements. The power requirements as depicted are calculated based on parameters of the suction and discharge streams. Since compressor design is very much company specific, the weight and footprint were directly picked from the *Elliot Compressor catalog* highlighted in *Appendix E.1*. Based on the power requirements the main compressor configuration was the frame 10 in either vertical or horizontal configuration depending on the pressure limits.

3.4 Piping

There are numerous factors that need to be considered when designing, constructing and operating a pipeline system. These differ in terms of onshore and offshore requirements. The pipeline systems in offshore processing plants can either be liquid, gas or multiphase pipeline systems. The total pressure drop required to transport a specified volume of fluid from point A to point B will consist of

- Frictional component
- Elevation component
- Pipe delivery pressure

The scope of this master thesis with respect to pipeline design is limited to determination of optimum diameter to achieve a specified flow velocity. This relates to the mechanical properties of the pipe specifically to wall thickness which in turn relates to weight of the pipe. The preliminary design is based on assumption of steady-state isothermal flow in gas pipelines. The scope of this research does not go into pipeline calculations taking into account elevations.

The calculator developed for the pipeline draws up a basic preliminary design calculation for the offshore gas processing platform based on pressure, flow rate as well as mechanical properties of pipe to obtain the *optimum diameter*, *wall thickness* and *pipe weight* based on optimum gas velocity.

3.4.1 General Flow Equation

The fundamental flow equation for the steady-state isothermal flow in a gas pipeline; as shown in *Figure 3.12*, is given from *equation 3.121* given that in the pipe segment from section 1 to section 2, the gas temperature T_f is assumed to be constant (isothermal flow).

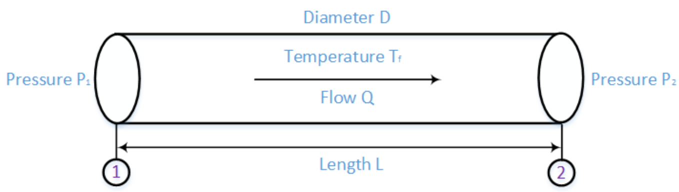


Figure 3.12: Pipeline illustration - steady flow

$$Q = 1.1494 \, x \, 10^{-3} \left(\frac{T_b}{P_b}\right) \left[\frac{(P_1^2 - P_2^2)}{(GT_f LZf)}\right] D^{2.5} \tag{3.121}$$

Equation 3.121 relates the capacity of a pipe segment of length L, based on an upstream pressure of P_1 and a downstream pressure of P_2 with the assumption that there is no elevation difference between upstream and downstream points; hence a horizontal pipe.

The *general flow equation* can be written in terms of transmission factor F, as depicted in *equation 3.122*; G represents the gas gravity

$$Q = 5.747 \, x \, 10^{-4} F \left(\frac{T_b}{P_b}\right) \left[\frac{(P_1^2 - P_2^2)}{(GT_f LZ)}\right] D^{2.5} \tag{3.122}$$

where the *Transmission factor*, F, and friction factor f are related by *equation 3.123*

$$F = \frac{2}{\sqrt{f}} \tag{3.123}$$

The transmission factor is inversely proportional to the friction factor. The friction factor indicates the resistance to flow a volume of gas through pipeline, whereas the transmission factor is a measure of the quantity of gas that can be transported through a pipeline.

When elevation difference between the ends of a pipe segment is included, the elevation should be incorporated in the length term in the general flow equation as in *equation 3.124*

$$Q = 5.747\ x\ 10^{-4} F \left(\frac{T_b}{P_b}\right) \left[\frac{(P_1^2 - e^s\ P_2^2)}{\left(GT_f L_e Z\right)}\right] D^{2.5} \tag{3.124}$$

Where e represents base of natural logarithms and takes the value 2.718.

$$L_e = \frac{L(e^s - 1)}{s} \tag{3.125}$$

$$s = 0.0684G \left[\frac{H_2 - H_1}{T_f Z}\right] \tag{3.126}$$

where s represents the elevation adjustment parameter; H_1 and H_2 represent the upstream and downstream elevation.

L_e in *equation 3.124* assumes a single slope between upstream point 1 and downstream point 2. For a series of slopes for a pipe segment L each individual subsegment that constitutes the pipe length from point 1 to point 2 is given by;

$$j = \frac{e^s - 1}{s} \tag{3.127}$$

j is calculated for each slope of each pipe subsegment of length L_1, L_2 etc that make up the total length L. The equivalent length $L_e$in *equation 3.124* is calculated by summing the individual slopes as defined below

$$L_e = j_1 L_1 + j_2 L_2 e^{s1} + j_3 L_3 e^{s2} + \cdots \tag{3.128}$$

The j terms are calculated from each s for each rise or fall in elevations of individual pipe subsegments. For the purpose of having a preliminary design; this thesis assumes the pipe segment to be horizontal. (Menon, 2005)

3.4.2 Compressibility factor

The compressibility factor, Z, is a measure of the deviation of a real gas from ideal gas. The compressibility factor is defined as the ratio of the gas volume at a given temperature and pressure to the volume the gas would occupy if it were an ideal gas at the same temperature and pressure.

There are several approaches to calculating the compressibility factor for a given gas temperature and pressure. The below are some of the methods used in the determination of Z. (Menon, 2005)

a. Standing-Katz Method
   Method utilises the critical temperature and critical pressure to obtain the pseudoreduced temperature and pressure where these are used to derive Z factor from Standing-Katz charts. Appendix F.
b. Dranchuk, Purvis and Robinson Method
c. California Natural Gas Association Method (CNGA); this is given by;

$$Z = \frac{1}{\left[1 + \left(\frac{P_{avg}\ 344{,}400(10)^{1.785G}}{T_f^{3.825}}\right)\right]} \tag{3.129}$$

Which is valid for average gas pressure of more than 6.9 barg (100psig)

P_{avg} = average gas pressure, psig; where
T_f = average gas temperature, °R
G = Gas gravity (air = 1.0)

Within this thesis, the compressibility factor is obtained from equations based on the Standing-Katz chart.

3.4.3 Velocity of Gas in Pipeline

The velocity of gas in a pipeline is a critical parameter to be determined at preliminary design. This is related to flowrate of the gas as an increase in the flowrate of the gas Q results in an increase in the velocity. The velocity of a gas at any point in a pipeline is given by equation

$$u = 14.7349\ \left(\frac{Q_b}{D^2}\right)\left(\frac{P_b}{T_b}\right)\left(\frac{ZT}{P}\right) \tag{3.130}$$

where

u = gas velocity , *m/s*
Q_b = gas flowrate at standard conditions, *m^3/day*
D = pipe inside diameter, *mm*
P_b = base pressure, kPa
T_b = base temperature, K
P = pressure, kPa
T = average gas flowing temperature, K
Z = gas compressibility factor at the flowing temperature

The velocity of the gas as indicated increases with flowrate. As velocity increases, vibration and noise occur. Higher velocities cause erosion of the interior of the pipe over a long period of time. Hence, the upper limit of gas velocity or maximum *erosional velocity* is determined from *equation 3.131*. An acceptable operational velocity is 50% of the erosional velocity. (Menon, 2005)

$$u_{max} = \frac{C}{\sqrt{\rho}} \tag{3.131}$$

where

u_{max} = maximum or erosional velocity, *m/s*
C = empirical factor, $kg^{0.5}m^{-0.5}s^{-1}$
ρ = gas density at flowing temperature, kg/m^3

The value of C is given for solids-free fluids based on continuous service and intermittent service. (Mokhatab, Poe, & Speight, 2006).

Further considerations of corrosion inhibition and the use of corrosion-resistant alloys for (API RP 14E, 1991) typical values of C are highlighted in *Table 3.3*.

Table 3.3: Empirical constants for erosional velocity (API RP 14E, 1991)

Continuous Service	100
Intermittent Service	125
Solids-free, No corrosion or CRA material (continuous service)	150 -200
Solids-free, No corrosion or CRA material (intermittent service)	150-250

Within this master thesis, the NORSOK standard for sizing of gas and liquid lines and determining the maximum erosional velocity is used. (NORSOK - Norwegian Oil Industry Association (OLF) & Standards Norway, 2006)

The gas lines are generally sized in order for the gas velocity not to exceed the acceptable noise level at the platform or create vibration problems. Per the standard this is given by *equation 3.132* (whichever is lowest);

$$V = 175\left(\frac{1}{\rho}\right)^{0.43} \quad or\ 60m/s \tag{3.132}$$

V - maximum velocity of gas to avoid noise, (m/s)
ρ - density of gas (kg/m^3)

For the sizing of liquid lines, per the NORSOK standard (NORSOK - Norwegian Oil Industry Association (OLF) & Standards Norway, 2006), maximum velocity is given by *Table 3.4*.
.

Table 3.4: Maximum velocities for sizing of liquid lines

Fluid	Maximum Velocities (m/s)			
	Carbon Steel	Stainless Steel	CuNi	GRP
Liquids [2]	6	7	3	6
Liquids with Sand [3]	5	7	N/A	6
Liquids with large quantities of mud or silt [3]	4	4	N/A	N/A
Untreated Seawater [1]	3	7	3	6
Deoxygenated Seawater [2]	6	7	3	6

Notes:

1) For pipe less than DN200 (8"), see BS MA-18 for maximum velocity limitations.
2) For Stainless Steels and Titanium the maximum velocities is limited by system design (available pressure drop/reaction forces).
3) Minimum velocity shall normally be 0.8 m/s
4) Minimum velocity for CuNi is 1.0 m/s.

With intermittent service, the velocity can be increased to 10 m/s. For CuNi the maximum velocity limit is 6 - 10 m/s depending on the duration and frequency of operation.

With corrosion inhibited fluids in carbon steel piping, the velocity is limited to wall shear stress of 40 N/m^2 to maintain the corrosion inhibiting film at the pipe wall, with the corresponding maximum velocity:

$$V_{max} = \sqrt{\frac{80}{f\rho}} \quad (m/s) \tag{3.133}$$

f - Fanning's Friction factor = ¼ of Dacy's friction factor (Moody diagram)
ρ - density of gas (kg/m^3)

For the purpose of the master thesis, the production flowrates are defined so as not to exceed optimum velocity given under *equation 3.132* for gas lines. Also, maximum velocities of condensate and liquid lines are evaluated based on information from *Table 3.4* for liquids with stainless steel (SS) with 7m/s as maximum velocity.

This would mean the limitations of achieving high production flowrates are due to erosional, vibration and noise limits on pipelines as well as operational envelopes on subsea processing equipment.

3.4.4 Friction Factor

Accurate predictions of friction are required to understand the relation of pressure drop along a pipe at a given flow rate. This project looks only at Darcy friction factor f and not the Fanning friction factor (where 4 times the Fanning Friction factor results in the Darcy friction factor). For laminar flow, the friction factor is inversely proportional to the Reynolds number;

$$f = \frac{64}{Re} \quad (3.134)$$

For turbulent flow, the friction factor is a function of the Reynolds number, pipe inside diameter and internal roughness of the pipe. Many empirical relationships are available for finding f . These could be;

- Colebrook-White equation / S. E Haaland
- American Gas Association (AGA) equation

For the purpose of this project, the friction factor would be derived from the modified Colebrook equation ; given as

$$\frac{1}{\sqrt{f}} \cong -1.8 \log\left[\frac{6.9}{Re} + \left(\frac{\varepsilon/D}{3.7}\right)^{1.11}\right] \quad (3.135)$$

3.4.5 Wall thickness (ANSI/ASME Standards)

The pipeline transmitting gas is subjected to various stresses. These include internal pressure from the fluid being transported, external stresses which could be as a result of hydrostatic pressure acting on the pipe in the case of subsea pipeline or pressure as a result of the weight from soil in the case of a buried pipeline.

In a subsea application the minimum wall thickness will be dictated predominantly by the internal pressure as well as the external pressure. The minimum wall thickness will depend on internal pressure, pipe diameter and the material of the pipe. The larger the pressure or diameter, the larger the wall thickness required. Steel pipes made of higher strength materials can withstand higher pressures hence will require less wall thickness as compared to low-strength materials.

In determining the wall thickness of the pipe, standards have been set out by the American Society of Mechanical Engineers (ASME) depending on the mode of application. *Table 3.5* below highlights the piping codes used within the oil and gas industry.

Table 3.5: ASME Piping Codes

ASME Piping Code	**Application**
ANSI/ASME Standard B31.1	Power Piping
ANSI/ASME Standard B31.3	Chemical plant and Petroleum Refinery Piping
ANSI/ASME Standard B31.4	Liquid Transportation Systems for Hydrocarbons, Liquid Petroleum Gas, Anhydrous Ammonia, and Alcohols. This standard applies to onshore oil pipeline facilities.
ANSI/ASME Standard B31.8	Gas Transmission and Distribution Piping Systems. This standard applies to gas transmission, gathering, and distribution pipelines onshore.

The thesis analyses the wall thickness based on ASME codes *B31.3*, *31.4* and *31.8*.
The wall thickness based on ASME 31.3 is given by *equation 3.136*

$$t = t_e + t_{th} + \left[\frac{Pd_o}{2(SE + PY)}\right]\left[\frac{100}{100 - T_{ol}}\right] \tag{3.136}$$

where

t - wall thickness , *mm*
t_e - corrosion allowance, *mm*
t_{th} - thread or groove depth, *mm (reference Appendix G.3)*
P - allowable internal pressure in pipe, *Pa*
d_o - outside diameter, *mm*
S - allowable Stress for Pipe, *Pa* (*reference Appendix G.4 and Appendix G.5*)
E - longitudinal weld-joint Factor (*reference Appendix G.6*)
Y - derating factor (0.4 for ferrous materials operating below 900°F)
T_{ol} - manufacturers allowable tolerance, % (12.5 pipe up to 20in. -OD, 10 pipe > 20 in OD, API 5L

The wall thickness given by ASME 31.4 is given by *equation 3.137*

$$t = \frac{Pd_o}{2(FES_Y)} \tag{3.137}$$

where

t - wall thickness , *mm*
P - internal pressure in pipe, *Pa*
d_o - outside diameter of pipe, *mm*
S_Y - allowable Stress for Pipe, *Pa* (*reference* Appendix G.7)
F - derating Factor, 0.72 for all locations
E - longitudinal weld-joint Factor (1.0 seamless, ERW, double submerged arc weld and flash weld; 0.80 electric fusion (arc) weld and electric fusion weld, 0.6 furnace butt weld

The wall thickness given by ASME 31.8 is given by *equation 3.138*

$$t = \frac{Pd_o}{2(FETS_Y)} \tag{3.138}$$

where

t - minimum design wall thickness , *mm*
P - internal pressure in pipe, *Pa*
d_o - Outside diameter of pipe, *mm*
S_Y - minimum yield stress for Pipe, *Pa* (*reference Appendix G.8*)
F - design factor (reference *Appendix G.9*)
E - Longitudinal weld-joint Factor (1.0 seamless, ERW, double submerged arc weld and flash weld; 0.80 electric fusion (arc) weld and electric fusion weld, 0.6 furnace butt weld (reference Appendix G.10)
T - temperature derating factor (*reference Appendix G.11*)

In this calculator developed for the pipeline, the wall thickness has been determined using the ASME standards as well as from pipeline standard data.

3.4.6 *Mechanical Design (Wall thickness and Weight) and Footprint*

In pipeline design, the weight off the pipe is required to ascertain the cost of the pipeline. This is dependent on the material of construction, the size of the pipe taking into account the wall thickness based on the application and the corrosion allowance.

From the determined outside diameter and wall thickness, a simple correlation to determine the weight of the pipe based on steel as material is given as;

$$w = 0.0246 \times t \times (d_o - t) \tag{3.139}$$

where

w - pipe weight, kg/m

d_o - pipe outside diameter, mm

t - pipe wall thickness, mm

The equation relates to pipes made of steel and incorporates the density of steel. For other pipe material, the ratio of densities can be applied to account for pipe weight for non-steel pipe. (Menon, 2005)

It is worth noting that for this master thesis, two general methods where investigated in determining the wall thickness. The first utilising the ANSI/ASME standards based on the application. The second, a more simplified approach, uses standard pipe parameters and optimum flowrates based on NORSOK standards (refer to *section 3.4.3*). Also, actual sectional pipeline lengths have not been considered or modelled in the absence of actual pipeline field data for comparison purposes. Weights of the pipeline have been presented as weight per metre (kg/m). Refer to *Appendix G.2*.

3.5 Pumps

3.5.1 Pump Design

There are various pump designs by numerous vendors for specific pumping applications. For the purpose of this master thesis, a basic centrifugal pump design is assumed to narrow down on basic design parameters needed to assess the performance and required parameters for the equipment analysis.

The pumps required for gas processing are seawater pumps to provide cooling for the shell and tube heat exchangers. With that in mind, the pump is needed to deliver an amount of seawater (flowrate) to achieve the necessary cooling or heat transfer for the process. Parameters to be investigated are;

- Head :- This includes the total differential head.
- Net Positive Suction Head Available (NPSHA) versus Net Positive Suction Head Required (NPSHR)
- Pump Power

Total Differential Head

The total differential head of a pump is determined by the flowrate of liquid being pumped and the systems through which the liquid flows. Frictional head losses exist in the system which work against the pump and the static head difference which is the difference in head between the discharge static head and the suction static head. This is given as

$$\boldsymbol{Total\ differential\ Head = Static\ Head\ difference + Frictional\ head\ losses}$$

Static Head Difference

The static head difference across the pump is given as the difference in head between the discharge static head and the suction static head. Given as;

$$Static\ Head\ difference = \ discharge\ static\ head - \ suction\ static\ head$$

Discharge Static Head – This is the sum of the pressure existing at the surface of the liquid in the discharge vessel in this case within the heat exchanger (expressed as *head*) and the difference in elevation between the discharge line and the centre line of the pump. Given as;

$$\begin{aligned} &Discharge\ static\ head \\ &\qquad = Discharge\ vessel\ gas\ pressure\ head + elevation\ of\ discharge\ pipe\ outlet \\ &\qquad - elevation\ of\ pump\ centre\ line \end{aligned}$$

Suction Static Head - The static Suction Head is the sum of the gas pressure at the surface of the liquid in the suction vessel (expressed as head) in this case this is assumed as a tank of seawater and the difference in elevation between the surface of liquid in the suction vessel and the centre line of the pump.

$$\begin{aligned} &Suction\ static\ head \\ &\qquad = Suction\ vessel\ gas\ pressure\ head + elevation\ of\ suction\ vessel\ liquid\ surface \\ &\qquad - elevation\ of\ pump\ centre\ line \end{aligned}$$

Frictional Head Losses

The Frictional losses in the system is comprised of the frictional losses in the suction piping and discharging piping system.

Friction in the piping system is as a result of viscous effects within the pipe. This is given from the Darcy-Weibasch factor (Cimbala & Cengel, 2008) in *equation 3.140*. The fittings, valves and bends contribute to the losses in the pipe. From the GPSA Engineering Data Book highlights equivalent lengths for valves and fittings for calculation of the losses within the piping system. This is captured under *Appendix I*. (Gas Processors Suppliers Association (U.S.), 2012)

$$\Delta P_L = f \frac{L}{D}\frac{\rho V_{avg}^2}{2} \tag{3.140}$$

where the pressure loss is dependent on

f - friction which is defined under

$$\frac{1}{\sqrt{f}} = -2.0 \log\left[\left(\frac{\varepsilon/D}{3.7}\right) + \frac{2.51}{Re\sqrt{f}}\right] \tag{3.141}$$

L - length of pipe, *m*
D - diameter of pipe, *m*
ρ - fluid Density, kg/m^3
V_{avg} - average Fluid velocity, *m/s*

Net Positive Suction Head Available (NPSHA)

The Net Positive Suction Head Available (NPSHA) is given as the difference between the absolute pressure at the pump suction and the vapour pressure of the liquid being pumped at the given temperature. The pressure at the suction needs to be above the vapour pressure to maintain the liquid being pumped in the liquid state and prevent the formation of vapour-filled bubbles. These bubbles could cause cavitation in the pumps which in turn causes undesirable noise, vibrations, reduction in efficiency and possible damage to the pump impeller blades.
The calculated NPSHA must exceed the Net Positive Suction Head Required (NPSHR) given from the manufacturer specification for the specific pump and indicated on the pump curve. The NPSHA is given as;

$$\text{NPSHA} = \text{Absolute Pressure Head at Suction} - \text{Liquid Vapour Pressure Head}$$

Pump Power

Driver selection for pumps could range from electric motors, diesel engines and steam turbines. The pump power is given by

$$P = \frac{\rho g Q H}{\eta} \tag{3.142}$$

Where
P - power, W or kW
ρ - fluid density, kg/m^3
g - acceleration due to gravity, m/s^2

Q	- flow rate, m^3/s
H	- total differential head, m
η	- pump efficiency

3.5.2 Submersible Pump (Seawater pump)

In the case of the master thesis, it should be noted that a submersible pump was modelled as this is normally the pump type used as a seawater pump for offshore installations. The principle behind the submersible pump is similar to that of a centrifugal pump where the *Total Dynamic Head* is determined by the pump levels and frictional losses.

The submersible pump is submerged below sea level placed in a protective casing or caisson. They do not require a pump room and are suspended from riser pipes in caissons mounted outside or integrated into the hull. With such a design the contribution of the submersible pump to footprint is negligible.

The seawater pumps in this application are used as source of power for the cooling media (sea water) in the heat exchangers during the processing of the gas. The seawater rates of the pump are determined by the heat transfer required in the heat exchanger of the shell side. *Figure 3.13* shows the pictorial view of the sea water pump layout.

The *Total dynamic head* (TDH) of the submersible pump is given by *equation 3.143*

$$TDH = Pumping\ Level + Vertical\ Rise + Friction\ Loss \tag{3.143}$$

The *vertical rise* is been assumed to be zero as the discharge (horizontal pipe) has been assumed to be on the same level as the heat exchanger.

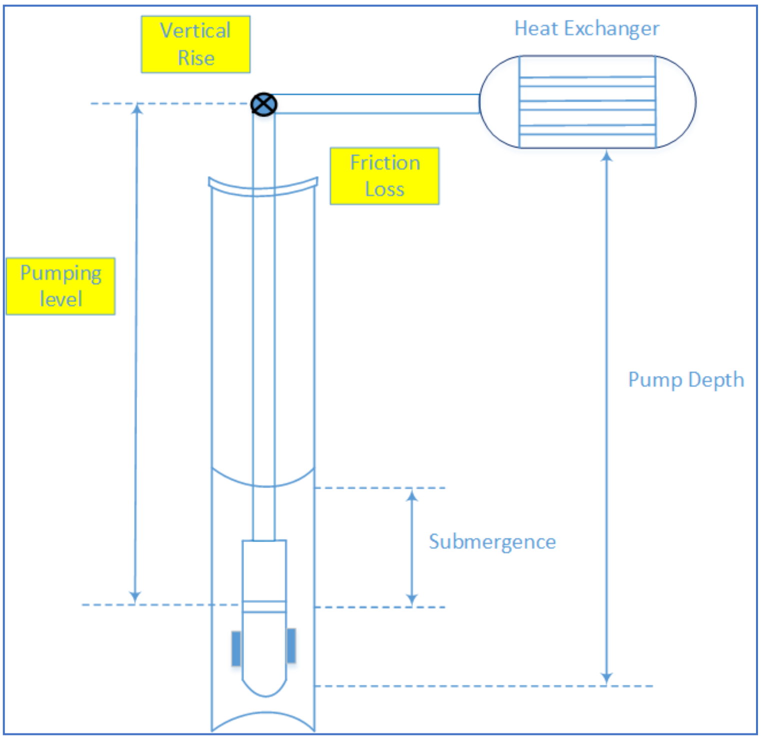


Figure 3.13: Submersible pump layout

The frictional losses are correlated from *Appendix I.2* knowing the pipe size and required flowrate.

$$Friction\ Loss = Total\ Length\ \times friction\ loss\ (pipe) factor + friction\ loss\ (fittings) \qquad (3.144)$$

The submersible pump to deliver the required flow rate and head is chosen from the vendor information (courtesy of Framo) given under *Appendix I.3.*

3.5.3 Mechanical Design (Wall thickness and Weight) and Footprint

There are many pump manufacturers and designs for specific applications. The pump design flowrate total head and pump weight are obtained from the manufacturer's catalog. As pump designs are manufacturer specific the designs needed for the study were taken from Framo Submersible Model pump indicated in *Appendix I.3.*

4 Simulation of Offshore Gas Processing Plant

This chapter highlights the model build-up of the offshore processing plant. ASPEN HYSYS was used to simulate the plant based on the case study scenario highlighted in *Chapter 1*. The platform inlet of 90 bara and temperature of 5°C was given. The hydrocarbon cricondenbar specification of 90 bara and export conditions of 200 bara and 15°C are also specified for the process with a feed flow rate of 5MMscmd.

The model simulates the main processes to achieve the specifications given. These include;

- Saturation of the gas to model wet gas
- Condensate stabilisation
- Dew point control/ Cooling and separation
- Gas dehydration using component splitter

The objective is to utilise the simulation to develop the process equipment calculators based on the theory highlighted in *Chapter 3*. In addition, the simulation is used to perform analysis with respect to different thermodynamic models specifically Soave-Redlich-Kwong (SRK) and Peng Robinson (PR). Other life-of-field parameters are compared which are captured in the chapters that follow. The simulation was used in comparing the equipment calculators with the simulation based on different thermodynamic models;

4.1.1 Saturation of Gas

The gas conditions from the case study given were simulated to saturate the well stream at 180 bar and 80°C prior to entering the plant at inlet separator.

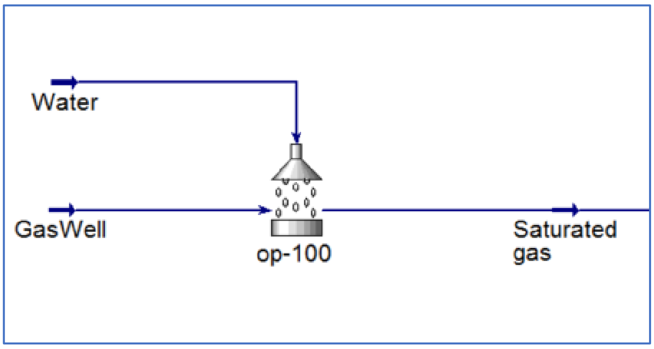


Figure 4.1: Saturated gas process

4.1.2 Condensate Stabilisation

The condensate stabilisation process is a 3-stage flash process. The liquid stream from the inlet separator is heated to aid in separation of the gas and liquid components which includes monoethylene glycol (MEG). The liquid component from the boot of the 1st stage 3-phase separator would be directed to the MEG unit for regeneration. The MEG regeneration stream is not included in the process simulation. The condensate stream undergoes further flashing from 77 bar to 8.8 bar and then to atmospheric conditions where stable condensate is obtained.

The 3-stage flash separation incorporates the use of:

i. 3-phase Horizontal Separator

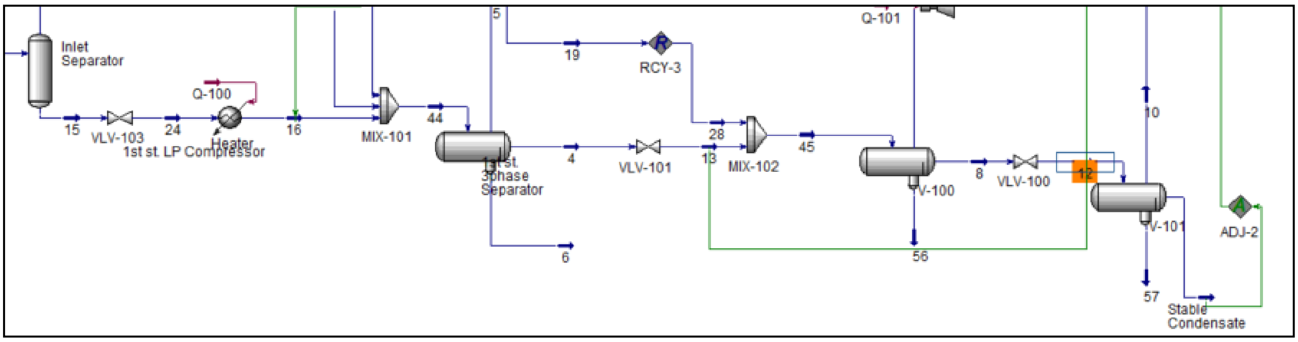


Figure 4.2: Condensate stabilisation layout

4.1.3 Hydrocarbon Dew Point Control

The hydrocarbon dew point of the process is achieved with the use of heat exchangers, scrubbers and compressors. By cooling the gas stream and separating out the heavy components the dew point of the mixture is controlled.

This is done to extract the liquid component for market and to prevent freeze out of the heavy components during transport. The cricondenbar specification of 90 bar was achieved with a Joule Thompson valve with pressure let down to 77 bar downstream of the valve.

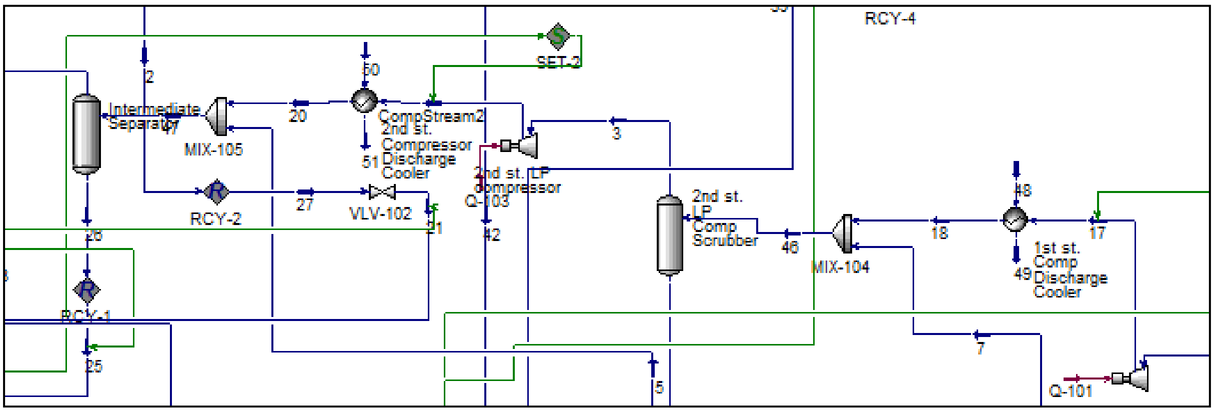


Figure 4.3: Hydrocarbon dew point control layout

4.1.4 Dehydration

The simulation developed on ASPEN HYSYS pertaining to the scope of this project does not include the absorption and regeneration portion of the processing plant. This could be considered for a future project. For the purpose of simulating the gas to meet the required water specification post the dehydration unit; a component splitter has been used to represent the dry gas specification from the dehydration unit. A water specification of 36 ppm has been simulated with the component as shown in *Figure 4.4* .

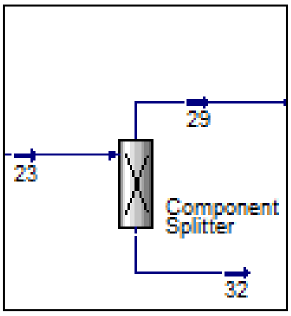


Figure 4.4: Component splitter (Dehydration specification)

4.1.5 Compression and cooling for export

The dry gas after meeting cricondenbar and dew point specification is compressed and routed via pipeline for further processing onshore. The compression and cooling process is undertaken in two stages with a heat exchanger and gas scrubber to remove any entrained liquids that could damage the export compressors. The dry gas is compressed and cooled to meet platform outlet specification of 200 bar and 15°C.

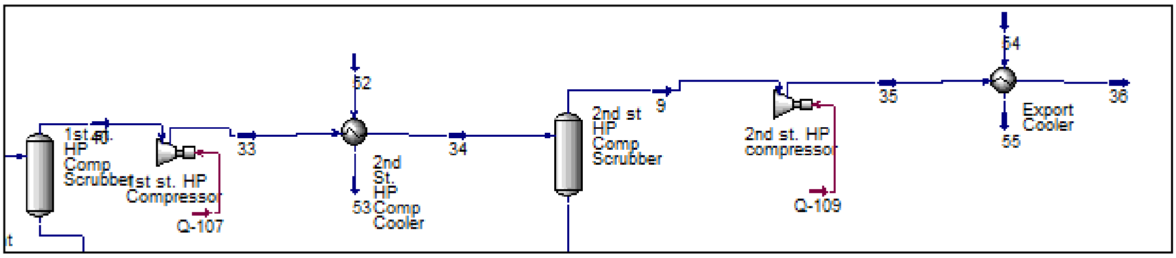


Figure 4.5: Export compression unit

The entire layout of the simulation performed on ASPEN HYSYS is shown in *Figure 4.6. Appendix J* shows the detailed stream and equipment property tables generated on ASPEN HYSYS.

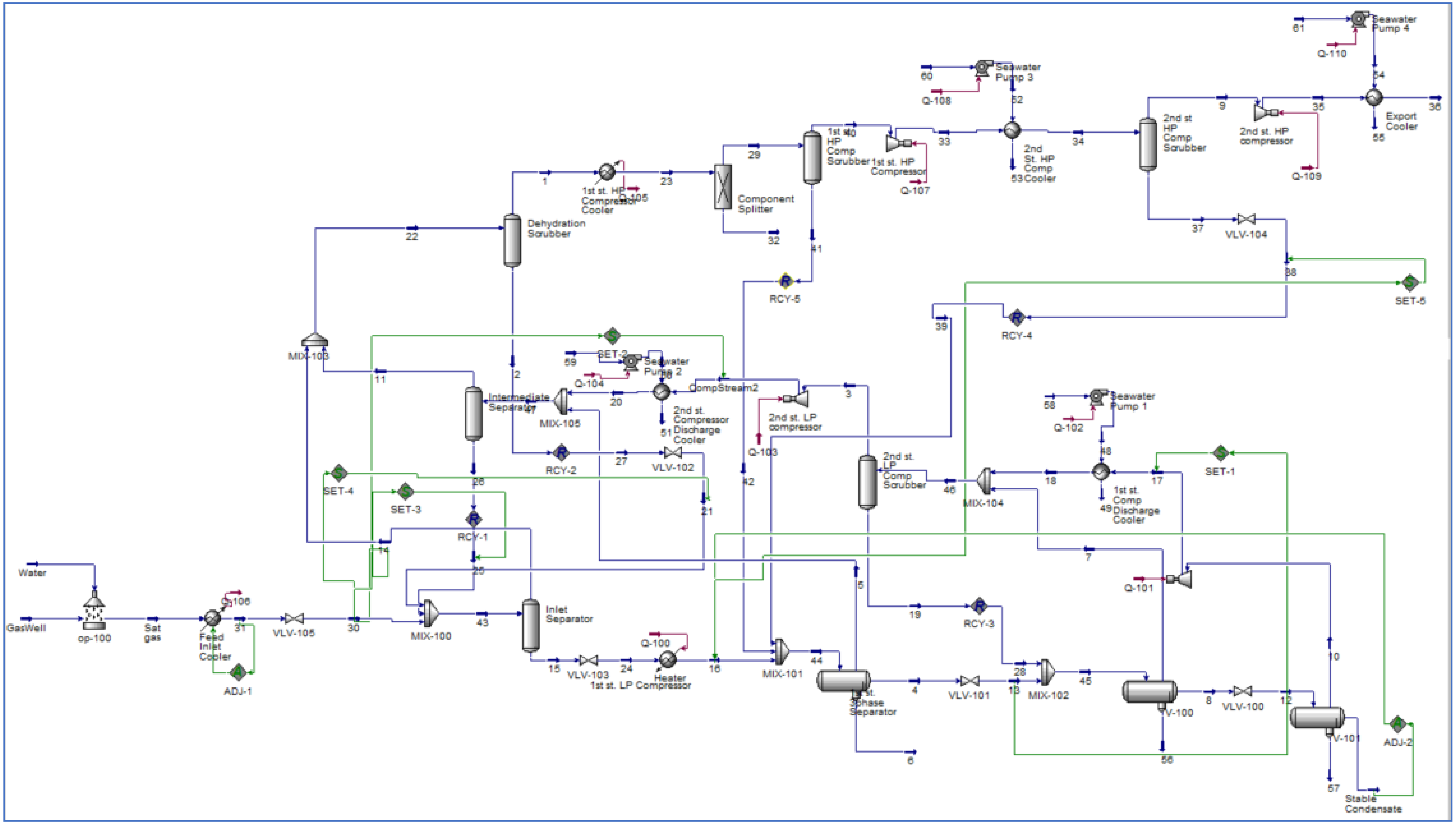


Figure 4.6: Offshore natural gas processing - Simulation of base case using ASPEN HYSYS

5 Evaluation of Equipment Calculator with HYSYS.

This chapter demonstrates the use of the theory highlighted in *Chapter 3* in sizing equipment for the offshore gas processing plant. The sizing design calculations were performed for the separation train for condensate stabilisation, compressor scrubbers, heat exchangers, compressors and pipeline. The design calculations herein referred to as "*sizing calculator*" were developed in MS Excel. The ASPEN HYSYS software was utilised in simulating the offshore processing plant (investigating both SRK and PR EoS) and used in conjunction with the calculator. The calculator developed is not designed for rigorous in-depth equipment design however gives a basis for preliminary design and sensitivity. The analysis gives an output of the different parameters of length, width, height, performance parameters, weight and footprint of the equipment.

5.1.1 Separation Equipment

The offshore processing platform utilises different types of separation equipment for different objectives. Separation equipment is utilised for condensate stabilisation for flash separation, scrubbers for liquid removal from gas and to ensure compressor safe operation. The design of the *Separator Calculator* is based on different theoretical methods and best practices for separator design from *Petroleum and Gas Field Processing by Abdel-Aal H.K.; Mohammed Aggour and Fahim M.A.* as well as *Design Two-Phase Separators Within the Right Limits* and *Successfully Specify Three-Phase Separators by Monnery W. D and Svrcek W. Y. Figure 5.1* and *Figure 5.2* give the typical design layout of the calculator for both 2-phase and 3-phase separator.

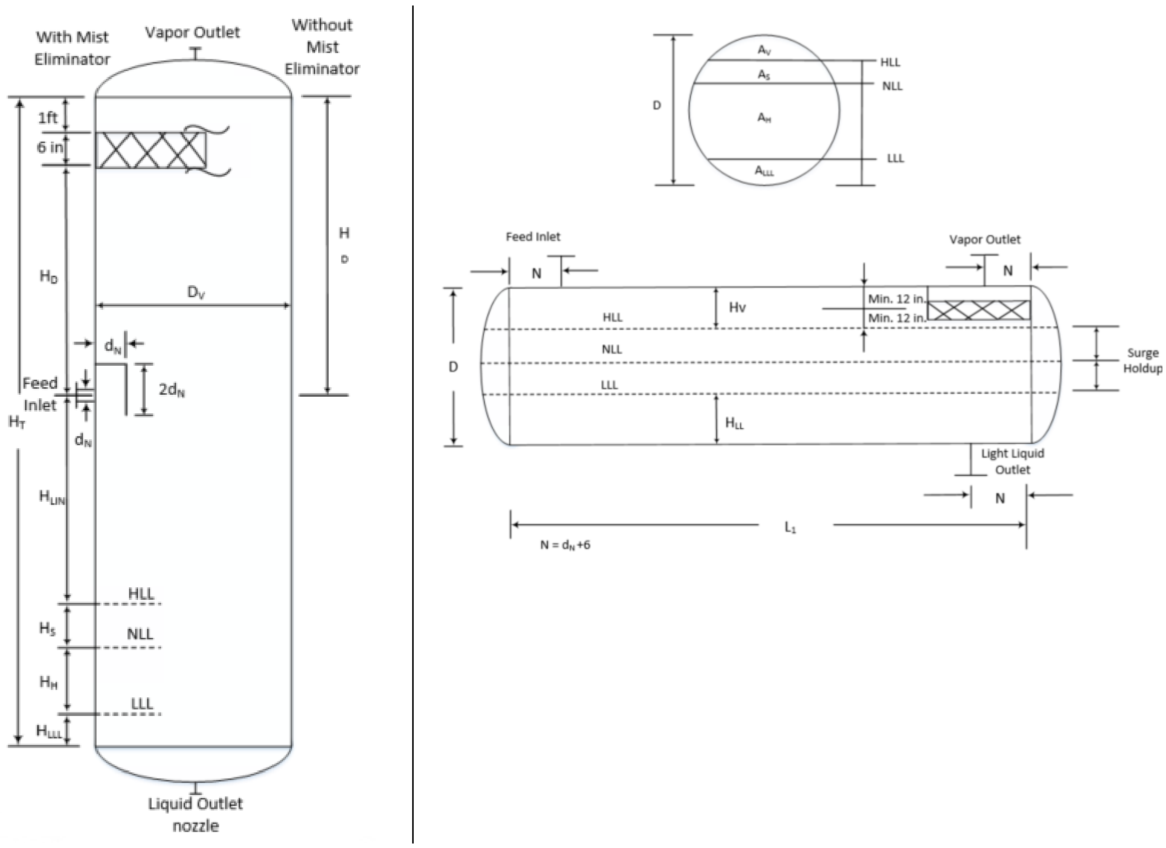


Figure 5.1: 2-Phase vertical and horizontal separator design layout

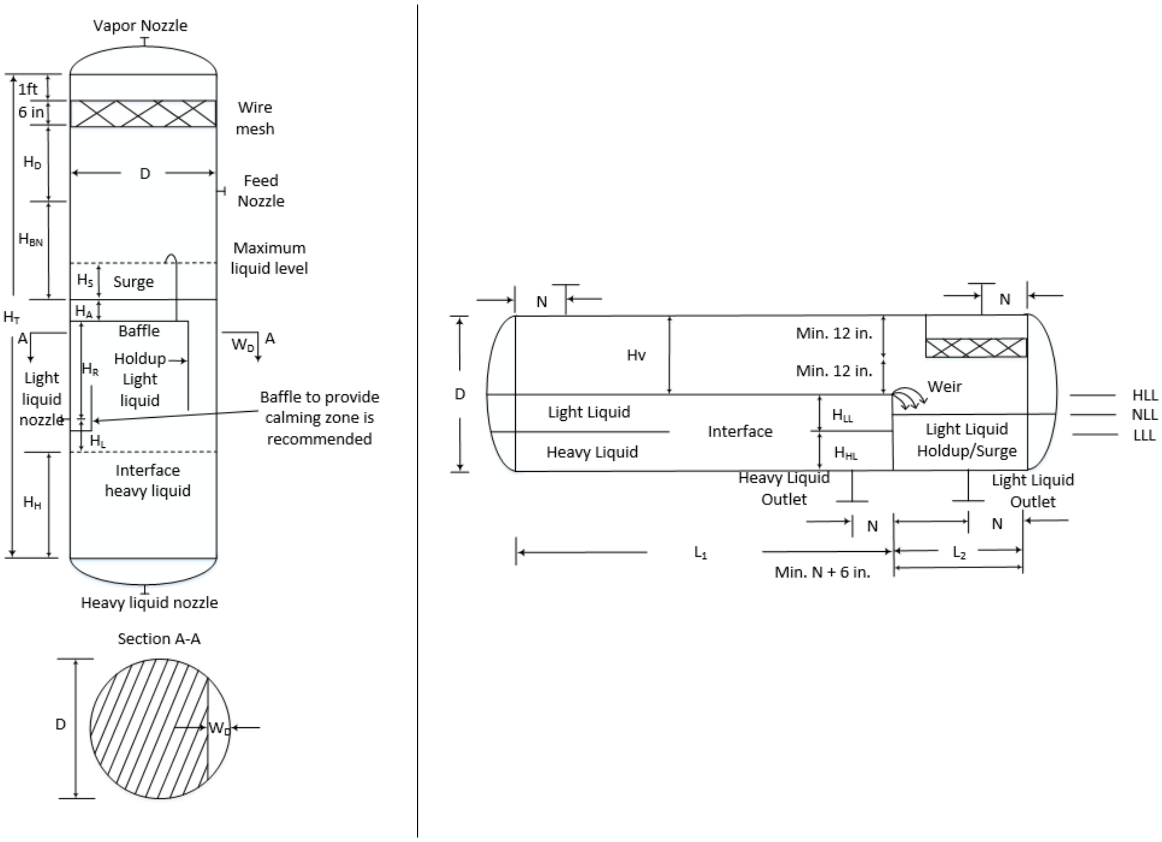


Figure 5.2: 3-Phase vertical and horizontal (weir configuration) separator design layout

The Separator calculator gives an overview design of the different types of separators within the modelled simulation. These are classified under;

5.1.1.1 Condensate Stabilisation

The separators utilised for condensate stabilisation within the process were a three-phase horizontal for the 1st stage flash separator to include MEG Regeneration. This configuration or type was assumed and designed as a separator with interface control with weir as depicted in *Figure 5.2*.

Appendix C.4 gives a representation of the calculator developed and incorporates the functional design for 3-phase horizontal separator labelled as *1stStage 3-phase Separator* which is the 1st stage flash separation. The 2nd and 3rd stage flash incorporate 3-phase horizontal separators and have been designed as such within the calculator as depicted in *Figure 5.2*. The 3rd stage flash upon investigation did not contain any water for the production flowrate investigated. Hence, this was actually modelled as 2-phase horizontal separator to take into account no liquid water. An example of the procedure in setting up the design is also explained in detail under *section 5.1.1.4*.

Upstream of the condensate stabilisation process is the receiving separator, termed *Inlet Separator*, which is a 2-phase vertical separator designed to separate large volumes of gas and entrained liquids. This is depicted in *Figure 5.1*. *Appendix C.1* gives the representation of the calculator. A detailed step-wise procedure in developing the calculator is presented under *section 5.1.1.3*.

The design of the separators was done to ascertain a length-to diameter ratio of 1.5 to 6.0 for the 2-phase horizontal and 3-phase horizontal separators. This was assumed as the optimum target for the equipment design. *Table 5.1* and *Table 5.2* show the results of the sizing calculator for the 3-stage flash separators utilising SRK and PR EoS for fluid characterisation.

5.1.1.2 Compressor Scrubbers and Liquid Removal (High Pressure Export Compressors)

Two-phase vertical separators are used upstream of the compressor as scrubbers for safety. The scrubbers are used to remove 3-5 volume % of liquid. The limits of the scrubbers have a carry-over specification of 13 litres/Msm^3. Within the HYSYS model the separators for the high pressure (*HP*) compressors have been simulated to have approximately no liquids entering the 1^{st} stage and 2^{nd} stage HP compressor scrubbers and gives no sizing relation in HYSYS. Within the calculator on the other hand, a conservative design flowrate of liquids has been assumed to account for liquids carryover. The separators have been given the nomenclature *1^{st} stage and 2^{nd} stage High Pressure Compressor Scrubber*. *Table 5.3* and *Table 5.4* captures the output of the calculator for the 2-phase HP compressor scrubbers.

Table 5.1 and *Table 5.2* highlight the base case 3-phase separator calculator design results utilising input parameters from SRK EoS and PR EoS.

Table 5.1: 3-Phase horizontal separator design parameters with Soave Redlich-Kwong EoS

THREE PHASE SEPARATORS	Flow Rate	Pressure	Temperature	Diameter	Length	L/D	Footprint	Volume	Weight
	sm^3/h	bar	°C	m	m		m^2	m^3	kg
3-Phase Horizontal									
1st Stage 3-Phase Separator	20,350	77	69.7	2.68	7.42	2.77	59.7	380	129,551
2nd Stage Separator	14,280	8.8	68.8	2.91	6.74	2.30	58.8	400	17,227
3rd Stage Separator*	7,464	1.013	20.0	1.58	2.38	1.51	18.8	78	1,157

Table 5.2: 3-Phase horizontal separator design parameters with Peng Robinson EoS

THREE PHASE SEPARATORS	Flow Rate	Pressure	Temperature	Diameter	Length	L/D	Footprint	Volume	Weight
	sm^3/h	bar	°C	m	m		m^2	m^3	kg
3-Phase Horizontal									
1st Stage 3-Phase Separator	19,580	77	68.8	2.64	7.35	2.78	58.8	366	124,917
2nd Stage Separator	13,618	9	67.9	2.88	6.70	2.32	58.1	393	16,935
3rd Stage Separator*	7,252	1.013	20.2	1.57	2.37	1.51	18.5	77	1,154

**In the sizing calculator this was modelled as a 2-phase horizontal separator as HYSYS simulation did not have parameters for heavy liquid phase water.*

Table 5.3 and *Table 5.4* highlight the 2-phase separator design specifications and makes a comparison of the calculator sizing design with the output parameters from HYSYS as input for the calculator. The output parameters from utilising SRK and PR are shown.

Table 5.3: Separator design parameters with SRK EoS

TWO PHASE SEPARATORS	**Flow Rate**	**Pressure**	**Temperature**	**Diameter**	**Height**	**Footprint**	**Volume**	**Weight**
	sm^3/h	**bar**	°C	**m**	**m**	**m^2**	**m^3**	**kg**
2-Phase Vertical								
Inlet Separator	214,200	77	0.7	1.70	7.6	14.5	180.2	52,603
2nd Stage LP Compressor Scrubber	8,677	8.8	24.4	0.78	2.9	3.1	16.2	881
Intermediate Separator	13,870	77	38.9	0.59	13.2	1.7	35.8	11,814
Dehydration Scrubber	202,800	77	2.6	1.57	2.3	12.3	54.2	14,383
1st St. HP Comp Scrubber**	202,800	77	25.1	1.64	2.3	13.3	59.4	15,495
2nd St. HP Comp Scrubber**	202,800	120	30.0	1.49	2.2	11.0	48.1	19,659

Table 5.4: Separator design parameters with Peng Robinson EoS

TWO PHASE SEPARATORS	**Flow Rate**	**Pressure**	**Temperature**	**Diameter**	**Height**	**Footprint**	**Volume**	**Weight**
	sm^3/h	**bar**	°C	**m**	**m**	**m^2**	**m^3**	**kg**
2-Phase Vertical								
Inlet Separator	213,500	77	0.4	1.68	7.5	14.1	174.1	50,776
2nd Stage LP Compressor Scrubber	8,144	8.8	24.1	0.75	2.9	2.8	15.2	846
Intermediate Separator	13,240	77	38.7	0.57	11.6	1.6	29.7	9,908
Dehydration Scrubber**	202,900	77	2.3	1.55	2.3	12.1	53.1	14,107
1st St. HP Comp Scrubber**	202,900	77	25.1	1.62	2.3	13.0	58.4	15,243
2nd St. HP Comp Scrubber**	202,900	120	30.0	1.47	2.2	10.7	47.1	19,259

***In the process, these separators do not have a liquid phase. The separator was given a conservative design assuming liquid phase of density 1000kg/m^3 and a flowrate of 0.000001m^3/s*

Based on the results of the separator design and as summarised in *Figure 5.3* and *Figure 5.4*, utilising different equations of state; either Soave-Redlich-Kwong or Peng Robinson parameters, has an effect on design sizing of the separator. This is due to the differences in PVT fluid characterisation derivation from the different equations of state. This creates differences input design parameters such as *density* and *flow rate* which in turn impacts the output design parameters such as diameter, height or length, footprint and weight required for effective separation of the gas and liquid phases. The figures show a relative difference between the two thermodynamic models with SRK as the reference. Majority of the parameters showed reduced results for PR as compared to SRK.

Due to these differences in PVT characterisation (example as seen from temperature in) as a result of different EoS as much as ~3.5% difference is observed in some design parameters such as weight in the 3-phase 1st Stage horizontal separator.

Another observation (from *Figure 5.4*) is the marked difference of ~35% in temperature specifically the Inlet separator (2-phase vertical) as well as the significant differences in the weight and volume calculations for the Intermediate separator. This noted difference could be as a result of the known varied liquid volumetric predictions between the SRK and PR EoS as these separators have large amount of liquids and also due to the fact that PR EoS underpredicts saturation pressure of reservoir fluids compared with the SRK EoS (Whitson et al., 2000).

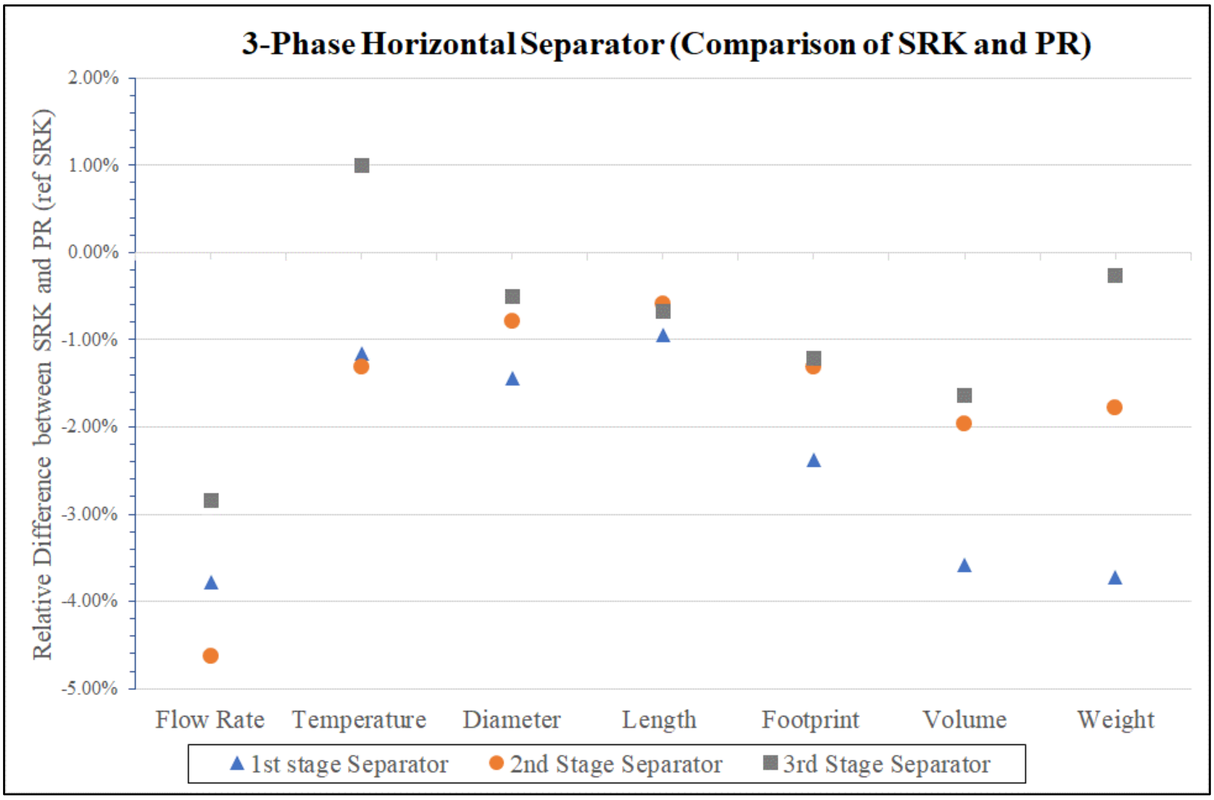


Figure 5.3: 3-Phase horizontal separator - Design comparison between SRK and PR EoS

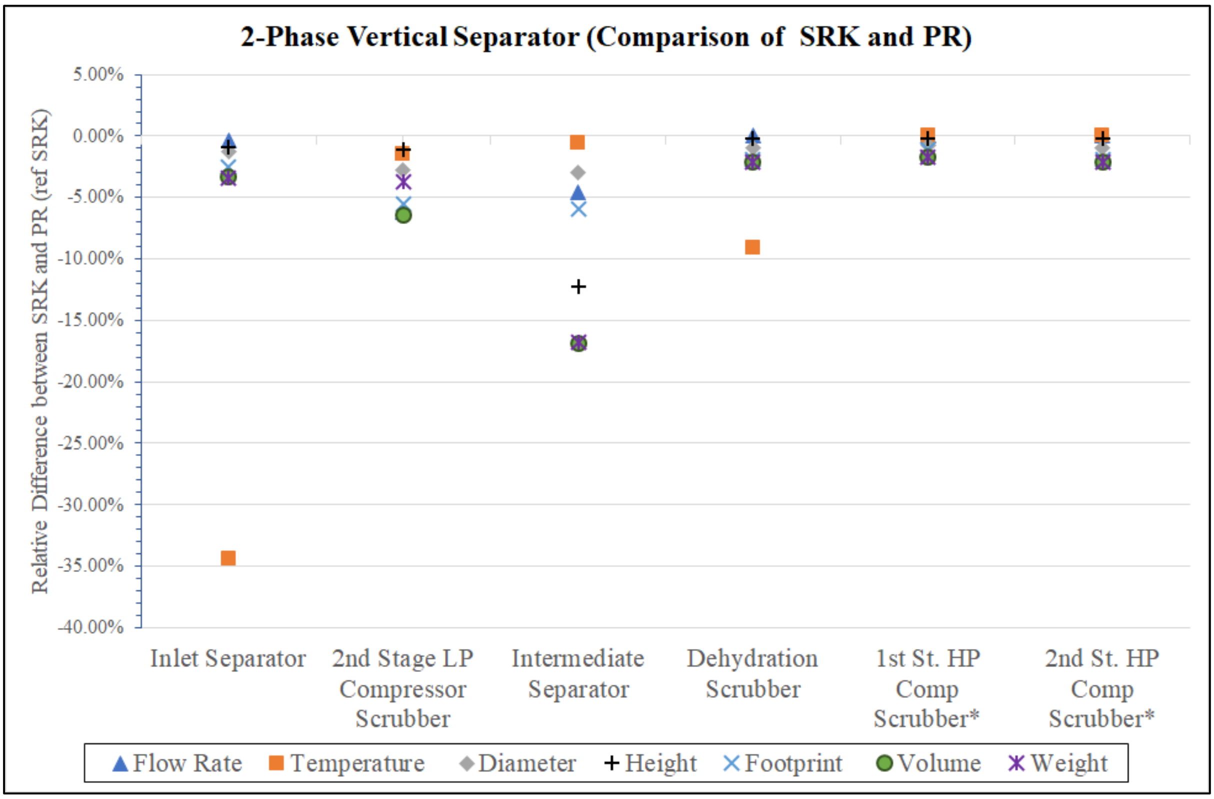


Figure 5.4: 2-Phase vertical separator- Design comparison between SRK and PR EoS

5.1.1.3 *Stepwise procedure for developing the 2-phase Vertical Separator calculator using API/GPSA sizing constant Ks (Svrcek & Monnery, 1993)*

Input parameters for the 2-phase vertical separator are obtained from the HYSYS model. Reference *Appendix C.12* for design sketch of the separator.

1. The input parameters are *liquid phase density, gas phase density, molecular weight of feed, gas flow rate in standard conditions* and *liquid actual volume flow rate*.
2. The sizing constant, K_s, is chosen from *Table 3.1* or from GPSA Standards for the application.
3. Obtain the *vapour mass velocity* from *equation 5.145.*
4. Determine the diameter of the separator from *equation 5.147* if there is a mist eliminator add 0.1524 m for allowance.
5. Determine the *holdup* and *surge* time from *Appendix C.7.* for the application of the separator.
6. Determine the holdup volume, V_H ,and surge volume, V_S, *from equations 5.148* and *5.149.*
7. Determine the low liquid level height, H_{LLL} from *Appendix C.8.*
8. Calculate the height from low liquid level to normal liquid level, H_H, and the normal liquid level to high liquid level (or high level alarm) H_S from *equations 5.151* and *5.152*.
9. Determine the height from high liquid level to the centreline of the inlet nozzle based on having an inlet diverter or not from *equation 5.153* where d_N is calculated from *5.154.*

10. Determine the disengagement height, from the centreline of the inlet nozzle to:
    a. The vessel top tangent line if there is no mist eliminator or
    b. the bottom of the demister pad from *equation 5.155.*
11. The total height of the vessel is determined from *equation 5.156.*

Weight Calculation

12. Determine the wall thickness and empty vessel weights from *equations 3.92, 3.93* and *3.94.*
13. The total weight of the vessel is calculated from *equation 5.157* knowing the weight of the internals, weight of the nozzles, piping, electrical and instrumentation and skid.

EQUATIONS

$v = K_s \left[\frac{\rho_l - \rho_g}{\rho_g}\right]^{0.5} \; m/s$ (5.145)	$u_v = 0.75K_s \left[\frac{\rho_l - \rho_g}{\rho_g}\right]^{0.5} \; m/s$ (5.146)
$D_{min} = \sqrt{\frac{4\, q_a}{\pi\, v}} \; m$ (5.147)	$V_H = T_H Q_L \, m^3$ (5.148)
$V_S = T_S Q_L \, m^3$ (5.149)	$V_S = T_S Q_L \, m^3$ (5.150)
$H_H = \frac{V_H}{(\pi/4)D_v^2} \; m$ (5.151)	$H_S = \frac{V_S}{(\pi/4)D_v^2} \; m$ (5.152)
$H_{LIN} = 12 + d_N$ (*with inlet diverter in metres*) $H_{LIN} = 12 + {}^{1}/_{2}\, d_N$ (*without inlet diverter in metres*) (5.153)	$d_N \geq \frac{1}{3.2808}\left(\frac{\frac{4Q_m}{\pi 60}}{\sqrt{\rho_m}}\right)^{0.5} \; m$ $Q_m = Q_L + Q_v \;\; m^3/s$ $\rho_m = \rho_l\lambda + \rho_v(1-\lambda) \;\; kg/m^3$ $\lambda = \frac{Q_L}{Q_L + Q_v}$ (5.154)
$H_D = 0.5\, D_V$ or a minimum of $H_D = 36 + {}^{1}/_{2}\, d_N$ (without mist eliminator)	$H_T = H_{LLL} + H_H + H_S + H_{LIN} + H_D + H_{ME}$ meters (5.156)

$H_D = 24 + {}^{1}/_{2}\, d_N$ (with mist eliminator) *Units in metres* (5.155)	
$t = 2.54\, x \dfrac{Pd}{2SE - 0.2P} \quad cm$ (3.92)	
$W_b = 3.47\ dt\ kg/m$ (3.93)	
$W_v = W_b L + W_I + W_N$ (3.94)	
$W_p = 0.4 * \ Weight\ of\ Empty\ Vessel$ (Weight of Piping) $W_E = 0.08 * \ Weight\ of\ Empty\ Vessel$ (Weight of Electrical and Instrument) $W_S = 0.1 * \ Weight\ of\ Empty\ Vessel$ (Weight of Skid steel) $Total\ weight = \ W_p + \ W_E + W_S \quad kg$	
(5.157)	

5.1.1.4 Stepwise procedure for developing the 3-phase Horizontal Separator (with weir) calculator

Input parameters for the 3-phase horizontal separator are obtained from the HYSYS model. Reference *Appendix C.11* for design sketch of the separator.

1. The input parameters are *density, viscosity flow rate holdup and surge times for different phases/fluids as well as pressure.*
2. The sizing constant, K_s, is chosen from *Table 3.1* or from GPSA Standards for the application.
3. Obtain the vertical terminal velocity, U_T from *equation 5.145.* and the conservative velocity, U_v, from *equation 5.146.*
4. Select the holdup and surge times from *Appendix C.7* and determine the holdup volume, V_H, and surge volume, V_S, *from equations 5.148* and *5.149.*
5. Obtain the initial L/D ratio from *Appendix C.9.*

6. Determine the diameter of the separator from *equation 5.158.* and the area, A_T, from *equation 5.159.*
7. Set the vapour space height H_v to to the larger of $0.2D$ or 0.6096 m.(0.3048m without a mist eliminator). Using *Appendix C.10* calculate A_V from H_v/D and A_v/A_T.
8. Calculate low liquid level height, H_{LLL}, from *equation 5.160* if $D \leq 1.22m$; then $H_{LLL} = 0.2286m$. Knowing H_{LLL}/D and from *Appendix C.10* can calculate A_{LLL}.
9. The weir height, H_W, is calculated from *equation 5.161.* If $H_W < 0.6096m$ increase *D and* repeat calculations from step 7.
10. Calculate the minimum length of the light liquid compartment to accommodate holdup/surge, L_2 from *equation 5.162.* The minimum length for L_2 to be $L_2 = d_N + 0.3048m$, where d_N is the nozzle diameter.
11. Set the interface at the height , $H_W/2$ to 50% of the separator height (or other). This is to define H_{HL} and H_{LL}.
12. For the liquid settling compartment, using the cross-sectional area of the heavy liquid H_{HL}/D determine A_{HL}, from *Appendix C.10* and from *equation 5.163* determine the cross-sectional area for light liquid, A_{LL}.
13. Calculate the settling velocity of the heavy liquid out of the light liquid phase, U_{HL}, and the light liquid out of the heavy liquid phase, U_{LH}, using *equations 5.164* and the assumed K_s.
14. Calculate the settling time for water to rise out of the oil, t_{HL}, and the settling time for oil to rise out of the water, t_{LH}, by dividing the known oil or water pad heights by the respective settling velocity. Ie $t_{LH} = H_{HL}/U_{LH}$ and $t_{HL} = H_{LL}/U_{HL}$.
15. Determine minimum L_1 based on *equation 5.165*
16. This gives $L = L_1 + L_2$
17. Liquid dropout, ϕ ,is calculated from *equation 5.166* and the actual vapour velocity, U_{VA}, from *equation 5.167.*
18. The minimum Length, L_{min}, required for vapour-liquid separation is calculated from *equation 5.168.*
19. If $L < L_{min}$ set $L = L_{min}$. If $L << L_{min}$ then increase H_V recalculate A_V and repeat calculations from step 7.

    If $L > L_{min}$ design is acceptable.

    If $L >> L_{min}$ (liquid Separation and Hold Up control) L can only be reduced and L_{min} increased if H_V is reduced. H_V may only be reduced if it is greater than minimum in L_2 calculation from step 10.
20. Determine L/D. If L/D<<1.5 then decrease D (unless already at minimum) if L/D >>6.0 increase D and repeat from step 6.
21. Determine the wall thickness as with *equation 3.93.*
22. Increase or decrease diameter by 0.1524 m and repeat calculations until L/D ranges between 1.5-6.0.
23. After obtaining optimum vessel size, calculate normal and high liquid levels from *equations 5.169* and H_{NLL} from *Appendix C.10.*

EQUATIONS

$D = \dfrac{1}{3.2808}\left(\dfrac{16(V_H + V_s)}{0.6\,\pi\,(L/D)}\right)^{\frac{1}{3}} \quad m$	$A_T = \dfrac{\pi D^2}{4} \quad m^2$
(5.158)	(5.159)
$H_{LLL} = 0.0254(0.5D + 7) \quad m$	$H_W = D - H_V \quad m$
(5.160)	(5.161)
$L_2 = \dfrac{V_H + V_S}{A_T - A_V - A_{LLL}} \quad m$	$A_{LL} = A_T - A_V - A_{HL} \ \ m^2$
(5.162)	(5.163)
$U_{HL} = \dfrac{K_s(\rho_H - \rho_L)}{\mu_L} \quad \mathrm{m/s}$ $U_{LH} = \dfrac{K_s(\rho_H - \rho_L)}{\mu_H} \quad \mathrm{m/s}$	$L_1 = \max\left(\dfrac{t_{LH}\,Q_{HL}}{A_{HL}}, \quad \dfrac{t_{HL}\,Q_{LL}}{A_{LL}}\right)$
(5.164)	(5.165)
$\phi = H_V / U_V$	$U_{VA} = Q_V / A_V$ m/s
(5.166)	(5.167)
$L_{min} = U_{VA}\,\phi \ \ m$	$H_{HLL} = D - H_V$ $A_{NLL} = A_{LLL} - V_H / L_2$
(5.168)	(5.169)

5.1.2 Heat Exchanger

The heat exchangers employed within offshore gas processing facility perform the main functions of dew point control by the method of cooling and separation and also to cool down the gas to meet export specifications.

The heat exchangers required to aid in the cooling separation process are the *1st stage compressor discharge cooler* and the *2nd stage Compressor Discharge Cooler*. The utility within the heat exchanger for cooling down the gas is sea water at 5°C and leaving the heat exchanger at maximum 20°C. As highlighted in *Chapter 3.2* the design incorporates assumptions made on the fouling factors and film transfer coefficients for natural gas and sea water at prevailing conditions. A counter current shell and tube (one pass) heat exchanger has been assumed. The fouling factor and film transfer coefficients used within the sizing calculator were assumed and are explained in *chapter 3.2* and given under *Appendix D.3.*

Appendix D.4 represents the design calculations for the designed heat exchangers.

Table 5.7 and *Table 5.8* show the output design sizing parameters for the calculator utilising the SRK and PR EoS. Based on the output of both designs as depicted from *Figure 5.5*; the evaluation showed differences as much ~7% in parameters such as duty. Differences as much as 3.2% were observed in parameters such as shell diameters and footprint covered by the heat exchanger with weight reaching a difference of 6%.

The largest differences in design parameters were observed in the 2nd stage discharge cooler.

Table 5.5: Heat exchanger design parameters with SRK EoS

HEAT EXCHANGER	HOT: T1	HOT: T2	COLD: T1	COLD: T2	LMTD	Shell Diameter	Footprint	Weight	Overall U	Duty	Massflow
	°C	°C	°C	°C	K	mm	m^2	kg	W/m²K	kW	kg/s
1st St. Compressor Discharge Cooler	112.4	25	5	20	47.3	258.5	1.6	617	408	195	1.04
2nd St. Compressor Discharge Cooler	163.5	30	5	20	67.8	420.5	3.8	2,310	408	1,142	3.11
1st St. HP Compressor Cooler	63.3	30	5	20	33.3	796.5	13.1	16,985	408	4,538	46.65
2nd St. HP Compressor Cooler	71.4	50	5	20	48.1	646.7	8.2	8,491	408	3,162	46.65

Table 5.6: Heat exchanger design parameters with PR EoS

HEAT EXCHANGER	HOT: T1	HOT: T2	COLD: T1	COLD: T2	LMTD	Shell Diameter	Footprint	Weight	Overall U	Duty	Massflow
	°C	°C	°C	°C	K	mm	m^2	kg	W/m2-K	kW	kg/s
1st St. Compressor Discharge Cooler	112.8	25	5	20	47.4	252.5	1.6	590	408	186	0.98
2nd St. Compressor Discharge Cooler	163.5	30	5	20	67.8	406.5	3.6	2,155	408	1,060	2.90
1st St. HP Compressor Cooler	62.9	30	5	20	33.2	792.7	13.0	16,807	408	4,475	46.78
2nd St. HP Compressor Cooler	70.9	50	5	20	47.9	637.1	8.1	8,238	408	3,048	46.78

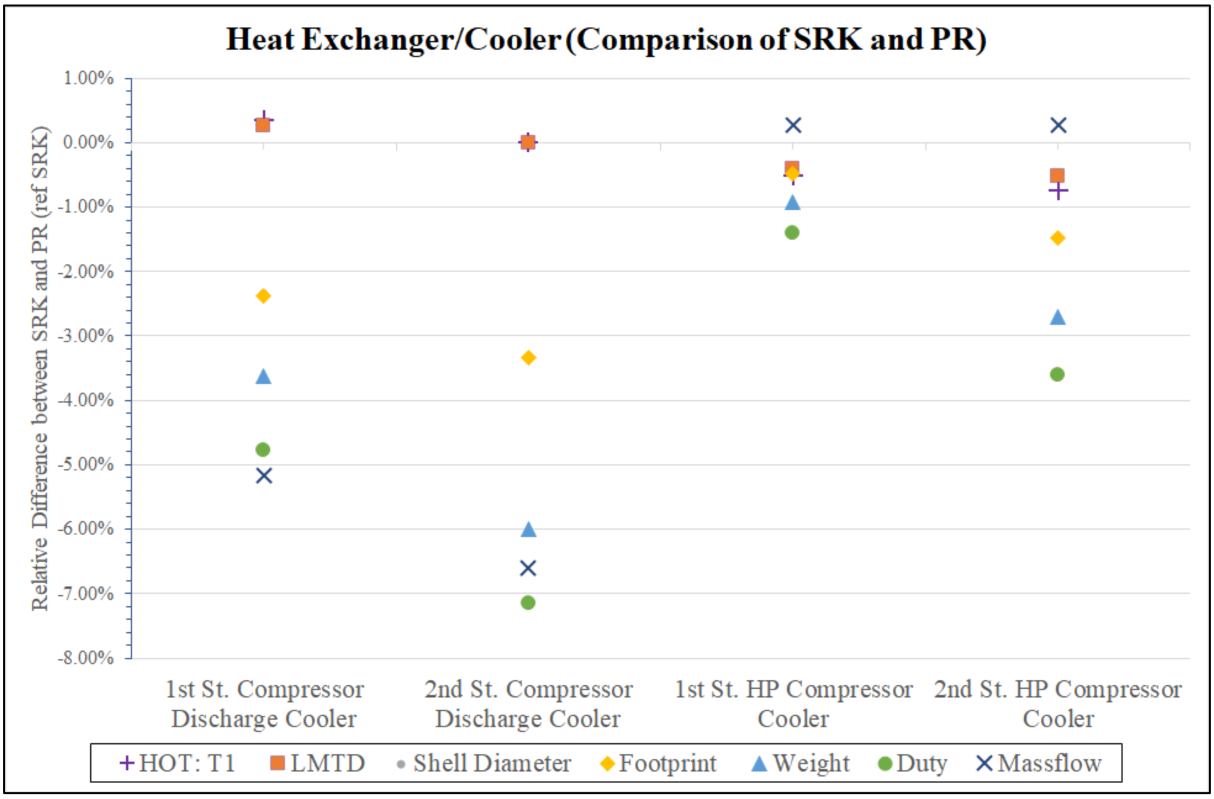


Figure 5.5: Heat exchanger design comparison between SRK and PR EoS

5.1.2.1 Stepwise procedure for developing the Heat Exchanger calculator

Input parameters for the heat exchanger are obtained from the HYSYS model.

1. The input parameters for both the tube and shell side are *density of fluid, mass flow rate* (tube fluid only), *specific heat capacity, temperature in, temperature out.*
2. For the purpose of the project the film transfer coefficients, h_f, and fouling factors, R_f, are obtained from empirical data from the *Heat Exchanger Design Handbook* highlighted in *Appendix D.3*
3. Determine the duty, Q, from *equation 3.96* which is equal on the tube and shell side. Based on this the *mass flow* of fluid on the shell side can be determined.
4. Calculate the corrected logarithmic temperature difference, $C - LMTD$, from *equation 5.170*
5. Select the tube parameters from *Appendix D.2. ie Wall thickness, OD, ID.*
6. Select the tube length based on step 7.
7. Assumptions made for the design are highlighted below;
    a. Tube length, L, for the 1st stage compressor discharge cooler set to 2m and shell side pressure rating set to 15 bar to meet TEMA and NORSOK minimum plate thickness.
    b. Tube length, L, for the 2nd stage compressor discharge cooler set to 3m and shell side pressure rating set to 15 bar to meet TEMA and NORSOK minimum plate thickness.
    c. Tube length, L, for the 1st stage HP compressor cooler set to 5m and shell side pressure rating set to 15 bar to meet TEMA and NORSOK minimum plate thickness.
    d. Tube length, L, for the 2nd stage HP compressor cooler (Export Cooler) set to 4m and shell side pressure rating set to 15 bar to meet TEMA and NORSOK minimum plate thickness.
    e. This is to ensure a length to diameter (L/D) ratio of 8-10.
8. The length of the heat exchanger is the tube length plus the head length.
9. Select the number of tube passes. (*One-pass* countercurrent assumed within this thesis).
10. The pitch ratio PR is assumed to be **1.25**. Determine the Tube Pitch, P_t, from *equation 3.103*
11. Determine the *cross-section area*, A_{cs}, of the tube knowing the ID.($A_{cs} = \pi\, ID^2/4$) and the area of a single tube , $A_{ST} = (2\,\pi\, ID/2)L$
12. Determine overall heat transfer co-efficient, U, from *equation 3.100. and the* total transfer area, A, from *equation 3.98.*
13. Calculate the number of tubes knowing the total area and area of a single tube (A/A_{ST})
14. Determine the fluid velocity per pass from *equation 5.171* and adjust the tube size or length to obtain the optimum fluid velocity.
15. Select the required tube pattern; ***Triangular*** or ***Square.*** Calculate the tube pattern area based on *equation 3.105*.
16. Determine the area of total tube bundle from *equation 3.106.*
17. Calculate the minimum shell diameter from *equation 3.107. As in step* 7 the length to diameter ratio to be approximately equal to 8.

Weight Calculation

18. The weight of the heat exchanger is determined from correlation for vessels (shell weight /separator weight), the weight of the tubes and the weight of the internals (baffles).
19. The number of baffle plates must be determined to determine the weight. This is obtained from *equation 3.108*. The baffle cut window (window height to ID - optimum between 25-35%) is assumed to be 30%. The baffle spacing is usually between 40-60% of the ID. This is assumed to be 50%. The weight of the baffle is determined from *equation 5.172*.
20. The tube weight is determined from *Appendix D.2* knowing the total number of tubes and the weight per meter.
21. The empty vessel weight of the heat exchanger is determined by adding the tube weight, baffle weight, flange weight and head weights.
22. For the total weight of the skid and vessel; the skid weights are determined from *equation 3.95.*

EQUATIONS

$Q = \dot{m}C_{pc}(T_{c,o} - T_{c,i}) = \dot{m}C_{p,h}(T_{h,i} - T_{h,o}) kW$	$Corr\ LMTD = \frac{\Delta T_1 - \Delta T_2}{ln\frac{\Delta T_1}{\Delta T_2}}.F$
(3.96)	(5.170)
$\frac{P_t}{OD} = 1.25$	$\frac{1}{U} = \left(\frac{1}{h_{gas}} + R_{f,gas}\right)\frac{d_o}{d_i} + \frac{d_o ln(d_o/d_i)}{k_w} + \left(\frac{1}{h_{seawater}} + R_{f,seawater}\right)$
(3.103)	(3.100)
$Q = UA(LMTD).F\ kW$	$Tot\ Tube\ area\ per\ pass = \frac{N_{tubes}}{A_{cs}}$ $Volumetric\ Flow\ (tube) = \frac{\dot{m}}{\rho}$ $Velocity = \frac{Volumetric\ flow}{Tube\ area}$
(3.98)	(5.171)
$Area_{tube,\ triangular} = 2\ (PRd_o)^2\frac{\sqrt{3}}{4}\ m^2$ (triangular) $Area_{tube,\ triangular} = (PRd_o)^2$ (Square)	$D_{tight} = 2\left(\frac{N_T Area_{tube}}{\pi}\right)^{0.5}\ m$
(3.105)	(3.104)

$A_{corrected} = D_{tight}\, d_o(n_p - 1) + (N_T Area_{tube})$	$D_{s,min} = 2\left(\frac{A_{corrected}}{\pi}\right)^{0.5} + 2d_o$
(3.106)	(3.107)
$N_b = \frac{L - L_{b,i} - L_{b,o}}{L_{b,c}} + 1$ $L_{b,c} = 50\%\ of\ D_{s,min}$ $L_{b,o} = L_{b,i} = 1.1\, L_{b,c}$	$Weight\ baffle = N_b\,(1 - 0.3) * \pi * L * \frac{(0.85*ID)^2}{4} * t * \rho$ $t - wall\ thickness$ $\rho - density\ of\ steel$
(3.108)	(5.172)
$W_b = 3.47\ dt\ kg/m$	
(3.92)	
$W_p = 0.4 * Weight\ of\ Empty\ Vessel$ (Weight of Piping) $W_E = 0.08 * Weight\ of\ Empty\ Vessel$ (Weight of Electrical and Instrument) $W_S = 0.1 * Weight\ of\ Empty\ Vessel$ (Weight of Skid steel) $Total\ weight = W_p + W_E + W_S\ kg$	
(5.157)	

5.1.3 Compressor System

The compressor calculator developed shows the sizing parameters used for the offshore centrifugal compressors. In determining the performance parameters, the stream parameters including the suction and required discharge pressures are required to determine parameters such as polytropic head, polytropic efficiency and total head.

As compressor design is supplier specific and trademarked, references from supplier equipment were used in determining the weight and footprint calculations. The ‘Frame type’ compressors from a supplier, Elliot Company, were used in defining weight and footprint. The supplier specifications are given under *Appendix E.1*.

The compressors utilised in the offshore platform can be divided into two. The compressors used in the liquid removal or cooling and separation process which are the *Low Pressure compressors (LP)* and the compressors required to meet export specifications termed *High Pressure compressors (HP).*

Table 5.7 and *Table 5.8* depict the performance parameters for the two main categories of compressors utilising the SRK and PR EoS.

It is observed from *Figure 5.5* (deviation from SRK) that PR EoS gives lower predictions of volumetric flowrates and power than SRK; however higher predictions of polytropic efficiency (except in the 2nd Stage LP Compressor). As information on compressor design is manufacturer specific, information from Elliot was used to obtain compressor configuration in relation to pressure limits, footprint and weight. Hence, information on weight and footprint is not represented for the different designs from the two EoS.

Table 5.7: Compressor design parameters with SRK EoS

COMPRESSOR	Inlet Flow Rate	P1	P2	T1	T2	Polytropic Efficiency	Power	Footprint	Weight	Frame 10 Configuration
	m^3/s	Bar	Bar	°C	°C	%	kW	m^2	kg	
1st st. LP Compressor	0.52	1.01	8.8	20.0	112.4	76	167	1.2	3,105	Horizontal
2nd st. LP Compressor	0.25	8.8	77	24.3	163.5	79	695	1.2	3,105	Vertical
1st st. HP Compressor	0.63	77	120	25.1	63.3	73	3,080	1.2	3,105	Vertical
2nd st. HP Compressor	0.39	120	200	30.0	71.4	76	3,547	1.2	3,105	Vertical

Table 5.8: Compressor design parameters with PR EoS

COMPRESSOR	Inlet Flow Rate	P1	P2	T1	T2	Polytropic Efficiency	Power	Footprint	Weight	Frame 10 Configuration
	m^3/s	Bar	Bar	°C	°C	%	W	m^2	kg	
1st st. LP Compressor	0.50	1.01	8.8	20.1	112.8	77	159	1.2	3,105	Horizontal
2nd st. LP Compressor	0.23	8.8	77	24.1	163.5	79	643	1.2	3,105	Vertical
1st st. HP Compressor	0.60	77	120	25.1	63.0	76	2,946	1.2	3,105	Vertical
2nd st. HP Compressor	0.37	120	200	30.0	70.9	77	3,321	1.2	3,105	Vertical

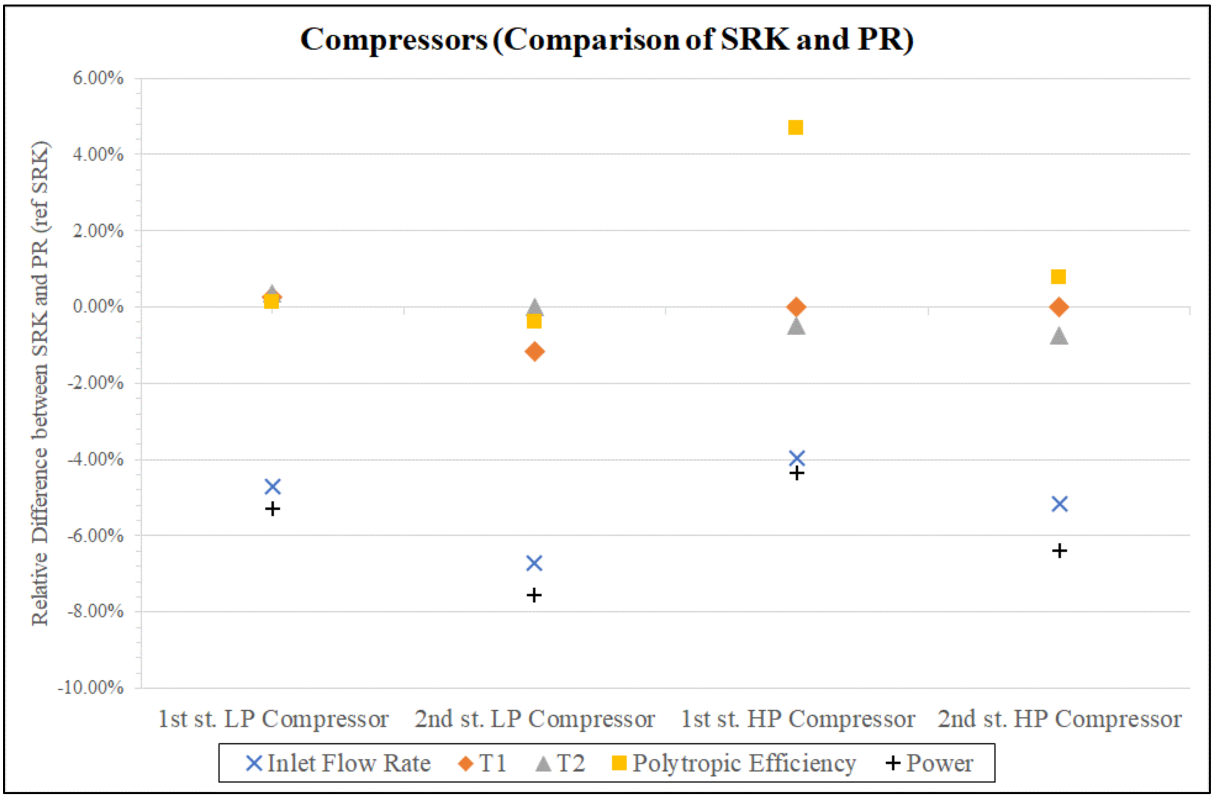


Figure 5.6: Compressor design comparison between SRK and PR EoS

5.1.3.1 Stepwise procedure for developing the Compressor calculator

The calculator is developed in parallel with ASPEN HYSYS model. Input parameters for the heat exchanger are obtained from the HYSYS model.

1. The input parameters for both the suction and discharge side are pressure, temperature, density of fluid, specific volume and enthalpy. Also, additional input parameters required on the suction side are flow rate, compressibility factor and molecular weight.
2. Determine the *polytropic exponent* from *equation 3.117*.
3. Determine the *pressure ratio* and calculate the *polytropic head*, H_p, from *equation 3.114* assuming the *correction factor*, f ,is 1.0.
4. Determine the *polytropic efficiency* from *equation 3.173.*
5. Determine the *total head*, H, from *equation 3.119.*
6. Determine the *Power* of the compressor from *equation 3.120.*

Weight and Footprint

The weight and footprint of the compressor was determined from data from manufacturer to be able to obtain as accurate result as possible. The compressor utilised was a frame 10 in either vertical or horizontal configuration as depicted under *Appendix E.1*.

EQUATIONS

$$n_v = \frac{ln\left(\frac{p_2}{p_1}\right)}{ln\left(\frac{v_1}{v_2}\right)}$$	$$H_p = f\ \frac{n_v}{n_v - 1}\ \frac{Z_1 R_o T_1}{MW}\left[\left(\frac{p_2}{p_1}\right)^{\frac{n_v-1}{n_v}} - 1\right]$$
(3.117)	(3.114)
$$\eta_p = \frac{H_p}{(h_2 - h_1)}$$	$$H = \frac{H_p}{\eta_p}$$
(3.173)	(3.119)
$$P_{fluid} = \dot{m}\,H = \rho_1 Q_1 H$$	
(3.120)	

5.1.4 Piping

The piping analysis done for the offshore processing plant takes into account sizing of the gas, liquid or multiphase line. The analysis focusses on designing the pipeline to within a velocity below the erosional velocity and/or to prevent liquid fallout using industry standards. The analysis does not take into account modelling using different thermodynamic models.

The wall thickness for the different pipe configurations are determined based on pressure rating and the required international standards for gas processing, specifically; ASME/ASTM and API standards. The wall thickness of the pipe is evaluated in the calculator based on different pipe codes. This comprises;

- ANSI/ASME Standard B31.1 Power Piping
- ANSI/ASME Standard B31.3 Chemical plant and Petroleum Refinery Piping
- ANSI/ASME Standard B31.4 Liquid Transportation Systems for Hydrocarbons, Liquid Petroleum Gas, Anhydrous Ammonia, and Alcohols. This standard applies to onshore oil pipeline facilities.
- ANSI/ASME Standard B31.8 Gas Transmission and Distribution Piping Systems. This standard applies to gas transmission, gathering, and distribution pipelines onshore.

In addition to these standards, the NORSOK standard for optimum pipeline specifications was utilised as highlighted in *section 3.4.3*. This provided the benchmark in defining the production rates for maximum velocity in both gas and liquid lines. The optimum maximum velocity data for the pipeline system under the simulation is captured in *Appendix G.2*. The input data; density and volume flowrate was taken from the HYSYS model.

Appendix G represents the alternative design calculations for *piping* using the ANSI/ASME standards. The calculator was developed to determine the optimum sizing for the pipeline inside diameter based on erosional velocity as a limitation. Also the wall thickness and the weight of the pipeline was determined based on the different ANSI/ASME codes which is dependent on the application.

5.1.5 *Evaluation Discussion*

The equipment evaluation looks at the comparison of the different calculators developed in parallel with the HYSYS model. The analysis is performed with different thermodynamic models which generate different input PVT fluid parameters from HYSYS. Based on these input PVT fluid parameters (such as density, temperature, viscosity, pressure etc) a detailed study is done to investigate the effect of the different thermodynamic models on equipment sizing.

The differences in the calculator arises out of the differences in thermodynamic models and the methods in characterising the reservoir fluids. As mentioned earlier, there could exist substantial liquid volumetric predictions difference between the SRK and PR EoS as well as the fact that PR EoS underpredicts saturation pressure of reservoir fluids in comparison to the SRK EoS hence requiring a somewhat larger hydrocarbon/hydrocarbon (C_1/C_7+) binary interaction parameters (BIP) for PR EoS (Whitson et al., 2000). There has also been some evidence that PR EoS gives slightly better performance around the critical point, making this EoS better suited for gas/condensate systems (Robinson et al., 1985).

These differences in fluid parameters arising from using different equations of state in designing processing equipment demands accurate predictions of fluid characterisation. As seen earlier in the chapter, notable differences in sizing parameters to design *Separators*, *Heat Exchangers* and *Compressors* evidently impact volume, weight and footprint which in turn reflect in the CAPEX and OPEX both in the preliminary design and operational phase.

From the analysis as shown in *Figure 5.7*, the total weight and footprint for the processing equipment was approximately 308 tons and 225 m^2 utilising the Soave-Redlich-Kwong Equation of State and 298 tons and 221 m^2 for the Peng Robinson Equation of State.

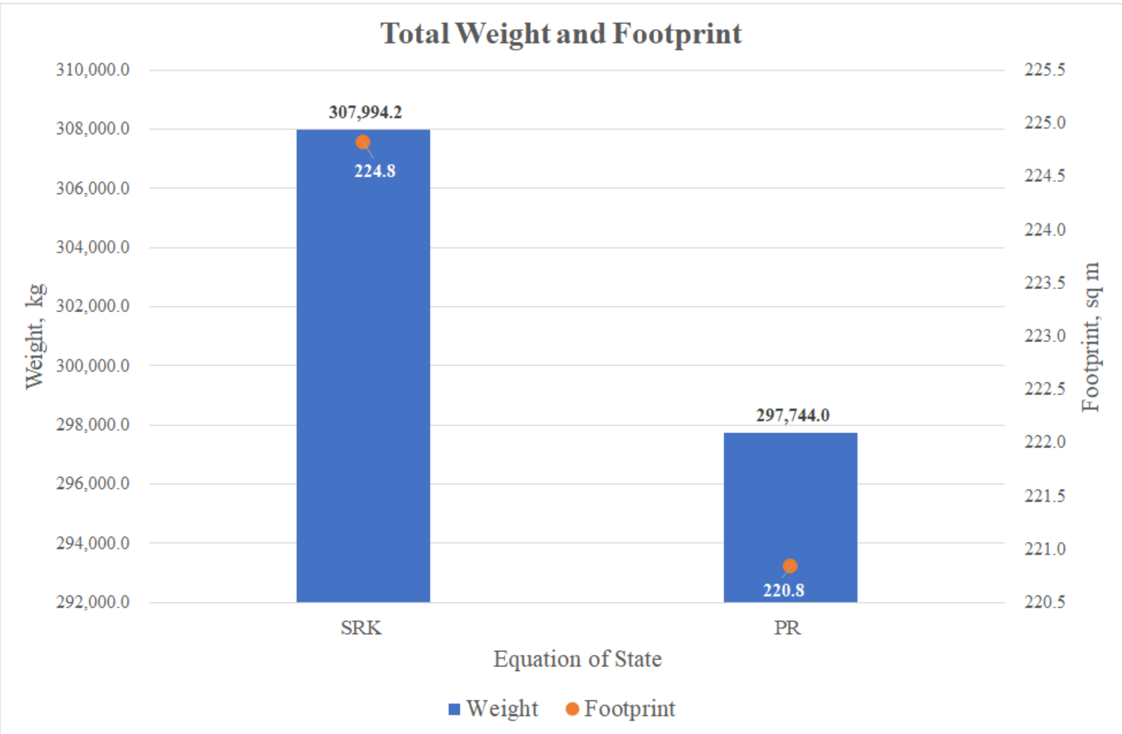


Figure 5.7: Total weight and footprint of processing equipment

The contribution of weight and footprint of the different equipment in the process plant are captured under *Figure 5.8* and *Figure 5.9*.

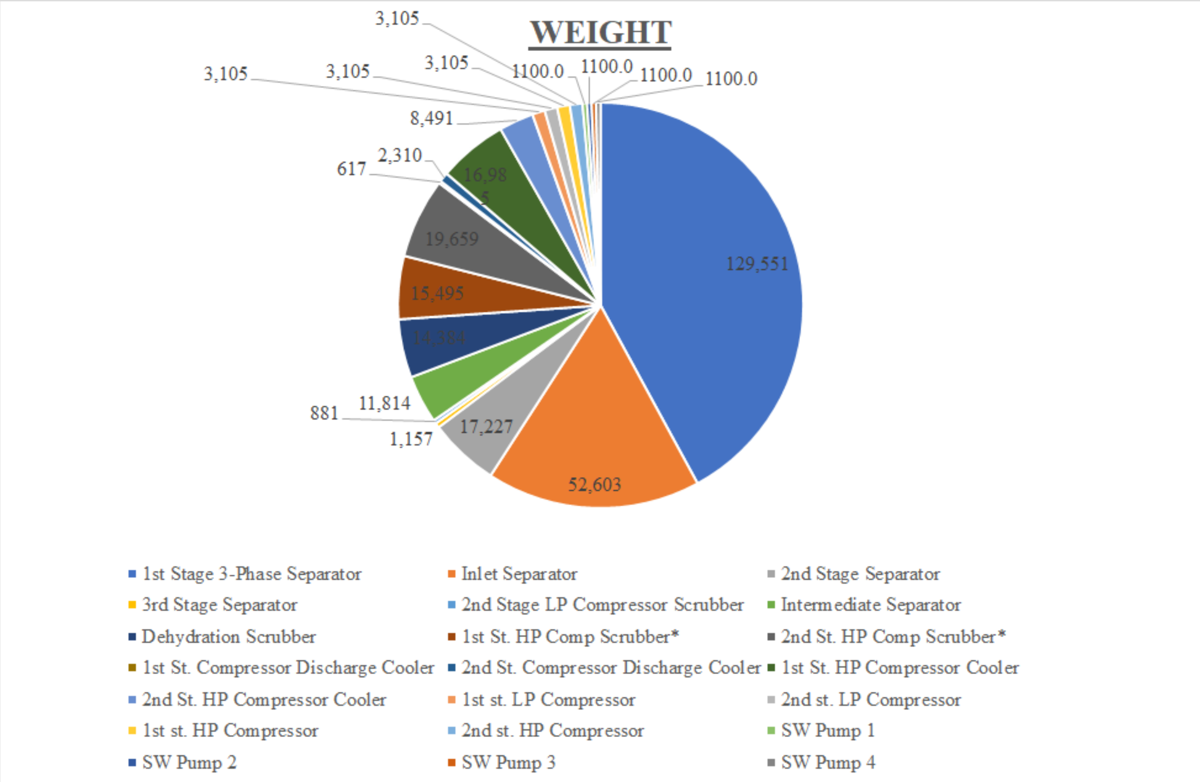


Figure 5.8: Weight contribution for different processing equipment (SRK)

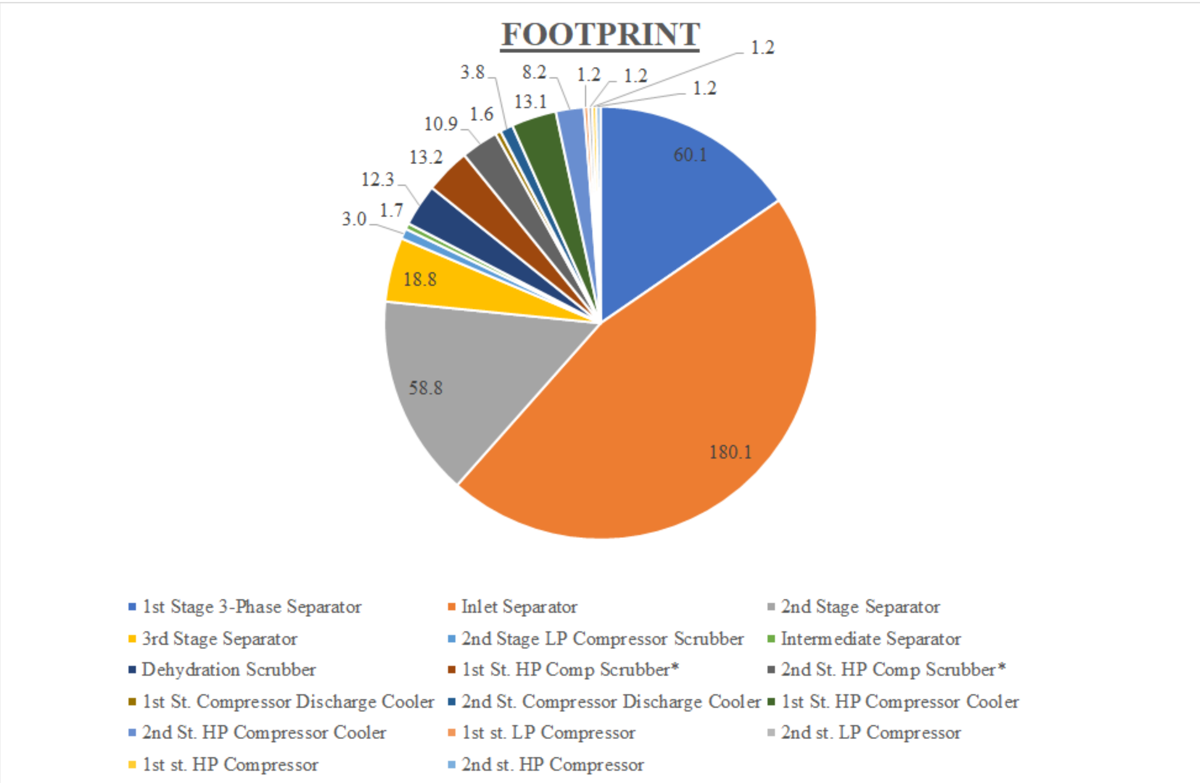


Figure 5.9: Footprint contribution for different processing equipment (SRK)

The results highlighted correspond to the base case well parameters given under *Table 1.1* and *Table 1.2*. A further evaluation performed comparing the calculators developed to the ASPEN HYSYS sizing models was done although this was not a major focus area. This is highlighted under *section 7.1.4*. Expected differences in output results of sizing between HYSYS model and the calculators were observed due to varied methods and empirical constants utilised. The following factors touch on a few of such differences;

Sizing Constant (K)

The sizing constant or empirical factor, K, within the calculator developed for both 2-phase and 3-phase separators were assumed based on API / GPSA standards as highlighted in *section 3.1*.

Holdup and Surge Times

Within the calculator developed, holdup and surge times were selected based on the service from *Appendix C.7*. Within the calculator the holdup-time of 5 minutes and surge time of 3 minutes was used.

Wall thickness

The wall thickness of the vessels both for the separator and heat exchanger as well as the pipeline were determined based on material of the vessel, the grade, the operating pressure of the conduit, working pressure of the material, the joint efficiency type and corrosion allowance. These calculations are taken from API and ASME standards as these differ based n ASME VIII Division I and Division 2 codes.

Heat Exchanger optimal design

The design of the heat exchanger involved various combinations of tube and shell dimensions. The optimal design was based on a length-to-diameter ratio of ~8-10 and a pitch ratio of 1.25 with a triangular tube pattern. An assumed and constant overall heat transfer co-efficient was set for the design. The internal baffle design specifically the cut and baffle spacing was taken to be 30% and 50% respectively and the clearance between baffles and shell taken as 85%.

6 Automation of Calculator with HYSYS.

The design calculators developed for each equipment was done independently of HYSYS equipment modelling. Inputs only were taken from HYSYS in order to obtain a detailed design analysis for each offshore equipment as explained in *Chapter 5*.

Further on, in order to perform any meaningful plant sensitivity analysis based on changing input parameters; be it from change in flowrate, pressure, temperature or to perform some economic analysis, it is imperative that interaction between ASPEN HYSYS and MS Excel (program used to develop the equipment calculators) is established. This was done utilising the

ASPEN Simulation Workbook (ASW) version 9. The workbook provided;

- an efficient user interface between HYSYS and MS Excel equipment design models
- a method to eliminate the need for writing lengthy programming code
- an interface for scenario study for process sensitivity analysis

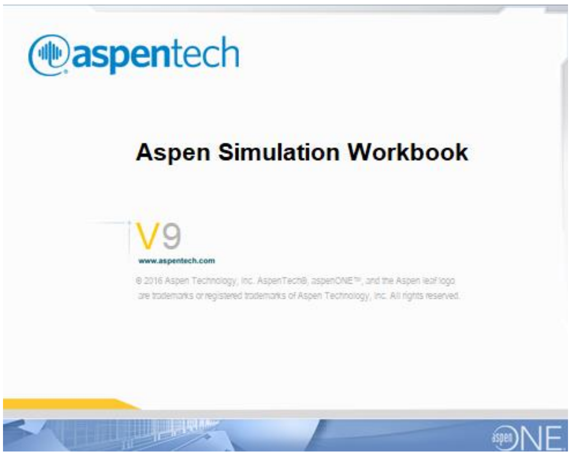


Figure 6.1: ASPEN simulation workbook

With such a tool various analysis could be performed to evaluate the impact of changes in life-of-field parameters to equipment size. This master thesis focuses on the change in annual production flowrates, with every other parameter being equal, and the resulting impact on equipment sizing to determine the optimal design for the processing plant. This is done by assuming a scenario of different production profiles during the plant life.

6.1.1 Creating a Scenario

Within MS Excel, the Add-in for Aspen Simulation Workbook must be enabled to activate the workbook. This is ***"Enabled"*** on the ASW ribbon. The simulation case of the plant design is then loaded - "***Connect"*** - to complete the interface between ASPEN HYSYS and MS Excel. (depicted by the '*red'* markings as shown in *Figure 6.2*)

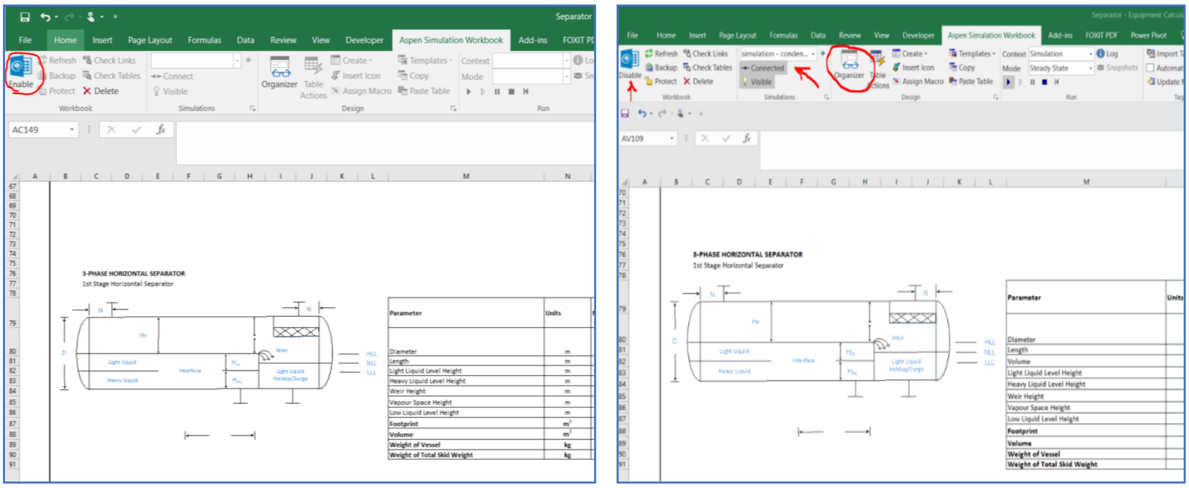
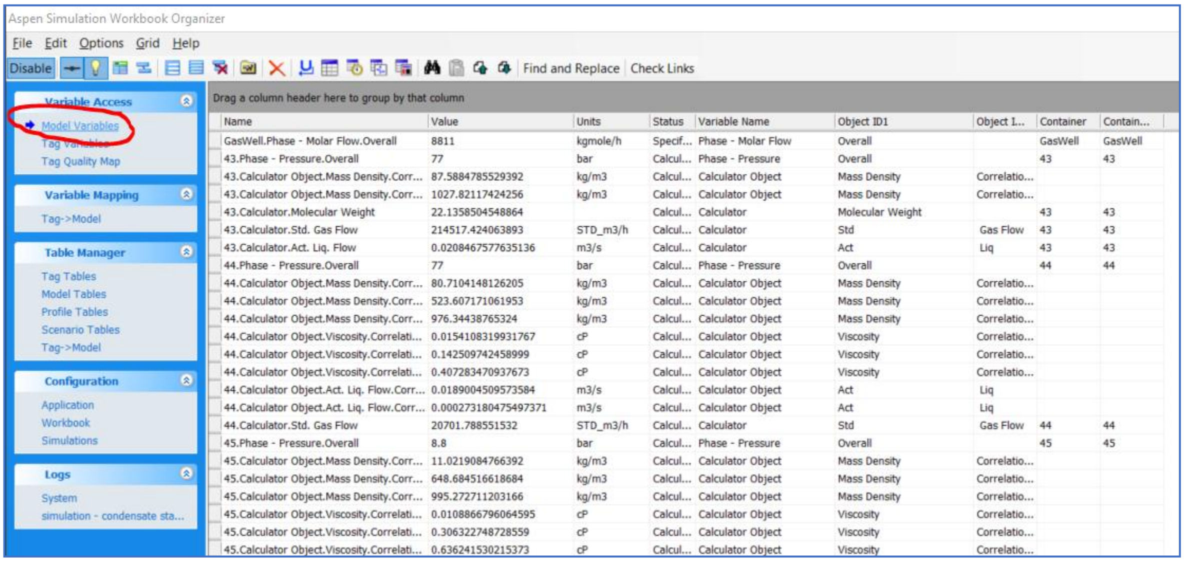


Figure 6.2: Aspen Simulation Workbook ribbon in Excel

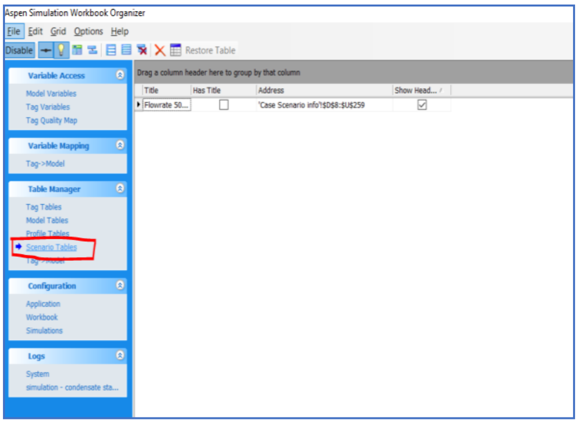
With the incorporated functionality of the workbook, various scenarios could be run from the HYSYS simulation model by stating the "***Model Variables***" via the simulation workbook ***"Organizer"*** as shown in *Figure 6.2* and *Figure 6.3.*



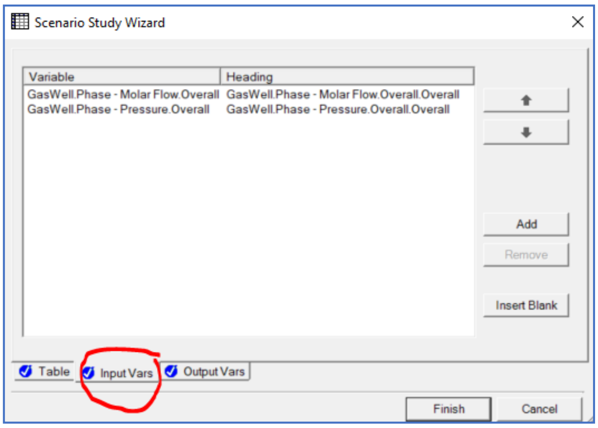
Name	Value	Units	Status	Variable Name	Object ID1	Object 1...	Container	Contain...
GasWell.Phase - Molar Flow.Overall	8811	kgmole/h	Specif...	Phase - Molar Flow	Overall		GasWell	GasWell
43.Phase - Pressure.Overall	77	bar	Calcul...	Phase - Pressure	Overall		43	43
43.Calculator Object.Mass Density.Corr...	87.5884785529392	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
43.Calculator Object.Mass Density.Corr...	1027.82117424256	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
43.Calculator.Molecular Weight	22.1358504548864		Calcul...	Calculator	Molecular Weight		43	43
43.Calculator.Std. Gas Flow	214517.424063893	STD_m3/h	Calcul...	Calculator	Std	Gas Flow	43	43
43.Calculator.Act. Liq. Flow	0.0208467577635136	m3/s	Calcul...	Calculator	Act	Liq	43	43
44.Phase - Pressure.Overall	77	bar	Calcul...	Phase - Pressure	Overall		44	44
44.Calculator Object.Mass Density.Corr...	80.7104148126205	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
44.Calculator Object.Mass Density.Corr...	523.607171061953	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
44.Calculator Object.Mass Density.Corr...	976.34438765324	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
44.Calculator Object.Viscosity.Correlati...	0.0154108319931767	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		
44.Calculator Object.Viscosity.Correlati...	0.142509742458999	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		
44.Calculator Object.Viscosity.Correlati...	0.407283470937673	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		
44.Calculator Object.Act. Liq. Flow.Corr...	0.0189004509573584	m3/s	Calcul...	Calculator Object	Act	Liq		
44.Calculator Object.Act. Liq. Flow.Corr...	0.000273180475497371	m3/s	Calcul...	Calculator Object	Act	Liq		
44.Calculator.Std. Gas Flow	20701.788551532	STD_m3/h	Calcul...	Calculator	Std	Gas Flow	44	44
45.Phase - Pressure.Overall	8.8	bar	Calcul...	Phase - Pressure	Overall		45	45
45.Calculator Object.Mass Density.Corr...	11.0219084766392	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
45.Calculator Object.Mass Density.Corr...	648.684516618684	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
45.Calculator Object.Mass Density.Corr...	995.272711203166	kg/m3	Calcul...	Calculator Object	Mass Density	Correlatio...		
45.Calculator Object.Viscosity.Correlati...	0.0108866796064595	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		
45.Calculator Object.Viscosity.Correlati...	0.306322748728559	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		
45.Calculator Object.Viscosity.Correlati...	0.636241530215373	cP	Calcul...	Calculator Object	Viscosity	Correlatio...		

Figure 6.3: ASW model variables

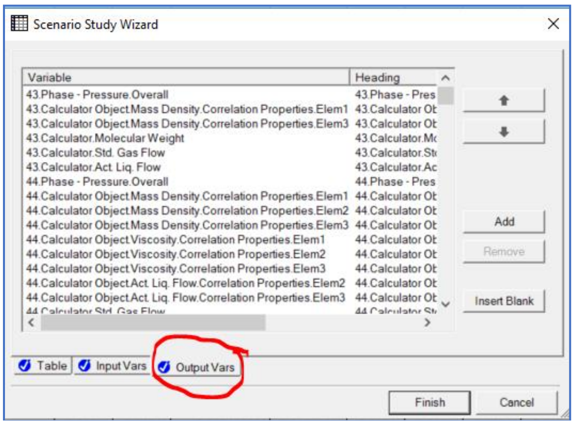
Within the *Organizer*, input and output parameters to run any sensitivity analysis can be defined under a scenario and multiple cases can be run for different input parameters. The multiple cases are run to generate the output results defined by the organizer under the ***"Scenario Study wizard"*** as depicted in *Figure 6.4.* The number of cases to be run are also defined under the study wizard.



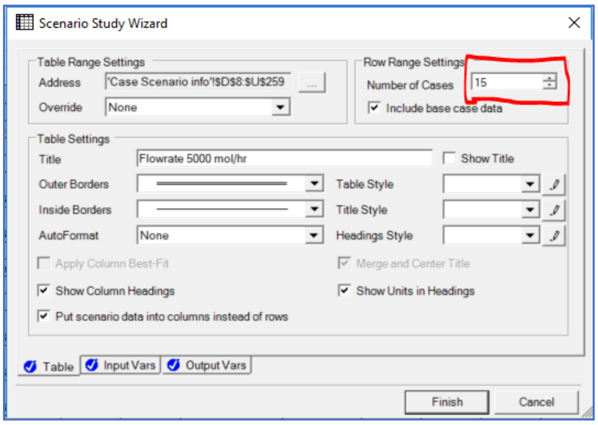
Creating Scenario table



Defining the input variables



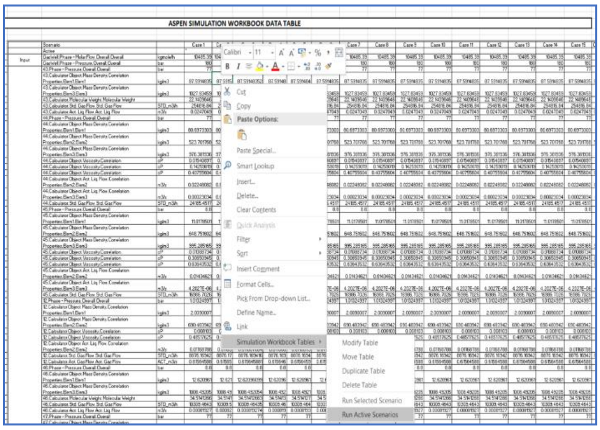
Defining the output variables



Defining the number of cases in a scenario

Figure 6.4: ASW organizer setup and scenario study wizard

The main parameters are created under the scenario table in excel. *Specific* or *multiple cases* can be run as shown in *Figure 6.5*. The outputs from the table are fed as inputs to the calculator to obtain the equipment plant design for each case. *Appendix K.1* gives an example of a complete scenario table with fifteen cases.



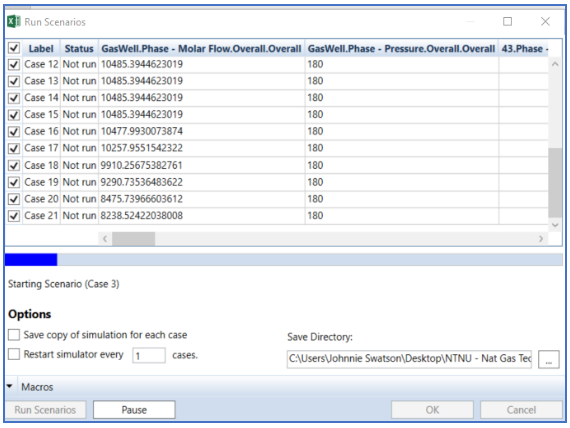


Figure 6.5: Running multiple cases under scenario table

6.1.2 Design Output Representation

Both the input and output parameters as defined under the Scenario Study Wizard are listed in the excel workbook when the "***Scenario Table***" is created. The different cases can be run altogether or selectively to generate the output parameters defined in the *scenario study wizard*. The *output parameters* produced are the *input parameters* for the equipment calculator sheets. These automatically generate the sizing, design and performance parameters of the separator, heat exchanger, compressor, pump and various piping.

Within this master thesis three different representations of the equipment design output have been presented;

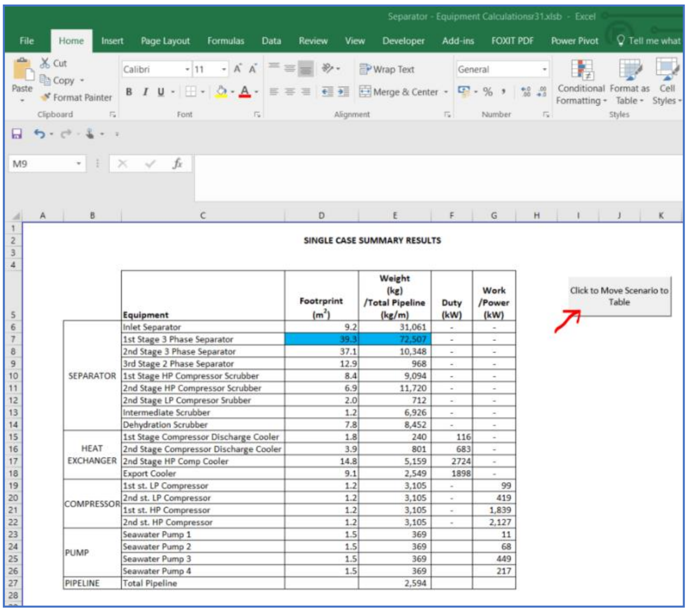
i. 2-D Graphical Layout of the plant by sections
ii. Single Case Summary
iii. Scenario Study Summary (which incorporates various cases)

This is shown under *Appendix K.4, Figure 6.6* and *Appendix K.5* respectively.

6.1.3 Scenario Study Recording

The output interface of the equipment calculator and ASW have been developed such that it gives the output design for the plant equipment based on a single case. In order to capture each single case and display the results, a macro was developed to record each single case output to generate a Scenario Study comprising different cases. Refer to *Appendix K.2* for the macro written to generate/record the sensitivity data.

The single case study is shown in *Figure 6.6*. Upon generating the single case study, the specific study is recorded and populated under the macro-enabled scenario by clicking the *"Click to Move Scenario To Table"* tab. The recorded data is populated under the *scenario study summary*. Refer to *Appendix K.5* for the fifteen case complete scenario generated for a production profile.



SINGLE CASE SUMMARY RESULTS

	Equipment	Footrprint (m^2)	Weight (kg) /Total Pipeline (kg/m)	Duty (kW)	Work /Power (kW)
SEPARATOR	Inlet Separator	9.2	31,061	-	-
	1st Stage 3 Phase Separator	39.3	72,507	-	-
	2nd Stage 3 Phase Separator	37.1	10,348	-	-
	3rd Stage 2 Phase Separator	12.9	968	-	-
	1st Stage HP Compressor Scrubber	8.4	9,094	-	-
	2nd Stage HP Compressor Scrubber	6.9	11,720	-	-
	2nd Stage LP Compresor Srubber	2.0	712	-	-
	Intermediate Scrubber	1.2	6,926	-	-
	Dehydration Scrubber	7.8	8,452	-	-
HEAT EXCHANGER	1st Stage Compressor Discharge Cooler	1.8	240	116	-
	2nd Stage Compressor Discharge Cooler	3.9	801	683	-
	2nd Stage HP Comp Cooler	14.8	5,159	2724	-
	Export Cooler	9.1	2,549	1898	-
COMPRESSOR	1st st. LP Compressor	1.2	3,105	-	99
	2nd st. LP Compressor	1.2	3,105	-	419
	1st st. HP Compressor	1.2	3,105	-	1,839
	2nd st. HP Compressor	1.2	3,105	-	2,127
PUMP	Seawater Pump 1	1.5	369		11
	Seawater Pump 2	1.5	369		68
	Seawater Pump 3	1.5	369		449
	Seawater Pump 4	1.5	369		217
PIPELINE	Total Pipeline		2,594		

Figure 6.6: Single case equipment summary

With a method to display a single case and record all cases an investigative analysis can be performed. The different cases can be set up to represent varied production flowrates for each year, varied fluid composition during the production lifecycle, pressure changes if any etc. In the case of the master thesis, an investigative analysis was performed using varied production flowrates during the lifecycle of the field. This is depicted under *Figure 6.7* where three (3) different production profile scenarios are considered and each case can be viewed and recorded with a drop-down selection (indicated by red arrow). The investigative analysis performed is detailed further under *Chapter 7.*

Scenario	Scenario 3

	Flowrate, q (sm3/d)		
Year (Cases)	Scenario 1	Scenario 2	Scenario 3
1	5000000	2000000	5000000
2	5000000	4000000	30000000
3	5000000	15000000	30000000
4	10000000	15000000	30000000
5	10000000	15000000	15000000
6	10000000	15000000	9000000
7	10000000	15000000	0
8	10000000	12000000	0
9	10000000	10000000	0
10	10000000	3000000	0
11	10000000	3000000	0
12	10000000	3000000	0
13	6000000	3000000	0
14	5000000	3000000	0
15	3000000	1000000	0

Case	Case 2

Figure 6.7: Case and scenario selection representing flowrate for each year

6.1.4 Flowchart for developing the Scenario Study

After setting up all equipment calculators for the offshore processing plant and having linked all the required output data from ASPEN HYSYS to the Equipment Calculators using ASPEN Simulation Workbook, the required scenario input data which would be the basis for the investigative analysis for the process can be defined.

Figure 6.8 gives a flowchart for developing the scenario study following the setup described in *sections 6.1.1* to *6.1.3*.

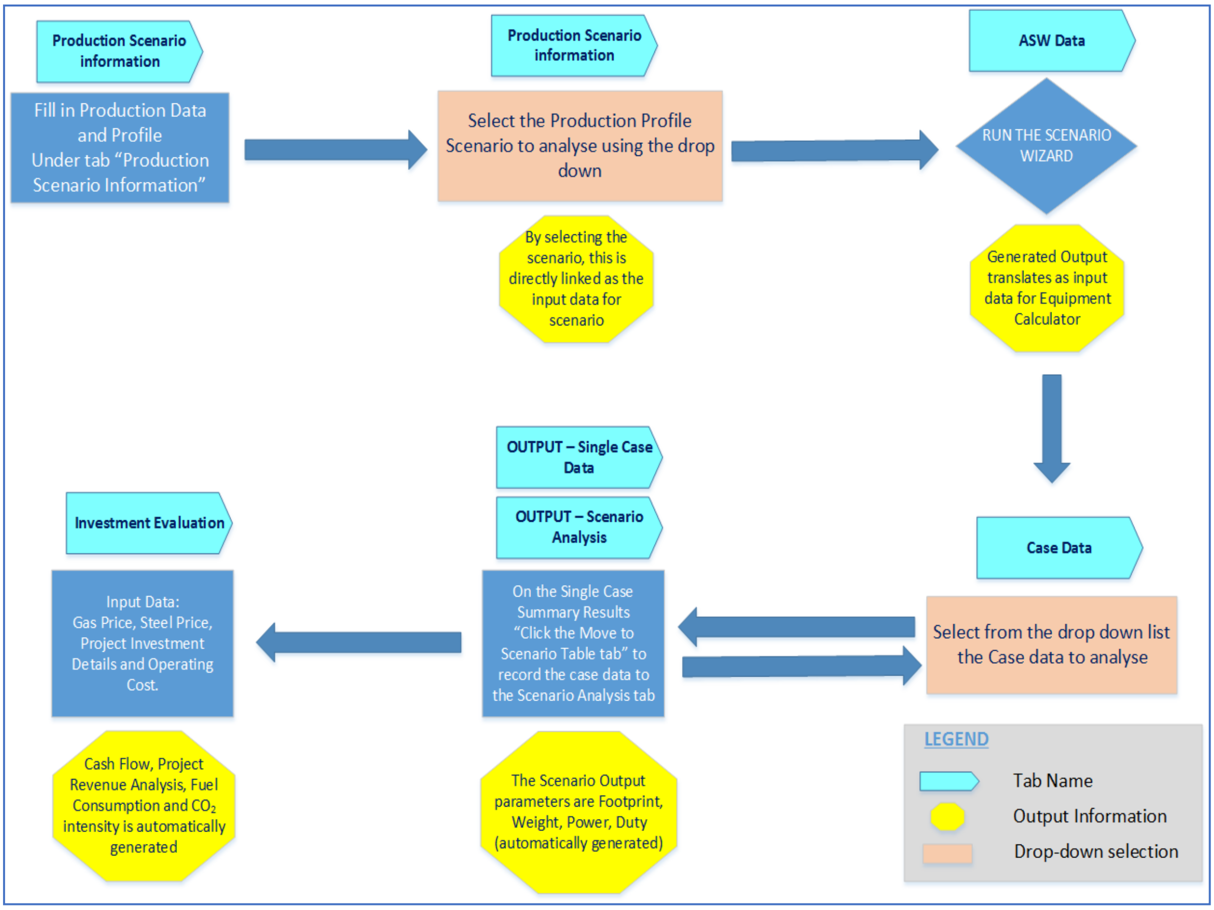


Figure 6.8: Flow chart for developing the scenario analysis

7 Investigative Analysis

After setting up the *Simulation model*, *Case analysis* and *Scenario study* various analysis pertaining to the project can be performed to ascertain optimum project deliverables. The analysis was performed to evaluate;

i. the optimum process equipment design based on a specific production profile for life of well.
ii. the carbon footprint of the process for the life of well.
iii. the profitability of the process plant based on a different scenario production profiles with subsequent effects on equipment design cost, carbon dioxide emissions and break-even analysis.
iv. comparison of the profitability of a project using different thermodynamic models keeping all other factors constant.

7.1.1 Equipment Design and Production profile

Equipment design varies based on different properties such as fluid composition, production flowrates etc. The production profile influences the design flowrates and operating parameters of equipment. For the scope of this master thesis, investigation in the change of the equipment design for the life of field is based on hypothetical three production profile scenarios as represented in *Figure 7.1* (with all scenarios giving the same total produced gas);

	Flowrate, q (sm^3/d)		
Year (Cases)	**Scenario 1**	**Scenario 2**	**Scenario 3**
1	5,000,000	2,000,000	5,000,000
2	5,000,000	4,000,000	30,000,000
3	5,000,000	15,000,000	30,000,000
4	10,000,000	15,000,000	30,000,000
5	10,000,000	15,000,000	15,000,000
6	10,000,000	15,000,000	9,000,000
7	10,000,000	15,000,000	
8	10,000,000	12,000,000	
9	10,000,000	10,000,000	
10	10,000,000	3,000,000	
11	10,000,000	3,000,000	
12	10,000,000	3,000,000	
13	6,000,000	3,000,000	
14	5,000,000	3,000,000	
15	3,000,000	1,000,000	

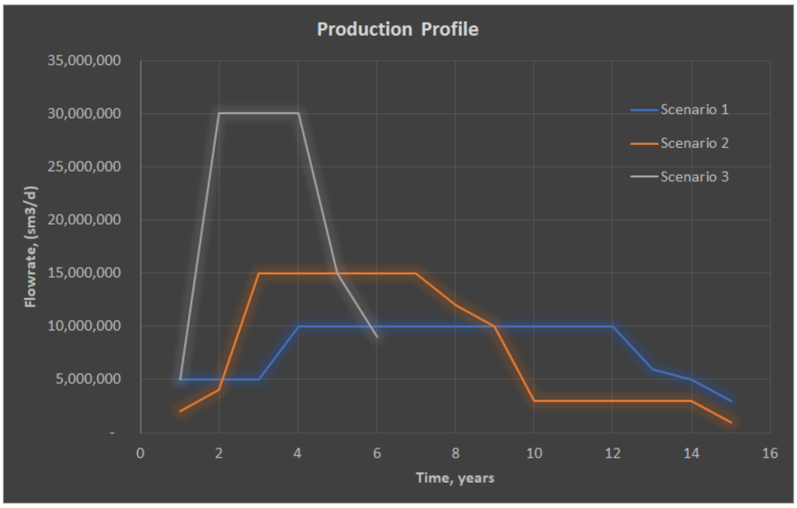


Figure 7.1: Scenario production profiles

A change in production flowrate gives rise to a change in design properties such as velocity, liquid and gas flowrates. This in turn affects the design of the process equipment; footprint, volume and weight as captured under *Chapters 3* and *5*.

The production profiles utilised are based on same gas volumes/reserves in place

Scenario 1	: Steady ramp up of production and longer production plateau and steady decline in production.
Scenario 2	: Steady ramp up of production and shorter production plateau and steady decline in production (ramp up and decline in production is steeper than scenario 1)
Scenario 3	: Steep ramp to high plateau for maximum production and sharp decline. This could be akin to extreme projects where it is desired to have maximum production at the earliest possible time.

As can be seen from *Figure 7.1*, the maximum equipment design/size will be defined by periods with corresponding high flow rates. This relates to year 4-12 for scenario 1, year 3-7 for scenario 2 and year 2-4 for scenario 3; as equipment would need to cater for high volumes within this period.

After running the scenarios from the model created, *Table 7.1* shows the process plant equipment design given the effect of change in flowrate along the process life of the plant for all three scenarios. The highlighted cells under *Table 7.1* give the values of the maximum parameters which correspond to the plateau/maximum production flowrates and would inform the design criteria for the plant.

Table 7.1: Scenario process plant design parameters

Scenario 1

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total Footprint (m^2)	224.8	224.8	224.8	396.5	396.5	396.5	396.5	396.5	396.5	396.5	396.5	396.5	260.0	224.9	150.5
Total Weight (tons)	307.9	307.9	307.9	635.2	635.2	635.2	635.2	635.2	635.2	635.2	635.2	635.2	369.9	308.1	188.7
Total Duty (MW)	9.03	9.03	9.03	18.07	18.07	18.07	18.07	18.07	18.07	18.07	18.07	18.07	10.84	9.03	5.42
Compressor Power (MW)	7.47	7.47	7.47	14.95	14.95	14.95	14.95	14.95	14.95	14.95	14.95	14.95	8.97	7.47	4.48
Daily Compressor Energy (MWh)	179.4	179.4	179.4	358.7	358.7	358.7	358.7	358.7	358.7	358.7	358.7	358.7	215.2	179.4	107.6
Pump Power (kW)	232.5	232.5	232.5	473.4	473.4	473.4	473.4	473.4	473.4	473.4	473.4	473.4	279.5	232.5	138.8
Condensate (bbl/d)	5,340	5,340	5,340	10,680	10,680	10,680	10,680	10,680	10,680	10,680	10,680	10,680	6,406	5,341	3,203

Scenario 2

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total Footprint (m^2)	111.3	188.5	558.9	558.9	558.9	558.9	558.9	461.9	396.1	150.6	150.6	150.6	150.6	150.6	68.2
Total Weight (tons)	132.8	247.7	992.9	992.9	992.9	992.9	992.9	774.2	634.7	188.8	188.8	188.8	188.8	188.8	79.2
Total Duty (MW)	3.6	7.2	27.1	27.1	27.1	27.1	27.1	21.7	18.1	5.4	5.4	5.4	5.4	5.4	1.8
Compressor Power (MW)	3.0	6.0	22.4	22.4	22.4	22.4	22.4	17.9	14.9	4.5	4.5	4.5	4.5	4.5	1.5
Daily Compressor Energy (MWh)	71.7	143.5	538.1	538.1	538.1	538.1	538.1	430.5	358.7	107.6	107.6	107.6	107.6	107.6	35.9
Pump Power (kW)	94.8	190.1	721.9	721.9	721.9	721.9	721.9	581.8	480.8	142.4	142.4	142.4	142.4	142.4	47.4
Condensate (bbl/d)	2,137	4,272	16,014	16,014	16,014	16,014	16,014	12,812	10,676	3,203	3,203	3,203	3,203	3,203	1,068

Scenario 3

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total Footprint (m^2)	224.8	1,022.1	1,022.1	1,022.1	559.0	362.9	-	-	-	-	-	-	-	-	-
Total Weight (tons)	308.0	2,198.6	2,198.6	2,198.6	993.0	565.74	-	-	-	-	-	-	-	-	-
Total Duty (MW)	9.03	54.21	54.21	54.21	27.10	16.26	-	-	-	-	-	-	-	-	-
Compressor Power (MW)	7.47	44.84	44.84	44.84	22.42	13.45	-	-	-	-	-	-	-	-	-
Daily Compressor Energy (MWh)	179.4	1,076.2	1,076.2	1,076.2	538.1	322.8	-	-	-	-	-	-	-	-	-
Pump Power (kW)	238.0	1,580.7	1,580.7	1,580.7	778.4	467.6	-	-	-	-	-	-	-	-	-
Condensate (bbl/d)	5,340	32,044	32,044	32,044	16,017	9,608	-	-	-	-	-	-	-	-	-

7.1.2 Carbon Footprint

Drivers such as gas turbines, in this case, are required in order to power the compressors on the offshore platform. In Norway, there exists a carbon tax which was introduced in 1991. This tax is levied on all combustion of gas, oil and diesel in petroleum operations on the Norwegian Continental Shelf (NCS) and on releases of carbon dioxide (CO_2) and natural gas which is in accordance with the *CO_2 Tax Act on Petroleum Activities.* The current tax rate (2018) at the time of writing this thesis is NOK 1.06 per standard cubic meters of gas (Norwegian Petroleum Directorate, 2018a). It is assumed that some of the processed gas would be utilised as fuel in the gas turbines to power the compressors.

For a gas turbine, as given in *equation 7.174*, the mass flowrate of fuel (Saravanamuttoo, 2009) is given as;

$$\dot{m}_f = \frac{W_N}{\eta_{GT}\, LHV} \tag{7.174}$$

where
$\dot{m}_f$ – Mass flowrate of fuel (kg/s)
W_N – Net work (kW), given as the difference in work of the turbine and compressor.
η_{GT} – Gas Turbine Efficiency/Cycle efficiency
LHV – Lower Heating Value of fuel (CH_4) which is 46,540 kJ/kg

For the ease of calculations as a gas turbine has not been modelled, an efficiency of 35% has been assumed. This efficiency is representative for gas turbine on the NCS. This indicates the percentage of heat supplied that translates into work for the compressor.

From the mass flowrate of fuel the volumetric flowrate, V_f, can be deduced and the CO_2 emissions cost per year determined from *equations 7.175* and *7.176*

$$V_f = \frac{\dot{m}_f R T_{sc}}{MW\, P_{sc}} \tag{7.175}$$

$$CO_2\ emissions\ cost/yr = CO_2\ Tax\ rate\ \times V_f\ \times 86400\ \frac{s}{day} \times operational\ days\ per\ year \tag{7.176}$$

where
V_f – Volumetric flowrate, m^3/s
R – Universal Gas Constant
T_{sc} – Temperature at standard conditions, 288.15 K
MW – Molecular Weight (CH_4) is 19.59
P_{sc} – Pressure at standard Conditions, 101,325Pa
Operational days set to 300 days per year to account for downtime and maintenance.

Table 7.2 shows the carbon footprint pertaining to the project for each year. The highlighted cell represents the maximum CO_2 emissions cost corresponding to the highest production rate as this would require increased fuel/power for the compression process.

One factor that is monitored with respect to carbon dioxide emissions on a project to project basis undertaken by Statoil and other companies on the NCS is the carbon intensity. This is measured as the weight of carbon dioxide per barrel of oil equivalent produced given as kilogram of CO_2 per BOE (barrel of oil equivalent). Statoil has set a strategy to create a low carbon advantage on the NCS. The current carbon intensity of projects in Norway is 9kg CO_2 per BOE. The target set out by Statoil by 2030 is 8kg CO_2 per BOE (Statoil ASA, 2017).

The carbon intensity is calculated as given below – assuming the fuel burned is processed methane from the plant;

For 1m^3 of Methane fuel ; Density of Methane = 0.657 kg/m^3
Methane (CH_4) contains = 12/16 = 75% of Carbon
Weight of Carbon = 75% of 0.657 kg = 0.4927 kg of Carbon per m^3 of methane

Assuming complete combustion of Carbon to carbon dioxide:

$$C + O_2 \xrightarrow{yields} CO_2$$

2.667kg of O_2 to 1kg of Carbon. This gives weight of O_2 to be 1.31416 kg to form CO_2.

Weight of CO_2 per m^3 of Methane = 0.4927 kg of Carbon + 1.31416 kg of O_2 = <u>1.8069 kg of CO_2 per m^3 of CH_4</u>

Knowing the amount of fuel consumed and given the conversion factor of 1m^3 of natural gas equals 0.00642857 BOE, the carbon intensity for the project can be obtained from *equation 7.177.*

$$Carbon\ Intensity\ (kg\ CO_2\ per\ BOE) = \frac{1.8069 \frac{kg\ CO_2}{m^3} \times V_f \times 86400 \frac{s}{day} \times \frac{300\ days}{yr}}{\Delta G_p \times 0.00642857\ \frac{BOE}{yr}} \tag{7.177}$$

Where ΔG_p is yearly produced gas and operational days is given as 300 days in a year to take into account downtime or maintenance days.

Table 7.2: Carbon footprint for scenario

Scenario 1

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Fuel Consumption (kg/s)	0.46	0.46	0.46	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.55	0.46	0.28
Fuel Consumption (sm^3/s)	0.55	0.55	0.55	1.11	1.11	1.11	1.11	1.11	1.11	1.11	1.11	1.11	0.67	0.55	0.33
CO_2 emissions cost per year (MM NOK /year)	15.24	15.24	15.24	30.49	30.49	30.49	30.49	30.49	30.49	30.49	30.49	30.49	18.29	15.24	9.15
CO_2 intensity (kg CO2 per BOE)	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70

Scenario 2

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Fuel Consumption (kg/s)	0.18	0.37	1.38	1.38	1.38	1.38	1.38	1.10	0.92	0.28	0.28	0.28	0.28	0.28	0.09
Fuel Consumption (sm^3/s)	0.22	0.44	1.66	1.66	1.66	1.66	1.66	1.33	1.11	0.33	0.33	0.33	0.33	0.33	0.11
CO_2 emissions cost per year (MM NOK /year)	6.10	12.19	45.73	45.73	45.73	45.73	45.73	36.58	30.49	9.15	9.15	9.15	9.15	9.15	3.05
CO_2 intensity (kg CO2 per BOE)	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70

Scenario 3

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Fuel Consumption (kg/s)	0.46	2.76	2.76	2.76	1.38	0.83	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fuel Consumption (sm^3/s)	0.55	3.33	3.33	3.33	1.66	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2 emissions cost per year (MM NOK /year)	15.24	91.46	91.46	91.46	45.73	27.44	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2 intensity (kg CO2 per BOE)	2.70	2.70	2.70	2.70	2.70	2.70	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

7.1.3 Cash Flow Analysis

For a project to be feasible, profitability must be determined. The revenues generated from the sale of gas and condensate as opposed to the cost associated with the project must be forecasted with its accompanying risk factors. Actual revenue depends not only on production rates but on current or contracted prices of oil and gas. Typically, the economics and risks associated with oil and gas field projects are ascertained using several price development scenarios. For the scope covered under this master thesis, the capital structure of the project has not been defined and assumed to be entirely financed internally without taking into account debt. *Figure 7.2* shows the revenue and cost factors in relation to the timeline for a typical gas project.

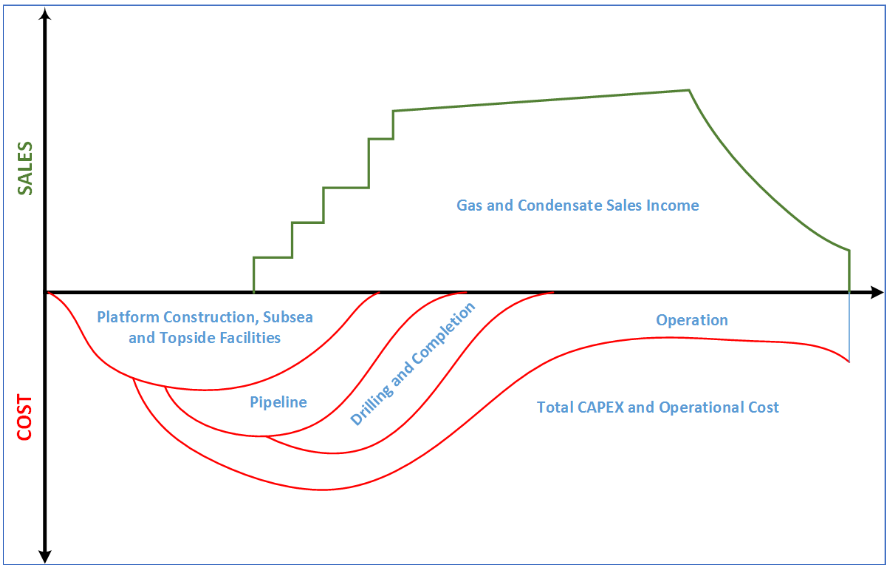


Figure 7.2: Typical gas field revenue and cost profiles

The accompanying risk factors mentioned are the assumed market price of risk, the *opportunity cost of capital*, which is the return that the market offers on investments with the same risk characteristics. (Wijst, 2013)

Revenue

The revenue generated from the project is the sale of 'dry' gas and condensate produced from the process plant. Forecasting price trends for the products is more uncertain than determining the gas field production profile. The market price of natural gas (NYMEX) as at the time of the master thesis is USD \$2.76 /MMBtu and USD \$64.94 per barrel (Bloomberg Energy, 2018). The conversion used for gas volume/price relationship is 1m^3 to 0.0411MMBtu.

Cost

Costs associated with a gas field development project are made up of Capital Costs and Operating costs. Capital costs fall under broad categories which include;

- Design and administration costs
- Equipment and materials purchase cost
- Fabrication costs
- Installation costs
- Commissioning cost
- Insurance spares cost
- Reinvestment cost

whereas operating costs include but are not limited to;

- Man-hour cost
- Spare parts and consumables consumption cost
- Logistic support cost
- Energy consumption cost
- Insurance cost
- Onshore support cost
- Cost of deferred production

Based on research data of Statoil operating fields, the overall cost breakdown of field development is categorised under *investment costs*, *operating costs*, *exploration costs*, *disposal and cessation* and *other costs*. In 2017, the total overall costs amounted to NOK 210 billion. This constituted 60% as *Investments*, 25% as *operating costs* with over 10% as exploration costs. (Norwegian Petroleum Directorate, 2017)

In the case of this master thesis, the costs captured have been narrowed down, for the capital cost, to include that associated with *Subsea development*, *Well and drilling* and *Production facility*.

The field development cost breakdown from data published in the OG 21 TTA4 Subsea Cost Report 2015 (OG21, 2015) gives the breakdown based on *field development concept* as represented in *Table 7.3;*

Table 7.3: Field development cost breakdown

FIELD DEVELOPMENT PROJECTS ON NCS			
	Subsea	**Well and Drilling**	**Production Facility**
Subsea tie-back	41%	35%	24%
Floating Installation	17%	37%	46%
Fixed Installation	3%	38%	69%

Source : OG 21 TTA4 Subsea Cost Report 2015(OG21, 2015)

This master thesis assumes the cost of the process equipment design and manufacture as well as installation correlated using a factor from the *average steel price*. This follows from the data given in *Table 7.3* and *Table 7.4* from NCS Subsea Cost Report from OG 21. It should be noted that following the decline in petroleum prices in 2014, costs related to offshore operations were affected and consistently change to adapt to market trends.

Table 7.4: Design and manufacture costs as a factor of Steel

Equipment Material Cost Factor	
Equipment Design and Manufacture Cost	3500%
Equipment Installation	4500%

The investment or initial capital cost was then determined by correlating the historical cost data on the NCS and the required weight of steel for processing equipment.
Other costs apart from capital cost considered under this master thesis has been broken down to include;

- Operating Costs – This has been lumped up to include all costs highlighted above. Based on Statoil historical data constitutes ***3%*** of capital cost.

- Tax - The tax rate is the percentage of corporate earnings imposed by government or federal state. Within Norway for the year 2018, the ordinary tax rate is pegged at 23% and special tax rate at 55% (Norwegian Petroleum Directorate, 2018b). Within the scope of the thesis, the tax rate has been assumed to be 78%. In addition, as the capital structure has not been defined to include debt, tax shields have not been taken to account in the profitability of the project.

- Carbon Tax – The carbon tax relates to regulating emissions to air from petroleum activities on the NCS. This is based on the *Carbon dioxide (CO_2) Tax Act on Petroleum Activities* which is levied on all combustion of gas, oil and diesel in petroleum operations on the continental shelf and on releases of CO_2 and natural gas. For 2018, the tax rate is NOK 1.06 per standard cubic metre of gas or per litre of oil or condensate. For combustion of natural gas, this is equivalent to NOK 453 per tonne of CO_2. (Norwegian Petroleum Directorate, 2017). A thorough explanation of this has been given under *Section 7.1.2.*

Cost of Capital
This is the rate that must be overcome to generate value for the project. It is company specific and depends on factors such as the company's operating history, profitability, creditworthiness etc. With respect to project financing, the capital structure is not defined under the scope of the project. Hence, it can be assumed the financing structure to be entirely from equity i.e. are obtained from internally generated funds. This implies the cost of capital can be used to discount the potential future cash flows from the project to estimate the Net Present Value. The Cost of Capital assumed for this project is ***8%***. All cash flows are discounted to year zero (0) which is the year the initial investment is made at the start of the project.

Depreciation
Costs of long-lived assets such as the gas processing equipment are spread over time. These have been assumed to be depreciated over a 6-year period for ease of comparison for each of the production profiles within the master thesis. This period has been selected based on data from the Norwegian Petroleum Directorate. (Norwegian Petroleum Directorate, 2017)

Net present Value
In order perform accurate profitability/investment decision analysis of the project, the present value of the cash flows need to be determined. The cash flows are discounted to a specific period (in this case to the time of investment or start of project) with the cost of capital assumed for the business. This takes into account the risk associated with the business including the time value of money. The cost of capital for upstream petroleum business in Norway is 8%. The present value of the cash flows for each year of the project is obtained by discounting the cash flows to the year of the capital expenditure, year zero. This is deduced from *equation 7.178* .

$$PV = \frac{FV}{(1+r)^n} \tag{7.178}$$

where;
PV - Present Value, *\$USD*
FV - Future Value, *\$USD*
r - Cost of Capital , 8*%*
n - time, *year*

The *Net Present Value* (NPV) is the sum of all the present value of the cash flows. A positive NPV indicates a profitable project as the cash inflows are greater than the cash outflows.

$$NPV = \sum PV\ of\ Cashflow\ , NPV > 0\ (project\ is\ feasible) \tag{7.179}$$

With an estimate of projected present value of cash flows, the payback period of the project can be estimated with cumulative present value of the cash flow. It should be noted again that for the purpose of this master thesis, this economic evaluation does not take into account equity and debt within the capital structure of the project, neither does it consider an economic evaluation if the project is deferred.

Table 7.5 sums up the economic and accounting input factors utilised in the cash flow analysis. *Appendix L.1*, *Appendix L.2* and *Appendix L.3* show the cash flow sheets for the three (3) production profiles under evaluation.

Table 7.5: Cost and Economic Factors

CURRENT ECONOMIC FACTORS		
Conversion (1m^3= 0.0411mmBtu)	0.0411	
Equipment Depreciation (years)	6	years
Average Gas Price	2.76	USD \$ per MMBtu
Average Condensate Price	64.94	USD \$ per bbl
Average Steel Price	660.00	per metric ton
Cost of Capital	8%	
Corporate Tax	*23%*	
Additional Tax Rate	*55%*	
Total Tax Rate	78%	

Based on the economic analysis performed *Figure 7.3*, *Figure 7.4* and *Figure 7.5* show the cumulative present value of the cash flow and break-even periods for each production profile based on a single well scenario. This shows scenario 1 giving the highest NPV followed by scenario 2 and then scenario 3. This is attributed to scenario 1 having the lowest plateau rate as compared to scenario 2 and 3. This implies the capacity/size of the processing equipment needed for the maximum production flowrate is lowest in scenario1. This in turn translates to lower CAPEX associated with equipment costs. In addition, although scenario 2 and 3 present higher production earlier on than in scenario 1, these revenues are largely eroded by the equipment capital cost based on their higher flowrates. This is extreme in the case of scenario 3. Another factor that is also accounted for is the discount rate; meaning, it is generally more profitable to produce today than tomorrow due to the ‘time value of money’. Although this theory of producing early is in favour of scenario 3, it is evident again that the very high production rates translate to larger capacity processing equipment hence much larger equipment capital cost which erodes into the high revenues generated under scenario 3.

Scenario 2, clearly falls in between scenario 1 and 3 where early higher production rates (meaning both higher revenues and larger capacity equipment) are observed with a later break-even period scenario 1.

As explained, from the results on the cash flow analysis, *scenario 3* generated high revenues (*Appendix L.3)* due to the high flowrates; however, the high flowrates required high capacity process equipment hence higher investment cost. The scenario in question showed no net profits for the production lifecycle, largely due to the high production rates as compared to the other scenarios. This presents the case where an optimum flowrate is required that would both meet the plant design criteria and the needed project profitability. This analysis of an optimum flowrate is detailed further on under *Section 7.1.5*.

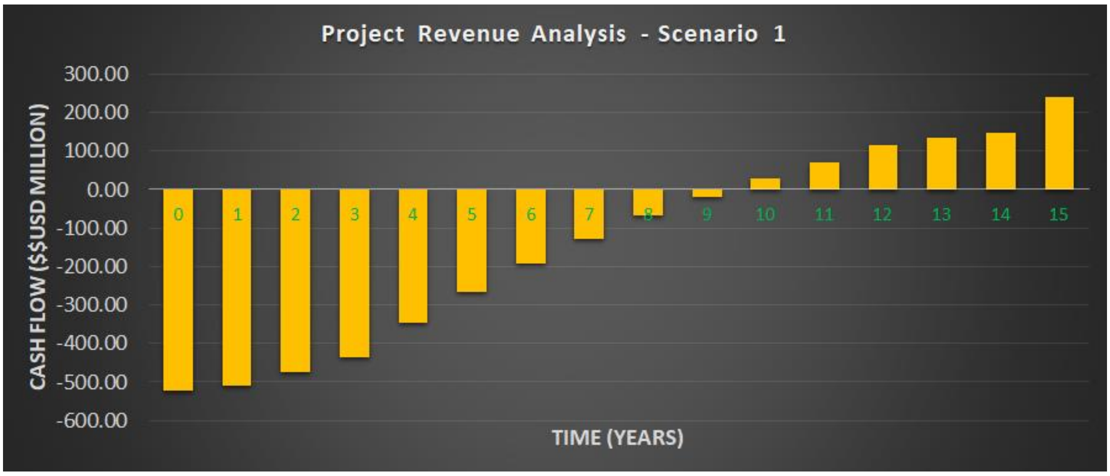


Figure 7.3: Economic evaluation - scenario 1

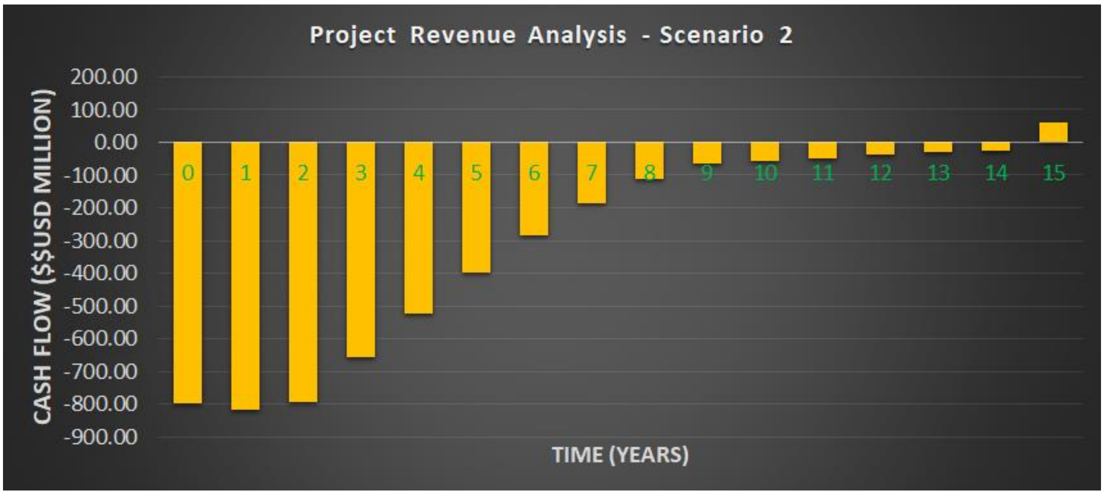


Figure 7.4: Economic evaluation - scenario 2

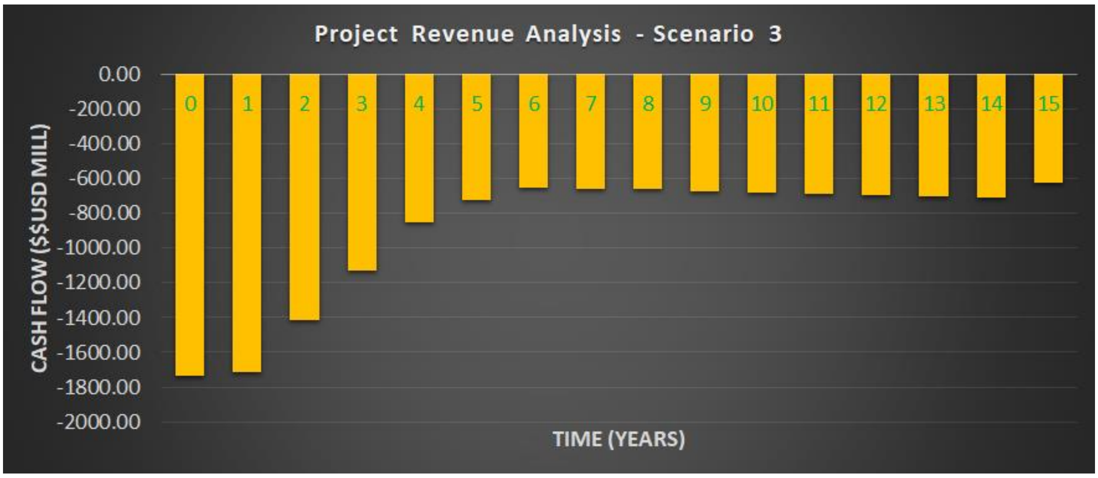


Figure 7.5: Economic evaluation - scenario 3

7.1.4 ASPEN HYSYS and ASPEN Process Economic Analyser (APEA)

This sub-chapter highlights the comparison of the process equipment sizing from the calculator developed and from ASPEN HYSYS. It also compares the cash flow analysis developed and the economic analyser, ASPEN Process Economic Analyser (APEA).

The sizing calculator developed takes into account fundamental concepts for analysing processing equipment. For more rigorous designs of each process equipment, more advanced design software can be used. *Figure 7.6* shows an example of the engineering design that can be performed on a separator in ASPEN HYSYS which in turn goes into the cost evaluation within the APEA.

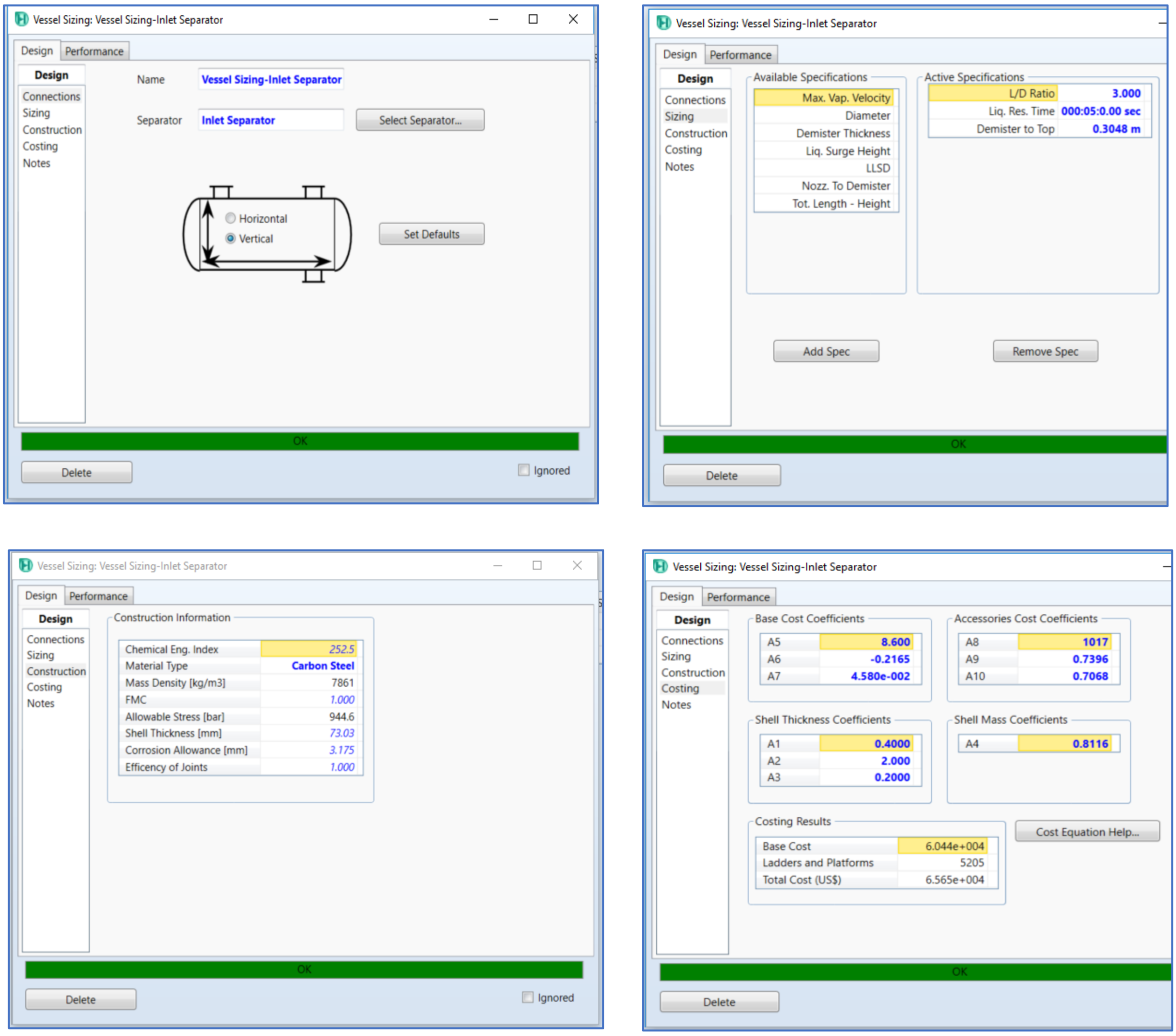


Figure 7.6: Vessel sizing evaluation

Figure 7.7 shows the relative differences of the results obtained in the calculator to the results in ASPEN HYSYS for some of the design parameters.

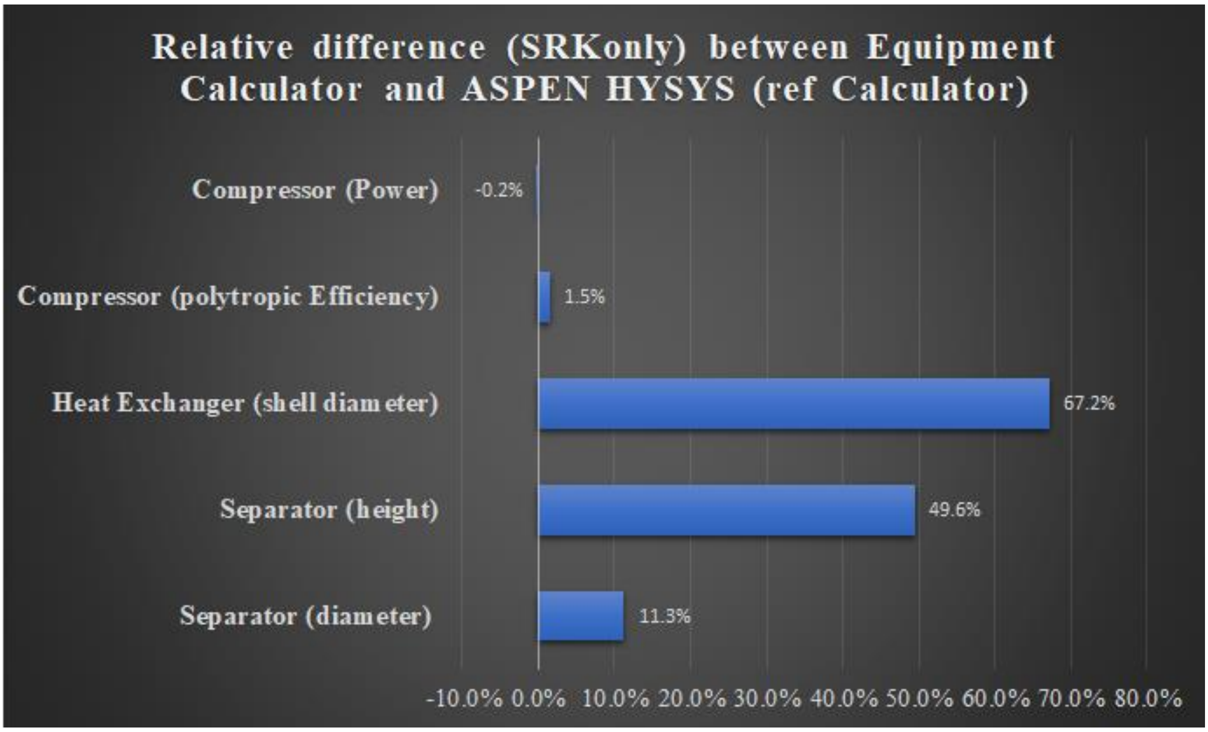


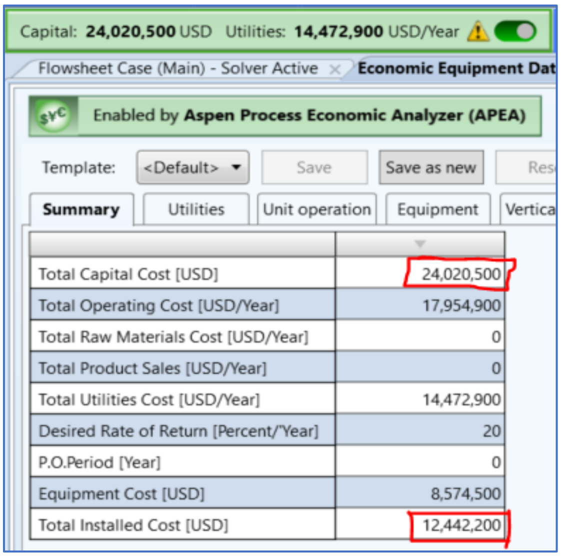
Figure 7.7: Relative difference between equipment calculator and ASPEN HYSYS

The APEA is a functionality on ASPEN HYSYS that allows a cost evaluation of a process simulation performed on ASPEN HYSYS. It takes into account an overall *capital expenditure* based on process equipment employed and a utility cost based on power requirements and consumption.

The APEA makes some assumptions based on a database of cost models. These models take into consideration factors such as

- operating Life of plant/ Operational hours
- length of plant startup
- equipment material and engineering specifications

For comparative purposes only, the APEA has been compared to the '*Calculator Economic model'* developed with the objective of matching both models. This comparison was also used as a check for the factors used to correlate the equipment cost to the steel cost shown under *Section 7.1.3*. *Figure 7.8* shows the final summary results of the process simulated for 10MMsm^3/d flowrate. The 'red markings' highlight the total processing equipment cost from the APEA model and the calculator economic model. This represented a relative difference of ~7.4% to the calculator economic model. This signifies a small difference between both models based on the assumed factors under *Section 7.1.3* and presents the calculator economic model as a good estimate for the cost evaluation of the processing plant.



Capital: **24,020,500** USD Utilities: **14,472,900** USD/Year

Flowsheet Case (Main) - Solver Active | **Economic Equipment Dat**

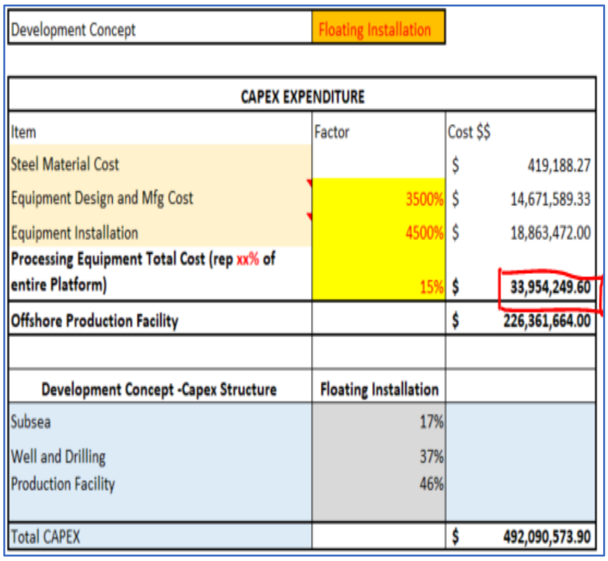
Enabled by **Aspen Process Economic Analyzer (APEA)**

Template: <Default> Save | Save as new | Res

Summary | Utilities | Unit operation | Equipment | Vertica

Total Capital Cost [USD]	24,020,500
Total Operating Cost [USD/Year]	17,954,900
Total Raw Materials Cost [USD/Year]	0
Total Product Sales [USD/Year]	0
Total Utilities Cost [USD/Year]	14,472,900
Desired Rate of Return [Percent/'Year]	20
P.O.Period [Year]	0
Equipment Cost [USD]	8,574,500
Total Installed Cost [USD]	12,442,200

ASPEN HYSYS Economic Model



Development Concept	Floating Installation	
CAPEX EXPENDITURE		
Item	Factor	Cost $$
Steel Material Cost		$ 419,188.27
Equipment Design and Mfg Cost	3500%	$ 14,671,589.33
Equipment Installation	4500%	$ 18,863,472.00
Processing Equipment Total Cost (rep xx% of entire Platform)	15%	**$ 33,954,249.60**
Offshore Production Facility		**$ 226,361,664.00**
Development Concept -Capex Structure	**Floating Installation**	
Subsea	17%	
Well and Drilling	37%	
Production Facility	46%	
Total CAPEX		**$ 492,090,573.90**

Calculator Economic Model

Figure 7.8: APEA model vs calculator economic model

From the comparison of the calculator economic model and the APEA model, it is evident that the model developed gives a more in-depth evaluation of internal components with respect to equipment such as separators, scrubbers and heat exchangers.

The cost structure for the calculator economic model has also taken into account business and economic factors in Norway for e.g. tax regimes, recent development concept breakdown factors for investment costs; be it for a floating installation, fixed platform or subsea installation.

As CAPEX and OPEX are dependent on costs factors that constantly change due to market structure, such as current commodity prices etc., a benchmark upon which to compare the cost models from ASPEN HYSYS and the model developed is not defined. However, the marginal differences of equipment cost from both models is a good indicator that an informed analysis from the calculator economic model can be made.

7.1.5 Case Study Suggestion

Process optimisation is one of the important procedures to undertake within complex processes to achieve best asset utilisation and performance. This is done to improve profitability of the plant or return on investment in a quantifiable manner. The benefits of this are to increase yield, reduce down-time, address off-specification production and to generally reduce energy costs (Poe & Mokhatab, 2017).

Various factors can be optimised depending on the objective within the gas processing plant. Within the scope of this master thesis, optimisation was performed by varying the production flowrates to determine the basis for the production profile that best gives a high return in revenue and low cost. However, the production rates of a gas field are affected by numerous technical and economic factors such as;

- Reservoir Characteristics
    - Water Coning
    - Sand production
    - Gas-Liquid Ratios (with increasing production)
- Equipment operational envelopes
    - Pressure
    - Temperature
    - Flowrate
- Pipeline flow or network limitations
    - Erosional velocity
    - Vibration
    - Noise
- Project profitability
    - Net Present value
    - Time to profitability

In effect, assumed production profiles as was done initially does not reflect the optimum production profile as the reservoir characteristics and other factors highlighted above have not been adequately represented in the profiles.

In practice, the production engineer provides the *production potential* or flowrates for life of field. These must be ascertained with the technical and economic factors highlighted. The suggested study case method employed considers a steady-state method for varied production profiles with a single objective to maximise Net Present Value.

It was noticed from the three production profiles that *scenario 1* gave the lowest maximum weight, relatively early break-even and the highest NPV as shown in *sections 7.1.1*, *7.1.2* and *7.1.3*. This indicates a high initial flowrate does not necessarily ensure high return as is the case with scenario 3; however, a sustainable plateau production gives a more technical and economically feasible life cycle of the project. From this, the deliverability of the field needs to be defined taking into account the mode of production and/or depletion during life of the field.

In oil and gas fields, production of hydrocarbon reserves can be done in two modes. Either producing via *constant rate, meaning* achieving plateau production or by *decline mode (full* *potential)*, which means producing as much as possible and as early as possible with a decline in production rates as depicted in *Figure 7.9*. Gas fields are predominantly produced using the constant rate mode where gas offtakes are bound by long-term sales agreements, production equipment limitations or regulatory control. This is important for economic analysis for an offshore development field. The constant rate mode is characterised by a plateau production rate and although there exists a constant rate this would eventually lead to production decline due to the fact that the plateau rate can no longer be sustained by the wellhead pressure sufficient to process and transport the gas and also due to depletion.

From historical data and by rule of thumb annual plateau rates normally within the range of *3.5% - 5%* of recoverable gas reserves for gas fields. (Golan & Whitson, 1996). The plateau rate is however flexible and it is a factor that can be controlled. However, the decline after a period of production is dependent on the reservoir characteristics. This means that production rates might be in some cases lower than the production potential but will follow a similar decline with time.

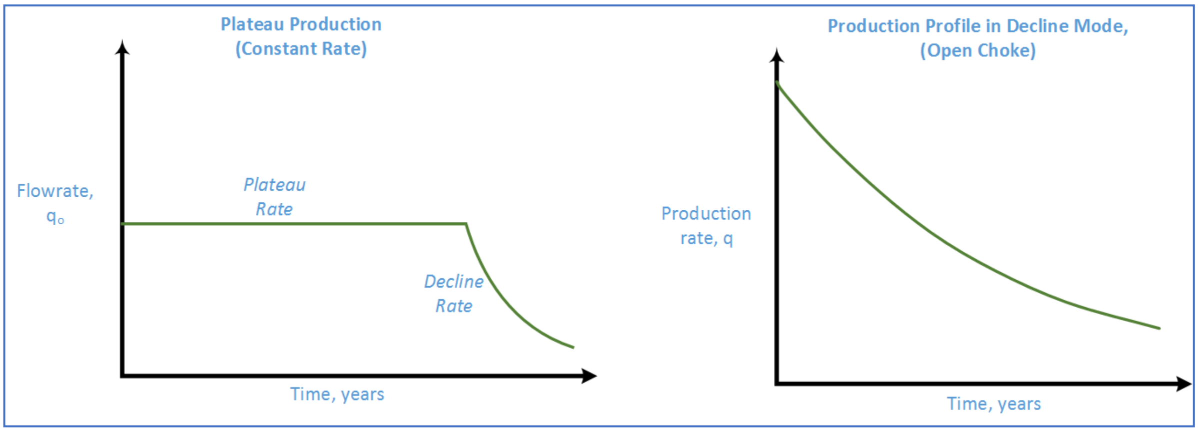


Figure 7.9: Oil and gas field production modes

The *production potential* of the well needs to be determined to ascertain the length of the plateau given. The production potential of the well is determined by performing material balance equations, flow equilibrium calculations based on a model of the well, flowline and separator parameters, information on the Inflow Performance Relationship (IPR), reservoir pressure and gas-liquid ratios (GLR). This information is obtained from the production engineer.

For the purpose of the master thesis, production potential has been deduced from the initial scenarios under investigation i.e. *scenario 1* and *2*. As mentioned, the production potential depends on the reservoir deliverability. It mainly depends on the amount of fluid withdrawn (not necessarily on time as is the case of water injected wells) from the reservoir. For a production system with a single well, assuming no changes to the production system in the life of the well and in a fully open choke production mode the production potential curve can be seen to be linear as shown in *Figure 7.10.*

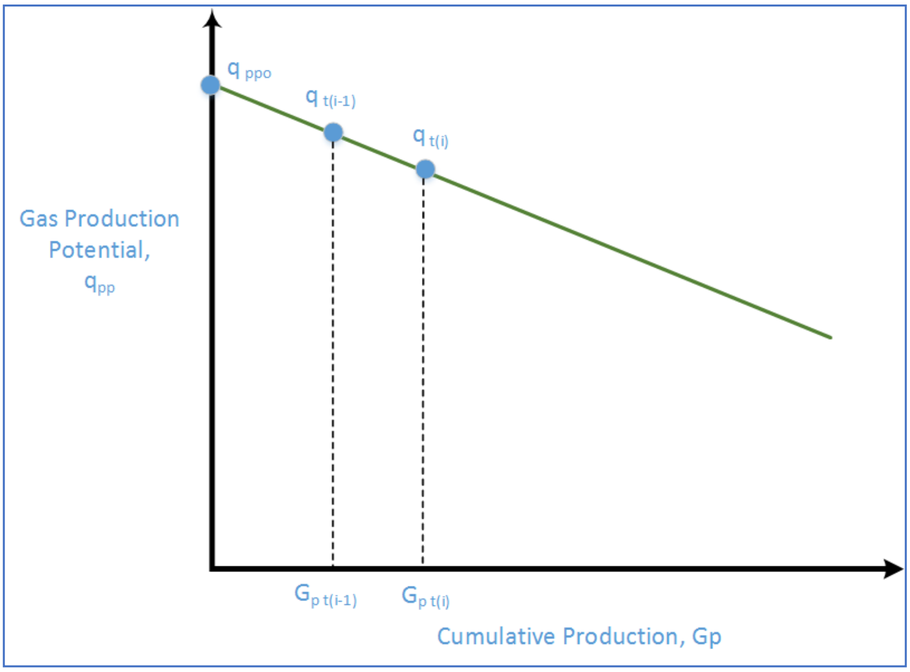


Figure 7.10: Production rate behaviour vs. cumulative production

The cumulative gas produced is given by*;*

$$G_p = \int_0^t q_g \; dt \tag{7.180}$$

where

G_p - Cumulative gas produced, *sm³*

q_g - Gas production rate, *sm³/d*

With the well producing all the time at its production potential, q_{pp}, (fully open choke) i.e. $q_g = q_{pp}$; the linear relationship between the production potential and cumulative production can be deduced from *equation 7.181*.

$$q_{pp} = -m\, G_p + q_{ppo} \tag{7.181}$$

where q_{ppo}, is the flowrate with no gas produced from the field.

Referencing *Figure 7.9* (point at which plateau ends and decline begins) and utilising *equation 7.181* a relationship for the well potential can be derived from both scenario 1 and scenario 2. Scenario 1 with a plateau rate (or q_{pp} in this case) of 10MMsm^3/d, cumulative production (G_p) at end of plateau at 31.5 Gsm^3 and scenario 2 with a plateau rate of 15MMsm^3/d, cumulative production at end of plateau at 24.3 Gsm^3; solving simultaneously for m and q_{ppo}, a linear relationship for the production potential is obtained and expressed as

$$q_{pp} = -0.694 \times 10^{-3}\, G_p + 31.86 \times 10^{6} \tag{7.182}$$

To determine the length of the plateau (for a plateau production mode) the cumulative gas produced for the years in production must be known. The production potential, q_{pp}, is representative of the open choke production i.e. assuming the well is allowed to flow at its full potential.

The production time, t, can be deduced from the same curve of gas flowrate, q_g, versus cumulative gas produced, G_p (*Figure 7.10).* This is given as *equation 7.180* where the time is determined based on trapezoid rule from *equation 7.183*

$$t^i = \frac{2\left(G_p^i - G_p^{i-1}\right)}{q_{pp}^i + q_{pp}^{i-1}} + t^{i-1} \tag{7.183}$$

The plateau rate can be sustained as long as the production potential is not exceeded, after which the production rate follows the decline of the production potential. (Nind, 1981)
The plateau rate, from historical data and as a rule of thumb, is approximately 3.5-5% of the recoverable reserves. Given in *equation 7.184*

$$q_{plateau} = \frac{0.5 * RF * IGIP}{Operational\ days} \tag{7.184}$$

where;
RF - Recovery Factor
$IGIP$ - Initial Gas In Place, sm^3

Within the thesis, the plateau rate of 5.95MMsm^3/d (based on the rule of thumb) was used as a starting point for determining the optimum plateau rate to achieve the maximum NPV. This plateau rate was analysed based on the production potential of the field. Increasing plateau rates were analysed with the objective to maximise NPV. The plateau flowrates analysed were;

- 5.95 MMsm^3/d
- 8 MMsm^3/d
- 10 MMsm^3/d
- 12 MMsm^3/d
- 15 MMsm^3/d
- 20 MMsm^3/d

The flowrates based on the production potential of the field gave a plateau length and decline (the lower the plateau rate the longer the plateau length) as presented in *Figure 7.11*.

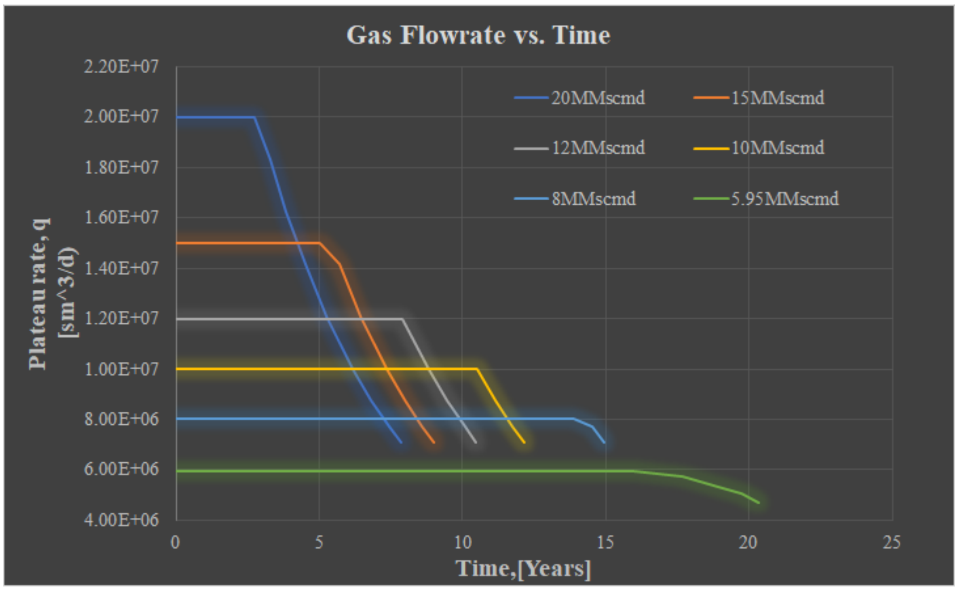


Figure 7.11: Gas Flowrate versus time for different plateau rates

For each assumed plateau rate, the NPV was investigated to determine which plateau rate gave the maximum NPV (reference *Appendix L.4* to *Appendix L.9*). This was found to be approximately 8MMsm^3/d with an NPV of USD $ 352 million as depicted in *Figure 7.12*. This is consistent with the analysis done for the initial 3 case scenario (scenarios 1, 2 and 3) whereby an increase in flowrate will result in increased production revenues. However, an increased flowrate would require an increase in capital expenditure for higher capacity equipment in line with that flowrate. Hence, production flowrate can be increased to an optimum point to obtain maximum possible revenues and lowest possible cost to obtain maximum profitability of the project. If Figure 7.12 had shown a flat increasing trend with flowrate, the results would be inconsistent as a limit would exist as to how much CAPEX can be added on to sustain an increase in flowrate. On the low side as well, a limit would exist as to how far production can be lowered to reduce cost and with the objective increasing profit from yearly production.

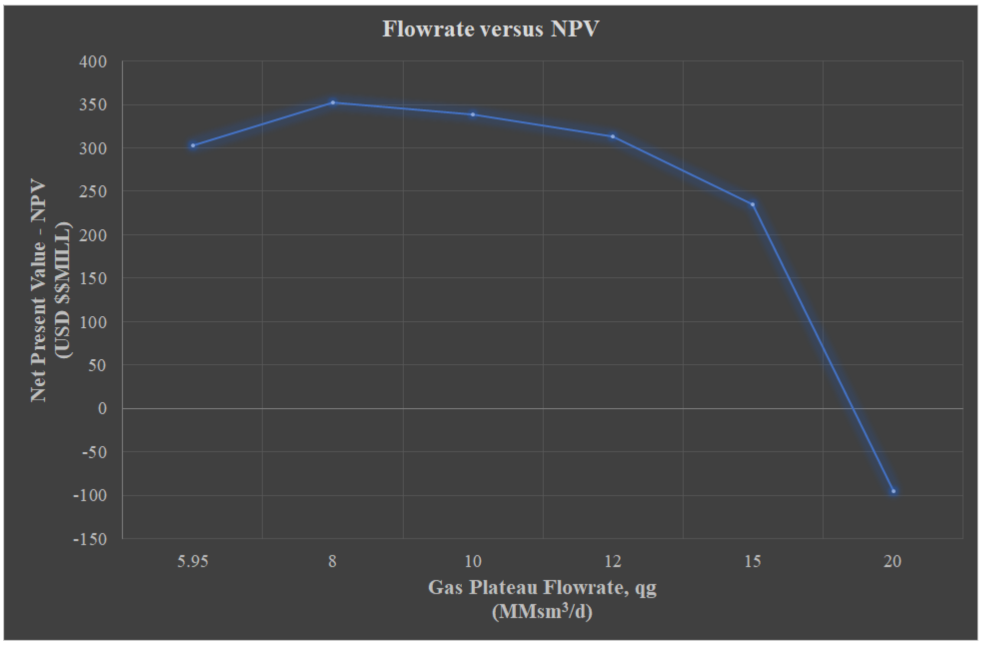


Figure 7.12: Plateau rate versus NPV

The length of the plateau rate also depends on the number of wells in production. An increase in the number of wells results in a longer plateau. An increase in the number of wells results in each well producing a smaller fraction of the total rate. Consequently, each well produces with a smaller pressure drawdown as compared to a single well producing at the total flow rate. This smaller pressure drawdown from having multiple wells reflects in a higher wellhead pressure which in turn results in a longer plateau period before reaching allowable wellhead pressure (Golan & Whitson, 1996). Determination of the allowable wellhead pressure has not been detailed under this research work.

Table 7.6, Figure 7.13 and *Figure 7.14* give the calculation of production potential and plateau length based on depletion and the corresponding time of plateau based on increasing the number of the wells for the optimum plateau rate of 8MMsm^3/d. *Appendix K.3* shows the visual basic code which was used for automatic interpolating of the production potential to the start of each year.

The results of the calculation reflect three cases;

- producing at a rate equal to the full potential of the well,
- producing from one single well
- producing from two wells and the corresponding length of the plateau

The results show that for a single well at open choke, the field would be depleted in approximately 7 years. Based on the same production potential, at a given plateau production of 8MMsm^3/d for a single well, production to depletion would take 15 years and for two wells this would take approximately 30 years.

Increasing the number of wells could be as a result of factors such as a change of gas sales contracts or agreements to deliver an estimated amount of gas for an extended period. It should be noted that increasing the number of wells will prolong the time till decline; however, this increases the capital expenditure as more wells would have to be drilled and completed and tied in to the processing facility. The impact of the increase in the number of wells to the profitability of the project has not been investigated in this thesis and is possibility for further work.

The single well case, plateau production of 8MMsm^3/d, is to be used as a *suggested case* for further study under a decision tree analysis. Further rigorous evaluation of this suggested case study could be performed to meet the reservoir, technical and economic requirements for managerial decisions to be made on project viability to meet gas sales contracts and agreements.

Table 7.6: Plateau length calculation

				PLATEAU				PLATEAU (2 wells)	
Cumulated gas produced, G_p	**Reservoir Pressure, P_R**	**Single Well Production Potential, qpp**	**Open Choke Production time (@ Qpp)**	**Single Well Plateau Production rate**	**Time (Single Well)**	**Field Production Potential, qpp**	**Field Open Choke Production Time (@ Qpp)**	**Field Plateau Production Rate**	**Time (Field)**
(sm^3)	(bara)	(sm^3/d)	(years)	(sm^3/d)	(years)	(sm^3/d)	(years)	(sm^3/d)	(years)
0.000E+00	225.00	3.186E+07	0.0	8.000E+06	0	3.186E+07	0	4.000E+06	0
1.500E+09	216.56	3.082E+07	0.2	8.000E+06	1	3.082E+07	0.16	4.000E+06	1
3.000E+09	207.36	2.978E+07	0.3	8.000E+06	1	2.978E+07	0.32	4.000E+06	3
4.500E+09	198.26	2.874E+07	0.5	8.000E+06	2	2.874E+07	0.50	4.000E+06	4
7.500E+09	180.97	2.666E+07	0.9	8.000E+06	3	2.666E+07	0.86	4.000E+06	6
1.050E+10	163.62	2.457E+07	1.2	8.000E+06	4	2.457E+07	1.25	4.000E+06	9
1.350E+10	146.74	2.249E+07	1.7	8.000E+06	6	2.249E+07	1.67	4.000E+06	11
1.650E+10	130.20	2.041E+07	2.1	8.000E+06	7	2.041E+07	2.14	4.000E+06	14
1.950E+10	113.89	1.832E+07	2.7	8.000E+06	8	1.832E+07	2.65	4.000E+06	16
2.250E+10	97.68	1.624E+07	3.2	8.000E+06	9	1.624E+07	3.23	4.000E+06	19
2.550E+10	81.47	1.416E+07	3.9	8.000E+06	11	1.416E+07	3.89	4.000E+06	21
2.850E+10	65.16	1.207E+07	4.7	8.000E+06	12	1.207E+07	4.65	4.000E+06	24
3.150E+10	48.64	9.989E+06	5.6	8.000E+06	13	9.989E+06	5.56	4.000E+06	26
3.330E+10	38.79	8.739E+06	6.2	8.000E+06	14	8.739E+06	6.20	4.000E+06	28
3.480E+10	30.33	7.698E+06	6.8	7.698E+06	15	7.698E+06	6.81	4.000E+06	29
3.570E+10	25.25	7.073E+06	7.2	7.073E+06	15	7.073E+06	7.22	4.000E+06	29.8

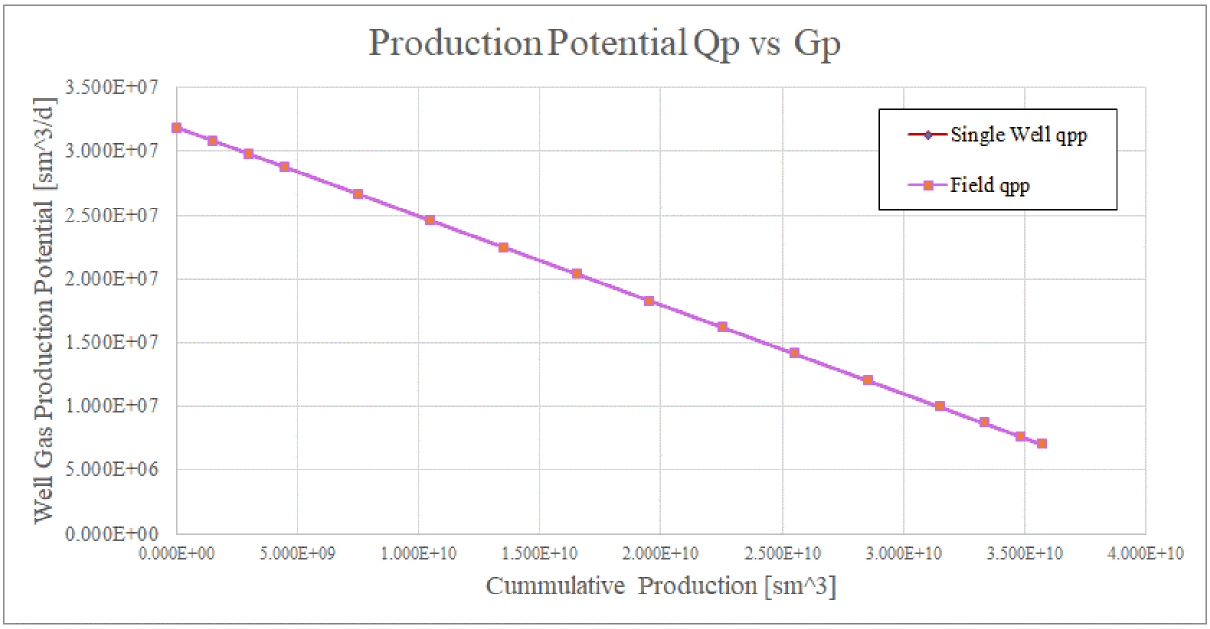


Figure 7.13: Suggested case production potential versus cumulative production

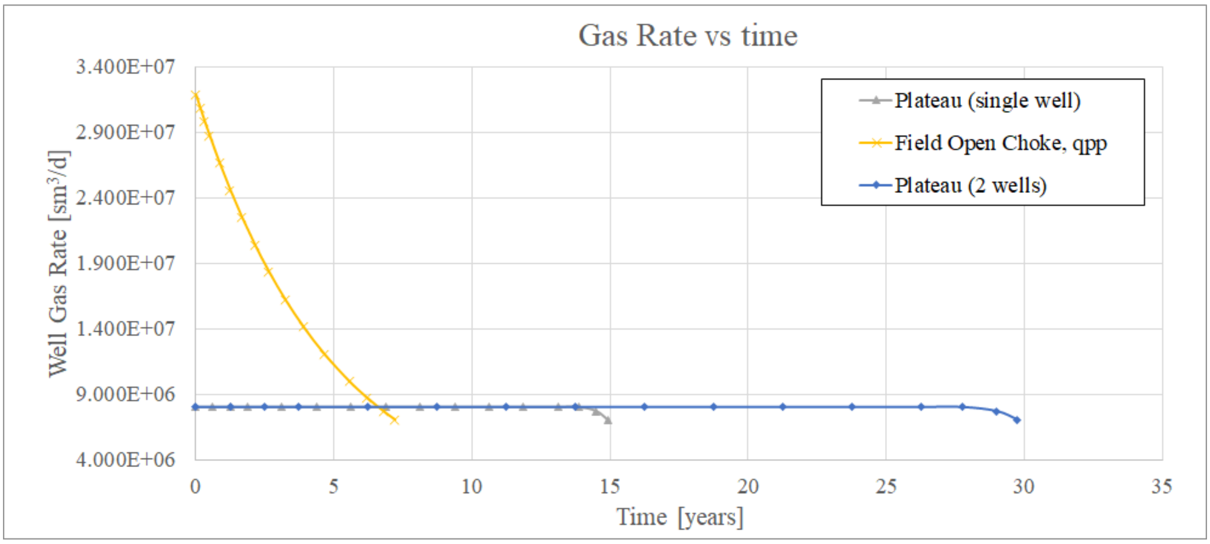


Figure 7.14: Production potential and plateau length

The production profile for the *suggested case* when run in the calculator gives the results for equipment design (weight and footprint), carbon footprint and cash flow analysis as displayed under *Table 7.7* and *Table 7.8.*

The NPV for the suggested case gives a higher value than that for scenario 1 (reference *Appendix L.1* and *Appendix L.5*) and payback period between 6 and 7 years. The highlighted cells show the years/case study under which maximum values for the lifecycle of the production was obtained.

Suggested Case Scenario

Table 7.7: Suggested case equipment results

Year/ Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total Footprint (m^2)	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	329.5	297.4
Total Weight (tons)	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	499.4	438.6
Total Duty (MW)	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	14.5	12.8
Compressor Power (MW)	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	12.0	10.6
Daily Compressor Energy (MWh)	287	287	287	287	287	287	287	287	287	287	287	287	287	287	254
Pump Power (kW)	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	375.8	331.1
Condensate (bbl/d)	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	8,544	7,555

Table 7.8: Suggested case carbon footprint

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Fuel Consumption (kg/s)	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.74	0.65
Fuel Consumption (sm^3/s)	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.78
CO_2 emissions cost per year (MM NOK /year)	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	24.39	21.57
CO_2 efficiency (kg CO_2 per BOE)	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.73

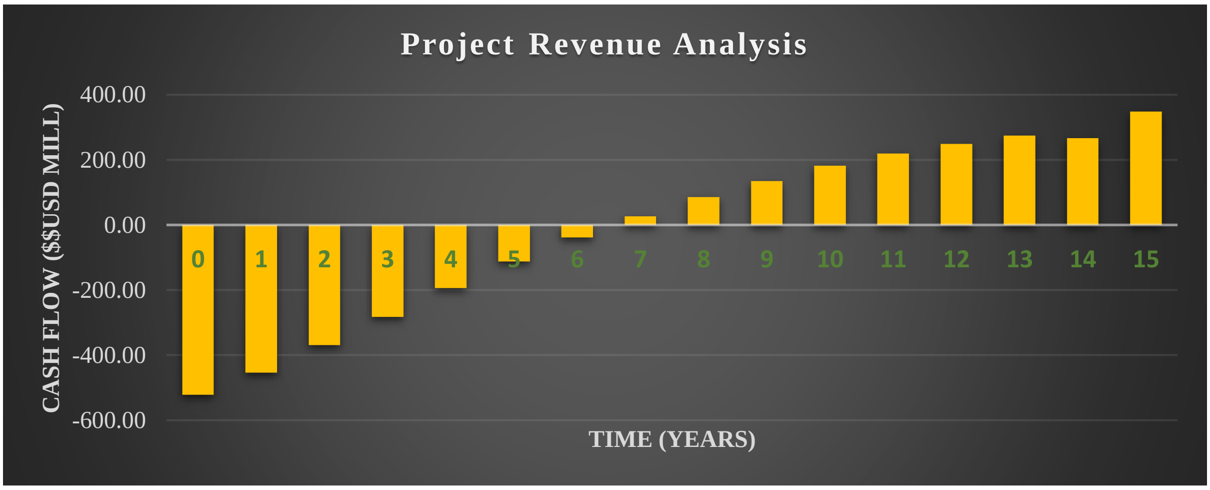


Figure 7.15: Economic evaluation - suggested case scenario

8 Discussion of Results

The focus of the thesis involved investigating two outcomes. Firstly, to evaluate the impact of gas process equipment design when utilising different thermodynamic models. This involved utilising two main thermodynamic models namely Soave-Redlich-Kwong (SRK) thermodynamic model and Peng Robinson to effectively characterise the fluid composition. This informed the parameters needed to size each processing equipment. By developing the equipment calculator (highlighted in *Chapter 3*) to effectively size all necessary equipment needed for gas processing i.e. separators, heat exchangers, compressors and pumps, an analysis was made into factors such as weight, footprint, volume and power requirements. These factors serve as indicators to make decisions based on processing facility requirements. From the calculators developed, the impact of utilising different thermodynamic models was highlighted in *Section 5.1.5*. The differences in results of the calculator arises out of the differences in thermodynamic models and the methods utilised by the models in characterising the reservoir fluids. As mentioned earlier, there could exist substantial liquid volumetric prediction differences between the SRK and PR Equations of State (EoS) as well as the fact that PR EoS underpredicts saturation pressure of reservoir fluids in comparison to the SRK EoS hence requiring a somewhat larger hydrocarbon/hydrocarbon (C_1/C_7+) binary interaction parameters (BIP) for PR EoS (Whitson et al., 2000). In addition, there is some evidence that PR EoS gives slightly better performance around the critical point, making the PR EoS better suited for gas/condensate systems (Robinson et al., 1985). Correlated historical and field data would ascertain the right thermodynamic models to be used.

Figure 8.1 summarises the results of the impact of thermodynamic models to gas processing equipment design. This represents a 3% and 2% difference in results for weight and footprint respectively using the SRK and PR thermodynamic models from the base case well scenario highlighted under *Chapter 1*. Such a difference could have a significant impact in the decision to be made on the facility requirements as well as the cost to the project depending on the scale.

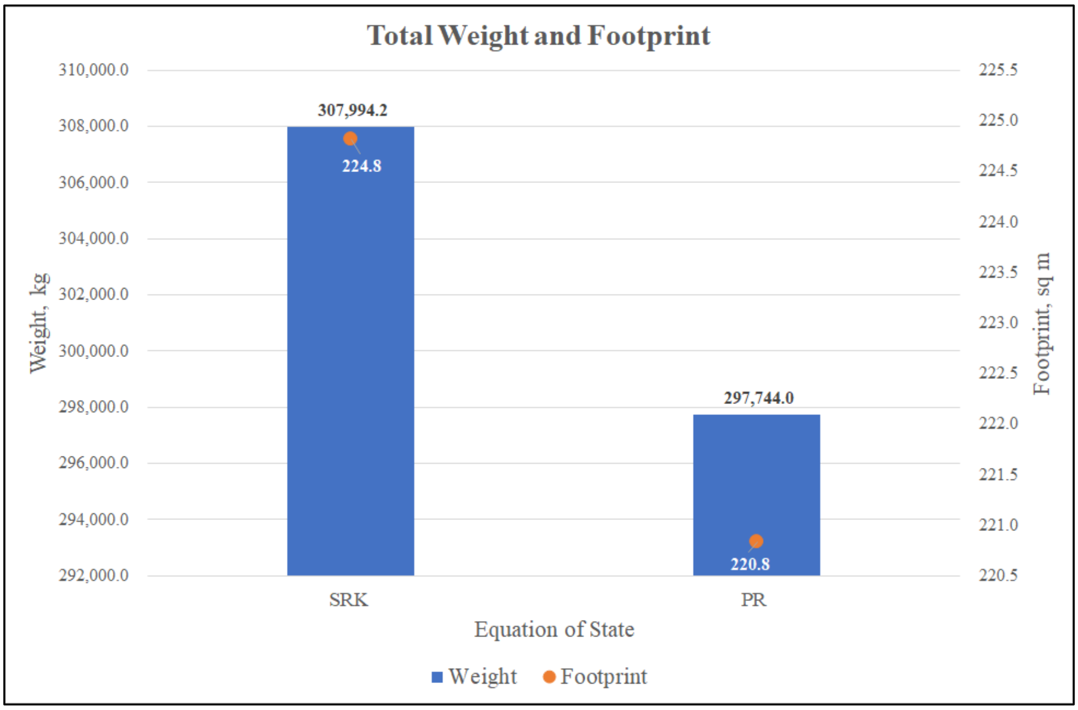
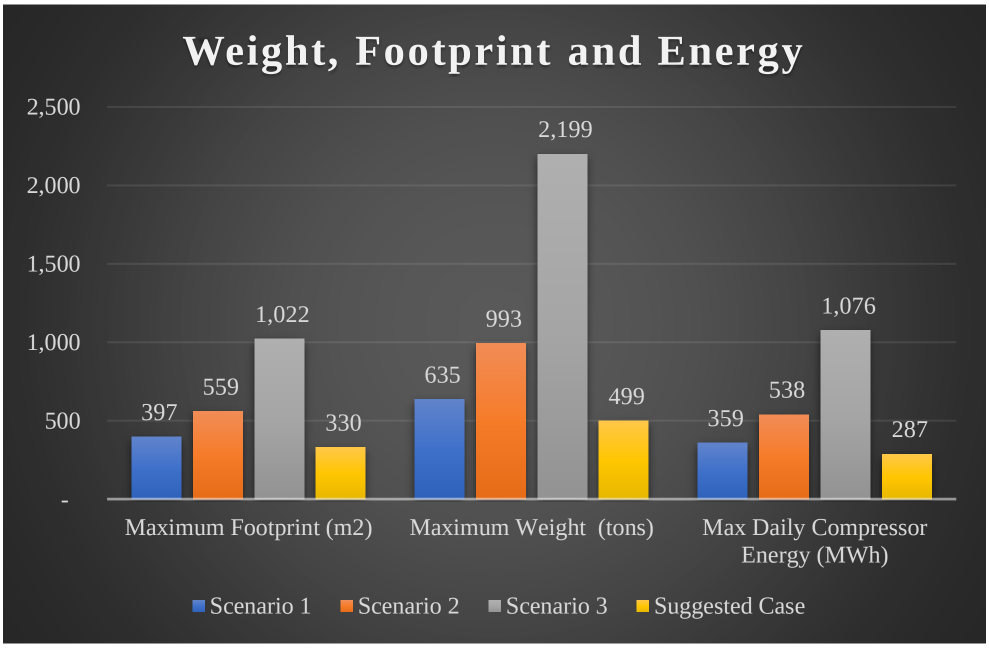
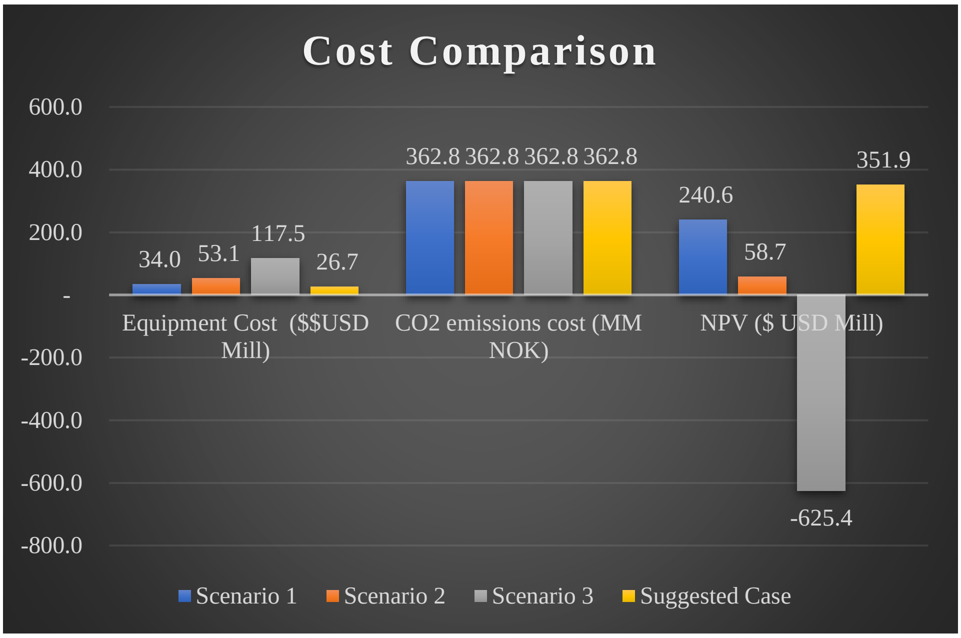


Figure 8.1: Weight and footprint analysis with SRK and PR thermodynamic models

The *second focus* of the master thesis is a build-up of the first objective which involves the automation of the gas processing plant. Many factors would impact the design of the offshore oil and gas field processing plant. An offshore gas processing plant could be developed based on different concepts; be it a subsea tie-back, floating installation or a fixed installation. The decision on the development concept impacts costs. In addition to the development concept factor is the change in operational parameters during the lifecycle of the field. This includes, but not limited to, production flowrates, pressure, and/or temperature which would affect the technical design and in turn affect capital structure and tax regime. Also, factors such as commodity price changes within the project lifetime would also affect its economic viability. In order to fully integrate all these factors in a working model, as was explained in *Chapters 0* and *7*, it is imperative to find the optimal design that incorporates both technically and economically feasible solutions that fits the objective of the project. The technical objective being the process design that meets the production lifecycle of minimal weight of equipment (which translates to cost) and the economic objective that maximises profitability (maximum NPV).

The automation was performed focussing on the three main production profiles (scenarios) explained in *Chapters 0* and *7*. This involved a sensitivity analysis using the production profiles within HYSYS to generate input parameters. These input parameters from HYSYS are incorporated directly into the calculator using the Aspen Simulation Workbook (ASW) interface. This automatically generates equipment design parameters for the production lifecycle which are recorded and tabulated to produce a technical design model. This technical design model then feeds into an economic model; a cash flow analysis which provides information on the project viability. The results of the initial scenario analysis based on the three production profiles (scenarios 1-3) explained under *Sections 7.1.1*, *7.1.2* and *7.1.3* are summarised under *Figure 8.2*.





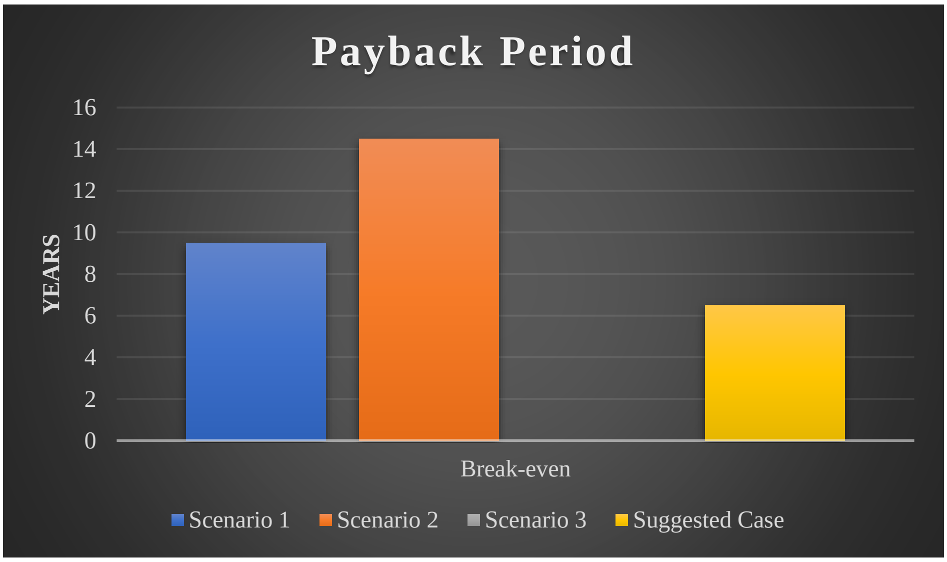


Figure 8.2: Scenario results comparison

The results showed that scenario 1; amongst the initial three scenarios investigated, represented the production profile that gave the minimum footprint, weight and energy consumption which translates to minimum CO_2 emission and minimum energy cost. This resulted in maximum net present value of the project utilising scenario 1 production profile. It can be inferred that a high production profile or a high early production rate (represented by scenario 3) which gives a high initial cash flow does not necessarily give an assurance of a profitable project as the costs associated with a high production rate are significantly increased, specifically with respect to increased equipment capacity. Due to these factors, scenario 3 showed a negative cashflow and no payback for the project. For cases such as scenario 3, more in-depth analysis needs to be performed to weigh the cost impacts against the revenue from high production rates as well as considering other development strategies such as drilling of multiple wells or tie-in to existing offshore processing facilities to offset some development costs.

Further to a comparison of the scenarios, an analysis of scenario 1 was performed utilising two different thermodynamic models; Soave-Redlich-Kwong and Peng Robinson. This was done to determine any significant differences to both the technical and economic parameters of the project when employing different thermodynamic models. The chart under *Figure 8.3* shows a comparative analysis of the study done. It is evident that the two different thermodynamic models give different predictions of equipment design and in turn different project profitability. The analysis showed a difference by a factor 3.5% in maximum weight of equipment in the plant and a factor of 5% in the case of NPV. This translates to a difference of 22 tons in weight and USD $12 million in NPV when utilising different thermodynamic models. PR EoS as shown earlier gives a lower weight approximation meaning lower equipment cost than SRK which translates to a higher NPV than SRK.

Based on these results, it is clear the significant impact that thermodynamic models have on evaluating equipment design and project profitability. It is imperative that as close as accurate thermodynamic models are used to evaluate the process as these could significantly impact the feasibility of offshore oil and gas field developments depending on the scale. In order, to ascertain the right thermodynamic model for a process design, correlating offset data or historical data from adjacent fields with previous models can give more accurate working models.

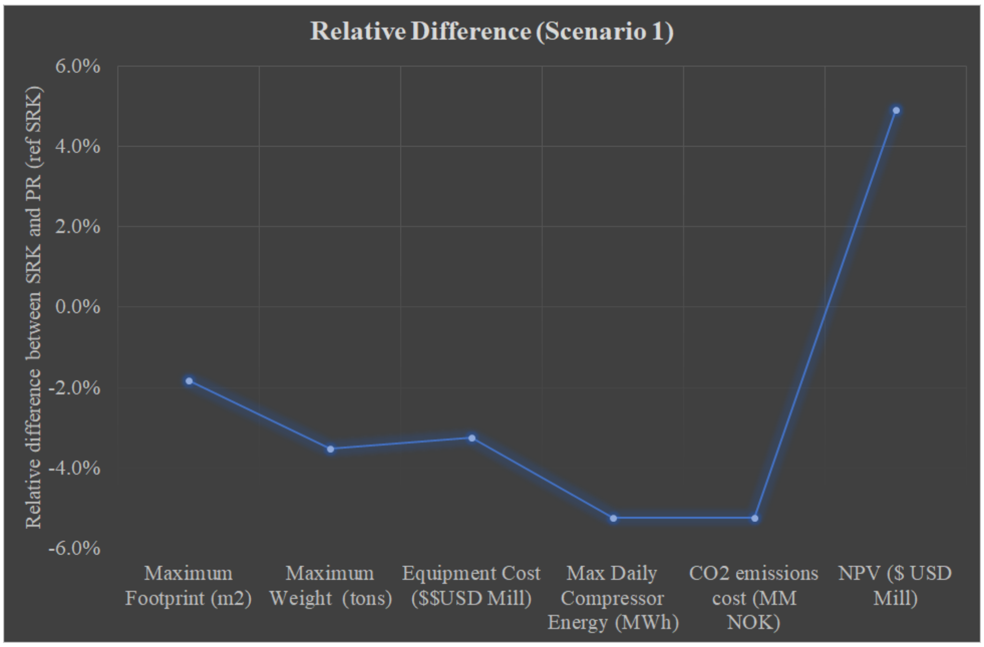
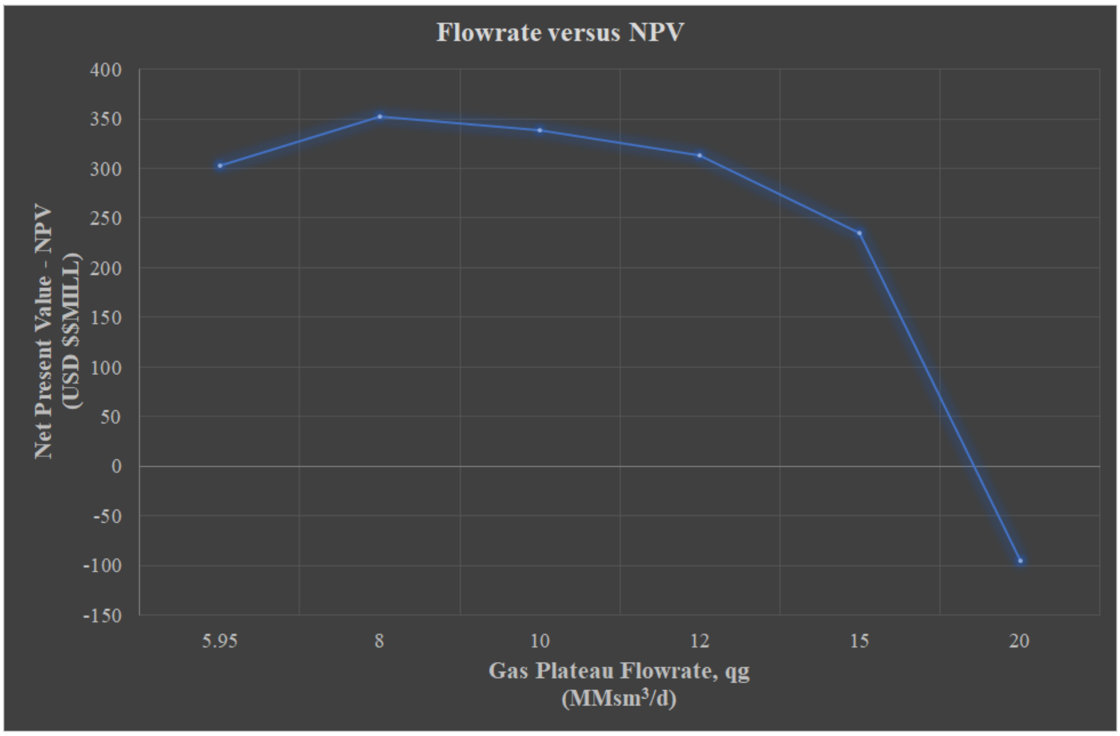


Figure 8.3: Relative difference with project indicators between SRK and PR

Appendix M gives the detailed breakdown of the sensitivity performed on scenario 1 using SRK and PR thermodynamic models.

The *suggested case* presented under *Section 7.1.5,* was based on further assumptions from scenario 1 incorporating the deliverability of the reservoir from the perspective of depletion. It highlighted a more realistic production profile depicting a plateau rate bounded by the field production potential. Different plateau production rates were investigated starting from a rate corresponding to 3.5-5% of the recoverable reserves (5.95 $MMsm^3/d$) which is based on a rule of thumb (explained under the same section). Incremental plateau rates were investigated and the net present value for each rate was evaluated. The results were consistent with the analysis performed for the initial three scenarios where an increase in flowrate results in increased production revenues. However, an increased flowrate would mean an increase in capital expenditure for higher capacity equipment. From the analysis of the plateau production, the optimum rate of 8 MMsm^3/d was seen to give the maximum NPV (*Figure 7.12* – highlighted below) which is indicative of maximum possible revenues and lowest possible cost.



The results from indicators based on the suggested case are shown in *Figure 8.2*. A further analysis was performed to show the effect of increasing the number of wells to the production life. The plateau rate could be sustained for much longer periods by increasing the number of wells; however, the profitability of the project would be impacted due to the increased cost of drilling and completing more wells as depicted in *Figure 7.14*.

Essentially, the master thesis has presented an automated tool capable of examining gas processing project indicators; comprising equipment weight/footprint/energy requirements, carbon footprint as well as cash flow analysis. It gives a preliminary design of gas processing equipment and provides the functionality of analysing the effect of different thermodynamic models to the design. Furthermore, it enables investigative analysis into changing parameters during the production lifecycle. These include, but are not limited to, production flowrates, increased water production, compositional changes, increase in well count etc. The master thesis looked critically at production flowrate as a basis to investigate project profitability. It presents a suggested production profile based on maximising NPV. The automated tool, thus allows evaluation of parameters such as equipment design (weight, footprint), energy requirements, carbon intensity and project profitability at the conceptual phase which would inform the decision process pertaining to offshore oil and gas field development which include;

- Offshore development concepts
- Facility weight limitations and requirements
- Equipment design and raw material costs
- Revenues from products

- Power and Utility consumption
- CO_2 emissions and CO_2 intensity targets and limits.
- Overall project economic viability

9 Further Work

The preceding chapters have focussed on the gas processing design and the impact of thermodynamic models (specifically with respect to Soave-Redlich-Kwong and Peng Robinson) on equipment sizing based on process equipment calculators.

Furthermore, the calculators have been interfaced with ASPEN HYSYS to allow automated functionality in performing sensitivity analysis on various gas project indicators such as offshore facility weight requirements, power and utility consumption, CO_2 emissions and intensity and overall project profitability based on different production profiles.

Further work in this regard is to analyse the dehydration unit incorporating the absorption column, regeneration column and ancillary equipment for a full offshore gas processing plant. This was not done due to time constraints and to limit the scope of the thesis.

The models for the master thesis could be utilised for other real life scenarios such as tie-in of additional wells to the gas processing facility. This could be used to show holistic optimum equipment design based on the field production lifecycle in the event of

- an increase in capacity,
- compositional changes
- increased water content in produced fluids

In addition, pressure decline analysis studies during life-of-field and corresponding fluid characteristic changes could be modelled to determine realistic product yield.

The models created from the thesis could be further investigated using other thermodynamic models such as Cubic Plus Association (CPA) and incorporating or interfacing to other simulation tools like NeqSim to create a functional automated interface.

A step further from the work done is to expand the model from an offshore platform and investigate aspects of optimising the process using more rigorous methods to ascertain product yield. A typical example is increasing yield of natural gas liquids to take advantage of increase in commodity price in a distillation process.

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Appendix A Physical properties of common petroleum reservoir fluids constituents

TABLE 1.1
Physical Properties of Common Petroleum Reservoir Fluid Constituents

Component	Formula	Molecular Weight (g/mol)	Melting Point (°C)	Normal Boiling Point (°C)	Critical Temperature (°C)	Critical Pressure (bar)	Acentric Factor	Density (g/cm^3) at 1 atm and 20°C
Inorganics								
Nitrogen	N_2	28.013	−209.9	−195.8	147.0	33.9	0.040	—
Carbon dioxide	CO_2	44.010	−56.6	−78.5	31.1	73.8	0.225	—
Hydrogen sulfide	H_2S	34.080	−83.6	59.7	100.1	89.4	0.100	—
				Paraffins				
Methane	CH_4	16.043	−182.5	−161.6	82.6	46.0	0.008	—
Ethane	C_2H_6	30.070	−183.3	−87.6	32.3	48.8	0.098	—
Propane	C_3H_8	44.094	−187.7	−42.1	96.7	42.5	0.152	—
Iso-butane	C_4H_{10}	58.124	−159.6	−11.8	135.0	36.5	0.176	—
n-Butane	C_4H_{10}	58.124	−138.4	−0.5	152.1	38.0	0.193	—
Iso-pentane	C_5H_{12}	72.151	−159.9	27.9	187.3	33.8	0.227	0.620
n-Pentane	C_5H_{12}	72.151	−129.8	36.1	196.4	33.7	0.251	0.626
n-Hexane	C_6H_{14}	86.178	−95.1	68.8	234.3	29.7	0.296	0.659
Iso-octane	C_8H_{18}	114.232	−109.2	117.7	286.5	24.8	0.378	0.702 (16°C)
n-Decane	$C_{10}H_{22}$	142.286	−29.7	174.2	344.6	21.2	0.489	0.730
				Naphthenes				
Cyclopentane	C_5H_{10}	70.135	−93.9	49.3	238.6	45.1	0.196	0.745
Methyl cyclopentane	C_6H_{12}	84.162	−142.5	71.9	259.6	37.8	0.231	0.754 (16°C)
Cyclohexane	C_6H_{12}	84.162	6.5	80.7	280.4	40.7	0.212	0.779
				Aromatics				
Benzene	C_6H_6	78.114	5.6	80.1	289.0	48.9	0.212	0.885 (16°C)
Toluene	C_7H_8	92.141	−95.2	110.7	318.7	41.0	0.263	0.867
o-Xylene	C_8H_{10}	106.168	−25.2	144.5	357.2	37.3	0.310	0.880
Naphthalene	$C_{10}H_8$	128.174	80.4	218.0	475.3	40.5	0.302	0.971 (90°C)

Source: Data from Reid, R.C., Prausnitz, J.M., and Sherwood, T.K. *The Properties of Gases and Liquids*, McGraw-Hill, New York, 1977.

Appendix B Compositions of reservoir fluids

Appendix B.1 Gas condensate mixture

Component	Mole Percentage	Molecular Weight	Density (g/cm³) at 1 atm and 15°C
N_2	0.53	—	—
CO_2	3.30	—	—
C_1	72.98	—	—
C_2	7.68	—	—
C_3	4.10	—	—
iC_4	0.70	—	—
nC_4	1.42	—	—
iC_5	0.54	—	—
nC_5	0.67	—	—
C_6	0.85	—	—
C_7	1.33	91.3	0.746
C_8	1.33	104.1	0.768
C_9	0.78	118.8	0.790
C_{10}	0.61	136	0.787
C_{11}	0.42	150	0.793
C_{12}	0.33	164	0.804
C_{13}	0.42	179	0.817
C_{14}	0.24	188	0.830
C_{15}	0.30	204	0.835
C_{16}	0.17	216	0.843
C_{17}	0.21	236	0.837
C_{18}	0.15	253	0.840
C_{19}	0.15	270	0.850
C_{20+}	0.80	391	0.877

Appendix B.2 Near-critical mixture

Component	Mole Percentage	Molecular Weight	Density (g/cm³) at 1 atm and 15°C
N_2	0.46	—	—
CO_2	3.36	—	—
C_1	62.36	—	—
C_2	8.90	—	—
C_3	5.31	—	—
iC_4	0.92	—	—
nC_4	2.08	—	—
iC_5	0.73	—	—
nC_5	0.85	—	—
C_6	1.05	—	—
C_7	1.85	95	0.733
C_8	1.75	106	0.756
C_9	1.40	121	0.772
C_{10}	1.07	135	0.791
C_{11}	0.84	150	0.795
C_{12}	0.76	164	0.809
C_{13}	0.75	177	0.825
C_{14}	0.64	190	0.835
C_{15}	0.58	201	0.841
C_{16}	0.50	214	0.847
C_{17}	0.42	232	0.843
C_{18}	0.42	248	0.846
C_{19}	0.37	256	0.858
C_{20+}	2.63	406	0.897

Appendix B.3 Black oil mixture

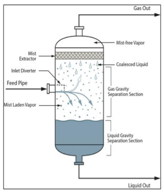
Component	Mole Percentage	Molecular Weight	Density (g/cm^3) at 1 atm and 15°C
N_2	0.04	—	—
CO_2	0.69	—	—
C_1	39.24	—	—
C_2	1.59	—	—
C_3	0.25	—	—
iC_4	0.11	—	—
nC_4	0.10	—	—
iC_5	0.11	—	—
nC_5	0.03	—	—
C_6	0.20	—	—
C_7	0.69	85.2	0.769
C_8	1.31	104.8	0.769
C_9	0.75	121.5	0.765
C_{10+}	54.89	322.0	0.936

Appendix C Separator calculation sheet

Appendix C.1 Two-phase vertical separator calculation sheet

HORIZONTAL SEPARATOR

SEPARATOR SIZING - Calculation Sheet



Colour Coding	
(yellow)	Input Parameter
(orange)	Drop down Input selection
(blue)	Empirical/Determined values
(green)	Output

DIAMETER CALCULATIONS

Liquid phase density	ρ_l	1000	kg/m3		
Gas phase density	ρ_g	74.35	kg/m3		
Molecular weight of Feed	MW	19.59	kg/kmol		
Sizing constant	$K_{S H}$	0.137	m/s		
Vapour mass velocity (Terminal velocity)	**Vgmax/ UT**	0.48339648	m/s		
Conservative Terminal Velocity	**Uv**	0.36254736			

**Based on Souders Brown equation*

Diameter of Vessel					
Gas flow rate	*Mscm*	4864800	scm/d	202,700	scm/h
Actual Gas flow rate	q_a	0.627562868	m3/s	2259.226324	m3/h
Vapour Mass velocity (Concservative)	*Vgmax*	0.36254736	m/s		
Mist Eliminator Diamter	*Dmist*	1.4844759			
Mist Eliminator allowance		YES		6inches	
Diameter of Separator	**Dmin**	**1.6368759**	**m**		

NB> Dmin is the mist eliminator diameter and 6in needs to be added to make up for Diameter of Vessel

Vessel Length (considering Mist Eliminators, Nozzle and Inlet Diverter - Svrcek and Monnery)					
Liquid Actual volume flow rate	qL	0.1709856	m3/d	0.000002	m3/s
Holdup Time **(reference Tables)**		5	min		
Holdup volume	V_L	0.001	m3		
Surge Time		3	min		
Surge Volume	V_s	0.000	m3		
Low Liquid Level Height	H_{LLL}	0.152	m		
Height from LLL to NLL	H_h	0.000	m	*1ft min*	
Height from NLL to HLL (High level alarm)	Hs	0.000	m	*6in min*	
Inlet Diverter		YES			
Height from HLL to Inlet nozzle centreline	H_{LIN}	0.612	m		
Disengagement Height	H_{dmin}	0.818			
Disengagement Height (with or without mist elim)	H_d	1.068	m		
Extra length	H_{ME}	0.457	m		
Height of Vessel		**2.2898**	**m**		

Nozzle Sizing					
Nozzle Sizing					
Mixture Volumetric Flowrate	Qm	0.628	m3/s	*22.16638767*	*ft3/s*
Mixture Liquid Fraction	λ	0.000			
Density of Mixture	ρ_m	74.525	kg/m3	*4.652453403*	*lb/ft3*
Nozzle Diameter	d_N	**0.307**	**m**	*1.007273491*	*ft*

EQUATIONS

$$V_{G\max} = K_S\sqrt{\left(\frac{\rho_L-\rho_G}{\rho_G}\right)}$$

$$D_{\min} = \sqrt{(4/\pi)q_a/(F_G V_{G\max})}$$

$$L_e = \sqrt{\frac{4tq_L}{\pi D^2 F_L}} \quad and \quad L = L_s + D$$

$$K_{SH} = K_{SV}(L_v/h_g) \quad \text{or} \quad K_{SH} = K_{SV}\left[\frac{(L_e/D)}{(h_g/D)}\right]$$

Wall thickness—2:1 ellipsoidal heads

$$t = \frac{Pd}{2SE - 0.2P}$$

$$W_V = W_b L + W_I + W_N$$

OUTPUT PARAMETERS			
Vertical Separator Diameter	**D**	**1.637**	**metres**
Vertical Separator Actual Height	**H**	**2.290**	**metres**
Vessel thickness	**t**	**2.780**	**in**
Weight of Vessel	**Wv**	**9.262**	**tons**
Weight of Total Skid Weight	**Wskid**	**14.951**	**tons**

NB. Weights do not take hydrostatic or fluid weights into consideration.

VESSEL WEIGHT CALCULATIONS

ANSI Class - Manways	-	150	
ANSI Class - Nozzles	-	150	
Nozzle Size	-	10	in
Vessel Diameter	-	1636.876	mm
Internal Description	-	Mist Eliminators	
Internal Type	-	Mist Mat	

Vessel Weight			
Internal Diameter	di	163.68759	cm
Height of Vessel	L	2.2898	m
Wall thickness (including corrosion allowance)	t	7.060439756	cm
Mass per unit length	Wb	4010.301097	kg/m
Weight of Internals	Wi	16	kg
Weight of Nozzles	WN	63.50300732	kg
Weight of Vessel	**Wv**	**9262.220054**	kg
Weight of Piping	WP	3704.888022	kg
Weight of Electrical & Instrument	WE	740.9776043	kg
Weight of Skid Steel	WS	926.2220054	kg
Weight of Manways		317	kg
Weight of Total Skid Steel	**Wskid**	**14951.30769**	kg

Vessel Thickness				
Material of vessel	-	**Carbon Steel Plates and Sheets**		
Specification number	-	**SA-516**		
Grade	-	**55**		
ASME Code		**Div1 (-20deg F to -650 deg F)**		
Working Pressure	P	1131.9	psi	77
Radius for Spherical shells	R			
Internal Radius of Shell	Ri	32.222	in	
External Radius of Shell	Ro			
Maximum Allowable Stress	S	13800	psi	
Joint Efficiency type	-	Fully radiographed - DW		
Joint Efficiency	E	1		
Vessel Shell Thickness	**t**	**2.7797**	in	
Ellipsoidal Head wall thickness	**teh**	0.0677		

Vessel Footprint			
Vessel Type		**Vertical**	
Skid Width		3.2737518	m
Skid Length		4.09218975	m
Skid Height		4.434673666	m
Footprint		13.39681356	m^2
Volume		59.4104963	m^3

VERTICAL GAS LIQUID SEPARATOR

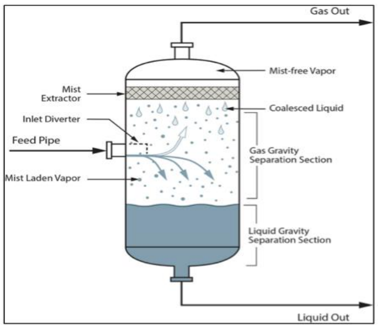
SEPARATOR SIZING - Calculation Sheet

Colour Coding

Colour	Meaning
(yellow)	**Input Parameter**
(orange)	**Drop down Input selection**
(blue-grey)	**Empirical/Determined values**
(green)	**Output**
(purple)	**iterative value**

INSTRUCTION

A. Fill in input parameters based on colour code
B. Determine **Drag co-efficient, Cd** by running the solver (error=0)---(Step 1) NB:-click on SOLVER)
C. Obtain **Minimum Allowable Vessel Diameter, Dmin**
D. From **Dmin**; obtain combinations of **D** and **H** (with an **SR**---between 3 and 5)
E. Manual operation of Dand H combinations is done manually or from ***OUTPUT REFERENCED TABLES***



INPUTS (IMPERIAL)

	NOMENCLATURE			METRIC	
Liquid phase density	ρ_l	53.05867665	lb/ft3	850	kg/m3
Gas phase density	ρ_g	3.745318352	lb/ft3	60.00	kg/m3
Gas Flow rate	Qg	15.010915	MMSCFD	0.43	MMscmd
Liquid Flow Rate	Q_l	3000	bbl/day	480	m^3/d
Gas viscosity	µg	0.013	cP		
Pressure	P	986.2584	psi	68	Bar
Temperature	T	519.678	R	288.71	K
Compressibility factor	Z	0.84	-		
Drag Co-efficient (Iterative)	Cd	1.166269	-		
Diameter of Particle	dm	100	µm		
Gas Retention time	tg	3	min		

STEP 1: DETERMINE Re, Cd, AND SETTLING VELOCITY

Settling velocity	u	0.398495419	ft/s	0.1215	m/s
Reynolds Number	Re	56.25547551			
Drag Co-efficient (Calculated)	Cd	1.166605671			
Error correction (Drag co-efficient)		0.000	SOLVER		

STEP 2: DETERMINE GAS CAPACITY CONSTRAINTS

Diameter squared	D^2	1000.161767	in^2		
Minimum Allowable Vessel Diameter	**Dmin**	**31.62533426**	in	0.8033	m

STEP 3: DETERMINE LIQUID CAPACITY CONSTRAINT

Retention time	t	0.002083333	day
D^2H^*		77085	in^3

**Determine combinations of D and H based on the min allowable vessel diameter*

MANUAL COMBINATION OF D AND H (refer to Output Table)

Diameter		32	in	0.8128	m
Height		75.27832031	in		

Slenderness Ratio Check (Between 3 and 4) - For D<36 in or For D>36 in

Height from combination	H	75.27832031	in		
Seam-to-seam length	Ls	12.60652669	ft	3.842	m
Slenderness Ratio	SR	4.72744751			

OUTPUT REFERENCES TABLES

Selected diameter (D)	Height of Separator (H)	Seam-to-Seam length (Ls)	Slenderness Ratio (SR)
in	in	ft	
32	75.28	12.61	4.73
34	66.68	11.89	4.20
36	59.48	11.29	3.76
38	53.38	10.95	3.46
40	48.18	10.68	3.20
42	43.70	10.47	2.99
44	39.82	10.32	2.81
46	36.43	10.20	2.66
48	33.46	10.12	2.53
50	30.83	10.07	2.42
52	28.51	10.04	2.32
54	26.44	10.04	2.23

EQUATIONS

$$u = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \quad \text{ft/s}$$

$$\mathrm{Re} = 0.0049\frac{\rho_g d_m u}{\mu_g}$$

$$C_d = 0.34 + \frac{3}{\mathrm{Re}^{0.5}} + \frac{24}{\mathrm{Re}}$$

$$D^2 = 5.058 Q_g\left(\frac{TZ}{P}\right)\left[\frac{\rho_g}{(\rho_o - \rho_g)}\frac{C_d}{d_m}\right]^{1/2} \ \text{in.}^2$$

$$D^2 H = 8.565 Q_o t \ \text{in.}^3$$

$D < 36$ in.

$$L_s = \frac{H + 76}{12} \ \text{ft}$$

$D > 36$ in.

$$L_s = \frac{H + D + 40}{12} \ \text{ft}$$

Notes

Gas Capacity Constraint - Determines the diameter
Smallest gas particle to be separated is 100µm
Retention time is between 1-3 minutes
Drag co-efficient is determined by iteration
Liquid Capacity Constraint - determines the height of the vessel

Typical retention times are as follows:

Natural gas-oil	1-3 minutes
Lean oil surge tanks	10-15 minutes
Fractionation feed surge tanks	8-15 minutes
Refrigerant surge tanks	4-7 minutes
Refrigerant economizers	2-3 minutes

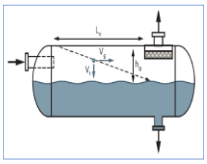
API 12J gives the following guidelines for gas-oil separation.

Oil Relative Density	Minutes
Below 0.85	1
0.85-0.93	1 to 2
0.93-1.0	2 to 4

Appendix C.2 Two-phase horizontal separator calculation sheet

HORIZONTAL SEPARATOR

SEPARATOR SIZING - Calculation Sheet



Colour Coding

Colour	Meaning
Yellow	Input Parameter
Orange	Drop down Input selection
Blue	Empirical/Determined values
Green	Output

DIAMETER CALCULATIONS

Liquid phase density	ρ_l	1086.860	kg/m3
Gas phase density	ρ_g	48.000	kg/m3
Sizing constant	$Ks\,H$	0.137	m/s
Vapour mass velocity (terminal velocity)	***Vgmax***	0.637	m/s
Conservative terminal velocity	***Uv***	0.478	m/s

**Based on Souders Brown equation*

Diameter of Vessel			
Gas flow rate	*Mscm*	5000000	scm/d
Fraction of Cross-section area for gas flow*	F_G	0.8	
Actual Gas flow rate	q_{av}	0.8439	m3/s
Vapour Mass velocity	*Vgmax*	0.6374	m/s
Minimum Gas Flow area	$A_{g,min}$	1.3241	
Diameter of Separator	**Dmin**	**1.4516**	**m**

**Fraction of cross section area available for gas flow (FG = 1 for vertical separators and is a function of liquid height for horizontal separators)*

Length of Vessel					
Liquid Actual volume flow rate	qL	3000	m3/d		
Holdup Time		5	min		
Holdup volume	V_L	10.417	m3		
Surge Time		5	min		
Surge Volume	V_s	10.417	m3		
L/D Ratio	L/D	6		Table	
Initial Diameter		2.246	m	7.3683	ft
Total Cross-sectional Area	A_T	3.962	m2		
Low Liquid Level Height	H_{LLL}	0.271	m	10.684	in
H_{LLL}/D ratio	H_{LLL}/D	0.187			
A_{LLL}/A_T ratio	A_{LLL}/A_T	0.088			
Low Liquid Level Area	A_{LLL}	0.347	m2		
Mist Eliminator Pad		YES	0.6096		
Height of Vapour Disengagement Area - Check		0.290	m	11.43	in
Height of Vapour Disengagement Area	H_v	0.610	m		
H_v/D ratio	H_v/D	0.420			
A_v/A_T ratio	A_v/A_T	0.217			
	A_v	0.860	m2		
Minimum Length - Holdup/Surge	L	4.030	m		
Liquid Dropout time	ϕ	1.275	s		
Actual Vapour Velocity	UVA	0.981	m/s		
Minimum Length for V-L Disengagement	Lmin	1.251	m		
Length of Vessel		**4.030**	m		
L/D Ratio		2.776			

NOTE: L/D Ratio must be between 1.5 to 6.0 (increase and decrease Diamter by 6in increments and repeat to obtain L/D ratio)

Area of Normal Liquid Level	A_{NLL}	2.9321	m2
A_{NLL}/A_T Ratio	A_{NLL}/A_T	0.7401	
H_{NLL}/D Ratio	H_{NLL}/D	-1.5511	
Height of Normal Liquid Level	H_{NLL}	-2.2516	m
Height of High Liquid Level	H_{HLL}	0.8420	m

EQUATIONS

$$V_{G\max} = K_S\sqrt{\left(\frac{\rho_L-\rho_G}{\rho_G}\right)}$$

$$D_{\min} = \sqrt{(4/\pi)q_a/(F_G V_{G\max})}$$

$$L_e = \sqrt{\frac{4tq_L}{\pi D^2 F_L}} \quad and \quad L = L_e + D$$

$$K_{SH} = K_{SV}(L_e/h_g) \quad \text{or} \quad K_{SH} = K_{SV}\left[\frac{(L_e/D)}{(h_g/D)}\right]$$

$$W_V = W_b L + W_I + W_K$$

Wall thickness—2:1 ellipsoidal heads

$$t = \frac{Pd}{2SE - 0.2P}$$

OUTPUT PARAMETERS			
Horizontal Separator Diameter	**D**	**1.452**	**metres**
Horizontal Separator Actual Length	**L**	**4.030**	**metres**
Vessel thickness	**t**	**1.232**	**in**
Weight of Vessel	**Wv**	**6.589**	**tons**
Weight of Total Skid Weight	**Wskid**	**10.955**	**tons**

NB. Weights do not take hydrostatic or fluid weights into consideration.

VESSEL WEIGHT CALCULATIONS

ANSI Class - Manways	-	300	
ANSI Class - Nozzles	-	300	
Nozzle Size	-	10	in
Vessel Diameter	-	1451.608	mm
Internal Description	-	Distillation Trays	
Internal Type	-	Normal	

Vessel Weight			
Internal Diameter	di	145.1607534	cm
Length of Vessel	L	4.0303	m
Wall thickness (including corrosion allowance)	t	3.129396464	cm
Mass per unit length	Wb	1576.301453	kg/m
Weight of Internals	Wi	159	kg
Weight of Nozzles	WN	77.1107946	kg
Weight of Vessel	**Wv**	**6589.078068**	kg
Weight of Piping	WP	2635.631227	kg
Weight of Electrical & Instrument	WE	527.1262454	kg
Weight of Skid Steel	WS	658.9078068	kg
Weight of Manways		544	kg
Weight of Total Skid Steel	**Wskid**	**10954.74335**	kg

Vessel Thickness			
Material of vessel	-	**Carbon Steel Plates and Sheets**	
Specification number	-	**SA-516**	
Grade	-	**55**	
ASME Code		**Div1 (-20deg F to -650 deg F)**	
Working Pressure	P	580	psi
Radius for Spherical shells	R		
Internal Radius of Shell	Ri	28.575	in
External Radius of Shell	Ro		
Maximum Allowable Stress	S	13800	psi
Joint Efficiency type	-	Fully radiographed - DW	
Joint Efficiency	E	1	
Vessel Shell Thickness	**t**	**1.2320**	in
Ellipsoidal Head wall thickness	**teh**	0.0306	

HORIZONTAL GAS LIQUID SEPARATOR

SEPARATOR SIZING - Calculation Sheet

Colour Coding

(yellow)	Input Parameter
(orange)	Drop down Input selection
(blue)	Empirical/Determined values
(green)	Output
(purple)	Iterative value

INSTRUCTION

A. Fill in input parameters based on colour code
B. Determine **Drag co-efficient, Cd** by running the solver (error=0)---(Step 1) NB:-click on SOLVER)
C. Obtain **Minimum Allowable Vessel Diameter, Dmin**
D. From **Dmin**; determine if separator is defined by Gas or Liquid constraint based on Lg and Lo
E. Manual operation of Dand H combinations can be done manually or from ***OUTPUT REFERENCED TABLES***

INPUTS (IMPERIAL)

	NOMENCLATURE			METRIC			
Liquid phase density	ρ_l	53.05867665	lb/ft3	850	kg/m3		
Gas phase density	ρ_g	3.745318352	lb/ft3	60.00	kg/m3		
Gas Flow rate	Qg	14.834316	MMSCFD	0.42	MMscmd		
Liquid Flow Rate	Qo	3000	bbl/day	480	m^3/d	11.77328	ft^3/min
Gas viscosity	µg	0.015	cP				
Pressure	P	986.2584	psi	68	Bar		
Temperature	T	519.678	R	288.71	K		
Compressibility factor	Z	0.84	-				
Drag Co-efficient (Iterative)	Cd	1.30158	-				
Diameter of Particle	dm	100	µm				
Gas Retention time	tg	3	min				
Fraction Occupied by Gas	-	0.5					
Fraction Occupied by Liquid	-	0.5					

STEP 1: DETERMINE Re, Cd, AND SETTLING VELOCITY

Settling velocity	u	0.377	ft/s	0.114976207	m/s
Reynolds Number	Re	46.151			
Drag Co-efficient (Calculated)	Cd	1.302			
Error correction (Drag co-efficient)		0.000	SOLVER		

STEP 2: DETERMINE GAS CAPACITY CONSTRAINTS

Diameter squared	D^2	1044.157977	in^2		
Minimum Allowable Vessel Diameter	**Dmin**	**32.31343339**	in	0.820761208	m
LgD		**87.11638325**	ft in		

STEP 3: DETERMINE LIQUID CAPACITY CONSTRAINT

D^2Lo		12841.38	in^3

MANUAL COMBINATION OF D AND H (refer to Output Table)

Diameter	D	33	in	0.8382	m
Constraint to satisfy design (compare Lg and Lo)		***Liquid Constraint***			

Reference Tables for Diameter and Height combinations

Length Based on Gas Constraint

Effective Length (gas constraint)	Lg	2.639890401	ft	0.80	m
Seam-to-seam length (Gas constraint)	Ls	5.389890401	ft	1.64	m
Slenderness Ratio	SR	0.163330012			

Length Based on Liquid Constraint

Effective Length (Liquid constraint)	Lo	11.79190083	ft	3.59400	m
Seam-to-seam length (Liquid constraint)	Ls	15.72253444	ft	4.79199	m
Slenderness Ratio	SR	5.717285249			

Slenderness Ratio Check (Between 3 and 5)

OUTPUT REFERENCES TABLES

Selected diameter (D)	Effective length-Gas (Lg)	Seam-to-Seam length-Gas (Ls-gas)	Effective length - liquid (Lo)	Seam-to-Seam length-Liquid (Ls-liq)	Constraint	Slenderness Ratio (SR)
in	ft	ft	ft	ft		
33	2.64	5.39	11.79	15.72	Liquid Constraint	5.72
35	2.49	5.41	10.48	13.98	Liquid Constraint	4.79
37	2.35	5.44	9.38	12.51	Liquid Constraint	4.06
39	2.23	5.48	8.44	11.26	Liquid Constraint	3.46
41	2.12	5.54	7.64	10.19	Liquid Constraint	2.98
43	2.03	5.61	6.95	9.26	Liquid Constraint	2.58
45	1.94	5.69	6.34	8.46	Liquid Constraint	2.25
47	1.85	5.77	5.81	7.75	Liquid Constraint	1.98
49	1.78	5.86	5.35	7.13	Liquid Constraint	1.75
51	1.71	5.96	4.94	6.58	Liquid Constraint	1.55
53	1.64	6.06	4.57	6.10	Liquid Constraint	1.38
55	1.58	6.17	4.25	5.66	Liquid Constraint	1.23
57	1.53	6.28	3.95	5.27	Liquid Constraint	1.11
59	1.48	6.39	3.69	4.92	Liquid Constraint	1.00
61	1.43	6.51	3.45	4.60	Liquid Constraint	0.91
63	1.38	6.63	3.24	4.31	Liquid Constraint	0.82

EQUATIONS

$$u = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \quad \text{ft/s}$$

$$\mathrm{Re} = 0.0049\frac{\rho_g d_m u}{\mu_g}$$

$$C_d = 0.34 + \frac{3}{\mathrm{Re}^{0.5}} + \frac{24}{\mathrm{Re}}$$

$$D^2 = 5{,}058 Q_g\left(\frac{TZ}{P}\right)\left[\frac{\rho_g}{(\rho_o - \rho_g)}\frac{C_d}{d_m}\right]^{1/2}$$

$$LD = 422\left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o - \rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \text{ ft in.}$$

$$D^2L = 1.428 Q_o t \text{ ft}^3$$

$$L_s = L_g + \frac{D}{12}\text{ ft}$$

$$L_s = \frac{4}{3}L_o\text{ ft}$$

Typical retention times are as follows:

Natural gas-oil	1-3 minutes
Lean oil surge tanks	10-15 minutes
Fractionation feed surge tanks	8-15 minutes
Refrigerant surge tanks	4-7 minutes
Refrigerant economizers	2-3 minutes

API 12J gives the following guidelines for gas-oil separation.

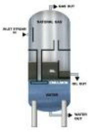
Oil Relative Density	Minutes
Below 0.85	<1
0.85-0.93	1 to 2
0.93-1.0	2 to 4

Notes
Theory Assumptions
Either the Gas Capacity Constrint or Liquid Capacity Constraint governs the design
Gas Capacity Constraint
Upward Gas velocity shouldnot exceed the the downward terminal velocity of the smallest oil droplet to be separated.
Iterate value of Cd to obtain settling velocity, u
Oil Capacity Constraint

Appendix C.3 Three-phase vertical separator calculation sheet

VERTICAL SEPARATOR

SEPARATOR SIZING - Calculation Sheet



Colour Coding

- Input Parameter
- Drop down Input selection
- Empirical/Determined values
- Output

DIAMETER CALCULATIONS

Gas phase density	ρg	48	kg/m3	2.9963 lb/ft3
Oil phase density	ρ_o	1086.86	kg/m3	67.844 lb/ft3
Water density	ρ_w	1100	kg/m3	
Gas viscosity	μg	0.015	cP	
Oil viscosity	μo	20	cP	
Water viscosity	μw	1	cP	
Oil Flow Rate	Q_O	800	m^3/d	5000 bbl/day
Water Flow Rate	Q_W	800	m^3/d	5000 bbl/day
Oil pad height	H_L	0.305	m	1 ft (assumed)
Water pad height	H_H	0.305	m	1 ft (assumed)
Holdup time	T_H	5	min	
Surge time	T_s	10		
Mixture Volumetric Flowrate	Qm	0.853		30.131 ft3/s
Mixture Liquid Fraction	λ	0.500		
Density of Mixture	ρm	567.430	kg/m3	35.424 lb/ft3
Nozzle diameter	d_N	0.595	m	1.9508 ft
Sizing constant	Ks_H	0.163	m/s	
Vapour mass velocity (vertical terminal vel)	***Vgmax (U_T)***	0.758	m/s	
Conservative vertical terminal velocity	***U_v***	0.569		

**Based on Souders Brown equation*

Diameter of Vessel			
Gas flow rate	*Mscm*	5000000	scm/d
Fraction of Cross-section area for gas flow*	*F_G*	0.8	
Actual Gas flow rate or Qv	*q_a*	0.844	m3/s
Vapour Mass velocity	*Vgmax*	0.758	m/s
Mist Eliminator Diamter	*Dmist*	1.374	
Mist Eliminator Installed		YES	
Diameter of Separator	***Dmin***	**1.527**	**m**

Settling Velocity				
Settling velocity of Heavy (water) out of Light liquid (oil)	U_{HL}	0.107	m/s	in/min
Rising velocity of Light liquid (oil) out of Heavy liquid(water)	U_{LH}	2.142	m/s	
Settling time for Water to rise out of Oil pad	t_{HL}	2.846	s	
Settling time for Oil to rise out of Water pad	t_{LH}	0.142	s	

Baffle Plate Calculation				
Density Difference (ρo-ρg)		1038.860	kg/m3	64.854 lb/ft3
Baffle Plate		YES		
Height from light liquid nozzle to baffle	H_R	0.229	m	9 in
Height from Liquid Interface to Baffle	$H_L + H_R$	0.533	m	
Allowable Downflow (downcomer)	G	8000.000	gphft2	8000 gphft2 ***read from the graph***
Downcomer cross-sectional area	A_{D1}	0.205	m^2	
Downcomer Chord Width	W_D	0.102	m	4.00 in assumed
	W_D/D	0.067		
Downcomer cross-sectional area to Area ratio (from Wd)	A_{D2}/A	0.024		
Area	A	1.831	m^2	
Downcomer cross-sectional area	A_{D2}	0.043	m^2	
Actual Downcomer cross-sectional Area	A_D	0.205	m^2	
Area of Baffle Plate	A_L	1.626	m^2	

Settling Time (Liquid)			
Settling Time - Oil	Θ_{LL}	53.538	secs
Settling Time - Water	Θ_{HL}	60.270	secs

****Note if $\Theta_{LL} < t_{HL}$ or $\Theta_{HL} < t_{LH}$ increase diameter and repeat procedure (Liquid separation is controlling factor)**

Vessel Height				
Actual Height from Oil nozzle to baffle (based on holdup time)	H_R	1.708	m	
Surge Height	H_s	6.068	m	*min 6in*
Liquid Level Above Baffle	H_A	0.152	m	
Minimum Check for H_{BN}	$0.5ft + H_s$	6.221	m	
Oil Height from Above Baffle Plate to Feed Nozzle	H_{BN}	6.518	m	
Minimum Check for H_D (0.5D)		0.763		
Minimum Check for H_D (with or without mist eliminator)		0.907		
Disengagement Height	H_D	0.907		
Extra Height with Mist Eliminator		0.457		
TOTAL Height of Vertical Separator	**H_T**	**8.873**	**m**	

EQUATIONS

$$V_{Gmax} = K_S\sqrt{\left(\frac{\rho_L - \rho_G}{\rho_G}\right)}$$

$$D_{min} = \sqrt{(4/\pi)q_a/(F_G V_{Gmax})}$$

$$t_{HL} = \frac{12H_L}{U_{HL}}$$

$$t_{LH} = \frac{12H_H}{U_{LH}}$$

$$\theta_{LL} = \frac{H_L A_L}{Q_{LL}}$$

$$\theta_{HL} = \frac{H_H A_H}{Q_{HL}}$$

$$H_R = \frac{Q_{LL} T_H}{A_L}$$

$$H_S = \frac{(Q_{LL} + Q_{HL})T_S}{A}$$

H_A = 6 in. minimum.
H_{BN} = ½ d_N + greater of (2 ft or H_S + 0.5 ft).
H_D = 0.5D or a minimum of:
36 in. + ½d_N (without mist eliminator), or
24 in. + ½d_N (with mist eliminator):

$$H_T = H_H + H_L + H_R + H_A + H_{BN} + H_D \quad (17)$$

Wall thickness—2:1 ellipsoidal heads

$$t = \frac{Pd}{2SE - 0.2P}$$

$$W_v = W_b L + W_I + W_K$$

$$t = \frac{PR_i}{SE - 0.6P} \qquad P = \frac{SEt}{R_i + 0.6t}$$

$$t = \frac{PR_o}{SE + 0.4P} \qquad P = \frac{SEt}{R_o - 0.4t}$$

OUTPUT PARAMETERS			
Vertical Separator Diameter	D	1.527	metres
Vertical Separator Actual Length	L		metres
Vessel thickness	t	1.296	in
Weight of Vessel	Wv	15.710	tons
Weight of Total Skid Weight	Wskid	25.365	tons

NB. Weights do not take hydrostatic or fluid weights into consideration.

VESSEL WEIGHT CALCULATIONS

ANSI Class - Manways	-	300	
ANSI Class - Nozzles	-	300	
Nozzle Size	-	10	in
Vessel Diameter	-	1526.854	mm
Internal Description	-	Distillation Trays	
Internal Type	-	Normal	

Vessel Weight			
Internal Diameter	di	152.6853517	cm
Length of Vessel	L	8.8728	m
Wall thickness (including corrosion allowance)	t	3.291612838	cm
Mass per unit length	Wb	1743.956291	kg/m
Weight of Internals	Wi	159	kg
Weight of Nozzles	WN	77.1107946	kg
Weight of Vessel	**Wv**	**15709.80835**	kg
Weight of Piping	WP	6283.923342	kg
Weight of Electrical & Instrument	WE	1256.784668	kg
Weight of Skid Steel	WS	1570.980835	kg
Weight of Manways		544	kg
Weight of Total Skid Steel	**Wskid**	**25365.4972**	kg

Vessel Thickness			
Material of vessel	-	**Carbon Steel Plates and Sheets**	
Specification number	-	**SA-516**	
Grade	-	**55**	
ASME Code		**Div1 (-20deg F to -650 deg F)**	
Working Pressure	P	580	psi
Radius for Spherical shells	R		
Internal Radius of Shell	Ri	30.056	in
External Radius of Shell	Ro		
Maximum Allowable Stress	S	13800	psi
Joint Efficiency type	-	Fully radiographed - DW	
Joint Efficiency	E	1	
Vessel Shell Thickness	**t**	**1.2959**	in
Ellipsoidal Head wall thickness	teh	0.0322	

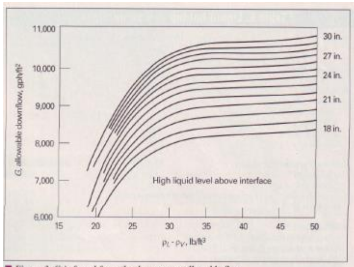


Figure 3. G is found from the downcomer allowable flow.

VERTICAL GAS LIQUID SEPARATOR

SEPARATOR SIZING - Calculation Sheet

Colour Coding

Colour	Meaning
Yellow	Input Parameter
Orange	Drop down Input selection
Grey	Empirical/Determined values
Green	Output
Purple	iterative value

INSTRUCTION

A. Fill in input parameters based on colour code
B. Determine **Drag co-efficient, Cd** by running the solver (error=0)---(Step 1) NB:-click on SOLVER)
C. Obtain **Minimum Allowable Vessel Diameter, Dmin**
D. From **Dmin**; obtain combinations of **D** and **H** (with an **SR**---between 3 and 5)
E. Manual operation of Dand H combinations is done manually or from ***OUTPUT REFERENCED TABLES***

INPUTS (IMPERIAL)	NOMENCLATURE			METRIC	
Liquid phase density	ρ_l	67.84394507	lb/ft3	1086.86	kg/m3
Gas phase density	ρ_g	2.996254682	lb/ft3	48.00	kg/m3
Specific Gravity water	γ_w	1			
Specific Gravity Oil	γ_o	0.87			
Gas Flow rate	Qg	176.599	MMSCFD	5.00	MMscmd
Oil Flow Rate	Q_O	5000	bbl/day	800	m^3/d
Water Flow Rate	Q_W	5000	bbl/day	800	m^3/d
Gas viscosity	μg	0.015	cP		
Oil viscosity	μo	20	cP		
Pressure	P	1305.342	psi	90	Bar
Temperature	T	500.67	R	278.15	K
Compressibility factor	Z	0.87	-		
Drag Co-efficient (Iterative)	Cd	1.275734123	-		
Diameter of Gas Particle	dm_g	100	µm		
Diameter of Liquid Particle	dm_l	500	µm		
Gas Retention time	tg	3	mins		
Oil retention time	to	10	mins		
Water retention time	tw	10	mins		
Fraction Occupied by Gas	-	0.5			
Fraction Occupied by Liquid	-	0.5			

STEP 1: DETERMINE Re, Cd, AND SETTLING VELOCITY				
Settling velocity	u	0.488	ft/s	
Reynolds Number	Re	47.813		
Drag Co-efficient (Calculated)	Cd	1.276		
Error correction (Drag co-efficient)		0.000	SOLVER	

STEP 2: DETERMINE WATER SETTLING CAPACITY CONSTRAINTS					
Minimum Allowable Vessel Diameter	D^2min	20572.30769	in^2		
Minimum Allowable Vessel Diameter	Dmin	143.4304978	in	3.643134644	m

STEP 3: DETERMINE GAS CAPACITY CONSTRAINTS					
Diameter squared	D^2min	3236.314846	in^2		
Minimum Allowable Vessel Diameter	Dmin	56.88861789	in	1.444970894	m

MANUAL COMBINATION OF D AND H (refer to Output Table)					
Constraint to satisfy design		Water Settling Constraint			
Diameter (Asssumed based on Step 2 or 3 Dmin)	D	144	in	3.6576	m

STEP 4: LIQUID RETENTION TIME CONSTRAINT			
$(Ho+Hw)D^2$		857600	in^3
(Ho+Hw)		41.35802469	in

*Determine combinations of D and (Ho+Hw) based on the min allowable vessel diameter

STEP 4: Check based on Seam-to-seam length, Ls

Slenderness Ratio Check (Between 3 and 4) - For D>36 in or For D<36 in					
Seam-to-seam length	Ls	18.77983539	ft	5.723814505	m
Slenderness Ratio	SR	0.130415524			

Slenderness Ratio Check (Between 3 and 5)

Notes

Gas Capacity Constraint or Water Settling Capacity - Determines the diameter
Smallest gas particle to be separated assumed to be 100µm (in absence of Lab data)
Smallest liquid particle to be separated assumed to be 500µm (in absence of Lab data)
Retention time is between 10-30 minutes for water separating from oil
Retention time is between 1-3 minutes for gas separating from liquid
Drag co-efficient is determined by iteration
Liquid Retention Constraint - determines the height of the vessel

OUTPUT REFERENCES TABLES

Water Settling Constraint diameter (Dmin)	Gas Capacity Constraint diameter (Dmin)	Design Constraint for Diameter	Height of Liquid level (Ho+Hw)	Seam-to-Seam length for Water Settling Constraint (Ls)	Seam-to-Seam length for Gas Capacity Constraint (Ls)	Water Settling Slenderness Ratio - (SR)	Gas Capacity Slenderness Ratio - (SR)
in	in		in	ft	ft		
144	57	Water Settling Constraint	41.36	18.78	11.53	1.56	2.43
146	59	Water Settling Constraint	40.23	18.85	11.60	1.55	2.36
148	61	Water Settling Constraint	39.15	18.93	11.68	1.53	2.30
150	63	Water Settling Constraint	38.12	19.01	11.76	1.52	2.24
152	65	Water Settling Constraint	37.12	19.09	11.84	1.51	2.19
154	67	Water Settling Constraint	36.16	19.18	11.93	1.49	2.14
156	69	Water Settling Constraint	35.24	19.27	12.02	1.48	2.09
158	71	Water Settling Constraint	34.35	19.36	12.11	1.47	2.05
160	73	Water Settling Constraint	33.50	19.46	12.21	1.46	2.01
162	75	Water Settling Constraint	32.68	19.56	12.31	1.45	1.97
164	77	Water Settling Constraint	31.89	19.66	12.41	1.44	1.93
166	79	Water Settling Constraint	31.12	19.76	12.51	1.43	1.90

EQUATIONS

$$u = 0.01186\left[\left(\frac{\rho_o - \rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \quad \text{ft/s}$$

$$\mathrm{Re} = 0.0049\frac{\rho_g d_m u}{\mu_g}$$

$$C_d = 0.34 + \frac{3}{\mathrm{Re}^{0.5}} + \frac{24}{\mathrm{Re}}$$

$$D_{\min}^2 = 5058 Q_g\left(\frac{TZ}{P}\right)\left(\frac{\rho_g}{\rho_o - \rho_g}\frac{C_d}{d_m}\right)^{1/2} \quad \text{in}^2$$

$$D_{\min}^2 = 6686\frac{Q_o \mu_o}{(\Delta\gamma)d_m^2} \quad \text{in.}^2$$

$$(H_o + H_w)D^2 = 8.576(Q_o t_o + Q_w t_w)$$

For $D > 36$ in.

$$L_s = \frac{1}{12}(H_o + H_w + D + 40) \quad \text{ft}$$

For $D < 36$ in.

$$L_s = \frac{1}{12}(H_o + H_w + 76) \quad \text{ft}$$

Typical retention times are as follows:

Natural gas-oil	1-3 minutes
Lean oil surge tanks	10-15 minutes
Fractionation feed surge tanks	8-15 minutes
Refrigerant surge tanks	4-7 minutes
Refrigerant economizers	2-3 minutes

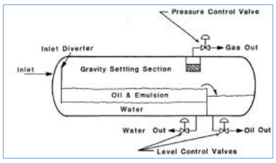
API 12J gives the following guidelines for gas-oil separation.

Oil Relative Density	Minutes
Below 0.85	<1
0.85-0.93	1 to 2
0.93-1.0	2 to 4

Appendix C.4 Three-phase horizontal separator calculation sheet

HORIZONTAL SEPARATOR

SEPARATOR SIZING - Calculation Sheet (3-Phase with Interface Control Weir)



Colour Coding
- Input Parameter
- Drop down Input selection
- Empirical/Determined values
- Output

Input Parameters

Gas Density	ρ_g	79.96	kg/m3	4.991	lb/ft3	
Oil phase density	ρ_o	519.8	kg/m3	32.447	lb/ft3	
Water density	ρ_w	967.6	kg/m3	60.400	lb/ft4	
Gas viscosity	μg	0.01571	cP			
Oil viscosity	μo	0.1391	cP			
Water viscosity	μw	0.3518	cP			
Oil Flow Rate	Q_O	1550.88	m^3/d	0.017950	m^3/s	
Water Flow Rate	Q_W	1645.92	m^3/d	0.019050	m^3/s	
Holdup time	T_H	5	min			
Surge time	T_s	5	min			
Mixture Volumetric Flowrate	Qm	0.52456905		18.52499864	ft3/s	
Mixture Liquid Fraction	λ	0.485135135				
Density of Mixture	ρm	293.3418378	kg/m3	18.31274425	lb/ft3	
Nozzle diameter	d_N	0.395336695	m	1.297020631	ft	
Gas flow rate	Q	5000000	sm^3/d			
Actual Gas flow rate	Q_V	0.50661905	m3/s			
DIAMETER CALCULATIONS						
Sizing constant	Ks H	0.163	m/s			
Vapour mass velocity	Vgmax (U_T)	0.38229496	m/s			
Vapour Mass velocity	U_v	0.28672122	m/s			
Operating Pressure	P	77	bar	1131.9	psi	
L/D Ratio	L/D	5				
Hold Up Volume	V_H	5.39	m^3	190.1696595	ft^3	
Surge Volume	V_s	5.39	m^3	190.1696595	ft^3	
Diameter	**D**	2.6344721	m	8.643176067	ft	
Total Cross-sectional Area	A_T	5.45101138	m^2			
Vessel Length						
Mist Eliminator Installed		YES				
Minimum Check for H_V(0.2D)		1.728635213				
Minimum Check for H_V(with or without mist eliminator)		0.6096				
Vapour Space Height	H_V	1.728635213	m			
H_V/D ratio	H_V/D	0.2				
A_V/A_T ratio	A_V/A_T	0.095320302				
	A_V	0.519592052	m^2			
Low Liquid Level Height	H_{HLL}	0.3048	m	12	in	CheckPoint
***Note if D<= 4ft then HLLL = 9in.						
H_{LLL}/D ratio	H_{LLL}/D	0.115696803				
A_{LLL}/A_T ratio	A_{LLL}/A_T	0.0482815				
Low Liquid Level Area	A_{LLL}	0.263183005	m^2			
Weir Height	H_W	0.905836887	m	2.9721	ft	CheckPoint
***Note if Hw< 2ft then increase D						
Minimum Length for Light Liquid compartment	L_2	2.307081145	m			
Min Length Check		0.700140402	m	2.29702	ft	
Interface (Hw/2) - Oil and Water Heights		50%				
	H_{LL}	0.452918443	m			
	H_{HL}	0.452918443	m			
	H_{HL}/D	0.17192				
	A_{HL}/A_T	0.079301888				
Cross-Sectional Area of Heavy Liquid	A_{HL}	0.432275494	m^2			
	A_{LL}	4.499143834	m^3			
Settling Velocity						
Settling velocity of Heavy (water) out of Light liquid (oil)	U_{HL}	0.0138664144	m/s	32.75534	in/min	
Rising velocity of Light liquid (oil) out of Heavy liquid(water)	U_{LH}	0.0054827125	m/s	12.95130	in/min	
Settling time for Water to rise out of Oil pad	t_{HL}	32.663	s			
Settling time for Oil to rise out of Water pad	t_{LH}	82.608	s			
Minimum Length for Light Liquid compartment	L_1	3.430270529	m			
Total Length	**L**	**5.737351674**	**m**			
Liquid Dropout time	ϕ	6.028975503	secs			
Actual Vapour Velocity	UVA	0.975032333	m/s			
Minimum Length for V-L Disengagement	**Lmin**	**5.878446051**	**m**			CheckPoint

***If L<Lmin set L=Lmin. If L<<Lmin then increase Hv recalculate Av.
***If L>Lmin design is acceptable.
If L>>Lmin (liquid Separation and Hold Up control) L can only be reduced and Lmin increased if Hv is reduced. Hv can only be reduced if it is greater than minimum in L2 calculation.

L/D Ratio		2.177799368	

**if L/D <<1.5 then decrease D(unless already minimum) if L/D >>6.0 increase D

Height of High Liquid Level	H_{HLL}	0.905836887	m
Area of Normal Liquid Level	A_{NLL}	2.597301166	m^2
A_{NLL}/A_T Ratio	A_{NLL}/A_T	0.476480599	
H_{NLL}/D Ratio	H_{NLL}/D	1.779665729	
Height of Normal Liquid Level	H_{NLL}	4.688479711	m

EQUATIONS

$$V_{Gmax} = K_S\sqrt{\left(\frac{\rho_L-\rho_G}{\rho_G}\right)}$$

$$D = \left(\frac{16(V_H+V_S)}{0.6\pi(L/D)}\right)^{1/3}$$

$$H_{LLL} = 0.5D + 7$$

$$H_W = D - H_V$$

$$L_2 = \frac{V_H+V_S}{A_T - A_V - A_{LLL}}$$

$$A_{LL} = A_T - A_V - A_{HL}$$

$$t_{HL} = 12H_{LL}/U_{HL}$$

$$t_{LH} = 12H_{HL}/U_{LH}$$

$$L_1 = \max\left(\frac{t_{LH}Q_{HL}}{A_{HL}}, \frac{t_{HL}Q_{LL}}{A_{LL}}\right)$$

$$L = L_1 + L_2$$

$$H_{HLL} = D - H_V$$
$$A_{NLL} = A_{LLL} + V_H/L_2$$

$$W_V = W_bL + W_I + W_N$$

Wall thickness—2:1 ellipsoidal heads

$$t = \frac{Pd}{2SE - 0.2P}$$

$$t = \frac{PR_i}{SE - 0.6P} \qquad P = \frac{SEt}{R_i + 0.6t}$$

$$t = \frac{PR_o}{SE + 0.4P} \qquad P = \frac{SEt}{R_o - 0.4t}$$

	Horizontal Vessels	Vertical Vessels
Skid Width	I.D. x 2	I.D. x 2
Skid Length	S/S x 1.5	I.D. x 2.5
Skid Height	I.D. x 2 + 1 meter	S/S x 1.5 + 1 meter

OUTPUT PARAMETERS			
Horizontal Separator Diameter	D	2.634	metres
Horizontal Separator Actual Length	L	5.878	metres
Vessel thickness	t	14.652	in
Weight of Vessel	Wv	196.174	tons
Weight of Total Skid Weight	Wskid	310.907	tons

NB. Weights do not take hydrostatic or fluid weights into consideration.

VESSEL WEIGHT CALCULATIONS

ANSI Class - Manways	-	600	
ANSI Class - Nozzles	-	600	
Nozzle Size	-	10	in
Vessel Diameter	-	8643.176	mm
Internal Description	-	Distillation Trays	
Internal Type	-	Normal	
Vessel Weight			
Internal Diameter	di	263.44721	cm
Length of Vessel	L	5.7374	m
Wall thickness (including corrosion allowance)	t	37.21534152	cm
Mass per unit length	Wb	34020.84429	kg/m
Weight of Internals	Wi	862	kg
Weight of Nozzles	WN	122.4700855	kg
Weight of Vessel	**Wv**	**196174.0181**	kg
Weight of Piping	WP	78469.60722	kg
Weight of Electrical & Instrument	WE	15693.92144	kg
Weight of Skid Steel	WS	19617.40181	kg
Weight of Manways		952	kg
Weight of Total Skid Steel	**Wskid**	**310906.9485**	kg
Vessel Thickness			
Material of vessel	-	Carbon Steel Plates and Sheets	
Specification number	-	SA-516	
Grade	-	55	
ASME Code		Div1 (-20deg F to -650 deg F)	
Working Pressure	P	1130	psi
Radius for Spherical shells	R		
Internal Radius of Shell	Ri	170.141	in
External Radius of Shell	Ro		
Maximum Allowable Stress	S	13800	psi
Joint Efficiency type	-	Fully radiographed - DW	
Joint Efficiency	E	1	
Vessel Shell Thickness	**t**	**14.6517**	in
Ellipsoidal Head wall thickness	teh	0.3568	
Vessel Footprint			
Vessel Type		Horizontal	
Skid Width		5.268944201	m
Skid Length		8.606027511	m
Skid Height		6.268944201	m
Footprint		45.34467874	m^2
Volume		284.2632608	m^3

HORIZONTAL GAS LIQUID SEPARATOR

SEPARATOR SIZING - Calculation Sheet

Colour Coding

- Input Parameter
- Drop down Input selection
- Empirical/Determined values
- Output
- Iterative value

INSTRUCTION

A. Fill in input parameters based on colour code
B. Determine **Drag co-efficient, Cd** by running the solver (error=0)---(Step 1) NB:-click on SOLVER)
C. Determine Liquid Area ratio Aw/A based on Oil pad thickness relative to Diameter Ho/D. (error=0)---(Step 2) NB:-click on SOLVER)
D. Obtain **Maximum Allowable Vessel Diameter, Dmax**
E. From **Dmax**; determine if separator length is defined by Gas or Liquid constraint based on Lg and Lo
F. Manual operation of Dand H combinations can be done manually or from ***OUTPUT REFERENCED TABLES***

INPUTS (IMPERIAL)

	NOMENCLATURE			METRIC			
Liquid phase density	ρ_l	67.84394507	lb/ft3	1086.86	kg/m3		
Gas phase density	ρ_g	2.996254682	lb/ft3	48.00	kg/m3		
Specific Gravity water	γ_w	1					
Specific Gravity Oil	γ_o	0.87					
Gas Flow rate	Qg	176.599	MMSCFD	5.00	MMscmd		
Oil Flow Rate	Qo	8062.5	bbl/day	1290	m^3/d	31.64069	ft^3/min
Water flowrate	Qw	3125	bbl/day	500	m^3/d	12.26383	ft^3/min
Gas viscosity	μg	0.015	cP				
Oil viscosity	μo	20	cP				
Pressure	P	1305.342	psi	90	Bar		
Temperature	T	500.67	R	278.15	K		
Compressibility factor	Z	0.89	-				
Drag Co-efficient (Iterative)	Cd	1.275734123	-				
Diameter of Gas Particle	dm_g	100	μm				
Diameter of Liquid Particle	dm_l	500	μm				
Gas Retention time	tg	3	mins				
Oil retention time	to	10	mins				
Water retention time	tw	10	mins				
Fraction Occupied by Gas	-	0.5					
Fraction Occupied by Liquid	-	0.5					

OUTPUT REFERENCES TABLES

Selected diameter (D)	Effective length-Gas (Lg)	Seam-to-Seam length-Gas (Ls-gas)	Effective length - liquid (Lo)	Seam-to-Seam length-Liquid (Ls-liq)	Design Constraint for Length	Gas Capacity Slenderness Ratio - (SR)	Water Settling Slenderness Ratio - (SR)
in	ft	ft	ft	ft			
36	17.16	20.16	123.36	164.47	Liquid Constraint	6.72	41.12
38	16.25	19.42	110.71	147.62	Liquid Constraint	6.13	34.96
40	15.44	18.77	99.92	133.22	Liquid Constraint	5.63	29.98
42	14.71	18.21	90.63	120.84	Liquid Constraint	5.20	25.89
44	14.04	17.70	82.58	110.10	Liquid Constraint	4.83	22.52
46	13.43	17.26	75.55	100.74	Liquid Constraint	4.50	19.71
48	12.87	16.87	69.39	92.52	Liquid Constraint	4.22	17.35
50	12.35	16.52	63.95	85.26	Liquid Constraint	3.96	15.35
52	11.88	16.21	59.12	78.83	Liquid Constraint	3.74	13.64
54	11.44	15.94	54.82	73.10	Liquid Constraint	3.54	12.18
56	11.03	15.70	50.98	67.97	Liquid Constraint	3.36	10.92
58	10.65	15.48	47.52	63.36	Liquid Constraint	3.20	9.83
60	10.29	15.29	44.41	59.21	Liquid Constraint	3.06	8.88
62	9.96	15.13	41.59	55.45	Liquid Constraint	2.93	8.05
64	9.65	14.98	39.03	52.04	Liquid Constraint	2.81	7.32
66	9.36	14.86	36.70	48.93	Liquid Constraint	2.70	6.67
68	9.08	14.75	34.57	46.10	Liquid Constraint	2.60	6.10
70	8.82	14.66	32.63	43.50	Liquid Constraint	2.51	5.59
72	8.58	14.58	30.84	41.12	Liquid Constraint	2.43	5.14
74	8.35	14.51	29.19	38.93	Liquid Constraint	2.35	4.73
76	8.13	14.46	27.68	36.90	Liquid Constraint	2.28	4.37
78	7.92	14.42	26.28	35.04	Liquid Constraint	2.22	4.04
80	7.72	14.39	24.98	33.31	Liquid Constraint	2.16	3.75
82	7.53	14.37	23.78	31.70	Liquid Constraint	2.10	3.48

STEP 1: DETERMINE Re, Cd, AND SETTLING VELOCITY

Settling velocity	u	0.488	ft/s	
Reynolds Number	Re	47.813		
Drag Co-efficient (Calculated)	Cd	1.276		
Error correction (Drag co-efficient)		0.000	SOLVER	

STEP 2: DETERMINE WATER SETTLING CAPACITY CONSTRAINTS

Area Ratio (A_w/A)	Aw/A	0.139664804	
Area Ratio (A_w/A) - Water droplet Settling Constraint	Aw/A	0.140093831	
Oil-pad Height to Diameter Ratio	Ho/D	0.255463086	
Error correction (Aw/A ratio)		0.000	SOLVER

STEP 3: DETERMINE MAXIMUM DIAMETER

Maximum Oil-Pad thickness	$H_{o,max}$	20.8	in	0.52832	m
Maximum Diameter	D_{max}	81.42076534	in	2.06808744	m
Diameter (Asssumed based on Dmax)	D	80	in	2.032	m

STEP 4: DETERMINE CONSTRAINT FOR LENGTH

Retention Time Constraint (Gas Capacity Constraint)

LgD		617.6467316	ft in		
Length (Calculated based on Dmax)	Lg	7.720584145	ft	2.353119215	m

Retention Time Constraint (Liquid Capacity Constraint)

D^2Lo		159869.375	in^2 ft		
Length (Calculated based on Dmax)	Lo	24.97958984	ft	7.61340745	m

MANUAL COMBINATION OF D AND H (refer to Output Table)

Constraint to satisfy design — Liquid Constraint

Length Based on Gas Constraint

Effective Length (gas constraint)	Lg	7.721	ft
Seam-to-seam length (Gas constraint)	Ls	14.387	ft
Slenderness Ratio	SR	2.158	

Length Based on Liquid Constraint

Effective Length (Liquid constraint)	Lo	24.980	ft
Seam-to-seam length (Liquid constraint)	Ls	33.306	ft
Slenderness Ratio	SR	4.996	

Slenderness Ratio Check (Between 3 and 5)

EQUATIONS

$$u = 0.01186\left[\left(\frac{\rho_o-\rho_g}{\rho_g}\right)\frac{d_m}{C_d}\right]^{1/2} \quad \text{ft/s}$$

$$\mathrm{Re} = 0.0049\frac{\rho_g d_m u}{\mu_g}$$

$$C_d = 0.34 + \frac{3}{\mathrm{Re}^{0.5}} + \frac{24}{\mathrm{Re}}$$

$$D^2 = 5{,}058 Q_g\left(\frac{TZ}{P}\right)\left[\frac{\rho_g}{(\rho_o-\rho_g)}\frac{C_d}{d_m}\right]^{1/2}$$

$$\frac{A_w}{A} = 0.5\frac{Q_w t_w}{Q_o t_o + Q_w t_w}$$

$$\frac{A}{A_w} = \left(\frac{1}{\pi}\right)\left[\cos^{-1}\left(\frac{2H_o}{D}\right) - \left(\frac{2H_o}{D}\right)\left(1-\frac{4H_o^2}{D^2}\right)^{-0.5}\right]$$

$$D_{\max} = \frac{H_{o,\max}}{H_o/D}$$

$$LD = 422\left(\frac{Q_g TZ}{P}\right)\left[\left(\frac{\rho_g}{\rho_o-\rho_g}\right)\left(\frac{C_d}{d_m}\right)\right]^{1/2} \text{ ft in.}$$

$$D^2L = 1.428 Q_o t \text{ ft}^3$$

$$L_s = L_g + \frac{D}{12} \text{ ft}$$

$$L_s = \frac{4}{3}L_o \text{ ft}$$

Typical retention times are as follows:

Natural gas-oil	1-3 minutes
Lean oil surge tanks	10-15 minutes
Fractionation feed surge tanks	8-15 minutes
Refrigerant surge tanks	4-7 minutes
Refrigerant economizers	2-3 minutes

API 12J gives the following guidelines for gas-oil separation.

Oil Relative Density	Minutes
Below 0.85	1
0.85-0.93	1 to 2
0.93-1.0	2 to 4

Notes

Water Settling Capacity - Determines the maximum diameter
Smallest gas particle to be separated assumed to be 100μm (in absence of Lab data)
Smallest liquid particle to be separated assumed to be 500μm (in absence of Lab data)
Retention time is between 10-30 minutes for water separating from oil
Retention time is between 1-3 minutes for gas separating from liquid
Drag co-efficient is determined by iteration
Ho/D is determined from SOLVER from Aw/A
Gas or Liquid Retention Constraint - determines the length of the vessel

Appendix C.5 Ks values for separator vessels

Reference - (Svrcek & Monnery, 1993)

Mist Eliminator (Pressure in Psia)	
1<=P<15	K = 0.1821 + 0.0029P + 0.0460Ln (P)
15<=P<=40	K = 0.35
40<=P<=5500	K = 0.430 -0.023Ln(P)

GPSA (Pressure in Psig)	
0<=P<=1500	K = 0.35 - 0.01(P-100/100)
Vapours under vacuum	K = 0.20
Glycol and Amine Solutions	Multiply K by 0.6-0.8
Vertical vessels without mist eliminators	Divide K by 2
For compressor suction scrubbers, mole sieve scrubbers and expander inlet separators multiply k by 0.7-0.8	Multiply K by 0.7-0.8

$$K = \sqrt{\frac{4gD_p}{3C_D}}$$

$C_D = \exp(Y)$

$Y = 8.411 - 2.243X + 0.273X^2 - 1.865E-2X^3 + 5.201E-4X^4$

$$X = Ln\left(\frac{0.95 + 8\rho_V D_p^3(\rho_L - \rho_V)}{\mu_V^2}\right)$$

Notes:
D_p, ft
ρ, lb/ft^3
μ, cP
1 micron = 3.28084×10^{-6} ft

Light Phase	Heavy Phase	Min Droplet Diameter, μm	Ks
Hydrocarbons		127	0.333
Sg at 60degF < 0.85	Water or Caustic	89	0.163
Sg at 60degF < 0.86	Water or Caustic	89	0.163
Water	Furfural	89	0.163

Appendix C.6 Separator vessel internals weight and nozzle weights

Vessel Internals Weight in pounds (Wi) Manways					
Vessel Diameter		Mist Eliminators		Distillation Trays	
		Vane	Mist Mat	Normal	Light Weight
mm	ft	kg	kg	kg	kg
0	0	6	5	32	23
616	2.0	6	5	32	23
770	2.5	8	7	48	34
924	3.0	10	9	73	50
1078	3.5	13	10	95	68
1232	4.0	15	12	127	91
1386	4.5	18	15	159	113
1540	5.0	21	16	200	141
1694	5.5	25	19	236	168
1848	6.0	27	21	284	200
2002	6.5	31	23	331	234
2156	7.0	34	25	386	272
2310	7.5	38	28	440	311
2464	8.0	42	31	504	354
2618	8.5	47	34	563	397
2772	9.0	53	36	635	445
2926	9.5	57	39	703	499
3080	10.0	62	42	794	553
3234	10.5	62	45	862	608

External Nozzle Weights in kg (WN)											
ANSI	Nominal Nozzle Sizes (DN)										
Class	2	3	4	6	8	10	12	14	16	18	20
150	10	30	45	65	100	140	185	240	320	345	410
300	15	30	55	95	130	170	245	325	440	565	670
400	20	40	70	100	150	205	295	370	490	580	705
600	20	40	75	120	180	270	330	485	675	825	1020.00

Appendix C.7 Liquid holdup and surge times

LIQUID HOLDUP AND SURGE TIMES			
SERVICES	Holdup Times NLL-HLL min	Surge Times NLL-LLL min	
A. UNIT FEED DRUM	10	5	
B. SEPARATORS			
1. Feed to Column	5	3	
2. Feed to other drum or Tankagge			
a. with pump or through exchanger	5	2	
b. without pump	2	1	
3. Feed to Fire Heater	10	3	
C. REFLUX OR PRODUCT ACCUMULATOR			
1. Reflux only	3	2	
2. Reflux and Product	3+	2+	
*based on reflux (3min) + appropriate holdup time of overhead product (per B.1,2,3)			
D. COLUMN BOTTOMS			
1. Feed to another column	5	2	
2. Feed to other drum or Tankagge			
a. with pump or through exchanger	5	2	
b. without pump	2	1	
3. Feed to Fired boiler	5-8	2-4	
*based on reboiler vapour expressed as liquid (3min) + appropriate holdup time for the bottom product (per D.1,2,)			
E. COMPRESSOR SUCTION /INTERSTAGE SCRUBBER			
3min between high liquid alarm (HLL/HLA) and high level shutdown (HLSD)			
10min from bottom tangent line to high liquid alarm			
F. FUEL GAS KNOCKOUT DRUM			
20ft slug in the incoming fuel gas line between NLL and HLSD			
F. FLARE KNOCKOUT DRUM			
20 to 30min to HLL			
Personnel	**factor**	**Instrumentation**	**factor**
Experienced	1.0	Well instrumented	1.0
Trained	1.2	Standard instrumented	1.2
Inxperienced	1.5	Poorly instrumented	1.5

Appendix C.8 Low liquid level height

Vessel Diameter	Vertical LLL		Horizontal LLL
	<300 psia	>300 psia	
<=4ft	15 in	6 in	9 in
6ft	15 in	6 in	10 in
8ft	15 in	6 in	11 in
10ft	6 in	6 in	12 in
12ft	6 in	6 in	13 in
16ft	6 in	6 in	15 in

Appendix C.9 L/D ratio guidelines

Vessel Operating pressure, psig	L/D
0<P<=250	1.5 - 3.0
250<P<500	3.0 - 4.0
500<P	4.0 - 6.0

Appendix C.10 Cylindrical height and area conversions

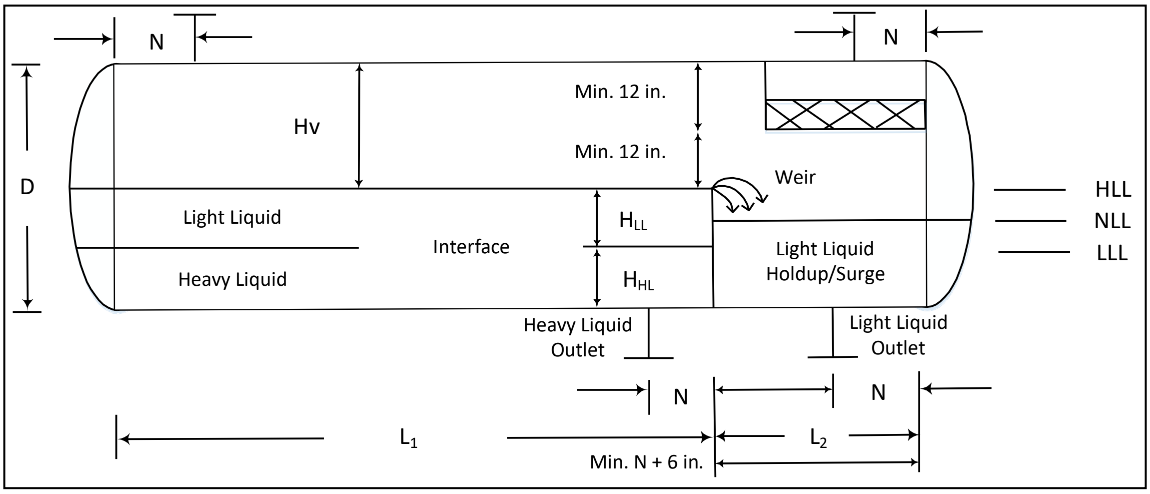
a	4.76E-05	0.00153756
b	3.924091	26.787101
c	0.174875	3.299201
d	-6.358805	-22.923932
e	5.668973	24.353518
f	4.018448	-14.844824
g	-4.916411	-36.999376
h	-1.801705	10.529572
i	-0.145348	9.892851

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

For **H/D to A/A$_T$** ; $y = \frac{A}{A_T}$ and $x = \frac{H}{D}$;

For **A/A$_T$** to **H/D**; $y = \frac{H}{D}$ and $x = \frac{A}{A_T}$

Appendix C.11 3-phase horizontal separator design

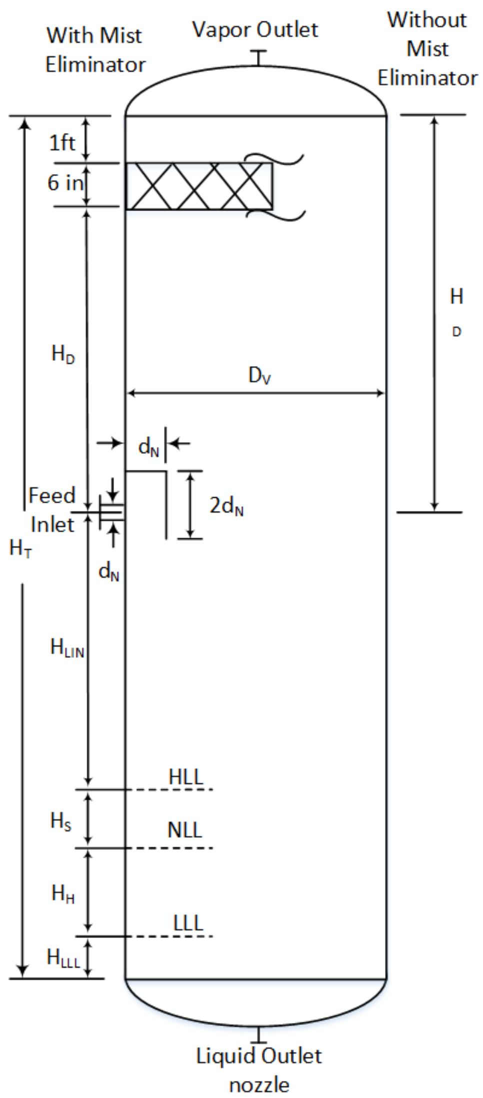


H_V	-	Vapour Space Height	**d_N**	-	Nozzle Diameter
H_{LLL}	-	Low Liquid Level Height	**D**	-	Diameter
H_W	-	Weir Height	**L_1**	-	Minimum Length for Light and heavy Liquid compartment
H_{LL}	-	Light Liquid Height	**L_2**	-	Minimum Length for Light Liquid compartment
H_{HL}	-	Heavy Liquid Height	**L**	-	Total Length

1st stage 3-phase separator design

3-Phase Horizontal (units in metres)	**H_V**	**H_{LLL}**	**H_W**	**H_{LL}**	**H_{HL}**	**d_N**	**D**	**L_1**	**L_2**	**L**
1st Stage 3-Phase Separator	1.76	0.30	0.92	0.46	0.46	0.16	2.68	4.02	3.41	7.43

Appendix C.12 2-phase vertical separator design



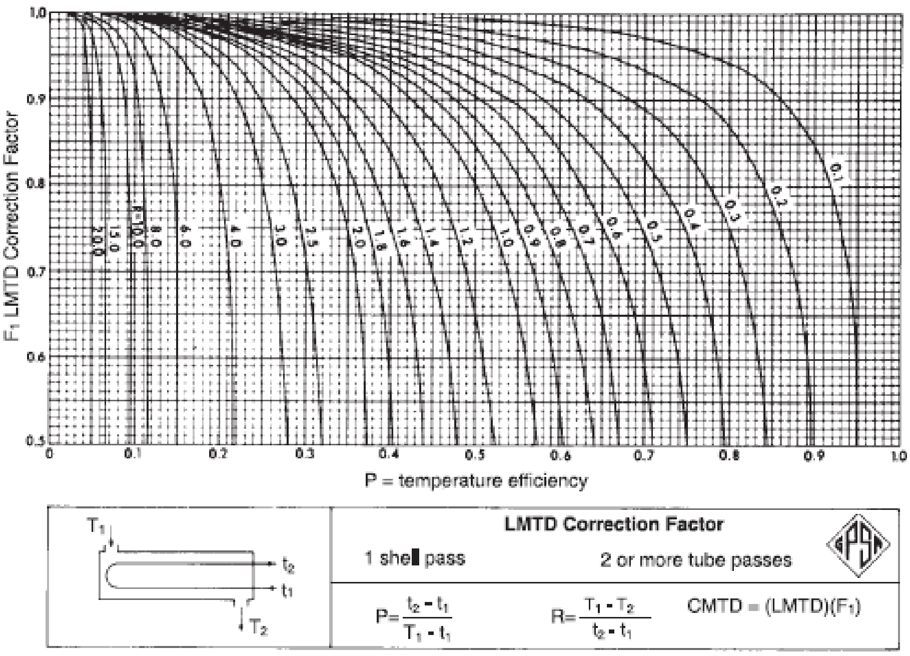
H_{LLL}	-	Low Liquid Level Height
H_H	-	Height from Normal Liquid Level (NLL) to High Liquid Level (HLL)
H_S	-	Surge Height
H_{LIN}	-	Height from HLL to Inlet nozzle centreline
H_D	-	Disengagement Height
H_{ME}	-	Extra length
d_N	-	Nozzle Diameter
D_V	-	Diameter
H_T	-	Total Height

2-Phase Vertical (units in metres)	H_{LLL}	H_H	H_S	H_{LIN}	H_D	H_{ME}	d_N	D	H_T
Inlet Separator	0.15	2.00	1.20	1.12	1.32	0.46	0.81	1.70	7.61
2nd Stage Separator	0.15	2.40	1.44	1.01	1.26	0.46	0.70	1.26	7.64
3rd Stage Separator	0.15	1.20	0.72	1.17	1.35	0.46	0.87	1.62	5.53
2nd Stage LP Compressor Scrubber	0.15	0.35	0.21	0.53	1.03	0.46	0.23	0.78	2.92
Intermediate Separator	0.15	4.00	2.40	0.66	1.09	0.46	0.35	0.59	13.24
Dehydration Scrubber*	0.15	0.00	0.00	0.60	1.06	0.46	0.29	1.57	2.27
1st St. HP Comp Scrubber*	0.15	0.00	0.00	0.61	1.07	0.46	0.31	1.64	2.29
2nd St. HP Comp Scrubber*	0.15	0.00	0.00	0.58	1.05	0.46	0.27	1.49	2.24

**In the process, these separators do not have a liquid phase. Separator was given a conservative design assuming liquid phase of density 1000kg/m^3 and a flowrate of 0.000001m^3/s*

Appendix D Heat exchangers

Appendix D.1 LMTD correction factor (1 shell pass; 2 or more tube passes)



Appendix D.2 Tubing characteristics (courtesy of TEMA)

Tube O.D. Inches	B.W.G. Gauge	Thickness Inches	Internal Area Sq. Inch	Sq Ft External Surface Per Foot Length	Sq Ft Internal Surface Per Foot Length	Weight Per Foot Length Steel Lbs*	Tube I.D. Inches	Moment of Inertia (Inches^4)	Section Modulus (Inches^3)	Radius of Gyration (Inches)	Constant C**	$\frac{\text{O.D.}}{\text{I.D.}}$	Transverse Metal Area Sq. Inch
1/4	22	0.028	0.0296	0.0654	0.0508	0.066	0.194	0.00012	0.00098	0.0791	46	1.289	0.0195
1/4	24	0.022	0.0333	0.0654	0.0539	0.054	0.206	0.00010	0.00083	0.0810	52	1.214	0.0158
1/4	26	0.018	0.0360	0.0654	0.0560	0.045	0.214	0.00009	0.00071	0.0823	56	1.168	0.0131
1/4	27	0.016	0.0373	0.0654	0.0571	0.040	0.218	0.00008	0.00065	0.0829	58	1.147	0.0118
3/8	18	0.049	0.0603	0.0982	0.0725	0.171	0.277	0.00068	0.0036	0.1166	94	1.354	0.0502
3/8	20	0.035	0.0731	0.0982	0.0798	0.127	0.305	0.00055	0.0029	0.1208	114	1.230	0.0374
3/8	22	0.028	0.0799	0.0982	0.0835	0.104	0.319	0.00046	0.0025	0.1231	125	1.176	0.0305
3/8	24	0.022	0.0860	0.0982	0.0867	0.083	0.331	0.00038	0.0020	0.1250	134	1.133	0.0244
1/2	16	0.065	0.1075	0.1309	0.0969	0.302	0.370	0.0021	0.0086	0.1555	168	1.351	0.0888
1/2	18	0.049	0.1269	0.1309	0.1052	0.236	0.402	0.0018	0.0071	0.1604	198	1.244	0.0694
1/2	20	0.035	0.1452	0.1309	0.1126	0.174	0.430	0.0014	0.0056	0.1649	227	1.163	0.0511
1/2	22	0.028	0.1548	0.1309	0.1162	0.141	0.444	0.0012	0.0046	0.1672	241	1.126	0.0415
5/8	12	0.109	0.1301	0.1636	0.1066	0.601	0.407	0.0061	0.0197	0.1865	203	1.536	0.177
5/8	13	0.095	0.1486	0.1636	0.1139	0.538	0.435	0.0057	0.0183	0.1904	232	1.437	0.158
5/8	14	0.083	0.1655	0.1636	0.1202	0.481	0.459	0.0053	0.0170	0.1939	258	1.362	0.141
5/8	15	0.072	0.1817	0.1636	0.1259	0.426	0.481	0.0049	0.0156	0.1972	283	1.299	0.125
5/8	16	0.065	0.1924	0.1636	0.1296	0.389	0.495	0.0045	0.0145	0.1993	300	1.263	0.114
5/8	17	0.058	0.2035	0.1636	0.1333	0.352	0.509	0.0042	0.0134	0.2015	317	1.228	0.103
5/8	18	0.049	0.2181	0.1636	0.1380	0.302	0.527	0.0037	0.0119	0.2044	340	1.186	0.089
5/8	19	0.042	0.2299	0.1636	0.1416	0.262	0.541	0.0033	0.0105	0.2067	359	1.155	0.077
5/8	20	0.035	0.2419	0.1636	0.1453	0.221	0.555	0.0028	0.0091	0.2090	377	1.126	0.065
3/4	10	0.134	0.1825	0.1963	0.1262	0.833	0.482	0.0129	0.0344	0.2229	285	1.556	0.259
3/4	11	0.120	0.2043	0.1963	0.1335	0.809	0.510	0.0122	0.0326	0.2267	319	1.471	0.238
3/4	12	0.109	0.2223	0.1963	0.1393	0.747	0.532	0.0116	0.0309	0.2299	347	1.410	0.219
3/4	13	0.095	0.2463	0.1963	0.1466	0.665	0.560	0.0107	0.0285	0.2340	384	1.339	0.195
3/4	14	0.083	0.2679	0.1963	0.1529	0.592	0.584	0.0098	0.0262	0.2376	418	1.284	0.174
3/4	15	0.072	0.2884	0.1963	0.1587	0.522	0.606	0.0089	0.0238	0.2411	450	1.238	0.153
3/4	16	0.065	0.3019	0.1963	0.1623	0.476	0.620	0.0083	0.0221	0.2433	471	1.210	0.140
3/4	17	0.058	0.3157	0.1963	0.1660	0.429	0.634	0.0076	0.0203	0.2455	492	1.183	0.126
3/4	18	0.049	0.3339	0.1963	0.1707	0.367	0.652	0.0067	0.0178	0.2484	521	1.150	0.108
3/4	20	0.035	0.3632	0.1963	0.1780	0.268	0.680	0.0050	0.0134	0.2531	567	1.103	0.079
7/8	10	0.134	0.2894	0.2291	0.1589	1.062	0.607	0.0221	0.0505	0.2662	451	1.442	0.312
7/8	11	0.120	0.3167	0.2291	0.1662	0.969	0.635	0.0208	0.0475	0.2703	494	1.378	0.285
7/8	12	0.109	0.3390	0.2291	0.1720	0.893	0.657	0.0196	0.0449	02736	529	1.332	0.262
7/8	13	0.095	0.3685	0.2291	0.1793	0.792	0.685	0.0180	0.0411	0.2778	575	1.277	0.233
7/8	14	0.083	0.3948	0.2291	0.1856	0.703	0.709	0.0164	0.0374	0.2815	616	1.234	0.207
7/8	15	0.072	0.4197	0.2291	0.1914	0.618	0.731	0.0148	0.0337	0.2850	655	1.197	0.182
7/8	16	0.065	0.4359	0.2291	0.1950	0.563	0.745	0.0137	0.0312	0.2873	680	1.174	0.165
7/8	17	0.058	0.4525	0.2291	0.1987	0.507	0.759	0.0125	0.0285	0.2896	706	1.153	0.149
7/8	18	0.049	0.4742	0.2291	0.2034	0.433	0.777	0.0109	0.0249	0.2925	740	1.126	0.127
7/8	20	0.035	0.5090	0.2291	0.2107	0.314	0.805	0.0082	0.0187	0.2972	794	1.087	0.092
1	8	0.165	0.3526	0.2618	0.1754	1.473	0.670	0.0392	0.0784	0.3009	550	1.493	0.433
1	10	0.134	0.4208	0.2618	0.1916	1.241	0.732	0.0350	0.0700	0.3098	656	1.366	0.365
1	11	0.120	0.4536	0.2618	0.1990	1.129	0.760	0.0327	0.0654	0.3140	708	1.316	0.332
1	12	0.109	0.4803	0.2618	0.2047	1.038	0.782	0.0307	0.0615	0.3174	749	1.279	0.305
1	13	0.095	0.5153	0.2618	0.2121	0.919	0.810	0.0280	0.0559	0.3217	804	1.235	0.270
1	14	0.083	0.5463	0.2618	0.2183	0.814	0.834	0.0253	0.0507	0.3255	852	1.199	0.239
1	15	0.072	0.5755	0.2618	0.2241	0.714	0.856	0.0227	0.0455	0.3291	898	1.168	0.210
1	16	0.065	0.5945	0.2618	0.2278	0.650	0.870	0.0210	0.0419	0.3314	927	1.149	0.191
1	18	0.049	0.6390	0.2618	0.2361	0.498	0.902	0.0166	0.0332	0.3367	997	1.109	0.146
1	20	0.035	0.6793	0.2618	0.2435	0.361	0.930	0.0124	0.0247	0.3414	1060	1.075	0.106
1 1/4	7	0.180	0.6221	0.3272	0.2330	2.059	0.890	0.0890	0.1425	0.3836	970	1.404	0.605
1 1/4	8	0.165	0.6648	0.3272	0.2409	1.914	0.920	0.0847	0.1355	0.3880	1037	1.359	0.565
1 1/4	10	0.134	0.7574	0.3272	0.2571	1.599	0.982	0.0742	0.1187	0.3974	1182	1.273	0.470
1 1/4	11	0.120	0.8012	0.3272	0.2644	1.450	1.010	0.0688	0.1100	0.4018	1250	1.238	0.426
1 1/4	12	0.109	0.8365	0.3272	0.2702	1.330	1.032	0.0642	0.1027	0.4052	1305	1.211	0.391
1 1/4	13	0.095	0.8825	0.3272	0.2775	1.173	1.060	0.0579	0.0926	0.4097	1377	1.179	0.345
1 1/4	14	0.083	0.9229	0.3272	0.2838	1.036	1.084	0.0521	0.0833	0.4136	1440	1.153	0.304
1 1/4	16	0.065	0.9852	0.3272	0.2932	0.824	1.120	0.0426	0.0682	0.4196	1537	1.116	0.242
1 1/4	18	0.049	1.0423	0.3272	0.3016	0.629	1.152	0.0334	0.0534	0.4250	1626	1.085	0.185
1 1/4	20	0.035	1.0936	0.3272	0.3089	0.455	1.180	0.0247	0.0395	0.4297	1706	1.059	0.134
1 1/2	10	0.134	1.1921	0.3927	0.3225	1.957	1.232	0.1354	0.1806	0.4853	1860	1.218	0.575
1 1/2	12	0.109	1.2908	0.3927	0.3356	1.621	1.282	0.1159	0.1545	0.4933	2014	1.170	0.476
1 1/2	14	0.083	1.3977	0.3927	0.3492	1.257	1.334	0.0931	0.1241	0.5018	2180	1.124	0.369
1 1/2	16	0.065	1.4741	0.3927	0.3587	0.997	1.370	0.0756	0.1008	0.5079	2300	1.095	0.293
2	11	0.120	2.4328	0.5236	0.4608	2.412	1.760	0.3144	0.3144	0.6660	3795	1.136	0.709
2	12	0.109	2.4941	0.5236	0.4665	2.204	1.782	0.2904	0.2904	0.6697	3891	1.122	0.648
2	13	0.095	2.5730	0.5236	0.4739	1.935	1.810	0.2586	0.2586	0.6744	4014	1.105	0.569
2	14	0.083	2.6417	0.5236	0.4801	1.701	1.834	0.2300	0.2300	0.6784	4121	1.091	0.500

* Weights are based on low carbon steel with a density of 0.2833 lbs/cu in. For other metals multiply by the following factors:

Aluminum	0.35	Aluminum Bronze	1.04	Nickel	1.13
Titanium	0.58	Aluminum Brass	1.06	Nickel-Copper	1.12
A.I.S.I. 400 Series Stainless Steels	0.99	Nickel-Chrome-Iron	1.07	Copper and Cupro-Nickels	1.14
A.I.S.I. 300 Series Stainless Steels	1.02	Admiralty	1.09		

** Liquid Velocity $= \dfrac{\text{Lbs Per (Tube} \cdot \text{Hour)}}{\text{(C) (Sp Gr of Liquid)}}$ in feet per sec (Sp Gr of Water at 60°F = 1.0)

Courtesy of TEMA

Appendix D.3 Typical film heat transfer co-efficients for shell and tube heat exchangers factor (Courtesy of HEDH : Heat Exchanger Design Handbook 2002)

3.1.4-4 3.1 INTRODUCTION TO HEAT EXCHANGER DESIGN / 3.1.4 Sizing of Shell-and-Tube Heat Exchangers

Table 2 Typical film heat transfer coefficients for shell-and-tube heat exchangers

Fluid conditions		α, W/m² K[a, b]	Fouling resistance, m² K/W[a]
Sensible heat transfer			
Water[c]	Liquid	5 000–7 500	1×10^{-4}–2.5×10^{-4}
Ammonia	Liquid	6 000–8 000	0–1×10^{-4}
Light organics[d]	Liquid	1 500–2 000	1×10^{-4}–2×10^{-4}
Medium organics[e]	Liquid	750–1 500	1.5×10^{-4}–4×10^{-4}
Heavy organics[f]	Liquid,		
	Heating	250–750	2×10^{-4}–1×10^{-3}
	Cooling	150–400	2×10^{-4}–1×10^{-3}
Very heavy organics[g]	Liquid,		
	Heating	100–300	4×10^{-4}–3×10^{-3}
	Cooling	60–150	4×10^{-4}–3×10^{-3}
Gas[h]	Pressure 100–200 kN/m² abs	80–125	0–1×10^{-4}
Gas[h]	Pressure 1 MN/m² abs	250–400	0–1×10^{-4}
Gas[h]	Pressure 10 MN/m² abs	500–800	0–1×10^{-4}
Condensing heat transfer			
Steam, ammonia	Pressure 10 kN/m² abs, no noncondensables[i, j]	8 000–12 000	0–1×10^{-4}
Steam, ammonia	Pressure 10 kN/m² abs, 1% noncondensables[k]	4 000–6 000	0–1×10^{-4}
Steam, ammonia	Pressure 10 kN/m² abs, 4% noncondensables[k]	2 000–3 000	0–1×10^{-4}
Steam, ammonia	Pressure 100 kN/m² abs, no condensables[i, j, k, l]	10 000–15 000	0–1×10^{-4}
Steam, ammonia	Pressure 1 MN/m² abs, no condensables[i, j, k, l]	15 000–25 000	0–1×10^{-4}
Light organics[d]	Pure component, pressure 10 kN/m² abs, no non-condensables[i]	1 500–2 000	0–1×10^{-4}
Light organics[d]	Pressure 10 kN/m² abs, 4% noncondensables[k]	750–1 000	0–1×10^{-4}
Light organics[d]	Pure component, pressure 100 kN/m² abs, no non-condensables[l]	2 000–4 000	0–1×10^{-4}
Light organics[d]	Pure component, pressure 1 MN/m² abs	3 000–7 000	0–1×10^{-4}
Medium organics[e]	Pure component or narrow condensing range, pressure 100 kN/m² abs[m, n]	1 500–4 000	1×10^{-4}–3×10^{-4}
Heavy organics	Narrow condensing range, pressure 100 kN/m² abs[m, n]	600–2 000	2×10^{-4}–5×10^{-4}
Light multicomponent mixtures, all condensable[d]	Medium condensing range, pressure 100 kN/m² abs[k, m, o]	1 000–2 500	0–2×10^{-4}
Medium multicomponent mixtures, all condensable	Medium condensing range, pressure 100 kN/m² abs[k, m, o]	600–1 500	1×10^{-4}–4×10^{-4}
Heavy multicomponent mixtures, all condensable[f]	Medium condensing range, pressure 100 kN/m² abs[k, m, o]	300–600	2×10^{-4}–8×10^{-4}
Vaporizing heat transfer[p, q]			
Water[r]	Pressure < 0.5 MN/m² abs, $\Delta T_{SH,max} = 25$ K	3 000–10 000	1×10^{-4}–2×10^{-4}
Water[r]	Pressure > 0.5 MN/m² abs, pressure < 10 MN/m² abs, $\Delta T_{SH,max} = 20$ K	4 000–15 000	1×10^{-4}–2×10^{-4}
Ammonia	Pressure < 3 MN/m² abs, $\Delta T_{SH,max} = 20$ K	3 000–5 000	$1 N 10^{-4}$–2×10^{-4}

(See footnotes on page 3.1.4-5.)

Appendix D.4 Heat exchanger calculator

HEAT EXCHANGER

HEAT EXCHANGER SIZING - Calculation Sheet

Colour Coding	
Input Parameter	
Drop down Input selection	
Empirical/Determined values	
Output	
Look up table	

STEP 1: INPUT PARAMETERS, SHELL AND TUBE SIDE CONDITIONS

		TUBE	UNITS	SHELL	
Fluid		**GAS**		**SEA WATER**	
Density of fluid	ρ	23.01	kg/m^3	1022	
Mass Flow	ṁ	1.901	kg/s	7.29	
Specific Heat Capacity	Cp	2.297	kJ/kg.K	4.451	
Temperature In	K	409.65	K	278.15	
Temperature Out	K	298.15	K	293.15	
Fouling Factor	Rf	1.00E-04		1.50E-04	Assumed const
Film Transfer Co-efficient	h	5.00E+02	W/m^2K	5.00E+03	Assumed const
Duty	Q	486.8755655	kW	486.8755655	

Logarithmic Temperature difference	LMTD	48.10629711	K
Correction Factor (countercurrent)	F	1	
Corrected LMTD	LMTD	48.10629711	

STEP 2: TUBE SIDE DESIGN PARAMETERS

Assumed overall U (Tube Side)					
Tube Size	OD	0.01905	m	0.75	inches
Wall Thickness	BWG	14			
Wall Thickness (Based on BWG selected)		0.002108	m		
Tube Size	ID	0.014834	m		
Tube Length		2.438	m	8ft assumption	
Number of Tube passes		1		Number of Passes (1-pass or even no. up to 14)	
Tube Length per pass		2.438			
Assumed Pitch Ratio (Pt/d)		1.25		assumed	
Tube Pitch		0.0238125	m		
Cross-Sectional Area of Tube		0.000172825	m^2		
Area of a single tube		0.113616616	m^2		

STEP 3: DETERMINE U AND AREA REQUIREMENTS

Assumed Overall Heat Transfer Co-efficient	U	408.1632653	W/m^2K	
Transfer Area	A	24.79602894	m^2	
Total Number of Tubes		218.2429814		
Number of Tubes per pass	N_{tubes}	219		
Total Tube Area per pass		0.037848663	m^2	
Volumetric Flow		0.082616254	m^3/s	
Fluid Velocity per pass		2.182805064	m/s	Adjust Tube Size/Length etc to obtain fluid velocity

STEP 4: SHELL SIDE DESIGN PARAMETERS

Tube Pattern		Triangular		
Area Tube		0.00049	m^2	
Diameter of Area	D_{tight}	0.37004		
Corrected Area	Acorrected	0.10754	m^2	
Minimum Shell Inside Diameter	$D_{s,min}$	0.40814	m	17 in
Head Lengths				
Shell Length		2.43800		
Length to Diameter Ratio	L/D	5.30561		

EQUATIONS

$$U_{o,fouled} = \frac{1}{\frac{d_o}{d_i\, h_i} + \frac{d_o\, R_{f,i}}{d_i} + \frac{d_o \ln(d_o/d_i)}{2\,k} + R_{f,o} + \frac{1}{h_o}}$$

$$Area_{1\ tube,\ triangular} = 2\,(PR\, d_o)^2\, \frac{\sqrt{3}}{4}$$

$$Area_{1\ tube,\ square} = (PR\, d_o)^2$$

$$D_{tight} = 2\left(\frac{N_t\, Area_{tube}}{\pi}\right)^{0.5}$$

$$A_{corrected} = D_{tight}\, d_o\,(n_p - 1) + (N_t\, Area_{tube})$$

$$D_{s,\min} = 2\left(\frac{A_{corrected}}{\pi}\right)^{0.5} + 2\,d_o$$

$$t_s = \frac{pD_s}{fJ - 0.6p} + c \qquad (2.1)$$

t_s=shell thickness

p= design pressure

D_s= Shell ID

f=Maximum allowable stress of the material of construction

J=Joint efficiency (usually varies from 0.7 to 0.9)

The minimum shell thicknesses should be decided in compliance with the nominal shell diameter including the corrosion allowance as specified by IS: 4503. Usually the

OUTPUT PARAMETERS

Tube Diameter	d_T	0.015	metres
Number of Tubes		219.00	
Shell Inside Diameter	D_s	0.408	metres
Heat Exchanger Length	L	2.438	in
Weight of Vessel	Wv	1.248	tons
Weight of Total Skid Weight	Wskid	2.441	tons

NB. Weights do not take hydrostatic or fluid weights into consideration.

VESSEL WEIGHT CALCULATIONS

Density of Steel	**kg/m^3**	**7841.717**

Tube Weight			
Weight Parameter		Weight Per meter	
Tube Weight per meter		0.8809848	
Total Tube Weight		468.7512	kg
Shell Weight			
Presssure Rating	P	90	bar
Shell Diameter	$D_{s,min}$	0.40814	m
Maximum Allowable Stress of Material	f	938.78	bar
Joint Efficiency type		Spot radiographed - DW	
Joint Efficiency	J	0.85	
Corrosion allowance (2mm)	C	0.002	m
Shell wall thickness	t	0.051374354	m
Length of Shell (seam-to-seam)		2.438	m
Shell ID		0.40814	m
Shell OD		0.45951	m
Shell Volume		0.085352382	m^3
Weight of Empty Shell		669.3092276	kg
Vessel Externals			
Head Weights		133.6224476	kg
Flange Weights		152	kg
Internals Weight (Baffles)			
Baffle Cut (window height to ID -25-35%)		30%	
Baffle Spacing (usually 40-60%) of ID)		50%	
Tube Length	L	2.4380	m
Central baffle Spacing	$L_{b,c}$	0.2041	m
Baffle Outlet Spacing	$L_{b,o}$	0.2245	m
Baffle Inlet Spacing	$L_{b,i}$	0.2245	m
Number of Baffles	N_b	11	
Clearance betweenn Baffles and Shell		85%	
Total Baffle Weight		293.2189676	kg
Nozzle Weight			
Empty Vessel Weight		**1248.150643**	**kg**
Weight of Piping	Wp	499.2602571	kg
Weight of Electrical & Instrument	WE	99.85205142	kg
Weight of Skid Steel	Ws	124.8150643	kg
Weight of Total Skid Steel	**Wskid**	**2440.8292**	
Skid Width		0.816	m
Skid Length		3.657	m
Skid Height		1.919	m

Appendix E Compressor data

Appendix E.1 Compressor specification data (courtesy of Elliot)

COMPRESSOR FRAME SUMMARIES

Frame	Max Flow		Impeller Diameter		Speed	Configuration	Casing Rating	
	m^3/hr	CFM	mm	in	RPM	M (Horizontal Split) MB (Vertical Split)	BARG	PSIG
10	10,900	6,400	264	10.4	19,800	M MB	69 345	1,000 5,000
15	14,400	8,500	303	11.9	17,300	M MB	69 689	1,000 10,000
20	19,000	11,200	348	13.7	15,000	M MB	69 689	1,000 10,000
25	25,000	14,700	401	15.8	13,100	M MB	69 689	1,000 10,000
29	33,000	19,400	461	18.2	11,400	M MB	69 689	1,000 10,000
32	43,500	25,600	530	20.9	9,900	M MB	69 345	1,000 5,000
38	57,400	33,800	610	24.0	8,600	M MB	69 345	1,000 5,000
46	75,900	44,700	701	27.6	7,500	M MB	69 217	1,000 3,150
56	100,200	59,000	806	31.7	6,500	M MB	69 217	1,000 3,150
60	132,500	78,000	927	36.5	5,600	M MB	69 138	1,000 2,000
70	175,000	103,000	1,066	42.0	4,900	M MB	52 103	750 1,500
78	231,100	136,000	1,226	48.3	4,300	M MB	41 103	600 1,500
88	305,800	180,000	1,410	55.5	3,700	M MB	41 69	600 1,000
103	404,400	238,000	1,622	63.8	3,200	M	26	380
110	535,200	315,000	1,865	73.4	2,800	M	7	100

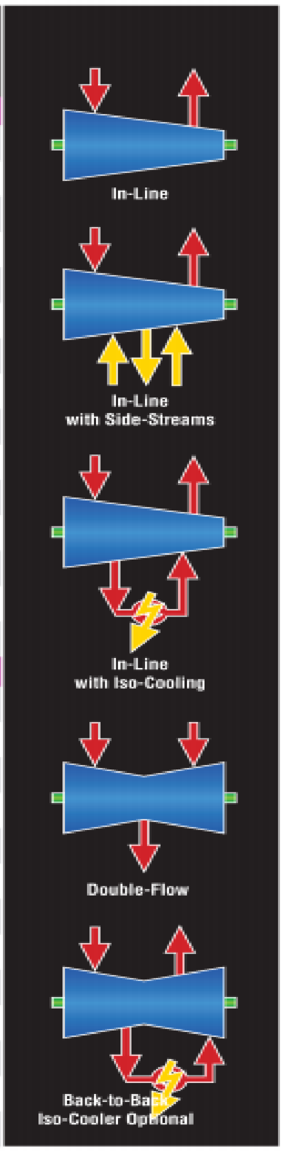
Note: Table does not include specifications for double flow configuration.

Standardization of Components

In developing the EDGE compressor product line, we focused on standardizing components and hardware to reduce costs and improve reliability across a wide array of applications. The EDGE product line consists of 15 standard frame sizes, which are scaled from the 38M median frame size. Casing bores and internal aerodynamic hardware, such as impellers, diaphragms, and shafts, are scaled. Scaling aerodynamic components improves performance predictability and increases reliability by preserving geometric similarity across frame sizes. Bearings and seals are selected from vendor standard sizes for each application.

Weights, Dimensions and Configurations

Frame Size	Minimum Rotor Length (in. / mm)	Maximum Rotor Length (in. / mm)	Casing Width (includes supports) (in. / mm)	Casing Height (exc. supports) (in. / mm)	Minimum Casing Weight (lb/kg)	Maximum Casing Weight (lb / kg)
Typical Weights and Dimensions for Elliott Horizontal Split Compressors*						
10M	35 / 890	64 / 1,625	37.3 / 947.4	36.25 / 921	4,700 / 2,130	9,000 / 4,080
15M	35 / 890	75 / 1,905	42.5 / 1,080	41.38 / 1,051	5,600 / 2,540	12,700 / 5,760
20M	40 / 1,015	80 / 2,030	48.5 / 1,232	47.85 / 1,215	8,200 / 3,720	18,000 / 8,165
25M	45 / 1,145	90 / 2,285	55.1 / 1,400	54.12 / 1,375	11,100 / 5,035	24,500 / 11,100
29M	50 / 1,270	110 / 2,795	65.4 / 1,661	59.25 / 1,505	14,000 / 6,350	32,000 / 14,500
32M	50 / 1,270	124 / 3,150	70.5 / 1,791	65.75 / 1,670	15,700 / 7,120	45,000 / 20,400
38M	55 / 1,400	135 / 3,430	76.3 / 1,938	70.62 / 1,794	23,000 / 10,430	62,000 / 28,100
46M	70 / 1,780	155 / 3,940	66.5 / 1,689	89.5 / 2,273	32,500 / 14,740	87,000 / 39,500
56M	80 / 2,032	175 / 4,445	76 / 1,930	93.38 / 2,372	51,500 / 23,360	127,000 / 57,600
60M	90 / 2,285	190 / 4,825	89.7 / 2,278	97 / 2,464	59,000 / 26,760	170,000 / 77,100
70M	100 / 2,540	230 / 5,840	103.5 / 2,629	113.88 / 2,893	71,000 / 34,000	210,000 / 95,250
78M	100 / 2,540	250 / 6,350	109.5 / 2,781	125.5 / 3,188	95,000 / 43,100	295,000 / 133,800
88M	115 / 2,920	275 / 6,985	133 / 3,378	137.5 / 3,492	130,000 / 59,000	380,000 / 172,400
103M	135 / 3,429	300 / 7,620	156 / 3,962	158.5 / 4,025	215,000 / 97,500	525,000 / 238,100
110M	140 / 3,556	325 / 8,255	182 / 4,623	182 / 4,630	270,000 / 122,470	690,000 / 312,980
Typical Weights and Dimensions for Elliott Vertical Split Compressors*						
10MB	35 / 890	62 / 1,575	43 / 1,092	42.5 / 1,080	7,000 / 3,175	13,000 / 5,900
15MB	35 / 890	72 / 1,830	46 / 1,168	48 / 1,219	8,400 / 3,810	17,500 / 7,940
20MB	40 / 1,015	80 / 2,030	50 / 1,270	53.75 / 1,366	12,000 / 5,440	25,000 / 11,340
25MB	45 / 1,145	88 / 2,235	58.5 / 1,486	62 / 1,575	18,400 / 8,345	36,000 / 163,030
29MB	50 / 1,270	105 / 2,670	64.3 / 1,633	64 / 1,626	23,000 / 10,435	49,000 / 22,225
32MB	50 / 1,270	120 / 3,050	71.7 / 1,821	76.5 / 1,943	28,500 / 12,900	69,000 / 31,300
38MB	55 / 1,400	130 / 3,300	78.5 / 1,994	83.25 / 2,115	36,500 / 16,560	89,000 / 40,400
46MB	70 / 1,780	150 / 3,810	96.5 / 2,451	86.5 / 2,197	47,500 / 21,500	115,000 / 52,200
56MB	80 / 2,030	170 / 4,320	104.2 / 2,647	102.12 / 2,594	70,000 / 31,750	160,000 / 72,600
60MB	90 / 1,525	185 / 4,700	113 / 2,870	112.5 / 2,858	90,000 / 41,000	200,000 / 91,000
70MB	100 / 2,540	225 / 5,715	115.2 / 2,926	120.96 / 3,064	100,000 / 45,350	251,000 / 113,900
78MB	100 / 2,540	245 / 6,225	120 / 3,048	140 / 3,556	125,000 / 56,700	315,000 / 143,000
88MB	115 / 2,920	265 / 6,730	137 / 3,480	148 / 3,759	205,000 / 93,000	465,000 / 211,000



Appendix E.2 Compressor calculator

Compressor Sizing

Colour Coding	
(yellow)	Input Parameter
(light blue)	Empirical/Determined values
(orange)	Drop down
(light green)	Output

COMPRESSOR SIZING - Calculation Sheet

Question: which enthalpy -- there is mass enthalpy and molarenthaply

273.15
136.5
409.65

INPUTS PARAMETERS

		Suction (1)	Units	Discharge (2)
Pressure	P	120	bar	200
Temperature	T	30	°C	70.86
		303.15	K	344.01
Density	ρ	127.2	kg/m^3	165.3
Specific Volume	v	0.007861635	m^3/kg	0.006049607
Enthalpy	h	-4416.0	kJ/kg	-4345.0
Flow Rate	Q	0.3677	m^3/s	
Compressibilty Factor	Z	0.7344	-	
Molecular Weight	MW	19.63	kg/kmol	
Gas Constant	R	8314	J/kmol.K	
Correction Factor	f	1.0		
Polytropic exponent	n	1.949706046		

PERFORMANCE PARAMETERS

Pressure Ratio	∏	1.67	-
Polytropic Head	H_p	54,689.50	J/kg
Polytropic Efficiency	η_p	77.0%	
Total Head	H	71,000.00	J/kg
Calculated Work	W	3,320.77	kW

COMPRESSOR TYPE

Frame		10	
Frame Configuration		Vertical Split	MB
Pressure Limit		**345**	**Barg**
Frame Selection - Confirmation (based on Pressure limit)		YES	*if no change frame*

Split Stream (based on Compressor pressure limits)		**1**	
Adjusted Flow rate		**0.3677**	m^3/s

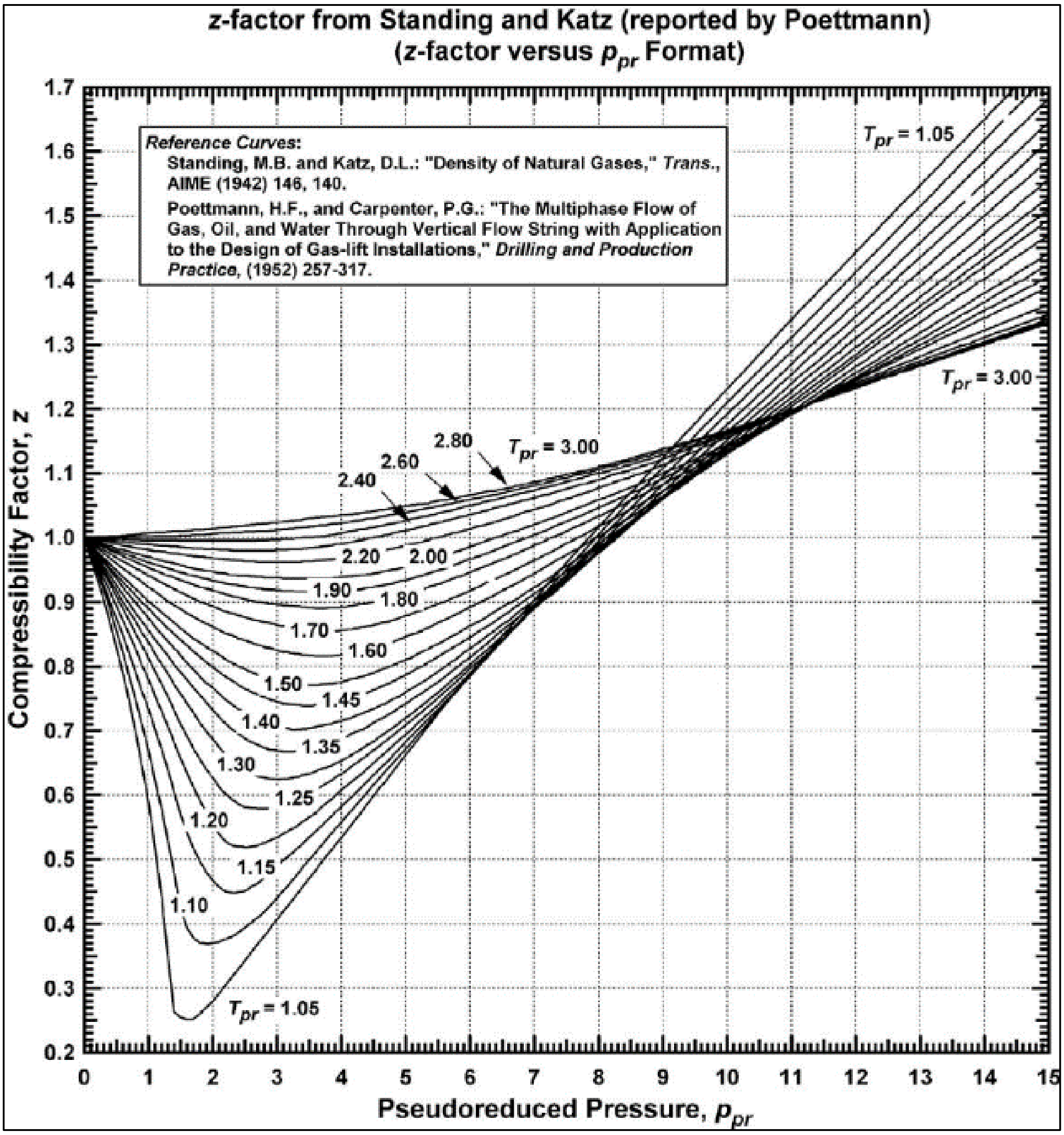
OUTPUT PARAMETERS			
Compressor Width	**W**	**1.092**	**metres**
Casing Height	**H**	**1.080**	**metres**
Average Footprint		**1.346**	$\mathbf{m^2}$
Weight of Vessel	$\mathbf{W_V}$	**4.538**	**tons**
Weight of Total Skid Weight	$\mathbf{W_{skid}}$	**4.538**	**tons**

NB. Weights do not take hydrostatic or fluid weights into consideration.

COMPRESSOR WEIGHT CALCULATIONS

Casing Width	-	1092.0	mm
Casing Height	-	1080.0	mm
Average Casing Weight	-	4537.5	kg
Average Footprint		**1.34589**	m^2
Average Casing Weight		**4537.5**	kg

Appendix F Standing-Katz chart (compressibility factor Z)



Appendix G Piping data

Appendix G.1 Piping calculator

Pipeline Sizing

Colour Coding	
(yellow)	Input Parameter
(blue)	Empirical/Determined values
(orange)	
(green)	Output

PIPELINE SIZING - Calculation Sheet

INPUT PARAMETERS			
Phase Flow		Liquid	

Wall thickness

Wall thickness (Generalised formula)

Hoop Stress in pipe wall	Hs	12000	psi
Length of Pipe	L		ft
Internal Pressure of the pipe	P	725	psi
Outside diameter of pipe	do	6.025	in
Pipe wall thickness	t	0.17163556	in

$$t = \frac{Pd_o}{2(H_s + P)}$$

ASME/ANSI Code B31.3

Nominal Pipe Size		2.5 - 20	in
Longitudinal Weld-joint type		Electric Resistance Weld (ERW)	
Pipe Grade		ASTM A206 &API 5L, Grade B	
Temperature limit		500	° F
Corrosion allowance	te		in
Thread or groove depth	tth	0.11	in
Allowable internal pressure	P	500	psi
Outside diameter of pipe	do	6.025	in
Allowable stress for pipe	S	18900	psi
Longitudinal Weld-joint factor	E	0.85	
Derating Factor*	Y	0.4	
Manufacturers allowable tolerance**	Tol	10.0%	
Minimum design wall thickness	**t**	**0.2027**	**in**

$$t = t_e + t_{th} + \left[\frac{Pd_o}{2(SE + PY)}\right]\left[\frac{100}{100 - T_{ol}}\right]$$

*0.4 for ferrous materials operaing below 900 °F

**12.5 pipe up to 20 in.-OD, 10 pipe > 20 in. OD, API 5L

ASME/ANSI Code B31.4

Internal Pressure of the pipe	P	725	psi
Outside diameter of pipe	do	5	in
Specification		API 5L, ASTM A 53, ASTM A 106	
Grade		B	
Weld Joint Type		Seamless	
Minimum Yield stress for pipe	SY	35000	psi
Derating Factor*	F	0.72	
Longitudinal Weld-joint type		Electric Fusion (Arc) Weld	
Longitudinal Weld-joint factor	E	0.80	
Minimum design wall thickness	**t**	**0.0899**	**in**

$$t = \frac{Pd_o}{2(FES_Y)}$$

ASME/ANSI Code B31.8

Internal Pressure of the pipe	P	725	psi
Outside diameter of pipe	do	5	in
Specification		API 5L	
Grade		A25	
Minimum Yield stress for pipe	SY	25000	psi
Derating Factor*		Unimproved public roads (without Casing)	
Location Class		1-Div 2	
Design Factor	F	0.6	
Specification		ASTM A 134	
Pipe Class		Electric Fusion Arc Welded	
Longitudinal Weld-joint type	E	0.80	
Temperature range		-20 to 250	
Temperture derating Factor	T	1	
Minimum design wall thickness	**t**	**0.1510**	in

$$t = \frac{Pd_o}{2FETS_Y}$$

OUTPUT PARAMETERS			
Wall Thickness (ASME Code)		**Wall thickness (Generalised formula)**	
		0.172	**in**
Line Type		**Multiphase Line**	
Pipe Internal Diameter		3.338	**in**
Velocity		20.705	**ft/sec**

Velocity Considerations (Governed by API RP 14)

Liquid Line Sizing

Pipe ID	d	16	in
Fluid Flow rate	QL	18870	B/D
Liquid Velocity*	V	14.1525	ft/sec

**where solids might be present or where water could settle out and create corrosion zones in low spots, a minimum velocity of 3 ft/sec is normally used. A maximum velocity of 15 ft/sec is often used to minimize the possibility of erosion by solids and water hammer caused by quickly closing a valve.*

Gas Line Sizing

Pipe ID	d	10	in
Gas Flow rate	Qg	176.5733	MMscf/D
Gas flowing temperature	T	552.87	R
Flowing Pressure	P	725	psi
Compressibility factor	Z	0.85	
Gas Velocity **	Vg	68.67208412	ft/sec

***velocity in gas lines should be less than 60 to 80 ft/sec to minimize noise and allow for corrosion inhibition. A lower velocity of 50 ft/sec should be used in the presence of known corrosives such as CO2. The minimum gas velocity should be between 10 and 15 ft/sec, which minimizes liquid fallout.*

Multiphase Line Sizing

Pressure	P	725	psi
Gas Constant	R	8.314	
Specific Gravity of the Liquid (relative to water)	SG	0.862	
Specific Gravity of the gas relative to air	S	0.67	
Temperature	T	552.87	R
Compressibility factor	Z	0.85	
Average density of the mxture	**rhom**	**52.48491718**	**kg/m3**
Empirical Constant	C	Solids-free, No corrosion or CRA material (cont. service) min	
Empirical Constant	C	150	
Erosional Velocity***	Ve	20.70494117	ft/ssec
Liquid-Flow rate	QL	18870	B/D
Pipe ID	d	3.337586317	in

****Recommended minimum velocity is 10 to 15 ft/sec. The maximum recommended velocity is 60 ft/sec to inhibit noise and 50 ft/sec for CO2 corrosion inhibition.*

Appendix G.2 Liquid and gas pipeline optimum velocity

PIPELINE OPTIMUM FLUID VELOCITY

60

PIPELINE	PHASE	Mass Density	Volume Flowrate	NOMINAL PIPE SIZE (in)	SCHEDULE	Wall thickness (m)	Inside Area (m^2)	Actual Velocity	Theoretical Max Velocity	Feasible	Weight (kg/m)	Assumed Length	Gross Weight
		kg/m^3	m^3/s	NPS			m^2	m/s	m/s		kg/m	m	
43	Wet Gas	106.55	1.0455	10	80	0.015062	0.046325	22.57	23.51	YES	95.73	7	670.13
15	Condensate	564.90	0.0415	2.5	5S	0.002108	0.003717	11.17	11.47	YES	3.68	7	25.73
14	Wet Gas	87.59	1.0040	10	120	0.021412	0.041608	24.13	25.57	YES	132.74	7	929.21
24	Condensate	564.90	0.0415	2.5	5S	0.002108	0.003717	11.17	11.47	YES	3.68	7	25.73
16	Condensate	297.32	0.0789	3	10s	0.003048	0.005382	14.66	15.12	YES	6.44	7	45.11
42	liquid	74.35	0.0000	8	XXS	0.022225	0.023938	0.00	27.44	YES	107.77	7	754.41
39	liquid	81.43	0.0000	8	XXS	0.022225	0.023938	0.00	26.39	YES	107.77	7	754.41
44	Condensate	297.32	0.0789	3	10s	0.003048	0.005382	14.66	15.12	YES	6.44	7	45.11
5	Wet Gas	80.70	0.0407	2	80 XS 80S	0.005537	0.001904	21.38	26.49	YES	7.47	7	52.29
6	Water	976.46	0.0004	1.5	XXS	0.010160	0.000613	0.63	8.00	YES	9.54	7	66.77
4	Condensate	523.76	0.0378	1.5	160	0.007137	0.000907	41.68	50.00	YES	7.23	7	50.63
13	Condensate	42.10	0.4701	12	100	0.021412	0.061996	7.58	8.00	YES	159.53	7	1116.72
28	Condensate	588.95	0.0014	1.5	XXS	0.010160	0.000613	2.24	8.00	YES	9.54	7	66.77
45	Condensate	43.54	0.4732	12	100	0.021412	0.061996	7.63	8.00	YES	159.53	7	1116.72
7	Wet Gas	11.02	0.4491	4	80 XS 80S	0.008560	0.007414	60.57	62.37	YES	22.29	7	156.05
56	Water	995.35	0.0000	2	XXS	0.011074	0.001144	0.01	8.00	YES	13.44	7	94.07
8	Condensate	648.78	0.0241	2.5	5S	0.002108	0.003717	6.49	8.00	YES	3.68	7	25.73
12	Condensate	14.84	1.0548	18	30	0.011100	0.148542	7.10	8.00	YES	121.98	7	853.89
10	Wet Gas	2.01	1.0351	4	40 Std 40S	0.006020	0.008209	126.10	129.65	YES	16.06	7	112.40
57	Water	757.59	0.0000	10	5S	0.003404	0.055645	0.00	8.00	YES	22.61	7	158.24
Stable Condensate	Condensate	690.52	0.0197	2	80 XS 80S	0.005537	0.001904	10.32	10.52	YES	7.47	7	52.29
17	Gas	14.07	0.1478	3	XXS	0.015240	0.002679	55.17	56.14	YES	27.65	7	193.55
18	Gas	43.80	0.0475	2	160	0.008712	0.001445	32.86	34.45	YES	11.07	7	77.50
58	Sea water	1022.25	0.0123	1.5	5S	0.001651	0.001587	7.75	8.00	YES	1.89	7	13.23
48	Sea water	1022.59	0.0123	1.5	5S	0.001651	0.001587	7.74	8.00	YES	1.89	7	13.23
49	Sea water	1011.50	0.0124	1.5	5S	0.001651	0.001587	7.83	8.00	YES	1.89	7	13.23
46	Wet gas	14.22	0.4940	4	10s	0.003048	0.009191	53.75	55.88	YES	8.35	7	58.44

PIPELINE	PHASE	Mass Density	Volume Flowrate	NOMINAL PIPE SIZE (in)	SCHEDULE	Wall thickness (m)	Inside Area (m^2)	Actual Velocity	Theoretical Max Velocity	Feasible	Weight (kg/m)	Assumed Length	Gross Weight
		kg/m^3	m^3/s	NPS			m^2	m/s	m/s		kg/m	m	
3	Wet Gas	12.62	0.4926	4	10s	0.003048	0.009191	53.60	58.83	YES	8.35	7	58.44
19	Condensate	588.94	0.0014	1.5	XXS	0.010160	0.000613	2.24	8.00	YES	9.54	7	66.77
CompStream2	Wet Gas	81.15	0.0766	3	160	0.011100	0.003492	21.94	26.43	YES	21.28	7	148.96
20	Wet Gas	276.92	0.0224	1.5	5S	0.001651	0.001587	14.15	15.59	YES	1.89	7	13.23
59	Sea water	1022.25	0.0442	3	5S	0.002108	0.005629	7.86	8.00	YES	4.51	7	31.56
50	Sea water	1022.59	0.0442	3	5S	0.002108	0.005629	7.85	8.00	YES	4.51	7	31.56
51	Sea water	1011.50	0.0447	3	5S	0.002108	0.005629	7.94	8.00	YES	4.51	7	31.56
47	Wet Gas	159.01	0.0598	2.5	10s	0.003048	0.003516	16.99	19.79	YES	5.25	7	36.77
26	Liquids	344.90	0.0120	1.5	5S	0.001651	0.001587	7.53	8.00	YES	1.89	7	13.23
11	Wet Gas	112.54	0.0478	2	10s	0.002769	0.002356	20.29	22.96	YES	3.93	7	27.50
22	Wet Gas	88.55	1.0539	10	120	0.021412	0.041608	25.33	25.45	YES	132.74	7	929.21
1	Wet Gas	88.55	1.0539	10	120	0.021412	0.041608	25.33	25.45	YES	132.74	7	929.21
23	Wet Gas	74.40	1.2542	10	80	0.015062	0.046325	27.07	27.43	YES	95.73	7	670.13
29	Dry gas	74.35	1.2550	10	80	0.015062	0.046325	27.09	27.44	YES	95.73	7	670.13
32	Water	1009.52	0.0000	8	XXS	0.022225	0.023938	0.00	8.00	YES	107.77	7	754.41
41	liquid	74.35	0.0000	8	XXS	0.022225	0.023938	0.00	8.00	YES	107.77	7	754.41
40	Dry gas	74.35	1.2550	10	80	0.015062	0.046325	27.09	27.44	YES	95.73	7	670.13
33	Dry gas	97.57	0.9563	10	120	0.021412	0.041608	22.98	24.41	YES	132.74	7	929.21
34	Dry gas	120.32	0.7755	10	160	0.028575	0.036591	21.19	22.31	YES	172.09	7	1204.64
60	Sea water	1022.22	0.1401	6	40 Std 40S	0.007112	0.018629	7.52	8.00	YES	28.23	7	197.61
52	Sea water	1022.39	0.1401	6	40 Std 40S	0.007112	0.018629	7.52	8.00	YES	28.23	7	197.61
53	Sea water	1011.29	0.1416	6	40 Std 40S	0.007112	0.018629	7.60	8.00	YES	28.23	7	197.61
9	Dry gas	120.32	0.7755	8	10s	0.003759	0.035134	22.07	22.31	YES	19.94	7	139.59
37	liquid	120.32	0.0000	8	XXS	0.022225	0.023938	0.00	8.00	YES	107.77	7	754.41
38	liquid	81.37	0.0000	8	XXS	0.022225	0.023938	0.00	8.00	YES	107.77	7	754.41
35	Dry gas	155.13	0.6015	8	40 Std 40S	0.008179	0.032259	18.65	20.00	YES	42.49	7	297.41
36	Dry gas	174.24	0.5355	8	80 XS 80S	0.012700	0.029445	18.19	19.03	YES	64.57	7	452.00
61	Sea water	1022.22	0.0950	6	XXS	0.021946	0.012145	7.82	8.00	YES	79.11	7	553.77
54	Sea water	1022.39	0.0949	6	XXS	0.021946	0.012145	7.82	8.00	YES	79.11	7	553.77
55	Sea water	1011.29	0.0960	6	XXS	0.021946	0.012145	7.90	8.00	YES	79.11	7	553.77
											2880.66	TOTAL	20164.63

Appendix G.3 Thread allowance calculations for threaded pipe-wall thickness, t_{th}

Nominal Pipe Size	t_{th}, in
0.25 - 0.375	0.05
0.5 - 0.375	0.06
1-2	0.08
2.5 - 20	0.11

Appendix G.4 Basic allowable stress for grade B and X-42 seamless pipe

Temperature, ° F	ASTM A206 &API 5L, Grade B	API 5L, Grade X42
-20 to 400	20000	20000
500	18900	N/A
600	17300	N/A
650	17000	N/A

Appendix G.5 Basic allowable stress for other grades of seamless pipe

Grade	Minimum Temperature	Allowable Stress Minimum Temperature to 100° F
API 5L-A	-20	16,000
API 5LX-42	-20	20,000
API 5LX-46	-20	21,000
API 5LX-52	-20	22,000
ASTM A-106-B	-20	20,000
ASTM A-333-6	-50	20,000
ASTM A-369-FPA	-20	16,000
ASTM A-369-FPB	-20	20,000
ASTM A-524-I	-20	20,000
ASTM A-524-II	-20	18,300

Appendix G.6 Longitudinal weld joint factor

Seamless	1.00
Electric Fusion Weld	0.95
Double Butt	0.95
Straight Seam	0.95
Spiral Seam APL 5L	0.95
Electric Resistance Weld (ERW)	0.85
Furnace Butt Weld	0.60

Appendix G.7 Minimum yield stress for pipe (courtesy ANSI/ASME)

Specification	Grade	Seamless	Furnace Butt Weld, Continuous Welded	Electric Resistance Weld (ERW) and Electric Flash welded	Electric Fusion Welded	Submerged Arc Weld
API 5L	A25	25,000	25,000	-	-	
API 5L, ASTM A 53, ASTM A 106	A	30,000	30,000	-	-	30,000
API 5L, ASTM A 53, ASTM A 106	B	35,000	35,000	-	-	35,000
API 5LU	U80	80,000	80,000	-	-	80,000
API 5LU	U100	100,000	100,000	-	-	100,000
API 5L	X42	42,000	42,000	-	-	42,000
API 5L	X46	46,000	46,000	-	-	46,000
API 5L	X52	52,000	52,000	-	-	52,000
API 5L	X56	56,000	56,000	-	-	56,000
API 5L	X60	60,000	60,000	-	-	60,000
API 5L	X65	65,000	65,000	-	-	65,000
API 5L	X70	70,000	70,000	-	-	70,000
ASTM A 106	C	40,000	-	-	-	-
ASTM A 524	I	35,000	-	-	-	-
ASTM A 524	H	30,000	-	-	-	-
API 5L, ASTM A 53, ASTM A 135	A	-	-	30,000	-	-
API 5L, ASTM A 53, ASTM A 135	B	-	-	35,000	-	-
ASTM A 134	-	-	-	-		-
ASTM A 139	A	-	-	-	30,000	-
ASTM A 139	B	-	-	-	35,000	-
ASTM A 671	-	-	-	-	-	-
ASTM A 671	-	-	-	-	-	-
ASTM A 672	-	-	-	-	-	-
ASTM A 672	-	-	-	-	-	-
ASTM A 381	Y35	-	-	-	-	35,000

Specification	Grade	Seamless	Furnace Butt Weld, Continuous Welded	Electric Resistance Weld (ERW) and Electric Flash welded	Electric Fusion Welded	Submerged Arc Weld
ASTM A 381	Y42	-	-	-	-	42,000
ASTM A 381	Y46	-	-	-	-	46,000
ASTM A 381	Y48	-	-	-	-	48,000
ASTM A 381	Y50	-	-	-	-	50,000
ASTM A 381	Y52	-	-	-	-	52,000
ASTM A 381	Y60	-	-	-	-	60,000
ASTM A 381	Y65	-	-	-	-	65,000

Appendix G.8 Specified minimum yield strength for steel pipe commonly used in pipe systems (courtesy ANSI/ASME – code B31.8)

Specification Number	Grade	Type	SMYS, psi
API 5L	A25	BW, ERW, S	25,000
API 5L	A	ERW, S, DSA	30,000
API 5L	B	ERW, S, DSA	35,000
API 5L	X42	ERW, S, DSA	42,000
API 5L	X46	ERW, S, DSA	46,000
API 5L	X52	ERW, S, DSA	52,000
API 5L	X56	ERW, S, DSA	56,000
API 5L	X60	ERW, S, DSA	60,000
API 5L	X65	ERW, S, DSA	65,000
API 5L	X70	ERW, S, DSA	70,000
API 5L	X80	ERW, S, DSA	80,000
ASTM A 53	Type F	BW	25,000
ASTM A 53	A	ERW, S	30,000
ASTM A 53	B	ERW, S	35,000
ASTM A 106	A	S	30,000
ASTM A 106	B	S	35,000
ASTM A 106	C	S	40,000
ASTM A 134	-	EFW	-
ASTM A 135	A	ERW	30,000
ASTM A 135	B	ERW	35,000
ASTM A 139	A	EFW	30,000
ASTM A 139	B	EFW	35,000
ASTM A 139	C	EFW	42,000
ASTM A 139	D	EFW	46,000
ASTM A 139	E	EFW	52,000
ASTM A 333	1	S, ERW	30,000

Specification Number	Grade	Type	SMYS, psi
ASTM A 333	3	S, ERW	35,000
ASTM A 333	4	S	35,000
ASTM A 333	6	S, ERW	35,000
ASTM A 333	7	S, ERW	35,000
ASTM A 333	8	S, ERW	75,000
ASTM A 333	9	S, ERW	46,000
ASTM A 381	Class Y-35	DSA	35,000
ASTM A 381	Class Y-42	DSA	42,000
ASTM A 381	Class Y-46	DSA	46,000
ASTM A 381	Class Y-48	DSA	48,000
ASTM A 381	Class Y-50	DSA	50,000
ASTM A 381	Class Y-52	DSA	52,000
ASTM A 381	Class Y-56	DSA	56,000
ASTM A 381	Class Y-60	DSA	60,000
ASTM A 381	Class Y-65	DSA	65,000

Appendix G.9 Basic design factor (F) for steel pipe construction in natural gas service (courtesy ANSI/ASME – code B31.8)

Facility	Location Class				
	1				
	1-Div 1	1-Div 2	2	3	4
Pipelines, mains and service lines	0.80	0.72	0.60	0.50	0.40
Private Roads (without Casing)	0.80	0.72	0.60	0.50	0.40
Unimproved public roads (without Casing)	0.60	0.60	0.60	0.50	0.40
Roads, Highways, Public streets with hard surface and railroads (without Casing)	0.60	0.60	0.50	0.50	0.40
Private Roads (with Casing)	0.80	0.72	0.60	0.50	0.40
Unimproved public roads (with Casing)	0.72	0.72	0.60	0.50	0.40

Facility	Location Class				
	1		2	3	4
	1-Div 1	1-Div 2			
Roads, Highways, Public streets with hard surface and railroads (with Casing)	0.72	0.72	0.60	0.50	0.40
Parallel encroachment - Private Roads	0.80	0.72	0.60	0.50	0.40
Unimproved public roads - Parallel encroachment	0.80	0.72	0.60	0.50	0.40
Roads, Highways, Public streets with hard surface and railroads - Parallel encroachment	0.60	0.60	0.60	0.50	0.40
Fabricated assemblies	0.60	0.60	0.60	0.50	0.40
Pipelines on bridges	0.60	0.60	0.60	0.50	0.40
Compression Station piping	0.50	0.50	0.50	0.50	0.40
Near concentration of people in Location Classes 1 and 2	0.50	0.50	0.50	0.50	0.40

Appendix G.10 Basic design longitudinal joint factor for steel pipelines in natural gas service (courtesy ANSI/ASME – code B31.8)

Specification Number	Pipe Class	E factor
ASTM A 53	Seamless	1.00
ASTM A 53	ERW	1.00
ASTM A 53	Furnace Butt Welded	0.60
ASTM A 106	Seamless	1.00
ASTM A 134	Electric Fusion Arc Welded	0.80
ASTM A 135	Electric Resistance Welded	1.00
ASTM A 139	Electric Fusion Welded	0.80
ASTM A 211	Spiral Welded Steel Pipe	0.80
ASTM A 333	Seamless	0.80
ASTM A 381	Double Submerged Arc-Welded	1.00
ASTM A 671	EFW - Class 13, 23, 33, 43, 53	0.80
ASTM A 671	EFW - Class 12, 22, 32, 42, 52	1.00
ASTM A 672	EFW - Class 13, 23, 33, 43, 53	0.80
ASTM A 672	EFW - Class 12, 22, 32, 42, 52	1.00
API 5L	Seamless	1.00
API 5L	Electric Resistance Welded	1.00
API 5L	Electric Flash Welded	1.00
API 5L	Submerged Arc-Welded	1.00
API 5L	Furnace Butt Welded	0.60

Appendix G.11 Basic design temperature derating factor for (T) for steel pipelines in natural gas service (courtesy ANSI/ASME – Code B 31.8)

Temperature, ° F	T
-20 to 250	1.000
300	0.967
350	0.933
400	0.900
450	0.867

Appendix H Maximum allowable stress (ASME Division 1 and 2)

Material	Spec No.	Grade	Div1 (-20deg F to -650 deg F)	Div2 (-20degF to 650degF)
Carbon Steel Plates and Sheets	SA-516	55	13,800	18,300
	SA-516	60	15,000	20,000
	SA-516	65	16,300	21,700
	SA-516	70	17,500	23,300
	SA-285	A	11,300	15,000
	SA-285	B	12,500	16,700
	SA-285	C	13,800	18,300
	SA-36	-	12,700	16,900
	SA-203	A	16,300	21,700
	SA-203	B	17,500	23,300
	SA-203	D	16,300	21,700
	SA-203	E	17,500	23,300
High Alloy Steel Plates	SA-240	304	1,200	20,000
	SA-240	304L	-	16,700
	SA-240	316	12,300	20,000
	SA-240	316L	10,200	16,700

Appendix I Pumps

Appendix I.1 Piping equivalent length of valves and fittings (feet)

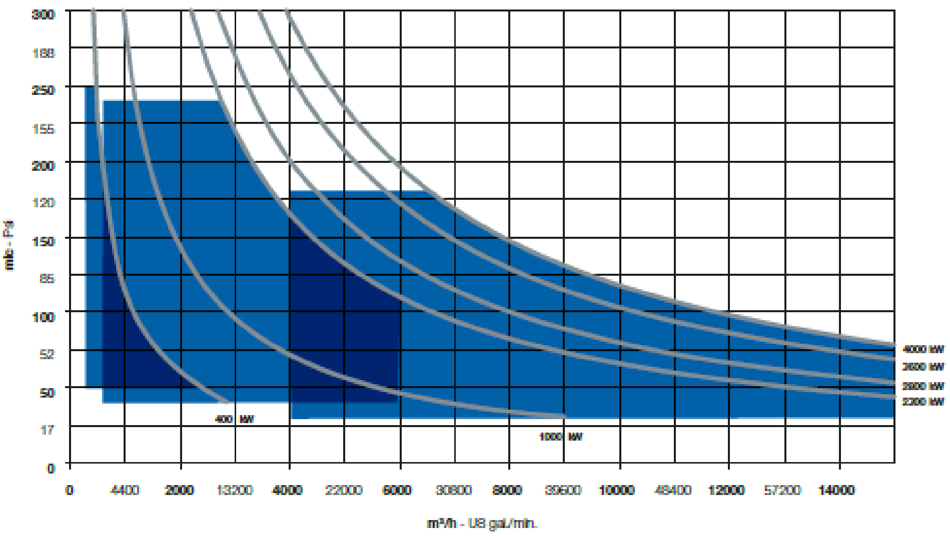
Nominal Pipe size in.	Globe valve or ball check valve	Angle valve	Swing check valve	Plug cock	Gate or ball valve	45° ell		Short rad. ell		Long rad. ell		Hard T		Soft T		90° miter bends			Enlargement					Contraction				
						Welded	Threaded	Welded	Threaded	Welded	Threaded	Welded	Threaded	Welded	Threaded	2 miter	3 miter	4 miter	Sudden			Std. red.		Sudden			Std. red.	
																			Equiv. L in terms of small d									
																			d/D = 1/4	d/D = 1/2	d/D = 3/4	d/D = 1/2	d/D = 3/4	d/D = 1/4	d/D = 1/2	d/D = 3/4	d/D = 1/2	d/D = 3/4
1½	55	26	13	7	1	1	2	3	5	2	3	8	9	2	3				5	3	1	4	1	3	2	1	1	–
2	70	33	17	14	2	2	3	4	5	3	4	10	11	3	4				7	4	1	5	1	3	3	1	1	–
2½	80	40	20	11	2	2	–	5	–	3	–	12	–	3	–				8	5	2	6	2	4	3	2	2	–
3	100	50	25	17	2	2		6		4		14		4					10	6	2	8	2	5	4	2	2	–
4	130	65	32	30	3	3		7		5		19		5					12	8	3	10	3	6	5	3	3	–
6	200	100	48	70	4	4		11		8		28		8					18	12	4	14	4	9	7	4	4	1
8	260	125	64	120	6	6		15		9		37		9					25	16	5	19	5	12	9	5	5	2
10	330	160	80	170	7	7		18		12		47		12					31	20	7	24	7	15	12	6	6	2
12	400	190	95	170	9	9		22		14		55		14		28	21	20	37	24	8	28	8	18	14	7	7	2
14	450	210	105	80	10	10		26		16		62		16		32	24	22	42	26	9	–	–	20	16	8	–	–
16	500	240	120	145	11	11		29		18		72		18		38	27	24	47	30	10	–	–	24	18	9	–	–
18	550	280	140	160	12	12		33		20		82		20		42	30	28	53	35	11	–	–	26	20	10	–	–
20	650	300	155	210	14	14		36		23		90		23		46	33	32	60	38	13	–	–	30	23	11	–	–
22	688	335	170	225	15	15		40		25		100		25		52	36	34	65	42	14	–	–	32	25	12	–	–
24	750	370	185	254	16	16		44		27		110		27		56	39	36	70	46	15	–	–	35	27	13	–	–
30	–	–	–	312	21	21		55		40		140		40		70	51	44										
36	–	–	–		25	25		66		47		170		47		84	60	52										
42	–	–	–		30	30		77		55		200		55		98	69	64										
48	–	–	–		35	35		88		65		220		65		112	81	72										
54	–	–	–		40	40		99		70		250		70		126	90	80										
60	–	–	–		45	45		110		80		260		80		190	99	92										

Appendix I.2 Frictional loss in pipes (schedule 40)

SCHEDULE 40 (STEEL PIPE) -- inches							
	Pipe Size in Inches *Friction Loss per 100ft*						
GPM	**2**	**3**	**4**	**6**	**8**	**10**	**12**
10	0.25	0.04					
12	0.34	0.05					
15	0.52	0.08					
20	0.87	0.13					
25	1.30	0.19					
30	1.82	0.26					
35	2.42	0.35					
40	3.10	0.44					
45	3.85	0.55					
70	8.86	1.22	0.35				
100	17.40	2.39	0.63				
150	38.00	5.14	1.32				
200	66.30	8.90	2.27	0.3	0.08		
250	90.70	14.10	3.60	0.49	0.13		
300		19.20	4.89	0.64	0.16	0.05	
350		26.90	6.72	0.88	0.23	0.07	
400		33.90	8.47	1.09	0.28	0.09	
450		42.75	10.65	1.36	0.35	0.11	
500		52.50	13.00	1.66	0.42	0.14	
550		63.20	15.70	1.99	0.51	0.16	
600		74.80	18.60	2.34	0.6	0.19	
650		87.50	21.70	2.73	0.69	0.22	
700		101.00	25.00	3.13	0.8	0.26	
750		116.00	28.60	3.57	0.91	0.29	
800		131.00	32.40	4.03	1.02	0.33	0.14
850		148.00	36.50	4.53	1.13	0.37	0.15
900		165.00	40.80	5.05	1.27	0.41	0.17
950		184.00	45.30	5.6	1.41	0.46	0.19
1000		204.00	50.20	6.17	1.56	0.50	0.21
1100				7.41	1.87	0.60	0.25
1200				8.76	2.2	0.70	0.30
1300				10.2	2.56	0.82	0.34
1400				11.8	2.95	0.94	0.40
1500				13.5	3.37	1.07	0.45
1600					3.82	1.21	0.51
1800					4.79	1.52	0.64
2000						1.86	0.78
2200						2.25	0.94

Appendix I.3 Framo submersible pump

Performance domain



Submerged lift pump

	Small		Medium			Large			
PUMP TYPE	SE200	SE225	SE280	SE315	SE355	SE400	SE450	SE500	SE560
Required caisson diameter	18"	26"	30"	34"	40"	46"	52"	58"	62"
Flow range [m³/h] (BEP)	200-500	300-1000	600-2400	700-3200	1400-6200	2800-8500	3000-10000	4000-12000	4000-15000
Pipestack diameter min/max	6"	10"/14"	10"/18"	14"/20"	18"/28"	24"/32"	24"/44"	24"/44"	24"/44"
Max power (50/60Hz) [kW]	175/220	400/400	800/1000	1000/1200	2100/2500	2200/2800	2900/3600	3300/4000	3800/4000
Max power (50/60Hz) [kW] 11kV	NA	NA	NA	NA	1400/1750	1800/2150	2200/2700	2600/3150	3800/4500
Voltage min/max [kV]	0.40/0.69	0.40/0.69	0.40/4.16	0.40/6.6	0.40/11	3.3/11	3.3/11	3.3/11	3.3/11
Weight pump/ motor unit max [kg]	900	1500	2700	5000	6800	8500	10200	12000	13500
Weight per 6m pipestack min/max dia [kg]	200	394/500	394/591	500/720	591/915	770/1150	770/1300	770/1300	770/1300
Weight top-bend and top-plate min dia/max dia [kg]	*	380/415	430/500	535/651	670/1050	1200/1250	1380/1500	1600/1700	1750/1850

Appendix J HYSYS simulation properties table

Stream Name	Pressure [bar]	Temperature [C]	Mass Flow [kg/s]	Std Ideal Liq Vol Flow [m3/s]	Molar Enthalpy [kJ/kgmole]
GasWell	**180**	**80,0**	53,44	0,1480	- 87 651,18
Water	**90**	**5,0**	0,20	0,0002	- 287 784,79
Stream 5	77	69,7	1,67	0,0044	- 92 390,93
Stream 6	77	69,7	0,19	0,0002	- 282 681,19
Stream 7	9	42,1	2,48	0,0058	- 99 946,80
Stable Condensate	1	20,0	6,77	0,0099	- 232 114,77
Stream 10	1	20,0	1,04	0,0020	- 118 377,96
Stream 12	**1**	20,0	7,81	0,0118	- 203 471,99
Stream 8	9	42,1	7,81	0,0118	- 203 471,99
Stream 4	77	69,7	9,88	0,0169	- 153 381,33
Sat gas	180	80,0	53,64	0,1482	- 88 351,56
Stream 15	77	0,7	11,75	0,0215	- 151 461,65
Stream 16	77	**69,7**	11,75	0,0215	- 141 596,20
Stream 17	9	112,4	1,04	0,0020	- 110 844,63
Stream 18	9	**25,0**	1,04	0,0020	- 130 709,51
CompStream2	77	163,5	3,11	0,0071	- 94 157,93
Stream 3	9	24,4	3,11	0,0071	- 101 540,45
Stream 20	77	**30,0**	3,11	0,0071	- 110 176,36
Stream 11	77	39,0	2,70	0,0070	- 95 662,32
Stream 14	77	0,7	43,97	0,1312	- 86 743,99
Stream 22	77	2,6	46,68	0,1382	- 87 139,59
Stream 23	77	**25,0**	46,67	0,1382	- 85 818,07
Stream 1	77	2,6	46,67	0,1382	- 87 133,40
Stream 2	77	2,6	0,00	0,0000	- 288 077,82
Stream 19	9	24,4	0,41	0,0007	- 171 507,94
Stream 21	77	2,6	0,00	0,0000	- 288 079,43
Stream 24	77	0,7	11,75	0,0215	- 151 461,65
Stream 25	**77**	**39,0**	2,08	0,0045	- 115 740,19
Stream 26	77	39,0	2,08	0,0045	- 115 694,36
Stream 27	**77**	**2,6**	0,00	0,0000	- 288 079,43

Stream Name	Pressure [bar]	Temperature [C]	Mass Flow [kg/s]	Std Ideal Liq Vol Flow [m3/s]	Molar Enthalpy [kJ/kgmole]
Stream 28	9	24,4	0,41	0,0007	- 171 542,55
Stream 31	90	5,0	53,64	0,1482	- 92 358,73
Stream 30	77	- 0,0	53,64	0,1482	- 92 358,73
Stream 13	9	44,0	9,88	0,0169	- 153 381,33
Stream 29	77	25,1	46,67	0,1382	- 85 789,60
Stream 32	77	25,1	0,01	0,0000	- 286 203,70
Stream 33	120	63,3	46,67	0,1382	- 84 503,10
Stream 34	120	30,0	46,67	0,1382	- 86 507,99
Stream 35	200	71,4	46,67	0,1382	- 85 019,91
Stream 36	200	50,0	46,67	0,1382	- 86 378,12
Stream9	120	30,0	46,67	0,1382	- 86 507,99
Stream 37	120	30,0	-	0,0000	- 86 508,04
Stream 38	77	12,5	-	0,0000	- 86 508,04
Stream 39	77	12,5	-	0,0000	- 86 530,17
Stream 40	77	25,1	46,67	0,1382	- 85 789,60
Stream 41	77	25,1	-	0,0000	- 85 789,68
Stream 42	77	25,1	-	0,0000	- 85 789,68
Stream 43	77	0,7	55,72	0,1527	- 92 897,07
Stream 44	77	69,7	11,75	0,0215	- 141 596,20
Stream 45	9	42,1	10,30	0,0176	- 154 182,58
Stream 46	9	24,4	3,52	0,0078	- 106 621,69
Stream 47	77	39,0	4,78	0,0115	- 102 706,41
Stream 48	1	5,0	6,59	0,0066	- 287 956,80
Stream 49	1	20,0	6,59	0,0066	- 286 755,70
Stream 50	1	5,0	22,71	0,0228	- 287 956,80
Stream 51	1	20,0	22,71	0,0228	- 286 755,70
Stream 52	1	5,0	71,63	0,0718	- 287 956,80
Stream 53	1	20,0	71,63	0,0718	- 286 755,70
Stream 54	1	5,0	48,52	0,0486	- 287 956,80
Stream 55	1	20,0	48,52	0,0486	- 286 755,70

	Separator								
Property	1st st. 3-phase Separator	3rd Stage Separator	Inlet Separator	2nd st. LP Comp Scrubber	2nd Stage Separator	Intermediate Separator	Dehydration Scrubber	2nd st HP Comp Scrubber	1st st. HP Comp Scrubber
Vessel Temperature [C]	69,7	20,0	0,7	24,4	42,1	39,0	2,6	30,0	25,1
Vessel Pressure [bar]	77,0	1,0	77,0	8,8	8,8	77,0	77,0	120,0	77,0
Vessel Pressure Drop [bar]	-	-	-	-	-	-	-	-	-
Vapour Outlet Pressure Drop [bar]	-	-	-	-	-	-	-	-	-
Tank Volume [m3]	2 479,4	5,2	26,7	2,1	7,8	3,3	32,8		
Liquid Volume [m3]	1 239,7	2,6	13,3	1,1	3,9	1,7	16,4		
Liquid Volume Percent [%]	50,0	50,0	50,0	50,0	50,0	50,0	50,0	50,0	50,0
Vessel Diameter [m]	0,3	1,1	2,1	0,9	1,2	0,9	2,3		
Vessel Length or Height [m]	32 767,0	5,9	7,5	3,2	6,7	5,0	8,0		

Heat Exchanger				
Property	**1st st. Compressor Discharge Cooler**	**2nd st. Compressor Discharge Cooler**	**2nd St. HP Comp Cooler**	**Export Cooler**
Duty [kcal/h]	378061,11	1302966,14	4109086,66	2783637,28
UA [kJ/C-h]	33427,14	88501,71	562085,44	247861,02
Control UA [kJ/C-h]	33427,14	88501,71	562085,44	247861,02
Tube Side Pressure Drop [bar]	0,00	0,00	0,00	0,00
Shell Side Pressure Drop [bar]	0,00	0,00	0,00	0,00
Tube Side Delta T [C]	-87,38	-133,48	-33,26	-21,39
Shell Side Delta T [C]	15,00	15,00	15,00	15,00
Uncorrected LMTD [C]	47,32	67,85	33,30	48,12
LMTD [C]	47,32	61,60	30,59	46,99
Ft Factor	1,00	0,91	0,92	0,98
Tube Side Zones	1,00	1,00	1,00	1,00
Shell Side Zones	1,00	1,00	1,00	1,00
Tube Side Volume [m3]	0,10	0,10	0,10	0,10
Shell Side Volume [m3]	0,10	0,10	0,10	0,10
Heat Trans. Area [m2]		60,32	60,32	60,32
Minimum Flow Scale Factor	0,00	0,00	0,00	0,00
Overall U [kJ/h-m2-C]		1467,24	9318,61	4109,20

Compressor				
Property	**1st st. LP Compressor**	**2nd st. LP compressor**	**1st st. HP Compressor**	**2nd st. HP compressor**
Compressor Speed [rpm]				
Power [kW]	166,63	697,92	3 064,45	3 544,60
Capacity (act feed vol flow) [ACT_m3/s]	0,52	0,25	0,63	0,39
Adiabatic Efficiency	75,00	75,00	75,00	75,00
Polytropic Efficiency	77,55	78,65	76,18	76,21
Compressor Volume [m3]	-	-	-	-
Delta T [C]	92,38	139,05	38,15	41,39
Delta P [bar]	7,79	68,20	43,00	80,00
Polytropic Head [m]	12 670,15	17 996,48	5 101,13	5 902,46
Adiabatic Head [m]	12 253,34	17 161,10	5 021,98	5 808,84
Dynamic Head [m]	12 670,15	17 996,48	5 101,13	5 902,46
Polytropic Fluid Head [kJ/kg]	124,25	176,49	50,03	57,88
Adiabatic Fluid Head [kJ/kg]	120,16	168,29	49,25	56,97
Dynamic Fluid Head [kJ/kg]	124,25	176,49	50,03	57,88
Polytropic Head Factor	1,01	1,00	1,00	1,00
Polytropic Exponent	1,11	1,16	1,63	2,01
Isentropic Exponent	1,08	1,11	1,47	1,79
Fluid Power [kW]	166,63	697,92	3 064,45	3 544,60
Duty [kcal/h]	143 371,61	600 506,28	2 636 722,86	3 049 854,18

Appendix K Automation

Appendix K.1 Scenario table with input and output parameters in MS excel

	Scenario		Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7	Case 8
	Active		'							
Input	GasWell.Phase - Molar Flow.Overall.Overall	kgmole/h	8811.26	8811.26	8811.26	17622.51	17622.51	17622.51	17622.51	17622.51
	GasWell.Phase - Pressure.Overall.Overall	bar	180	180	180	180	180	180	180	180
	43.Phase - Pressure.Overall.Overall	bar	77	77	77	77	77	77	77	77
	43.Calculator Object.Mass Density.Correlation Properties.Elem1.Elem1	kg/m3	92.10569413	87.5893	87.58925499	87.589255	87.590446	87.59083933	112.6372281	112.6372281
	43.Calculator Object.Mass Density.Correlation Properties.Elem3.Elem3	kg/m3	1028.099624	1027.83	1027.832009	1027.832	1027.8337	1027.833331	999.702233	999.702233
	43.Calculator.Molecular Weight.Molecular Weight		22.09665502	22.1413	22.14133618	22.141336	22.140824	22.14051283	36.13424836	36.13424836
	43.Calculator.Std. Gas Flow.Std. Gas Flow	STD_m3/h	213383.0808	1284790	1284789.954	1284790	642390.54	385438.4257	8731.181645	8731.181645
	43.Calculator.Act. Liq. Flow.Act. Liq. Flow	m3/s	0.020006815	0.12458	0.12457732	0.1245773	0.0622755	0.037358878	0.010746629	0.010746629
	44.Phase - Pressure.Overall.Overall	bar	77	77	77	77	77	77	77	77
	44.Calculator Object.Mass Density.Correlation Properties.Elem1.Elem1	kg/m3	84.01327022	80.6944	80.69438375	80.694384	80.693059	80.69405605	364.2401322	364.2401322
	44.Calculator Object.Mass Density.Correlation Properties.Elem2.Elem2	kg/m3	529.4635129	523.787	523.7871642	523.78716	523.78557	523.722956	966.0143817	966.0143817
	44.Calculator Object.Mass Density.Correlation Properties.Elem3.Elem3		977.1334983	976.478	976.4783482	976.47835	976.45735	976.3754702		

	Scenario		Case 9	Case 10	Case 11	Case 12	Case 13	Case 14	Case 15
	Active								
Input	GasWell.Phase - Molar Flow.Overall.Overall	kgmole/h	17622.51	17622.51	17622.51	17622.51	10573.51	8811.26	5286.75
	GasWell.Phase - Pressure.Overall.Overall	bar	180	180	180	180	180	180	180
	43.Phase - Pressure.Overall.Overall	bar	77	77	77	77	77	77	77
	43.Calculator Object.Mass Density.Correlation Properties.Elem1.Elem1	kg/m3	112.6372281	112.6372281	112.6372281	112.6372281	112.6372281	112.6372281	112.6372281
	43.Calculator Object.Mass Density.Correlation Properties.Elem3.Elem3	kg/m3	999.702233	999.702233	999.702233	999.702233	999.702233	999.702233	999.702233
	43.Calculator.Molecular Weight.Molecular Weight		36.13424836	36.13424836	36.13424836	36.13424836	36.13424836	36.13424836	36.13424836
	43.Calculator.Std. Gas Flow.Std. Gas Flow	STD_m3/h	8731.181645	8731.181645	8731.181645	8731.181645	8731.181645	8731.181645	8731.181645
	43.Calculator.Act. Liq. Flow.Act. Liq. Flow	m3/s	0.010746629	0.010746629	0.010746629	0.010746629	0.010746629	0.010746629	0.010746629
	44.Phase - Pressure.Overall.Overall	bar	77	77	77	77	77	77	77
	44.Calculator Object.Mass Density.Correlation Properties.Elem1.Elem1	kg/m3	364.2401322	364.2401322	364.2401322	364.2401322	364.2401322	364.2401322	364.2401322
	44.Calculator Object.Mass Density.Correlation Properties.Elem2.Elem2	kg/m3	966.0143817	966.0143817	966.0143817	966.0143817	966.0143817	966.0143817	966.0143817
	44.Calculator Object.Mass Density.Correlation Properties.Elem3.Elem3								

Appendix K.2 Macro recording for scenario study sensitivity analysis

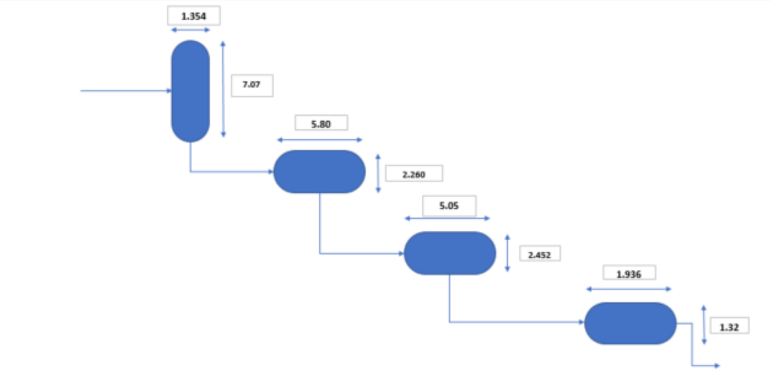
```vba
Sub MoveCase()
'
' MoveCase Macro
' MoveCase
'

'
    Range("D6:D27").Select
    Selection.Copy
    Sheets("OUTPUT - Cases Analysis").Select
    Range("A5").Select
    Selection.End(xlToRight).Select
    Selection.End(xlToRight).Select
    ActiveCell.Offset(0, 1).Range("A1").Select
    Selection.PasteSpecial Paste:=xlPasteValues, Operation:=xlNone, SkipBlanks _
        :=False, Transpose:=False
    ActiveWindow.SmallScroll Down:=18
    ActiveCell.Offset(22, 0).Range("A1").Select
    Sheets("OUTPUT - Single Case Graph Data").Select
    Range("E6:E27").Select
    Application.CutCopyMode = False
    Selection.Copy
    Sheets("OUTPUT - Cases Analysis").Select
    Selection.PasteSpecial Paste:=xlPasteValues, Operation:=xlNone, SkipBlanks _
        :=False, Transpose:=False
    ActiveWindow.SmallScroll Down:=18
    ActiveCell.Offset(22, 0).Range("A1").Select
    Sheets("OUTPUT - Single Case Graph Data").Select
    Range("F6:F27").Select
    Application.CutCopyMode = False
    Selection.Copy
    Sheets("OUTPUT - Cases Analysis").Select
    Selection.PasteSpecial Paste:=xlPasteValues, Operation:=xlNone, SkipBlanks _
        :=False, Transpose:=False
    ActiveWindow.SmallScroll Down:=15
    ActiveCell.Offset(22, 0).Range("A1").Select
    Sheets("OUTPUT - Single Case Graph Data").Select
    Range("G6:G27").Select
    Application.CutCopyMode = False
    Selection.Copy
    Sheets("OUTPUT - Cases Analysis").Select
    ActiveWindow.SmallScroll Down:=9
    Selection.PasteSpecial Paste:=xlPasteValues, Operation:=xlNone, SkipBlanks _
        :=False, Transpose:=False
    ActiveWindow.SmallScroll Down:=12
    ActiveCell.Offset(22, 0).Range("A1").Select
    Sheets("OUTPUT - Single Case Graph Data").Select
    Range("H6").Select
End Sub
```

Appendix K.3 VBA code for automatic interpolation of production potential

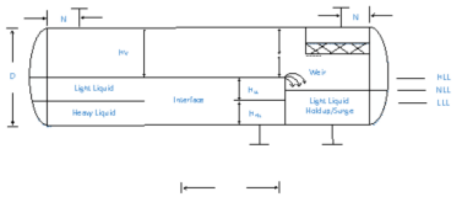
```vba
Function tabinterpol(x, col, Matrix As Range)
'function to perform linear interpolation in tables for properties VAR1, VAR2,
'INPUT:
'-x value for which the property BOP is required
'-col: column number in which the property is located
'Matrix: table organized in the following manner:

'   VAR1  VAR2  VAR3  VAR4
'x1
'x2
'x3
'.

'Reading the dimensions of the matrix
'Number of rows
m = Matrix.Rows.Count
'Number of columns
N = Matrix.Columns.Count

'Checking is value p is within the Matrix ranges
If x < Matrix(1, 1) Or x > Matrix(m, 1) Then
    Err = 1
End If
'If values are not in the matrix range return a message
If Err = 1 Then
    tabinterpol = "x NOT IN RANGE"
Else
    'Searching through the column
    For i = 1 To m - 1
                If x >= Matrix(i, 1) And x <= Matrix(i + 1, 1) Then
                    'Interpolating
                    tabinterpol = Matrix(i, col) + ((Matrix(i + 1, col) - Matrix(i, col)) * (x - Matrix(i, 1)) / (Matrix(i + 1, 1) - Matrix(i, 1)))
                Exit For
                Else
                End If
    Next
End If

End Function
```

Appendix K.4 Graphical layout of case analysis

CONDENSATE STABILISATION



3-PHASE HORIZONTAL SEPARATOR
1st Stage Horizontal Separator



Parameter	Units	Nomenclatu	1st Stage 3 Phase Separator	2nd Stage 3 Phase Separator	3rd Stage 2 Phase Separator
Diameter	m	D	2.26	2.45	1.33
Length	m	L	5.80	5.05	1.94
Light Liquid Level Height	m	H_{LL}	0.39	0.42	0.23
Heavy Liquid Level Height	m	H_{HL}	0.39	0.42	-
Weir Height	m	H_W	0.78	0.84	-
Vapour Space Height	m	H_V	1.48	1.61	0.61
Low Liquid Level Height	m	H_{LLL}	0.28	0.30	0.72
Footprint	**m^2**	**F**	**39.3**	**37.1**	**12.9**
Volume	**m^3**	**V**	**216.9**	**219.2**	**47.0**
Weight of Vessel	**kg**	**W_V**	**45,288**	**5,947**	**268**
Weight of Total Skid Weight	**kg**	**W_{skid}**	**72,507**	**10,348**	**968**

Appendix K.5 Case by case scenario study generation

CASE BY CASE SENSITIVITY ANALYSIS OF PLANT DESIGN

			Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7	Case 8	Case 9	Case 10	Case 11	Case 12	Case 13	Case 14	Case 15
FOOTPRINT (m^2)	SEPARATOR	Inlet Separator	14.505045542851	14.51	14.51	14.51	27.51	27.51	27.51	27.51	27.51	27.51	27.51	27.51	9.16	9.16	9.16
		1st Stage 3 Phase Separator	60.14	60.14	60.14	60.14	108.38	108.38	108.38	108.38	108.38	108.38	108.38	108.38	39.29	39.29	39.29
		2nd Stage 3 Phase Separator	58.78	58.78	58.78	58.78	110.54	110.54	110.54	110.54	110.54	110.54	110.54	110.54	37.12	37.12	37.12
		3rd Stage 2 Phase Separator	18.77	18.77	18.77	18.77	31.18	31.18	31.18	31.18	31.18	31.18	31.18	31.18	12.86	12.86	12.86
		1st Stage HP Compressor Scrubber	13.23	13.23	13.23	13.23	25.02	25.02	25.02	25.02	25.02	25.02	25.02	25.02	8.38	8.38	8.38
		2nd Stage HP Compressor Scrubber	10.89	10.89	10.89	10.89	20.48	20.48	20.48	20.48	20.48	20.48	20.48	20.48	6.93	6.93	6.93
		2nd Stage LP Compresor Srubber	3.01	3.01	3.01	3.01	5.34	5.34	5.34	5.34	5.34	5.34	5.34	5.34	2.02	2.02	2.02
		Intermediate Scrubber	1.71	1.71	1.71	1.71	2.91	2.91	2.91	2.91	2.91	2.91	2.91	2.91	1.18	1.18	1.18
		Dehydration Scrubber	12.31	12.31	12.31	12.31	23.23	23.23	23.23	23.23	23.23	23.23	23.23	23.23	7.81	7.81	7.81
	HEAT EXCHANGER	1st Stage Compressor Discharge Cooler	2.27	7872298943	2.27	2.27	3.05	3.05	3.05	3.05	3.05	3.05	3.05	3.05	1.84	1.84	1.84
		2nd Stage Compressor Discharge Cooler	4.92	4.92	4.92	4.92	6.77	6.77	6.77	6.77	6.77	6.77	6.77	6.77	3.92	3.92	3.92
		2nd Stage HP Comp Cooler	18.88	18.88	18.88	18.88	26.32	26.32	26.32	26.32	26.32	26.32	26.32	26.32	14.83	14.83	14.83
		Export Cooler	11.49	11.49	11.49	11.49	15.96	15.96	15.96	15.96	15.96	15.96	15.96	15.96	9.06	9.06	9.06
	COMPRESSOR	1st st. LP Compressor	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19
		2nd st. LP Compressor	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19
		1st st. HP Compressor	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19
		2nd st. HP Compressor	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19	1.19
	PUMP	Seawater Pump 1	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52	1.52
		Seawater Pump 2	1.52	1.52	1.52	1.52	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	1.52	1.52	1.52
		Seawater Pump 3	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	1.52	1.52	1.52
		Seawater Pump 4	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	2.42	1.52	1.52	1.52
	PIPELINE	Total Pipeline															

Appendix L Investigative analysis

Appendix L.1 Cashflow analysis (scenario 1)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		170.15	170.15	170.15	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	204.18	170.15	102.09
Production Revenue - Condensate Sales ($$US mill)		104.03	104.03	104.03	208.07	208.07	208.07	208.07	208.07	208.07	208.07	208.07	208.07	124.80	104.05	62.40
Operating Cost (3% of CAPEX)	$ -14.76	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98
CO_2 Cost ($$US mill)		-1.96	-1.96	-1.96	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-2.35	-1.96	-1.17
Operating Income		271.24	271.24	271.24	543.48	543.48	543.48	543.48	543.48	543.48	543.48	543.48	543.48	325.65	271.27	162.33
Process Equipment Capital Cost	33.96															
Equipment Depreciation Per year	5.66															
Depreciation		-5.66	-5.66	-5.66	-5.66	-5.66	-5.66	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		265.58	265.58	265.58	537.82	537.82	537.82	543.48	543.48	543.48	543.48	543.48	543.48	325.65	271.27	162.33

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Tax		-207.15	-207.15	-207.15	-419.50	-419.50	-419.50	-423.91	-423.91	-423.91	-423.91	-423.91	-423.91	-254.01	-211.59	-126.62
Income after Tax		58.43	58.43	58.43	118.32	118.32	118.32	119.57	119.57	119.57	119.57	119.57	119.57	71.64	59.68	35.71
Income after Tax		58.43	58.43	58.43	118.32	118.32	118.32	119.57	119.57	119.57	119.57	119.57	119.57	71.64	59.68	35.71
Depreciation		5.66	5.66	5.66	5.66	5.66	5.66	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-492.14															
Cash Flow	-522.14	14.09	39.09	49.09	120.98	118.98	117.98	111.57	109.57	97.57	101.57	100.57	108.57	56.64	36.68	295.71
PV Cash Flow		13.04	33.51	38.97	88.92	80.98	74.35	65.10	59.19	48.81	47.04	43.13	43.11	20.83	12.49	93.22
Sum of Present Value	762.69															
Net Present Value	**240.56**															
Cumulative PV of Cash flow	-522.14	-509.09	-475.58	-436.61	-347.69	-266.72	-192.37	-127.27	-68.08	-19.27	27.78	70.91	114.02	134.85	147.33	240.56

Appendix L.2 Cashflow analysis (scenario 2)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		68.06	136.12	510.46	510.46	510.46	510.46	510.46	408.37	340.31	102.09	102.09	102.09	102.09	102.09	34.03
Production Revenue - Condensate Sales ($$US mill)		41.63	83.22	311.99	311.99	311.99	311.99	311.99	249.61	207.99	62.40	62.40	62.40	62.40	62.40	20.80
Operating Cost (3% of CAPEX)	$ -23.08	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54
CO_2 Cost ($$US mill)		-0.78	-1.56	-5.87	-5.87	-5.87	-5.87	-5.87	-4.69	-3.91	-1.17	-1.17	-1.17	-1.17	-1.17	-0.39
Operating Income		107.37	216.24	815.05	815.05	815.05	815.05	815.05	651.74	542.85	161.78	161.78	161.78	161.78	161.78	52.91
Process Equipment Capital Cost	53.08															
Equipment Depreciation Per year	8.85															
Depreciation		-8.85	-8.85	-8.85	-8.85	-8.85	-8.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		98.52	207.39	806.20	806.20	806.20	806.20	815.05	651.74	542.85	161.78	161.78	161.78	161.78	161.78	52.91
	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Tax		-76.85	-161.77	-628.84	-628.84	-628.84	-628.84	-635.74	-508.36	-423.42	-126.19	-126.19	-126.19	-126.19	-126.19	-41.27
Income after Tax		21.67	45.63	177.36	177.36	177.36	177.36	179.31	143.38	119.43	35.59	35.59	35.59	35.59	35.59	11.64
Income after Tax		21.67	45.63	177.36	177.36	177.36	177.36	179.31	143.38	119.43	35.59	35.59	35.59	35.59	35.59	11.64
Depreciation		8.85	8.85	8.85	8.85	8.85	8.85	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-769.26															
Cash Flow	-799.26	-19.48	29.47	171.21	183.21	181.21	180.21	171.31	133.38	97.43	17.59	16.59	24.59	20.59	12.59	271.64
PV Cash Flow		-18.04	25.27	135.91	134.67	123.33	113.56	99.96	72.06	48.74	8.15	7.12	9.77	7.57	4.29	85.63
Sum of Present Value	857.98															
Net Present Value	**58.72**															
Cumulative PV of Cash flow	-799.26	-817.30	-792.03	-656.11	-521.45	-398.12	-284.56	-184.60	-112.53	-63.80	-55.65	-48.53	-38.77	-31.20	-26.91	58.72

Appendix L.3 Cashflow analysis (scenario 3)

	YEAR															
	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Production Revenue - Gas ($$US mill)		170.15	1020.92	1020.92	1020.92	510.46	306.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Production Revenue - Condensate Sales ($$US mill)		104.03	624.28	624.28	624.28	312.04	187.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Operating Cost (3% of CAPEX)	$ -51.10	-3.41	-3.41	-3.41	-3.41	-3.41	-3.41	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2 Cost ($$US mill)		-1.96	-11.73	-11.73	-11.73	-5.87	-3.52	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Operating Income		268.82	1630.07	1630.07	1630.07	813.23	486.54	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Process Equipment Capital Cost	117.54															
Equipment Depreciation Per year	19.59															
Depreciation		-19.59	-19.59	-19.59	-19.59	-19.59	-19.59	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		249.23	1610.48	1610.48	1610.48	793.64	466.95	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Tax		-194.40	-1256.17	-1256.17	-1256.17	-619.04	-364.22	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income after Tax		54.83	354.30	354.30	354.30	174.60	102.73	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income after Tax		54.83	354.30	354.30	354.30	174.60	102.73	-	-	-	-	-	-	-	-	-
Depreciation		19.59	19.59	19.59	19.59	19.59	19.59	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-1703.41															
Cash Flow	-1733.41	24.42	348.89	358.89	370.89	189.19	116.32	-8.00	-10.00	-22.00	-18.00	-19.00	-11.00	-15.00	-23.00	260.00
PV Cash Flow		22.61	299.12	284.90	272.62	128.76	73.30	-4.67	-5.40	-11.01	-8.34	-8.15	-4.37	-5.52	-7.83	81.96
Sum of Present Value	1,108.00															
Net Present Value	**-625.41**															
Cumulative PV of Cash flow	-1733.41	-1710.80	-1411.68	-1126.78	-854.16	-725.40	-652.10	-656.77	-662.17	-673.18	-681.51	-689.66	-694.03	-699.55	-707.38	-625.41

Appendix L.4 Cashflow analysis (5.95 MMsm^3/d)

	YEAR																				
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**
Production Revenue - Gas ($$US mill)		202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.48	202.34	198.09	191.38	179.41
Production Revenue - Condensate Sales ($$US mill)		123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.74	123.66	121.11	117.04	109.65	100.08
Operating Cost (3% of CAPEX)	$ -8.53	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41	-0.41
CO_2 Cost ($$US mill)		-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.56	-0.54	-0.53	-0.49	-0.45
Operating Income		325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.18	322.50	314.20	300.13	278.64
Process Equipment Capital Cost	19.61																				
Equipment Depreciation Per year	3.27																				
Depreciation		-3.27	-3.27	-3.27	-3.27	-3.27	-3.27	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		322.00	322.00	322.00	322.00	322.00	322.00	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.26	325.18	322.50	314.20	300.13	278.64

	YEAR																				
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**
Tax		-251.16	-251.16	-251.16	-251.16	-251.16	-251.16	-253.71	-253.71	-253.71	-253.71	-253.71	-253.71	-253.71	-253.71	-253.71	-253.64	-251.55	-245.07	-234.10	-217.34
Income after Tax		70.84	70.84	70.84	70.84	70.84	70.84	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.54	70.95	69.12	66.03	61.30
Income after Tax		70.84	70.84	70.84	70.84	70.84	70.84	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.56	71.54	70.95	69.12	66.03	61.30
Depreciation		3.27	3.27	3.27	3.27	3.27	3.27	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	-5	-5	-5	-5	-5	-5
Investment	-284.18																				
Cash Flow	-314.18	24.11	49.11	59.11	71.11	69.11	68.11	63.56	61.56	49.56	53.56	52.56	60.56	56.56	48.56	66.56	66.54	65.95	64.12	61.03	56.30
PV Cash Flow		22.32	42.10	46.92	52.27	47.03	42.92	37.09	33.26	24.79	24.81	22.54	24.05	20.80	16.53	20.98	19.42	17.82	16.05	14.14	12.08
Sum of Present Value	621.44																				
Net Present Value	**307.26**																				
Cumulative PV of Cash flow	-314.18	-291.86	-249.76	-202.84	-150.57	-103.54	-60.62	-23.53	9.72	34.52	59.32	81.86	105.91	126.71	143.24	164.22	183.65	201.47	217.52	231.66	243.74

Appendix L.5 Cashflow analysis (suggested case- 8MMsm^3/d)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	272.25	238.22
Production Revenue - Condensate Sales ($$US mill)		166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	166.45	147.19
Operating Cost (3% of CAPEX)	$ -11.61	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77	-0.77
CO_2 Cost ($$US mill)		-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.75	-0.66
Operating Income		437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	383.97
Process Equipment Capital Cost	26.70															
Equipment Depreciation Per year	4.45															
Depreciation		-4.45	-4.45	-4.45	-4.45	-4.45	-4.45	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		432.73	432.73	432.73	432.73	432.73	432.73	437.18	437.18	437.18	437.18	437.18	437.18	437.18	437.18	383.97
Tax		-337.53	-337.53	-337.53	-337.53	-337.53	-337.53	-341.00	-341.00	-341.00	-341.00	-341.00	-341.00	-341.00	-341.00	-299.50

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Income after Tax		95.20	95.20	95.20	95.20	95.20	95.20	96.18	96.18	96.18	96.18	96.18	96.18	96.18	96.18	84.47
Income after Tax		95.20	95.20	95.20	95.20	95.20	95.20	96.18	96.18	96.18	96.18	96.18	96.18	96.18	96.18	84.47
Depreciation		4.45	4.45	4.45	4.45	4.45	4.45	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-386.91															
Cash Flow	-416.91	49.65	74.65	84.65	96.65	94.65	93.65	88.18	86.18	74.18	78.18	77.18	85.18	81.18	73.18	344.47
PV Cash Flow		45.97	64.00	67.20	71.04	64.42	59.02	51.45	46.56	37.11	36.21	33.10	33.83	29.85	24.91	108.59
Sum of Present Value	773.25															
Net Present Value	**356.34**															
Cumulative PV of Cash flow	-416.91	-370.94	-306.94	-239.75	-168.71	-104.29	-45.27	6.18	52.74	89.85	126.06	159.16	192.98	222.83	247.75	356.34

Appendix L.6 Cashflow analysis (10 MMsm^3/d)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	304.50	241.06	240.69	0.00	0.00
Production Revenue - Condensate Sales ($$US mill)		204.83	204.83	204.83	204.83	204.83	204.83	204.83	204.83	204.83	204.83	183.44	145.19	144.99	0.00	0.00
Operating Cost (3% of CAPEX)	$ -14.76	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98
CO_2 Cost ($$US mill)		-0.93	-0.93	-0.93	-0.93	-0.93	-0.93	-0.93	-0.93	-0.93	-0.93	-0.84	-0.66	-0.66	0.00	0.00
Operating Income		543.22	543.22	543.22	543.22	543.22	543.22	543.22	543.22	543.22	543.22	486.12	384.60	384.03	-0.98	-0.98
Process Equipment Capital Cost	33.95															
Equipment Depreciation Per year	5.66															
Depreciation		-5.66	-5.66	-5.66	-5.66	-5.66	-5.66	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		537.56	537.56	537.56	537.56	537.56	537.56	543.22	543.22	543.22	543.22	486.12	384.60	384.03	-0.98	-0.98
Tax		-419.30	-419.30	-419.30	-419.30	-419.30	-419.30	-423.71	-423.71	-423.71	-423.71	-379.17	-299.99	-299.54	0.77	0.77

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Income after Tax		118.26	118.26	118.26	118.26	118.26	118.26	119.51	119.51	119.51	119.51	106.95	84.61	84.49	-0.22	-0.22
Income after Tax		118.26	118.26	118.26	118.26	118.26	118.26	119.51	119.51	119.51	119.51	106.95	84.61	84.49	-0.22	-0.22
Depreciation		5.66	5.66	5.66	5.66	5.66	5.66	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-492.09															
Cash Flow	-522.09	73.92	98.92	108.92	120.92	118.92	117.92	111.51	109.51	97.51	101.51	87.95	73.61	69.49	-23.22	259.78
PV Cash Flow		68.45	84.81	86.47	88.88	80.94	74.31	65.06	59.16	48.78	47.02	37.72	29.23	25.55	-7.90	81.89
Sum of Present Value	870.36															
Net Present Value	**348.27**															
Cumulative PV of Cash flow	-522.09	-453.64	-368.83	-282.37	-193.49	-112.55	-38.24	26.82	85.99	134.76	181.78	219.50	248.73	274.28	266.38	348.27

Appendix L.7 Cashflow analysis (12 MMsm^3/d)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		408.37	408.37	408.37	408.37	408.37	408.37	408.37	408.37	403.75	319.64	59.32	0.00	0.00	0.00	0.00
Production Revenue - Condensate Sales ($$US mill)		249.67	249.67	249.67	249.67	249.67	249.67	249.67	247.01	195.39	154.75	147.23	0.00	0.00	0.00	0.00
Operating Cost (3% of CAPEX)	$ -18.00	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20	-1.20
CO_2 Cost ($$US mill)		-1.12	-1.12	-1.12	-1.12	-1.12	-1.12	-1.12	-1.11	-0.88	-0.70	-0.66	0.00	0.00	0.00	0.00
Operating Income		655.72	655.72	655.72	655.72	655.72	655.72	655.72	653.07	597.06	472.49	204.68	-1.20	-1.20	-1.20	-1.20
Process Equipment Capital Cost	41.40															
Equipment Depreciation Per year	6.90															
Depreciation		-6.90	-6.90	-6.90	-6.90	-6.90	-6.90	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		648.82	648.82	648.82	648.82	648.82	648.82	655.72	653.07	597.06	472.49	204.68	-1.20	-1.20	-1.20	-1.20
Tax		-506.08	-506.08	-506.08	-506.08	-506.08	-506.08	-511.46	-509.40	-465.71	-368.55	-159.65	0.94	0.94	0.94	0.94

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Income after Tax		142.74	142.74	142.74	142.74	142.74	142.74	144.26	143.68	131.35	103.95	45.03	-0.26	-0.26	-0.26	-0.26
Income after Tax		142.74	142.74	142.74	142.74	142.74	142.74	144.26	143.68	131.35	103.95	45.03	-0.26	-0.26	-0.26	-0.26
Depreciation		6.90	6.90	6.90	6.90	6.90	6.90	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-600.06															
Cash Flow	-630.06	99.64	124.64	134.64	146.64	144.64	143.64	136.26	133.68	109.35	85.95	26.03	-11.26	-15.26	-23.26	259.74
PV Cash Flow		92.26	106.86	106.88	107.79	98.44	90.52	79.51	72.22	54.70	39.81	11.16	-4.47	-5.61	-7.92	81.88
Sum of Present Value	924.03															
Net Present Value	**293.96**															
Cumulative PV of Cash flow	-630.06	-537.80	-430.94	-324.06	-216.27	-117.83	-27.32	52.19	124.41	179.11	218.93	230.09	225.62	220.00	212.08	293.96

Appendix L.8 Cashflow analysis (15 MMsm^3/d)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		510.46	510.46	510.46	510.46	510.46	510.46	446.29	353.31	187.29	0.00	0.00	0.00	0.00	0.00	0.00
Production Revenue - Condensate Sales ($$US mill)		312.10	312.10	312.10	312.10	312.10	272.84	215.99	170.98	147.10	0.00	0.00	0.00	0.00	0.00	0.00
Operating Cost (3% of CAPEX)	$ -23.09	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54
CO_2 Cost ($$US mill)		-1.40	-1.40	-1.40	-1.40	-1.40	-1.23	-0.97	-0.77	-0.66	0.00	0.00	0.00	0.00	0.00	0.00
Operating Income		819.62	819.62	819.62	819.62	819.62	780.54	659.77	521.98	332.19	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54
Process Equipment Capital Cost	53.11															
Equipment Depreciation Per year	8.85															
Depreciation		-8.85	-8.85	-8.85	-8.85	-8.85	-8.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		810.77	810.77	810.77	810.77	810.77	771.69	659.77	521.98	332.19	-1.54	-1.54	-1.54	-1.54	-1.54	-1.54
Tax		-632.40	-632.40	-632.40	-632.40	-632.40	-601.92	-514.62	- 407.15	-259.11	1.20	1.20	1.20	1.20	1.20	1.20

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Income after Tax		178.37	178.37	178.37	178.37	178.37	169.77	145.15	114.84	73.08	-0.34	-0.34	-0.34	-0.34	-0.34	-0.34
Income after Tax		178.37	178.37	178.37	178.37	178.37	169.77	145.15	114.84	73.08	-0.34	-0.34	-0.34	-0.34	-0.34	-0.34
Depreciation		8.85	8.85	8.85	8.85	8.85	8.85	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-769.76															
Cash Flow	-799.76	137.22	162.22	172.22	184.22	182.22	172.62	137.15	104.84	51.08	-18.34	-19.34	-11.34	-15.34	-23.34	259.66
PV Cash Flow		127.06	139.08	136.72	135.41	124.02	108.78	80.03	56.64	25.55	-8.49	-8.29	-4.50	-5.64	-7.95	81.86
Sum of Present Value	980.26															
Net Present Value	**180.50**															
Cumulative PV of Cash flow	-799.76	-672.70	-533.62	-396.90	-261.49	-137.48	-28.70	51.33	107.97	133.52	125.03	116.74	112.23	106.59	98.65	180.50

Appendix L.9 Cashflow analysis (20 MMsm^3/d)

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		680.62	680.62	680.62	658.98	521.69	413.01	326.97	87.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Production Revenue - Condensate Sales ($$US mill)		416.07	416.07	402.83	318.89	252.46	199.96	158.32	147.15	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Operating Cost (3% of CAPEX)	$ -31.94	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13
CO_2 Cost ($$US mill)		-1.87	-1.87	-1.81	-1.43	-1.13	-0.90	-0.71	-0.66	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Operating Income		1092.69	1092.69	1079.51	974.31	770.89	609.94	482.44	231.52	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13
Process Equipment Capital Cost	73.46															
Equipment Depreciation Per year	12.24															
Depreciation		-12.24	-12.24	-12.24	-12.24	-12.24	-12.24	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		1080.45	1080.45	1067.27	962.06	758.65	597.70	482.44	231.52	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13	-2.13
Tax		-842.75	-842.75	-832.47	-750.41	-591.75	-466.20	-376.31	-180.59	1.66	1.66	1.66	1.66	1.66	1.66	1.66

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Income after Tax		237.70	237.70	234.80	211.65	166.90	131.49	106.14	50.94	-0.47	-0.47	-0.47	-0.47	-0.47	-0.47	-0.47
Income after Tax		237.70	237.70	234.80	211.65	166.90	131.49	106.14	50.94	-0.47	-0.47	-0.47	-0.47	-0.47	-0.47	-0.47
Depreciation		12.24	12.24	12.24	12.24	12.24	12.24	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-1064.70															
Cash Flow	-1094.70	199.94	224.94	232.04	220.90	174.15	137.74	98.14	40.94	-22.47	-18.47	-19.47	-11.47	-15.47	-23.47	259.53
PV Cash Flow		185.13	192.85	184.20	162.37	118.52	86.80	57.26	22.12	-11.24	-8.55	-8.35	-4.55	-5.69	-7.99	81.82
Sum of Present Value	1,044.69															
Net Present Value	**-50.01**															
Cumulative PV of Cash flow	-1094.70	-909.56	-716.71	-532.51	-370.14	-251.62	-164.82	-107.56	-85.44	-96.68	-105.24	-113.59	-118.14	-123.83	-131.82	-50.01

Appendix M Scenario 1 - SRK and PR comparison

Appendix M.1 Equipment design

Soave-Redlich-Kwong

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total Footprint (m^2)	224.8	224.8	224.8	396.5	396.5	396.5	396.5	396.5	396.5	396.5	396.5	396.5	260.0	224.9	150.5
Total Weight (tons)	307.9	307.9	307.9	635.2	635.2	635.2	635.2	635.2	635.2	635.2	635.2	635.2	369.9	308.1	188.7
Total Duty (MW)	9.03	9.03	9.03	18.07	18.07	18.07	18.07	18.07	18.07	18.07	18.07	18.07	10.84	9.03	5.42
Compressor Power (MW)	7.47	7.47	7.47	14.95	14.95	14.95	14.95	14.95	14.95	14.95	14.95	14.95	8.97	7.47	4.48
Daily Compressor Energy (MWh)	179.4	179.4	179.4	358.7	358.7	358.7	358.7	358.7	358.7	358.7	358.7	358.7	215.2	179.4	107.6
Pump Power (kW)	232.5	232.5	232.5	473.4	473.4	473.4	473.4	473.4	473.4	473.4	473.4	473.4	279.5	232.5	138.8
Condensate (bbl/d)	5,340	5,340	5,340	10,680	10,680	10,680	10,680	10,680	10,680	10,680	10,680	10,680	6,406	5,341	3,203

Peng Robinson

Year/Case	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Total FootPrint (m2)	220.9	220.9	220.9	389.3	389.3	389.3	389.3	389.3	389.3	389.3	389.3	389.3	255.6	220.9	148.0
Total Weight (kg)	297.7	297.7	297.7	612.9	612.9	612.9	612.9	612.9	612.9	612.9	612.9	612.9	357.5	297.9	182.8
Total Duty (kW)	8.76	8.76	8.76	17.52	17.52	17.52	17.52	17.52	17.52	17.52	17.52	17.52	10.51	8.76	5.26
Compressor Power (kW)	7.08	7.08	7.08	14.16	14.16	14.16	14.16	14.16	14.16	14.16	14.16	14.16	8.50	7.08	4.25
Daily Compressor Energy (MWh)	170.0	170.0	170.0	339.9	339.9	339.9	339.9	339.9	339.9	339.9	339.9	339.9	204.0	170.0	102.0
Pump Power (kW)	230.5	230.5	230.5	469.8	469.8	469.8	469.8	469.8	469.8	469.8	469.8	469.8	277.4	230.5	137.8
Condensate (bbl/d)	5,251	5,251	5,251	10,502	10,502	10,502	10,502	10,502	10,502	10,502	10,502	10,502	6,303	5,254	3,151

Appendix M.2 Carbon footprint

Soave-Redlich-Kwong

Year/Case	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Fuel Consumption (kg/s)	0.46	0.46	0.46	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.92	0.55	0.46	0.28
Fuel Consumption (sm^3/s)	0.55	0.55	0.55	1.11	1.11	1.11	1.11	1.11	1.11	1.11	1.11	1.11	0.67	0.55	0.33
CO_2 emissions cost per year (MM NOK /year)	15.24	15.24	15.24	30.49	30.49	30.49	30.49	30.49	30.49	30.49	30.49	30.49	18.29	15.24	9.15
CO_2 intensity (kg CO2 per BOE)	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70	2.70

Peng Robinson

Year/Case	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Fuel Consumption (kg/s)	0.44	0.44	0.44	0.87	0.87	0.87	0.87	0.87	0.87	0.87	0.87	0.87	0.52	0.44	0.26
Fuel Consumption (sm3/s)	0.53	0.53	0.53	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	0.63	0.53	0.32
CO2 emissions cost per year (MM NOK /year)	14.45	14.45	14.45	28.89	28.89	28.89	28.89	28.89	28.89	28.89	28.89	28.89	17.33	14.45	8.67
CO2 efficency (kg CO2 per BOE)	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56	2.56

Appendix M.3 Cash flow analysis

Soave-Redlich-Kwong

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		170.15	170.15	170.15	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	204.18	170.15	102.09
Production Revenue - Condensate Sales ($$US mill)		104.03	104.03	104.03	208.07	208.07	208.07	208.07	208.07	208.07	208.07	208.07	208.07	124.80	104.05	62.40
Operating Cost (3% of CAPEX)	$ -14.76	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98	-0.98
CO_2 Cost ($$US mill)		-1.96	-1.96	-1.96	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-3.91	-2.35	-1.96	-1.17
Operating Income		271.24	271.24	271.24	543.48	543.48	543.48	543.48	543.48	543.48	543.48	543.48	543.48	325.65	271.27	162.33
Process Equipment Capital Cost	33.96															
Equipment Depreciation Per year	5.66															
Depreciation		-5.66	-5.66	-5.66	-5.66	-5.66	-5.66	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		265.58	265.58	265.58	537.82	537.82	537.82	543.48	543.48	543.48	543.48	543.48	543.48	325.65	271.27	162.33

	YEAR															
	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Tax		-207.15	-207.15	-207.15	-419.50	-419.50	-419.50	-423.91	-423.91	-423.91	-423.91	-423.91	-423.91	-254.01	-211.59	-126.62
Income after Tax		58.43	58.43	58.43	118.32	118.32	118.32	119.57	119.57	119.57	119.57	119.57	119.57	71.64	59.68	35.71
Income after Tax		58.43	58.43	58.43	118.32	118.32	118.32	119.57	119.57	119.57	119.57	119.57	119.57	71.64	59.68	35.71
Depreciation		5.66	5.66	5.66	5.66	5.66	5.66	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-492.14															
Cash Flow	-522.14	14.09	39.09	49.09	120.98	118.98	117.98	111.57	109.57	97.57	101.57	100.57	108.57	56.64	36.68	295.71
PV Cash Flow		13.04	33.51	38.97	88.92	80.98	74.35	65.10	59.19	48.81	47.04	43.13	43.11	20.83	12.49	93.22
Sum of Present Value	762.69															
Net Present Value	**240.56**															
Cumulative PV of Cash flow	-522.14	-509.09	-475.58	-436.61	-347.69	-266.72	-192.37	-127.27	-68.08	-19.27	27.78	70.91	114.02	134.85	147.33	240.56

Peng Robinson

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Production Revenue - Gas ($$US mill)		170.15	170.15	170.15	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	340.31	204.18	170.15	102.09
Production Revenue - Condensate Sales ($$US mill)		102.31	102.31	102.31	204.61	204.61	204.61	204.61	204.61	204.61	204.61	204.61	204.61	122.79	102.36	61.39
Operating Cost (3% of CAPEX)	$ -14.25	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95
CO_2 Cost ($$US mill)		-1.85	-1.85	-1.85	-3.71	-3.71	-3.71	-3.71	-3.71	-3.71	-3.71	-3.71	-3.71	-2.22	-1.85	-1.11
Operating Income		269.66	269.66	269.66	540.26	540.26	540.26	540.26	540.26	540.26	540.26	540.26	540.26	323.80	269.71	161.42
Process Equipment Capital Cost	32.76															
Equipment Depreciation Per year	5.46															
Depreciation		-5.46	-5.46	-5.46	-5.46	-5.46	-5.46	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Income before Tax		264.20	264.20	264.20	534.80	534.80	534.80	540.26	540.26	540.26	540.26	540.26	540.26	323.80	269.71	161.42

	YEAR															
	0	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Tax		-206.07	-206.07	-206.07	-417.14	-417.14	-417.14	-421.40	-421.40	-421.40	-421.40	-421.40	-421.40	-252.56	-210.38	-125.91
Income after Tax		58.12	58.12	58.12	117.66	117.66	117.66	118.86	118.86	118.86	118.86	118.86	118.86	71.24	59.34	35.51
Income after Tax		58.12	58.12	58.12	117.66	117.66	117.66	118.86	118.86	118.86	118.86	118.86	118.86	71.24	59.34	35.51
Depreciation		5.46	5.46	5.46	5.46	5.46	5.46	-	-	-	-	-	-	-	-	-
Change in Working Capital	-30	-50	-25	-15	-3	-5	-6	-8	-10	-22	-18	-19	-11	-15	-23	260
Investment	-474.85															
Cash Flow	-504.85	13.58	38.58	48.58	120.12	118.12	117.12	110.86	108.86	96.86	100.86	99.86	107.86	56.24	36.34	295.51
PV Cash Flow		12.58	33.08	38.57	88.29	80.39	73.80	64.68	58.81	48.45	46.72	42.83	42.83	20.68	12.37	93.16
Sum of Present Value	757.24															
Net Present Value	**252.38**															
Cumulative PV of Cash flow	-504.85	-492.28	-459.20	-420.63	-332.34	-251.95	-178.15	-113.46	-54.65	-6.20	40.52	83.34	126.18	146.85	159.22	252.38

Appendix N Risk Assessment

NTNU	Hazardous activity identification process	Prepared by	Number	Date
HSE		HSE section	HMSRV2601E	09.01.2013
		Approved by		Replaces
		The Rector		01.12.2006

Unit: Energy and Process Engineering Department **Date:** 6th June 2018

Line manager: Even Solbraa

Participants in the identification process (including their function): John Swatson – Masters Student

Short description of the main activity/main process: Master project for student John Swatson. Thesis Title: Automated Process Design in Oil and Gas Field Development

Is the project work purely theoretical? (YES/NO): YES *Answer "YES" implies that supervisor is assured that no activities requiring risk assessment are involved in the work. If YES, briefly describe the activities below. The risk assessment form need not be filled out.*

Signatures: *Responsible supervisor:* Even Solbraa *Student:*

ID nr.	Activity/process	Responsible person	Existing documentation	Existing safety measures	Laws, regulations etc.	Comment
	Master thesis involves purely theoretical research and performing simulations. No experimental work nor risk assement required.					