Optimization

of

Offshore Oil&Gas Separation Train

Master Thesis Project

K10k-3-F17

Aalborg University Esbjerg

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Title Page

Title: Optimization of Offshore Oil&Gas Separation Train

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Abstract

The aim of this project was to develop optimization strategies to find the optimal operating pressure and temperature values for an offshore Oil&Gas separation train, in order to maximize the recovery of C_{3+} hydrocarbons in a stable Crude Oil stream at storage conditions, and the recovery of $C_{1\&2}$ hydrocarbons in Gas phase.

Three different kinds of petroleum reservoir fluids were chosen for this study: Gas Condensate, Near Critical and Black Oil mixtures (ascending molecular weight order). Moreover, 60 bar and 70°C, 30 bar and 40°C were selected as feed stream flowing conditions, while 1.2 bar and 15°C were fixed as stock tank pressure and temperature. The Original Optimizer of the Aspen HYSYS process simulator was used as optimization tool.

Two different simulation flowsheets were developed: a separation train composed by 2 two – phase separators and a sequence of 3 two – phase separators, before the storage tank. It was assumed that pressure and temperature of the first separator of the train are fixed and equal to the reservoir fluids flowing conditions. The stability constraint of the optimization was decided to be expressed by the bubble point pressure of the Crude Oil delivered to the stock tank: this pressure must be equal or lower than the 70% of the stock pressure condition (0.85 bar).

Performing the constrained optimization by mean of the BOX method subroutine of Aspen HYSYS, it results that the production of Crude Oil, matching at the same time the recovery requirements and the stability constraint, can be realized setting the pressure of the last separator equal to the tank pressure and the temperature higher than the one at stock tank. Moreover, when using a sequence of 3 separators, large pressure drops should be realized between the first and the second separator.

As the number of separators increases, the recovery of C_{3+} hydrocarbons increases according to the kind of mixture and the first separator operating conditions: for Gas Condensate – like fluids the recovery increases from 19.1% to 25.3%, for Near Critical – like streams from 6.2% to 7.7%, and for Black Oil – like mixtures from 0.01% up to 2.4%.

The simulation of the separation train does not highlight a significant change of the optimal pressures and temperatures, when varying the composition of the reservoir fluid processed. On the other hand, the heavier the fluid processed, the higher the recovery of C_{3+} hydrocarbons in Crude Oil and the lower the amount of propane and heavier hydrocarbons to be recovered from the Gas phase.

Preface & Acknowledgments

This is a 10th semester Master Thesis project, which was written in the spring semester of 2017 by a student enrolled in the Chemical Engineering Master's programme at Aalborg University Esbjerg. The list of references can be found on page 45. References are enclosed into round brackets, showing authors and publication year.

In this report, decimals are separated by a dot.

After References, Appendix can be found. The Appendix includes the tables containing the data generated during the different study cases developed.

*Dedicated to **Albano Mariarosaria**,*
your love is always cornerstone of my life

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Chapter 1

1.1. Introduction

Starting from the 19th century, fossil fuels became the main source of energy (by mean of combustion), due to the development and rise of new technology in many fields, but especially in transportations, and because of their relative low price. Fossil fuels include Crude Oil, Gas and Coal supplying the 41%, 24% and 23% of the energy consumption worldwide respectively (The Ecology Global Network, 2010).

Furthermore, several commonly used materials are produced by processing chemicals derived from Crude Oil. Important examples are plastics, which are high molecular weight organic polymers, where the main chain often has a petrochemical as principal compound. One of the most widely used polymers is polyester (ester functional group in the main chain), commonly employed to produce fabrics for clothes. Other examples are fertilizers and common chemicals like detergents, solvents and adhesives. The former consists in compounds that can improve the quality of the soil enhancing plants growth. Urea ($CO(NH_2)_2$) is an important fertilizer obtained by processing ammonia (NH_3), which is produced through hydrogen derived from hydrocarbons. The latter are derived from olefins, which are products of the fluid catalytic cracking of Crude Oil fractions (Petrarulo, 2016).
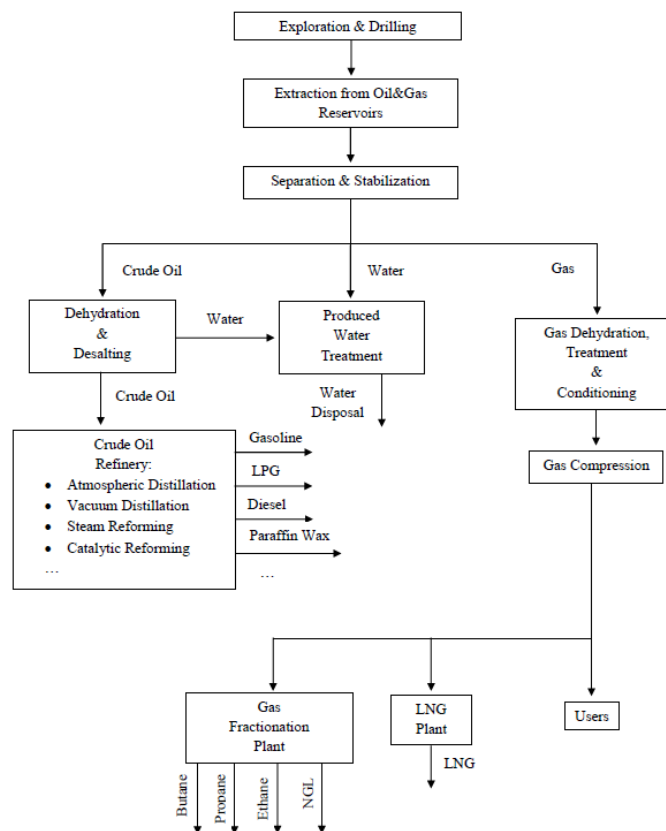

Figure 1. Oil&Gas Industry (adapted from image of (Piping Engineering, 2015))

As shown in Figure 1, Oil&Gas Industry involves the operations of exploration and drilling of wells, which are spread worldwide, and the possible consequent extraction of the so – called ***petroleum reservoir fluids*** and their processing for the delivery of different products.

North Sea Oil&Gas industry began in the 1960's with the discovery of the first significant reservoirs, but that area became a key non – OPEC (*Organization of the Petroleum Exporting Countries*) region just between 1980's and 1990's. Indeed, in those years major projects for the exploitation of wells came on stream. Because of the large depths of the North Sea reservoirs and the climate inhospitality of the region, the Oil&Gas industry requires high – technology offshore facilities in this area. Therefore, production results to have a relatively high cost. Despite that, the political stability and the ideal position respect to the European market allowed this area to become a major producer in the Oil&Gas market. Currently, North Sea production of Crude Oil and Natural Gas reached a plateau and a long – term decline is expected to begin. According to *CIA World Factbook*, in 2008 North Sea proven Crude Oil reserves accounted for about 11.2 billion of oil barrel: 59% is owned by Norway, the 30% belongs to the UK and the 7% is disposed by Denmark. The remaining 4% is shared between Germany and Holland. In the near future, the improvement of the Crude Oil recovery, together with new projects and the discovery of new sizeable volumes of reservoir fluids, will delay the downward trend of Oil&Gas Industry in the North Sea (Offshore Center Danmark, 2010).

1.2. Petroleum Reservoir Fluids

Petroleum reservoir fluids are mixtures of thousands of components, consisting mainly in hydrocarbons and small amounts of inorganic chemicals, among which water, carbon dioxide (CO_2), nitrogen (N_2), hydrogen sulphide (H_2S) are the most common (Pedersen & Christensen, 2007).

Those fluids originated from the bodies of single celled aquatic organisms, which are the progenitors of the current algae, plankton etc. The process of petroleum formation starts with the burial of these organisms under layers of clay and sediments, forming a matrix made up of organic matter and clay. This matrix is slowly turned into a new material called *kerogen*. If the deposition of clay and sediments continues, kerogen is entombed deeper and deeper, with a corresponding increase of pressure and temperature. For periods of time long enough (millions of years) and between 760 m and 4900 m depth, kerogen transformation into oil and gas occurs. Because of differences of pressure, petroleum fluid starts to flow laterally, following the path formed by permeable rocks, until reaching dome structures, called *anticlines,* where it is trapped. Consequently, petroleum starts to accumulate over time (formation of true reservoirs can thus take eons). Buoyancy forces in the reservoirs produce a separation in the fluids, where water disposes in the bottom while natural gas floats on the top. If pressure is high enough, part of the gas is dissolved in petroleum fluids (Tabak, 2009).

Petroleum reservoir fluids can be classified according to the critical temperatures of those hydrocarbon multicomponent mixtures compared to the temperature of the reservoir. Indeed, considering a phase envelope in a PT diagram, critical temperature is the meeting point of bubble and dew pressure branches.
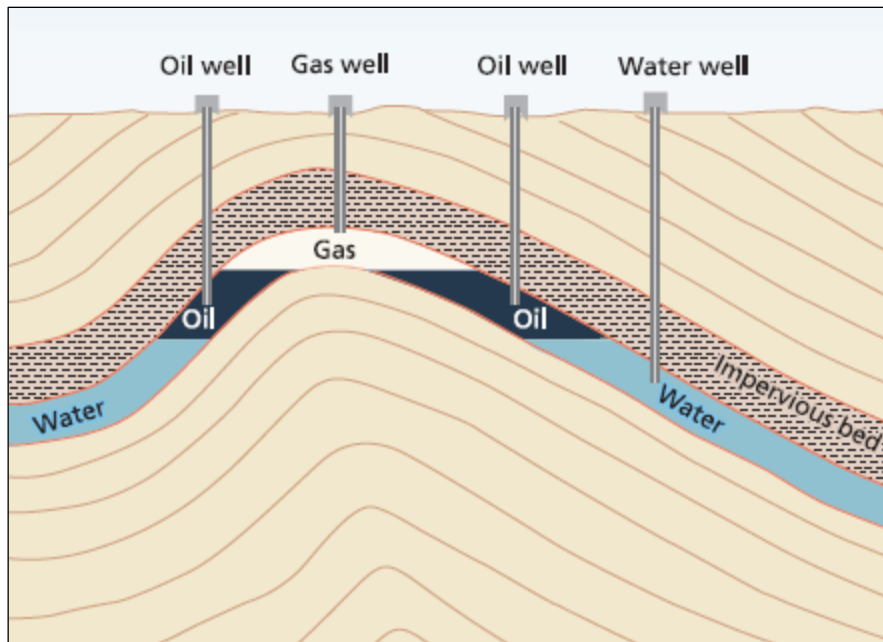
Figure 2. Structure of a Typical Petroleum Reservoir Fluid Well (Tabak, 2009)

Therefore, varying the composition of a well fluid, the critical temperature and the branches of the phase envelope change accordingly (Figure 3). While reservoir temperature (T_{res} and T'_{res} in Figure 3) remains almost constant during exploiting, extraction of fluids from wells produces a decrease of pressure inside the reservoir. Thus, extracting matter from the reservoir, the position of the system in the PT diagram moves approaching one of two branches. This means that the number of phases in the reservoir can change. Four main groups of fluids can be identified (Pedersen & Christensen, 2007):

- *Natural Gas*: the reservoir fluid is gas phase and pressure decrease does not produce the formation of a new phase
- *Gas Condensate*: the reservoir fluid is a gas phase mainly. Pressure decrease leads to meet the dew point pressure branch, therefore the formation of a liquid phase occurs.
- *Near Critical*: the reservoir fluid is mainly a liquid phase. If the well temperature is smaller than mixture critical temperature, decreasing pressure approaches bubble point pressure branch, meaning a gas phase is forming. If well temperature is higher than mixture critical temperature, decreasing pressure approaches dew point pressure branch, meaning a second liquid phase is forming.
- *Black Oil*: the reservoir fluid is mainly a liquid phase and pressure decrease always leads to meet the bubble point pressure branch, and thus to the formation of a gas phase.

Generally speaking, the lightest hydrocarbons present in this kind of fluids are methane (CH_4), ethane (C_2H_6) and propane (C_3H_8), which are referred to as C_1, C_2 and C_3 respectively. Heavier hydrocarbons are similarly referred to the same way using the number of carbon atoms.
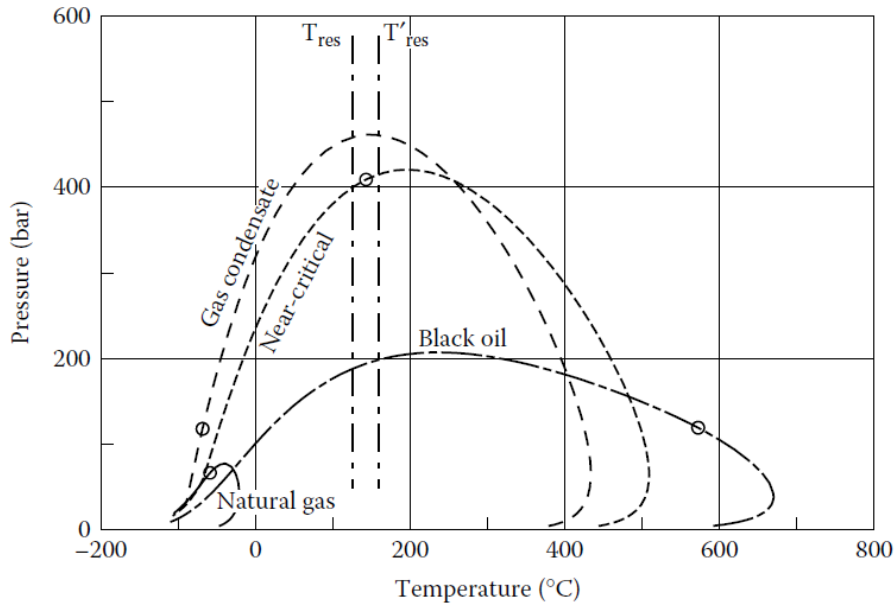
Figure 3. PT Phase Envelope Diagram for Natural Gas, Gas Condensate, Near – Critical and Black Oil Reservoir Fluids (Pedersen & Christensen, 2007)

However, since the high number of different hydrocarbons composing reservoir mixtures, a complete compositional analysis of those fluids is almost impossible to be performed. Moreover, reservoir fluids' hydrocarbons can belong to three components classes (Pedersen & Christensen, 2007):

- *Paraffins*, also known as *alkanes*, are hydrocarbons where carbons are linked by single bonds (Figure 4). *Normal* – paraffins are straight – chain compounds, while *iso* – paraffins contains at least one side chains.
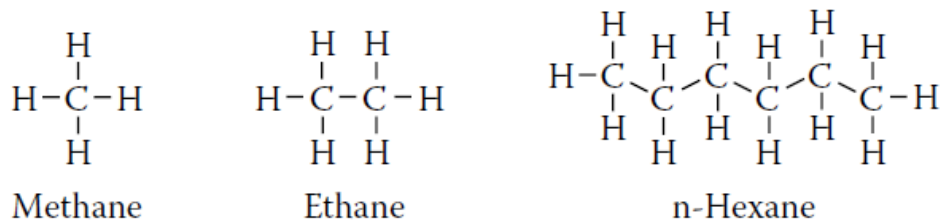
Figure 4. Examples of Chemical Structures of Paraffins (Pedersen & Christensen, 2007)

- *Naphthenes*, also known as *cycloalkanes*, are hydrocarbon compounds similar to paraffins, but show a cyclic structure (Figure 5). Carbon atoms forming the ring structures are linked by single bonds. Petroleum reservoir fluids commonly contain naphthenic rings with 5, 6 or 7 carbons.
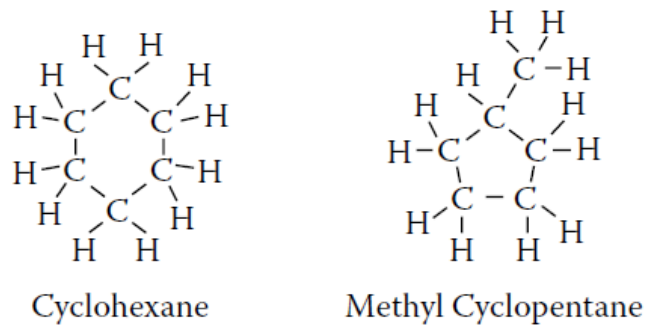
Figure 5. Examples of Chemical Structures of Naphtenes (Pedersen & Christensen, 2007)

- *Aromatics* are hydrocarbon chemicals containing one or more hexagonal structures, where carbon atoms are connected by alternating double and single bonds (Figure 6). The simplest aromatic compound is benzene (C_6H_6), but in reservoir fluids it is common to find polycyclic aromatics like naphthalene ($C_{10}H_8$).
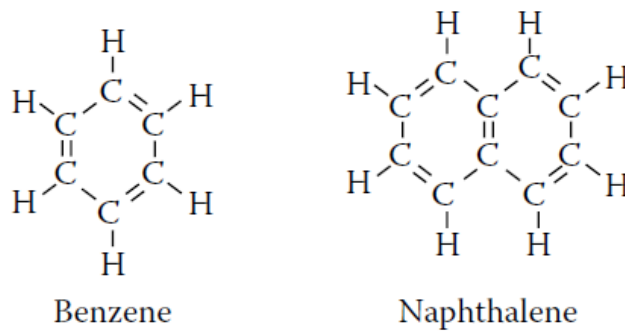
Figure 6. Examples of Chemical Structures of Aromatics (Pedersen & Christensen, 2007)

Therefore, starting from C_6, it is possible to find different kinds of hydrocarbons with the same number of carbon atoms. Generally speaking, densities of hydrocarbons with the same number of carbon atoms increase according to the scheme:

$$P \rightarrow N \rightarrow A$$
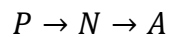
Moreover, considering the data regarding properties of some reservoir fluids constituents listed in Table 1, physical properties of reservoir fluids components can vary in a wide range.

Table 1. Physical Properties of some Reservoir Fluids Components

Component	Formula	Molecular Weight (g/mol)	Normal Boiling Point (°C)	Density at 1 atm and 20°C (g/cm³)	Critical Temperature (°C)	Critical Pressure (bar)	Acentric Factor
Nitrogen	N_2	28.01	-195.8	0.0012	-147	33.9	0.04
Carbon Dioxide	CO_2	44.01	-78.5	0.0019	31.1	73.8	0.225
Methane	CH_4	16.04	-161.6	0.0007	-82.6	46	0.008
Ethane	C_2H_6	30.07	-87.6	0.0028	32.3	48.8	0.098
Propane	C_3H_8	44.09	-42.1	0.002	96.7	42.5	0.152
n – Hexane	C_6H_{14}	86.17	68.8	0.659	234.3	29.7	0.296
Cyclohexane	C_6H_{12}	84.16	80.7	0.779	280.4	40.7	0.212
Benzene	C_6H_6	78.11	80.1	0.885	289	48.9	0.212
n – Decane	$C_{10}H_{22}$	142.28	174.2	0.730	344.6	21.2	0.489
Naphthalene	$C_{10}H_8$	128.17	218	0.971	475.3	40.5	0.302

A reliable characterization of petroleum reservoir fluids can be obtained grouping hydrocarbons heavier than nC_5 into cuts named *carbon number fractions*, based on their normal boiling points. General cut points are fixed using n – paraffins boiling points: C_6 fraction includes hydrocarbons whose boiling point is between 0.5°C above the boiling point of nC_5 and 0.5°C above the boiling point of nC_7. Similarly, C_7 fraction includes hydrocarbons whose boiling point is between 0.5°C above the boiling point of nC_6 and 0.5°C above the boiling point of nC_8, etc. The content of paraffins (P), naphthenes (N) and aromatics (A) per each carbon number fraction cut is called PNA distribution, and it can be used as reference to have an idea of the density of cuts and vice versa: the higher the density, the more aromatic the cut and the higher the content of aromatics, the denser the carbon faction.

Compositional analysis is carried out analysing the sample of the reservoirs, which can be taken in the bottom of the well (single phase) or after a first separation made after the extraction (gas and liquid sample). Both kind of samples are then flashed at standard conditions (1.01 bar, 288.15 K) and the two phases are analysed separately. *Gas – Chromatography (GC)* and *True Boiling Point (TBP)* are common analytical analysis used for the characterization of reservoir fluids samples. The former is ideal for the gas phase: a small sample of the gas is injected in a GC column, which separates the components showing a peak per each one on a chromatogram. Using proper detectors and performing the analysis the right way, it is possible to identify and quantify each one component of the gas phase. The TBP analysis is well suited for the characterization of the liquid phase: the samples' components are split based on their boiling points, identifying the different cuts whose molecular weight and density is then measured. The distillation is performed at atmospheric pressure up to C_{10}, afterwards the pressure is lowered to 2.26 mbar, allowing the separation of the heavier hydrocarbons and avoiding their thermal decomposition. The measurement of the molecular weight is performed through the study of the freezing point depression phenomenon, using a suitable solvent. On the other hand, density is usually measured by mean of densitometers (Pedersen & Christensen, 2007).

Pressure and temperature inside reservoirs depend on their depths and the heat exchange through their walls, and needs to be measured case by case. Generally speaking, wells pressure can vary between 150 bar and 410 bar, while temperature rarely is higher than 100°C. At these conditions, the liquid phase to be extracted contains water and absorbed gases (Society of Petroleum Engineers, s.d.).

After extraction, operating pressure and temperature of well fluids are lower than in the reservoir, and the flow is a mixture of oil, gas and water. Therefore, a separation process is required to split the flow into the different phases for the following processing.

1.3. Offshore Facilities

Offshore Crude Oil and Gas production involves facilities which allow the extraction of petroleum reservoir fluids from beneath the sea, and to process them to separate Crude Oil from the Gas, whose further treatment leads to the production of LNG and other products like propane, butane etc.
Oil&Gas production from subsea began in the 1950s, exploiting shallow reservoirs (200 m seabed depth) using platforms built on the seabed using concrete and metal legs. Since the 1990s, companies began to extract petroleum fluids from deeper reservoirs (300 m – 1500 m seabed depth). Nowadays, reservoirs fluids are extracted from wells deep up to 3000 m – 3500 m under the seabed, accounting for the 30% of the worldwide Crude Oil production and the 27% of the global Natural Gas production (Total S.A., 2015).
Different kinds of offshore facilities can be employed, depending on size and the depth of seabed. Common offshore structures are the following (Devold, 2013):

- *Shallow Water Complex Platform*: it is used for water depths up to 100 m and it is composed by independent platforms connected using gangway bridges (Figure 7). Each one of those units includes a different part of the whole process and utilities**.**

Figure 7. Shallow Water Complex Platform (Csanyi, 2017)

- *Gravity Base Platform*: it consists in a platform placed on fixed and large steel reinforced concrete structures, built on the seabed (Figure 8). It is used in the case of large fields with a water depth between 100 m and 500 m and it includes Crude Oil storage tanks placed on the bottom of the concrete support legs.

Figure 8. Gravity Base Platform (Csanyi, 2017)

- *Floating Production Systems*: these are production sites where facilities are placed on floating structure (Figure 9). The most common are the following:

  - *Floating Production Storage and Offloading (FPSO)*: it is an independent structure, where the Crude Oil produced is offloaded periodically through a shuttle tanker. They can produce between 10000 and 200000 barrels of Crude Oil per day.

  - *Tension Leg Platform (TLP)*: it is a floating structure hold in a specific position by tensioned tendons, fixed on the seabed through several piles. It can be used for a wide range of water depths up to 2000 m.

  - *SPAR*: it is a structure consisting in a floating cylindrical hull which supports a fixed platform.

- *Subsea Production Systems*: these structures are placed on the sea floor and can only extract and transport reservoir fluids towards other facilities for the separation, through undersea pipelines (Figure 9). Subsea systems are commonly used for water depths of 500 m or more.
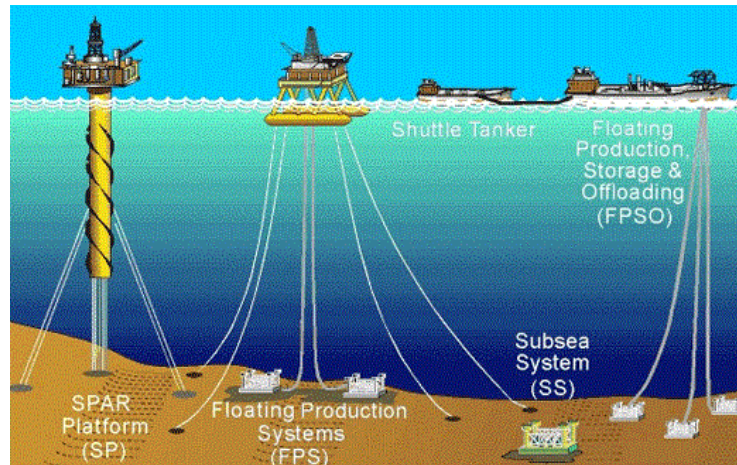
Figure 9. Floating & Subsea Production Systems (American Petroleum Institute, 2017)

Offshore production facilities have different constraints concerning costs, space availability and environmental issues. Indeed, each platform is built on purpose, based on the reservoir fluid(s) expected to receive as feed and the kind of well to be exploited, making investments for billions of dollars. Therefore, choices regarding production equipment need to be carefully considered. Moreover, since space is limited on offshore platforms and enlargements are not likely possible after building, the equipment has to be as compact and lightweight as possible. Furthermore, blowout risks with consequent environmental damage are important possible events on this kind of facilities. It is thus important to foresee potential issues and plan emergency cases solutions and constant vigilance ((Magalhães, et al., 2015), (Total S.A., 2015)).

1.3.1. Separation Process

The target of the separation of reservoir fluids from undersea wells is to produce a Gas stream as much free as possible of C_{3+} hydrocarbons and a Crude Oil stable at storage conditions. Indeed, the Crude Oil produced must not undergo to any vaporization when delivered into the storage tank and in the event of small variations of storage pressure and/or temperature.

As it is shown in Figure 10, those aims are often accomplished by progressively reducing pressure and temperature of the fluid, through a multistage separation consisting in a sequence of two or three separators, which forms a configuration named *separation train*. The first separator is usually a vessel where liquid (Crude Oil), Gas and water are separated mainly by density, at reservoir fluid pressure and temperature, unless different conditions of pressure and temperature are required. Gas phase goes to the top of the vessel and is sent to the gas treatment unit, where dehydration, C_{3+} condensate removal and other processes are performed. Water of the reservoir fluid processed goes to the bottom of the separator and then routed to the produced water treatment unit, before it is available for reinjection in the well or disposal in the sea. Crude Oil from the first separator flows to the second separator, where pressure and/or temperature are reduced. This way a minor further fraction of gas is released by the vaporization of the light hydrocarbons still present in the liquid phase. Moreover, a further small amount of water is removed from the liquid and is combined with the water coming out of the first separator. A third separator might be required for a further separation to meet the liquid product requirements, regarding stability and recovery, and further reduce water content. Gas released from the second and, eventually, third separator is compressed and mixed with the gas coming out from the first separator. The Crude Oil produced is then stored into tanks, before being routed towards the oil treatment unit for the removal of the remaining water and salts, and then to the refinery plant (Piping Engineering, 2015).
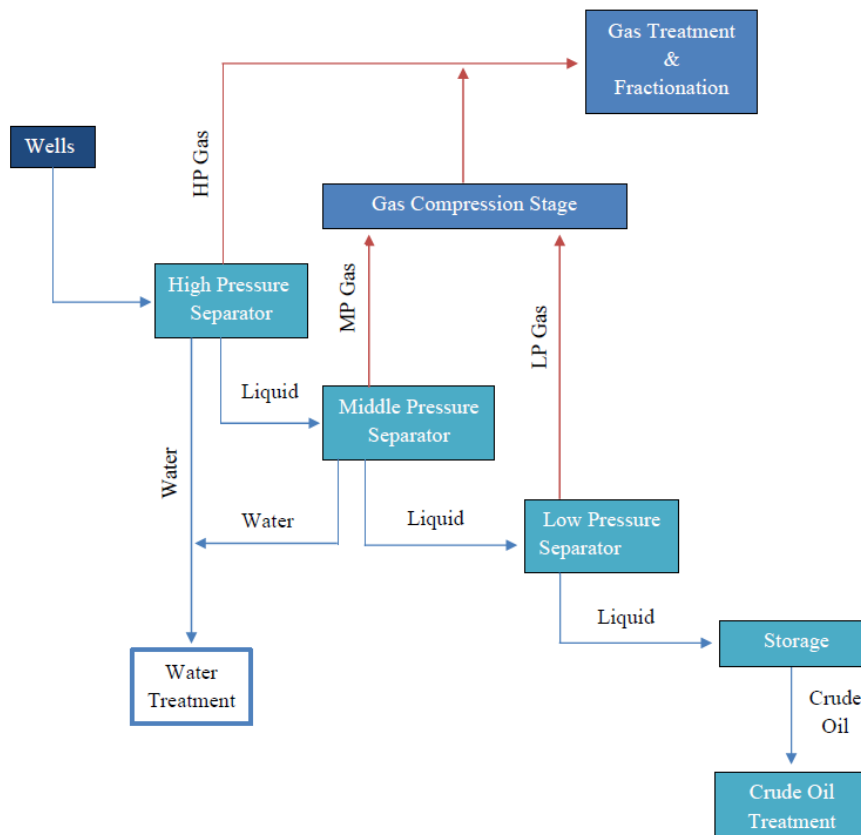
Figure 10. Oil&Gas Separation Train

Separators to be used are primarily classified depending on the number of phases they can manage. According to this principle, it is possible to distinguish two kinds of separators commonly used in Oil&Gas Industry (Piping Engineering, 2015):

- *Two Phase Separator*: it separates the reservoir fluid flow into a liquid and a gas phases.

- *Three Phase Separator*: it splits the well fluid into oil, gas and water flows, mainly by density.

Furthermore, separators can be distinguished into two main different classes depending on their configuration (Piping Engineering, 2015):

- *Horizontal Separator*: it is a common choice for three phase separations and reservoir fluids with a low amount of gas with respect to the amount of oil (*Gas Oil Ratio – GOR*). It is easy to install and provides a sufficient residence time for the liquid – liquid split, with a large area for the liquid phase that reduces the turbulence through a better foam dispersion. However, this kind of vessel is difficult to clean from deposits of sand, wax, paraffins etc., the liquid level control results to be more critical and only a portion of the vessel shell is available for gas release.

- *Vertical Separator*: it is a well – suited choice in the event of reservoir fluids with a high GOR and for two – phase separations. Despite the difficulty of the installation, the bigger diameter required for the same amount of gas managed by a horizontal separator and the higher cost, the separation of the oil from gas is easier due to the complete vessel diameter available for the gas flowing to the top and liquid level control is less critical. Moreover, the removal of deposits results to be easier.

1.3.2. Oil Treatment & Storage

After the separation train, Crude Oil still contains up to the 2% of water together with soluble and insoluble salts. Additional treatments are required to remove those salts and water traces since they are likely to form a more and more stable emulsion and sediments over time, causing issues as corrosion, incrustation and bubbles in the pipeline during transportation of the oil.

Crude Oil **dehydration** on offshore facilities is commonly performed by injecting the proper amount of a demulsifier chemical and introducing the oil into an electrostatic coalescer. The target is to reduce water content up to **0.5%**. The demulsifier reduces the interfacial tension between oil (continuous phase) and water (dispersed phase), destabilizing the emulsion. At the same time, the electrostatic coalescer applies an electrostatic field to the fluid, charging the water droplets and forming dipoles that enhance water particles coalescence. Moreover, fluid heating allows to reduce oil viscosity, allowing water droplets to move more easily and collide with higher forces. This dehydration technique is the most widely used since it allows to minimize the space and weight required for the operation equipment, and the amount and cost of the demulsifier to be used.

Crude Oil **desalting** is carried out by mean of mixing the proper amount of fresh water with dehydrated oil, dissolving the salt crystals. Afterwards, a second dehydration produces an oil containing a less saline and proper amount of water. In particular, salts content needs to be between **4.5 kg** and **13.6 kg** per thousand Crude Oil barrels.

Physical properties and some compositional aspects of the Crude Oil after separation and treatment can vary in a wide range. Those physical and compositional features are included in the crude assays, which are reports used for sales purposes by the producers. Typical main characteristics listed in those assays are the following (Piping Engineering, 2015):

- *Specific gravity (SG) & API gravity*: they are two expressions of Crude Oil density commonly used in oil industry

$$SG = \frac{\rho_{Crude\ Oil}}{\rho_{Water}}$$

Equation 1

$$°API = \frac{141.5}{SG} - 131.5$$

Equation 2

Where $\rho_{CrudeOil}$ and ρ_{Water} are the density of Crude Oil and water respectively, expressed in g/cm^3 (density of water is set at 1 g/cm^3). Specific gravity is measured experimentally with a hydrometer at 15°C, then it is converted into API gravity. Those parameters may vary in a wide range: light crudes are gas – like mixtures with 65° API, while heavy crudes are semisolid materials with about 10° API.

- *Sulphur content*: it is the weight percentage of sulphur contained in the Crude Oil, depending on which it is possible to distinguish the oil produced in two classes

  - *Sweet Crude*: < 0.5%
  - *Medium Sour Crude*: 0.5% − 1.5%
  - *Sour Crude*: > 1.5%

- ***Reid Vapour Pressure (RVP)***: it is a measure of the volatility of the Crude Oil produced, and it is defined as the absolute vapour pressure exerted by a substance at 100°F (37.8°C). RVP is an experimental measure that follows the standard test method ASTM D – 323. Typical values of RVP for a Crude Oil are 0.69 and 0.83 bar (10 – 12 psi).

Crude Oil can be stored into different kinds of tanks, by shape and size, throughout its production and distribution. The choice of storage equipment should consider many different aspects, like storage capacities in relation to foreseen production, pressure and temperature of the oil, design issues, etc. On most of production facilities, Crude Oil is sent directly to refineries or tanker terminals through the pipeline, where 10 to 100 tanks of varying size and volume can store up to 50 million Crude Oil barrels. If the offshore platform does not have a direct pipeline system, Crude Oil is stored on board into storage tanks, which are regularly offloaded by shuttle tankers (every two weeks mostly), storing several million barrels. On gravity base platforms, oil is stored into cells around the concrete shafts, while on floating production sites the crude can be stored in onboard tanks or separate storage tanks. The most widely used tanks are vertical cylinders, with capacity ranging between 100 to 1.5 million barrels where pressure and temperature are usually nearly – atmospheric. Moreover, storage tanks are equipped with gauging systems, as floats and level radars, to measure the level reached by the fluid and converting those data to the volume of Crude Oil stored ((Devold, 2013), (Magalhães, et al., 2015), (Society of Petroleum Engineers, 2012)).

1.3.3. Gas Treatment

Gas stream produced by the separation train is composed by methane, ethane and traces of higher hydrocarbons. Moreover, small fractions of water vapour, carbon dioxide, hydrogen sulphide, nitrogen and other impurities are present. Therefore, this raw Gas stream must be treated to meet the specifications required to produce NGL etc (Figure 11). The first step is the removal of hydrocarbons heavier than methane and ethane and acid gases removal. The former is performed cooling the gas below its dew point and routing the resulting gas – liquid mixture into a high pressure three – phase separator to split water, gas and liquid hydrocarbons. The latter uses an amine – based absorption for the removal of carbon dioxide and hydrogen sulphide, through an absorber, amine regenerator and corresponding accessory equipment. Afterwards, the gas stream is dehydrated using an absorption using glycols, pressure swing absorber or membranes. Additional treatments are the removal of mercury through molecular sieves and the rejection of nitrogen traces, by mean of a cryogenic distillation, whose result is a high concentration nitrogen gas stream. The water removed by the condensate removal and dehydration is sent to the produced water treatment unit, while the hydrocarbon condensate removed is addressed to the oil refinery (Devold, 2013).

1.3.4. Water Treatment

The water recovered through the separation train and gas processing contains traces of oil and other impurities. Produced water stream can be reused through reinjection in the reservoir to enhance production, or it is discharged in the sea. In the latter event, impurities and oil must be removed by mean of several processing. In particular, oil concentration must be lowered below 40 mg/L.

The first step is the sand cyclone which removes sand impurities, that are afterwards washed again before being discharged. Then, the water stream is sent to a hydrocyclone to remove oil droplets: a steady vortex forces the separation of the oil phase in the middle from the water on the side. The oil phase recovered is usually recycled to the third separator of the separation train. As last treating step, a degassing drum is used to further remove oil droplets from water: the gas dispersed in the water stream start to rise to the surface, dragging oil droplets with them. Thus, the oil film produced is drained and water can be discharged into the sea (Devold, 2013).
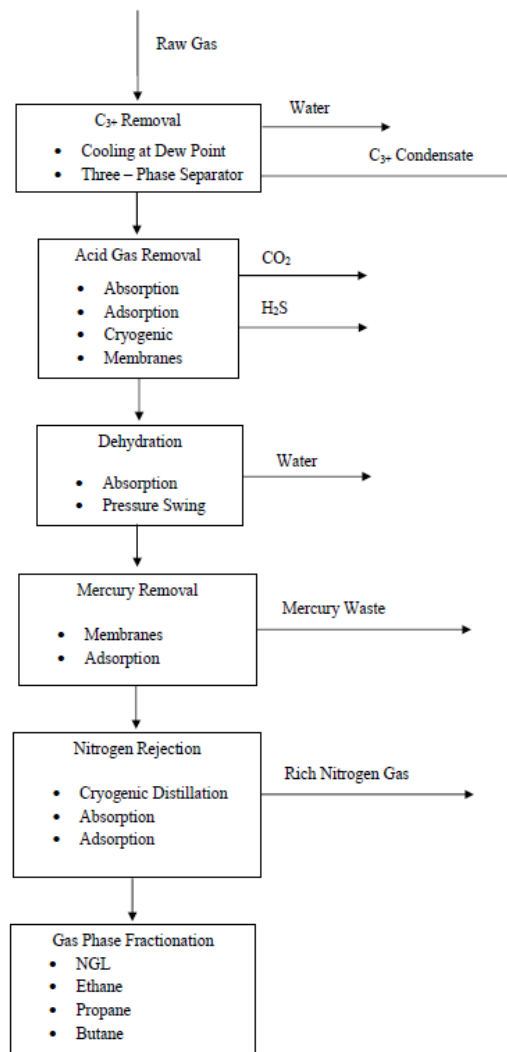
Figure 11. Block Flow Diagram of Gas Treating Process

1.4. Project Delimitation

Over the last 30 years, North Sea has had a key role in the production of Crude Oil and Gas in the European frame, representing an important tool to avoid a complete dependence from Middle – East Crude Oil import. North Sea oil and gas production reached the peak of production during the first four years of the last decade, and a progressive decline of the largest and oldest fields in this area began ever since. Therefore, Oil&Gas Industry in the North Sea has to deal with important challenges in the future to maintain its key role in the European Oil&Gas market. The main efforts in delaying the decline of petroleum reservoirs over the North Sea are being done by the development of new technology for the exploration and drilling of new sizeable oil volumes, and enhancement of the oil and gas production from the existing reservoirs (Institute for the Analysis of Global Security, 2004). Offshore platforms are complex facilities aimed to exploit subsea oil and gas reservoirs, requiring billions of dollars of investments. Those funding are granted based on cost competitive projects, developed case – by – case for specific petroleum wells. Therefore, the optimization of both new and old offshore oil and gas facilities in order to maximize the profit is a crucial aim in Oil&Gas Industry for the near future. Indeed, on the one hand the operating conditions of the units on old facilities may be adjusted in response to changes in the feed stream conditions to be treated, since a single platform can receive fluids from different wells, and those streams can vary their operating conditions, compositions and flow rates. On the other hand, it is profitable designing new offshore platforms aiming to maximize the amount and quality of oil and gas produced, and allowing the adjustment of the operating parameters according to the variations of compositional and operating conditions of the feed.

The main core of the oil and gas production process is the separation of the well fluids into water, gas and oil through a sequence of three – phase and/or two – phase separators, where a progressive reduction of pressure and temperature occurs. That separation aims to produce a gas stream as much free as possible of propane and heavier hydrocarbons, delivering a stable oil. The oil phase is defined stable when it does not vaporize when it is delivered to the tank, and if storage pressure and/or temperature fluctuations occur. The optimization of the oil and gas production thus involves the investigation of the optimal operating pressures and temperatures of the separators to be used on the platforms, in order to meet the requested targets. Moreover, the variation of those optimal conditions depending on the number of separators used and the feed conditions is of particular interest.

1.5. Project Definition

Based on the project delimitation, the main focus of the project is the following:

Develop simulation models of a petroleum reservoir fluids separation process to optimize separators pressure and temperature values, in order to maximize the separation of heavy hydrocarbons (C_{3+}) in the Crude Oil, and light hydrocarbons ($C_{1\&2}$) in the Gas stream produced.

The specific aims set to achieve the above – stated goal of the project include:

i. Choose the suitable thermodynamic model to be used in the simulation models for the characterization of the phase equilibria of multicomponent mixtures, based on the operating conditions of the process.

ii. Develop optimization strategies to evaluate the optimal separation train operating pressures and temperatures, varying the number of separators to be used (1, 2, 3) before the storage tank.

iii. Evaluate the variation of optimal pressures and temperatures of the separators depending on the reservoir fluid features:

- Composition
- Pressure and temperature

Chapter 2

2.1. Oil&Gas Separation Train Optimization

As first step, the optimization of the Oil&Gas separation process requires the definition of the details of the feed stream. Therefore, composition, pressure and temperature of the petroleum reservoir fluid need to be defined.

As explained in section 1.2, different kinds of well fluids exist, depending on their composition which determines the critical properties, and thus their phase behaviour inside the reservoir during extraction. As the aim of these project is investigating the separation of hydrocarbon components between Crude Oil and Gas, reservoir fluid composition(s) is(are) considered on free – water basis. Since no literature has been found regarding the change in composition of the feed of this separation process, and having the aim of making this project widely effective, three different kinds of reservoir fluids compositions are considered:

- Gas Condensate reservoir fluid mixture (Ørke, et al., 1983)

Table 2. Gas Condensate Mixture Composition

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)
N_2	0.87	28.01	0.0012	C_{10}	0.31	133	0.809
CO_2	0.84	44.01	0.0019	C_{11}	0.18	146	0.808
C_1	73.57	16.04	0.0007	C_{12}	0.14	160	0.819
C_2	9.96	30.07	0.0028	C_{13}	0.14	176	0.840
C_3	6.30	44.1	0.002	C_{14}	0.12	187	0.850
iC_4	0.96	58.12	0.0025	C_{15}	0.11	200	0.856
nC_4	1.76	58.12	0.0025	C_{16}	0.07	212	0.862
iC_5	0.60	72.15	0.616	C_{17}	0.09	230	0.844
nC_5	0.63	72.15	0.626	C_{18}	0.05	243	0.844
C_6	0.70	85	0.678	C_{19}	0.04	256	0.853
C_7	0.93	90	0.747	C_{20+}	0.20	336	0.869
C_8	0.95	102	0.776	Total	100	26.48	0.036
C_9	0.48	121	0.801				

- Near Critical reservoir fluid mixture (Rogne & Fjareide, 1983)

Table 3. Near Critical Mixture Composition

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	0.87	28.01	0.0012	C_{10}	0.92	134	0.788
CO_2	0.96	44.01	0.0019	C_{11}	0.86	148	0.791
C_1	57	16.04	0.0007	C_{12}	0.94	162	0.801
C_2	9.28	30.07	0.0028	C_{13}	0.81	177	0.818
C_3	5.82	44.1	0.002	C_{14}	0.67	188	0.830
iC_4	1.19	58.12	0.0025	C_{15}	0.75	201	0.836
nC_4	2.19	58.12	0.0025	C_{16}	0.48	215	0.841
iC_5	0.99	72.15	0.616	C_{17}	0.55	234	0.839
nC_5	1.07	72.15	0.626	C_{18}	0.52	250	0.843
C_6	1.52	85	0.677	C_{19}	0.46	264	0.852
C_7	2.38	92	0.734	C_{20+}	5.53	496	0.913
C_8	2.72	104	0.756	**Total**	100	67.44	0.197
C_9	1.72	119	0.775				

- Black Oil reservoir fluid mixture (Pedersen & Christensen, 2007)

Table 4. Black Oil Mixture Composition

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	0.49	28.01	0.0012	C_{15}	2.03	206	0.842
CO_2	0.31	44.01	0.0019	C_{16}	1.67	222	0.849
C_1	44.01	16.04	0.0007	C_{17}	1.38	237	0.845
C_2	3.84	30.07	0.0028	C_{18}	1.36	251	0.848
C_3	1.12	44.1	0.002	C_{19}	1.19	263	0.858
iC_4	0.61	58.12	0.0025	C_{20}	1.02	275	0.863
nC_4	0.72	58.12	0.0025	C_{21}	0.89	291	0.868
iC_5	0.69	72.15	0.616	C_{22}	0.78	305	0.873
nC_5	0.35	72.15	0.626	C_{23}	0.72	318	0.877
C_6	1.04	86.18	0.655	C_{24}	0.64	331	0.881
C_7	2.87	96	0.738	C_{25}	0.56	345	0.885
C_8	4.08	107	0.765	C_{26}	0.53	359	0.889
C_9	3.51	121	0.781	C_{27}	0.48	374	0.893
C_{10}	3.26	134	0.792	C_{28}	0.46	388	0.897
C_{11}	2.51	147	0.796	C_{29}	0.45	402	0.900
C_{12}	2.24	161	0.810	C_{30+}	9.96	449.1	0.989
C_{13}	2.18	175	0.825	**Total**	100	125.9	0.418
C_{14}	2.07	190	0.836				

Flowing pressure and temperature of the stream processed in a separation train may vary case by case. Kim (Kim, et al., 2014) assumes a feed stream flowing at about 48.4°C and 29 bar, while Magalhães (Magalhães, et al., 2015) assigns reservoir fluid operating conditions of 62°C and 50 bar. Thus, two set of operating conditions are considered to extend the validity of this study:

- 60 bar, 70°C
- 30 bar, 40°C

As stated in section 1.3.1, Oil&Gas separation process involves successive splits of different phases using suitable separators, by mean of density difference and/or progressive reduction of pressure and temperature. Since compositions of the reservoir fluids considered are on free – water basis, the separation process involves successive splits of a vapour/gas phase from a liquid phase (Figure 12). Each single separation stage accomplishes a *flash separation*. A flash is defined as a system where overall composition, pressure and temperature of the feed stream are known and, if a phase split occurs, compositions of the phases at equilibrium can be calculated, together with the relative amount of the two phases.

The well fluids considered are composed by similar non – polar components (hydrocarbons) and slightly – polar components (carbon dioxide) and nitrogen (inert).
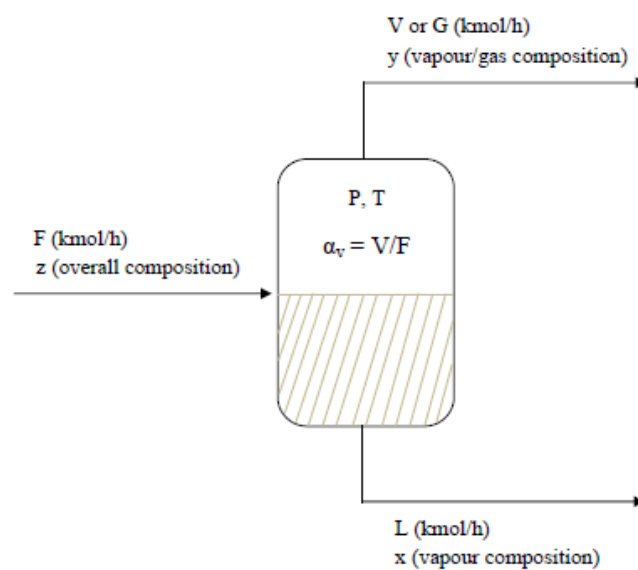
Figure 12. Flash Separation Stage

Under those conditions, when pressure is low (close to atmospheric), the vapour/gas and liquid phase can be considered ideal and the equilibrium is described using the following equations:

- *Raoult's Law*: Vapour – Liquid Equilibria (VLE)

$$y_i = \frac{P_{s_i}(T)}{P} x_i = K_i x_i$$

Equation 3.

$$for \ i = 1, \dots, c$$

where Ps is vapour pressure, K is the distribution coefficients and c is the number of components of the mixture.

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- ***Henry's Law***: Gas – Liquid Equilibria (GLE)

$$y_i = \frac{H_{L_i}(T)}{P} x_i = K_i x_i$$

Equation 4.

$$for \ i = 1, \dots, c$$

where H_L is the Henry's constant K is the distribution coefficients and c is the number of components of the mixture.

Volatile components have distribution coefficients higher than 1, while the K – ratios of heavy components vary between 0 and 1.

As the pressure of the system is far from atmospheric pressure, under the above – stated conditions of compositions reservoir fluid considered, both vapour/gas and liquid phases are not ideal. Those phases are at equilibrium when the following condition is fulfilled:

$$f_{L_i}(P, T, x_1, \dots, x_{c-1}) = f_{V_i}(P, T, y_1, \dots, y_{c-1})$$

$$\downarrow$$

$$P x_i \varphi_{L_i}(P, T, x_1, \dots, x_{c-1}) = P y_i \varphi_{V_i}(P, T, y_1, \dots, y_{c-1})$$

$$\downarrow$$

$$y_i = \frac{\varphi_{L_i}(P, T, x_1, \dots, x_{c-1})}{\varphi_{V_i}(P, T, y_1, \dots, y_{c-1})} x_i = K_i x_i$$

$$for \ i = 1, \dots, c$$

where f_L, φ_L and f_V, φ_V are the ***fugacity*** and ***fugacity coefficients*** of liquid and vapour phase respectively, while c is the number of components of the mixture. Therefore, the phase equilibria calculations require to apply models for the determination of vapour and liquid phases fugacity coefficients. Soave – Redlich Kwong and Peng – Robinson Cubic Equations of State (CEOS) are the most used thermodynamic models to describe non – ideal vapour/gas and liquid phases.

Since the operating conditions of the three reservoir fluids chosen for this study are far from the atmospheric, the Peng – Robinson CEOS is chosen for the characterization of the phase equilibria of multicomponent mixture:

$$P = \frac{RT}{\underline{V_m} - b_m} - \frac{\theta_{PRm}(T)}{\underline{V_m}\left(\underline{V_m} + b_m\right) + b_m\left(\underline{V_m} - b_m\right)}$$

Equation 5.

$$Z^3 + (B_m - 1)Z^2 + (A_m - 3B_m^2 - 2B_m)Z + (B_m^3 + B_m^2 - A_m B_m) = 0$$

Equation 6.

$$\ln \varphi_i = \frac{B_i}{B_m}(Z - 1) - \ln(Z - B_m) - \frac{A_m}{2\sqrt{2}B_m}\left(\frac{2\sum_{j=1}^{c} z_j A_{ij}}{A_m} - \frac{B_i}{B_m}\right)\ln\left[\frac{Z + \left(1 + \sqrt{2}\right)B_m}{Z + \left(1 - \sqrt{2}\right)B_m}\right]$$

Equation 7.

where $\underline{V}_m$ and b_m are molar volume (m³/mol) and co – volume (m³/mol) of the mixture respectively, Z is the compressibility factor, R is the gas constant (8.314 m³ Pa K⁻¹ mol⁻¹). A, B, A_m, B_m, θ_{PRm} are parameters of the equation, which are function of pressure and temperature of the system, critical pressure, critical temperature and acentric factor of the components, composition of the mixture.

As stated in section 1.2, since petroleum reservoir fluids are mixtures composed by thousand components of different kinds, the characterization of those multicomponent mixtures is carried out identifying cuts of components, whose boiling points are included in specific ranges, measuring at the same time the corresponding molecular weights and density. These experimental data can be used to determine critical pressure, critical temperature and acentric factor of the hydrocarbon cuts, using specific property correlations (Pedersen & Christensen, 2007):

$$T_c = c_1\rho + c_2 \ln M_w + c_3 M_w + \frac{c_4}{M_w}$$

Equation 8.

$$\ln P_c = d_1 + d_2\rho^{d_5} + \frac{d_3}{M_w} + \frac{d_4}{M_w^2}$$

Equation 9.

$$m = e_1 + e_2 M_w + e_3\rho + e_2 M_w^2$$

Equation 10.

$$m = 0.37464 + 1.54226\omega + 0.26992\omega^2$$

Equation 11.

where ρ is density in g/cm³ and M_w is molecular weight in g/mol, P_c, T_c and ω are critical pressure (atm), critical temperature (K) and acentric factor (dimensionless). Correlation coefficients are shown in Table 5.

Table 5. Property Correlation Coefficients (Pedersen & Christensen, 2007)

Subindex/Coefficient	1	2	3	4	5
c	$7.34043 \cdot 10$	$9.73562 \cdot 10$	$6.18744 \cdot 10^{-1}$	$-2.05932 \cdot 10^3$	-
d	$7.28462 \cdot 10^{-2}$	2.18811	$1.6391 \cdot 10^2$	$-4.04323 \cdot 10^3$	0.25
e	$3.73765 \cdot 10^{-1}$	$5.49269 \cdot 10^{-3}$	$1.17934 \cdot 10^{-2}$	$-4.93049 \cdot 10^{-6}$	-

Using the data of density and molecular weights listed in Table 2, Table 3 and Table 4, it is possible to determine the critical properties and the acentric factors of three reservoir fluids chosen for this study (Table 6, Table 7, Table 8).

Table 6. Properties of Gas Condensate Mixture's Components

Component	P_c (bar)	T_c (K)	ω	Component	P_c (bar)	T_c (K)	ω
N_2	33.9	126.2	0.04	C_9	25.4	583.5	0.366
CO_2	73.8	304.2	0.225	C_{10}	23.7	602.3	0.395
C_1	46	190.6	0.008	C_{11}	22.1	620.7	0.426
C_2	48.8	305.4	0.098	C_{12}	20.8	640.3	0.457
C_3	42.5	369.8	0.152	C_{13}	19.7	662.2	0.491
iC_4	36.5	408.1	0.176	C_{14}	19.1	676.4	0.514
nC_4	38	425.2	0.193	C_{15}	18.3	692.1	0.539
iC_5	33.8	460.4	0.227	C_{16}	17.8	706.2	0.562
nC_5	33.7	469.5	0.251	C_{17}	16.8	724.7	0.594
C_6	31.2	510.7	0.271	C_{18}	16.3	738.6	0.616
C_7	31.3	525.7	0.286	C_{19}	15.9	752.8	0.637
C_8	28.7	550.2	0.318	C_{20+}	14.2	831.9	0.745

Table 7. Properties of Near Critical Mixture's Components

Component	P_c (bar)	T_c (K)	ω	Component	P_c (bar)	T_c (K)	ω
N_2	33.9	126.2	0.04	C_9	25.3	578.5	0.361
CO_2	73.8	304.2	0.225	C_{10}	23.2	602.2	0.398
C_1	46	190.6	0.008	C_{11}	21.6	622.2	0.43
C_2	48.8	305.4	0.098	C_{12}	20.4	641.6	0.461
C_3	42.5	369.8	0.152	C_{13}	19.4	661.9	0.493
iC_4	36.5	408.1	0.176	C_{14}	18.8	676.1	0.516
nC_4	38	425.2	0.193	C_{15}	18.1	691.8	0.541
iC_5	33.8	460.4	0.227	C_{16}	17.4	708.1	0.657
nC_5	33.7	469.5	0.251	C_{17}	16.6	728.7	0.601
C_6	31.2	510.6	0.271	C_{18}	16	745.9	0.627
C_7	30.4	528.6	0.291	C_{19}	15.7	760.9	0.649
C_8	27.9	552.2	0.323	C_{20+}	12.7	974	0.858

Table 8. Properties of Black Oil Mixture's Components

Component	P_c (bar)	T_c (K)	ω	Component	P_c (bar)	T_c (K)	ω
N_2	33.9	126.2	0.04	C_{14}	18.7	678.9	0.52
CO_2	73.8	304.2	0.225	C_{15}	17.9	698	0.551
C_1	46	190.6	0.008	C_{16}	17.2	716.4	0.58
C_2	48.8	305.4	0.098	C_{17}	16.5	732.3	0.606
C_3	42.5	369.8	0.152	C_{18}	16	747.3	0.629
iC_4	36.5	408.1	0.176	C_{19}	15.7	760.4	0.648
nC_4	38	425.2	0.193	C_{20}	15.4	772.8	0.666
iC_5	33.8	460.4	0.227	C_{21}	15.1	789	0.689
nC_5	33.7	469.5	0.251	C_{22}	14.8	802.9	0.707
C_6	30.3	511.4	0.274	C_{23}	14.6	815.6	0.724
C_7	29.5	536.5	0.302	C_{24}	14.4	828.1	0.739
C_8	27.4	558	0.331	C_{25}	14.1	841.4	0.755
C_9	25.1	582.1	0.366	C_{26}	13.9	854.4	0.769
C_{10}	23.3	602.5	0.398	C_{27}	13.8	868.2	0.784
C_{11}	21.8	621.2	0.428	C_{28}	13.6	880.9	0.796
C_{12}	20.6	641	0.459	C_{29}	13.5	893.5	0.807
C_{13}	19.6	659.9	0.489	C_{30+}	13.6	940.5	0.839

As stated in section 1.5, the target of the separation train optimization is to maximize at the same time the recovery of C_{3+} hydrocarbons in Crude Oil, and the recovery $C_{1/2}$ hydrocarbons in the Gas produced.
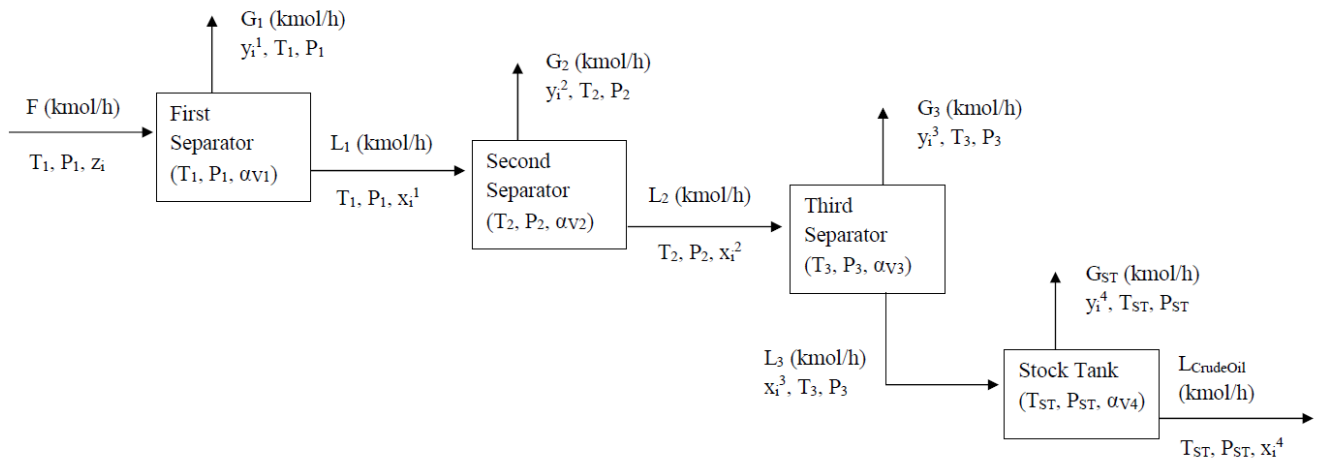
Figure 13. Block Flow Diagram of Separation train

Looking at the scheme in Figure 13, the above – mentioned recoveries are defined as follows:

$$R_{oil} = \frac{L_{Crude\ Oil} \sum_{i=C_3}^{C_n} x_i^4}{F \sum_{k=C_3}^{C_n} z_k} = \left(1 - \alpha_{V_1}\right)\left(1 - \alpha_{V_2}\right)\left(1 - \alpha_{V_3}\right)\left(1 - \alpha_{V_4}\right)\frac{\sum_{i=C_3}^{C_n} x_i^4}{\sum_{k=C_3}^{C_n} z_k}$$

Equation 12.

$$R_{gas} = \frac{G_1 \sum_{i=C_1}^{C_2} y_i^1}{F \sum_{k=C_1}^{C_2} z_k} + \frac{G_2 \sum_{i=C_1}^{C_2} y_i^2}{F \sum_{k=C_1}^{C_2} z_k} + \frac{G_3 \sum_{i=C_1}^{C_2} y_i^3}{F \sum_{k=C_1}^{C_2} z_k} =$$

$$= \frac{\alpha_{V_1} \sum_{i=C_1}^{C_2} y_i^1}{\sum_{k=C_1}^{C_2} z_k} + \frac{\alpha_{V_2}(1 - \alpha_{V_1}) \sum_{i=C_1}^{C_2} y_i^2}{\sum_{k=C_1}^{C_2} z_k} + \frac{\alpha_{V_3}(1 - \alpha_{V_1})(1 - \alpha_{V_2}) \sum_{i=C_1}^{C_2} y_i^3}{\sum_{k=C_1}^{C_2} z_k}$$

Equation 13.

Equation 12 and Equation 13 can be generalized, deriving the following equations:

$$R_{oil} = \left(\prod_{j=1}^{N+1}(1 - \alpha_{V_j}) \right) \frac{\sum_{i=C_3}^{C_n} x_i^{N+1}}{\sum_{k=C_3}^{C_n} z_k}$$

Equation 14.

$$R_{gas} = \frac{\sum_{i=1}^{N} \alpha_{V_i} \prod_{j=1}^{i}\left(1 - \alpha_{V_{1-j}}\right) \sum_{k=C_1}^{C_2} y_k^i}{\sum_{k=C_1}^{C_2} z_k}$$

Equation 15.

where R_{oil} and R_{gas} are the recoveries in Crude Oil and Gas respectively (values included between 0 and 1), N is the number of separators and n is the higher carbon number fraction in the reservoir fluid. Since the optimization aims to maximize both recoveries at the same time, those variables need to be combined into one objective function. Many kinds of combinations are possible: linear, exponential, etc. Since no valid reason was found to promote a specific kind of function, a linear combination of the recovery in Crude Oil and Gas is used (Equation 16). The constants to be used in the objective function are determined using the prices of Oil and Gas (Equation 17 and Equation 18). Indeed, this way the importance of the two recoveries is weighted during the optimization. The prices considered for Crude Oil and Gas in this optimization are 326 US\$/m^3 and 0.388 US\$/m^3 respectively (U.S. Department of Energy, 2008). Since both the constants of the linear combination and the recoveries values are included between 0 and 1, the objective function varies in the same range.

$$\Phi\left(R_{oil}, R_{gas}\right) = c_1 R_{oil} + c_2 R_{gas}$$

Equation 16.

$$c_1 = \frac{Price_{oil}}{Price_{oil} + Price_{gas}} = 0.99$$

Equation 17.

$$c_2 = 1 - c_1 = 0.01$$

Equation 18.

The optimization of the separation process is constrained by the need to produce a stable Crude Oil. As explained in section 1.3.1, the oil phase produced from the last separator must stay in liquid phase when pressure and/or temperature are reduced to stock conditions and the product is delivered to the storage tank. Moreover, no vaporization must occur if storing conditions undergo to slight variations. This stability constraint is thus expressed as bubble point pressure of the Crude Oil, at the storage temperature, must be equal or lower than a certain percentage of the pressure in the storage tank:

$$P_B(x_1^{N+1}, \dots, x_c^{N+1}, T_{stock}) \leq \theta P_{stock}$$

Equation 19.

$$with \ 0 < \theta < 1$$

where N is the number of separators and c is the number of components in the mixture. The closer θ to 0, the stricter the stability constraint.

The storing conditions can vary case by case. Both Kim (Kim, et al., 2014) and Kylling (Kylling, 2009) suggest a nearly – atmospheric storage pressure and temperature. Therefore, stock conditions are fixed at 1.2 bar and 15°C.

2.2. Optimization Strategies

The development of models for the simulation of the Oil&Gas separation process can be performed through different tools.

Chemical process simulators are softwares widely used for the design and/or evaluation of whole or sections of already existing or not yet existing chemical plants. One of the most important process simulators is Aspen HYSYS, by AspenTech. This software allows wide choices regarding the built – in thermodynamic models and routines, in order to simulate different kinds of processes and optimize their operating conditions. Moreover, it results to be a very well tested tool in process engineering, thus their results are considered very reliable.

Alternatively, MATLAB can be used for the development of codes that can simulate the separation process by mean of successive flash calculations. On the one hand, self – built programs allow to have great flexibility regarding the implementation of flash calculations methods and optimization strategies. On the other hand, MATLAB codes development may be time consuming and it may not return reliable results.

In this project, Aspen HYSYS V9 process simulator (full licence provided by Aalborg University) is used because of the reliability of the software and its wide use in Oil&Gas companies.

Firstly, three component lists have to be created to model the three reservoir fluids chosen for this study. Nitrogen, carbon dioxide and hydrocarbons from C_1 to nC_5 are already present into the HYSYS dataset. The carbon fraction cuts of the selected fluids are modelled as hypothetical components, specifying "hydrocarbon" as components class. The data shown in tables from 2 to 4 and from 6 to 8 are given as input for characterization of the components.

Figure 14. Aspen HYSYS Hypothetical Components Input Screen

Secondly, different flowsheets are developed to study the separation process, using the three different feed streams selected and varying the number of separators. As stated in section 1.3.1, unless otherwise required, pressure and temperature of the first separator of a separation train are the flowing conditions of the feed stream. In this study, the separation process is simulated as a sequence of at first 2, and then 3 separators, assuming that the reservoir fluids are flashed at their flowing pressure and temperature in the first separator. Additionally, the Crude Oil tank is simulated by mean of a separator. Since only vapour/gas – liquid phase splits are considered in this project, two – phase separators are modelled in the simulator. Figure 15 shows the screen for the configuration of a two – phase separator in the HYSYS simulation environment.
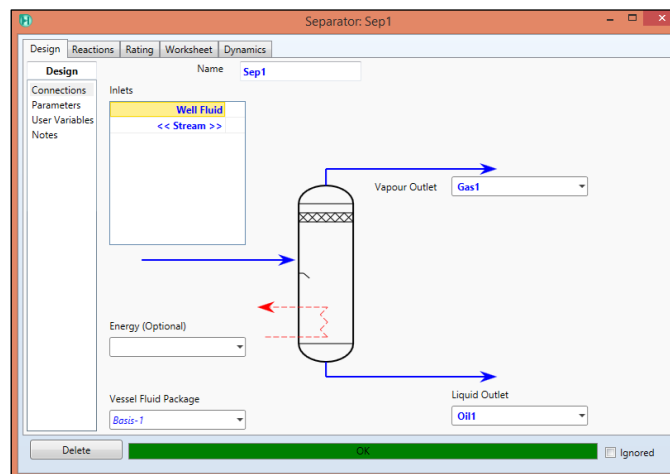
Figure 15. Two – Phase Separator Setup Screen

The optimization of the operating pressure and temperature is carried out by mean of the HYSYS Original Optimizer. The first step (Figure 16) is the selection of the variables to be adjusted to meet the optimization aims needed, the range in which those variables should be varied (lower and upper bound). In this project, the variables to be optimized are the pressures and temperatures of the streams entering the separators. Afterwards, the parameters needed for the optimization must be specified in the spreadsheet (Figure 17).
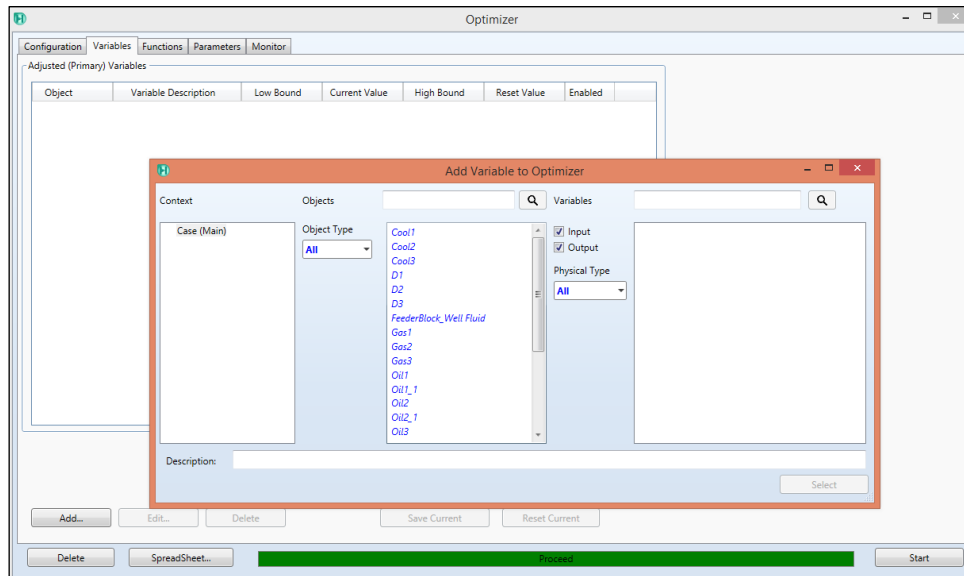
Figure 16. Variable Tab of the Original Optimizer

Therefore, vapour fractions, molar compositions, recoveries in Crude Oil and Gas, objective function and bubble point Crude Oil at stock temperature are stated and connected to the main flowsheet.

Figure 17. Original Optimizer Spreadsheet

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Afterwards, the cell containing the objective function is specified in the Functions tab (Figure 18) and the option "maximize" is chosen. Then, the cell containing the parameters used as constraints are indicated and the relation between them (lower, equal or higher) is stated. In this project, the constraint is the bubble point pressure of the Crude Oil stored in the tank and it must be lower than a fixed percentage of the tank pressure. Literature suggests that, at the storage temperature (15°C), the boiling pressure of the stored Crude Oil should be between 0.68 bar and 0.88 bar (Piping Engineering, 2015). Therefore, in this project the constraint for the optimization is that the bubble point pressure of the Crude Oil is equal or lower than 0.85 bar (70% of the tank pressure).
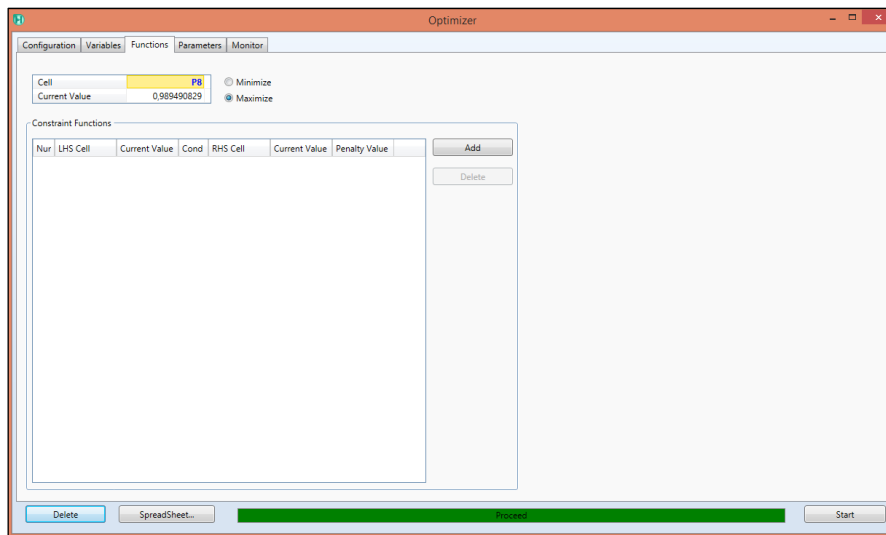
Figure 18. Functions Tab of the Original Optimizer

In the end, the Parameter tab allows to choose the details of the optimization (Figure 19). The main detail is the optimization method to be used. HYSYS allows several choices according to some aspects of the optimization to be performed (linear or non – linear objective function, equality and/or inequality constraints):
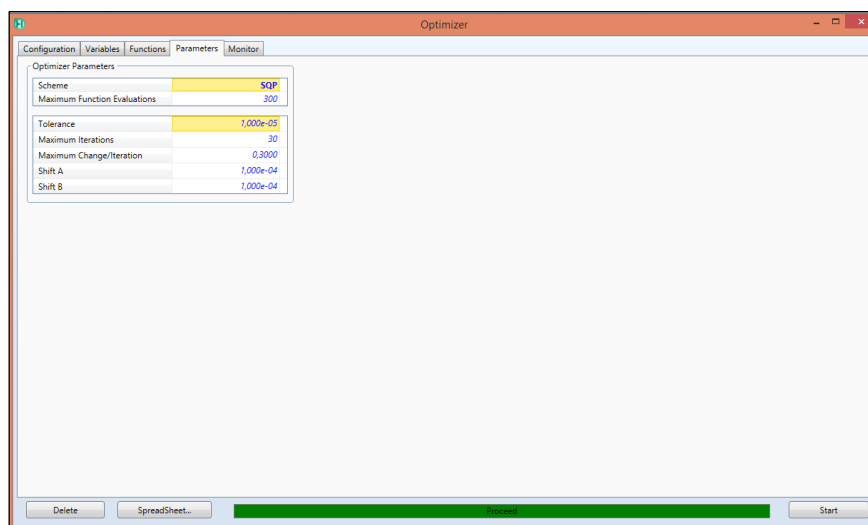
Figure 19. Parameter Tab of the Original Optimizer

- *BOX Method*: it is a sequential search technique to solve problems with non – linear objective functions, which are subjected to non – linear inequality constraints, without using derivatives. It is not very efficient and requires many iterations to converge to the global solution, but if applicable it results to be very robust. This method evaluates the function in different points around a suitable starting point and within a suitable range. Once having found a point that optimize the objective function and satisfy the constraints at same time, an extrapolation is used to find the next point. If the new point satisfies the requirements, one more extrapolation is performed, otherwise new points are generated around the last suitable point and the search is repeated.

- *Sequential Quadratic Programming (SQP)*: it is a optimization method, which that generates a Lagrangian function using the objective function, the constraints and the Lagrange's multipliers. Then, an appropriate search direction is found by solving a quadratic subproblem, where a quadratic linear approximation of the Lagrangian function and its second derivative matrix are used. SQP is considered the most efficient optimization method for non – linear objective functions, with linear and/or non – linear equality and/or inequality constraints. However, it does not handle large number of variables and it is mostly used for local optimizations

- *Mixed Method*: this method uses at first the BOX method until converge to the global constrained optimum of the objective function. Then the SQP method is used to locate the final solution. This way the global optimization of the BOX method is combined with the efficiency of the SQP method.

- *Fletcher – Reeves Method*: it is an optimization method which is effective for general problems with no constraints. Given a starting point, the program evaluates the gradient of the function with respect to the primary variables and the search direction as conjugated of the gradient. A one – dimensional search is performed until funding the local optimum.

- *Quasi – Newton Method*: this method does not handle constraints and determines a search direction by approximating the inverse of the Hessian matrix of the objective function.

The descriptions of the optimization methods suggest that a suitable choice to find the global optimal values for pressures and temperatures of the Oil&Gas separation process is the BOX method.

3.1. Optimization Results and Discussion

As explained in section 2.1, the flowsheets developed for the simulation of the Oil&Gas separation in the HYSYS simulator environment include sequences of at first 2, and then 3 two – phase separators. One additional separator is used to simulate the Crude Oil storage tank.

3.1.1. Case Study 1: Sequence of Two Separators

Figure 20 shows the flowsheet developed to evaluate the optimal recovery conditions of a separation train composed by a sequence of two separators. In the simulation, the cooler model is used to lower both pressure and temperature. In a real case, a relief valve is used to reduce the pressure, and afterwards a heat exchanger lowers the temperature.

Figure 20. Flowsheet of the Separation Train with 2 Separators

Since pressure and temperature of the first separator are fixed, it is possible to evaluate the trends of the objective function and the recovery of C_{3+} hydrocarbons in Crude Oil, varying the operating conditions of the second separator.

Figure 21 shows the above – mentioned trends when the feed stream is flowing at 60 bar and 70°C. It can be noticed that small changes occur in the objective function and the recovery of C_{3+} with pressure and temperature, until reaching nearly – atmospheric pressure. Indeed, those variables have a decrease of 43%, 27% and 6.4% for Gas Condensate, Near Critical and Black Oil respectively (data available in the Appendix, page 53 and 56).
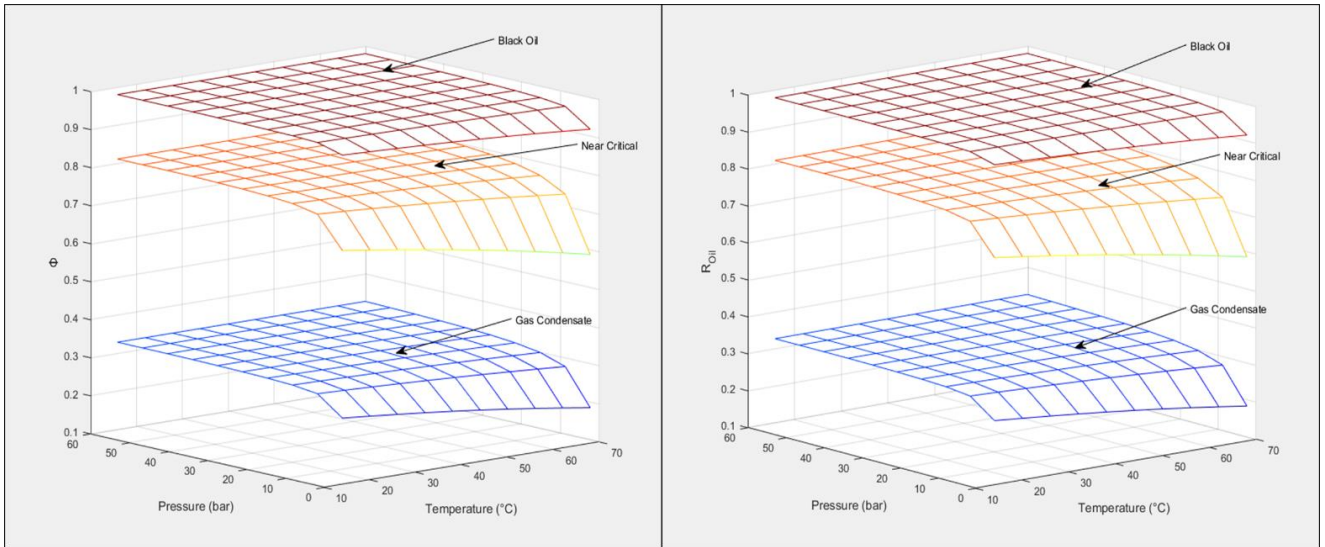
Figure 21. Feed Conditions: 60 bar, 70°C;
Left Panel: Trend of the Objective Function (Φ); Right Panel: Trend of the Recovery of C₃₊ in Crude Oil (Roil)
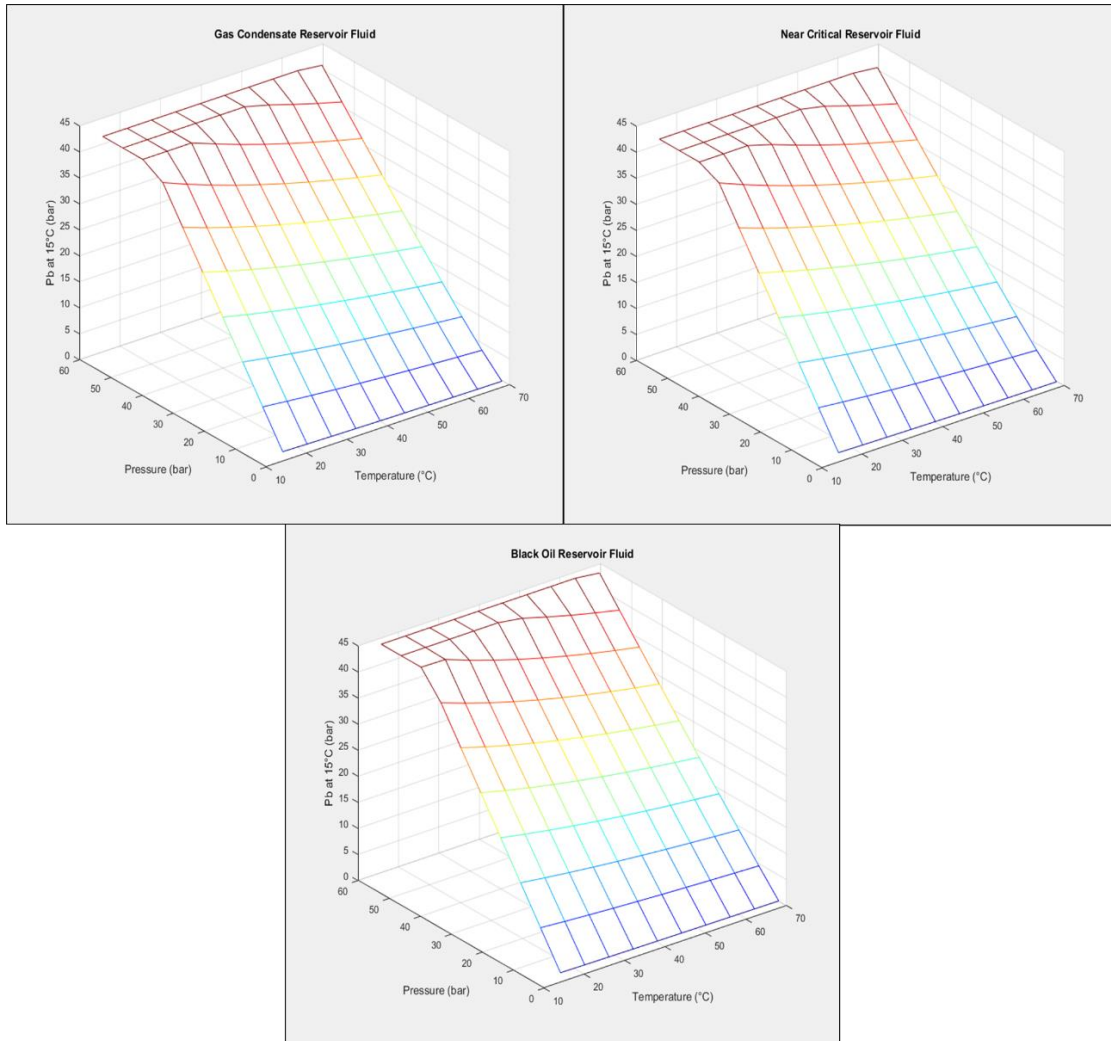
Figure 22. Bubble Point Pressure of Crude Oil produced (Pb, bar) as function Pressure (bar) and Temperature (°C) of the Second Separator

Figure 22 shows the trend of the bubble point pressure at stock temperature (15°C) of the Crude Oil in the stock tank, as function of operating conditions of the second separator. It is possible to notice that, as the second stage pressure decreases, the bubble point pressure decreases. Thus, it is reasonable to expect that the stability constraint is met when the pressure of the second separator is very close to the tank pressure (1.2 bar). The data regarding the trend of Crude Oil bubble point pressure are shown in the Appendix (page 59).

An optimization is performed to maximize the objective function, respecting the volatility constraint. As it is shown in Table 9, in the case of two separators before the storage tank, a stable Crude Oil can be produced reducing the pressure of the liquid from the first separator up to stock pressure. Moreover, temperature needs to be almost halved in the case of Gas Condensate and Near Critical reservoir fluids, while a reduction of about 7°C is required for Black Oil mixture.

Table 9. Results of the Optimization for a Sequence of Two Separators

Parameters \ Fluids	Gas Condensate	Near Critical	Black Oil
P_1 (bar)	60	60	60
T_1 (°C)	70	70	70
α_{V1} (Vapour Fraction)	0.93	0.66	0.36
P_2 (bar)	1.2	1.2	1.2
T_2 (°C)	39	39	53
α_{V2} (Vapour Fraction)	0.46	0.36	0.24
Φ	0.24	0.665	0.937
Roil	0.23	0.664	0.937
Rgas	0.999	0.999	0.999
Pb Crude Oil at 15°C (bar)	0.81	0.84	0.85

In all three cases, the recovery of C_1 and C_2 hydrocarbons in Gas phase is nearly complete (99%), while the recovery of C_3 and heavier hydrocarbons in Crude Oil is very low for the Gas Condensate (23%), while it is higher for Near Critical mixture (66.4%) and Black Oil fluid (93.7%). The compositions of the Crude Oil streams produced are shown in the Appendix (page 47).

Considering 30 bar and 40°C as pressure and temperature of the feed streams and the first separator, Figure 23 shows the trends of the objective function and the recovery of C_{3+} hydrocarbons in Crude Oil, varying pressure and temperature of the second separator. As it can be seen, those two variables change slightly varying the operating conditions of the second separator, until reaching nearly – atmospheric pressures. Indeed, decline of the 27%, 16% and 2% occur for Gas Condensate, Near Critical and Black Oil fluids respectively when approaching stock pressure (data available in the Appendix, page 53 and 56).

Figure 24 shows the trend of the bubble point pressure at stock temperature (15°C) of the stream delivered to the storage tank, as function of operating conditions of the second separator. As in the previous case, the bubble point pressure decreases as the second separator pressure decreases. Therefore, it is reasonable to expect that the stability constraint is met when the pressure of the second separator is very close to the tank pressure (1.2 bar).
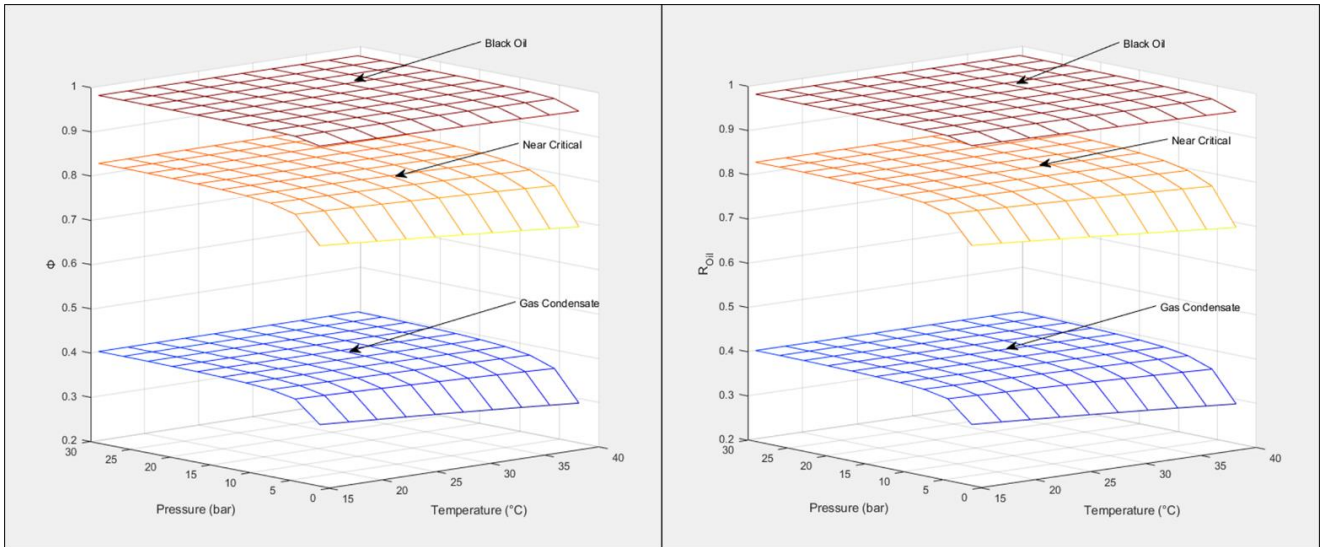
Figure 23. Feed Conditions: 30 bar, 40°C;
Left Panel: Trend of the Objective Function (Φ); Right Panel: Trend of the Recovery of C₃₊ in Crude Oil (Roil)
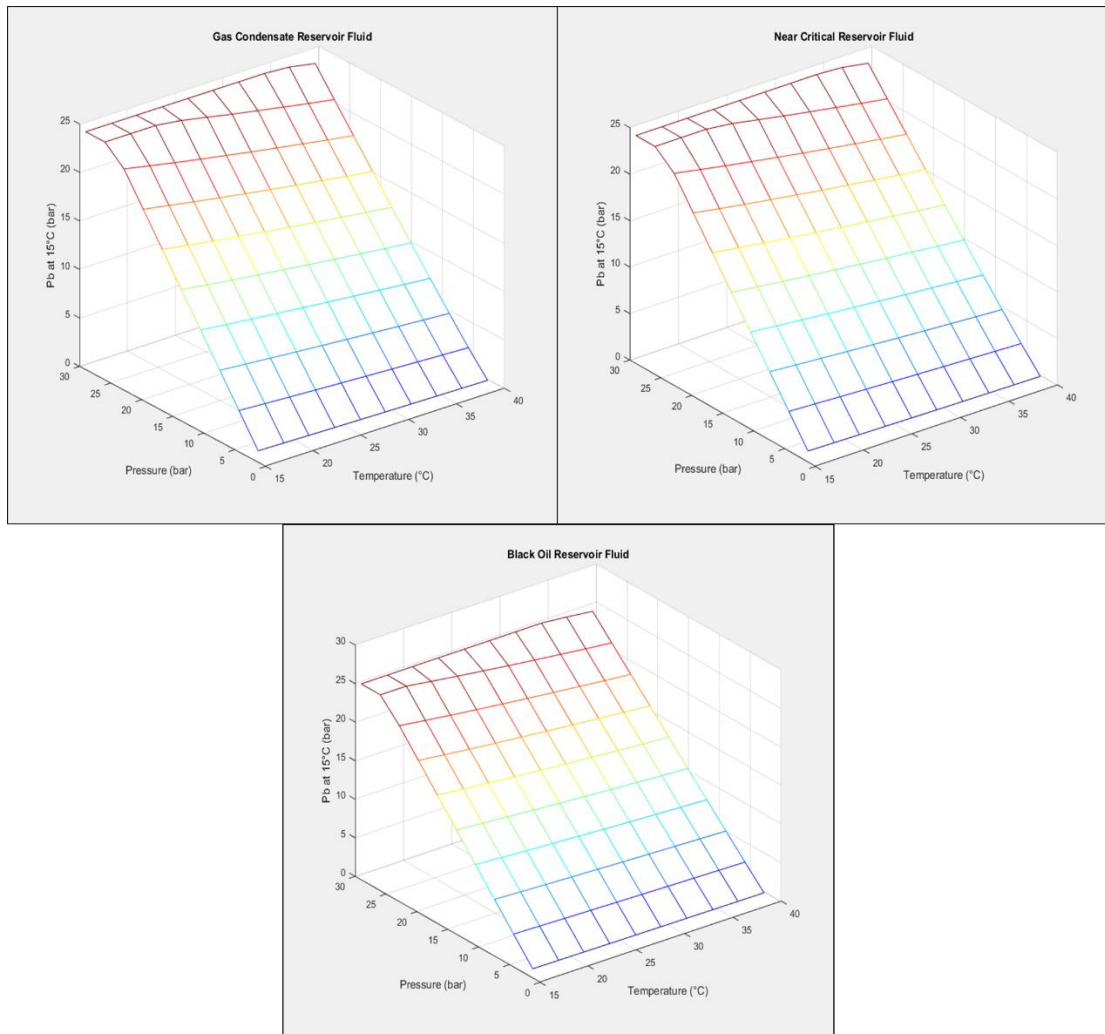
Figure 24. Bubble Point Pressure of Crude Oil produced (Pb, bar) as function Pressure (bar) and Temperature (°C) of the Second Separator

The results of an optimization to maximize the objective function and respecting the volatility constraint are shown in Table 10. The optimal pressure of the second separator to produce a stable Crude Oil is equal to the one of the stock tank for all the three compositions considered. Moreover, the optimal temperature needs to be lowered up to 34°C for the Gas Condensate mixture and to 36°C for the Near Critical fluid. It is possible to notice that the optimal conditions for the Black Oil bring to a stable Crude Oil, but its bubble point exceeds the fixed constraint of about the 11.5%.

Table 10. Results of the Optimization for a Sequence of Two Separators

Fluids / Parameters	Gas Condensate	Near Critical	Black Oil
P_1 (bar)	30	30	30
T_1 (°C)	40	40	40
α_{V1} (Vapour Fraction)	0.93	0.69	0.42
P_2 (bar)	1.2	1.2	1.2
T_2 (°C)	34	36	36
α_{V2} (Vapour Fraction)	0.37	0.28	0.15
Φ	0.309	0.703	0.96
Roil	0.308	0.702	0.96
Rgas	0.999	0.999	0.997
Pb Crude Oil at 15°C (bar)	0.84	0.84	0.96

In all three cases, the recovery of C_1 and C_2 hydrocarbons in Gas phase is nearly complete (99%), while the recovery of C_3 and heavier hydrocarbons in Crude Oil is low for the Gas Condensate fluid (30.8%), it is middle for Near Critical mixture (70.2%) and quite high for the Black Oil fluid (96%). The composition of Crude Oil produced at the optimal conditions are shown in the Appendix (page 47).

The results of the above – shown optimizations suggest that a configuration where a separation train is composed by a sequence of two separators (the first one operating at the flowing pressure and temperature) can deliver a stable Crude Oil. However, the recovery of C_{3+} hydrocarbons is not satisfactory. Indeed, when processing Gas Condensate and Near Critical mixtures, the amount of initial C_3 and heavier hydrocarbons in the Gas produced ranges between the 28% and the 77%. This means that, before being addressed to treating processes, a consistent amount of condensate needs to be removed from the gas, to achieve an effective recovery of C_{3+} hydrocarbons.

3.1.2. Case Study 2: Sequence of Three Separators

Figure 25 shows the flowsheet developed to evaluate the optimal recovery conditions of a separation train composed by a sequence of three separators. As in the previous case study, in the simulation the cooler model is used to lower both pressure and temperature. In a real case, relief valves are used to reduce the pressure, and afterwards heat exchangers lower the temperature.
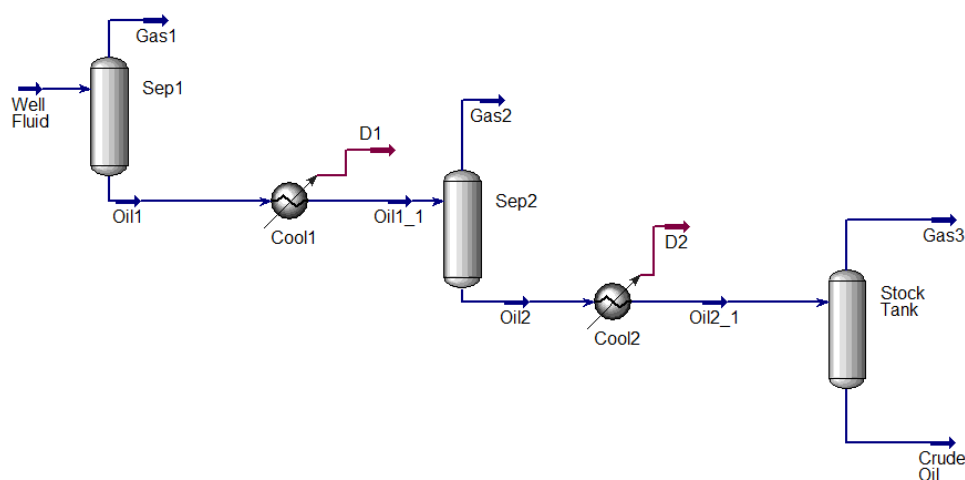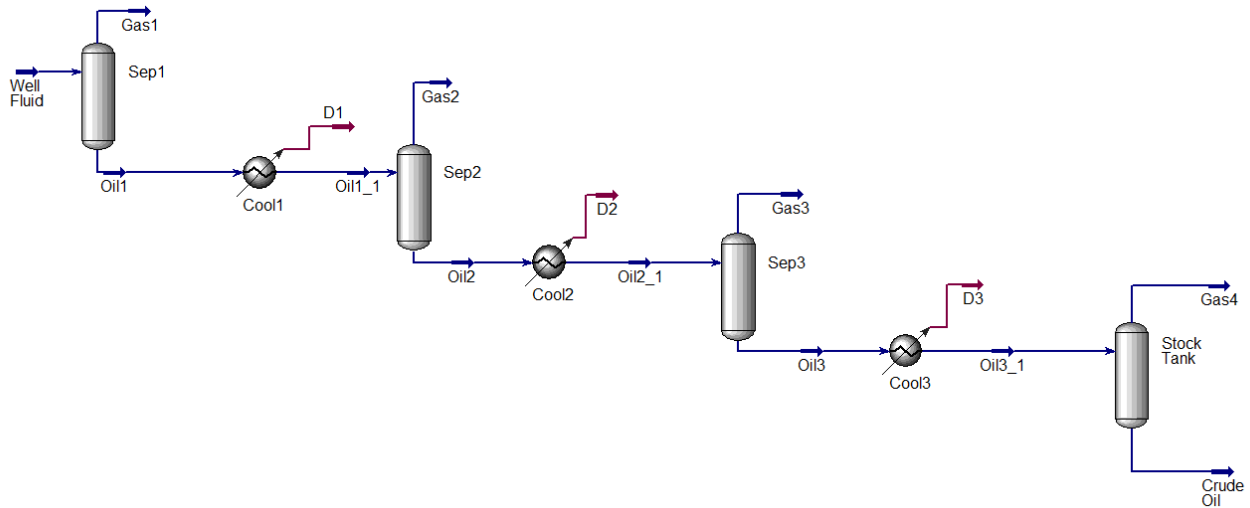
Figure 25. Flowsheet of the Separation Train with 3 Separators

Considering 60 bar and 70°C as feed stream flowing conditions and operating pressure and temperature of the first separator,

Table 11 shows the results of the optimization of the operating conditions of the second and third separator. For the Gas Condensate, Near Critical and Black Oil fluids, a large pressure drop is required between the first and the second separator (from 60 bar to 8 bar). Moreover, for Gas Condensate and Near Critical mixtures, temperatures are almost equally distanced between separators. For the Black Oil reservoir fluid, temperatures of second and third separator are close to the first separator temperature condition.

Table 11. Results of the Optimization for a Sequence of Three Separators

Fluids / Parameters	Gas Condensate	Near Critical	Black Oil
P_1 (bar)	60	60	60
T_1 (°C)	70	70	70
α_{V1} (Vapour Fraction)	0.93	0.66	0.36
P_2 (bar)	8	8	8
T_2 (°C)	52	53	53
α_{V2} (Vapour Fraction)	0.3	0.24	0.18
P_3 (bar)	1.2	1.2	1.2
T_3 (°C)	35	38	51
α_{V3} (Vapour Fraction)	0.13	0.1	0.23
Φ	0.275	0.72	0.97
Roil	0.274	0.72	0.97
Rgas	0.999	0.998	0.998
Pb Crude Oil at 15°C (bar)	0.78	0.77	0.79

It is possible to notice that recovery on C_{3+} hydrocarbons in Crude Oil increases when using a third separator for all the three compositions considered. Indeed, a 16%, 8% and 3% increase is achieved for Gas Condensate, Near Critical and Black Oil fluids respectively. On the other hand, the recovery of C_1 and C_2 in the Gas phase is almost complete (99%). The compositions of the Crude Oil streams produced are shown in the Appendix (page 47).

Changing the operating conditions of the first separator to 30 bar and 40°C and performing a new optimization, the optimal pressures and temperatures for second and third separator results to be as shown in Table 12. The data show that, for all the three well fluids, pressure and temperature in the second separators are lower than 10 bar. Then, in the third separator, pressure is lowered directly to stock pressure. Moreover, temperatures of the second and third separators result to be close to feed condition.

Looking at the recovery of C_{3+} components in Crude Oil, a very slight increase is achieved using an additional separator, compared to the case of two separators in sequence. Indeed, recovery in Crude Oil is low for the Gas Condensate fluid (34%), it assumes a middle – high value for the Near Critical mixture (75%) and a high value for the Black Oil fluid (97.9%). On the other hand, the recovery of C_1 and C_2 in Gas phase in practically complete (99%). Moreover, for the Black Oil fluid the optimal operating pressures and temperatures of the separators do not allow to meet the stability constraint fixed, which is exceeded of about the 8.6%.

Table 12. Results of the Optimization for a Sequence of Three Separators

Parameters \ Fluids	Gas Condensate	Near Critical	Black Oil
P_1 (bar)	30	30	30
T_1 (°C)	40	40	40
α_{V1} (Vapour Fraction)	0.93	0.69	0.42
P_2 (bar)	7	4	5
T_2 (°C)	35	37	37
α_{V2} (Vapour Fraction)	0.18	0.184	0.11
P_3 (bar)	1.2	1.2	1.2
T_3 (°C)	33	36	36
α_{V3} (Vapour Fraction)	0.35	0.27	0.15
Φ	0.34	0.75	0.979
Roil	0.34	0.75	0.979
Rgas	0.999	0.999	0.999
Pb Crude Oil at 15°C (bar)	0.79	0.77	0.93

The results of the optimization highlight that a separation train configuration composed by three separators before the tank can deliver a stable Crude Oil, but a satisfactory separation of C_{3+} hydrocarbons in liquid phase cannot be achieved. Indeed, the gas phase will contain between the 2.1% and 64% of propane and heavier hydrocarbons, which must be recovered as condensate while treating the Gas stream to enable a satisfactory Crude Oil production. The compositions of the Crude Oil streams produced are shown in the Appendix (page 47).

Conclusions

The results of the Oil&Gas separation train optimization shown in the previous chapter allow the draw some conclusions.

Assuming 1.2 bar and 15°C as storage conditions, the production of Crude Oil matching the fixed stability constraint (bubble point pressure of Crude Oil at stock temperature equal or lower than 70% of tank pressure) requires that the last separator before the stock tank operates at storage pressure. Moreover, the temperature of the last separator needs to be higher than the tank temperature, depending on the molecular weight of the fluid processed (the heavier the fluid, the higher the temperature).

Analysing the recovery of C_{3+} hydrocarbons in Crude Oil and C_1 and C_2 hydrocarbons in the Gas phase, it results that the latter is practically complete (99%) both using two and three separators before the stock tank. The former results to be very low for Gas Condensate – like mixtures (23% − 27.5% when using two separators and 30.8% − 34% when using a sequence of three separators), it has a moderate value processing Near Critical – like reservoir fluids (66.4% − 70.3% when using two separators and 72% − 75% when using a train of three separators), while it reaches high values for Black Oil – like well mixtures (93.7% − 97% when using two separators and 96% − 97.9% when using a train of three separators). Therefore, it is fair to state that a separation train composed by a sequence of three separators allows to better recover C_3 and heavier hydrocarbons in Crude Oil and $C_{1\&2}$ hydrocarbons in the Gas stream produced.

The composition of the petroleum reservoir fluid to be processed slightly influence the operating conditions of the separators. Since as the molecular weight of the hydrocarbon feed decreases the recovery of C_{3+} in Crude Oil decreases, the more volatile the well mixture extracted, the higher the amount of C_{3+} condensate that must be recovered from the Gas phase. Furthermore, maximizing the recovery of propane and heavier hydrocarbons brings to large pressure drops between the first separator and the second one, leading to a high energy consumption at the compressors, to pressurize the Gas streams from the second and third separator to mix them with the Gas stream from the first separator.

Recommendations

After having completed this Master's Thesis project, some recommendations can be suggested for future developments of Oil&Gas separation train optimization.

An interesting aspect for future evaluations is the influence of the reservoir fluid stream operating conditions to be fed to the first separator. Indeed, a reduction of pressure and temperature before the first separator might lead to increase the recovery in Crude Oil and Gas phase and to smaller pressure drops between separators.

Moreover, the development of self – built MATLAB programs to simulate the separation process might be useful. Indeed, it could give the possibility to extend the complexity the optimization, including the energy consumption in compressors, liquid – liquid equilibria calculations (the composition of reservoir fluids includes water) and size of the equipment.

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Appendix

Crude Oil Compositions at Optimal Conditions

- **Sequence of Two Separators**

  o *First Separator at 60 bar & 70°C*

Table A1. Composition from Gas Condensate Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$4.29 \cdot 10^{-6}$	28.01	0.0012	C_9	11.06	121	0.801
CO_2	0.01	44.01	0.0019	C_{10}	7.71	133	0.809
C_1	0.22	16.04	0.0007	C_{11}	4.7	146	0.808
C_2	0.42	30.07	0.0028	C_{12}	3.77	160	0.819
C_3	1.76	44.1	0.002	C_{13}	3.85	176	0.840
iC_4	0.96	58.12	0.0025	C_{14}	3.33	187	0.850
nC_4	2.67	58.12	0.0025	C_{15}	3.07	200	0.856
iC_5	2.48	72.15	0.616	C_{16}	1.96	212	0.862
nC_5	3.34	72.15	0.626	C_{17}	2.52	230	0.844
C_6	7.89	85	0.678	C_{18}	1.41	243	0.844
C_7	12.78	90	0.747	C_{19}	1.13	256	0.853
C_8	17.3	102	0.776	C_{20+}	5.64	336	0.869

Table A2. Composition from Near Critical Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$6.06 \cdot 10^{-6}$	28.01	0.0012	C_9	7.62	119	0.775
CO_2	0.002	44.01	0.0019	C_{10}	4.19	134	0.788
C_1	0.2	16.04	0.0007	C_{11}	3.96	148	0.791
C_2	0.41	30.07	0.0028	C_{12}	4.36	162	0.801
C_3	1.36	44.1	0.002	C_{13}	3.77	177	0.818
iC_4	0.79	58.12	0.0025	C_{14}	3.12	188	0.830
nC_4	2.01	58.12	0.0025	C_{15}	3.49	201	0.836
iC_5	1.85	72.15	0.616	C_{16}	2.24	215	0.841
nC_5	2.34	72.15	0.626	C_{17}	2.57	234	0.839
C_6	5.08	85	0.677	C_{18}	2.43	250	0.843
C_7	8.92	92	0.734	C_{19}	2.15	264	0.852
C_8	11.33	104	0.756	C_{20+}	25.8	496	0.913

Table A3. Composition from Black Oil Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)
N_2	$7.65 \cdot 10^{-6}$	28.01	0.0012	C_{14}	4.27	190	0.836
CO_2	0.01	44.01	0.0019	C_{15}	4.19	206	0.842
C_1	0.33	16.04	0.0007	C_{16}	3.45	222	0.849
C_2	0.25	30.07	0.0028	C_{17}	2.85	237	0.845
C_3	0.28	44.1	0.002	C_{18}	2.81	251	0.848
iC_4	0.33	58.12	0.0025	C_{19}	2.46	263	0.858
nC_4	0.49	58.12	0.0025	C_{20}	2.11	275	0.863
iC_5	0.78	72.15	0.616	C_{21}	1.84	291	0.868
nC_5	0.44	72.15	0.626	C_{22}	1.61	305	0.873
C_6	1.76	86.18	0.655	C_{23}	1.49	318	0.877
C_7	5.32	96	0.738	C_{24}	1.32	331	0.881
C_8	7.96	107	0.765	C_{25}	1.16	345	0.885
C_9	7.07	121	0.781	C_{26}	1.09	359	0.889
C_{10}	6.65	134	0.792	C_{27}	0.99	374	0.893
C_{11}	5.15	147	0.796	C_{28}	0.95	388	0.897
C_{12}	4.61	161	0.810	C_{29}	0.93	402	0.900
C_{13}	4.49	175	0.825	C_{30+}	20.54	449.1	0.989

- ***First Separator at 30 bar & 40°C***

Table A4. Composition from Gas Condensate Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm^3)
N_2	$2.6 \cdot 10^{-6}$	28.01	0.0012	C_9	9.91	121	0.801
CO_2	0.01	44.01	0.0019	C_{10}	6.58	133	0.809
C_1	0.17	16.04	0.0007	C_{11}	3.88	146	0.808
C_2	0.47	30.07	0.0028	C_{12}	3.04	160	0.819
C_3	2.48	44.1	0.002	C_{13}	3.06	176	0.840
iC_4	1.46	58.12	0.0025	C_{14}	2.62	187	0.850
nC_4	4.13	58.12	0.0025	C_{15}	2.41	200	0.856
iC_5	3.64	72.15	0.616	C_{16}	1.53	212	0.862
nC_5	4.77	72.15	0.626	C_{17}	1.97	230	0.844
C_6	9.53	85	0.678	C_{18}	1.09	243	0.844
C_7	14.48	90	0.747	C_{19}	0.88	256	0.853
C_8	17.51	102	0.776	C_{20+}	4.38	336	0.869

Table A5. Composition from Near Critical Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$3.97 \cdot 10^{-6}$	28.01	0.0012	C_9	7.41	119	0.775
CO_2	0.02	44.01	0.0019	C_{10}	4.02	134	0.788
C_1	0.16	16.04	0.0007	C_{11}	3.77	148	0.791
C_2	0.46	30.07	0.0028	C_{12}	4.14	162	0.801
C_3	1.85	44.1	0.002	C_{13}	3.57	177	0.818
iC_4	1.1	58.12	0.0025	C_{14}	2.95	188	0.830
nC_4	2.75	58.12	0.0025	C_{15}	3.3	201	0.836
iC_5	2.31	72.15	0.616	C_{16}	2.12	215	0.841
nC_5	2.84	72.15	0.626	C_{17}	2.42	234	0.839
C_6	5.49	85	0.677	C_{18}	2.29	250	0.843
C_7	9.28	92	0.734	C_{19}	2.03	264	0.852
C_8	11.32	104	0.756	C_{20+}	24.37	496	0.913

Table A6. Composition from Black Oil Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$6.05 \cdot 10^{-6}$	28.01	0.0012	C_{14}	4.17	190	0.836
CO_2	0.015	44.01	0.0019	C_{15}	4.09	206	0.842
C_1	0.34	16.04	0.0007	C_{16}	3.36	222	0.849
C_2	0.4	30.07	0.0028	C_{17}	2.78	237	0.845
C_3	0.51	44.1	0.002	C_{18}	2.74	251	0.848
iC_4	0.56	58.12	0.0025	C_{19}	2.39	263	0.858
nC_4	0.79	58.12	0.0025	C_{20}	2.05	275	0.863
iC_5	1.05	72.15	0.616	C_{21}	1.79	291	0.868
nC_5	0.57	72.15	0.626	C_{22}	1.57	305	0.873
C_6	1.94	86.18	0.655	C_{23}	1.44	318	0.877
C_7	5.57	96	0.738	C_{24}	1.29	331	0.881
C_8	8.07	107	0.765	C_{25}	1.13	345	0.885
C_9	7.01	121	0.781	C_{26}	1.07	359	0.889
C_{10}	6.54	134	0.792	C_{27}	0.97	374	0.893
C_{11}	5.05	147	0.796	C_{28}	0.93	388	0.897
C_{12}	4.51	161	0.810	C_{29}	0.91	402	0.900
C_{13}	4.39	175	0.825	C_{30+}	20.01	449.1	0.989

• Sequence of Three Separators

○ *First Separator at 60 bar & 70°C*

Table A7. Composition from Gas Condensate Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$6.8 \cdot 10^{-7}$	28.01	0.0012	C_9	9.92	121	0.801
CO_2	0.01	44.01	0.0019	C_{10}	6.85	133	0.809
C_1	0.09	16.04	0.0007	C_{11}	4.16	146	0.808
C_2	0.53	30.07	0.0028	C_{12}	3.33	160	0.819
C_3	3.38	44.1	0.002	C_{13}	3.39	176	0.840
iC_4	1.79	58.12	0.0025	C_{14}	2.94	187	0.850
nC_4	4.73	58.12	0.0025	C_{15}	2.71	200	0.856
iC_5	3.47	72.15	0.616	C_{16}	1.73	212	0.862
nC_5	4.36	72.15	0.626	C_{17}	2.23	230	0.844
C_6	8.28	85	0.678	C_{18}	1.24	243	0.844
C_7	12.74	90	0.747	C_{19}	0.99	256	0.853
C_8	16.16	102	0.776	C_{20+}	4.97	336	0.869

Table A8. Composition from Near Critical Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$6.06 \cdot 10^{-6}$	28.01	0.0012	C_9	7.62	119	0.775
CO_2	0.002	44.01	0.0019	C_{10}	4.19	134	0.788
C_1	0.2	16.04	0.0007	C_{11}	3.96	148	0.791
C_2	0.41	30.07	0.0028	C_{12}	4.36	162	0.801
C_3	1.36	44.1	0.002	C_{13}	3.77	177	0.818
iC_4	0.79	58.12	0.0025	C_{14}	3.12	188	0.830
nC_4	2.01	58.12	0.0025	C_{15}	3.49	201	0.836
iC_5	1.85	72.15	0.616	C_{16}	2.24	215	0.841
nC_5	2.34	72.15	0.626	C_{17}	2.57	234	0.839
C_6	5.08	85	0.677	C_{18}	2.43	250	0.843
C_7	8.92	92	0.734	C_{19}	2.15	264	0.852
C_8	11.33	104	0.756	C_{20+}	25.8	496	0.913

Table A9. Composition from Black Oil Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$2.54 \cdot 10^{-6}$	28.01	0.0012	C_{14}	4.16	190	0.836
CO_2	0.015	44.01	0.0019	C_{15}	4.08	206	0.842
C_1	0.24	16.04	0.0007	C_{16}	3.36	222	0.849
C_2	0.46	30.07	0.0028	C_{17}	2.78	237	0.845
C_3	0.61	44.1	0.002	C_{18}	2.74	251	0.848
iC_4	0.63	58.12	0.0025	C_{19}	2.39	263	0.858
nC_4	0.86	58.12	0.0025	C_{20}	2.05	275	0.863
iC_5	1.08	72.15	0.616	C_{21}	1.79	291	0.868
nC_5	0.57	72.15	0.626	C_{22}	1.57	305	0.873
C_6	1.93	86.18	0.655	C_{23}	1.45	318	0.877
C_7	5.53	96	0.738	C_{24}	1.29	331	0.881
C_8	8.02	107	0.765	C_{25}	1.13	345	0.885
C_9	6.98	121	0.781	C_{26}	1.07	359	0.889
C_{10}	6.52	134	0.792	C_{27}	0.97	374	0.893
C_{11}	5.03	147	0.796	C_{28}	0.93	388	0.897
C_{12}	4.49	161	0.810	C_{29}	0.91	402	0.900
C_{13}	4.38	175	0.825	C_{30+}	20	449.1	0.989

o ***First Separator at 30 bar & 40°C***

Table A10. Composition from Gas Condensate Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$3.72 \cdot 10^{-7}$	28.01	0.0012	C_9	9.16	121	0.801
CO_2	0.009	44.01	0.0019	C_{10}	6.06	133	0.809
C_1	0.63	16.04	0.0007	C_{11}	3.57	146	0.808
C_2	0.52	30.07	0.0028	C_{12}	2.79	160	0.819
C_3	3.76	44.1	0.002	C_{13}	2.8	176	0.840
iC_4	2.13	58.12	0.0025	C_{14}	2.41	187	0.850
nC_4	5.72	58.12	0.0025	C_{15}	2.21	200	0.856
iC_5	4.28	72.15	0.616	C_{16}	1.41	212	0.862
nC_5	5.37	72.15	0.626	C_{17}	1.81	230	0.844
C_6	9.52	85	0.678	C_{18}	1	243	0.844
C_7	14.09	90	0.747	C_{19}	0.8	256	0.853
C_8	16.5	102	0.776	C_{20+}	4.02	336	0.869

Table A11. Composition from Near Critical Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$5.69 \cdot 10^{-7}$	28.01	0.0012	C_9	7.07	119	0.775
CO_2	0.02	44.01	0.0019	C_{10}	3.81	134	0.788
C_1	0.06	16.04	0.0007	C_{11}	3.57	148	0.791
C_2	0.51	30.07	0.0028	C_{12}	3.91	162	0.801
C_3	2.94	44.1	0.002	C_{13}	3.37	177	0.818
iC_4	1.66	58.12	0.0025	C_{14}	2.79	188	0.830
nC_4	3.92	58.12	0.0025	C_{15}	3.13	201	0.836
iC_5	2.77	72.15	0.616	C_{16}	2	215	0.841
nC_5	3.26	72.15	0.626	C_{17}	2.29	234	0.839
C_6	5.62	85	0.677	C_{18}	2.17	250	0.843
C_7	9.21	92	0.734	C_{19}	1.92	264	0.852
C_8	10.94	104	0.756	C_{20+}	23.05	496	0.913

Table A12. Composition from Black Oil Petroleum Reservoir Fluid

Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)	Component	Mole Fraction (%)	Mw (g/mol)	Density (g/cm³)
N_2	$2.03 \cdot 10^{-6}$	28.01	0.0012	C_{14}	4.11	190	0.836
CO_2	0.02	44.01	0.0019	C_{15}	4.03	206	0.842
C_1	0.26	16.04	0.0007	C_{16}	3.31	222	0.849
C_2	0.67	30.07	0.0028	C_{17}	2.74	237	0.845
C_3	0.83	44.1	0.002	C_{18}	2.7	251	0.848
iC_4	0.77	58.12	0.0025	C_{19}	2.36	263	0.858
nC_4	1.02	58.12	0.0025	C_{20}	2.02	275	0.863
iC_5	1.18	72.15	0.616	C_{21}	1.77	291	0.868
nC_5	0.62	72.15	0.626	C_{22}	1.55	305	0.873
C_6	1.98	86.18	0.655	C_{23}	1.43	318	0.877
C_7	5.58	96	0.738	C_{24}	1.27	331	0.881
C_8	8.02	107	0.765	C_{25}	1.11	345	0.885
C_9	6.94	121	0.781	C_{26}	1.05	359	0.889
C_{10}	6.46	134	0.792	C_{27}	0.95	374	0.893
C_{11}	4.98	147	0.796	C_{28}	0.91	388	0.897
C_{12}	4.44	161	0.810	C_{29}	0.89	402	0.900
C_{13}	4.32	175	0.825	C_{30+}	19.72	449.1	0.989

Results of the Objective Function (Φ)
Sequence of Two Separators

- ## *First separator at 60 bar & 70°C*

  - ### *Gas Condensate Mixture*

Table A13. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333
52.6	0.332	0.332	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333
46.2	0.330	0.331	0.331	0.332	0.332	0.333	0.333	0.333	0.333	0.333
39.7	0.327	0.329	0.329	0.330	0.331	0.332	0.332	0.333	0.333	0.333
33.3	0.325	0.326	0.327	0.328	0.329	0.330	0.331	0.331	0.332	0.332
26.9	0.321	0.323	0.324	0.326	0.327	0.328	0.329	0.330	0.331	0.331
20.5	0.315	0.317	0.319	0.321	0.323	0.325	0.326	0.328	0.329	0.330
14.1	0.304	0.308	0.311	0.314	0.317	0.319	0.321	0.323	0.325	0.326
7.6	0.283	0.288	0.293	0.297	0.302	0.306	0.309	0.313	0.316	0.319
1.2	0.189	0.201	0.211	0.222	0.231	0.240	0.248	0.256	0.263	0.270

  - ### *Near Critical Mixture*

Table A14. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815
52.6	0.812	0.813	0.814	0.815	0.815	0.815	0.815	0.815	0.815	0.815
46.2	0.809	0.811	0.812	0.813	0.813	0.814	0.815	0.815	0.815	0.815
39.7	0.806	0.808	0.809	0.810	0.811	0.812	0.813	0.814	0.814	0.815
33.3	0.802	0.804	0.805	0.807	0.809	0.810	0.811	0.812	0.813	0.813
26.9	0.795	0.798	0.801	0.803	0.805	0.807	0.808	0.809	0.811	0.812
20.5	0.786	0.790	0.793	0.796	0.799	0.801	0.804	0.806	0.807	0.809
14.1	0.770	0.775	0.780	0.785	0.789	0.792	0.796	0.799	0.801	0.804
7.6	0.736	0.744	0.751	0.758	0.765	0.771	0.776	0.782	0.786	0.791
1.2	0.592	0.608	0.624	0.638	0.652	0.665	0.677	0.688	0.699	0.710

o *Black Oil*

Table A15. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985
52.6	0.984	0.984	0.984	0.984	0.985	0.985	0.985	0.985	0.985	0.985
46.2	0.983	0.983	0.984	0.984	0.984	0.984	0.984	0.985	0.985	0.985
39.7	0.982	0.983	0.983	0.983	0.984	0.984	0.984	0.984	0.984	0.984
33.3	0.981	0.982	0.982	0.983	0.983	0.983	0.983	0.984	0.984	0.984
26.9	0.980	0.981	0.981	0.982	0.982	0.982	0.983	0.983	0.983	0.984
20.5	0.978	0.979	0.980	0.980	0.981	0.981	0.982	0.982	0.983	0.983
14.1	0.975	0.976	0.977	0.978	0.979	0.979	0.980	0.981	0.981	0.982
7.6	0.967	0.969	0.970	0.972	0.974	0.975	0.976	0.977	0.978	0.979
1.2	0.922	0.928	0.934	0.939	0.943	0.948	0.951	0.955	0.958	0.961

- *First Separator at 30 bar & 40°C*

o *Gas Condensate Mixture*

Table A16. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.407	0.407	0.407	0.407	0.407	0.407	0.407	0.407	0.407	0.407
25.9	0.406	0.406	0.406	0.407	0.407	0.407	0.407	0.407	0.407	0.407
22.8	0.405	0.405	0.405	0.406	0.406	0.406	0.406	0.406	0.407	0.407
19.7	0.403	0.404	0.404	0.404	0.405	0.405	0.405	0.406	0.406	0.406
16.6	0.401	0.402	0.402	0.403	0.403	0.404	0.404	0.404	0.405	0.405
13.6	0.399	0.399	0.400	0.401	0.401	0.402	0.402	0.403	0.403	0.404
10.5	0.394	0.395	0.396	0.397	0.398	0.399	0.399	0.400	0.401	0.401
7.4	0.385	0.387	0.389	0.390	0.392	0.393	0.394	0.395	0.396	0.397
4.3	0.366	0.369	0.372	0.374	0.376	0.379	0.381	0.383	0.385	0.387
1.2	0.299	0.304	0.309	0.314	0.318	0.323	0.327	0.332	0.336	0.340

o *Near Critical Mixture*

Table A17. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832
25.9	0.831	0.831	0.831	0.831	0.832	0.832	0.832	0.832	0.832	0.832
22.8	0.829	0.830	0.830	0.830	0.831	0.831	0.831	0.831	0.832	0.832
19.7	0.827	0.828	0.828	0.829	0.829	0.829	0.830	0.830	0.830	0.831
16.6	0.825	0.825	0.826	0.827	0.827	0.828	0.828	0.829	0.829	0.829
13.6	0.821	0.822	0.823	0.823	0.824	0.825	0.826	0.826	0.827	0.827
10.5	0.815	0.816	0.817	0.818	0.820	0.821	0.822	0.823	0.823	0.824
7.4	0.804	0.806	0.808	0.809	0.811	0.813	0.814	0.816	0.817	0.819
4.3	0.779	0.783	0.786	0.789	0.792	0.794	0.797	0.800	0.802	0.804
1.2	0.697	0.703	0.708	0.714	0.719	0.724	0.729	0.734	0.739	0.744

o *Black Oil Mixture*

Table A18. Trend of the Objective Function, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986
25.9	0.985	0.985	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986
22.8	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.986	0.986	0.986
19.7	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985
16.6	0.984	0.984	0.984	0.985	0.985	0.985	0.985	0.985	0.985	0.985
13.6	0.983	0.984	0.984	0.984	0.984	0.984	0.984	0.984	0.985	0.985
10.5	0.982	0.983	0.983	0.983	0.983	0.983	0.984	0.984	0.984	0.984
7.4	0.980	0.981	0.981	0.981	0.982	0.982	0.982	0.982	0.983	0.983
4.3	0.976	0.977	0.977	0.978	0.978	0.979	0.979	0.980	0.980	0.980
1.2	0.958	0.960	0.961	0.962	0.964	0.965	0.966	0.967	0.968	0.969

Recovery of C₃₊ Hydrocarbons in Crude Oil (R_{oil})
Sequence of Two Separators

- *First Separator at 60 bar & 70°C*

  o *Gas Condensate Mixture*

Table A19. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.332	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333	0.333
52.6	0.331	0.331	0.332	0.332	0.333	0.333	0.333	0.333	0.333	0.333
46.2	0.329	0.330	0.330	0.331	0.332	0.332	0.332	0.333	0.333	0.333
39.7	0.327	0.328	0.329	0.329	0.330	0.331	0.331	0.332	0.332	0.332
33.3	0.324	0.325	0.326	0.327	0.328	0.329	0.330	0.331	0.331	0.332
26.9	0.320	0.322	0.323	0.325	0.326	0.327	0.328	0.329	0.330	0.330
20.5	0.314	0.316	0.319	0.321	0.322	0.324	0.325	0.327	0.328	0.329
14.1	0.304	0.307	0.310	0.313	0.316	0.318	0.321	0.322	0.324	0.326
7.6	0.282	0.287	0.292	0.297	0.301	0.305	0.309	0.312	0.315	0.318
1.2	0.188	0.200	0.211	0.221	0.230	0.239	0.247	0.255	0.262	0.269

  o *Near Critical Mixture*

Table A20. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815	0.815
52.6	0.812	0.813	0.814	0.815	0.815	0.815	0.815	0.815	0.815	0.815
46.2	0.809	0.811	0.812	0.813	0.813	0.814	0.815	0.815	0.815	0.815
39.7	0.806	0.807	0.809	0.810	0.811	0.812	0.813	0.814	0.814	0.815
33.3	0.801	0.803	0.805	0.807	0.808	0.810	0.811	0.812	0.813	0.813
26.9	0.795	0.798	0.800	0.803	0.805	0.806	0.808	0.809	0.810	0.811
20.5	0.786	0.790	0.793	0.796	0.799	0.801	0.803	0.805	0.807	0.809
14.1	0.770	0.775	0.780	0.784	0.788	0.792	0.795	0.798	0.801	0.804
7.6	0.736	0.743	0.751	0.758	0.764	0.770	0.776	0.781	0.786	0.791
1.2	0.592	0.608	0.623	0.638	0.651	0.664	0.676	0.688	0.699	0.710

o *Black Oil Mixture*

Table A21. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985
52.6	0.984	0.984	0.984	0.984	0.985	0.985	0.985	0.985	0.985	0.985
46.2	0.983	0.983	0.984	0.984	0.984	0.984	0.984	0.985	0.985	0.985
39.7	0.982	0.983	0.983	0.983	0.984	0.984	0.984	0.984	0.984	0.984
33.3	0.981	0.982	0.982	0.983	0.983	0.983	0.983	0.984	0.984	0.984
26.9	0.980	0.981	0.981	0.982	0.982	0.982	0.983	0.983	0.983	0.984
20.5	0.978	0.979	0.980	0.980	0.981	0.981	0.982	0.982	0.983	0.983
14.1	0.975	0.976	0.977	0.978	0.979	0.979	0.980	0.981	0.981	0.982
7.6	0.967	0.969	0.970	0.972	0.974	0.975	0.976	0.977	0.978	0.979
1.2	0.922	0.928	0.934	0.939	0.943	0.948	0.951	0.955	0.958	0.961

• **First Separator at 30 bar & 40°C**

o *Gas Condensate*

Table A22. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.406	0.406	0.406	0.406	0.406	0.406	0.406	0.406	0.406	0.406
25.9	0.405	0.405	0.406	0.406	0.406	0.406	0.406	0.406	0.406	0.406
22.8	0.404	0.404	0.405	0.405	0.405	0.405	0.406	0.406	0.406	0.406
19.7	0.403	0.403	0.403	0.404	0.404	0.404	0.405	0.405	0.405	0.405
16.6	0.401	0.401	0.402	0.402	0.403	0.403	0.403	0.404	0.404	0.404
13.6	0.398	0.399	0.399	0.400	0.401	0.401	0.402	0.402	0.403	0.403
10.5	0.393	0.394	0.395	0.396	0.397	0.398	0.399	0.399	0.400	0.401
7.4	0.385	0.386	0.388	0.389	0.391	0.392	0.393	0.394	0.396	0.397
4.3	0.365	0.368	0.371	0.373	0.376	0.378	0.380	0.382	0.384	0.386
1.2	0.298	0.303	0.308	0.313	0.318	0.322	0.326	0.331	0.335	0.339

o *Near Critical Mixture*

Table A23. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832	0.832
25.9	0.831	0.831	0.831	0.831	0.832	0.832	0.832	0.832	0.832	0.832
22.8	0.829	0.829	0.830	0.830	0.830	0.831	0.831	0.831	0.831	0.832
19.7	0.827	0.828	0.828	0.829	0.829	0.829	0.830	0.830	0.830	0.831
16.6	0.824	0.825	0.826	0.826	0.827	0.827	0.828	0.828	0.829	0.829
13.6	0.821	0.821	0.822	0.823	0.824	0.825	0.825	0.826	0.827	0.827
10.5	0.814	0.816	0.817	0.818	0.819	0.821	0.822	0.822	0.823	0.824
7.4	0.803	0.805	0.807	0.809	0.811	0.813	0.814	0.816	0.817	0.818
4.3	0.779	0.782	0.785	0.789	0.791	0.794	0.797	0.799	0.802	0.804
1.2	0.697	0.702	0.708	0.713	0.719	0.724	0.729	0.734	0.739	0.744

o *Black Oil Mixture*

Table A24. Trend of C$_{3+}$ Hydrocarbons Recovery in Crude Oil, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986
25.9	0.985	0.985	0.986	0.986	0.986	0.986	0.986	0.986	0.986	0.986
22.8	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.986	0.986	0.986
19.7	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985	0.985
16.6	0.984	0.984	0.984	0.985	0.985	0.985	0.985	0.985	0.985	0.985
13.6	0.983	0.984	0.984	0.984	0.984	0.984	0.984	0.984	0.985	0.985
10.5	0.982	0.982	0.983	0.983	0.983	0.983	0.984	0.984	0.984	0.984
7.4	0.980	0.981	0.981	0.981	0.982	0.982	0.982	0.982	0.983	0.983
4.3	0.976	0.977	0.977	0.978	0.978	0.979	0.979	0.980	0.980	0.980
1.2	0.958	0.960	0.961	0.962	0.964	0.965	0.966	0.967	0.968	0.969

Bubble Point Pressure of Crude Oil at 15°C
Sequence of Two Separators

- ## First Separator at 60 bar & 70°C

  - ### *Gas Condensate Mixture*

Table A25. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	41.46	41.99	41.99	41.99	41.99	41.99	41.99	41.99	41.99	41.99
52.6	36.60	37.89	39.28	40.76	41.99	41.99	41.99	41.99	41.99	41.99
46.2	31.78	32.95	34.20	35.54	36.98	38.53	40.21	41.99	41.99	41.99
39.7	27.00	28.04	29.16	30.35	31.63	33.00	34.48	36.09	37.83	39.73
33.3	22.26	23.18	24.15	25.19	26.30	27.49	28.78	30.17	31.67	33.31
26.9	17.59	18.36	19.18	20.06	21.00	22.00	23.09	24.25	25.52	26.89
20.5	12.99	13.61	14.27	14.98	15.74	16.55	17.42	18.36	19.37	20.47
14.1	8.52	8.98	9.46	9.98	10.54	11.14	11.79	12.48	13.23	14.04
7.6	4.27	4.54	4.83	5.14	5.47	5.83	6.23	6.66	7.12	7.62
1.2	0.51	0.56	0.62	0.67	0.74	0.81	0.89	0.98	1.09	1.20

  - ### *Near Critical Mixture*

Table A26. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	41.04	41.54	41.54	41.54	41.54	41.54	41.54	41.54	41.54	41.54
52.6	36.30	37.63	39.05	40.57	41.54	41.54	41.54	41.54	41.54	41.54
46.2	31.59	32.78	34.06	35.42	36.88	38.46	40.15	41.54	41.54	41.54
39.7	26.91	27.96	29.09	30.29	31.58	32.97	34.46	36.07	37.83	39.73
33.3	22.26	23.17	24.15	25.18	26.30	27.49	28.78	30.17	31.67	33.31
26.9	17.66	18.42	19.24	20.10	21.04	22.04	23.11	24.27	25.53	26.89
20.5	13.12	13.72	14.37	15.06	15.81	16.60	17.46	18.38	19.38	20.47
14.1	8.68	9.12	9.59	10.09	10.63	11.21	11.84	12.52	13.25	14.04
7.6	4.41	4.67	4.94	5.24	5.56	5.91	6.28	6.69	7.14	7.62
1.2	0.56	0.60	0.65	0.71	0.77	0.84	0.91	1.00	1.09	1.20

o *Black Oil Mixture*

Table A27. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	69	63	57	51	45	39	33	27	21	15
59	43.77	44.35	44.35	44.35	44.35	44.35	44.35	44.35	44.35	44.35
52.6	38.85	39.90	41.03	42.27	43.62	44.35	44.35	44.35	44.35	44.35
46.2	33.96	34.89	35.91	37.01	38.20	39.51	40.94	42.52	44.25	44.35
39.7	29.10	29.92	30.80	31.76	32.80	33.95	35.19	36.56	38.07	39.73
33.3	24.27	24.96	25.71	26.53	27.42	28.39	29.46	30.62	31.90	33.31
26.9	19.46	20.03	20.65	21.32	22.06	22.86	23.73	24.68	25.73	26.89
20.5	14.68	15.13	15.61	16.14	16.71	17.33	18.01	18.75	19.57	20.47
14.1	9.94	10.26	10.60	10.98	11.38	11.83	12.31	12.83	13.41	14.04
7.6	5.26	5.45	5.65	5.86	6.10	6.35	6.63	6.93	7.26	7.62
1.2	0.74	0.78	0.82	0.86	0.91	0.96	1.01	1.07	1.13	1.20

• **First Separator at 30 bar & 40°C**

o *Gas Condensate Mixture*

Table A28. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	23.86	24.37	24.54	24.54	24.54	24.54	24.54	24.54	24.54	24.54
25.9	21.21	21.67	22.14	22.63	23.13	23.65	24.19	24.54	24.54	24.54
22.8	18.56	18.98	19.41	19.85	20.31	20.78	21.26	21.77	22.29	22.82
19.7	15.92	16.29	16.68	17.08	17.49	17.91	18.34	18.79	19.25	19.73
16.6	13.29	13.62	13.96	14.31	14.67	15.04	15.42	15.82	16.22	16.64
13.6	10.67	10.96	11.25	11.55	11.86	12.18	12.51	12.85	13.20	13.56
10.5	8.08	8.32	8.56	8.81	9.07	9.33	9.60	9.88	10.17	10.47
7.4	5.53	5.72	5.90	6.09	6.29	6.50	6.71	6.92	7.15	7.38
4.3	3.06	3.18	3.30	3.43	3.56	3.70	3.84	3.98	4.13	4.29
1.2	0.76	0.80	0.84	0.89	0.93	0.98	1.03	1.08	1.14	1.20

o *Near Critical Mixture*

Table A29. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	23.87	24.37	24.54	24.54	24.54	24.54	24.54	24.54	24.54	24.54
25.9	21.23	21.69	22.16	22.65	23.15	23.66	24.20	24.54	24.54	24.54
22.8	18.60	19.01	19.44	19.88	20.33	20.80	21.28	21.78	22.29	22.82
19.7	15.98	16.34	16.72	17.11	17.52	17.93	18.36	18.80	19.26	19.73
16.6	13.36	13.68	14.01	14.36	14.71	15.07	15.45	15.83	16.23	16.64
13.6	10.76	11.03	11.32	11.61	11.91	12.22	12.54	12.87	13.20	13.56
10.5	8.18	8.40	8.63	8.87	9.12	9.37	9.63	9.90	10.18	10.47
7.4	5.63	5.80	5.98	6.16	6.35	6.54	6.74	6.94	7.16	7.38
4.3	3.15	3.26	3.37	3.49	3.61	3.73	3.87	4.00	4.14	4.29
1.2	0.80	0.84	0.88	0.92	0.96	1.00	1.05	1.10	1.15	1.20

o *Black Oil*

Table A30. Trend of Bubble Point Pressure (bar) of Crude Oil at 15°C, varying Pressure and Temperature of Second Separator

Temperature (°C) / Pressure (bar)	39	36.3	33.7	31	28.3	25.7	23	20.3	17.7	15
29	24.70	25.11	25.41	25.41	25.41	25.41	25.41	25.41	25.41	25.41
25.9	22.03	22.40	22.78	23.17	23.58	24.01	24.46	24.92	25.41	25.41
22.8	19.36	19.69	20.03	20.38	20.75	21.13	21.53	21.94	22.37	22.82
19.7	16.70	16.99	17.29	17.60	17.92	18.25	18.60	18.96	19.34	19.73
16.6	14.05	14.29	14.55	14.81	15.09	15.37	15.67	15.98	16.31	16.64
13.6	11.39	11.60	11.81	12.03	12.26	12.50	12.75	13.01	13.28	13.56
10.5	8.75	8.91	9.08	9.26	9.44	9.63	9.83	10.03	10.24	10.47
7.4	6.12	6.24	6.36	6.49	6.62	6.76	6.91	7.06	7.21	7.38
4.3	3.50	3.58	3.66	3.74	3.82	3.91	4.00	4.09	4.19	4.29
1.2	0.94	0.96	0.99	1.02	1.04	1.07	1.10	1.13	1.17	1.20