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TKP4170(1) PROCESS DESIGN PROJECT

Title: Subsea Separation	**Keyword (3-4):** Subsea, separation, boosting, remote
Written by: Mandar Thombre, Marlene Lund and Hanne Betten	**Work period:** Autumn 2015
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Summary:

This project studies the feasibility of implementing subsea separation to a low pressure, high water cut, and remote oil field. Water and sand is separated subsea and injected to a disposal reservoir. Four different cases regarding cost of transport and location of the separation of oil and gas are studied. In the chosen case, the oil and gas is boosted with a multiphase pump and transported through 150 km pipelines with direct electrical heating to a Floating Production, Storage and Offloading (FPSO) unit for further separation.

The total investment of the project is found to be 1.3 bill. USD, the net present value (NPV) is found to be 1.88 bill. USD over a ten-year period, and the payback time is 3.7 years. For the project to be economically feasible, the oil price can drop about 60% of the current value.

Conclusions and recommendations:

The pressure and long distance transport issues are solved with the use of multiphase boosting. Flow assurance challenges due to the low temperature are dealt with by using heating of pipelines and chemical injection. Because of increasing water cuts, limited water handling capacity topside, and sand production, the separation of water and sand is done subsea. In terms of NPV, internal rate of return (IRR), investment on return (IOR), payback time and the sensitivity analysis, the project is economically feasible. However, not being able to establish the project risk and the many rough assumptions made is leading to inaccurate results from the investment analysis.

To implement this plant, further research and development of equipment used for pipeline heating and online measurements of oil in water is necessary.

Date and signature:

Acknowledgements

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Finally, we would like to thank Adriaen Verheyleweghen for his support throughout the project.

Abstract

The design basis for this project was a low energy oil field (26°C and 90 bar), 150 km away from the nearest receiving facility. Subsea separation, sand handling and water handling were chosen to avoid bringing water and sand topsides. Four possible design solutions regarding the boosting and transport of the oil and gas were modelled and cost estimated. Multiphase boosting and multiphase transport were found to be the best alternatives, as they provided the simplest design with low cost and power consumption, compared to the other possibilities. This design was also the most mature in terms of technical development.

The total investment of the chosen case was estimated to be 1.3 bill. USD. The annual power consumption was on average 4 MW, which together with the estimated maintenance costs lead to an annual operating cost of 17 mill. USD on average. The annual revenues from oil and gas sales together with the mentioned costs gave a total net present value of 1.88 bill. USD over a 10 year period. The break even oil price for this project was found to be about 23 USD/bbl.

Contents

1 Introduction

"One thing is clear: the era of easy oil is over". These were the words of then-CEO of the energy company Chevron, Dave O'Reilly in 2005 [1]. Remaining oil fields have difficulties that we have managed to avoid until today. Waters are deeper, fields smaller, distances longer, water cuts higher, oil more viscous, the environment more harsh but at the same time more sensitive. These are all key motivations to move more of the current oil- and gas processing down to the seabed. For instance, to produce remote- and low energy oil- and gas fields, it is necessary to boost the produced fluids subsea, in order for them to reach their final destination at a platform, an FPSO (Floating Production, Storage and Offloading) or a shore facility. Boosting or compression is also playing a role in increased oil- and gas recovery, especially for low pressure fields. Higher water cuts raise a demand for more efficient solutions for the handling of produced water. Separating out the produced water at the seabed could remove or reduce the demand of topsides produced water cleaning.

Subsea production systems are not a new invention. Already in the 1970s, subsea production of oil and gas was tested on the Norwegian continental shelf [2]. In the coming centuries, several underwater productions were installed and the technology was used all over the world. For instance, installing subsea production turned out to be economically beneficial for smaller discoveries that could not justify the building and operation of a platform installation [3]. Along the way, the idea of moving oil- and gas processing to the seabed has developed as a feasible solution for the new key issues of the industry. Today, a number of subsea boosting-, separation- and compression plants have been built.

This report studies the feasibility of combining these solutions, going a step further to the complete subsea production- and processing plant, referred to as the "Subsea Factory" by Statoil [4].

2 Background

This chapter will cover some of the subsea process units and utilities that are used today, coping with the several challenges regarding subsea operation. This includes subsea separation, boosting, gas compression, produced water- and sand handling, pressure safety, flow assurance and utility supply.

2.1 Subsea Separation

In a subsea oil well, there is usually a water layer beneath the oil called formation water. The main objective of subsea separation is to separate out the water from the oil, in order to avoid bringing it to the receiving facility. Throughout the production, the water cut will increase, and the topsides water handling facilities might reach its limitations. Other important advantages are reduced power consumption for fluid transportation, and reduced hydrate formation risk. The latter is described in closer detail in Chapter 2.6.

The concept of gravity separation, where sand, water, oil and gas separates in a pressure vessel due to density differences, may be used for this purpose. This method is usually used in topsides installations. Over the past decades, subsea separation has been employed at several fields, and different separation technologies have been used. For example, at the Statoil Tordis plant, a horizontal separator is used to separate the water from the oil. The separator is 17 meters long, has a diameter of 2.1 meters and a liquid retention time of 3 minutes. It can handle up to 100,000 barrels of water and 50,000 barrels of oil per day [5]. The separator is provided by FMC Technologies (Fig. 2.1).

Figure 2.1: Horizontal separator used at the Tordis plant [5].

A proprietary pipe separator system, provided by FMC Technologies, is used at the Petrobas Marlim plant for subsea separation. On receiving the mixture of oil, gas, water and sand, this system first separates the gas and then the water. The entire separation module can be retrieved to the surface and thus the maintenance and replacement is cheaper and more efficient [6].

Another common oil and water separating system is the hydrocyclone. A hydrocyclone separates the dense liquid, the water, from the less dense liquid, the oil, by use of centrifugal force. The water is pushed to the wall of the hydrocyclone, and taken out at one end of the system, while the oil is centered at the middle of the hydrocyclone, and exited through another opening. The water exiting a hydrocyclone has low content of oil, and can be discharged [7]. The separated oil, and some water, is injected to the part of the well stream which is taken to the receiving facility.

2.2 Subsea Boosting and Gas Compression

Over time, there will be a decline in pressure in the produced reservoirs. Water- or gas injection is often used for pressure support to ensure sufficient pressure for free flow of the production to the receiving facilities during the field lifetime. Subsea boosting- or gas compression is an energy efficient alternative option, especially in cases with low initial reservoir pressure and long tie-back distances. Additionally, the use of boosting or compression could contribute to increased oil recovery.

Currently, there are several existing boosting- and compression projects in manufacturing and operation. The two first full size subsea compression systems in the world are the Gullfaks Wetgas Compression system and the Åsgard Gas Compression system, which have both started operation in 2015. The Åsgard project consists of two compressor trains with 10 MW compressors [8], while the Gullfaks system has two 5 MW compressors [9]. Single- and multiphase boosting are slightly more developed, with for instance the Statoil Lufeng (5x0.4 MW single phase pumps) and Total CLOV (2x2.3 MW multiphase pumps) [10].

The pumps used for boosting in subsea operations are chosen according to the conditions specific to the processing plants. An important factor to consider when choosing a pump, along with the needed differential pressure, is the amount of gas it can handle. A single phase pump is preferred for water injection and oil boosting, due to the lower unit cost, compared to other kinds of pumps. For the boosting of liquid and gas together, or for variable gas volume fractions (GVF), a multiphase pump (MPP) is used. For lower GVF, it is also possible to use a hybrid pump, which is a combination of the two types of pumps. Subsea compressors are used for high GVF. For gas reservoirs, small amounts of condensate and water will be produced together with the gas, so a wet gas compressor can be chosen. Fig. 2.2 shows the types of pumps and compressors suitable for different GVF. Subsea pumps and compressors need to be enclosed in a pressure vessel to protect them from the surroundings at large water depths [11].

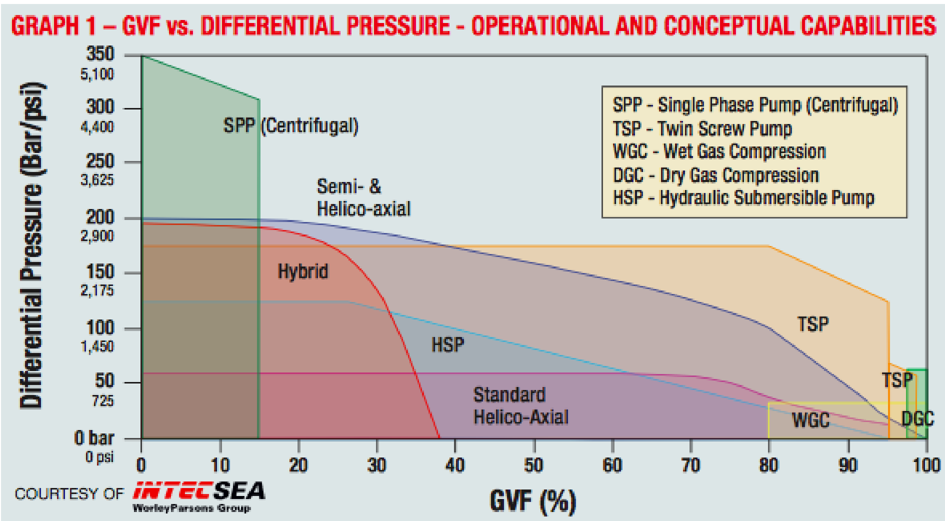

Figure 2.2: Suitable types of pumps and compressors at different differential pressures and GVF (Gas Volume Fractions) [10].

2.3 Produced Water Handling

The liquid which is pumped from a well is a mixture of hydrocarbons and the produced water. The produced water contains several dissolved salts, injected chemicals, and dispersed oil [12]. After separating it from the oil, the produced water is discharged. The water can either be pumped down in the reservoir to restore its pressure and achieve maximum oil recovery, or it can be injected to a separate discharge reservoir. This could be both energy and cost efficient, in addition to solving limited water handling capacity topsides. However, for produced water to be discharged to sea, there are strict rules regarding the content of oil in the water, since oil is very toxic to the environment. In Norway, the oil content in the discharged water should not be over 30 ppm [13]. There are currently no solutions for subsea discharge of water directly to the sea.

2.4 Sand Handling

In subsea processing, the production of sand is a common issue. Substantial quantities of produced sand can affect the operations of the various equipment. For example, the pumps, pipelines and compressors can get worn out or damaged by erosion, and the separators may get filled up. This calls for efficient sand handling techniques to limit the sand flowing out of the reservoir as well as the separation of any sand that may pass through with the oil and gas into the downstream vessels.

Sand production in subsea processing is typically not more than 10 ppm by weight [14]. For processing 10 million litres of oil per day, this corresponds to sand handling of 100 kg on a daily basis and 30-40 tons on an annual basis [3]. Typical equipment used for sand handling in subsea processing are hydrocyclone desanders, hammer mills, coalescing plate interceptors and other proprietary technologies [5].

At the Statoil Tordis station, any sand that comes from the well is deposited to the bottom of the separator tank. A 'sand jetting system', which uses specially designed nozzles to flush out the sand at regular intervals is the primary sand removal mechanism. A cyclonic sand removal system is also installed and can be used as a backup for the main sand removal system [15]. Both of these systems are provided by FMC Technologies. The flushed sand is taken to a gravity desander and a sand accumulator vessel in batches. This accumulated sand is then pressurized and discharged along with the produced water into the injection well using the water injection pump [5].

At the Marlim station, a multiphase inline desander, provided by FMC Technologies, shown in Fig. 2.3, is used as the initial sand separation system at the inlet [16].

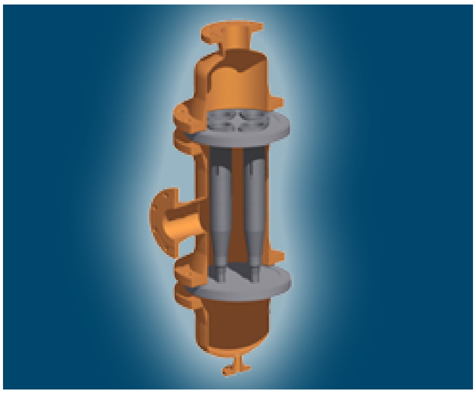

Figure 2.3: Inline desander provided by FMC Technologies [5].

This prevents large quantities of sand settling downstream in the separators. The 'sand jetting system' is used for flushing out whatever sand settles in the downstream vessels. Finally another inline desander is used to separate the remaining sand particles from the water, in order to protect the water injection well. At the Marlim station, the separated sand is taken to the topside facility along with the oil [5].

2.5 Subsea Design Pressure and Pressure Safety

The design pressure is defined as the maximum pressure pipes and equipment are designed to handle. It is set to the pressure at the most severe conditions (temperature and pressure) expected for the system [17]. This could for instance be determined by the maximum settle-out pressure. This is the equalized pressure obtained in the system in case of, for instance, a compressor trip [18]. In oil- and gas production, the shut-in pressure is also important to consider. Shut-in pressure occurs when there is production into the system from the reservoir, but no fluid outflow from the system. In subsea installations, the external pressure from the seawater bulk also plays an important role. This pressure is given by the hydrostatic pressure relation;

$$P_{ext} = \rho g h \tag{2.1}$$

Here, P_{ext} is the external pressure, ρ is the water density, g is the gravitational constant and h is the water depth.

If the internal pressure of a pressure vessel is low at some point, for instance when it is brought down to the seabed, the external pressure exerted by the water might cause hydrostatic imploding of the vessel.

The strength of a vessel or a pipeline, or its ability to handle pressure, is determined by several factors. First of all, it is affected by the strength of the material it is built from. Diameter and shell/wall thickness are also important [19].

On platforms and FPSOs, the system that protects against pressurizing equipment above the design pressure is the pressure relief system, where gas is removed and flared at the top of a tower to lower the pressure. Subsea, it is not an alternative to discharge the gas, as it is flammable and harmful to the environment. Instead, Safety Instrumented Systems (SIS) are used. An example of a SIS used subsea is the High Integrity Pressure Protection System (HIPPS). This system has the objective to shut down the pressure sources, which are the producing wells, by automatically closing one or more valves if high pressure is detected [20].

2.6 Flow Assurance and Chemical Injection

Flow Assurance refers to ensuring effective and economical flow of hydrocarbons from the reservoir to delivering the products to the market [21]. Several common operational issues related to flow assurance are possibly solved by chemical injection. Some of the most important of these are listed below.

- Gas hydrate formation
- Wax formation and deposition
- Inorganic scale deposition
- Corrosion

The following sections will introduce each of the phenomena and give examples of methods to protect against them.

2.6.1 Gas Hydrate Formation

Gas hydrates are ice- or snow-like solid structures that form when water and light hydrocarbons are mixed at high pressures and low temperatures. The hydrate formation temperature is the temperature where hydrates begin to form. Above this temperature, the risk of hydrate formation is significantly reduced. The hydrate formation temperature is estimated from the following relation;

$$T_{hydrate}[^{\circ}F] = 8.9P[psi]^{0.285} \tag{2.2}$$

Here, $T_{hydrate}$ is the hydrate formation temperature (given in Fahrenheit), and P is the pressure (given in Pounds per Square Inch) [22].

Hydrates can restrict or block the flow, lead to erosion in pipelines, damage compressors and even act as projectiles, the latter presenting a threat both to equipment and people.

There are three common ways of protecting against hydrates: Injection of hydrate inhibitors, heating of flowlines, and depressurization of flowlines. Thermodynamically inhibiting chemicals, like methanol (MeOH) and mono ethylene glycol (MEG), decrease the hydrate formation temperature. MEG is often the preferred chemical due to MeOH contamination of oil and gas and the toxicity of MeOH [23].

2.6.2 Wax Formation and Deposition

Waxes are long-chained hydrocarbons in the oil phase. They have high melting points, and can precipitate out as the liquid phase is cooled down. Wax particles in the oil phase will increase its viscosity, hence increase pumping costs. Deposition of wax on pipe walls will reduce the flow capacity, and could in the worst case plug the pipeline.

Wax control strategies used in industry include mechanical pigging of pipelines - using a device that runs through the pipeline and removes deposited wax, temperature control and injection of wax inhibitors. Wax inhibitor chemicals alter the surface of wax crystals, restraining them from sticking to solid surfaces [24].

2.6.3 Inorganic Scale Deposition

Inorganic scale is deposition of inorganic salts from produced water on pipeline walls and in equipment. Layers of salt crystals build up, and gradually reduce flow and productivity. Most salts have lower solubility at low temperatures, meaning that decreasing temperature will increase the scale issue. Use of scale inhibitors, which prevent the crystals from forming or growing, is the most common way to deal with the problem.

Scale inhibitors are usually injected continuously into wellstreams and re-injection water streams. Many scale inhibitors are harmful to the environment, and it is critical to find an environment-friendly and effective chemical. Polyaspartate is an example of such a chemical [25].

2.6.4 Corrosion

Carbon steel is a widely used material for pipelines in the oil and gas industry, and as long as water is present, corrosion will be a problem. Corrosion inhibitor chemicals are commonly used, and prevent corrosion by adsorbing onto a metal surface, forming a protective film [26].

2.7 Umbilicals and Power Supply

The umbilical cables transfer injection chemicals, hydraulic fluids, barrier fluids, communication in the form of fiber optics and also often electrical power from the receiving facility to the subsea installation. The cross-section of a typical umbilical cable is illustrated in Fig. 2.4.

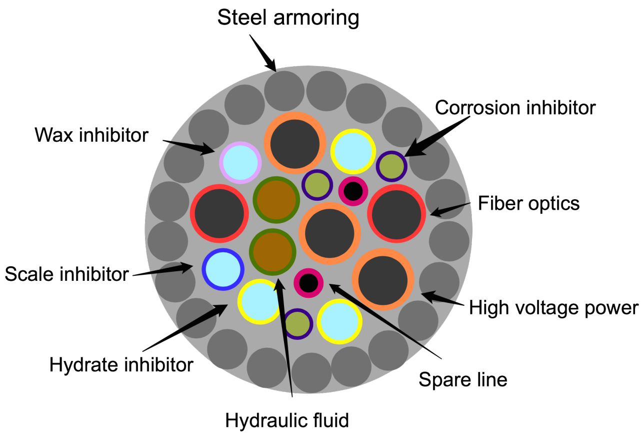

Figure 2.4: Illustration of the cross-section of an umbilical cable with power supply [27].

Choosing between separate or joint power and utility/communication umbilicals is a trade-off between reduced cost and avoiding common current transport issues. Long transport distances give significant voltage drop, which gives rise to the need for large power cables (cross-sectional areas). In such cases, it can be beneficial to have separate high voltage cables instead of using one large joint umbilical. Also, cross-talk (the current in the power cables disturbs the fiber optic communication signal) is a common issue that is avoided using separate cables [28].

For power supply, equipment controlling the power and the power distribution is needed; Variable Speed Drive (VSD), Switchgear and Transformer. For long tie-back distances and limited space on the topside facilities, it could be preferred to locate such equipment subsea [29].

3 Design Basis

The basis for the field development handled in this project was a low energy oil field, meaning that the starting point was a reservoir of low temperature and pressure. At the production start-up, the pressure is at its highest, declining with production time. During the production time, the water cut will increase and oil production rates are reduced. The production dynamics were taken into account by considering two different scenarios in time; early and late production (see Table 3.1). High water production (late production) was assumed from the start of year 7. For investment analysis, the time horizon of 10 years was used, although an oil field is expected to be in operation more than twice this time.

The development was assumed to be a tie-in to an FPSO which already received production from other fields. These frames were set to make the plant cost independent of the capital expenditures (CAPEX) and operating expenditures (OPEX) of the FPSO itself. In addition, an already producing FPSO will have a limited capacity for produced water handling and electrical power delivery. In this case, limited water handling was assumed from the first year of production. The power for the plant was assumed to be delivered by gas turbines on the FPSO. Three to four small gas turbines (60 MW or below) are typically used offshore [30, 31]. In this particular case, three gas turbines of 30 MW each were assumed. Since the FPSO delivers power to several production sites, it was assumed that the new subsea processing plant could utilize maximum 20 MW of the total 90 MW.

The location was assumed to be in arctic areas close to Norway. This information was used to give reasonable estimates in cost calculations. For instance, the electricity price is based on the current Norwegian industrial rate (0.09 USD/kWh)[32]. The oil price is based on the current rate of North Sea Brent Crude (48.6 USD/Barrel) [33]. Table 3.1 shows the complete design basis- and boundary data, and Table 3.2 shows the composition of the well stream.

Table 3.1: Design basis- and design boundary data.

Boundary Specification	**Value**
Gas Oil Ratio	108
Reservoir pressure, early production [34]	90 bar
Reservoir pressure, late production	50 bar
Oil production, early production [34]	7000 Sm^3/day
Water production, early production [34]	900 Sm^3/day
Oil production, late production [34]	400 Sm^3/day
Water production, late production [34]	8500 Sm^3/day
Reservoir temperature [34]	26°C
Wax content [34]	4.5wt%
Wax appearance temperature [34]	27°C
Distance plant to FPSO [34]	150 km
Water depth [34]	500 m
Sand production [3]	100 kg/day
Max. electricity delivery[31]	20 MW
Electricity price [32]	0.09 USD/kWh
Oil price [33]	48.6 USD/bbl.
Gas price [33]	2.56 USD/MMBtu

Table 3.2: Composition of the well stream.

Component	Mole fraction
Nirogen	0.0047
CO_2	0.0005
Methane	0.4900
Propane	0.0242
Ethane	0.0323
i-Butane	0.0054
n-Butane	0.0117
i-Pentane	0.0068
n-Pentane	0.0056
Hexane	0.0099
Heptane	0.0169
Octane	0.0217
Nonane	0.0174
C10+	0.3528

4 Process Description

The objective of the subsea separation plant is to avoid bringing produced water to the surface for processing, and to ensure safe and effective transportation of the produced oil and gas to the FPSO. The latter includes making up for pressure losses in pipelines and decreasing pressure in the reservoir, avoiding deposition of solids in pipelines and equipment, as well as phase stabilization of the fluids.

A schematic flowsheet of the different parts of the process is shown in Fig. 4.1. The wellstream that enters the plant contains oil (mainly heavy hydrocarbons), gas (mainly light hydrocarbons) and saline water. In the first part of the process, oil, gas, water and sand are separated. The oil and the gas proceeds to oil and gas treatment, which is intended to stabilize the two phases in order to avoid phase transitions and solids formation in the flowlines. The produced fluids are transported 150 km on the seabed, before they are brought half a kilometer up to the FPSO. To ensure that the product fluids have sufficient energy to move all the way from the wells to the FPSO, pressure boosting is necessary. The water undergoes removal of oil to meet the requirements for reservoir injection, and the sand production is handled.

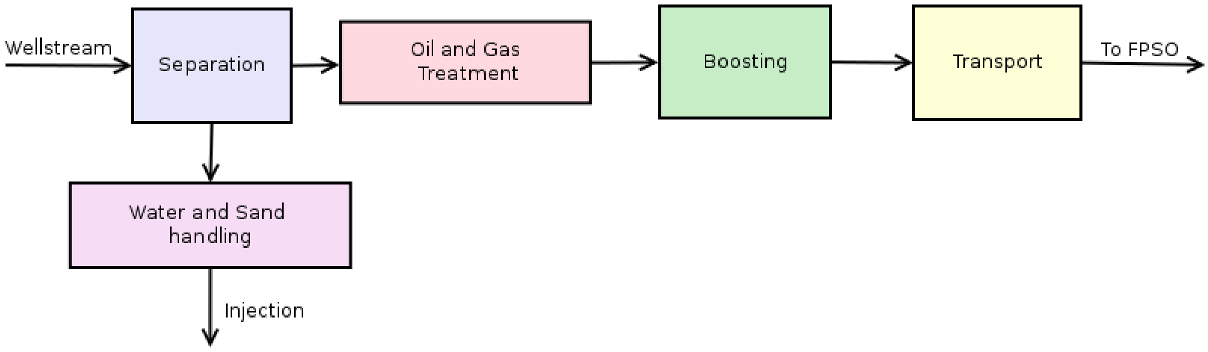

Figure 4.1: A general overview of the different parts of the process.

Several possible design solutions exist for the different blocks in Fig. 4.1. This is the case especially for boosting and fluid transport. The main question here is whether or not to boost and transport the vapour and liquid phases separately. To study this problem in further detail, four different cases were considered and compared in terms of cost and operation;

- Case 1: Multiphase pumping upstream of separation, single phase pumping and compression downstream of separation; and separate gas/oil flowlines and risers.
- Case 2: Single phase pumping and compression downstream of separation; and separate gas/oil flowlines and risers.
- Case 3: Multiphase pumping downstream of separation, a single set of flowline and riser; and separation of oil and gas topsides.
- Case 4: Single phase pumping and compression downstream of separation, a single set of flowline and riser; and separation of oil and gas topsides.

This chapter will give descriptions of chosen technology and solutions based on Chapter 2. First, chosen solutions which are common for all four studied cases will be given; separation, sand and water handling, chemical injection, and power and chemical supply. Then solutions for boosting and transportation of production fluids for the four different cases will be described in detail.

4.1 Separation

Separation of oil, gas, sand and water is done in a subsea 4-phase gravity separator. The separator itself was chosen to be a regular separator of the same type that is used topsides. This choice has both advantages and disadvantages. With this technology, the separator becomes large and heavy, which is less preferable when it comes to installation and retrieving of the vessel from the seabed for maintenance. The main advantages is that the large separator volume allows for slug-catching to a larger extent than compact- and pipe separators, in addition to the valuable experience already in the industry on separators of this kind.

4.2 Sand and Produced Water Handling

In this project, the sand handling system is modelled based on the one used on the Statoil Tordis substation (Chapter 2.4) i.e. the sand is discharged into a disposal reservoir after separation. Since the sand goes to the discharge side of the water injection pump, the pump itself does not need to handle large quantities of sand. The alternative, where the sand is carried topsides along with the oil and gas, may cause equipment damage in case of large sand particles escaping downstream. The produced water is treated with a hydrocyclone, and injected, along with the sand, to a separate reservoir for disposal. The pressure drop over the hydrocyclone creates the need for a pressure increase of the contaminant oil stream (overflow) before joining it together with the oil stream. Therefore, an ejector is installed. Injecting the water and sand to a disposal reservoir would cause the least costs for handling of the water. For this particular plant, it is assumed that the content of oil in the injected produced water must not be over 1000 ppm. This is a much higher tolerance than if the water were to be re-injected to the original reservoir, due to the risk of damaging the formation. For re-injection, it is assumed that the oil content should not be higher than 50 ppm, which would require further treatment of the produced water.

4.3 Material Selection

According to NORSOK, duplex stainless steel of type 22 Cr (2205) is suitable for subsea flowlines carrying well fluids, produced water and gas [35]. The same material can be used for subsea equipment such as separators [19]. It is assumed that this material is suitable for the entire subsea plant.

A few useful properties for this material are given in the table below.

Table 4.1: Properties of 22 Cr duplex stainless steel.

Property	
Density [36]	7800 kg/m^3
Composition [36]	Cr 22%, Ni 5%, Mo 3.2%
Cost [37]	1.56 x Carbon steel
Upper temp. limit [38]	315°C

4.4 Chemical Injection

The design basis for the plant includes low reservoir temperature (26°C) and pressure (90 bar), and relatively high water cut. From Equation 2.2, the hydratate formation temperature was approximated to 20.4°C at 90 bar. Even though the production temperature is above hydrate formation temperature, the design was given robustness against pressure and temperature changes. A small and continuous inhibitor injection at the wellhead was chosen to protect the wellstream and the part of the process upstream of transportation.

MEG was chosen as the hydrate inhibitor chemical due to the contamination risk of using MeOH.

The wax formation temperature for the well fluids was assumed to be 27°C [34]. The reservoir temperature is just below this level, meaning that injecting wax inhibitor into the wells is necessary. Direct Electrical Heating (DEH) was chosen as the solution to keep the products out of both the hydrate and wax formation envelopes during the long transportation to the FPSO.

Continuous scale inhibitor injection into the wells was also included as a part of the design, due to the high salinity of the produced water and the low temperature.

The chosen material for subsea pipelines and equipment was duplex stainless steel. This material has a high corrosion resistance, but given the highly corrosive conditions, it was assumed that additional protection was needed both subsea and topsides, where the chosen steels are likely of lower quality. Corrosion inhibitor was decided to be injected into the wells to protect all equipment and pipelines.

4.5 Umbilicals and Power Supply

The transfer distances for supplies for the particular plant handled are about 150 km. This means that the advantages of choosing two separate umbilical cables are present (see Chapter 2.7). Based on this statement, one high voltage cable and one umbilical containing injection chemicals, hydraulic fluids, barrier fluids and fiber optics was chosen.

4.6 Case 1

In Case 1, the transportation of oil and gas is done separately through two pipelines. The well stream is pumped through a multiphase pump and separated into four streams, oil, gas, water and sand, in a gravity separator. The gas stream is cooled so that remaining liquid can be separated out before the dry gas is then compressed and transported through a 150 km pipeline and a 510 m riser to the FPSO. The produced water is treated in a hydrocyclone to separate out most of the contaminants. The sand is removed through a sand jetting system, and together with the clean water it is injected to a disposal reservoir. The oil stream is pumped through a single phase pump and then transported through a 150 km long pipeline and a 510 m riser to the FPSO. The process flow diagram of Case 1 is shown in Fig. 4.2. The sand and water handling is the same for all four cases.

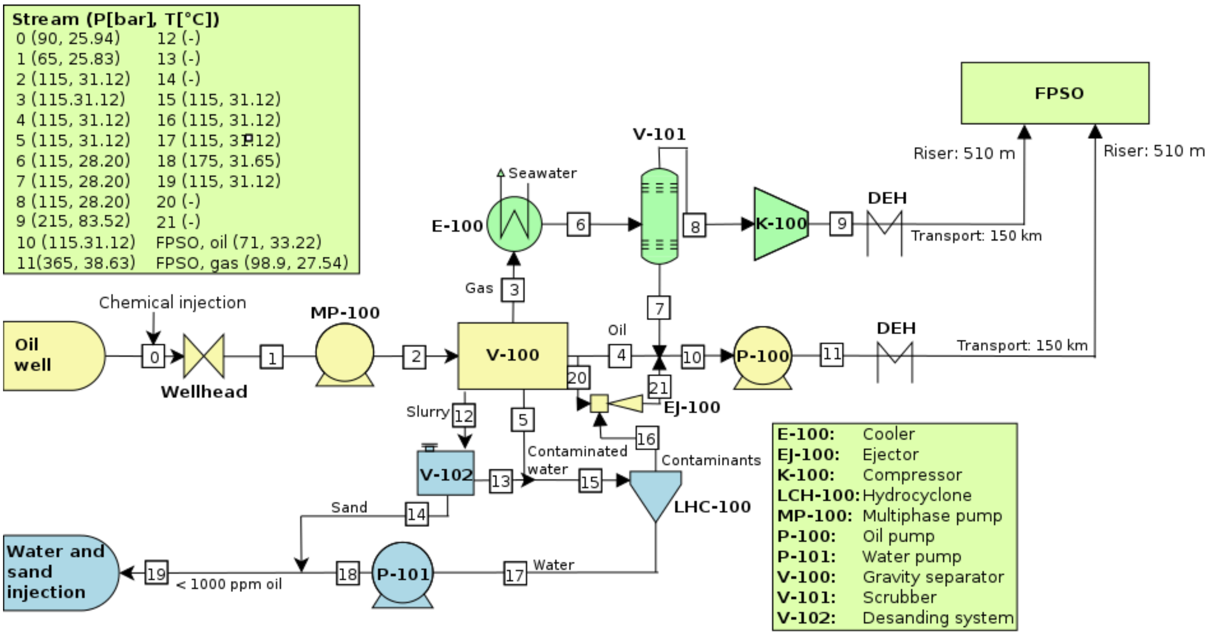

Figure 4.2: A process flow diagram of Case 1, with a multiphase pump and two separate risers for oil and gas.

4.7 Case 2

Case 2 is equal to Case 1 concerning number of transportation pipelines, but there is no multiphase pumping of the well stream before the gravity separator. The transport of oil and gas is done in two separate pipelines. The process flow diagram of Case 2 is shown in Fig. 4.3.

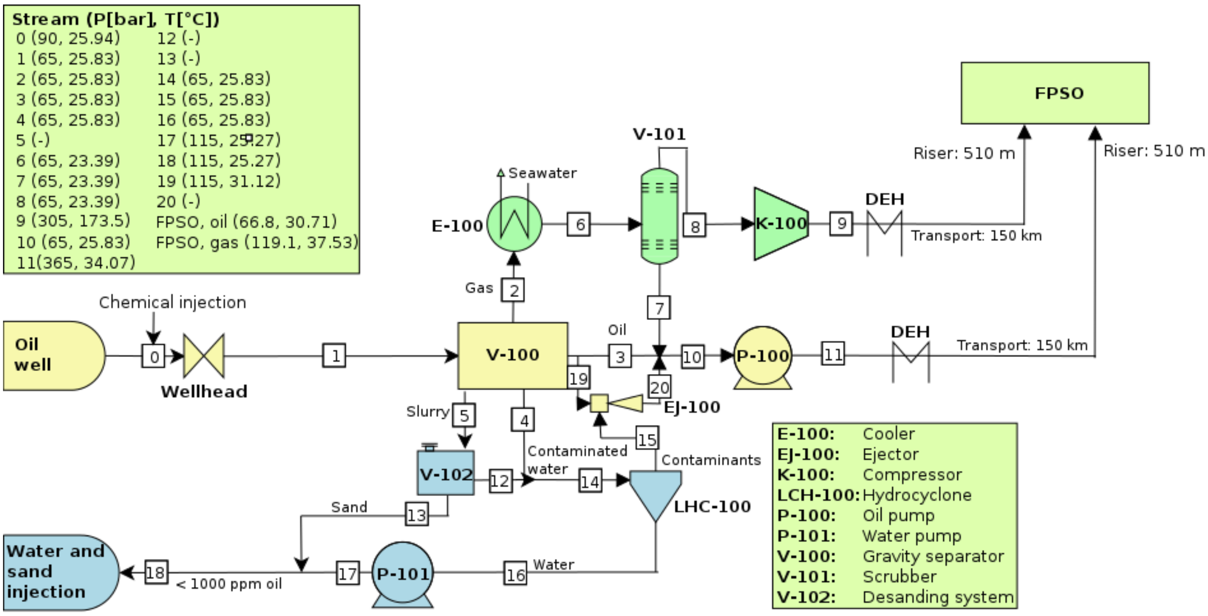

Figure 4.3: A process flow diagram of Case 2, with no multiphase pump and two separate risers for oil and gas.

4.8 Case 3

Case 3 describes a plant where the oil and gas is transported in a joint pipeline to the FPSO. After separating out the sand and water in the gravity separator, the oil and gas phases are joined together, pressurized through two multiphase pumps in series and transported through a 150 km pipeline and a 510 m riser up to the FPSO. The process flow diagram of Case 3 is shown in Fig. 4.4.

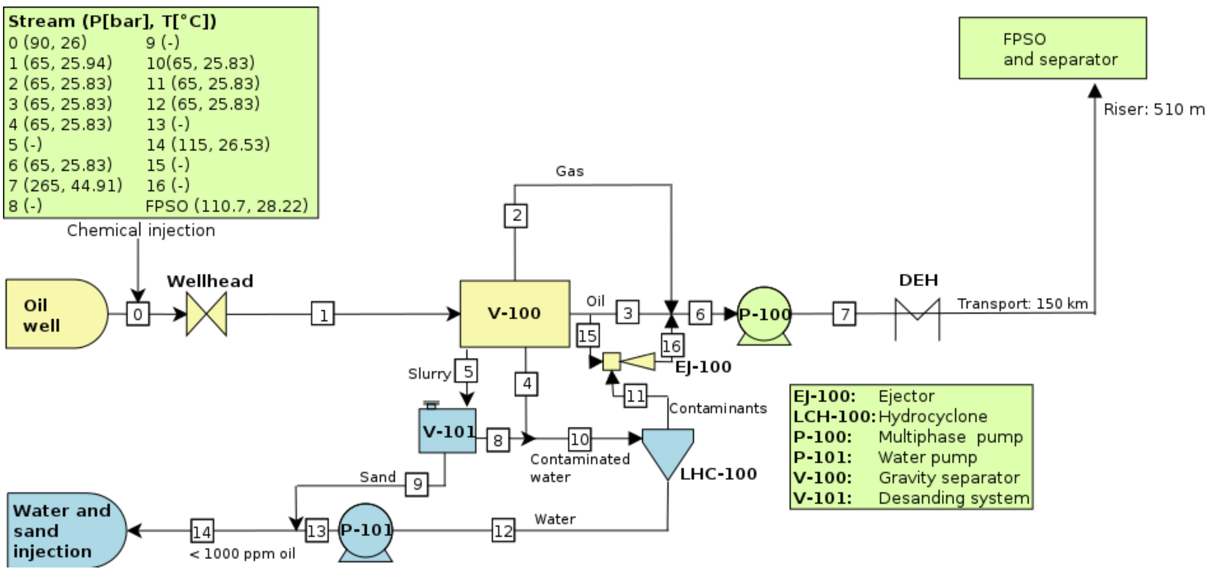

Figure 4.4: A process flow diagram of Case 3, with a multiphase pump and one riser for the oil and gas, which is to be separated at the top facility.

4.9 Case 4

Case 4 differs from Case 3 in terms of the pressurizing of the oil and gas. After separating out the water, the oil and gas are pressurized separately before joining the two phases and transporting them through a 150 km pipeline and a 510 m riser up to the FPSO. The process flow diagram of Case 4 is shown in Fig. 4.5.

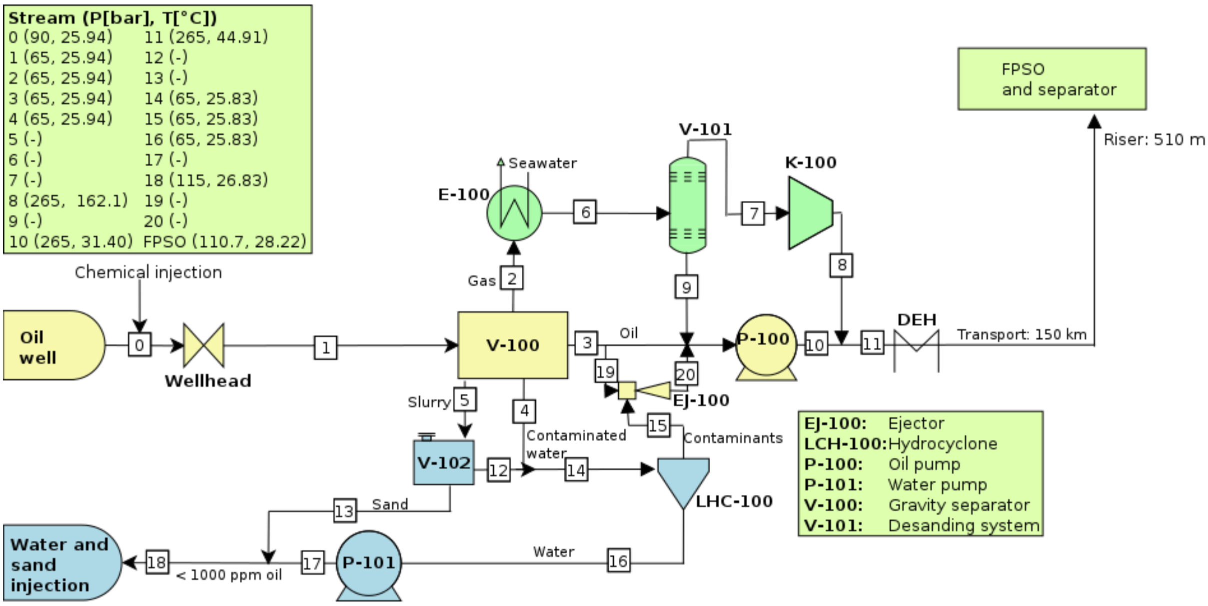

Figure 4.5: A process flow diagram of Case 4, with no multiphase pump and one riser for the oil and gas, which is to be separated at the top facility.

5 Flowsheet Calculations

The different plant cases are modelled using Aspen HYSYS. The models are simplified compared to the actual design. The main differences and assumptions are;

- Pressure drop only occurs over the wellhead, and in the transport pipelines. Therefore, the ejector used to pressurize the oil stream from the hydrocyclone is not included.
- Heat loss only occurs in the transport pipelines.
- A multiphase pump is modelled as a single phase pump and a compressor. Several multiphase pumps in series are modelled as one set of pump and compressor. This makes the model for Case 3 and Case 4 equal.
- The hydrocyclone is modelled as a 3-phase separator with no gas stream (liquid-liquid separation).
- The sand handling system is not included.
- The chemical injection system is not included.

- The models assume constant stream size and composition equal to the early production case given in the Design Basis chapter. For Case 3, both early and late production rates- and compositions are modelled.

5.1 Case 1

The flow diagram of the modelling of Case 1 in Aspen HYSYS is shown in the figure below. A larger version of the diagram is given in Appendix D. Table 5.1 shows selected stream data from the model (molar and mass flowrate, pressure, temperature and power duty).

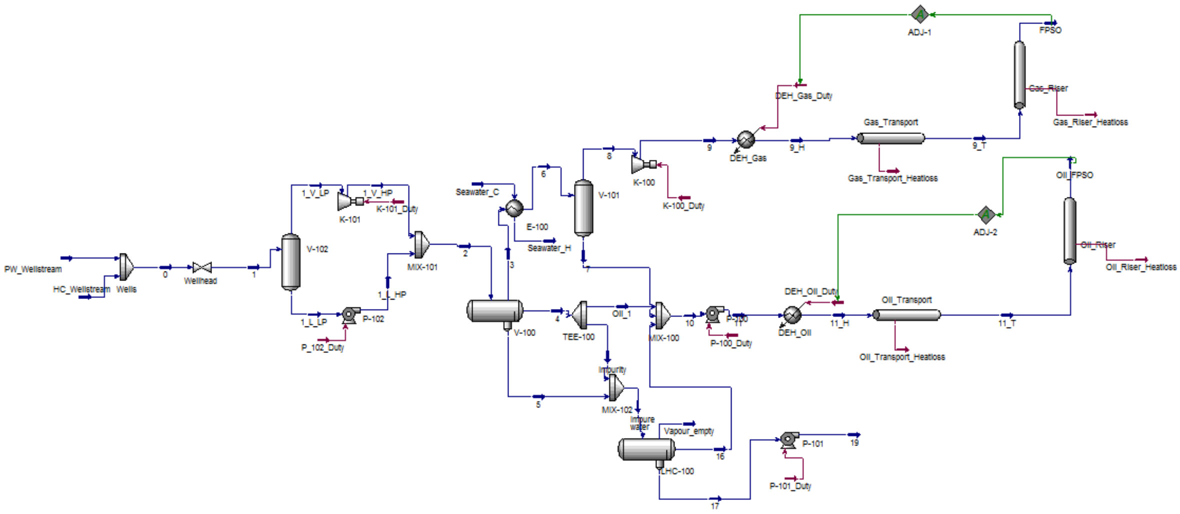

Figure 5.1: Flow diagram from the HYSYS modelling of Case 1.

Table 5.1: Stream data from the flowsheet calculations for Case 1, early production (maximum oil).

Stream	**Flowrate [kmol/h]**	**Flowrate [kg/h]**	**Pressure [bar]**	**Temperature [°C]**	**Duty [kW]**
PW_Wellstream	2 077	37 420	90	26	-
HC_Wellstream	1 907	2.189 e5	90	26	-
0	3 984	2.563 e5	90	25.94	-
1	3 984	2.563 e5	65	25.83	-
1_V_LP	735.6	12 720	65	35.83	-
1_V_HP	735.6	12 720	115	78.85	-
1_L_LP	3 249	2.436 e5	65	25.83	-
1_L_HP	3 249	2.436 e5	115	26.96	-
2	3 984	2.563 e5	115	31.12	-
3	537.5	9 337	115	31.12	-
4	1 370	2.096 e5	115	31.12	-
5	2 076	37 410	115	31.12	-
6	537.5	9 337	115	28.20	-
7	0.04885	0.9608	115	28.20	-
8	537.4	9 336	115	28.20	-
9	537.4	9 336	215	83.52	-
9_H	57.4	933.6	215	104.2	-
9_T	537.4	9 336	103.7	32.93	-
Impurity	27.41	4 192	115	31.12	-
Oil_1	1 343	2.054 e5	115	31.12	-
10	1 370	2.096 e5	115	31.12	-
11	1 370	2.096 e5	365	38.36	-
11_H	1 370	2.096 e5	365	47.89	-
11_T	1 370	2.096 e5	109	35.21	-
Impure water	2 104	41 600	115	31.12	-
16	27.41	4 192	115	31.12	-
17	2 076	37 410	115	31.12	-
19	2 076	37 410	175	31.65	-
Oil_FPSO	1 370	2.096 e5	71	33.22	-
Gas_FPSO	537.4	9 336	98.9	27.54	-
Gas_Riser_Heatloss	-	-	-	-	16
Gas_Transport_Heatloss	-	-	-	-	380
Oil_Riser_Heatloss	-	-	-	-	13.3
Oil_Transport_Heatloss	-	-	-	-	2 969
P-100_Duty	-	-	-	-	2 344.4
P-101_Duty	-	-	-	-	82.6
P-102_Duty	-	-	-	-	521.1
K-100_Duty	-	-	-	-	358.9
K-101_Duty	-	-	-	-	280.8
DEH_Gas_Duty	-	-	-	-	165.5
DEH_Oil_Duty	-	-	-	-	1 066.9

5.2 Case 2

The flow diagram of the modelling of Case 2 in Aspen HYSYS is shown in the figure below. A larger version of the diagram is given in Appendix D. Table 5.2 shows selected stream data from the model (molar and mass flowrate, pressure, temperature and power duty).

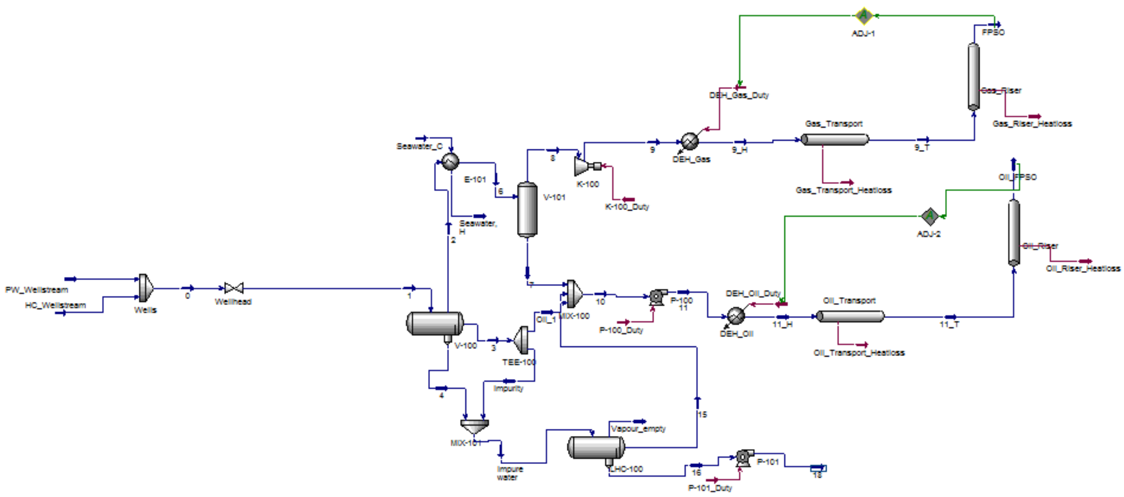

Figure 5.2: Flow diagram from the HYSYS modelling of Case 2.

Table 5.2: Stream data for the flowsheet calculations for Case 2, early production (maximum oil).

Stream	Flowrate [kmol/h]	Flowrate [kg/h]	Pressure [bar]	Temperature [°C]	Duty [kW]
PW_Wellstream	2 077	37 420	90	26	-
HC_Wellstream	1 907	2.189 e5	90	26	-
0	3 984	2.563 e5	90	25.94	-
1	3 984	2.563 e5	65	25.83	-
2	735.6	12 720	65	25.83	-
3	1 172	2.062 e5	65	25.83	-
4	2 076	37 410	65	25.83	-
6	735.6	12 720	65	23.39	-
7	0.0644	1.16	65	23.39	-
8	735.5	12 720	65	23.39	-
9	735.5	12 720	305	173.5	-
9_H	735.5	12 720	305	173.5	-
9_T	735.5	12 720	124.8	43.05	-
10	1 149	2.021 e5	65	25.83	-
Impurity	23.44	4 124	65	25.83	-
Oil_1	1 172	2.062 e5	65	25.83	-
11	1 172	2.062 e5	365	34.07	-
11_H	1 172	2.062 e5	365	39.66	-
11_T	1 172	2.062 e5	109.8	31.09	-
14	2 100	41 530	65	25.83	-
15	23.44	4 124	65	25.83	-
16	2 076	37 410	65	25.83	-
18	2 076	37 410	115	26.27	-
Oil_FPSO	1 172	2.062 e5	66.8	30.71	-
Gas_FPSO	735.5	12 720	119.1	37.53	-
Gas_Riser_Heatloss	-	-	-	-	24.3
Gas_Transport_Heatloss	-	-	-	-	115
Oil_Riser_Heatloss	-	-	-	-	11.8
Oil_Transport_Heatloss	-	-	-	-	2 484
P-100_Duty	-	-	-	-	2 714.7
P-101_Duty	-	-	-	-	68.7
K-100_Duty	-	-	-	-	1 135.3
DEH_Gas_Duty	-	-	-	-	0
DEH_Oil_Duty	-	-	-	-	599.7

5.3 Case 3&4

Case 3 and 4 are modelled the same way in HYSYS, due to the fact that a multiphase pump is modelled as a combination of a single phase pump and a compressor.

The flow diagram of the modelling of Case 3 and 4 in Aspen HYSYS is shown in the figure below. A larger version of the diagram is given in Appendix D. Table 5.3 and 5.4 show selected stream data from the early production and late production models, respectively.

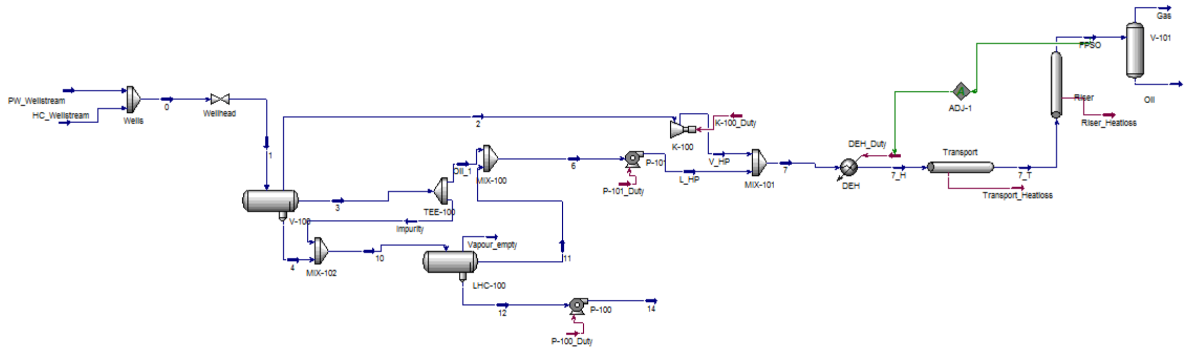

Figure 5.3: Flow diagram from the HYSYS modelling of Case 3 and Case 4.

Table 5.3: Stream data from the flowsheet calculations for Case 3 and 4, early production (maximum oil).

Stream	Flowrate [kmol/h]	Flowrate [kg/h]	Pressure [bar]	Temperature [°C]	Duty [kW]
PW_Wellstream	2 077	37 420	90	26	-
HC_Wellstream	1 907	2.189 e5	90	26	-
0	3 984	2.563 e5	90	25.94	-
1	2 984	2.563 e5	65	25.83	-
2	735.6	12 720	65	25.83	-
3	1 172	2.062 e5	65	25.83	-
4	2 076	37 410	65	25.83	-
Impurity	23.44	4 124	65	25.83	-
Oil_1	1 149	2.021 e5	65	25.83	-
6	1 172	2.062 e5	65	25.83	-
L_HP	1 172	2.062 e5	265	31.40	-
V_HP	735.6	12 720	265	162.1	-
7	1 908	2.189 e5	265	44.91	-
7_H	1 908	2.189 e5	265	56.99	-
7_T	1 908	2.189 e5	142.1	30.50	-
10	2 100	41 530	65	25.83	-
11	23.44	4 124	65	25.83	-
12	2 076	37 440	65	25.83	-
14	2 076	37 440	115	26.53	-
FPSO	1 908	2.189 e5	110.7	28.22	-
Riser_Heatloss	-	-	-	-	14.5
Transport_Heatloss	-	-	-	-	3 744.4
P-100_Duty	-	-	-	-	68.7
P-101_Duty	-	-	-	-	1 809.7
K-100_Duty	-	-	-	-	1 016.4
DEH_Duty	-	-	-	-	1 480

Table 5.4: Stream data from the flowsheet calculations for Case 3 and 4, late production (maximum water).

Stream	Flowrate [kmol/h]	Flowrate [kg/h]	Pressure [bar]	Temperature [°C]	Duty [kW]
PW_Wellstream	19 620	3.535 e5	50	26	-
HC_Wellstream	109	12 510	50	26	-
0	19 730	3.660 e5	50	26	-
1	19 730	3.660 e5	25	26.52	-
2	52.59	929.1	25	26.52	-
3	56.38	11 580	25	26.52	-
4	19 620	3.353 e5	25	26.52	-
Impurity	1.128	231.6	25	26.52	-
Oil_1	55.25	11 350	25	26.52	-
6	56.38	11 580	25	26.52	-
L_HP	56.38	11 580	195	31.12	-
V_HP	52.59	929.1	195	229.9	-
7	109	12 510	195	56	-
7_H	109	12 510	195	204.4	-
7_T	109	12 510	194.3	30.39	-
10	19620	3.537 e5	25	26.52	-
11	1.128	231.6	25	26.52	-
12	19620	3.538 e5	25	26.52	-
14	19620	5.338 e5	75	27.21	-
FPSO	109	12 510	153.6	27.06	-
Riser_Heatloss	-	-	-	-	13.64
Transport_Heatloss	-	-	-	-	1 385.6
P-100_Duty	-	-	-	-	650
P-101_Duty	-	-	-	-	85.47
K-100_Duty	-	-	-	-	119.42
DEH_Duty	-	-	-	-	1 205.8

6 Case Discussion

6.1 Cost

Comparison of the four cases in terms of cost was based on cost calculations of electric power, flowlines, multiphase pumps, single phase pumps, compressors, spare equipment (pumps and compressors) and an additional topside separator in the cases of one flowline from the subsea station to the FPSO. The parts of the plant that are the same for all four cases, like the sand- and water handling system and power/utility umbilicals, are left out of the cost comparison. The equipment sizing is done for early production (maximum oil production). Spare equipment for pumps and compressors are included, as the mean time between failure is assumed to be shorter than the economical lifetime of 10 years. The size and cost estimations are shown in Appendix A and B, respectively.

The cost of the equipment and the duty costs of each case is given in Tables 6.1 - 6.8. For the duty costs, the number of operation hours per year is assumed to be 8000. This correspond to the plant running 91% of the time.

Table 6.1: Equipment overview and estimated cost for Case 1.

Unit	**Cost [USD]**
Multiphase pump (MPP)	16 000 000
Spare MPP	10 000 000
Oil pump	2 808 585
Spare oil pump	1 041 206
Compressor	4 572 678
Spare compressor	1 695 194
Gas flowlines	93 675 000
Gas riser	714 000
Oil flowline	129 165 000
Oil riser	1 438 200
Total cost	**272 955 858**

Table 6.2: Duty overview and estimated cost for Case 1.

Duty	**Cost [USD/year]**
MPP duty	633 744
Oil pump duty	1 687 680
Compressor duty	202 176
Oil DEH duty	768 240
Gas DEH duty	119 160
Total duty cost	**3 411 000**

Table 6.3: Equipment overview and estimated cost for Case 2.

Unit	**Cost [USD]**
Oil pump	3 045 354
Spare oil pump	1 128 981
Compressor	22 761 554
Spare compressor	8 438 219
Gas flowlines	93 675 000
Gas riser	714 000
Oil flowline	129 165 000
Oil riser	1 438 200
Total cost	**272 212 303**

Table 6.4: Duty overview and estimated cost for Case 2. The HYSYS model for Case 2 gives that no heating of the gas is required to obtain desired temperature out of the plant (Gas DEH duty is zero).

Duty	Cost [USD/year]
Oil pump duty	1 954 800
Compressor duty	817 200
Oil DEH duty	431 784
Gas DEH duty	0
Total duty cost	**3 203 784**

Table 6.5: Equipment overview and estimated cost for Case 3.

Unit	Cost [USD]
MPP (2 in series)	32 000 000
Spare MPP	10 000 000
Flowline	154 500 000
Riser	2 177 700
Topside separator	462 183
Total cost	**207 389 071**

Table 6.6: Duty overview and estimated cost for Case 3.

Duty	Cost [USD/year]
MPP duty	2 034 792
DEH duty	1 065 600
Total duty cost	**3 100 392**

Table 6.7: Equipment overview and estimated cost for Case 4.

Unit	Cost [USD]
Oil pump	2 365 921
Spare oil pump	877 100
Compressor	22 035 465
Spare compressor	8 169 041
Flowline	154 500 000
Riser	2 177 700
Topside separator	462 183
Total cost	**198 836 600**

Table 6.8: Duty overview and estimated cost for Case 4.

Duty	Cost [USD/year]
Oil pump duty	1 302 984
Compressor duty	731 808
DEH duty	1 065 600
Total duty cost	**3 100 392**

Looking at the total cost, Case 1 is the most expensive, and Case 4 is the least expensive. Multiphase pumps are expensive compared to the possible differential pressure they can make. Having two multiphase pumps (as in Case 3) will cost more than having a single phase pump and a compressor (Case 4). Also, an additional flowline contributes to the total costs of Case 1 and 2 being higher than that of Case 3 and 4, which only have one flowline.

In terms of CAPEX, it is clear that Case 4, with only one flowline and no multiphase pump, is the least expensive alternative. However, there are some more aspects which need to be considered when it comes to the operational part of the plant.

6.2 Operation

A subsea plant should be simple and robust, to minimize the need for maintenance and inspection of the units. In Case 1 especially, there are a lot of units on seabed. This would require several spare units in case some units need to be changed or fixed. Case 1 and 2 also have two separate flowlines for the gas and oil. This means that there is twice the length of pipelines to be considered regarding maintenance and possible leaks along the way to the FPSO, in comparison to Case 3 and 4.

In Case 3 and 4 there is only one pipeline. The gas flow in the pipeline could contribute to the rise of the oil phase with the gas lift effect, which would then lower the pressure needed to transport the well fluid to the FPSO. However, there is need for a topside separator, which would require a certain space at the FPSO. This could be difficult to install on a vessel with limited space capacity. In addition, all units at seabed require topside equipment, and additional room is needed for the utility, control and power system for each unit.

Having a multiphase pump at seabed would decrease the number of units at the seabed by one, since there would not be need for both a compressor and a pump. However, the multiphase pumping of the gas and oil phase could result in an emulsification of the two phases, thus making it harder to separate them at a later stage. Having the multiphase pump before the gravity separator, which is Case 1, could affect the separation quality. In Case 3, however, the transportation pipeline is so long that there is assumed to be a separation effect throughout the transportation, so that multiphase pumping would not effect the topside separation.

Case 1, 2 and 4 all have a compressor unit in the design. These compressors have duties between 0.3 and 1.1 MW. Compared to the compressors used in the Åsgard Gas

Compression system (10 MW) mentioned in Chapter 2.2, these compressors are most likely too small to justify the installation. If they were to be installed regardless of this, they would need to undergo a qualification process.

6.3 Case Conclusion

Because of the few units at seabed in Case 3, as well as only having one riser, it seems to be the best alternative in terms of operation. It is also the second cheapest alternative in terms of CAPEX, and it avoids the issues with a very small compressor for the gas pressurization. Case 3 was therefore chosen to be the alternative for this plant.

7 Cost Estimation

A full cost estimation was only performed for Case 3, which will be presented in this chapter.

7.1 Capital Expenditures (CAPEX)

7.1.1 Cost Data of Relevant Projects

Cost data of subsea equipment are not easily obtained. Subsea operation belongs to relatively modern time, and such information is usually a well kept secret. However, it is possible to find costs for contracts awarded in projects, and what they include. Cost data for relevant projects are shown in Table 7.1.

It is not possible to directly compare these projects, as they are differently placed in time. Engineering costs and development of technology are playing large roles in contract cost for a project. This is easily seen when comparing the Åsgard and the Ormen Lange project costs. The Ormen Lange pilot project was built upon entirely new technology and about 90 000 engineering hours were spent, while the Åsgard project benefited from already tested subsea technology [39].

For the cost calculations, these data were used to estimate the cost of DEH cables, multiphase pumps and umbilicals. In addition, they were used to scale equipment costs to the correct order of magnitude for subsea installations.

From the Fossekall Dompap DEH contract, an installation cost of 11 million USD were assumed (half of the total cost). This leaves 440 USD per meter (in 2011) of piggyback cable. From the Valhall project, a cost estimate of 125 USD per meter of power cable (in 2006) was made based on the assumption that the power cable cost is one third of the total cost. The total cost includes power cable, fiber optic cable, land- and offshore equipment and installation. For the umbilicals (utility and communication), the Pazflor project was used as a basis. This contract only included delivery of the umbilicals, and the umbilicals delivered contained power cables. It is therefore assumed that the "power cable fraction" of the umbilical cost is 10%, and the resulting umbilical cost is 375 USD per meter (in 2008).

Table 7.1: Contract costs and descriptions for various subsea projects over the past few years. The project values are from the year the contract was signed, and they are not adjusted for inflation or the time value of money.

Project	Year	Description	Cost [mill. USD]
Johan Sverdrup [40]	2015	Semi-submersible drilling rig and drilling operations.	670
"Offshore North-Africa [41]"	2015	Subsea production system 330 and installation.	
Fossekall Dompap [42]	2011	DEH piggyback cable (25 km) and installation	22
Pazflor [43]	2008	Three umbilicals of 11.8 km each.	15
Valhall [44]	2006	292 km subsea power cable from shore, fiber optic cable, land- and plattform equipment and installation.	109
Åsgard Gas Compression [8]	2012	Three compressor trains, manifold, power distribution system, control system, topsides equipment, spare compressor, transport and installation.	185
Tordis [45, 46]	2005	Separator, desander, PLIM, one multiphase pump + spare, one single phase pump + spare, 12 km power umbilical, 12 km control umbilical, process control system, water injection subsea tree and installation.	97
Draugen Field [47]	2012	Power and control umbilical, manifold, one multiphase pump + spare and installation.	100
Girassol [48]	2012	Power and control system, 4 multiphase pumps + 2 spare with new technology with differential pressure up to 120 bar.	200
Ormen Lange Gas Compression Pilot [39, 49]	2006	One compression train, control and power system, and installation	130

Multiphase pumps are a part of the contracts for the Tordis, Draugen Field and Girassol projects. From these contracts, it is reasonable to assume that the cost of a multiphase pump module that is able to handle a maximum differential pressure of 120 bar, is 10 million USD. Here, it is assumed that the cost of the pump modules are 20% of the total contract cost, which also includes power- and control system both subsea and topsides, in addition to installation.

Also, these data were used to set a cost factor of 3 for adding engineering, module cost and custom design, and construction for subsea conditions to the costs of the compressors, single phase pumps, hydrocyclone and subsea pressure vessels. The size of this factor was determined by extracting reasonable costs for compressor modules and single phase pump modules, and comparing these with the cost calculations from Sinnot&Towler (see Appendix B).

7.1.2 Separators and Desander

Procedures for size estimation of separators and pressure vessels in general, as well as the specific data basis for the size estimations, are shown in Appendix A. The resulting dimensions, shell thicknesses and shell masses from the size estimations are shown in Table 7.2.

Table 7.2: Dimensions, thickness and shell mass for the different pressure vessels in Case 3.

Vessel	**Type**	**D_v [m]**	**L_v [m]**	**t_w [m]**	m_{shell} **[kg]**
Topsides separator	Horizontal	2.3323	11.661	0.0664	14 082.7
4-phase Separator, early stage	Horizontal	2.419	12.094	0.1	22 815.7
4-phase Separator, late stage	Horizontal	2.392	11.962	0.1	-
Desander	Vertical	0.7380	2.2139	0.1	1 274.3

The largest 4-phase separator size (for early production) is chosen. Cost of a pressure vessel is a function of the shell mass. The detailed cost relations for horizontal and vertical pressure vessels are shown in Appendix B.4. Here, the procedure of scaling the cost to final and installed cost in current time is also shown. Since the size of the separator is approximately the same for early and late stage of the production, the largest separator is chosen.

The final and installed costs of 2014 for the pressure vessels in Table 7.2 are shown in Table 7.3.

Table 7.3: Final costs of 2014 for all pressure vessels in Case 3. The cost includes engineering, design, material (22 Cr Duplex stainless steel), installation, piping, structure, coating, electrical work, and instrumentation and control.

Unit	**Installed Cost [USD]**
Topsides separator	462 182
4-phase Separator	675 795
Desander	63 360

7.1.3 Pumps

Cost of a single phase pump is divided into two; pump cost and motor cost. The pump cost is a function of the handled liquid flowrate, and the motor cost varies with the motor driver power. These data are obtained from the flowsheet calculations. The cost relations for a single phase pump and a motor, as well as the relevant data are shown in Appendix B.6.

Multiphase pumps are relatively new on the market, and there exist no cost relations for these. An approximate fixed price for uninstalled multiphase pump modules are extracted from Table 7.1 to be 10 mill. USD.

The resulting installed cost for the produced water pump and the the two multiphase pumps are given in Table 7.4. The spare pumps, one MPP and one produced water (PW) pump, are not installed, which gives a lower total cost for these units.

Table 7.4: Final installed costs of pumps and costs for spare pumps of 2014.

Unit	Installed Cost [USD]
MPP (2 units)	32 000 000
Spare MPP (1 unit)	10 000 000
PW pump	418 489
Spare PW pump	155 143

7.1.4 Flowlines and Risers

The cost estimation procedure for flowlines and risers are shown in Appendix B.3. A fixed price per meter of rigid or flexible pipelines is given, and a diameter size factor is used. Coating costs and DEH costs are added as a fixed price per meter. The DEH cost is extracted from Table 7.1 and discussed in chapter 7.1.1. The resulting installed flowline cost for Case 3 is given in Table 7.5.

Table 7.5: Final installed costs of 2014 for transportaton flowlines and risers in Case 3.

Unit	Installed Cost [USD]
Transportation flowline	154 500 000
Flexible riser	2 177 700

7.1.5 Umbilicals and Power Cables

The price per unit length of service- and communication umbilicals and power cables are discussed in Chapter 7.1.1. The used data and the resulting costs are shown in Appendix B.7. The resulting installed cost of 2014 is shown in Table 7.6.

Table 7.6: Final installed costs of 2014 for umbilicals and power cables.

Unit	Installed Cost [USD]
Umbilicals	18 893 379
Power cables	65 279 724

7.1.6 Hydrocyclone

The hydrocyclone cost is affected by the total liquid flowrate that comes into the hydrocyclone. The cost relation to calculate the basic cost of a hydrocyclone is shown in Appendix B.8. Then the basic cost is scaled for purchase year, material, and different installation factors in the same way as for pressure vessels, compressors and single phase pumps, and the final installed cost of 2014 is given in Table 7.7.

Table 7.7: Final installed costs of 2014 for the hydrocyclone in Case 3.

Unit	Installed Cost [USD]
Hydrocyclone	351 480

7.1.7 Total Equipment Cost

Equipment cost of all installed and spare units, as well as the total CAPEX is shown in Table 7.8. The costs are on a US Gulf Cost 2014 basis, and they all include engineering and design, material (22 Cr Duplex stainless steel), installation, structure, coating, electrical work, and instrumentation and control.

Table 7.8: Final installed cost of 2014 for all major equipment in Case 3. The bottom row shows the total equipment cost (CAPEX).

Unit	Installed Cost [USD]
Topsides separator	462 182
4-phase Separator	675 795
Desander	63 360
MPP (2 units)	32 000 000
Spare MPP (1 unit)	10 000 000
PW pump	418 489
Spare PW pump	155 143
Transportation flowline	154 500 000
Flexible riser	2 177 700
Umbilicals	18 893 379
Power cables	65 279 724
Hydrocyclone	351 480
Total	**293 226 441**

7.2 Operating Expenditures (OPEX)

The operating expenditures consist of power consumption, consumption of chemicals, labor, and maintenance. For a subsea installation, chemical consumption is approximately only 2% of the total OPEX, and therefore, the chemical consumption was not included in this cost study [50].

For a onshore processing plant, annual maintenance costs are typically 3-5% of the Inside Battery Limits (ISBL) investment costs [51]. For a subsea plant, maintenance and modification projects are rarely executed and very expensive, compared to an onshore plant. The availability to the equipment on the seabed is limited, and retrieving of the units to do maintenance topsides is usually necessary. This is both work intensive and time consuming, resulting in both high maintenance costs and lost production. On the other hand, the investment of a subsea plant is several times as high as for the onshore/topsides plant. Considering this, 5% of the investment costs are assumed to be sufficiently accurate for the purpose of the cost calculations in this project.

The plant is assumed to be in operation 8000 hours per year, which correspond to the plant running 91% of the time.

The total workload of operation of the subsea installations and the FPSO increases because of the complexity of the subsea plant. A few extra operators are likely needed, but compared to the annual maintenance cost, this cost is relatively small, and is therefore neglected in the profitability analysis.

For the first 6 years of production, there is assumed a fixed maximum oil production, giving a fixed power consumption. For the last 4 years of the total economical lifetime of 10 years, there is correspondingly a fixed production of maximum water, giving another power consumption rate.

The power consumption obtained from the flowsheet calculations, for the two cases are shown in Table 7.9.

Table 7.9: Power consumption in the cases of early and late production.

Unit	Early prod. power [kW]	Late prod. power [kW]
MPP (2 units)	2826.1	1929.12
PW pump	68.7	650
DEH	1480	1205.80

In Table 7.10, the operating expenditures (power and maintenance) are shown for each year of the economical lifetime.

Table 7.10: Operating expenditures for the cases of early (Year 1-6) and late (Year 7-10) production.

Year	Power cost [USD]	Maintenance cost [USD]	OPEX [USD]
1-6	3 149 856	14 661 322	17 811 178
7-10	2 725 142	14 661 322	17 386 464

8 Investment Analysis

The investment analysis relies on the following assumptions;

- To obtain realistic result on profitability evaluations, the investment costs which are not a part of the project scope (drilling and subsea production system) are assumed to be a total of 1 bill. USD. This number is obtained by considering the costs of the Johan Sverdrup drilling contract and the subsea production system contract awarded to OneSubsea outside the North-African coast, both contracts from 2015 (see Table 7.1).
- The project is financed with 100% equity.

- The discount rate used for NPV calculations is 10%.
- The corporate rate of taxation is assumed to be 35%.
- Working capital is assumed to be zero, since the new field is connected to an already existing production.
- The equipment is assumed to have no second-hand value.
- Depreciation is not taken into account.
- The economical lifetime is set to 10 years.

8.1 Profitability Evaluation

The probability evaluations done for this project consist of calculation of several profitability indicators: Net Present Value (NPV), payback time, Return on Investment (ROI) and Internal rate of return (IRR). The procedures and calculations used to obtain these values are described in Appendix C. The resulting values are shown in the table below.

Table 8.1: Profitability indicators for the project. The NPV and the IRR is the sum of discounted pre-tax cash flows. The payback time is calculated on the basis of uneven discounted after tax cash flows. The ROI is based on average after tax cash flows.

Profitability indicator	**Value and unit**
NPV	1.879 bill. USD
Payback time	3.66 years
ROI	22.29%
IRR	51.28%

A cash flow diagram is shown in Figure 8.1. The point where the curve intersects the x-axis represents the point in time where all investments are payed back by the incoming revenue (payback time). The colored areas represent the total investment (below the x-axis) or total profits (above the x-axis).

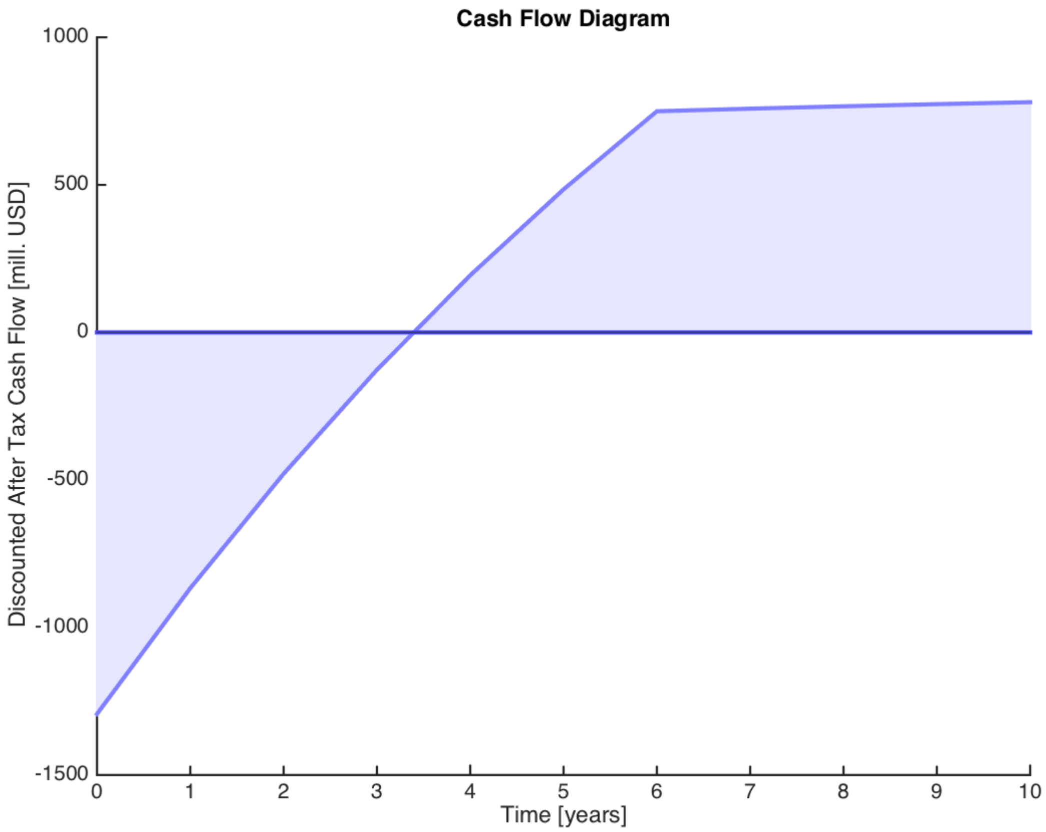

Figure 8.1: Cash flow diagram.

8.2 Sensitivity Analysis

Sensitivity analysis is calculations on how sensitive the profitability of the project is for changes or uncertainty in different parameters. The sensitivity analysis in this project studied the effect of changes in the oil price, the initial investments (CAPEX) and the operating expenditures (OPEX). A graphical representation of the sensitivity analysis is shown in Figure 8.2. The dashed axes indicate the base case profitability.

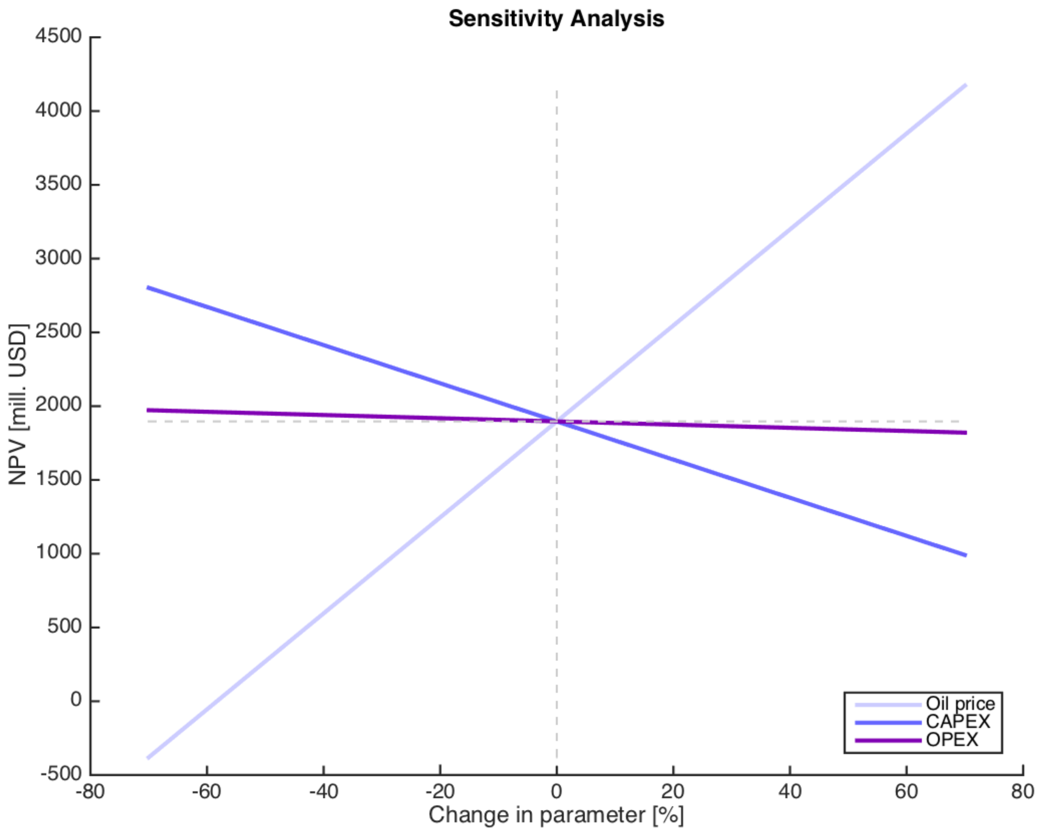

Figure 8.2: Graphical representation of the sensitivity analysis. The sensitivity in project NPV is tested against changes (± 70%) in oil price, CAPEX and OPEX.

From this figure, it is seen that the break even oil price is about 23.4 USD/bbl (limit to negative NPV). OPEX and CAPEX does not reach the break even point within the range of ± 70%.

9 Discussion

The design of Case 3 is evaluated as the simplest and most robust design for the scope of the plant, as well as the most technically mature. It has few units on the seabed, reducing the investment cost, the cost of maintenance activities and increasing the reliability of the plant. The main technology behind the plant is already tested and developed. A downside with the design is the requirement for available space topsides for the topsides separation.

Since there is no MPP unit available in HYSYS, the MPP is assumed to be a 'black box' which when given a certain input produces a certain output. For the purpose of the simulation, this 'black box' can thus be modelled as a compressor+pump unit in HYSYS. The input and output conditions for this unit are the same as those for the real model consisting of two MPPs in series, meaning that this assumption is not considered to be a source of error. The pressure drop and heat loss in all equipment are neglected in the model. Certainly, this is not realistic, and the real power consumption is higher than calculated. The real power consumption is still not expected to exceed the maximum capacity of 20 MW, because of the low modelled value of about 4 MW. The hydrocyclone is for simplicity modelled as a 3-phase separator. A 3-phase separator will not give a realistic model for the concentration of oil in the water that goes to the water injection well. The pressure drop, which is especially important for the hydrocyclone unit, is also left out. Another issue with the injection stream is the possibility to monitor the oil in water content, which is important to ensure that the content is within the range determined by the reservoir engineers. Currently, the oil content in produced water is measured with lab analysis, and there exists no fast and efficient solution for subsea oil in water analysis.

The two multiphase pumps have the power consumption of 1.4 MW each for early production (largest power consumption). This value is not far from the duties of the multiphase pumps of the Total CLOV project (2.3 MW), mentioned in Chapter 2. Correspondingly, the single phase pump for water injection has duty of 0.65 MW for late production (largest power consumption). This is near the value of the Lufeng project with 0.4 MW per pump. From this information, the conclusion is that multiphase- and single phase boosting is already tested for a production of this size.

The length of the multiphase pipeline in the plant design is 150 km. The longest multiphase pipeline in production today is the Snøhvit subsea to shore transport of 140 km [52]. In other words, no issues are foreseen with implementing the transport itself. When it comes to heating of the pipeline, a piggy-back cable for direct electrical heating of this length is not possible with today's technology. The longest DEH-equipped pipeline existing today is only 44 km (the Tyrihans project) [53]. More research and improvement of technology and equipment are necessary before such a facility can be installed. Alternatively, a MEG regeneration plant could be installed instead of using pipeline heating for hydrate protection. A facility for MEG injection is already incorporated in the design, but it is assumed that the consumed amounts of MEG in case of pipeline heating is too small to justify a regeneration process.

The NPV of the project is 1.88 billion USD. Since the NPV is positive, the project should be executed. Also, this is based on an economical lifetime of 10 years. Usually, oil and gas projects have significantly longer operation lifetimes, meaning that the total value of the project could be remarkably higher. The IRR was found to be 51.3%, which is larger than the used discount rate of 10%. This indicates that the project should be executed, in conformity with the NPV calculations. The ROI calculation reveals that the annual return on investments in the project is 22.3%. Whether or not this is an acceptable return is dependent on the project risk, which is hard to establish based on the limited information provided for this project. The payback time of 3.7 years is relatively short for a project of this size, especially compared to the possible lifetime of the plant. All of these aspects indicates that the project is economically feasible, and the probability of a loss is insignificant.

According to the sensitivity analysis, the profitability is quite sensitive to changes in the oil price, which is seen from the slope of the line in Fig.8.2. The break even oil price is about 23.4 USD/bbl, meaning that the oil price could be reduced about 60% of the current level. Considering that the current oil price is at one of its lowest levels in recent times, the economics of the project are robust against changes. This is also the case for changes in CAPEX and OPEX. The NPV is only reduced to about 1 bill. USD for a 70% increase in CAPEX. The profitability is close to independent to changes in OPEX.

Given these profitability indicators, the project seems to be economically feasible. Still, it is important to emphasize that these calculations are based upon rough assumptions. The factorial method of equipment cost estimations as well as cost estimations from existing projects both contribute to significant inaccuracy in the calculated investment. Project costs are for instance highly dependent of the maturity of the technology used. For the maintenance cost, a percentage of this investment was used. Further on, several assumptions were made about the financial frames of the project, such as the taxation rate, the discount rate and the capital structure of the project. If the project was partly financed by a bank loan, the project would benefit from the tax advantage of debt.

10 Conclusions and Recommendations

The basis for the plant is a remote low energy oil field. The pressure and long distance transport issues are solved with the use of multiphase boosting. Flow assurance challenges due to the low temperature are dealt with by using heating of pipelines and chemical injection. Because of increasing water cuts, limited water handling capacity, and sand production, the separation of water and sand is done subsea. In terms of NPV, IRR, IOR, payback time and the sensitivity analysis, the project is economically feasible. However, not being able to establish the project risk and the many rough assumptions made lead to inaccurate results from the investment analysis.

To implement this plant, further research and development of equipment used for pipeline heating and online measurements of oil in water is necessary. In addition, some of the equipment might need to go trough a qualification process before being installed on the seabed, for instance the subsea VSD.

The project's combination of the different existing technologies could be the next step on the way towards the complete subsea factory. It also shows that subsea processing is a feasible solution to many of the new challenges within oil and gas production.

List of Symbols- and Abbreviations

Table 10.1: Description of symbols used in the report.

Symbol	Description
a	Constant cost factor
A_l	Liquid cross-sectional area
b	Coefficient for variable cost
$CAPEX_i$	Capital expenditures of period i
C	Cost
C_l	Cost per length
C_{misc}	Miscellaneous costs
C_{ref}	Reference cost
C_0	Basic cost
CF_{avg}	Average after tax cash flow
CF_{fp}	Value of first positive cash flow
CF_i	After tax cash flow of period i
CF_{ln}	Abs. value of last negative cash flow
D_v	Vessel diameter
f_c	Installation factor, civil work
f_{el}	Installation factor, electrical work
f_{er}	Installation factor, erection
f_l	Installation factor, lagging, insulation and paint
f_i	Installation factor, instrumentation and control
f_{inst}	Merged installation factor
f_m	Installation factor, material
f_{MPP}	Installation factor for multiphase pumps
f_r	Room factor
f_s	Installation factor, structure
f_{size}	Size factor
f_{SUB}	Installation factor, subsea engineering
f_t	Type factor
g	Gravitational constant
h	Water depth
I	Investment
I_y	CEPCI index of year y
L	Length
L_v	Vessel length
m_{shell}	Shell mass
n	Exponential cost factor
NPV	Net present value
NI_i	Net income of period i

$OPEX_i$	Operantional expenditures of period i
P	Pressure
P_{comp}	Compressor driver power
P_{ext}	External pressure
P_{pump}	Pump driver power
q_s	Sand flowrate
q_l	Liquid volumetric flowrate
q_v	Vapour volumetric flowrate
r	Discount rate
R_i	Revenue of period i
ROI	Return on investment
S	Size
S_s	Maximum allowable stress
S_{ref}	Reference size
t_{holdup}	Hold-up time
$t_{holdup,desired}$	Desired hold-up time
$T_{hydrate}$	Hydrate formation temperature
t_{ln}	Last period of negative cash flow
t_{pb}	Payback time
t_r	Rate of taxation
t_w	Wall thickness
u_s	Stationary velocity
u_t	Settling velocity
V_{holdup}	Hold-up volume
v_v	Vapour velocity
V_v	Vessel volume
ρ	Density
ρ_l	Liquid density
ρ_m	Metal density
ρ_s	Wet sand density
ρ_v	Vapour density
τ_{act}	Actual residence time
τ_{req}	Required residence time

Table 10.2: Description of abbreviations used in the report.

Abbreviation	Description
CAPEX	Capital Expenditure
CEPCI	Chemical Engineering Plant Cost Index
DEH	Direct Electrical Heating
FPSO	Floating Production Storage and Offloading
GVF	Gas Volume Fractions
HIPPS	High Integrity Pressure Protection System
IRR	Internal Rate of Return
ISBL	Inside Battery Limits
MEG	Mono Ethylene Glycol
MEOH	Methanol
MPP	Multiphase Pump
NPV	Net Present Value
OPEX	Operating Expenditure
PW	Produced Water
ROI	Return on Investment
SIS	Safety Instrumentet System
VSD	Variable Speed Drive

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A Equipment Size Estimation

A.1 Separators

A.1.1 Size of Horizontal Separators

A horizontal separator is chosen when the liquid fraction in the inlet stream is high. The calculation procedure is taken from Sinnot&Towler [51].

Using a demister pad, the stationary velocity is equal to the settling velocity.

$$u_s = u_t = 0.07\sqrt{\frac{(\rho_l - \rho_v)}{\rho_v}} \tag{A.1}$$

Here, ρ_l is the liquid density and ρ_v is the vapour density.

The required residence time (τ_{req}) for the droplets to settle is given by the ratio of the liquid level, assumed to be half the vessel diameter (D_v), and the stationary velocity (u_s).

$$\tau_{req} = \frac{0.5D_v}{u_s} \tag{A.2}$$

The actual residence time (τ_{act} is given by the ratio between the vessel length (L_v) and the vapour velocity (v_v).

$$\tau_{act} = \frac{L_v}{v_v} \tag{A.3}$$

, where L_v is the vessel length, and the vapour velocity is given by;

$$v_v = \frac{8q_v}{\pi D_v^2} \tag{A.4}$$

, where q_v is the volumetric flow of vapour.

For pressures lower than 20 bar, the chosen length is 3 times the vessel diameter. Equivalently, pressures higher than 35 bar gives 5 times the diameter. For all pressures in between, the length to diameter factor is 4. The desired actual residence time should be equal to the required residence time. Solving for A.3 = A.2 gives the vessel diameter. The liquid level is assumed to be at half the vessel diameter.

When the dimensions are calculated, a check whether or not they give the desired hold-up time for the liquid is necessary.

The hold-up volume is found from;

$$V_{holdup} = A_l Lv \tag{A.5}$$

, where the liquid cross-sectional area, A_l is given by;

$$A_l = \frac{\pi D_v^2}{8} \tag{A.6}$$

The hold-up time is then calculated.

$$t_{holdup} = \frac{V_{holdup}}{q_l} \tag{A.7}$$

Here, q_l is the volumetric flow of liquid.

If this hold-up time is not equal to the desired hold-up time, the vessel diameter is scaled by a factor;

$$f = \left(\frac{t_{holdup,desired}}{t_{holdup}}\right)^{0.5} \tag{A.8}$$

A.1.2 Shell Mass

Pressure vessels like separators are often priced in terms of the mass of steel needed to produce it.

$$m_{shell} = \pi D_v L_v t_w \rho_m \tag{A.9}$$

Here, t_w is the thickness of the shell and ρ_m is the metal density.

The shell thickness is a function of internal pressure of the shell for a standard separator vessel.

$$t_w = \frac{PD_v}{2S_s - P} \tag{A.10}$$

, where S_s is the maximum allowable stress.

The necessary shell thickness for a subsea separator is affected by the chosen vessel diameter, the design pressure and the external pressure. The above thickness relation assumes atmospheric external pressure. In the DNV RP F-301 [19], necessary thickness to avoid vessel collapse with zero internal pressure (vacuum inside and maximum differential pressure over the shell) at different water depths is calculated. The thickness for 1000 m water depth, or 110 bar differential pressure, should be at least 80 mm for a vessel with diameter of 2.1 m. The same vessel should have a thickness of 100 mm with differential pressure of 180 bar [19].

The external pressure at 500 m water depth is 50 bar (see chapter 2.5). In this case, the design pressure is unknown. According to the flowsheet calculations, the highest pressure obtained in Case 3 is 265 bar. This pressure is obtained for the gas phase in a limited part of the plant, and the amounts of gas is relatively small compared to the liquid phase. Based on this, it is unlikely that the internal pressure will exceed 200 bar. At the lowest possible pressure in the vessel, the differential pressure is 50 bar. At the assumed design pressure, the differential pressure is 150 bar. The thickness should be larger than 80 mm, and the diameter taken into consideration (see Separator Sizing Results), 100 mm shell thickness is chosen.

A.1.3 Separator Sizing Results

All used data and the resulting dimensions from size calculations for the topside- and 4-phase separators, are shown in Table A.1 and A.2.

Table A.1: Data used for the calculations and the resulting calculated dimensions and shell mass of the topside separator.

Property	Value
Type	Horizontal
ρ_l	834.8 kg/m^3
ρ_v	81.47 kg/m^3
q_l	0.0692 m^3/s
q_v	0.0359 m^3/s
P	110.7 bar
S	2000 bar
ρ_m	7800kg/m^3
t_{holdup}	359.7 s
D_v	2.8 m
L_v	13.2 m
t_w	66.4 mm
m_{shell}	14082 kg

Table A.2: Data and calculated dimensions and shell mass of the subsea 4-phase separator. The data is for the early production case, which gives the largest separator volume.

Property	Value
Type	Horizontal
ρ_l	869.0 kg/m^3
ρ_v	52.74 kg/m^3
q_l	0.0779 m^3/s
q_v	0.0670m^3/s
ρ_m	7800kg/m^3
t_{holdup}	356.6 s
D_v	2.4 m
L_v	12.1 m
t_w	100 mm
m_{shell}	22816 kg

A.2 Desander

The desander is a pressure vessel which stores the sand for a determined period or hold-up time. It is sized similarly as a vertical separator, except for the separation calculation of the settling of drops. As for the 4-phase separator, the length is chosen as a pressure determined factor times the diameter.

The required volume of the vessel (V_v) is found by;

$$V_v = \frac{f_r q_s t_{holdup}}{\rho_s} \tag{A.11}$$

, where f_r is a factor increasing the vessel volume such that some room is left in the desander when the hold-up time is reached. q_s is the sand production per unit time, and ρ_s is the wet sand density.

The volume of a cylindrical shaped vessel is given by;

$$V_v = \frac{\pi D_v^2 L_v}{4} \tag{A.12}$$

Inserting for the vessel length and setting Equation A.11 equal to Equation A.12, the vessel diameter can be solved for.

The hold-up time, or the period between emptying the desander is set to 14 days. The vessel is set to be 30% (by volume) larger than the necessary volume to contain 14 days of sand production. Since the desander could be exposed to the same pressures as the subsea separator, the thickness is chosen to be 100 mm.

All used data and the resulting dimensions from the size calculations are shown in the table below.

Table A.3: Data and calculated dimensions and shell mass for the desander.

Property	**Value**
Type	Vertical
ρ_s	1922 kg/m^3
q_s	100 kg/day
t_{holdup}	14 days
f_r	1.3
ρ_m	7800kg/m^3
D_v	0.74 m
L_v	2.21 m
t_w	100 mm
m_{shell}	1274 kg

B Equipment Cost Estimation

B.1 Installation Cost Factors

The different installation factors used to calculate the ISBL Cost of each unit in the plant, are given in Table B.1.

Table B.1: Installation factors used to calculate the ISBL Cost [51].

Factor	Value [-]	Description
f_m	1.56	Material factor 22Cr Duplex Stainless Steel
f_{er}	0.3	Equipment erection factor
f_p	0.8	Piping factor
f_i	0.3	Instrumentation and Control factor
f_{el}	0.2	Electrical work factor
f_c	0.3	Civil Engineering work factor
f_s	0.2	Structures and buildings factor
f_l	0.1	Lagging, insulation and paint factor
f_{sub}	3	Factor including subsea construction and engineering

The factors are merged together to form a single installation factor (f_{inst}). This factor is used for all equipment which is cost estimated using cost relations for onshore, uninstalled equipment in carbon steel.

$$f_{inst} = (1 + f_p)f_m + f_{er} + f_i + f_{el} + f_c + f_s + f_l + f_{sub} \tag{B.1}$$

The uninstalled cost for the multiphase pumps are estimated from table 7.1. This means that material, insulation and paint, structures, civil engineering work and subsea engineering are already included in the cost. Therefore, an installation factor for the multiphase pumps (f_{MPP}), including only what is not already taken into account, is used.

$$f_{MPP} = f_p + f_{er} + f_i + f_{el} \tag{B.2}$$

The values of f_{inst} and f_{MPP} are shown in table B.2.

Table B.2: Merged installation factors.

Factor	Value [-]
f_{inst}	4.208
f_{MPP}	1.600

B.2 Chemical Engineering Plant Cost Index (CEPCI)

Markets are constantly changing, and the Chemical Engineering business is no exception. To adjust for price changes in process equipment, the Chemical Engineering Plant Cost Index (CEPCI) is often used. An overview of the CEPCI values for the recent years is given in Table B.3.

Table B.3: Chemical Engineering Plant Cost Index for the recent years [54, 55].

Year	CEPCI
2006	499.6
2007	525.4
2008	575.4
2009	521.9
2010	550.8
2011	585.7
2012	584.6
2013	567.3
2014	579.8

B.3 Flowlines and Risers

For cost estimation of flowlines and risers, the cost model from Subsea Engineering Handbook was used [50].

$$C = f_t f_{size} C_0 L + C_{misc} L \tag{B.3}$$

Here, C is the flowline cost, f_t is a type factor (rigid or flexible), f_{size} is a size factor, C_0 is basic cost per unit length, L is the length of the pipeline and C_{misc} is miscellaneous costs per unit length (for instance pipe coating).

The average basic cost is 230 USD/m for rigid pipe and 2300 USD/m for flexible pipe. Size factor, coating costs and DEH costs for different pipe diameters are given in the table below. Here, the type factor is included in the basic costs and the size factor. The size factor includes the pressure rating. The DEH cable cost is assumed to be 440 USD per meter, and is only possible to install on rigid pipelines. Table B.4 shows a summary of the size factors and costs for different pipe diameters- and types.

Table B.4: Size factors, DEH costs and coating costs (miscellaneous) for different pipe diameters and types.

Size [in.]	Type	f_{size}	$C_{coating}$ [USD/m]	DEH [USD/m]
4	Rigid	0.15	150	440
4	Flexible	0.50	150	-
6.625	Flexible	1	-	-
8	Flexible	1.10	-	-
10	Rigid	1	360	436
10	Flexible	1.70	440	-
12	Rigid	1.20	400	440
16	Rigid	1.60	480	440
20	Rigid	2.20	590	440

Data and estimated costs for the sets of flowlines and risers for the four studied cases are given in the tables below.

Table B.5: Overview of data and cost estimations for flowlines and risers for cases 1 and 2.

	Description	**Type**	**Size [in.]**	**Length [m]**	**Cost [USD]**
Case 1&2	Gas flowline	Rigid	4	150 000	93 675 000
Case 1&2	Oil flowline	Rigid	8	150 000	129 165 000
Case 1&2	Gas riser	Flexible	4	510	714 000
Case 1&2	Oil riser	Flexible	8	510	1 438 200
Total Cost					**236 838 195**

Table B.6: Overview of data and cost estimations for flowlines and risers for cases 3 and 4.

	Description	**Type**	**Size [in.]**	**Length [m]**	**Cost [USD]**
Case 3&4	Flowline	Rigid	10	150 000	154 500 000
Case 3&4	Riser	Flexible	10	510	2 177 700
Total Cost					**164 926 889**

B.4 Separators and Desander

Cost of a horizontal pressure vessel in carbon steel is a function of shell mass. The cost is on an US Gulf Coast 2007 basis [51].

$$C_0 = a + bS^n = 8800 + 27m_{shell}^{0.85} \tag{B.4}$$

The corresponding relation for a vertical separator is;

$$C_0 = 10000 + 29m_{shell}^{0.85} \tag{B.5}$$

The final installed cost is obtained by the use of several installation factors.

$$C = C_0 f_{sub} f_{inst} \tag{B.6}$$

Estimated costs for all pressure vessels is given in the table below.

Table B.7: Pressure vessel data, installation factors and estimated costs. The final cost is the 2014 cost including engineering, design, material, piping, installation, electrical work, instrumentation and control, coating and structures.

Pressure vessel	**Shell mass [kg]**	$\mathbf{f}_{sub}$	$\mathbf{f}_{inst}$	$\mathbf{I}_{2014}/\mathbf{I}_{2007}$	**Installed C, 2014 [USD]**
4-phase separator, All cases	22 816	3	4.208	1.1035	675 795
Desander, All cases	1 274	3	4.208	1.1035	63 360
Topsides separator, Case 3&4	14 082	3	4.208	1.1035	462 183

B.5 Compressors

The cost of a centrifugal compressor in carbon steel is a function of the driver power [51].

$$C_0 = 490000 + 16800P_{comp}^{0.6} \tag{B.7}$$

The power consumption, installation factors and the total installed cost of 2014 for all compressors are shown in table B.8. Spare equipment is not installed, and is adjusted only for subsea construction, engineering and design, material and production year.

Table B.8: Compressor data, installation factors and estimated costs. The final cost is the 2014 cost including engineering, design, material, piping, installation, electrical work, instrumentation and control, coating and structures.

Compressor	**Power [kW]**	$\mathbf{f}_{sub}$	$\mathbf{f}_{inst}$	$\mathbf{I}_{2014}/\mathbf{I}_{2007}$	**Installed C, 2014 [USD]**
Compressor, Case 1	280.8	3	4.208	1.1035	4 572 678
Spare compressor, Case 1	280.8	3	1.56	1.1035	1 695 194
Compressor, Case 2	1135.3	3	4.208	1.1035	22 761 555
Spare Compressor, Case 2	1135.5	3	1.56	1.1035	8 438 219
Compressor, Case 4	1050.4	3	4.208	1.1035	22 035 465
Spare compressor, Case 4	1050.4	3	1.56	1.1035	8 169 041

B.6 Pumps

The cost of a single-stage centrifugal pump (single phase pump) is a function of the liquid flowrate (in liters per second).

$$C_0 = 6900 + 206q_l^{0.6} \tag{B.8}$$

The cost of the motor is given by the following relation;

$$C_0 = -950 + 1770P_{pump}^{0.6} \tag{B.9}$$

, where P_{pump} is the power consumed by the pump.

Multiphase pumps are relatively new on the market, and there is currently no existing cost relations for these. Extracting information from the different projects in Table 7.1 give an approximate fixed cost for a multiphase pump module of 10 mill. USD without umbilicals, power- and control system and installation. Including these costs, the installed multiphase pumps is approximately 16 mill. USD (see Chapter B.1). Data, installation factors and cost for all single phase pumps are shown in the table below.

Table B.9: Data, installation factors and cost for single phase pump all pumps. The final cost is 2014 cost including engineering, design, material, piping, installation, electrical work, instrumentation and control, coating and structures

Pump	**Flow [L/s]**	**Power [kW]**	$\mathbf{f}_{sub}$	$\mathbf{f}_{inst}$	$\mathbf{I}_{2014}/\mathbf{I}_{2007}$	**C, 2014 [USD]**
Oil pump, Case 1	70.33	2 344	3	4.208	1.1035	2 808 586
Spare oil pump, Case 1	70.33	2 344	3	1.560	1.1035	1 041 206
Oil pump, Case 2	68.98	2 715	3	4.208	1.1035	3 045 354
Spare oil pump, Case 2	68.98	2 715	3	1.560	1.1035	1 128 981
Oil pump, Case 4	68.98	1 719	3	4.208	1.1035	2 365 922
Spare oil pump, Case 4	68.98	1 719	3	1.560	1.1035	877 100
PW pump, All cases	10.39	82.60	3	4.208	1.1035	454 942
Spare PW pump, Case 1	10.39	82.60	3	1.560	1.1035	168 657

B.7 Umbilicals and Power Cables

The umbilicals and Power Cables are cost estimated with a price per length constant (C_l) extracted from the project data in table 7.1.

$$C = C_l \cdot L \tag{B.10}$$

The extracted price per length constant (for the project year), the cable length, and the final installed equipment cost of 2014 adjusted for price changes over time, for both the umbilical cable and the high voltage power cable, are shown in Table B.10.

Table B.10: Data and costs for the umbilical and the power cable.

Cable	$\mathbf{C}_l$ **[USD/m]**	**Length [m]**	$\mathbf{I}_{2014}/\mathbf{I}_{2006}$	$\mathbf{I}_{2014}/\mathbf{I}_{2008}$	**C, 2014 [USD]**
Umbilical	125	150 000	-	1.0076	18 893 379
Power Cable	375	150 000	1.1605	-	65 279 724

B.8 Hydrocyclone

The hydrocyclone is cost estimated using historic cost data [56]

$$C_0 = C_{ref} \left(\frac{S}{S_{ref}} \right)^n \tag{B.11}$$

, where S is a size parameter.

For a hydrocyclone, the size parameter is inlet flowrate given in L/s. Inserted for the reference case;

$$C_0 = 38000 \left(\frac{q_l}{50} \right)^{0.35} \tag{B.12}$$

The reference hydrocyclone is in carbon steel, and the cost is given for a CEPCI of 1 000. In 2014, CEPCI was 579.8 [54]. The final cost of the hydrocyclone is given in the table below.

Table B.11: Cost for the subsea hydrocyclone. The final cost is the 2014 cost including engineering, design, material, piping, installation, coating and structures.

Unit	Flow [L/s]	f_{sub}	f_{inst}	I_{2014}/I_x	C, 2014 [USD]
Hydrocyclone	97.58	3	4.208	0.5798	351 480

C Profitability Calculations

C.1 After Tax Cash Flows

The net income of period i (NI_i) is given by;

$$NI_i = R_i - CAPEX_i - OPEX_i \tag{C.1}$$

, where R_i is the revenues of period i, $CAPEX_i$ is the capital expenditures of period i and $OPEX_i$ is the operating expenditures of period i. [51]

The after tax cash flow for period i before tax is given by;

$$CF_i = NI_i(1 - t_r) \tag{C.2}$$

, where t_r is the rate of taxation.

Capital expendiures are assumed to be a one-time investment in year 0. Values for CAPEX and OPEX for each year are given in chapter 7.1 and 7.2, respectively. The revenues are calculated from sales income of oil and gas. The oil and gas prices used in the calculations are stated in the the Design Basis chapter (3), and are 58.6 USD/bbl and 2.56 USD/MMBtu for oil and gas, respectively. The calculated annual revenues are shown in Table C.1.

Table C.1: Annual production and income data of oil and gas for early and late production, and calculated annual revenues.

i	Oil prod. [bbl]	Oil Income [USD]	Gas prod. [MMBtu]	Gas income [USD]	R_i [USD]
Early	12 439 104	728 931 494	4 073 026	10 426 946	739 358 440
Late	741 717	43 464 604	139 711	357 661	43 822 265

The net income of period i is calculated from Equation C.1, and the net income is used to calculate the after tax cash flow of period i from Equation C.2. The data used and the resulting after tax cash flows for early and late production are shown in the table below.

Table C.2: Net income, rate of taxation, and the after tax cash flows for the cases of early and late production.

i	NI_i [USD]	t_r [%]	CF_i [USD]
Early	721 547 262	35	469 005 720
Late	26 435 801	35	17 183 271

C.2 Net Present Value (NPV)

For a given discount rate (or cost of capital) r, the Net Present Value (NPV) of a project is the sum of the discounted cash flows from each year in the projects economical lifetime. [51]

$$NPV = \sum_{i=0}^{t} \frac{CF_i}{(1+r)^i} \tag{C.3}$$

i denotes the period and t is the economical lifetime.

The discount rate is often chosen as the opportunity cost of capital, or the expected return if the capital was invested in another project with comparable size and risk. Projects with positive NPV will result in increased wealth and should be executed.

Based on the after tax cash flows shown in Table C.2, and a discount rate of 10%, the project NPV was calculated to be 1.897 bill. USD by the use of spreadsheet calculations.

C.3 Internal Rate of Return (IRR)

The Internal Rate of Return (IRR) is the discount rate that gives NPV = 0. Projects with IRR larger than the discount rate should be executed.

The projects internal rate of return was calculated to be 51.28% by the use of a spreadsheet solver.

C.4 Return on Investment (ROI) and Payback Time

The return on investment is the ratio between the annual net profit from the project and the initial investment in the project.

$$ROI = \frac{CF_{avg}}{I} \tag{C.4}$$

, where CF_{avg} is the average annual net profit or cash flow and I is the total investment.

Payback time is the time it takes to earn back the invested amount of capital.

$$t_{pb} = \frac{I}{CF_{avg}} \tag{C.5}$$

If the annual cash flows are uneven, another approach is used;

$$t_{pb} = t_{ln} + \frac{CF_{ln}}{CF_{fp}} \tag{C.6}$$

, where t_{ln} is the last period with negative cash flow, CF_{ln} is the absolute value of the last negative cash flow and CF_{fp} is the value of the first positive cash flow.

The results of the ROI and payback time is presented in the table below.

Table C.3: An overview of the total investment, the average annual net profit, return on investment and the resulting payback time of the plant.

I **[mill. USD]**	CF_{avg} **[mill. USD]**	**ROI [%]**	t_{pb} **[years]**
1 293 226 441	288 276 741	22.29	3.66

D Full Size HYSYS Flow Diagrams

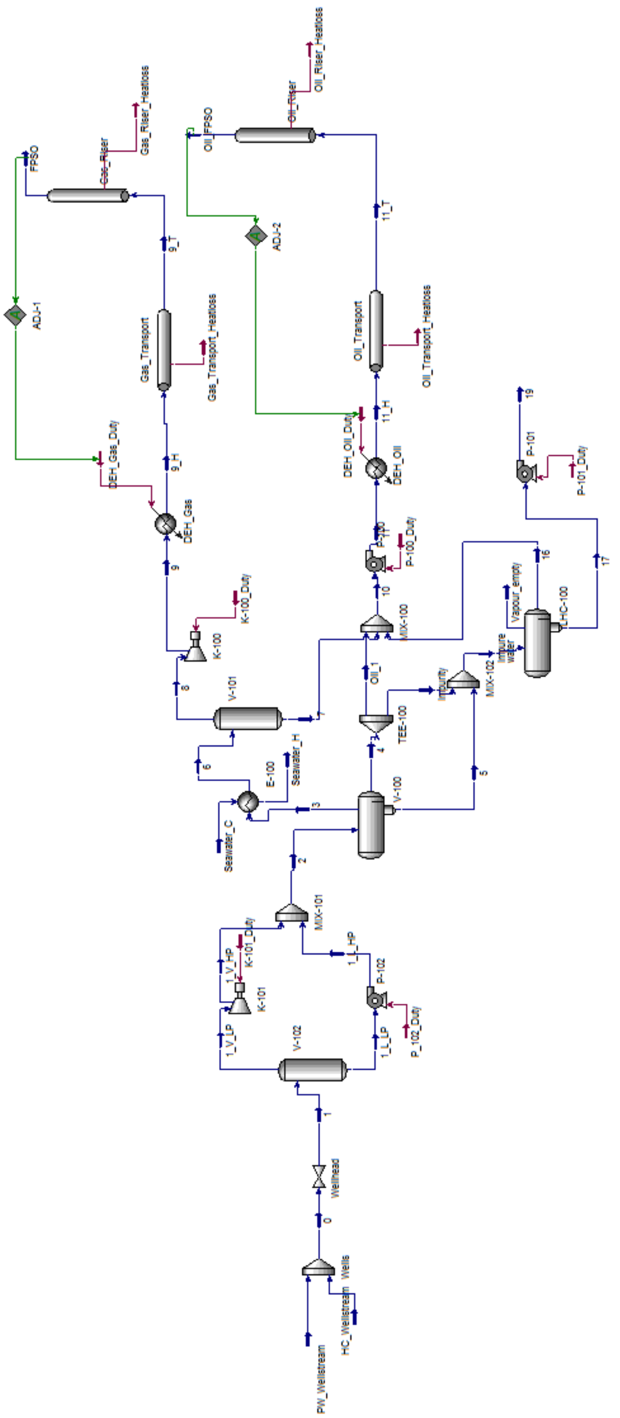

Figure D.1: Full size HYSYS flow diagram from the HYSYS model of Case 1.

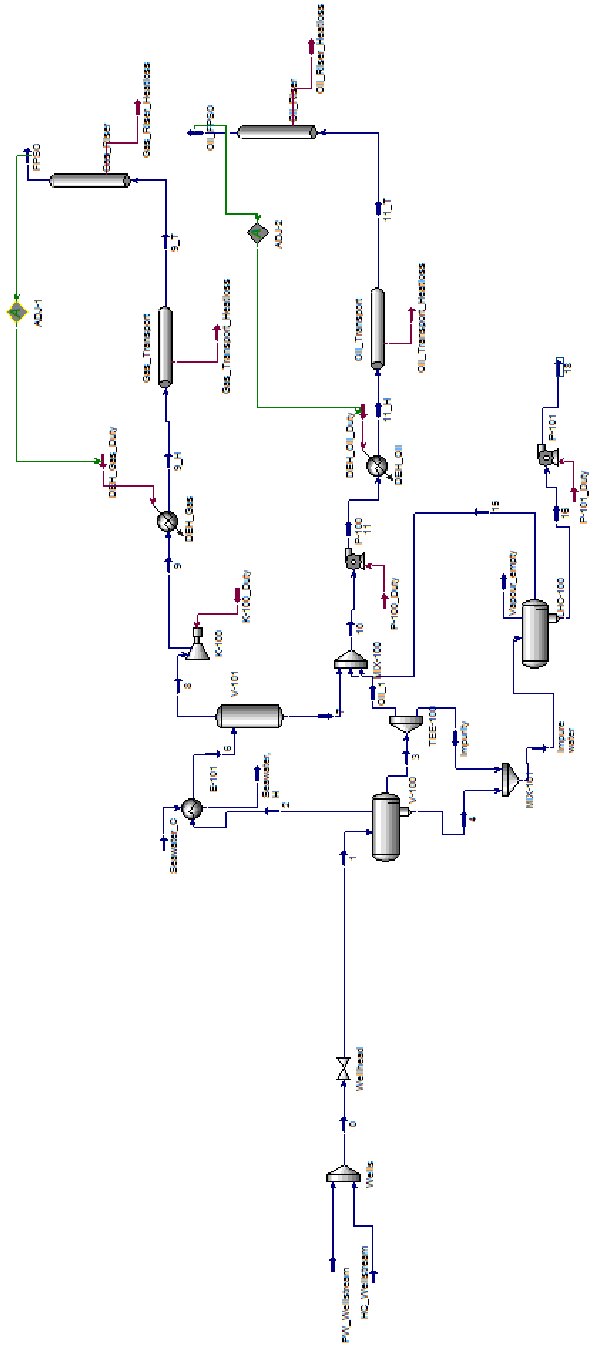

Figure D.2: Full size HYSYS flow diagram from the HYSYS model of Case 2.

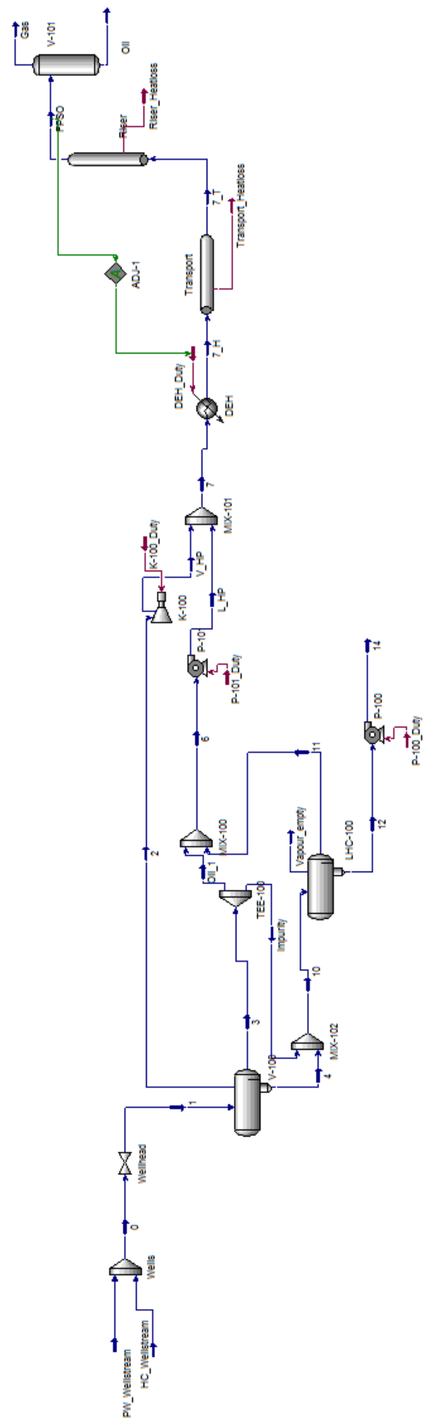

Figure D.3: Full size HYSYS flow diagram from the HYSYS model of Case 3&4.