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# Further development of solutions for subsea liquid hydrocarbon processing

Eiliv Kraabøl

Mechanical Engineering
Submission date: June 2016
Supervisor: Jostein Pettersen, EPT

Norwegian University of Science and Technology
Department of Energy and Process Engineering

NTNU
Norwegian University
of Science and Technology

Department of Energy
and Process Engineering

EPT-M-2016-71

# MASTER THESIS

for

Student Eiliv Kraabøl

Spring 2016

Further development of solutions for subsea liquid hydrocarbon processing

*Videreutvikling av løsninger for subsea prosessering av hydrokarbonvæske*

**Background and objective**

In order to realize full subsea petroleum processing solutions, gas and liquid products must be delivered in a transportable form either to standard market specifications or to specifications that can be accepted in a downstream system. This gives requirements for subsea stabilization of liquid hydrocarbon products, as well as gas processing to a rich or dry gas specification.

Earlier work by Hove (2013) and Kraabøl (2015) looked at full and partial liquid stabilization, respectively. The complexity of full stabilization makes this challenging subsea, but partial stabilization for pressurized pipeline or ship export seems feasible.

The process solutions suggested by Kraabøl (2015) were based on adaptation of a "standard" two-stage stabilization process, with various gas recompression schemes. This solution could be realized subsea, but there could still be room for simplification and optimization, for instance by use of ejectors instead of re-compressors, and by consideration of process schemes that use a common glycol system for hydrate inhibition and gas dehydration. Also, field depletion scenarios over time and realistic compressor performance data could be used to introduce more realism and field lifetime data in the modelling. Finally, adaptation of the process schemes to other feed parameters and reconsideration of the potential for full subsea stabilization can be made. Design and operation with process pressure below seafloor water pressure needs attention in some of the scenarios.

The objective of the master thesis is thus to further develop and evaluate process solutions for subsea liquid hydrocarbon processing, focusing on simplification, reduced utilities needs (power, heat, glycol), and increased flexibility of operation.

**The following tasks are to be considered:**

1. Brief review of earlier work and relevant background information, including update on further technology options that can enable or simplify subsea liquid hydrocarbon processing.

2. Definition of basis for design and analysis, including upstream and downstream assumptions, framework of study (integration with production system), equipment data and options, and general assumptions.

3. Development of system solutions, and modelling and analysis of selected processing systems to clarify performance data (product rates, power/utility needs), system complexity, and risk factors including potential operational issues.

4. Design considerations for critical equipment units, e.g. separators, compressors, heat exchangers, contactors, ejectors.

5. Evaluation of all results, data, and findings, leading to recommendations for subsea liquid processing system design and integration, review of risk factors, and identification of research and technology development needs. The evaluation may also include considerations of how the proposed solutions can be adapted to unmanned topside installation.

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

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

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

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

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

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

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

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

Department of Energy and Process Engineering, 14 January 2016

Olav Bolland

Olav Bolland
Department Head

Jostein Pettersen

Jostein Pettersen
Academic Supervisor

Research Advisor: Dr. Arne O Fredheim, Statoil

# Sammendrag

I olje- og gassbransjen er det nå en felles visjon kjent som "Alt subsea". Prosessløsninger som presenteres i denne oppgaven tar et stort skritt mot denne visjonen, og vil være en betydelig utvidelse av dagens løsninger. De er i stand til å prosessere hydrokarboner til produkter som kan transporteres og prosesseres videre i et nedstrøms prosessanlegg. Prosessløsningene inkluderer væske stabilisering, gass cricondenbar/cricondenterm kontroll, og gass dehydrering. I tillegg er det en kort evaluering av subsea varme og kraftproduksjon, som viser at det kan være mulig å spare importert kraft til prosessanlegget.

Tidligere forskning viser at delvis stabilisering av væske gir en betydelig forenkling av systemet, og en reduksjon i kraftforbruket. Dette bekreftes av analysen gjort i **Kapittel 6.1**. Et flertall av prosessløsningene i denne studien er basert på delvis stabilisering.

Videreutvikling av prosessløsninger er basert på en tradisjonell prosess med tre likevektstrinn, og en to-trinns undervannsløsning utviklet av Kraabøl (2015), se **Seksjon 4.1** og **4.2.2**. Hvilket type system som brukes for rekomprimering av gass fra stabiliseringsprosessen er funnet å ha betydelig innvirkning på systemkompleksitet. Dette har vært vektlagt ved design av prosessløsninger.

Det er gjenomført en utstyrs vurdering i **Kapittel 5**. Hvor ejektorer og skruekompressorer ble funnet egnet for rekomprimering av gas. Sentrifugal kompressorer og pumper, kan brukes for eksport av produkter.

For tørking av gas er absorpsjon, adsorpsjon, selektiv membran, og ekspansjon/kjøling evaluert. Den anbefalte løsningen er å bruke et glykol absorpsjon system. Systemet har to medstrøms glycol mixere med påfølgende separatorer, se **Seksjon 7.2**. Dette systemet er robust og gir det laveste glykolforbruket av alle systemene som er analysert. Regenerering av glycol er ikke utført subsea for å redusere kompleksiteten og strømforbruk for subsea anlegget.

Det anbefales å bruke et to-trinns system med en dobbel skrukompressor løsning for rekompremering av gas fra væske stabilsieringen. Dette systemet har høy fleksibilitet, og er i stand til å operere med både lette og tunge fødestrømmer, se **Seksjon 6.2.6**. For lette fødestrømmer, med relativt lav gassrate fra stabiliseringsprosessen,  kan et veldig enkelt, robust og kompakt system brukes. I dette systemet brukes blant annet en høytrykks separator og to ejektorer, se **Seksjon 6.2.5**. Ulempen er at dette systemet ikke kan operere med tunge hydrokarbon blandinger.

# Abstract

Subsea processing has grown into a common vision for the industry known as “All subsea”. Process solutions presented take a big step towards this vision, and will be a significant extension from today’s solutions. The systems developed are able to process hydrocarbons from the wells into products with acceptable specification for transportation and further processing in a downstream system. The processes solutions includes liquid stabilisation, gas cricondenbar/cricondenterm control, and gas dehydration. In addition there is a brief evaluation of subsea heat and power production.

Earlier research shows that partial stabilisation provides a significant simplification of the system, and a reduction in power consumption. This is confirmed by analysis presented in **Section 6.1** of this master thesis. The majority of process solutions presented in this study are therefore based on partial stabilisation.

Further development of process solutions is done based on conventional topside processing and the two-stage subsea solution developed by Kraabøl (2015), see **Section 4.1** and **4.2.2**. The configuration of the recompression system for flash gas from the stabilisation process, is found to have significant impact on system complexity, and has been in centre of design.

There is done an assessment of equipment in **Chapter 5**. Where ejectors and screw compressors where found suitable for flash gas recompression. Centrifugal compressors and pumps are found suitable for export of products.

Use of absorption, adsorption, selective membrane, and expansion/cooling is evaluated for gas dehydration. The recommended solution is to use a glycol absorption system. The system has two co-current contactors with downstream separators, see **Section 7.2**. This system is robust and provides the lowest glycol circulation rate of the systems analysed. Lean glycol is imported from a host, and exported back as rich glycol for regeneration, to reduce complexity and power consumption subsea.

In the end a system using dual screw compressors, with high operational flexibility is recommended, see **Section 6.2.6**. This system is able to operate with both lean and rich feeds, and can be also be used on an unmanned topside installation.

For lean feeds, having a relatively low flash gas rate, a very simple, robust, and compact system, including a high pressure separator and dual ejectors can be used, see **Section 6.2.5**. The drawback is that this system cannot operate with rich well streams.

# Preface

This master thesis is written at the Department of Energy and Process Engineering, at the Norwegian University of Science and Technology, with supervising from Statoil Research Centre in Trondheim, Norway.

Throughout my studies, industrial process technology with focus on oil and gas processing has been my research area. When I got an opportunity to cooperate with world leading researchers on subsea processing at Statoil Research Centre, my path for project and master thesis was set. In the pre-project work, the task was to develop a feasible subsea process for partial stabilisation of liquid. Based on the pre-project the objective of this master thesis was developed, with focus on further development of solutions for subsea hydrocarbon processing.

Working with this master thesis has been interesting, educational, and given me great insight in industrial processing in a subsea environment. It has been an honour to learn from the skilled professors here at NTNU, and to cooperate with researchers at Statoil Research Centre.

I would like to express a special thanks to my supervisor Jostein Pettersen and research advisor Dr. Arne Olav Fredheim at Statoil Research Centre, for their superior support, time and effort to guide me through this project. Your support has been remarkable and given me great knowledge about oil and gas processing, technologies and possibilities.

Thanks to my mother Gunn Gihle Riisehagen and brother Simen Kraabøl for help on proofreading of the report. Thanks to my father Eirik Otto Kraabøl and brother Torleif Kraabøl for giving me motivation and support through the whole process. Thanks to my stepfather Terje Kleppestø for support and god advices. And last but not least, a special thanks to my partner Siri Søtvik Høy for here patience and great support.

Trondheim June, 2016.

Eiliv Kraabøl

# Contents

# Nomenclature

## Definitions

| | |
|---|---|
| Completely stabilised | Refers to a liquid product from the stabilisation process with a vapour pressure less than 1 bar at 37.8°C |
| Liquid | Mixture of hydrocarbons in liquid state |
| Partial stabilised | Refers to a liquid product from the stabilisation process with a vapour pressure higher than 1 bar at 37.8°C, but less than the well stream. |
| Stabilisation | Removing volatile hydrocarbons from the liquid to get a lower vapour pressure |
| Topside installation | Offshore platform above sea level. |

## Abbreviations

| | |
|---|---|
| ASME | American Society of Mechanical Engineers |
| BM | Base model |
| BS | Bottom sediment |
| CAPEX | Capital expenditure |
| DRCS | Double recompression with scrubber |
| DWRC | Double wet gas recompression |
| EOS | Equation of state |
| GE | General electric |
| HP | High pressure |
| IP | Intermediate pressure |
| LNG | Liquified natural gas |
| LP | Low pressure |
| MEG | Mono ethylene glycol |
| n/a | Not applicable |
| OPEX | Operating expenditure |
| PFD | Process flow diagram |
| RG | Rich gas |
| SCR | Scrubber |
| SPP | Singel phase pump |
| SRCP | Single recompression and pump |
| SWRC | Single wet gas recompression |
| TVP | True vapour pressure |
| TEG | Triethylene glycol |
| UA | Overall heat transfer coefficient times area |
| W | Wat |

# 1. Introduction

## 1.1 Background

In the last decades “All subsea” has grown into a shared vision for the oil and gas industry. The goal for this vision is to remove the need for a topside facility completely, or at least reduce it to a minor unmanned host platform. Drivers for development of subsea processing is to increase recovery, increase profitability, lower the costs, and enable development of fields that earlier were left undeveloped due to technical and/or economic constraints. (Ruud, Idrac, McKensie, & Høy, 2015)

To reach the all subsea vision, there is need for a subsea system that can process gas and liquid into products which can be transported and accepted in downstream systems. This will be a significant extension of today’s subsea solutions, and development of new technology and adaption of already operating technology will most likely be required to make it feasible.

Subsea stabilisation of liquid hydrocarbons is the next step towards the all subsea vision. Hove (2013) considered full stabilisation of the liquid, which resulted in a complex system which is challenging subsea. To get a simpler system Kraabøl (2015) did research on partial stabilisation, and came up with solutions based on a two stage separation system, which seems feasible for subsea implementation. Kraabøl (2015) found that recompression was a challenging part of the system, and gas dehydration, and hydrate inhibition was only briefly discussed. Further development and evaluation of these systems is needed to enable subsea processing as a feasible alternative in field development

## 1.2 Objective

The objective of this master thesis is further development and evaluation of solutions for subsea processing of hydrocarbons, with focus on simplicity, utility need(power, heat, glycol), and operational flexibility. The subsea process shall be able to produce gas and liquid products, which have acceptable specifications for transportation and further processing in a downstream process.

## 1.3 Scope of work

This master thesis is based on processing fundamentals and earlier research of subsea processing. A literature review is done to evaluate technology and system solutions. Focus is set on simplicity, flexibility and power consumption for equipment and the overall system.

Research and evaluation of system solutions for liquid stabilisation and gas processing, including alternative solutions and improvements to earlier work is considered. Gas dehydration is evaluated. This is a technological gap in subsea processing today.

Impact of transportation method for the liquid product is evaluated by looking at partial versus full stabilisation.

In liquid stabilisation, the recompression system for flash gas is challenging due to complexity and technological constraints. Further development and evaluation of solutions for recompression is addressed. Use of equipment such as ejector and different compressor types is evaluated for subsea implementation.

## 1.4 Outline of report

**Chapter 2** provides a system overview, with a process block diagram and description of the systems considered in this study. The Framework is presented in **Chapter 3**, including motivation for subsea processing, the subsea design philosophy, and basis for analysis and design. These two chapters provide the fundament for this master thesis.

The process solutions developed, is further developments of selected processes presented in **Chapter 4**. Since these processes already is in operation, or has been evaluated in earlier research, the processes selected from the screening should provide a good foundation for further development.

To make improvements it is important to know what kind of equipment that is available and their characteristics. In **Chapter 5** an assessment of equipment is presented, where both a selection and description of equipment is given. The emphasis of the assessment is on ejectors, compressor types and pumps. This is due to the difficult recompression part of the system, and possibilities for lowering power consumption and complexity. Separator types, heaters and coolers are also addressed.

In **Chapter 6** design and analysis of processes is presented, giving a foundation for evaluation and recommendation of process solutions. Since the processes must be able to produce products with specifications acceptable for transportation, the liquid product vapour pressure will depend on transportation method. The least restricted scenario is pipeline transport (TVP<10bara), which is used for most of the system simulations in this study. An analysis is done to see the impact of lowered vapour pressure on the process design and process parameters. Due to the difficult recompression part this is the main focus for process designs and equipment selections. In **Chapter 7** there are analysed solutions for gas dehydration, with emphasis on finding a simple system with low glycol circulation rate.

To see if there is possible to produce power and heat subsea, some solutions are presented in **Appendix A**. If heat is needed in the process system, producing this subsea without electrical power will have direct impact on the power consumption of the plant.

To connect the dots throughout the master thesis a discussion and evaluation is done in **Chapter 8.** This discussion provides the foundation for recommendations given in the conclusion in **Chapter 9**.

# 2. System overview

The first step in oil and gas processing is to make transportable products that can be accepted in a downstream process plant. The minimum number of systems needed is shown in **Figure 2.1**. This section gives a more detailed description of the process plant needed to achieve required product specifications set in **Table 3.5** and **Table 3.6**.

Subsea processing of hydrocarbons to products which can be transported and accepted in a downstream process plant is a complex system. To get an overview of the systems evaluated in this study, the process plant is divided into smaller subsystems as shown in the process block diagram in **Figure 2.1**. These systems are described in the following section of this chapter. Be aware that field specific factors, like impurities($CO_2$, $H_2S$, mercury, etc.) in the hydrocarbon mixture  can make need for additional systems.

In **Figure 2.1** systems evaluated are placed inside the large rectangle. Utilities provided from a host facility are placed in the small rectangle. The utilities are used for different parts of the subsea plant to provide power, control and glycol to wherever it is needed in the process. Subsea systems for heat and power production are evaluated to lower the need of imported power.

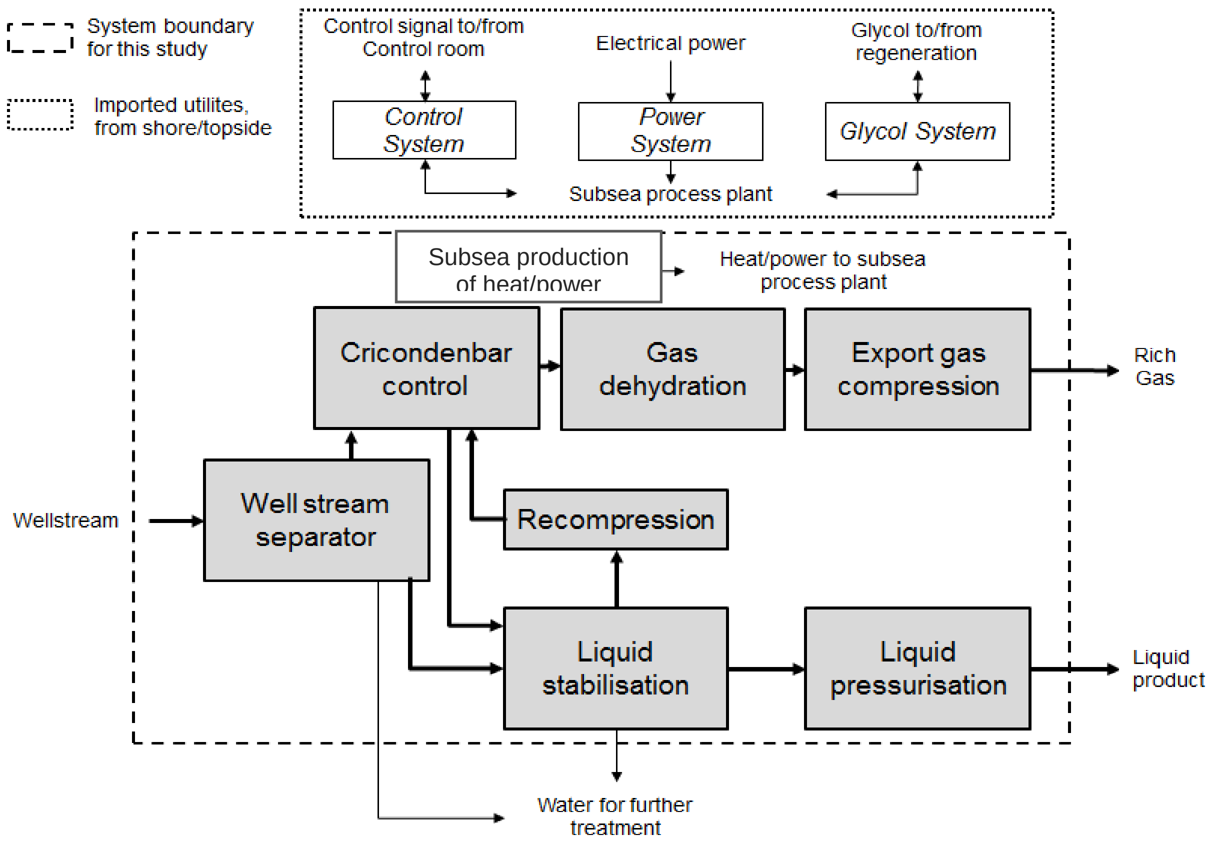



**Figure 2.1 Process block diagram**

Produced water is sent for treatment either subsea or topside before disposal or reinjection to the field. Water treatment system is not evaluated in the study as this technology already is available subsea.

## 2.1 Wells

Hydrocarbons flow from the wells through the wellhead and further on to the process plant. This flow is driven by the pressure difference between the wells and the process plant inlet.

Hydrocarbons are coming from the wells as multiphase flow, which can give challenges with flow assurance in the production pipeline, and flow rate variations and flow pattern entering the process plant. Flow assurance problems, like hydrate formation and wax deposition, can be avoided by heating, insulation, injection of chemicals or other technical solutions. Action needed depends on factors such as well stream composition, temperature, pressure, and distance from the well to the process plant. The wells are mature subsea technology and is not further evaluated in this study.

## 2.2 Inlet separator (Well stream separator)

Controlling the flow into the process plant can be done by using an inlet separator which break the flow pattern and separate the flow. Separation of the well stream into gas, liquid and free water provides single phase flow for further processing. Hydrocarbon liquid is sent to stabilisation, lowering the liquids vapour pressure to an acceptable value for transportation. Flash gas from the stabilisation process need to be recompressed and mixed with the gas stream.

The inlet separator must be able to break momentum and have capacity to handle liquid slugs coming from the pipeline. The inlet separator is also used as a buffer in the process to assure steady flow through the rest of the system. Pressure in the inlet separator can be set at the same pressure as used in the cricondenbar control. This will be the highest pressure possible upstream the export compressor and gives maximum flash gas at high pressure lowering power needed for recompression.

## 2.3 Gas processing

The gas needs systems for cricondenbar control, gas dehydration and export compression to achieve rich gas specifications. In addition systems for gas sweetening and mercury removal can be needed, but in this study it is assumed that the gas is within rich gas specifications without removing sour gases, mercury or other impurities.

Cricondenbar and cricondenterm specification can be reached by lowering the pressure and/or temperature of the gas, condensing heavy hydrocarbons and separating them out. This gives a cricondenbar and cricondenterm that is acceptable for rich gas pipeline transport, assuring that no hydrocarbon liquid will condense out in the transportation to shore.

The gas is saturated with water which will condense and provide an environment for hydrate formation, so keeping the temperature above the hydrate curve is important. Use of hydrate inhibitor will move the hydrate curve towards lower temperatures, but this will then be lost to the water phase or it will need to be recovered onshore or topside. Hydrate inhibitor will allow lower temperature and a higher pressure in the cricondenbar control, lowering power need in the export compressor. In this study it is assumed that no hydrate inhibitor is used, and the temperature is kept above 25°C to prevent hydrate formation, this is found to be at least 5°C above the hydrate formation curve, see **Figure 3.4**.

The wet gas need to be dehydrated, lowering the water dew point to a level where no water is condensed out during transportation. Dehydration of gas can be done in a glycol absorption process. It is assumed that the glycol is regenerated at a host facility, to avoid complex and energy intensive systems subsea. Glycol can also be used for hydrate inhibition, but if it ends up in the produced water, environmental concerns for disposal, expenses for regeneration and/or replacement of lost glycol must be considered.

When the gas is within rich gas transport specifications, see **Table 3.6**, it is pressurised and transported by pipeline to a downstream process facility.

## 2.4 Liquid stabilisation

Hydrocarbon liquid needs a system to lower the vapour pressure to a specification acceptable for transport by ship or pipeline. Lowering the vapour pressure can be done in a stabilisation process where light hydrocarbons are separated from the liquid. This can be done by lowering the pressure and/or heating the liquid to higher temperatures, with other words moving towards more vapour inside the phase envelope and then separating this mixture of gas and liquid. Free water can also be taken out in this system if needed. After stabilisation the liquid can be pressurised for transport or subsea storage.

## 2.5 Treatment of produced water

Systems for treatment of produced water are already in operation subsea and are only briefly discussed in this study. If produced water contains large amounts of glycol it is assumed that treatment is unavailable subsea.

## 2.6 Utilities (Power and glycol supply)

Utilities necessary to run the process plant can be imported from shore or a nearby topside facility. Imported utilities will typically be lean glycol and electrical power. It is also systems available and under development for power production offshore that can be an alternative for the imported power. Heat is so far only produced in electrical heaters subsea. Therefore reduction in heat requirement is important for power consumption in a subsea system. Alternative heating systems is evaluated to see if there are solutions that can lower the power consumption.

# 3. Framework

## 3.1 Motivation for subsea processing

In the last decade there has been a significant development of subsea processing. The oil and gas industry has now great belief in subsea processing, and a shared vision of “All subsea” is established. The “All subsea” vision is moving production towards a full subsea process plant, where hydrocarbons are processed directly from the well to the market. Drivers for the vision is factors like maximize recovery, reducing CAPEX and OPEX, and enabling development of fields that would have been left undeveloped due to technological and/or economic constraints. In addition development and subsea implementation of new technology is increasing operational possibilities subsea, see **Figure 3.1** and **Appendix F** for status on subsea processing. (Ruud, Idrac, McKensie, & Høy, 2015)

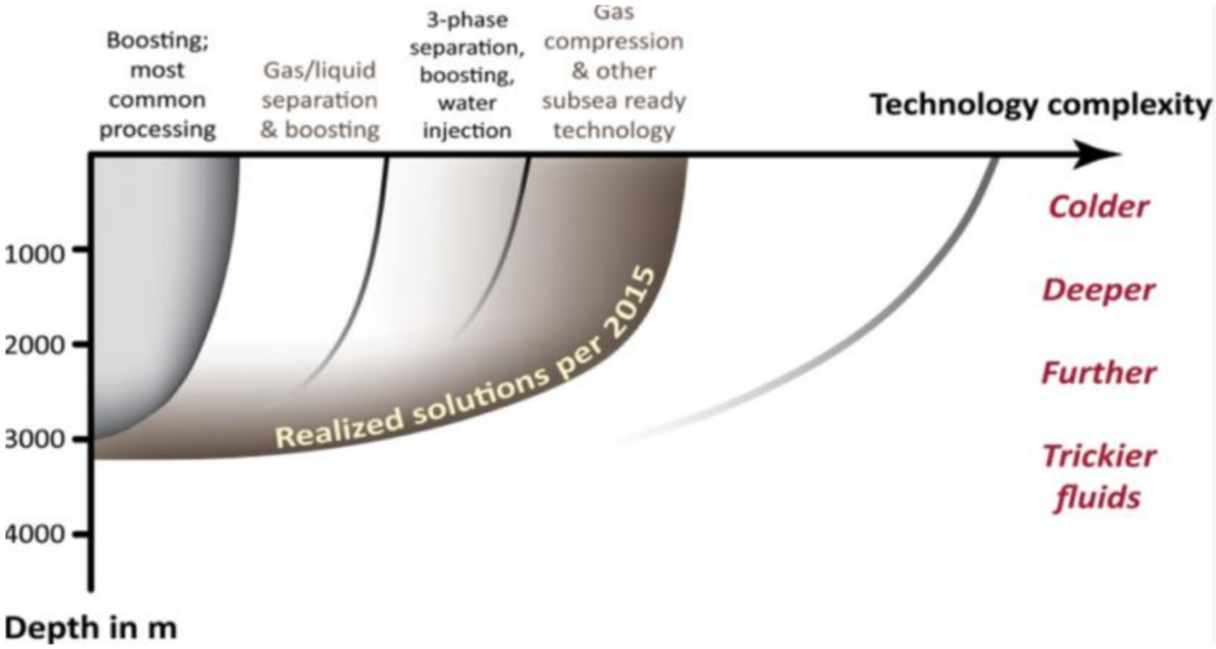



**Figure 3.1 Subsea processing solutions** (Ruud, Idrac, McKensie, & Høy, 2015)

Other factors driving development on subsea processing is that production is moving towards deeper water and colder climate. In this scenarios use of conventional topside installations is less attractive. In cold climates a topside installation must be winterised to meet low temperatures and icing. Winterisation will increase CAPEX compared to a conventional topside installation, making subsea processing more attractive.

In special for deep water, recovery will increase with use of subsea processing since the well can produce against the back pressure at the sea floor. As shown in **Figure 3.2** subsea boosting can increase both plateau production and tail production for the field increasing the total recovery.

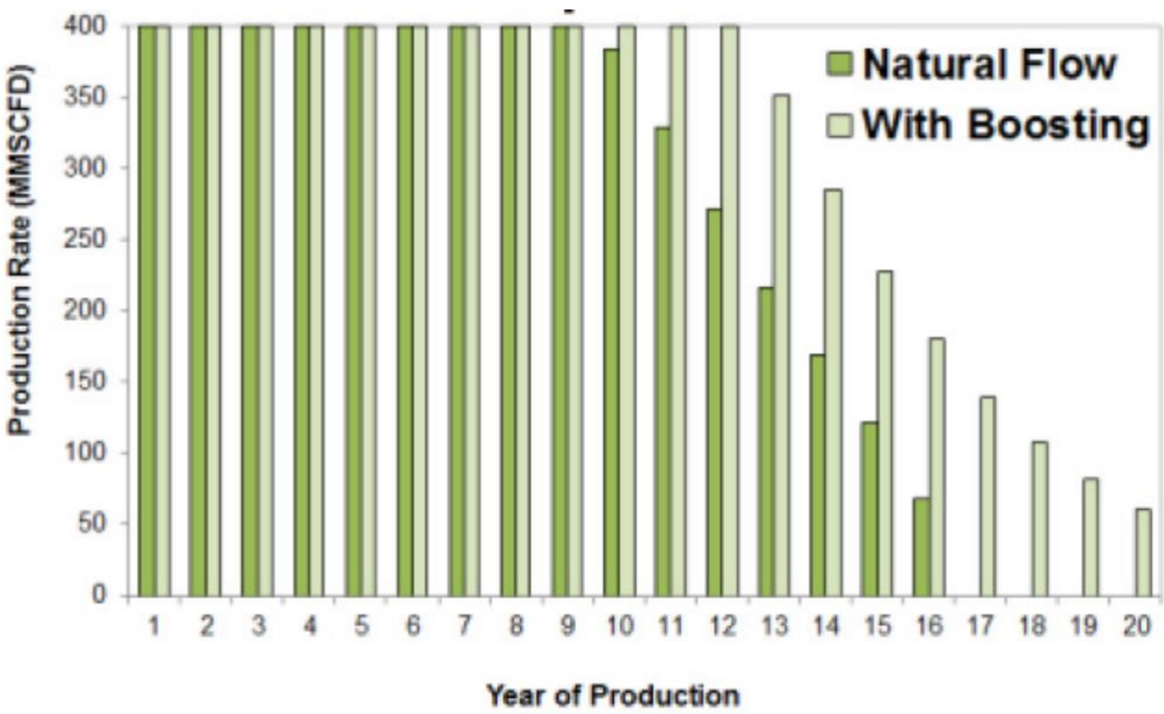



**Figure 3.2 Production profile with Natural flow versus Subsea Boosting**

(Sandy & Hasan, 2016)

## 3.2 Subsea design philosophy

In subsea processing simplicity and robustness is important to get a low maintenance system. Subsea maintenance is expensive and complicated. Putting effort into low maintenance design will be pay off by lower OPEX and decreased downtime for the plant. Use of equipment with moving parts should be held on a minimum, as this is complex equipment, which is challenging and expensive to provide a low maintenance design.  Equipment that is sensitive to impurities such as sand and fouling should be avoided. But if it gives clear advantages, effort on equipment design to handle impurities should be considered.

Installation is challenging subsea and a ship will be needed for transportation and installation of equipment. This can be made easier by building modules that can be connected on the seabed. Lowering size and weight for equipment and the total module will lower requirements of the installation vessel, which should provide lower installation costs.

Heat and power is not produced subsea at this point, meaning that all heat input will be produced by electrical heating if no alternative heating is developed. Lowering the need for heat input, consider heat integration and use of alternative heat sources has direct impact on power consumption. Putting effort on lowering the total power demand will lower CAPEX and OPEX for the plant.

Choosing technology already in operation subsea, will lower development costs and provide important know how and improve reliability. If new technology is needed it will be preferred to adapt mature technology used offshore.

Other challenges in subsea processing is commissioning, hydrate prevention, leakages to the surroundings, and negative pressure difference that can give leaks into the system.

In development of subsea processing some of the most important factors are:

- Robustness and reliability to avoid downtime and increase operational safety.
- Maintenance needs should be low, as it most likely needs to be done automatically or by a remotely operated vehicle.
- Complexity should be as low as possible to limit the number of fail sources and increase operational control.
- Power consumption, keeping the power consumption low will lower installation and operational costs.
- Compactness is important for module weight and size.
- Operational flexibility to handle variation in process parameters.
- Water resistant, as it will be surrounded with water.
- Maturity of the technology affects development costs and reliability, see **Figure 4.2** for maturity status for subsea processing technologies.

## 3.3 Basis for analysis and design

General assumptions used for calculations, simulations and evaluation is stated in **Table 3.1**

**Table 3.1 Assumptions**

| Designation and unit | Specification |
|---|---|
| Sea water temperature (°C) | 5 |
| Pressure drop in simulated equipment | 0 |
| Adiabatic efficiency for pumps (%) | 75 |
| Polytropic efficiency for compressors (%) | 75 |
| Maximum pressure ratio for compressors and ejectors | 4 |
| Maximum temperature in compressors (°C) | 130 |
| Allowed internal pressure in vessel type separators (bara) | 100 |
| Minimum temperature difference in heat exchangers (°C) | 10 |
| Minimum temperature allowed in the system (°C) | 25 |
| Overall heat transfer coefficient, U ($W/m^2K$) based on active coolers (Pettersen, 2016), | 800 |

### *3.3.1 Equation of state*

Peng Robinson (PR) is the equation of state used for HYSYS simulations. The HYSYS Peng Robinson is modified compared to the original equation, it has temperature range down to -271°C and pressure range up to 100000 kPa, which is applicable for this study. (HYSYS, 8.6)

To predict water dew point the Kabadi Danner (SRK-KD) is applied. This is an improvement of the Soave-Redlich-Kwong (SRK) equation to get better equilibrium calculations in water-hydrocarbon systems, in special for dilute mixtures. SRK-KD is found to be in compliance with the water prediction diagram shown in **Figure E.7**, **Appendix E**, for dilute mixtures.

For dehydration processes with TEG, the glycol package in HYSYS is applied. The glycol package is tuned for dehydration of natural gas with TEG, and is based on the Twu-Sim-Tassone (TST) equation (HYSYS, 8.6).

### *3.3.2 Feed*

In **Table 3.2** the well stream composition used for simulations is shown on a dry basis (without water content). The real composition is saturated with water and in addition the well produces a free water stream of 100 $m^3$ per day.

**Table 3.2 Dry well stream composition (Pettersen, 2016)**

| Component | Lean case (mol %) | Rich case (mol %) |
|---|---|---|
| Methane | 88.4 | 71.5 |
| Ethane | 5.9 | 9.0 |
| Propane | 1.6 | 4.4 |
| i-Butane | 0.4 | 0.8 |
| n-Butane | 0.4 | 1.7 |
| i-Pentane | 0.1 | 0.7 |
| n-Pentane | 0.1 | 0.8 |
| n-Hexane | 0.2 | 1.0 |
| n-Heptane | - | 1.2 |
| n-Octane | - | 1.3 |
| n-Nonane | - | 0.9 |
| C7+ | 0.5 | - |
| C10+ | - | 3.9 |
| Nitrogen | 0.5 | 0.3 |
| CO2 | 1.9 | 2.5 |
| H2O | - | - |
| **Note:** | C7+ is a hypothetical component with Molecular weight 130 kg/kmol and ideal liquid density 803 kg/$m^3$<br>C10+ is a hypothetical component with Molecular weight 223 kg/kmol and ideal liquid density 830 kg/m3 | |

Phase envelopes and hydrate curves for the well stream composition is shown in **Figure 3.3**. The rich case has a clearly larger phase envelope than the lean case, due to the larger fraction of heavy hydrocarbons in the rich feed. In **Figure 3.4** hydrate curves is presented. Hydrates will form to the right of these lines, so operating the plant above the hydrate curve will prevent hydrate formation. If hydrate inhibitor is used the line is moved to the left, making it possible to operate at lower temperatures without hydrate formation.

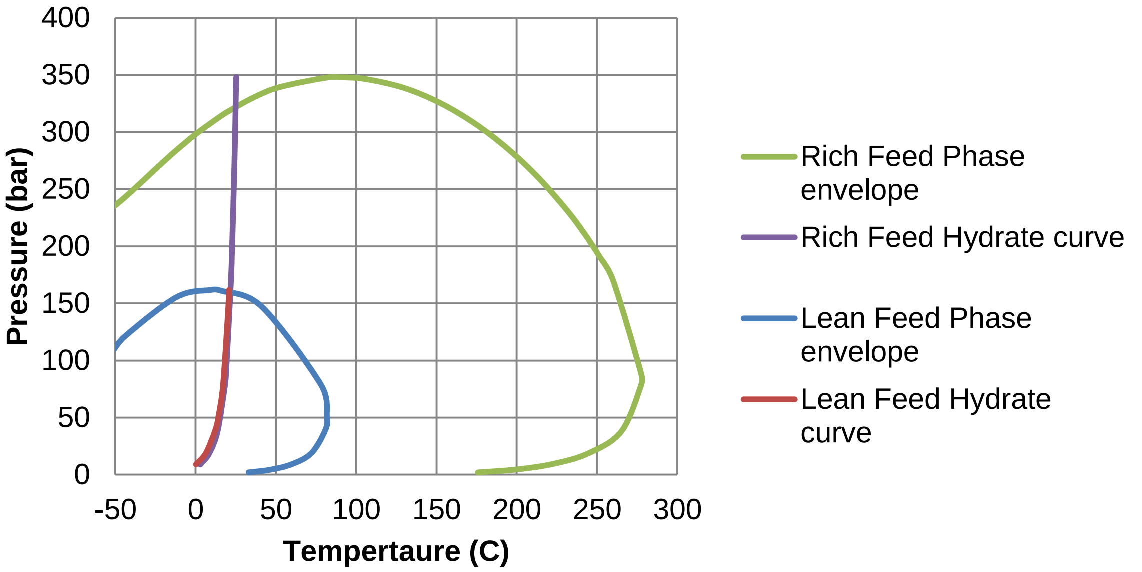



**Figure 3.3 Feed phase envelopes and hydrate curves**

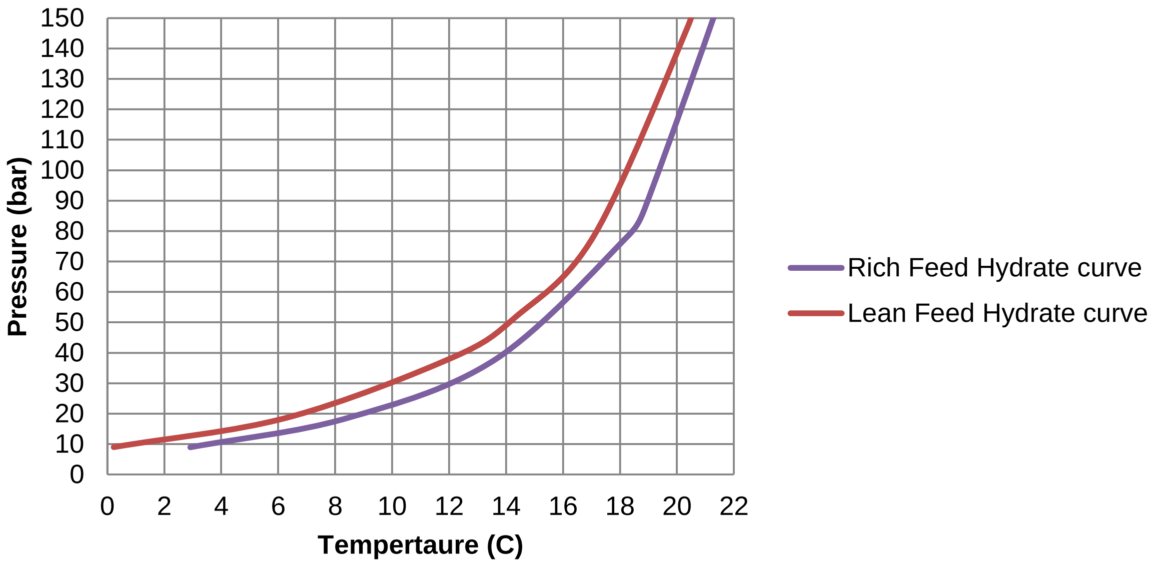



**Figure 3.4 Hydrate Curves**

In this study the same volumetric flow rate at 1atm and 15°C is used for both lean and rich feed. **Table 3.3** shows the conversion to mass flow, which is used in simulations.

**Table 3.3 Well flow rate on dry basis**

| Designation and unit | Lean mass flow | Rich mass flow |
|---|---|---|
| Actual volume flow rate at 1atm and 15°C ($Sm^3/d$) | $3*10^6$ | $3*10^6$ |
| Flow rate assumed ideal gas at 1atm and 15°C ($Sm^3/d$) | 3 014 765 | 3 252 476 |
| Mass flow plotted in HYSYS (tonne/d) | 2431.6 | 4572.6 |
| Mass flow (kg/s) | 28,14 | 52,9 |
| Actual density at 1 atm, 15°C ($kg/Sm^3$) | 0.8105 | 1,5242 |
| Molar weight (kg/kmol) | 19.07 | 33.23 |

#### *3.3.2.1 Adding water in HYSYS simulations*

The system used for adding water to the dry feed stream in HYSYS simulations is shown in **Figure 3.5**. The dry feed enters a saturation chamber, saturating the well stream. When the feed is saturated with water an additional 100 $m^3$ per day of water is added in a water mixer. The wet feed downstream the mixer is then saturated with water and in addition it has a free water stream of 100 $m^3$ per day.

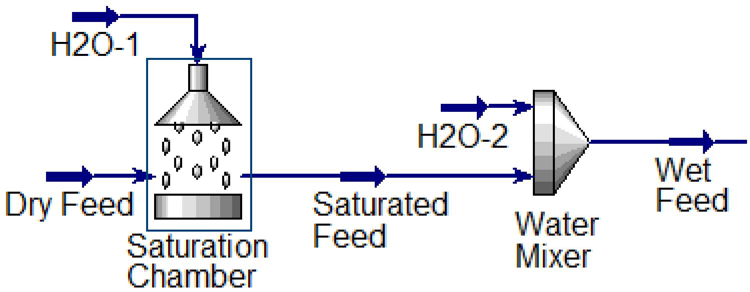



**Figure 3.5 System used for adding water to a dry feed stream in HYSYS**

### 3.3.3 Products

Valuable products are liquid hydrocarbons and gas which need to be transported to shore for further processing. Transportation method and downstream processing sets specifications for the products. In addition there will be produced water that needs treatment before reinjected or disposed to the sea. If glycol is used in the process this need to be regenerated topside or on shore.

#### 3.3.3.1 Liquid product

Hydrocarbon liquid can be transported by a pipeline, or stored and transported by ship to shore. Choice of transportation method depends on factors like existing infrastructure, liquid production rate, tanker availability and subsea processing.

Transportation by ships is traditionally done in large tankers without pressurisation or cooling, this transport need full liquid stabilised to a TVP less than 1 bara. To reach full stabilisation low pressure or high temperatures will be needed to remove light hydrocarbons in the subsea process. Using partial stabilisation will put less restriction on the subsea process, and the liquid product can be transported in semi-pressurised tankers or by pipeline.

Semi-pressurised tankers are normally used for LPG transport, but are also used for transportation of $CO_2$ (Seamanship, 2012) (IPCC, 2005). Data for semi-pressurised ships operating at two different temperatures is found in **Table 3.4**

**Table 3.4 Semi-Pressurised ships (Seamanship, 2012)**

| Designation and unit | Semi-pressurised Fully refrigirated | Semi-pressurised Semi-refrigerated |
|---|---|---|
| Transportation pressure (bara) | 5-8 | 5-8 |
| Minimum transport temperature (°C) | -48 | -10 |
| Capacity for ships in operation ($m^3$) | 15000 | 5000 |

Transportation by pipeline can be done at higher pressure than ship transport, giving even less restriction on TVP. No cooling will be needed as the transport is taken place subsea at the same temperature as the surrounding sea water. In this study the specification is set from the Norpipe oil pipeline, with TVP<10 (Pettersen, 2016).

Other specifications for the liquid product is water content, temperature and pressure. Water has no value and is not wanted in a downstream process. In **Table 3.5** Specifications used for the liquid product is stated based on full stabilisation, Norpipe oil pipeline and semi-pressurised ship transport.

**Table 3.5 Liquid product specifications**

| Designation and unit | Completly Stabilised | Stabilised for Norpipe | Stabilised for ship | Note |
|---|---|---|---|---|
| True vapour pressure at 37,8ºC (bara) | TVP<1.0 | TVP<10 | TVP<5 | 1 |
| Basic Sediment and Water (vol%) | BS&W<0.5 | BS&W<2.0 | BS&W<0.5 | |
| Export pressure (bara) | - | 100 | 41 | 2,3 |
| Note 1: TVP, which is the bubble point pressure for the fluid, is used as the vapour pressure criteria.<br>Note 3: Pressure and temperature into Norpipe oil pipeline is assumed values. | | | | |

#### *3.3.3.2 Rich gas*

On the Norwegian continental shelf most of the gas is transported in shared infrastructure owned by Gassled and operated by Gassco. The infrastructure is divided into areas with different product specifications. The most common is to process gas to rich gas specifications offshore and then transport it to shore for further processing. Rich gas specifications used in this study is taken from Åsgard Transport, Gassled area B, shown in **Table 3.6**.

**Table 3.6 Rich Gas entry specifications Area B, Åsgard Transport (Gassled, 2014)**

| Designation and unit | Specification |
|---|---|
| Gas export pressure used in this study (bara) | 200 |
| Gas export temperature used in this study (°C) | 60 |
| Maximum cricondenbar pressure (bara) | 105 |
| Maximum cricondentherm temperature (°C) | 40 |
| Maximum water dewpoint (°C at 70 bara) | -18 |
| Max. daily average glycol content (litres/MSm$^3$) | 8 |

### 3.3.4 Absorption basics and glycol types (MEG and TEG)

TEG is the most common glycol used in absorption, and can reach the water dew point specification of Rich gas, see **Figure E.8** in **Appendix E** for water dew points using TEG. TEG has also a high thermal decomposition temperature of about 206°C, see **Figure E.9** in **Appendix E**, which makes it relatively easy to regenerate to the high purity. It is found that most designs use a circulation rate of 15-14liter TEG/kg $H_2O$, which is near the economical optimum (Campbell, 1992).

In subsea processing MEG is often used for hydrate inhibition, and it would provide a huge advantage if MEG could be used for subsea gas dehydration as well. Some of the reasons why TEG is applied instead of MEG is that MEG has a low thermal dehydration temperature (165°C), making it hard to recover to high purity, higher power consumption, and more carryover due to higher gas solubility than TEG. There is systems that claims to regenerate MEG to 99,5% purity, which should be sufficient for gas dehydration (CAMERON, 2015).

Equation (1) is developed from Dalton's law and Raoult's law, the principle of this equation is significant for absorption and distillation processes, including glycol systems (Genakopolis, 2014). The equation states that increasing total pressure, and/or decreasing the pure vapour pressure, which can be done by decreasing the temperature, will decrease the mole fraction in the gas phase. In other words absorption of water is favoured by high pressure and low temperature.

$$\frac{p_i}{P} = \frac{y_i}{x_i} \tag{1}$$

Where $p_i$ (bara) vapour pressure of pure i, P (bara) is total pressure, $y_i$ is mole fraction of component i in vapour phase, $x_i$ is mole fraction of component i in the liquid phase.

If water is going to be absorbed in the glycol there need to be a high enough partial pressure of water in the gas and a low enough concentration of water in the glycol to have sufficient driving forces for mass transfer. If the mixture is given sufficient contact time it will eventually reach equilibrium. Henry's law, Equation (2), provides the basic principle to estimate equilibrium relation between partial pressure and liquid mole fraction for a given component. This relation can often be used for low concentrations. Henry's law states that the partial pressure in the gas phase is proportional to the concentration in the liquid phase. The constant in Henry's law depends on type of fluids mixed, fluid properties and temperature. (Genakopolis, 2014)

$$p_{H2O} = Hx_{H20} \quad (2)$$

Where $p_{H2O}$ (bara) is partial pressure of $H_2O$ in the gas phase, H(bara) is Henry's law constant, $x_{H20}$ is mol fraction of water in the glycol.

#### *3.3.4.1 General data used in analysis of gas dehydration systems*

In **Table 3.7** data used in simulations of dehydration processes is presented. The data is taken from HYSYS simulations done on the dual screw compressor system shown in **Figure 6.17**, but in the recompression **Section 6.2**it is seen that the different system configurations has low impact on gas production. Water content in the dry gas is estimated by use of the Kabadi Danner (SRK-KD) equation of state, as stated in **Section 3.3.1** this is assumed to provide a good dew point specification for this dilute mixture. Using the diagram for water content of sweet natural gas, **Figure E.7**, **Appendix E,** 30kg/MSm$^3$ is found for the dry gas.

**Table 3.7 General data used for simulations of dehydration processes**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Gas flow rate entering dehydration (MSm$^3$/d) | 3 | 2.9 |
| Dew point specification at 70 bara (°C) | -18 | -18 |
| Dry gas water content (kg/MSm$^3$)<br>EOS-Kabadi Danner (SRK-KD) | 28 | 29.5 |
| Wet gas water content (kg/MSm$^3$) | 465 | 441 |
| Water fraction removed (%) | 94 | 93 |
| Water flow rate removed (tonne/d) | 1.3 | 1.2 |

To estimate the needed purity of TEG the concentration chart shown in **Figure E.8**, **Appendix E**, can be used. It is found that the lean TEG concentration needs to be at least 98,8wt% when it is in equilibrium with the gas at 25°C. But since the mixer is unable to reach equilibrium 99,5wt% TEG was used in this analysis. MEG is also simulated with 99,5wt%, but also with 98,5wt% since there are concerns about how high concentration it is possible to get from the regeneration system. But there is claimed that MEG can reach 99,5wt% for some available technologies (CAMERON, 2015).

All the simulations is done with 100% separation efficiency and equilibrium mixing. This will not affect comparison between the process alternatives, which is the purpose of the evaluation. The method will have some impact on glycol purity and circulation rate, but since there are uncertainties in results from these simulations there will be need for a deeper study to predict glycol circulation rate more accurately.

### 3.3.5 Basis for ejector analysis

For a two phase ejector the key parameters in thermodynamic evaluation are ejector efficiency, suction pressure ratio and mass entrainment ratio (Hafner, Banasiak, & Andersen, 2012) (Elbel & Hrnjak, 2007). These parameters are defined in equation (3), (4) and (5) below.

Mass entrainment ratio (Elbel & Hrnjak, 2007):

$$\Phi_m = \frac{m_s}{m_M} \tag{3}$$

$\Phi_m$ mass entrainment ratio, $m_s(\frac{kg}{s})$ suction mass flow, $m_M(\frac{kg}{s})$ motive mass flow

Suction pressure ratio (Elbel & Hrnjak, 2007):

$$\Pi_S = \frac{P_s}{P_P} \tag{4}$$

$\Pi_S$ suction pressure ratio, $P_s$ $(bara)$ suction pressure, $P_P(bara)$ product pressure.

In **Figure 3.6** the ejector motive and suction flow is illustrated. Motive and suction flow expands to the mixing pressure, before both are compressed to the diffuser exit pressure. The net effect for an ejector is compression of the suction flow to the exit pressure, as shown on the right side in **Figure 3.6**.

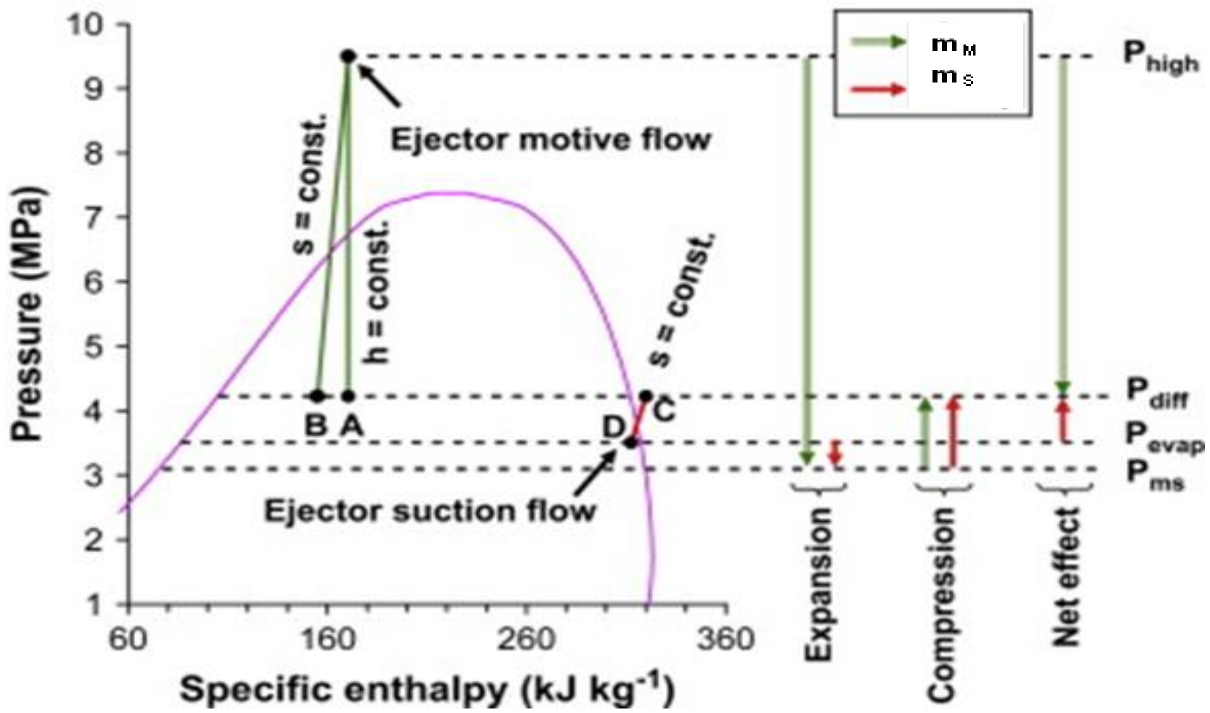



**Figure 3.6 Expansion and compression inside a two phase ejector (Elbel & Hrnjak, 2007)**

Determination of efficiency for individual parts in the ejector is difficult due to unknown parameters such as specific enthalpy in the mixing section. The efficiency defined in equation (5) is more convenient and it gives total ejector efficiency in one calculation. This definition compares actuall work recovered to pressurise the suction stream, to the theoretically maximum expansion work from expanding the motive stream. The maximum expansion work can be found by an isentropic expansion of the motive stream from the motive inlet pressure to the ejector outlet pressure.

Ejector efficiency (Elbel & Hrnjak, 2007):

$$\eta_{ejec} = \frac{W_{rec}}{W_{recmax}} \tag{5}$$

$\eta_{ejec}$ Ejector efficiency, $W_{rec}(kW)$ recovered expansion power, $W_{recmax}\,(kW)$ maximum possible expansion power.

Using **Figure 3.6** the ejector efficiency can be calculated from equation (6), see **0** for development of this equation:

$$\eta_{ejec} = \frac{m_S(h_D - h_C)}{m_M(h_A - h_B)} \tag{6}$$

$\eta_{ejec}$ Ejector efficiency, $m_s(\frac{kg}{s})$ is suction mass flow, $m_S(\frac{kg}{s})$ is motive mass flow, $h\left(\frac{kJ}{kg}\right)$ enthalpy (Point A-B-C-D is shown in **Figure 3.6**)

In this study ejector analysis is done by use of the thermodynamic method presented in **Figure 3.6**, **Section 5.1.4** . In **Figure 3.6** the motive flow goes through an expansion to the exit pressure, while the suction flow is compressed to the same exit pressure. This can be simulated in HYSYS as shown in **Figure 3.7** by using a expander on the motive flow, compressor on the suction flow and a mixer to find the exit condition. The expander and the compressor need to have the same outlet pressure and power, as the exit pressure is equal and the recoverable expansion power is used to compress the suction flow. In simulations the compression is done adiabatic while the expansion is done using the ejector efficiency as adiabatic efficiency. This gives the smallest amount of recoverable power and the adiabatic efficiency will be the same as the ejector efficiency found from equation (6) and (7).

$$m_M(h_A - h_B)\eta_{ejec} = m_S(h_D - h_C) \tag{7}$$

$\eta_{ejec}$ Ejector efficiency, $m_s(\frac{kg}{s})$ is suction mass flow, $m_S(\frac{kg}{s})$ is motive mass flow, $h\left(\frac{kJ}{kg}\right)$ enthalpy (Point A-B-C-D is shown in **Figure 3.6**)

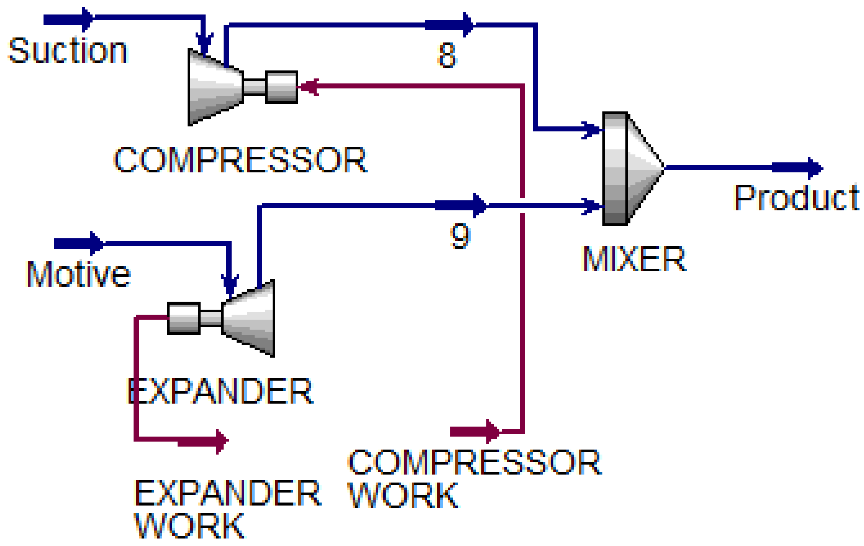



**Figure 3.7 HYSYS ejector simulation**

Realistic ejector data was provided from supervisor (Pettersen, 2016) for calibration of the HYSYS model. These data points with the efficiency found in HYSYS is presented in **Figure 3.8**. For definition of mass entrainment ratio see Equation (3), suction pressure ratio see Equation (4), ejector efficiency see Equation (6).

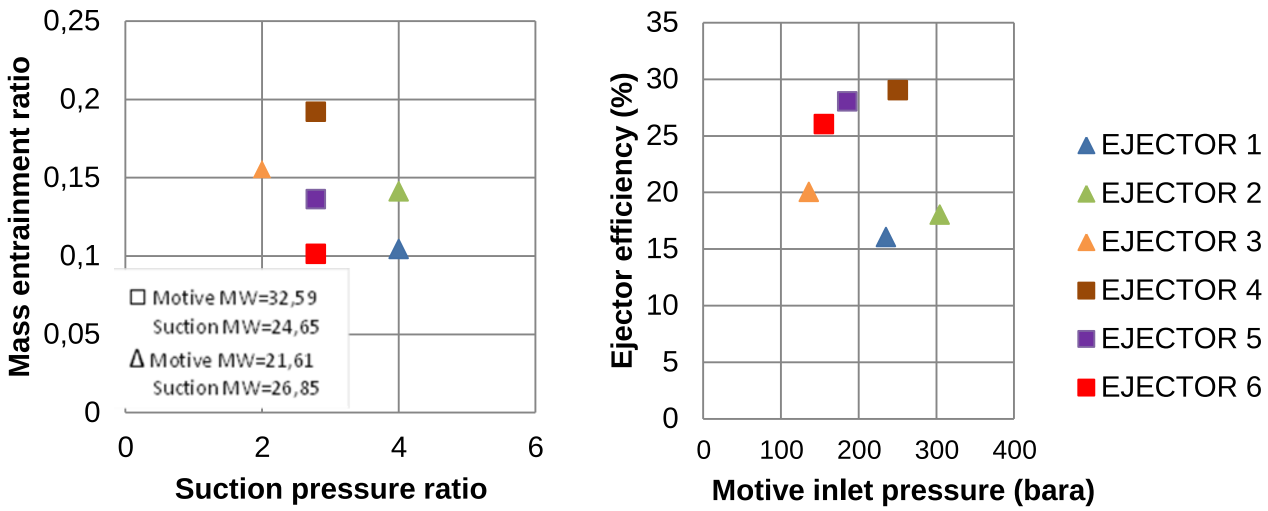



**Figure 3.8 Ejector performance based on data from supervisor** (Pettersen, 2016)

As seen in **Figure 3.8** the efficiency has some variation between 15-30%. Based on these results an ejector efficiency of 20% was chosen for this study. In addition maximum suction pressure ratio (defined in Equation (4)) is set to 4 as the ejector efficiency seems to decrease with increased suction pressure ratio.

**Table 3.8 Ejector limitations**

| Designation and unit | Specification |
|---|---|
| Ejector/expander efficiency (%) | 20 |
| Maximum suction pressure ratio | 4 |

# 4. Screening of process solutions

In this chapter process solutions found in literature and earlier research is presented. This will provide a foundation for further development of subsea processing solutions.

## 4.1 Conventional topside type process

When designing a subsea process plant adapting and simplification of conventional topside process design will provide operational experience and maturity to the process design. In the conservative oil and gas industry this approach will most likely be easier to implement than use of a new type of design without operational experience. It is expected that adapting a conventional topside process will lower development costs and provide operational reliability compared to a new process design.

One conventional three-stage process is operating at the Kristin field in the Norwegian Sea. The Kristin process PFD is shown in **Figure 4.1**. Liquid is marked with brown/dark lines and gas is marked as yellow/bright lines.

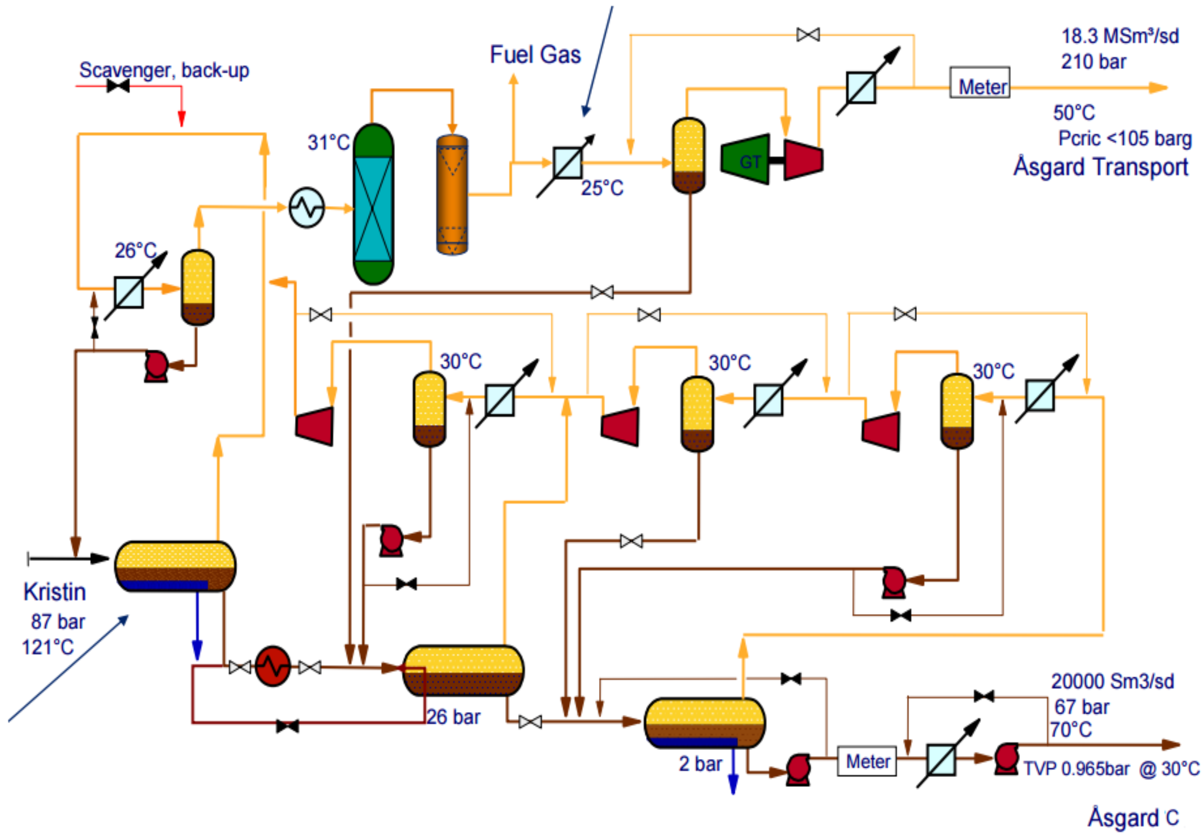



**Figure 4.1 Kristin Process PFD** (Fordal, 2005)

In the Kristin process showed in **Figure 4.1** the well stream pressure is reduced to 87bar before entering the inlet separator. In the inlet separator the bulk of gas and liquid is separated. Liquid stabilisation is done in a three stage separation process, where the pressure is reduced to lower the liquids vapour pressure by flashing gas which can be separated by gravitational separators. There is also used a heater between HP and MP separator to boil of gas, and ease separation of water and oil by reducing viscosity. Recompression of flash gas is done by use of three compressors with intercooling and separation of condensed liquid. In the LP separator the pressure is only 2 bar to reach a full stabilised liquid (TVP<1bara). Subsea this pressure will be lower than the surrounding hydrostatic pressure in most cases.

In **Figure 4.1** all the gas is lead to a cooler to condense heavy hydrocarbons which is separated from the gas in a downstream separator, lowering cricondenbar to transport specifications ($P_{crit}$<105bar). After the cricondenbar control, gas enters a glycol dehydration process to lower the gas water dew point before it is pressurised and sent to shore through Åsgard transport pipe.

## 4.2 Subsea processing systems

Subsea processing is so far mainly used for pressure boosting, and water treatment and injection. In **Figure 4.2** the technologies is placed based on maturity, from proven subsea technologies on the left side, to no identified concepts to the right. Pumps, separators and coolers are of the most mature subsea technology, compressors are also installed subsea in the Norwegian Sea at Åsgard and Gullfaks. Heaters, that are placed on the fare right, are only used for hydrate prevention at this point. In **Appendix F** more about the status for subsea processing is presented. Liquid stabilisation, gas cricondenbar control and dehydration are not yet done subsea, this will be a more complex system than what has been done subsea so fare.

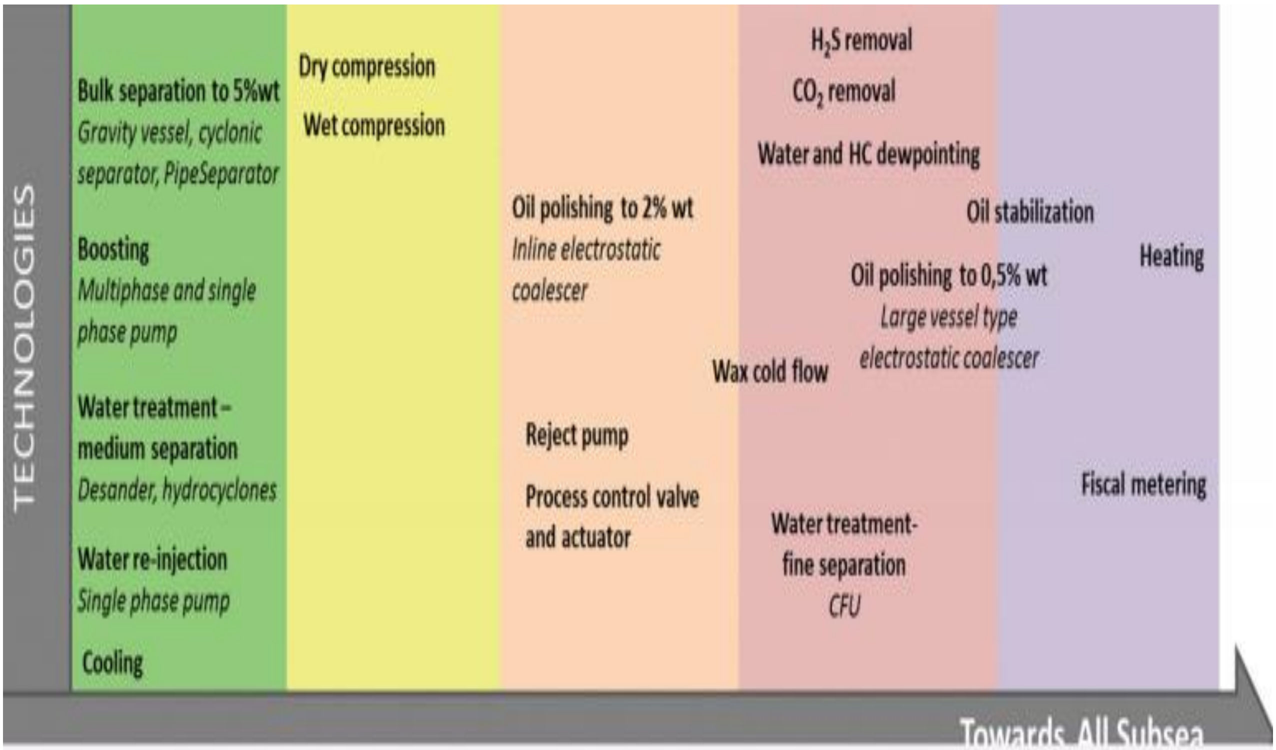



**Figure 4.2 Maturity status of subsea processing technologies**
**(Ruud, Idrac, McKensie, & Høy, 2015)**

### 4.2.1 Subsea solution utilising stabiliser column

Subsea stabilisation is not available at this time, and the development is still on a research stage. A system proposed for subsea stabilisation of hydrocarbon liquid is shown in **Figure 4.3** (Hove, 2013). The well stream is assumed dehydrated upstream this process, so no water is coming into the process. This assumption seems not to be very realistic, as the well stream will be saturated with water and also produce free water with the hydrocarbon flow. Anyhow the total dehydrated well flow rate is set to be 20 $MSm^3/d$ (mole flow ideal gas 3525 kmol/h), and has a molar weight of 21.8 kg/kmol in the case study done on the system in **Figure 4.3**. The upper figure shows the overall system with an inlet separator for bulk separation of gas and liquid. (Hove, 2013)

Liquid from the inlet separator is sent to the stabilisation process, which is shown in the lower part of **Figure 4.3**. As seen from this figure there is a liquid flow rate going to the stabilisation of 2706 kmol/h or 15.4 $MSm^3/d$ ideal gas. The liquid pressure is reduced from 100 to 10 bar, lowering the liquid vapour pressure by flashing of light hydrocarbons and separating them out in a gravitational separator. Actually more than half of the flow flashes in this separator, and sent to recompression. Liquid is further sent to a column where light hydrocarbons is boiled of to get a full stabilised liquid (TVP<1 bara). For the given process parameters, the columns reboiler use 4.25 MW and a heater upstream the column use 9.34 MW to boil of the light hydrocarbons. This heat will most likely come from electrical power in a subsea process. (Hove, 2013)

It is important to be aware that the gas is not processed in this system, only pressurised and transported in a pipeline. In a more realistic scenario this gas will be above the cricondenbar criteria and saturated with water. So it can therefore be expected that hydrocarbons and water will condense out in the pipeline. There are low temperatures in the system, down to -21°C, this will be an environment where both ice and hydrates can form in the system, see **Figure 4.5** and **Figure 3.4** for example of hydrate curves. This means that it can be expected additional heat input or hydrate inhibitor upstream the stabilisation process to avoid ice and hydrate formation.

Compared to subsea systems operating today, the system proposed in **Figure 4.3** is complex, has a large number of units and high power consumption. Use of a distillation column subsea seems a bit optimistic, regarding its complexity, operational reliability and maintenance needs. Use of a subsea column is therefore not addressed further in this study. To avoid use of a column and reduce heat input it can be an option to use partial stabilisation of the liquid, where less light hydrocarbons need to be flashed or boiled of the liquid compared to a full stabilisation.

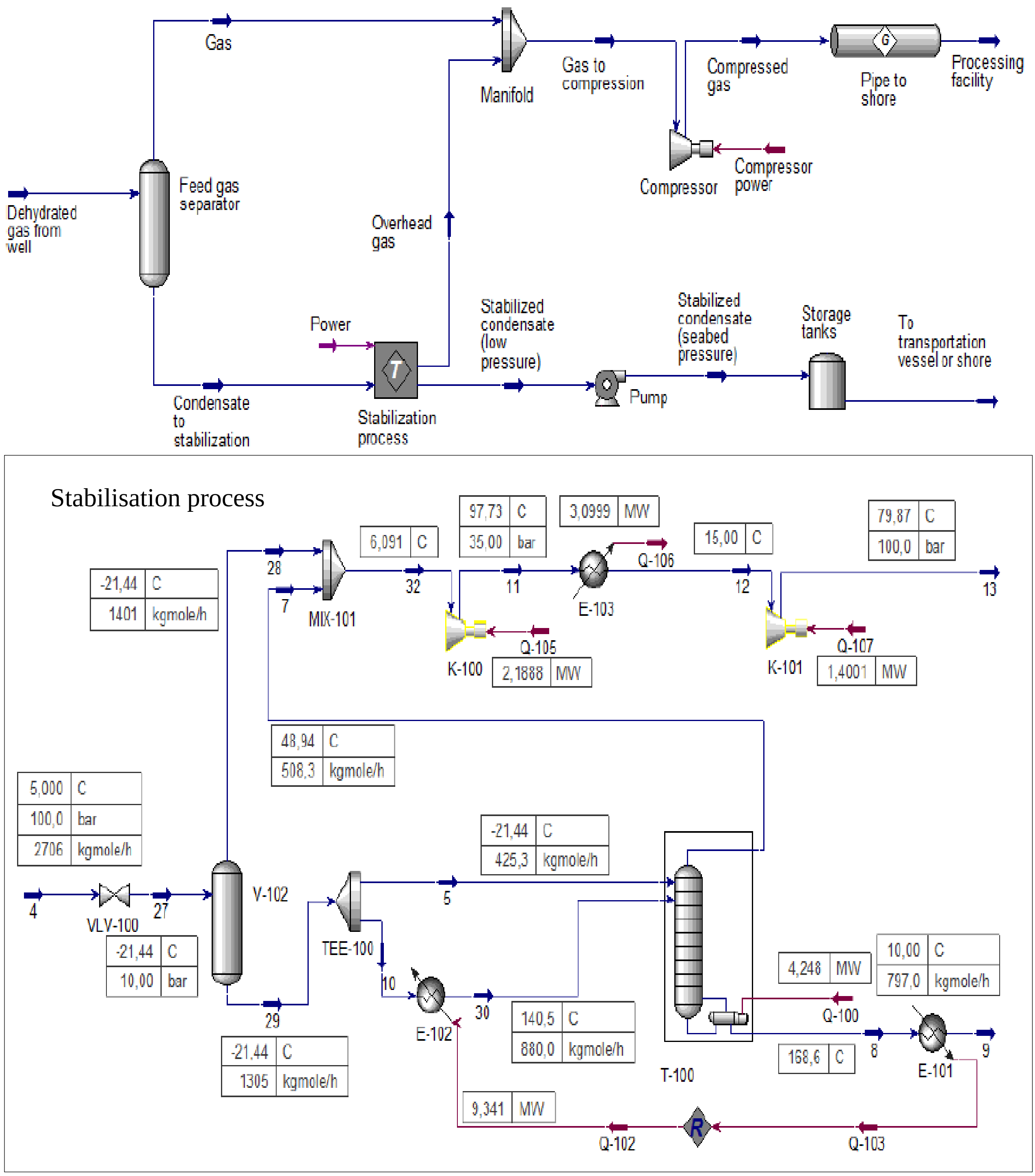



**Figure 4.3 Subsea liquid stabilisation process, Overall system design is shown in the upper figure, Stabilisation process is shown in the lower figure. (Hove, 2013)**

### 4.2.2 Two-stage subsea solution

The solution shown in **Figure 4.4** is a simplification of a conventional topside process (Kraabøl, 2015). In the inlet separator (HP), water, hydrocarbon liquid and gas is separated. The water separated will need further processing, but is left out as it don't affect the process and water treatment is already available subsea, see **Figure 4.2** and **Appendix F**.

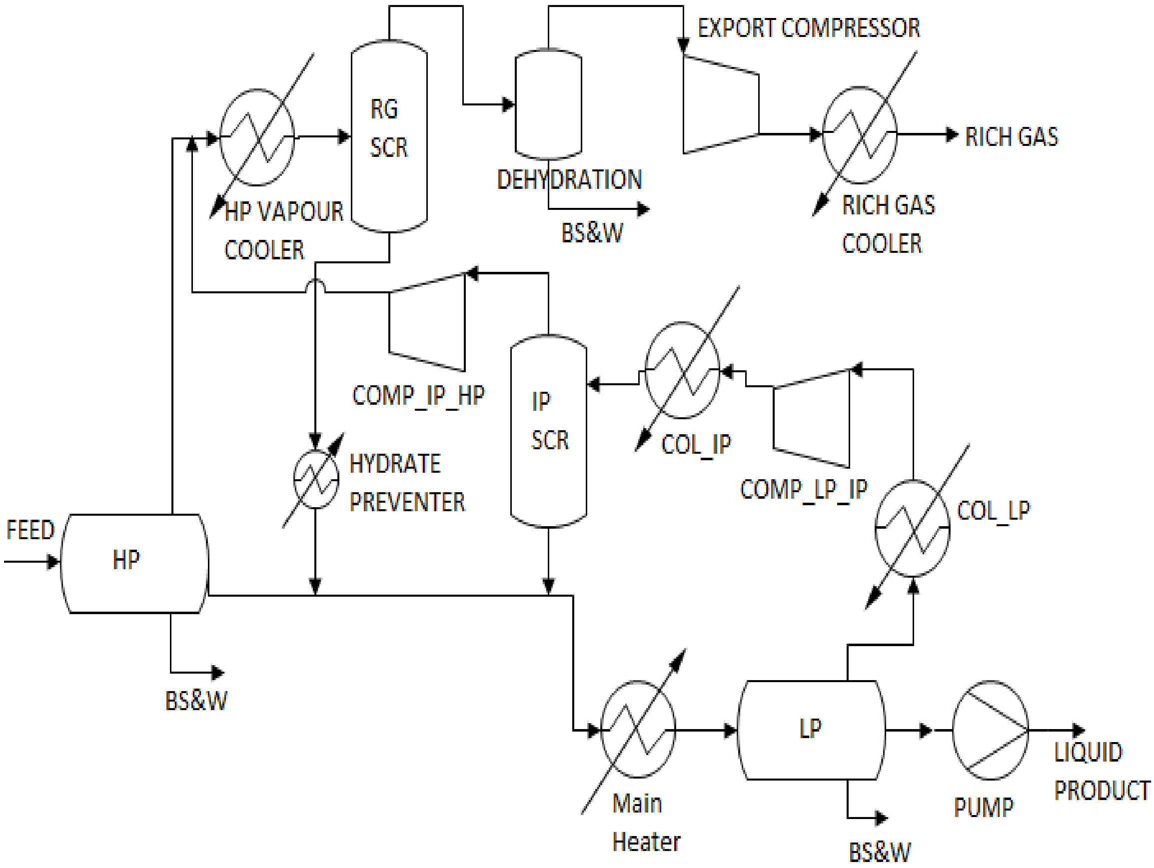



**Figure 4.4 Two-stage subsea process (Kraabøl, 2015)**

Hydrocarbon liquid is taken further down in pressure and heated, boiling and flashing of light hydrocarbons lowers the liquid vapour pressure. The LP separator is also a three phase separator to get a liquid that has an acceptable vapour pressure and water content.

One of the most challenging parts of the process is recompression of gas from the LP separator. This was found challenging due to high pressure ratios and temperatures giving need of coolers to keep the temperature down, see **Table 4.1**. This cooling will condense some of the gas, and in a conventional system scrubbers upstream the compressors is used to avoid liquid entering the compressors, see **Figure 4.1**. As seen for the Kristin process in **Figure 4.1**, when a scrubber is used between the LP separator and the first compressor, a condensate pump is needed to get the liquid back to the LP separator. Subsea it is important to lower number of units and in special rotating equipment, therefore it is assumed that compressors in **Figure 4.4** can handle 5vol% liquid. To be sure not to break the liquid restriction for the compressors a scrubber is placed upstream the second compressor in the recompression train. This solution adds only this scrubber, as the liquid is at IP pressure the liquid will flow back to LP separator without need of a pump. (Kraabøl, 2015)

In **Figure 4.4** the gas is first going to the cricondenbar control. Here the gas is cooled down (HP Vapour cooler) to condense heavy hydrocarbons and separating them out in a downstream scrubber (RG SCR). After this process the cricondenbar and cricondenterm specification for Rich gas transport is reached, see **Table 3.6** for specifications.

After the cricondenbar control the gas is still saturated with water and need to be dehydrated. The dehydration process system is missing in this solution, and is only represented by a separator to show that water need to be removed. Further development on this will be needed to get a feasible subsea process solution. In the study done on this system, a fictive component remover was used in HYSYS to remove water. (Kraabøl, 2015)

In **Table 4.1** data from HYSYS simulations done on the process solution shown in **Figure 4.4** is presented. The liquid is partial stabilised for transportation by ships or an oil pipeline, giving lower heat consumption or higher pressures than for a full stabilisation case. For the rich feed case lowering the TVP from 10bara to 5bara yields a significant increase in power consumption for the heater and a significant reduction in operating pressure for the LP separator. (Kraabøl, 2015)

The framework used in development and simulations of this system was almost the same as what is used in this master thesis. The only difference is that it is used a limitation of 5vol% liquid in the compressors, which give need for a scrubber in the recompression part.

**Table 4.1 Data from simulations of the Two-stage subsea process (Kraabøl, 2015)**

| PIPE \| SHIP | Rich (TVP<10bara) | Lean (TVP<10bara) | Rich (TVP<5bara) | Lean (TVP<5bara) |
|---|---|---|---|---|
| Dry feed compostion | Rich feed | Lean feed | Rich feed | Lean feed |
| Feed temperature (°C) | 100 | 100 | 100 | 100 |
| Feed pressure (bara) | 250 | 250 | 250 | 250 |
| Liquid product transport spec (Note3) | Norpipe (TVP<10bara) | Norpipe (TVP<10bara) | Ship (TVP<5bara) | Ship (TVP<5bara) |
| Gas transport spec (Note 4) | Åsgard | Åsgard | Åsgard | Åsgard |
| Volume flow at 1atm/15°C ($MSm^3/d$) | 3 | 3 | 3 | 3 |
| Feed mass flow - dry basis (tonne/d) | 4572.6 | 2431.6 | 4572.6 | 2431.6 |
| Feed mass flow - wet basis (tonne/d) | 4683 | 2543 | 4683 | 2543 |
| Rich gas product ($MSm^3/d$) | 2,9 | 3,0 | 2,9 | 3,0 |
| Liquid product (Actual $m^3/d$) | 2894 | 119 | 2852 | 117 |
| Rich gas product (tonne/d) | 2553 | 2348 | 2582 | 2349 |
| Liquid product (tonne/d) | 2021 | 84 | 1992 | 83 |
| Comp_LP_IP Suction Volume flow($m^3/h$) | 463 | 20 | 1121 | 45 |
| Comp_LP_IP Suction mass flow(tonne/d) | 322 | 10.5 | 603 | 16 |
| Recompression work (kW) | 405 | 18 | 733 | 30 |
| Recompressor pressure ratio | 2,2 | 2,1 | 2,8 | 2,8 |
| Export compressor work (kW) | 3878 | 5142 | 4011 | 5143 |
| Export compressor pressure ratio | 2,6 | 2,9 | 2,6 | 2,9 |
| Highest compressor T (Note 1) (°C) | 104(E) | 124(E) | 106(R) | 124(E) |
| Max liquid in compressor (vol%) | 2,9 | 1,1 | 4,8 | 1,5 |
| Rich gas cooler($T_o$ 60°C) (kW) | 3981 | 5292 | 4122 | 5293 |
| HP Vapour cooler ($T_o$ 25°C) (kW) | 5563 | 3522 | 5755 | 3530 |
| Recompression cooling (Note 2) (kW) | 1691 | 54 | 3828 | 100 |
| Liquid product cooler ($T_o$ 100°C) (kW) | - | - | - | - |
| Heater outlet temperature (°C) | 100 | 100 | 100 | 100 |
| Main heating utlity (kW) | 2744 | 209 | 4665 | 246 |
| Liquid product outlet pressure (bara) | 100 | 100 | 41 | 41 |
| Pump work (kW) | 376 | 16 | 137 | 6 |
| Pressure steps, HP\|LP (bara) | 78\|17 | 70\|16 | 78\|10 | 70\|9 |
| Notation: In **Figure 4.4** Recompression is Comp_LP_HP and Comp_IP_HP ; Recompression cooling is Col_LP and Col_IP;<br>Note 1: The highest compressor temperature is marked with E for Export compression and R for Recompression<br>Note 2: All coolers has an outlet temperature of 25°C<br>Note 3: Liquid product specifications is given in **Table 3.5**<br>Note 4: Gas transport specifications is given in **Table 3.6**<br>Note 5: Compositions used is shown in **Table 3.2** | | | | |

## 4.3 Gas dehydration

The gas coming from the well is always saturated with water, and need to be dried in a dehydration process to avoid water knock out in pipelines. If hydrocarbon gas is mixed with free water, hydrates can form and block process equipment and pipelines. In **Figure 4.5** the behaviour of propane mixed with water is shown. The hydrate curve is shown as a solid line, to the left of this line hydrates will form and to the right hydrates is not going to form, the same behaviour is found for mixtures (Campbell, 1992). To avoid water knock out rich gas transportation specification on the Norwegian continental shelf has typically a maximum water dew point of -18°C at 70bara, see **Table 3.6**.

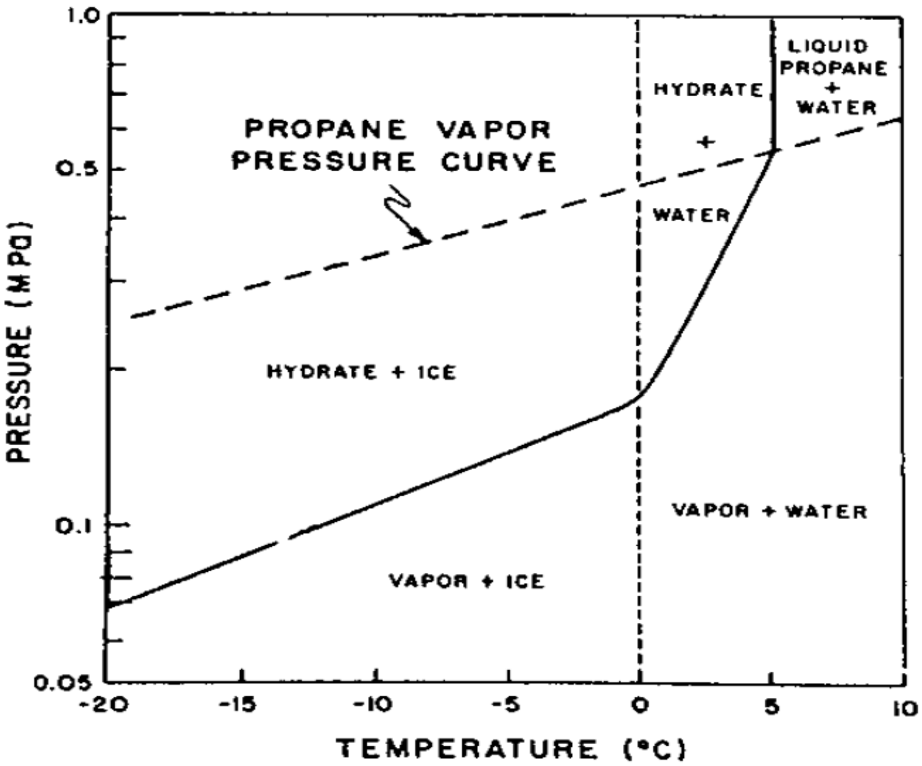



**Figure 4.5 General behaviour for propane with water, hydrate curve (Campbell, 1992)**

On the Norwegian continental shelf the most common dehydration process is absorption by glycol. This process is able to reach the rich gas water dew point specification and glycol is often available for use as hydrate inhibitor as well. Other dehydration processes is adsorption, use of selective membranes, and cooling and/or expansion.

### 4.3.1 Glycol absorption

#### 4.3.1.1 Conventional glycol absorption process

The most common dehydration method in gas processing is use of glycol that absorbs water from the gas. In **Figure 4.6** a conventional glycol absorption process is shown. In addition to the components shown in **Figure 4.6**, it is important to use a scrubber upstream this process to remove free liquid from the feed gas stream (Eimer, 2014). If liquid hydrocarbons enter the absorber foaming can occur. Foam will lower performance of the absorber by decreasing contact between glycol and gas, and increase glycol carryover.

The absorber in **Figure 4.6** is a counter-current multiple stage contactor, and has trays or packings that ensure contact between glycol and gas. Lean glycol enters at the top of the column and flows downwards, while the gas enters at the bottom flowing upwards. This ensures that it is driving forces available for dehydration through the whole column. For a tray column each tray is close to equilibrium leaving the gas dryer for each tray, while a packed column has a more continues operation.

At the bottom of the column water rich glycol leaves and is sent for regeneration. To recover glycol the water is boiled of at high temperature, but the temperature is limited to thermal decomposition of the glycol (About 200°C for TEG), and in addition stripping gas is often needed to get the water content left in the glycol low enough.

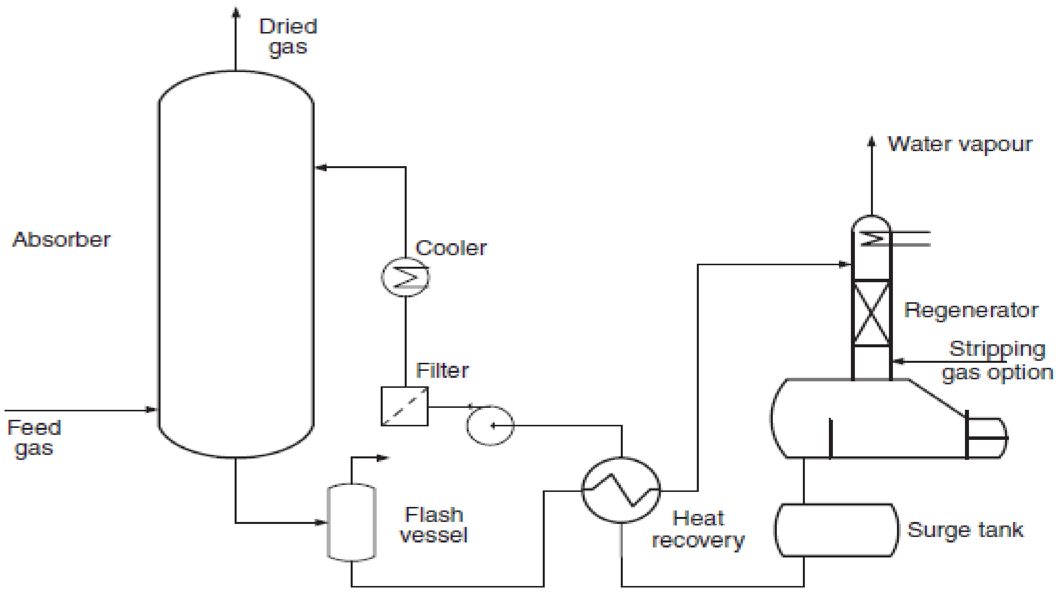



**Figure 4.6 Conventional glycol adsorption process (Eimer, 2014)**

#### *4.3.1.2 Co-current gas/glycol contactor process (single-stage contactor)*

The traditional counter-current contactor is known to be very large and heavy, making them less attractive offshore (ExxonMobil, Cullinane, Grave, & Freeman, 2014). For moderate dew point reduction, a single-stage contactor will give significant size and weight reduction (SINTEF, 2016).

In **Figure 4.7** theoretical equilibrium stages for a single stage and multi-stage process is shown. A co-current contactor has maximum one theoretical equilibrium stage, while a typical counter-current gas/glycol contactor has 2-4 theoretical equilibrium stages. Gas leaving the single-stage contactor will be in equilibrium with the rich glycol, while gas leaving the multi-stage contactor will be in equilibrium with the lean glycol at the last stage (SINTEF, 2016). As long as it is sufficient driving forces to make $H_2O$ go from the gas to the glycol, number of equilibrium stages will decide how dry the gas will be at the outlet. If one single-stage contactor is insufficient to reach the water dew point specification, adding more units can be an option.

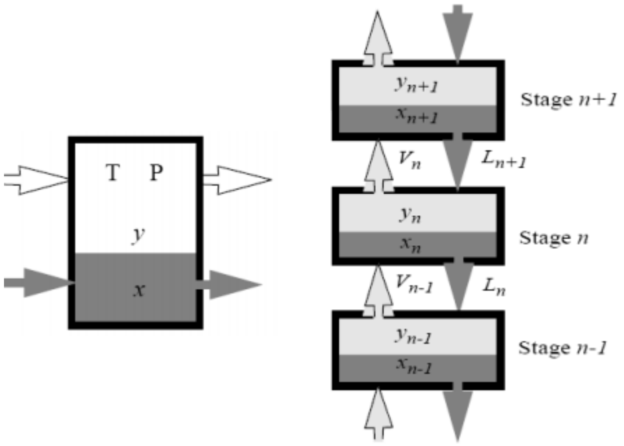



**Figure 4.7 Single-stage (left) and multi-stage (right) equilibrium process** (SINTEF, 2016)

To make a contactor that can approach equilibrium, contact area and time between gas and glycol is very important. Low contact means low transfer of $H_2O$, no equilibrium between gas and glycol, and in the end the gas will not be dried sufficiently. Improving contactor design, increasing retention time or adding more stages can options to reach a lower water dew point. Glycol purity is also very important, and is typical above 99% to have sufficient driving forces to reach the water dew point specification for a rich gas (-18°C/70bar).

In **Figure 4.8** dehydration process using single stage co-current contactors (Mixer 1 and 2) is shown. This is a process mainly designed for use subsea, and to lower complexity the glycol regeneration is not done subsea but at a topside installation nearby. Feed gas is first mixed with a semi lean glycol in mixer 1, which removes some of the water. In mixer 2 the gas is mixed with lean glycol to reach rich gas specifications. After each mixer there is a separator to separate the water and gas. It is found that the co-current contactor process is most efficient when the feed gas has low temperature and/or high pressure. High pressure and/or low temperature provide reduced TEG circulation rate, and number of mixers/equilibrium stages needed to reach the water dew point specification. (Fredheim, Johnsen, Johannessen, & Kojen, 2016).

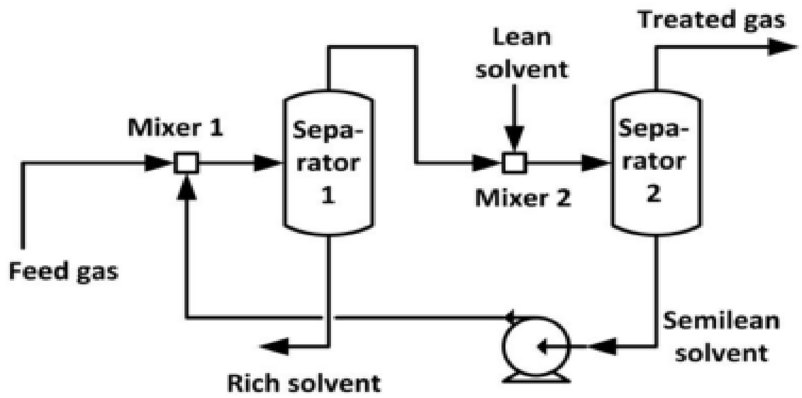



**Figure 4.8 Co-current contactor process with two stages**

(Fredheim, Johnsen, Johannessen, & Kojen, 2016)

There are different co-current single-stage contactors available which can be used as mixers in **Figure 4.8**, see **Figure 4.10** and **Figure 4.9** for some examples. The ProDry contactor in **Figure 4.9** can reduce the water dew point with about 30°C for a single mixer, it can tolerate entrained liquid in the feed gas, no flooding or foaming issues, it has a high interfacial area between gas and glycol giving it high efficiency. (SINTEF, 2016)

**Figure 4.9 ProPure C100W (ProDry) Co-current contactor** (SINTEF, 2016)

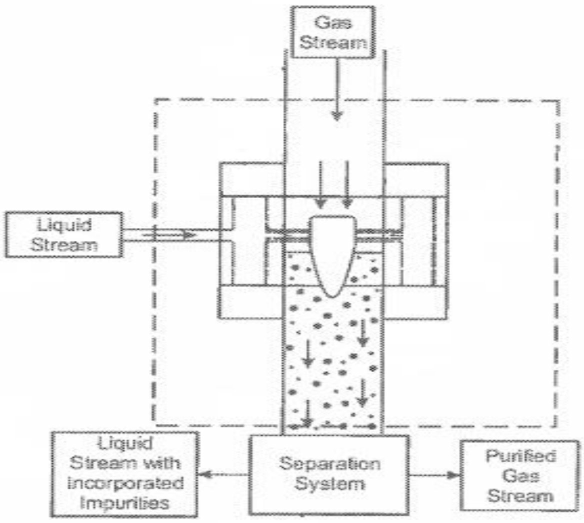



**Figure 4.10 Vertical oriented co-current contactor** (ExxonMobil, Cullinane, Grave, & Freeman, 2014)

### *4.3.2 Adsorption*

Adsorption uses solids where water is adsorbed on the surface and in pores of the adsorbent. By using adsorption the water dew point can get very low. In general adsorption processes is seen expensive and this technology is most common in processes with low dew point requirement. Actually adsorption can be used to practically remove all the water and is therefore used in LNG process plants.

To operate an adsorption process at least two columns in parallel will be needed, one in adsorption mode and one in regenerating mode, see **Figure 4.11**. This is because the column will get saturated with water after some time in operation. To regenerate the column a stripping gas is used to desorb the adsorbed water. This is energy demanding and the stripping gas need to have a high temperature, typical above 200°C for zeolites (Eimer, 2014).

In a subsea system heat for regeneration must most likely come from an electrical powered heater, giving a direct impact on the plants energy consumption which is a subsea challenge. If there is fouling gas going through the column this can reduce or in worst case destroy the adsorbents dehydrating capabilities, reduce the systems reliability and increase maintenance needs. Therefore use of this system directly on a well stream process can give operational difficulties.

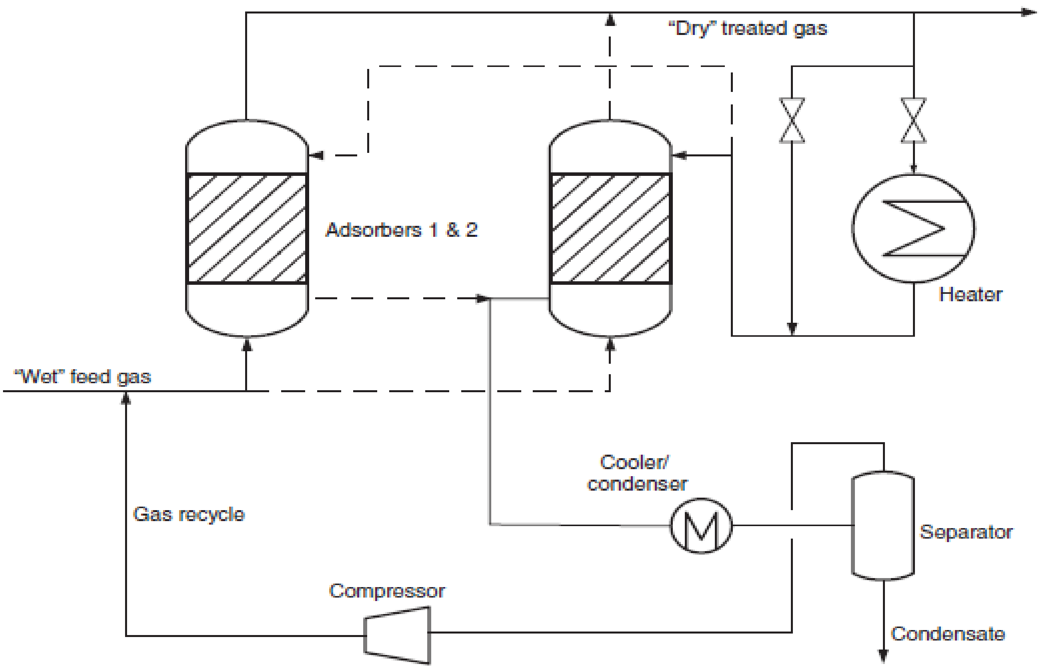



**Figure 4.11 General flowsheet of an adsorption dehydration process (Eimer, 2014)**

### *4.3.3 Dew pointing by cooling and/or expansion*

Dehydration by cooling and/or expansion lowers the gas temperature and provides condensation. When using expansion, pressure is reduced giving increased power consumption for the export compressor. If these processes are operating below the hydrate curve, hydrate inhibitor like glycol will be needed.

One process utilising expansion is the supersonic twister technology shown in **Figure 4.12**. Here the gas is accelerated to supersonic velocity in a naval nozzle reducing the temperature, which creates condensate that is separated out, and the gas stream is further sent through a diffuser recovering some of the pressure energy. As seen in **Figure 4.12** the temperature is much lower than for a normal Joule Thompson valve, so the twister can reach a lower dew point. Since the temperature is very low, far to the left of the hydrate curve, hydrate inhibitor will be needed.

The cricondenbar control in **Figure 4.1** and **Figure 4.4** is a cooling process where both water and hydrocarbons are condensed, but to avoid hydrate the temperature is kept above the hydrate curve. With this high temperature the processes is not sufficient to reach the rich gas water dew point specification, only the cricondenbar can be reached at this stage.

The cooling and/or expansion solutions seems as a good alternative for the cricondenbar control and to reduce the water dew point, but for lowering the water dew-point to the rich gas specification it is found less attractive. The main reason is that it will increase power consumption and size for the export compression, since the pressure must be reduced. Another argument is that glycol will be needed for hydrate inhibition, making the glycol absorption process to a more feasible alternative.

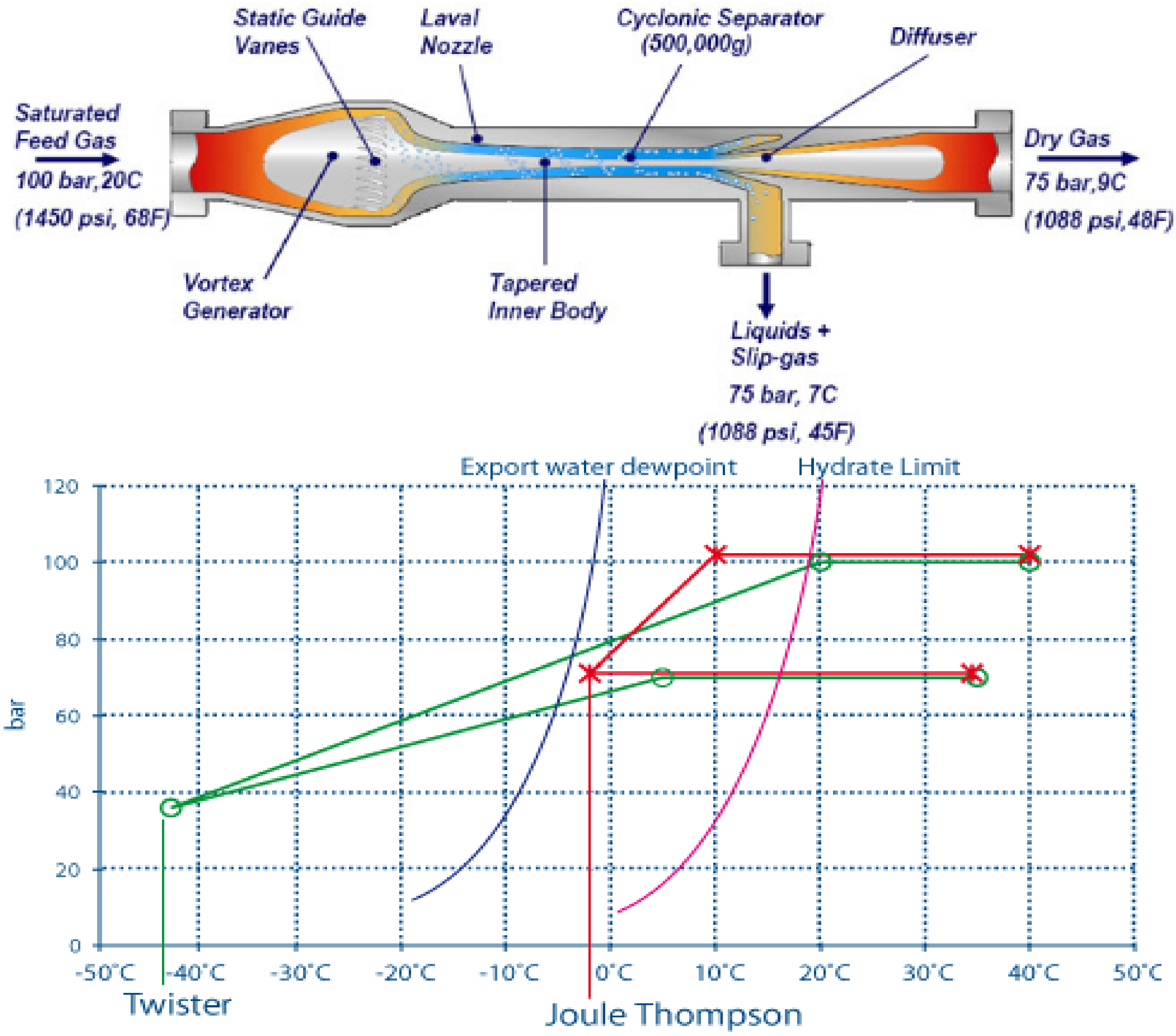



**Figure 4.12 Twister supersonic separator** (Twister, 2016)

### *4.3.4 Membrane dehydration*

Membrane dehydration uses a selective membrane that allows water molecules to pass through but not the hydrocarbon gas. The advantage of a membrane separation is that it has no power consumption and a relatively low number of units. Membranes for gas processing are in a development stage, and probably a large qualification program and significant topside testing is needed if it should be made available for use subsea. One issue is that the membrane must be tolerant for some hydrocarbon liquid. Also for membranes fouling gas will be challenging, and decrease reliability and increase maintenance for cleaning or replacing the membrane. Due to lack of operational experience, state of the technology and concerns related to maintenance, membrane technology is not evaluated further in this study.

# 5. Assessment of equipment

When designing a new process plant equipment selection is critical. Choices made at this stage will affect the whole operation and have direct impact on CAPEX and OPEX. Technology chosen for this study is done in regards to the subsea design philosophy, the process block diagram in **Figure 2.1**, earlier research, presented process solutions, and new knowledge about subsea processing.

## 5.1 Pumps and compressors

In the process block diagram shown in **Figure 2.1**, there are fluids moving from low pressure to high pressure regions. Moving fluid from low to high pressure is possible with use of equipment like a pump or compressor to increases pressure of the low pressure fluid. In **Appendix D** classification of compressor and pump technologies are presented. The two main technologies used are positive displacement and dynamic machines.

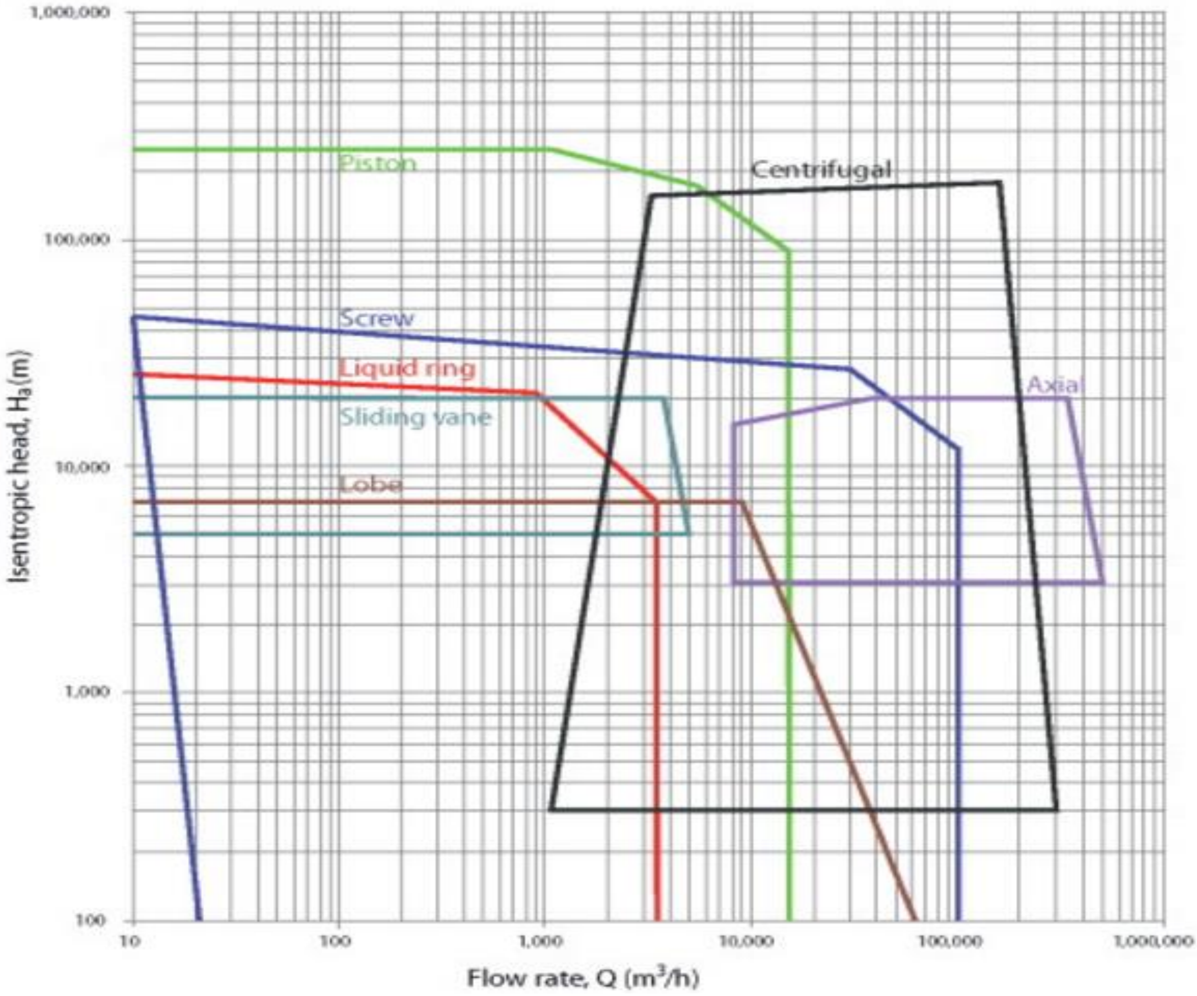



**Figure 5.1 Compressor selection chart** (Larralde & Ocampo, 2014)

Selection of pump and compressor technology depends on factors such as flow rate, pressure increase/ratio, fluid properties, and temperature. In **Figure 5.1** a rough estimate of compressor application range for some of the most common compressor types are shown. New technology are pushing and moving on these boundaries as stated in the following section, but **Figure 5.1** gives an indication on which compressors that may be applicable for a given flow rate and head.

Earlier research on the subsea process plant in **Figure 2.1** shows large differences in flow rates and pressure ratios for compressors and pumps. There is also found multiphase flow or wet gas in the recompression section when cooling flash gas below dew point, see **Table 4.1**. (Kraabøl, 2015)

Technologies selected for further evaluation is based on earlier research, available technology, the compressor selection chart in **Figure 5.1** and important factors for subsea processing. Important factors for subsea processing are found in the design philosophy, **Section 3.2**. The main technologies selected for this study is centrifugal, screw, ejector and counter rotating axial compressor. More details about the technologies and background for the selections are presented in the following sections.

### 5.1.1 Compressor and pump data

**Table 5.1** shows vendors performance data for pumps and compressors available for oil and gas production subsea and topside. **Table 5.2** shows operational performance data for subsea pumps and compressors which is or has been in operation subsea.

**Table 5.1 Vendor data for a selection of pumps and compressors**

| Manufacturer - Type | Type of technology | Highest discharge pressure (bar) | Unit flow rate ($m^3/h$) | Differential pressure (bar) | Gas volume fraction | Unit Power (MW) | Main application area |
|---|---|---|---|---|---|---|---|
| Bornemann – SMPC | Twin-screw Multiphase pump | | 50-2500 | Up to 150 | 0-100% | | Subsea |
| GE - Blue C | Centrifugal compressor | 175 | Up to 18 000 | | | Up to 15 | Subsea |
| Siemens - STC-ECO | Centrifugal compressor | 220 | 250-40 000 | | | Up to 20 | Topside, Subsea |
| MAN – HOFIM | Centrifugal compressor | 303 | 35 700 | | | Up to 20 | Topside, Subsea |
| Howden | Oil-Free screw compressor | 15 | Up to 26 000 | | | | Topside, Onshore |
| Kobelco | Oil-Free screw compressor | 45 | Up to 142 000 | | | | Topside, Onshore |
| MAN- CP type | Oil-free screw compressor | 50 | 200-20 000 | | | | Topside, Onshore |
| MAN- CP type | Oil-free screw compressor | 16 | 4 000-100 000 | | | | Topside, Onshore |
| References: (GE, 2016) ; (Siemens, 2016) ; (MAN-Disel&Turbo, 2016) ; (Howden, 2016) ; (Kobelco, 2016) ; (Bornemann, 2016) | | | | | | | |

**Table 5.2 Operational data for a selection of subsea pumps and compressors**

| Field/ Region | Unit manufacturer/ Type of technology | Water depth (m) | Unit flow rate ($m^3$/h) | Differential pressure (bar) | Gas volume fraction | Unit Power (MW) | Operative units/ start-end or present (month. year) |
|---|---|---|---|---|---|---|---|
| Åsgard/ Offshore Norway | MAN/ Centrifugal compressor | 300 | 10000 | 50 | n/a | 11.5 | 2 units/ 09.2015-05.2016 |
| Gullfaks/ Offhsore Norway | OneSubsea/ Contra rotating wet gas compressor | 135 | 4800 | 30 | 95% | 5 | 2 units/ 10.2015-05.2016 |
| Ceiba C3+C4/ Equatorial Guinea | OneSubsea/ Helico-Axial multiphase pump | 750 | 300 | 45 | 75% | 0.85 | 2 units/ 10.2002-05.2016 |
| Mutineer/ NW Shelf Australia | OneSubsea/ Helico-Axial multiphase pump | 145 | 600 | 30 | 0-40% | 1.1 | 2 units/ 03.2005-05.2016 |
| Barracuda / Campos Basin Brazil | OneSubsea/ Helico-Axial multiphase pump | 1040 | 280 | 70 | 35-60% | 1.5 | 1 unit/ 07.2012-05.2016 |
| Girassol/ Blk17 Angola | OneSubsea/ Helico Axial multiphase pump | 1350 | 600 | 130 | 20-50% | 2.5 | 4 units/ Q2.2015-05.2016 |
| King/ US GOM | Bornemann/ Twin-screw multiphase pump | 1700 | 500 | 50 | 0-95% | 1.3 | 2 units/ 11.2007-02.2009 |
| Marlim/ Campos Basin Brazil | Leistitz/ Twin-screw multiphase pump | 1900 | 500 | 60 | 0-100% | 1.2 | 1 unit/ |
| Lufeng 22/1 Field | OneSubsea/ Centrifugal Single phase pump | 330 | 135 | 35 | 3% | 0.4 | 5 units/ 01.1998-07.2009 |
| Julia/ US GOM | TBD | 2287 | 166 | 175 | 10% | 3 | 2 units/ Start: mid-2016 |
| References: (INTECSEA & Magazine, INTECSEA.com, 2016) ; (Souzea, et al., 2013), (Bibet, Huet, & Åsmul, 2016) ; (Vinterstø, Birkeland, Ramberg, Davis, & Hedne, 2016) ; (Davis, Kelly, Kierulf, Normann, & Homstvedt, 2009) | | | | | | | |

### 5.1.2 *Centrifugal compressor (Åsgard)*

The centrifugal compressor is one of the most common technologies found in upstream gas processing. Operational experience is a significant advantage for this technology, providing reliability and possibilities for optimization of well-known designs. In 2015 a centrifugal compressor was set in operation subsea at the Åsgard field in the Norwegian Sea.

Centrifugal compressors can handle a wide range of flow rates (250-300 000 $m^3/h$), and relatively large discharge and differential pressures, see **Figure 5.1**, **Table 5.1** and **Table 5.2**. This machine is relatively robust and can be made for low maintenance, using for example magnetic bearings. Some draw backs is that it can’t handle much liquid, often a scrubber will be needed upstream the compressor, and it needs a surge control to avoid surge in the compressor. With the possibilities of large flow rates, high pressures, and no entrained liquiud the centrifugal compressor will be a great candidate as an export compressor for the process shown in **Figure 2.1**.

#### *5.1.2.1 Description of centrifugal compressor technology*

In a centrifugal compressor the fluid is exposed to centrifugal forces which provide kinetic energy to the fluid. Then by efficiently reduction of the fluid velocity, kinetic energy is converted to pressure energy. (Perry & Green, 1997)

**Figure 5.2** shows a schematic of a centrifugal compressor with two steps and intercooler. If the cooled gas between the stages reaches saturation a scrubber will be needed as well. Gas enters the impellers axially and leaves radially after the centrifugal force has acted on the fluid.

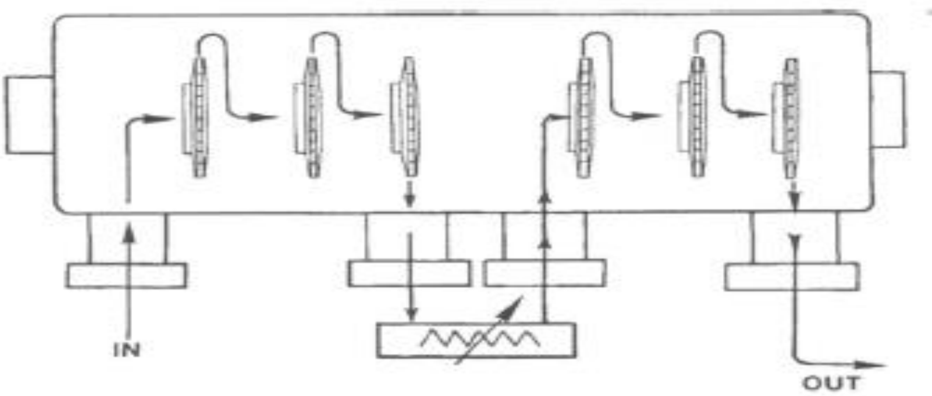



**Figure 5.2 Centrifugal compressor two stage arrangement with intercooling (Brown, 2005)**

In 2015 two subsea centrifugal compressors was installed at the Åsgard field in the Norwegian Sea. The compressors is of the type “High-Speed Oil Free Integrated Motor-Compressor (HOFIM)” delivered by MAN Diesel & Turbo, see **Figure 5.3**. This is a robust and wear-free design, magnetic bearings eliminates need for a lubricating oil system. The two compressors operating at Åsgard is on 11,5MW each and has a design pressure of 210 bar and design flow rate of 10,5MSm$^3$/d each. They are designed for a lifetime of 30 years. HOFIM compressors are available from 3MW to 18 MW integrated motor, and can deliver discharge pressures up to 303bar. (MAN-Disel&Turbo, 2016) (Vinterstø, Birkeland, Ramberg, Davis, & Hedne, 2016)

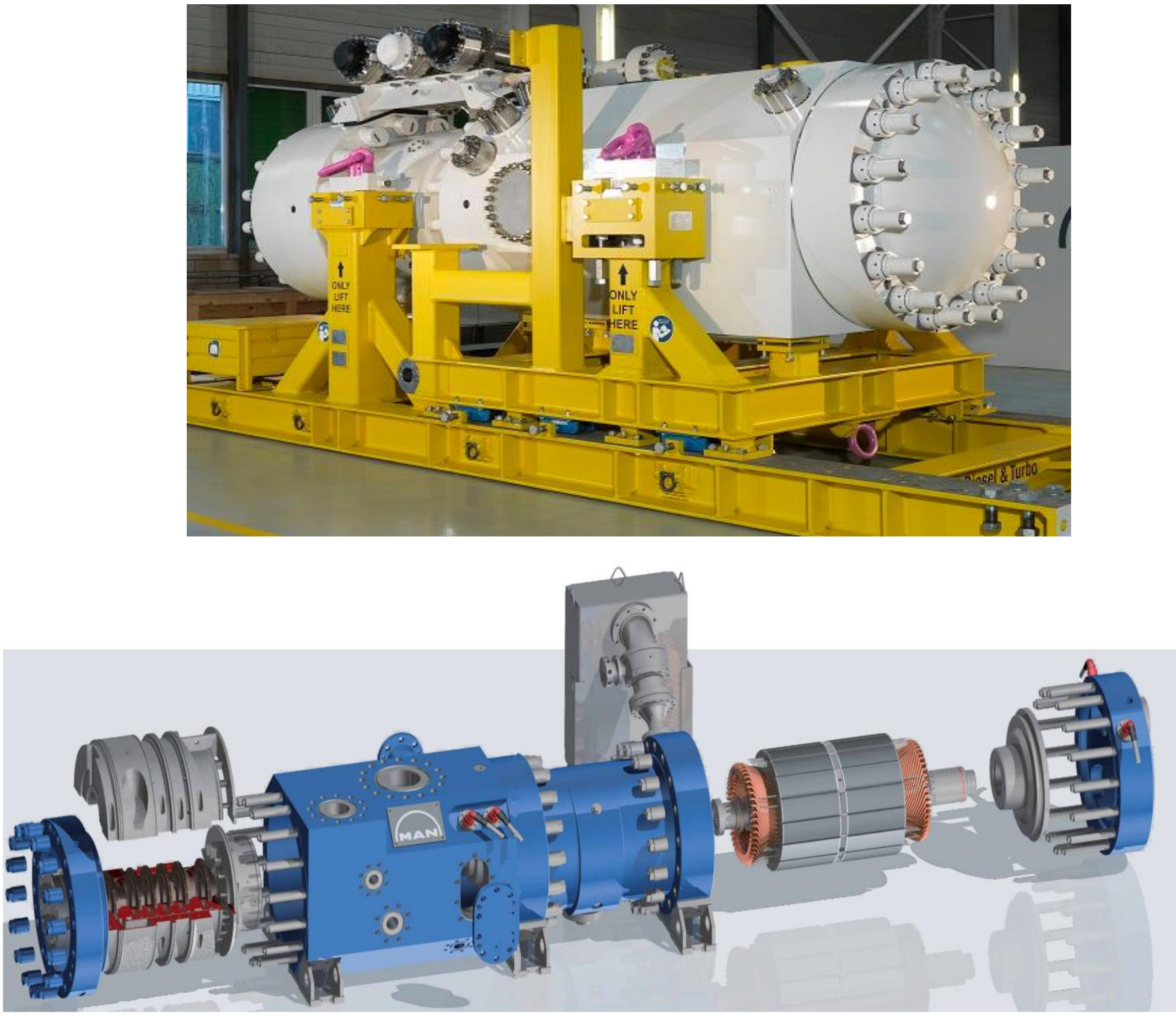



**Figure 5.3 High-Speed Oil Free Integrated Motor-Compressor (HOFIM)**

(MAN-Disel&Turbo, 2016)

The compressor can only handle small amounts of liquid. During testing it was found that it could operate fine with 5 weight % liquid (Vinterstø, Birkeland, Ramberg, Davis, & Hedne, 2016). In **Figure 5.4** a flow schematic of the Åsgard compressor station is shown. To avoid high concentrations of liquid entering the compressor an upstream scrubber is needed. The liquid is pumped and mixed back with the gas downstream of the compressor. There is also two coolers keeping the temperature down. Surge control is also needed for the centrifugal compressor. (Time & Torpe, 2016)

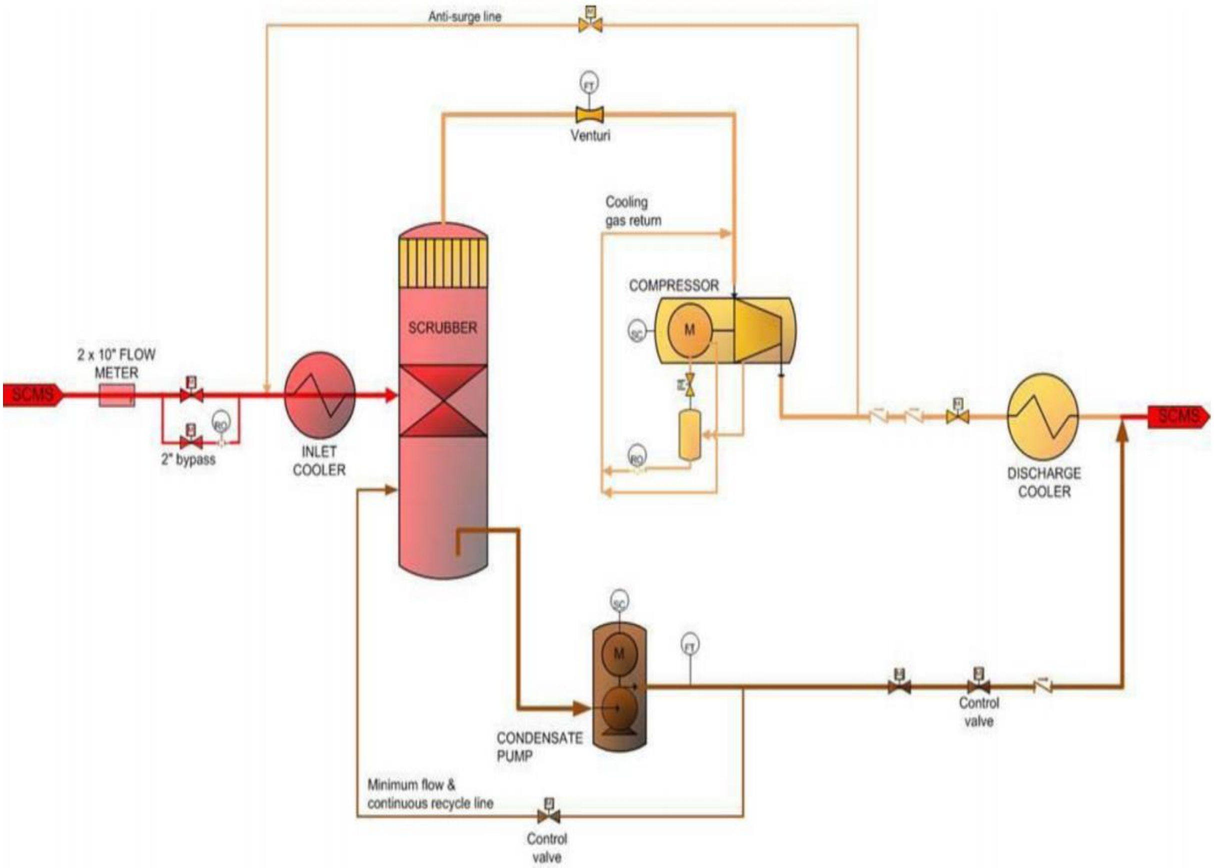



**Figure 5.4 Flow schematic of compressor train Åsgard** (Time & Torpe, 2016)

Another centrifugal subsea compressor available is the Blue-C developed by GE, see **Figure 5.5**. This is a robust machine packed in a single-sealed housing to withstand high pressures and temperatures. It is equipped with magnetic bearings to avoid wear, lower maintenance and increase reliability. To handle wet gas compression a diffuser separation system and a dust removal device is implemented in the compressor.

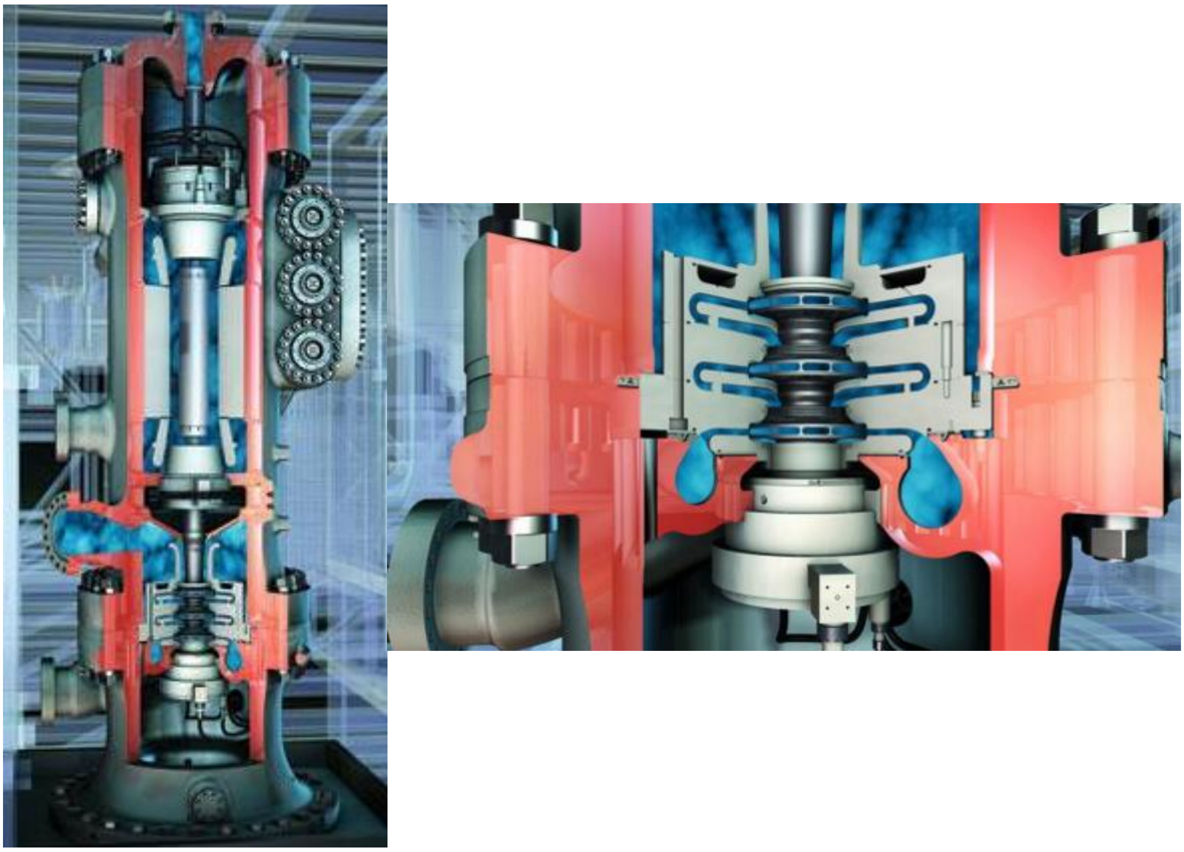


**Figure 5.5 Blue-C Subsea compressor** (GE, 2016)

### 5.1.3 Contra rotating wet gas compressor (Gullfaks)

The contra rotating axial compressor developed for use subsea at the Gullfaks field in the Norwegian Sea, developed by OneSubsea, is an interesting machine. By using contra rotating impellers, rotating the inner shaft and outer shield in opposite direction, the machine gets compact and can handle up to 0-100% liquid. The axial contra rotating compressor can be used for large flow rates and low differential pressures.

One of the drawbacks with the contra rotating machine is that it has limited differential pressure in the same way as other axial compressors. There is also limited experience in use of these machines in the industry, but it is one of two subsea compressors in operation.

Since this machine is working with high flow rates and low differential pressure, it is only at a candidate as an export compressor. The contra rotating compressor is also found to be more complex, have lower flexibility, and less operational experience than centrifugal compressors. Based on this discussion the subsea philosophy, maturity status, and since there is no entrained liquid found in the export compressor, the centrifugal compressor is the preferred option.

#### *5.1.3.1 Description of contra rotating wet gas compressor technology*

Conventional axial compressors are designed to compress gas, and are composed by rotating impellers and static diffusers. For these compressors presence of liquid will significantly reduce performance. When gas and liquid separates the gas streamlines are disturbed and performance is reduced, concentration of liquids can make imbalance and provide mechanical issues. To avoid liquid separation contra rotating impellers can be used, see **Figure 5.6**. This design provides such a good inter-stage mixing of the phases the process fluid can be considered single phase with equivalent fluid properties. The contra rotating design is also more compact and allowing a shorter shaft than a conventional axial compressor. (FramoEngineering, Torkildsen, Vikre, & Kjellnes, 2012)

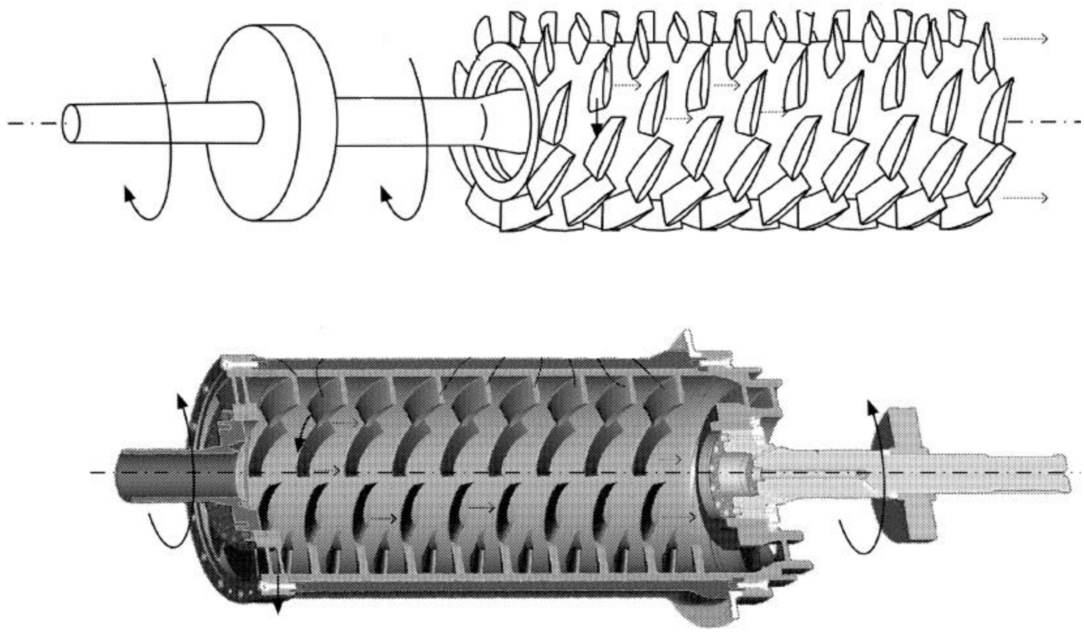


**Figure 5.6 Contra rotating wet gas compressor** (FramoEngineering, Torkildsen, Vikre, & Kjellnes, 2012)

In 2015 a contra rotating wet gas compressor was installed subsea at the Gullfaks field on the Norwegian continental shelf. The compressor can handle 0-100% liquid without mechanical issues, typically it is operating with a gas volume fraction of 95-100%. One drawback is that it needs barrier fluid from a host to provide overpressure protection, lubrication and cooling of the compressor. In **Figure 5.7** a picture of two 5MW compressor units for the Gullfaks field is shown. (OneSubsea, 2015) (Vinterstø, Birkeland, Ramberg, Davis, & Hedne, 2016)

**Figure 5.7 Two 5MW contra rotating subsea multiphase compressors for the Gullfaks subsea wet gas compression.**

(OneSubsea, 2016)

### *5.1.4 Ejector*

Ejectors can be used to compress or pressurise a fluid by use of a high pressure motive flow, see **Figure 5.8**. The ejector is robust equipment with no moving parts, giving it low maintenance needs and high reliability. In addition it is compact and has no power consumption. Ejectors are already qualified for subsea use and in operation subsea at the Marlim field, which will lower technological development costs and time.

The drawback is that available motive flow limits operational flexibility, making it necessary to evaluate different configuration to see if use of an ejector is feasible. To increase operational flexibility use of multiple ejectors in parallel and/or series can be an option, but this will add units and complexity to the subsea system.

Ejector technology fulfils most of the important factors for subsea use, with no moving parts, robustness and low maintenance need. In **Table 4.1** it is clear that the flash gas stream from the LP separator in **Figure 4.4** is relatively small, and there are larger streams available that can be used as motive flow since they already is going down in pressure. This makes ejectors an alternative for recompression of flash gas from the stabilisation process, and is selected for further evaluation in this study.

#### *5.1.4.1 Description of ejector technology*

In an ejector high pressure motive flow is accelerated through a nozzle converting pressure energy into kinetic energy. At the nozzle outlet the pressure is lower than the suction flow pressure, making mixing of the two streams possible in the mixing section shown in **Figure 5.8**. After the two streams are mixed there is a diffuser that slows the flow velocity down and recovers pressure energy. Due to irreversible processes in the ejector, such as friction, mixing losses and shock waves in supersonic flow, ejector efficiency is low. Efficiency can be increased by cooling and/or condensing the suction stream, which also will reduce ejector size and consumption of motive flow. (Perry & Green, 1997)

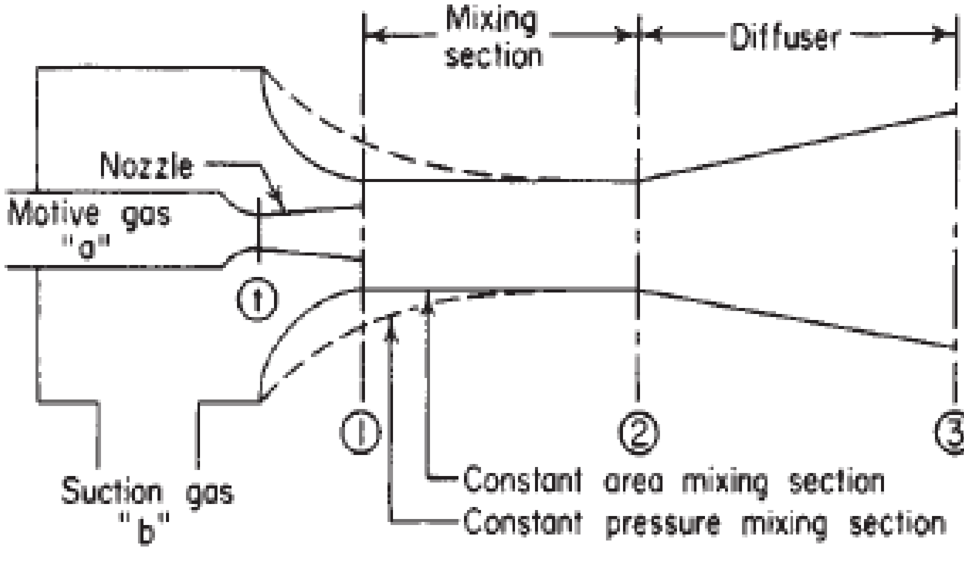



**Figure 5.8 Ejector (Perry & Green, 1997)**

### *5.1.5 Screw compressors*

The screw compressor utilises positive displacement for pressurisation of fluid and is a robust type rotary compressor. The screw compressor can operate with lower flow rates than a centrifugal compressor, see **Figure 5.1**. If a conical screw is used there is almost no limit for how low the flow can be, see **Figure 5.13** where the flow rate is barely above $2m^3/h$. The oil-free screw compressor can also handle entrained liquid in the gas, something most compressors cant.

There is found to be low flow rates and entrained liquid in recompression of flash gas from the stabilisation process, see **Table 4.1**. In addition robustness, operational flexibility, and low maintenance need, makes the screw compressor a candidate for recompression of flash gas from the stabilisation process. Use of the screw compressor will probably eliminate the scrubbers in the recompression system, lowering the number of units which is important for the subsea process plant.

#### *5.1.5.1 Description of screw compressor technology*

In a screw compressor gas is entrapped in a space that decreases from the inlet to the outlet. **Figure 5.9** shows a dual shaft screw compressor where the gas is entrapped between the screws, and as they rotate the spaces between them decreases and compress the gas. (Bloch & Soares, 1998)

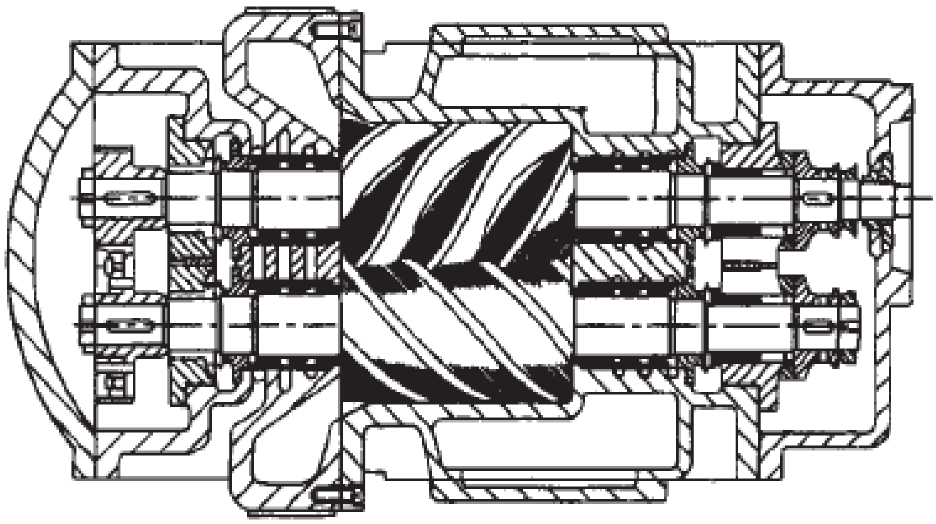


**Figure 5.9 Twin screw compressor** (Perry & Green, 1997)

The main screw compressor types are oil-free or flooded, which has twin screw or single screw for pressurisation, but there is also hybrid systems available, see **Figure 5.10**. The flooded type is more robust and oil is cooling the gas inside the compressor increasing the possible pressure ratio for one step compression (Brown, 2005). The drawback with the oil flooded type is that the oil must have lubricating abilities, so an additional oil treatment system will be needed and oil lost to the gas must be replaced. The quality of the lube oil need to be expensive synthetic oil like ISO grade VG68. When compressing heavy hydrocarbon gas (C6+) the lube oil can be diluted. The heavy hydrocarbon gas makes the oil viscosity decrease which reduces the lubricating effect. This will increase wear and may cause compressor failure. Oil flooded screw compressor is therefore seldom used for heavy hydrocarbon gas found in well stream processing. (Fujimatsu, 2009)

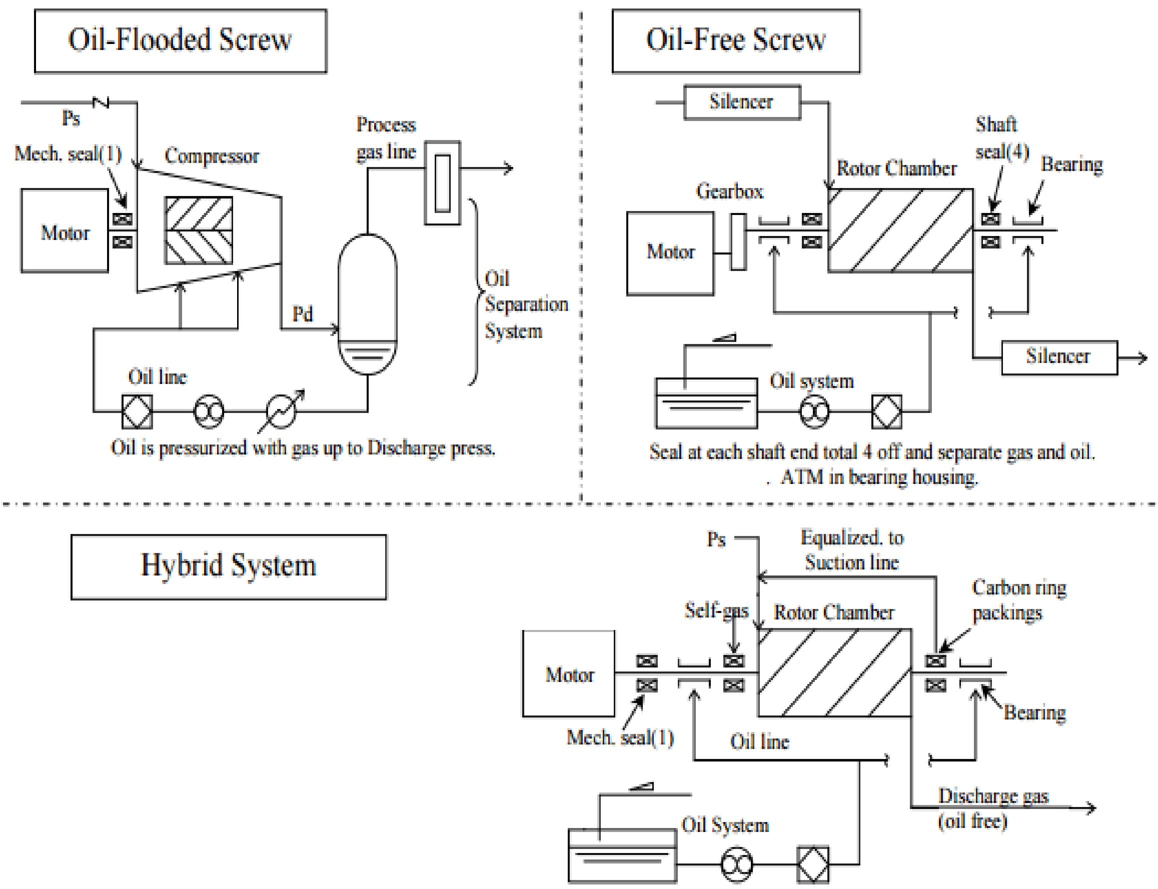



**Figure 5.10 Schematic diagram for oil-free and flooded screw compressor**

(Fujimatsu, 2009)

For the oil-free compressor one of the largest advantages is that it can handle some liquid. It is possible to add liquid in the compressor which cools the gas during compression. Clearances need to be kept small compared to a flooded compressor, to avoid large backflow of gas in the compressor and get a good efficiency. Some fouling will actually be an advantage for the oil-free compressor as it will increase efficiency by minimizing clearances. The drawback with the oil-free type is that it needs a gearbox for timing the screws and a silencer to avoid high frequency pulsation. (Brown, 2005)

The hybrid screw compressor contains special futures of the oil-free and flooded compressor. Lube oil is stored in a tank under suction pressure, which reduces the amount of heavy hydrocarbons diluted in the lube oil, compared to the flooded type where the lube oil is contained at the outlet pressure. Lube oil for the hybrid compressor can be mineral oil, which is cheaper than the synthetic oil. As for the oil-free compressor there is smaller clearance than in the flooded compressor, which reduces quantity of lube oil used. (Fujimatsu, 2009)

The screw compressor can operate in a wide range of flow rates and pressures. This compressors has a range from about 200m$^3$/h to 100 000m$^3$/h, see **Figure 5.11**. (MAN-Disel&Turbo, 2016)

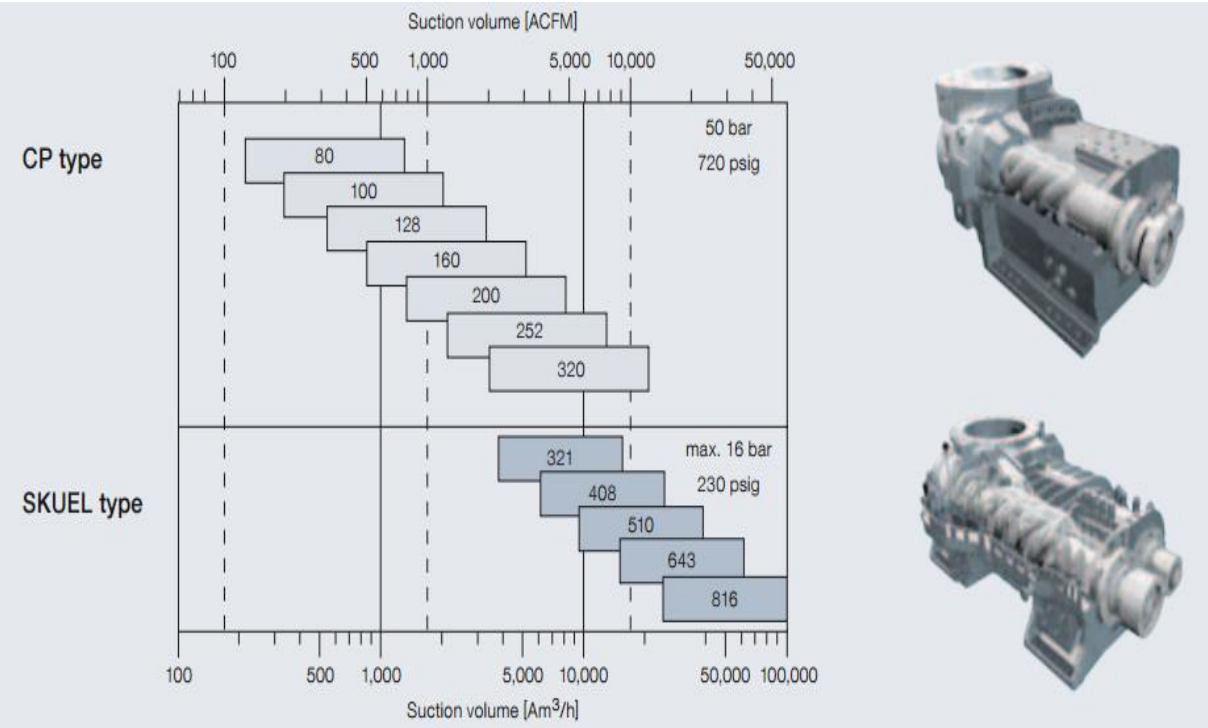



**Figure 5.11 Performance for MAN oil-free screw compressors** (MAN-Disel&Turbo, 2016)

### *5.1.5.2 Conical screw compressor*

To improve efficiency of screw compressors, in special small compressors, reducing backflow is important. To improve efficiency a new single screw compressor design with conical shape is under development, see **Figure 5.12**. This compressor has a much smaller backflow rate than conventional twin-screw compressor, giving it improved efficiency. The volume of the working chambers is at least half the size of a twin-screw compressor with the same rotor size. Both smaller size and less back flow makes this design applicable for low flow rates.

**Figure 5.12 Conical screw compressor, internal and external screw elements.**

(Dmitriev & Tabota, 2014)

One compressor tested for this design is a oil-flodded type, HRC version MK4, which can compresse air from atmospheric pressure up to 23bara at 1000rpm. In **Figure 5.13** oultett pressure and flow rate is presented for the HRC version MK4 conical screw compressor. Further development is in progress to make an oil-free machine similar to the MK4 but with a compression ratio of 1:4. (Dmitriev & Tabota, 2014)

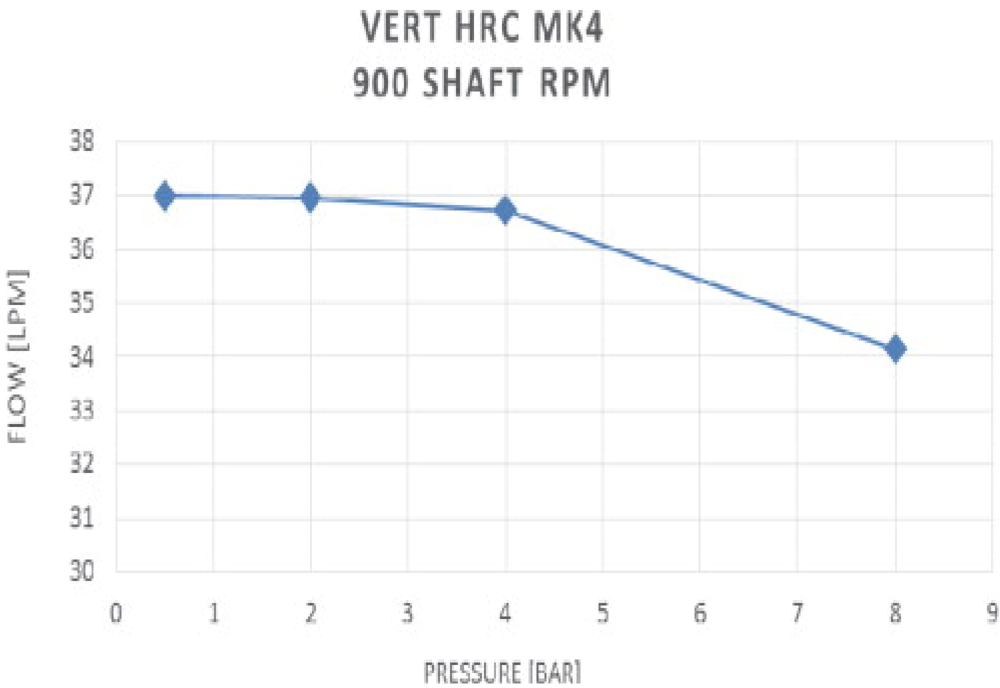



**Figure 5.13 Conical screw compressor pressure flow diagram, compression of atmospheric air, type: HRC version MK4,** (Dmitriev & Tabota, 2014)

### *5.1.6 Reciprocating compressors*

Reciprocating compressors can handle low flow rates and high pressure ratios better than a centrifugal compressor. One drawback is that it has a large number of moving parts which is subject to wear, giving low reliability and high maintenance (Brown, 2005). Large size and weight is also a drawback compared to kinetic technologies (Campbell, 1992).

Use of a reciprocating compressors subsea seems less attractive than other alternatives due to lower reliability, higher maintenance, larger size/weight and need for technological development. Therefor reciprocating compressors will not be further evaluated in this study.

### 5.1.7 Twin-screw multiphase pump

The twin-screw is based on the same principle as conventional screws, but can handle multiphase fluids. This can be an option to use for the recompression part shown in **Figure 2.1**, but the low flow rates can be an issue, see **Table 4.1**. One option can be to develop a new type of subsea screw compressor/pump for low flow rates of multiphase fluid, based on the conventional screw and twin-screw design.

#### 5.1.7.1 Description of twin-screw multiphase pump

Boosting of well stream pressure has increased and is pushing development on multiphase pumps forward. One benefit with a multiphase pump is that number of units decreases, e.g separators upstream the pump is unnecessary. Multiphase twin-screw pumps are robust equipment and can handle rougher conditions than many other pump and compressor technologies. A twin-screw pump is shown **Figure 5.14**. (Morrison, et al., 2014)

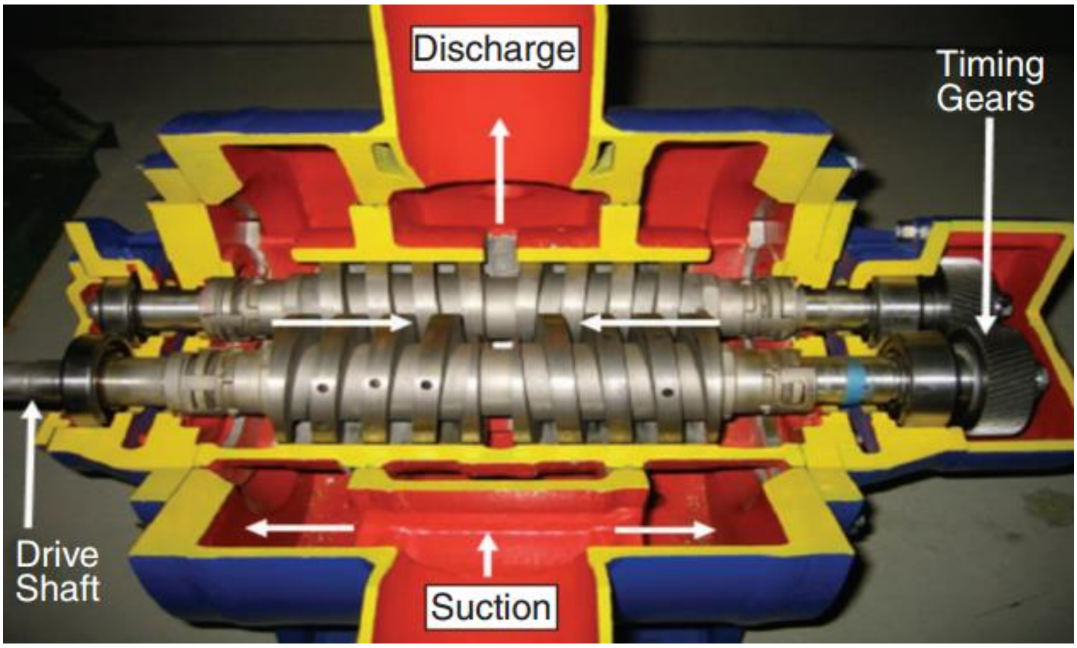



**Figure 5.14 Twin-screw pump** (Morrison, et al., 2014)

The subsea twin-screw pump SMPC series 4 are shown in **Figure 5.15**. This is a robust machine that is designed to operate for 5years without maintenance. The pump can operate with flow rates from about 100-5000m$^3$/h with gas content 0-95% without influencing the capacity. If it is up to 100% gas content in the feed recirculating or adding some liquid can be an option.  In **Figure 5.15** the performance data for five pump sizes (6 to10) selected for subsea implementation is show, but the vendor Bornemann has smaller and larger pumps used topside (size 2 to 12).

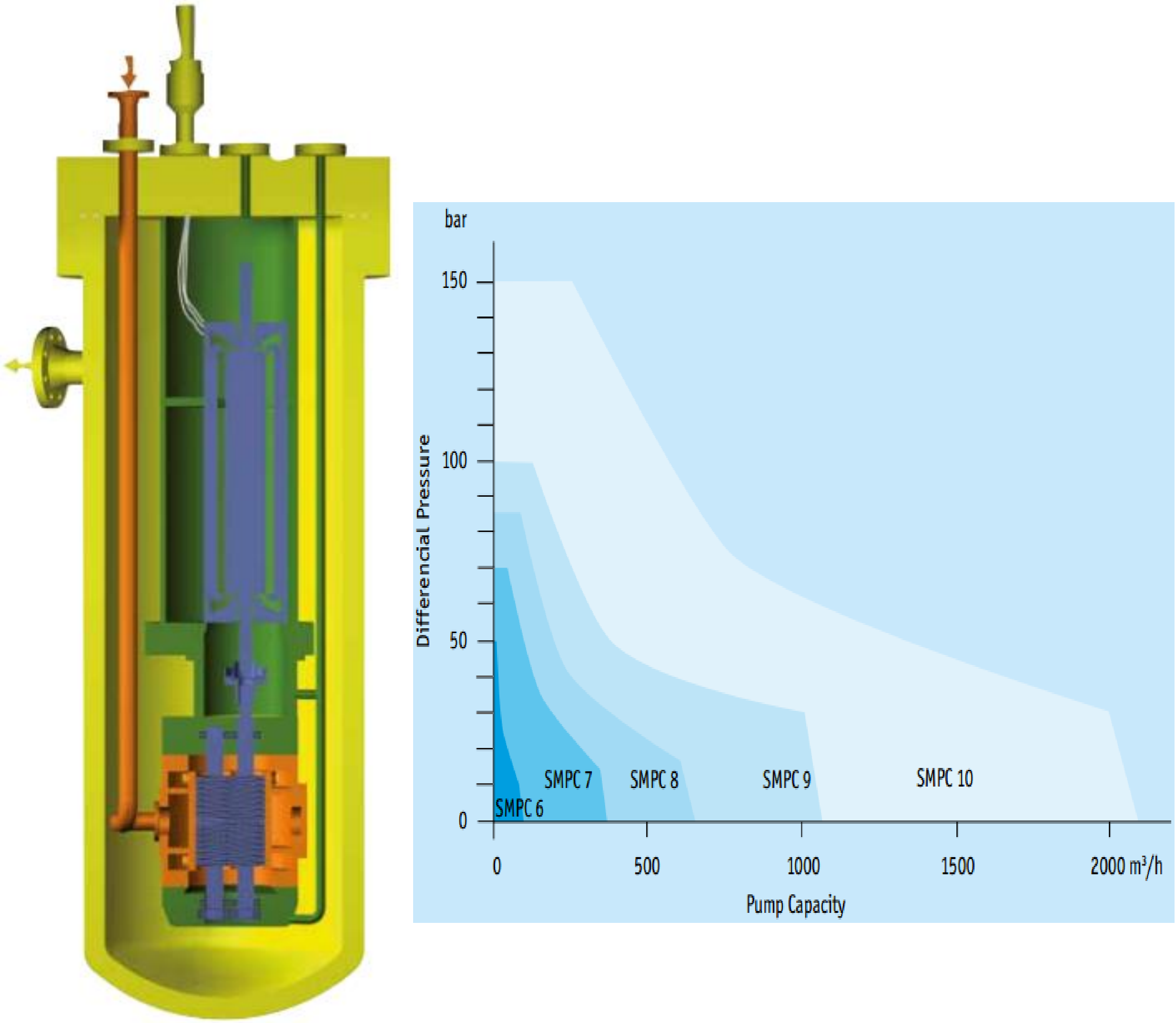



**Figure 5.15 Subsea Multiphase twin-screw Pump Compressor with performance data for the SMPC series 4** (Bornemann, 2016)

### *5.1.8 Helico-axial multiphase pump*

The helico-axial multiphase pump is the most used multiphase pump at this time. The pumps is operating with flow rates in the range 280m$^3$/h to 600m$^3$/h, this range is found satisfying for recompression in the rich case but not the lean case in **Table 4.1**. Limited pressure difference has been an issue for this pump, but development has made it possible to reach a differential pressure as high as 130bar. Since the machine has problems operating with low flow rates, using a screw compressor will probably be a more suitable machine for the recompression part. But the operational experience will be an advantage if the recompression flow rate is found large enough.

### *5.1.8.1 Description of helico-axial multiphase pump*

Helico-axial multiphase pumps have impellers providing kinetic energy to the gas and diffusers which increases the pressure. This is similar to an axial compressor or a centrifugal compressor, but due to the unique design of the helico-axial pump gas-liquid separation and gas-locking phenomenon is avoided, allowing the pump to operate with higher gas fractions than conventional pumps. Typical design of a helical-axial multiphase pump is shown in **Figure 5.16**. To control the gas fraction at the inlet partial liquid extraction and recirculation is needed. (Kuchpil, et al., 2013)

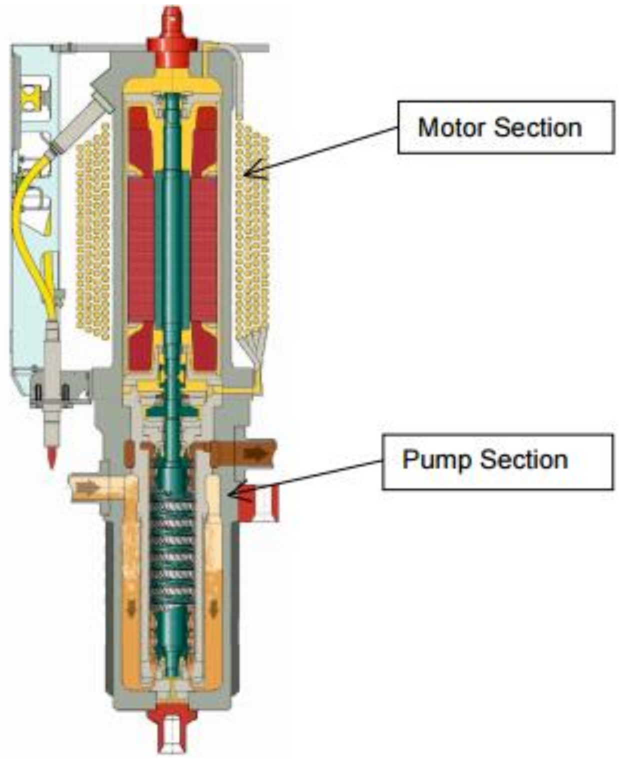



**Figure 5.16 Helico-axial multiphase pump impeller, FRAMO** (Kuchpil, et al., 2013)
**Typical subsea multiphase pump cross Section, OneSubsea** (Souzea, et al., 2013)

Compared to a twin-screw multiphase pump, the helico-axial multiphase pump has a simpler mechanical design, making them more tolerant for sand and is more compact for the same flow conditions. On the other hand operational performance of a helico-axial multiphase pump is limited for fluids with high gas fractions or high viscosity. (Kuchpil, et al., 2013)

The first subsea helico-axial multiphase pump was installed in 1994 at the Draugen field in the North Sea. This pump was designed for 35 bar differential pressure, capacity of 210m$^3$/h with a gas fraction of 30vol%. (Souzea, et al., 2013)

One of the most severe drawbacks with the helico-axial pump has been that the maximum differential pressure is in the range 40-60 bar. Improved design of the helico-axial multiphase pump has been done with focus on higher pressure boost. The design was tested and found to deliver 160bar differential pressure with a gas fraction of 60vol%. The first pilot is installed at the Barracuda field offshore Brazil in 2012, with a limitation of 70 bar differential pressure and 70vol% gas fraction, and 64-126 bara suction pressure. In 2015 helico-axial multiphase pumps where set in operation subsea at Girassol offshore Angola. The pumps have differential pressure limitation of 130 bar, design gas fraction of 20-50vol%, at a flow rate of 600m$^3$/h operating power of 2,5MW. (Souzea, et al., 2013) (Bibet, Huet, & Åsmul, 2016) (Rodallec & Delourme, 2016)

### 5.1.9 Pumps

Subsea pumping is the most mature technology for subsea pressure boosting, and can operate in a wide range of process parameters. The main subsea technology for single phase pumping is a centrifugal pump, and will be the preferred choice for pressurisation of liquid in the process shown in **Figure 2.1**. There are also hybrid pumps available with both helico-axial and centrifugal technologies available. The hybrid pump can be used for fluids with low to medium gas volume fractions as shown in **Figure 5.17**.

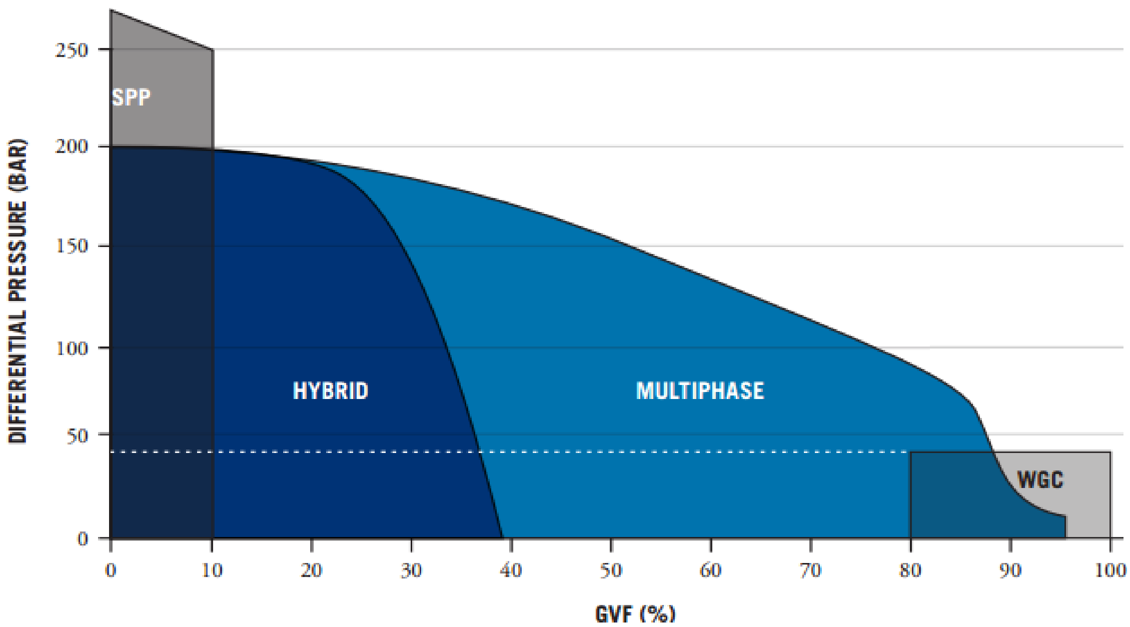



**Figure 5.17 Pumps and wet gas compressors, differential pressure for given gas volume fraction** (OneSubsea, 2016)

## 5.2 Separators

In this study it is focused on use of simple gravitational separator to provide robustness and low maintenance need to the subsea process plant. The separators must be able to handle two or three phase separation and solids/sand coming into the separator. Separator selection must be considered individual for each field development, and will depend on general factors for subsea development presented earlier, and special factors like water depth, internal/external pressure, separator efficiency and number of phases. There are a large number of separator designs available and only a small selection is presented in this section and in **Appendix F**.

### 5.2.1 Conventional vessel separator

Most of gravitational separators in operation are types of vessel separators, which can separate two or three phases. These separators consist mainly of a large vessel with internal equipment to ease separation, see **Figure 5.18**. Three-phase vessel separators are in operation subsea to separate produced water from the well stream in the North Sea fields Troll C and Tordis, and offshore Brazil at the Marlim field. (Olson, Grave, Juarez, & Anderson, 2014)

The separator in **Figure 5.18** is designed for use subsea at shallow water, and is now qualified for subsea use, see **Table 5.3** for data. At the separator inlet there is an inlet vane diffuser which breaks momentum and provides bulk separation, without making small droplets that is difficult to separate. The next internal equipment is perforated baffles, which straighten the oil and water flow pattern, avoiding turbulent mixing of the phase. Due to reliability concerns, use of internals downstream the perforated baffles is avoided in this design. In the bottom of the separator it is sand-removal devices. The water-retaining weir separates the oil/water outlets, and a dome with cyclones separates out the last droplets of the gas. (Olson, Grave, Juarez, & Anderson, 2014)

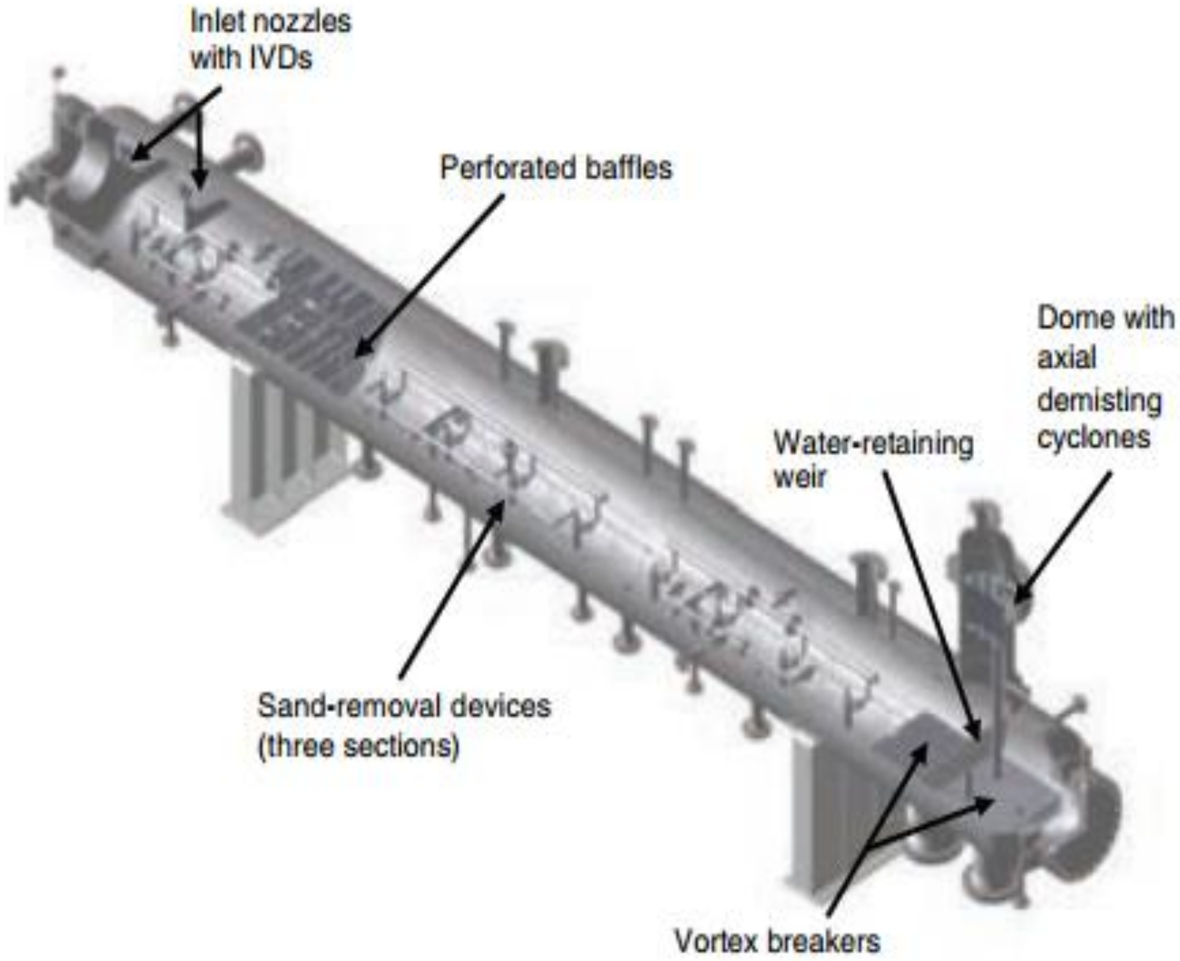



**Figure 5.18 Three-phase separator for use at shallow water (Olson, Grave, Juarez, & Anderson, 2014)**

**Table 5.3 Data for the three phase separator in Figure 5.18**

| Designation and unit | Design | Operating |
|---|---|---|
| **Pressure (barg)** | 240 | 45 |
| **Temperature (°C)** | 93 | 60 |
| **Water depth (m)** | 1-500 | |

### 5.2.2 T-Separator

The T-Separator in **Figure 5.19** is intended to use as an subsea inlet separator, separating the bulk of gas and liquid from in the well stream (Statoil, Holm, Bakke, & Gunnerød, 2014). It has a simple and compact design, making it robust and suitable for high pressures separation subsea. The drawback with this design is low separation efficiency, and no possibility for separation of water and liquid.

It is seen as a possibility to use the T-Separator to take out high pressure gas that can be used as motive gas in an ejector.

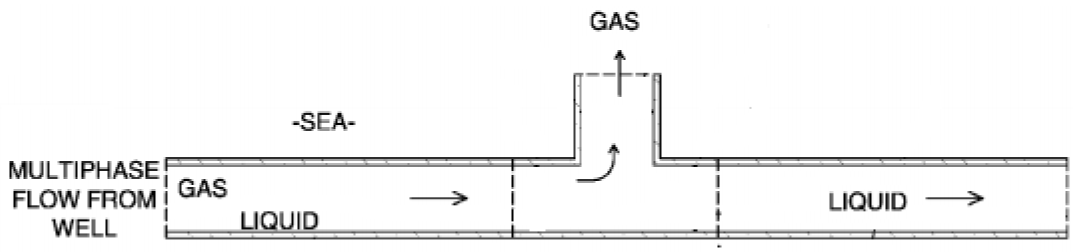



**Figure 5.19 T-Separator** (Statoil, Holm, Bakke, & Gunnerød, 2014)

### 5.2.3 Pipe separators

In subsea developments pipe separators has gained popularity. Compared to a conventional vessel separator, a pipe separator requires much smaller wall thickness for the same pressure. If using standard pipe sizes the manufacturing costs can be reduced, as they are easily available and easy to fabricate. Two types of two-phase pipe separators is shown in **Figure 5.20**. (Prescott, Mantha, Kundu, & Swenson, 2016)

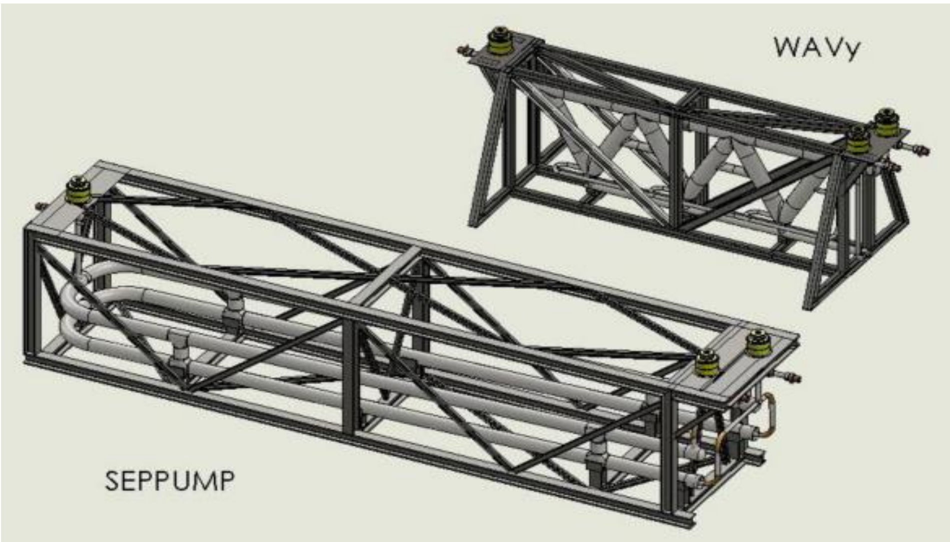



**Figure 5.20 Pipe Separators (SEPPUMP and WAVy)** (Prescott, Mantha, Kundu, & Swenson, 2016)

## 5.3 Subsea heating and cooling solutions

### *5.3.1 Subsea heating*

Subsea heating is mainly done for hydrate privation so far, and in **Figure 4.2** it is clear that sufficient subsea heating systems is not yet operating subsea. So it can be expected that development on subsea heaters is needed for use in the stabilisation process.

The subsea heater that was found available is a medium voltage heater developed by Gaumer Process. The use of medium voltage means that smaller cables can be used, which is an advantage for a subsea system powered from shore. The heater has a large duty ranging from 0.5MW to 30MW. (GaumerProcess, 2016)

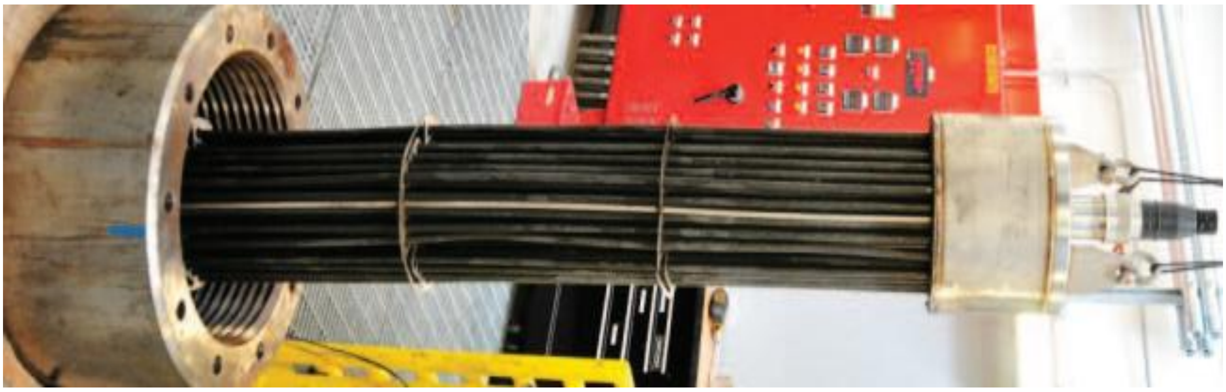


**Figure 5.21 Subsea Heater, 5000V, 3500m depth**

### *5.3.2 Subsea cooling*

For subsea cooling there is now both active and passive coolers available, see **Figure 5.22**. In the subsea process plant there is found to be a large need for cooling, see **Table 4.1**, and probably both active and passive coolers will be used in the system.

The active cooler shown on the left side in **Figure 5.22** is a conventional shell and tube heat exchanger based on forced convection created by a pump. The typical operation for this type of cooler is about 10MW to 20MW cooling capacity, and it can cool the process fluid down to 10°C above seabed temperature (KongsbergMaritime, 2016). Compared to a passive cooler, the active cooler with forced convection provides better temperature control, and lower heat transfer area due to increased heat transfer coefficient. Temperature control within the process will be important to reach product specifications, and to avoid hydrate formation by keeping the temperature above the hydrate line.

The passive cooler shown on the left side of **Figure 5.22** has a simpler and more robust design than the active cooler. Here it is natural convection providing heat transfer, as the surrounding water is heated it rises and is replaced with colder water due to density differences (KongsbergMaritime, 2016). The passive cooler has limited temperature control, assuming that there is not possible to control the internal flow in a large degree, the only way to control the temperature will be to add or remove heat transfer areal/units. The passive cooler can be a good alternative for cooling the product streams in a subsea process plant, such as the rich gas cooler in **Figure 4.4**, where there is only an upper temperature limit that must be reached.

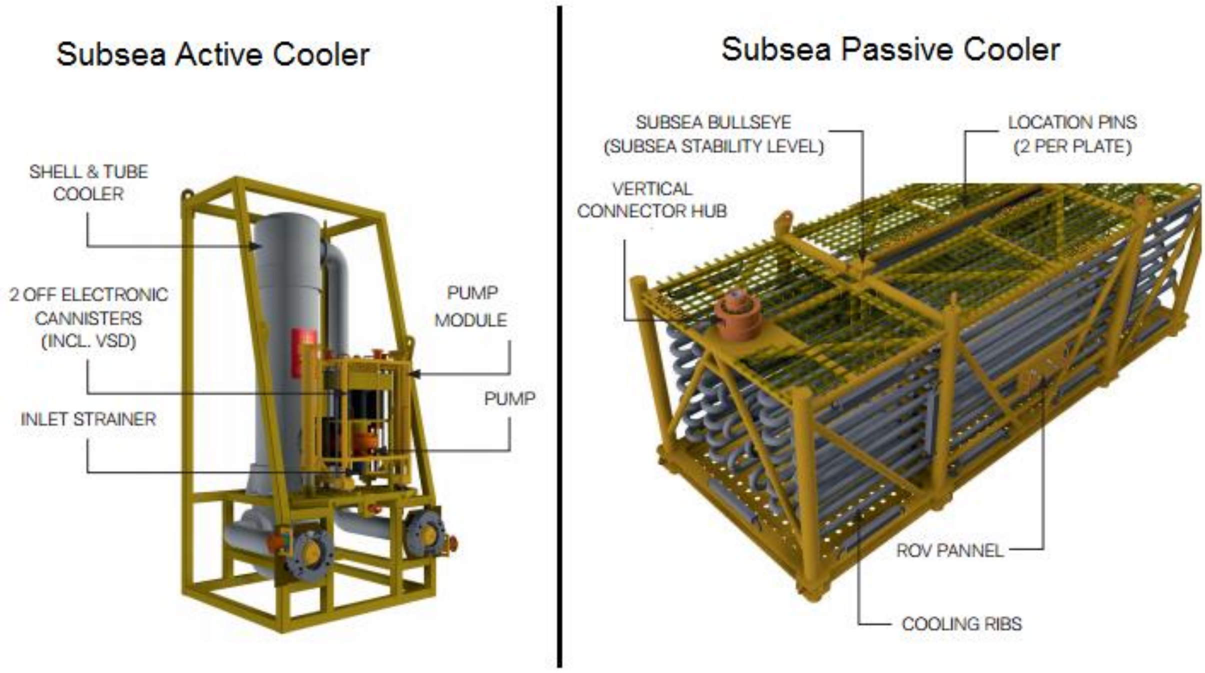



**Figure 5.22 Subsea Cooler Systems** (KongsbergMaritime, 2016)

# 6. Process design and analysis

In this chapter process designs and analysis is presented. The framework for design and analysis is given in **Chapter 3**. The process designs has it foundation from a conventional topside processing and earlier research on subsea processing, which is presented and discussed in **Chapter 4**.

One significant specification for the process design and process parameters it the liquid vapour pressure specification. The impact of this specification is presented in **Section 6.1**. This sensitivity analysis provides insight in why partial stabilisation is an interesting approach for subsea processing.

Earlier research on partial stabilisation of liquids has found feasible systems, but further development and evaluation is needed to improve these systems. The two-stage solution presented in **Section 4.2.2** is selected for further development. This two-stage process is a simplified version of a conventional topside process, with regards on subsea processing. The solution presented in **Section 4.2.1** with a subsea distillation column is seen to complex and energy intensive for use subsea. In addition an extensive qualification program will be needed to implement this system subsea. Selection of equipment for the system is discussed in **Chapter 5** and further addressed in this chapter.

One of the most severe issues in stabilisation of liquid hydrocarbons are recompression of flash gas from the stabilisation process. Different designs for the recompression systems are seen to have significant impact on overall system complexity and operational flexibility. Therefore the design is done with emphasis on recompression of flash gas, see **Section 6.2.**

There is also need for gas dehydration, as the gas is saturated with water, which needs to be removed to avoided condensation and hydrate formation. Different solutions is presented in **Section 4.3**. The co-current glycol dehydration process presented in **Section 4.3.1.2** is selected for further development and analysis for gas dehydration. This system is found to be in compliance with the subsea design philosophy, and is able to reach Rich gas specifications. The reason for this selection and Further analysis of the absorption technology is given in **Chapter 7**.

## 6.1 Impact of liquid product vapour pressure

In this section a brief analysis of variation in process parameters is presented. The analysis is done using a system with two screw compressors and a glycol dehydration system with two equilibrium stages, see **Figure 6.1**. In this system there is also applied a heater upstream of the LP separator. Together with pressure reduction this can be used to reach lower vapour pressure for the liquid product, by boiling of light hydrocarbons.

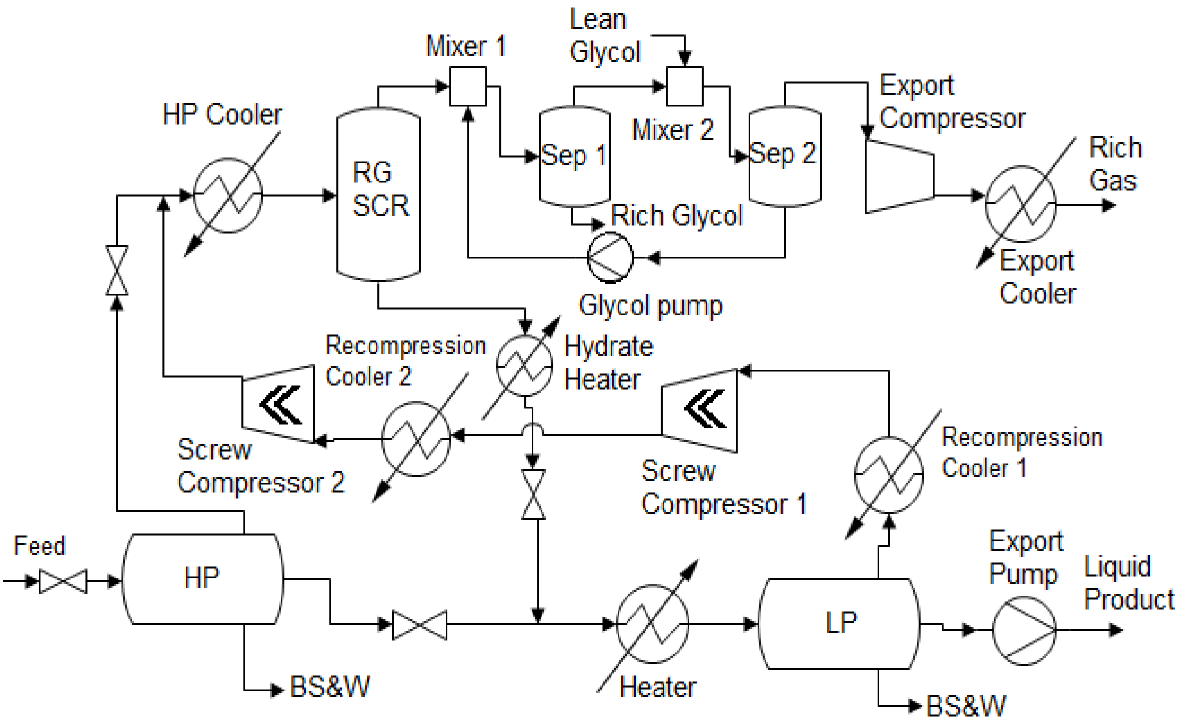



**Figure 6.1 Process used for analysis of process parameters, Dual screw compressor with Heater**

In **Table 6.1** the system in **Figure 6.1** is tested for changes in liquid product vapour pressure. The three different scenarios are discussed earlier in **Section 3.3.3.1** and are presented in **Table 3.5**. One scenario is that the liquid is completely stabilised for transport in regular oil tankers at atmospheric pressure, this is the strictest specification with TVP<1bara. There is also available ships that can transport liquid under pressure and use cooling to keep the pressure down during transport, this restriction is set to TVP<5bara. The last and least restricted scenario is that the liquid is transported at high pressures by pipelines to shore, then the vapour specification is set to TVP<10bara.

For the lean feed scenario all the different vapour pressure specifications seems feasible. For complete stabilisation, the low pressure of 3bara in the LP separator is a concern in a subsea environment with high external pressure. Low internal pressure can make it operate below hydrostatic pressure subsea which will be an issue in design of the subsea separator. Another problem with low pressure is that sea water can leak into the system and damage components and the liquid product. The differences between other processes tested with TVP<10bara is found small for scenario with lean feed, see **Table 6.1** and **Figure 8.3**. The reason for this small difference is the low liquid production, making the lean feed less sensitive to the stabilisation process and vapour pressure specification. The total power consumption is just above 5MW, with less than 10% power increase from the other solutions tested for lean feed, the only real additional power consumption is because of the Heater. The same behaviour is found for the cooling load. The cooling load will increase with increasing heat input and power consumption for compressors. In this system there is not added a cooler on the liquid product line, which reduces some of the impact the Heater has on cooling load.

In the rich feed scenario the differences is much larger. In **Section 6.2,** solutions are tested without the Heater, and with TVP<10bara. The total power consumption is found to be around 5MW, and the total cooling load around 10MW in the rich feed cases. Adding a heater to reach a vapour pressure of TVP<10 bara, increases the pressure in the LP separator, but it also increases total power consumption with about 50% when operating the LP separator at 100°C, see **Figure 8.4**. The reason for this large increase in the rich feed compared to the lean feed case, is that there is a much larger liquid flow rate when the feed is rich. Stabilisation to a TVP<5bara seems feasible, see **Table 6.1**. It may be considered to lower the pressure in the LP separator to reduce heat input and cooling load.

Complete stabilisation of rich feed to TVP<1bara, is not feasible with a two-stage system, see **Table 6.1**. The reason for this is the equilibrium relation in the LP separator. There are only two equilibrium stages, and a lot of light hydrocarbons in the liquid is going to the LP separator. When mixing in even more, coming from the cricondenbar control, the mixture into the LP separator gets even lighter. To flash these light components, pressure can be reduced and/or the temperature can be increased, according to equation (1) in **Section 3.3.4**. This makes also heavier hydrocarbons flash of going back to the cricondenbar control, condenses and is sent back to the LP separator. In the end HYSYS seems to accumulate a large stream going in this loop between the LP separator and the cricondenbar control, giving the extremely large numbers in **Table 6.1**. Another system with three equilibrium stages is able to handle complete stabilisation much better, see **Section 6.1.1**.

**Table 6.1 Impact of liquid product TVP (PFD Figure 6.1, Dual screw compressor with heater)**

| Designation and unit | Complete stabilisation | Stabilised for ship transport | Stabilised for pipe transport | Complete stabilisation | Stabilised for ship transport | Stabilised for pipe transport |
|---|---|---|---|---|---|---|
| Feed composition | Lean feed | Lean feed | Lean feed | Rich Feed | Rich Feed | Rich Feed |
| Feed pressure \| temperature (bara \| °C) | 250 \| 100 | 250 \| 100 | 250 \| 100 | 250 \| 100 | 250 \| 100 | 250 \| 100 |
| Liquid product TVP@37.8°C (bara) | 1 | 5 | 10 | 1 | 5 | 10 |
| Rich gas product (Sm3/d) | 3 000 446 | 2 998 662 | 2 997 684 | 2 894 907 | 2 929 476 | 2 903 782 |
| Liquid product (tonne/d) | 77 | 81 | 83 | 1885 | 1954 | 1995 |
| Total power consumption (kW) | 5666 | 5425 | 5390 | 391 275 | 9336 | 7420 |
| Total Cooling load (kW) | 9138 | 8897 | 8855 | 396 717 | 13 121 | 11 176 |
| Total Cooling area (Note 1;Note 2) ($m^2$) | 207 | 200 | 199 | 11 586 | 346 | 284 |
| Power consumption compressors (kW) | 5196 | 5169 | 5159 | 73 490 | 4789 | 4407 |
| Heating utility (kW) | 454 | 240 | 215 | 317 370 | 4159 | 2643 |
| Outlet temperature Heater (°C) | 100 | 100 | 100 | 100 | 100 | 100 |
| Glycol circulation rate (99,5%MEG) $m^3/h$ | 0.6 | 0.6 | 0.6 | 0.5 | 0.5 | 0.5 |
| Pressure levels HP \| LP (bara) | 70 \| 3 (Note 5) | 70 \| 9 | 70 \| 15 | 78 \| 3 (Note 5) | 78 \| 10 | 78 \| 17 |

Note 1: Sea water temperature is assumed 5°C constant through the whole cooling process.
Note 2: Assumed total heat transfer coefficient, U=800 $W/m^2K$ for active subsea coolers. All coolers is assumed active in this calculations
Note 3: Temperatures is below 130°C in all compressors
Note 4: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.
Note 5: The large pressure difference in the complete stabilisation process gives pressure ratio for the compressor close to 5:1, which is higher than the limit set at 4:1 for this study.

### 6.1.1 *Three-stage system for complete stabilisation of liquid*

The two stage system is not feasible for complete stabilisation of rich feed, due to the equilibrium relation in the LP separator. To decrease the amount of light hydrocarbons in the LP separator a MP separator is added to the system, see **Figure 6.2**. In addition three screw compressors are used for recompression of flash gas to keep the pressure ratios below 4:1. This system is much more complex than the other systems presented, with 4 compressors, and in total 21 units.

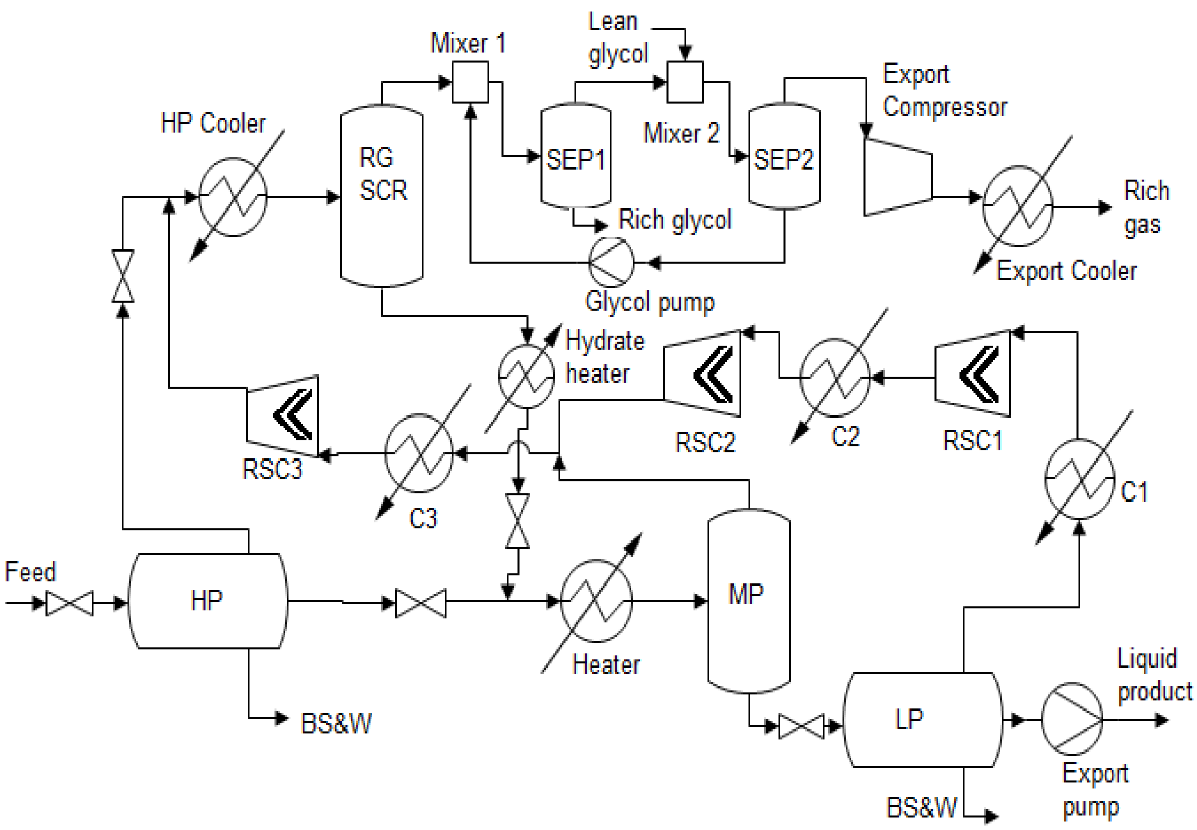



**Figure 6.2 Three-stage System for complete stabilisation of liquid**

Data from a simulation done on this system with rich feed gas is presented in **Table 6.2**. It is clear that this is much more realistic than the two-stage system for complete stabilisation. The power consumption is almost the same as for the scenario with TVP<5bara, and the cooling load and area is increased with about 10% for this system. For the systems with no heater and liquid produced to pipe specifications (TVP<10bara), the power increase is about 80% and the cooling load is about 40% higher for complete stabilisation (TVP<1bara).

**Table 6.2 Data for full stabilisation of rich feed (PFD Figure 6.2, Three-stage system)**

| **Designation and unit** | **Complete stabilisation** |
|---|---|
| Feed composition | Rich feed |
| Feed pressure \| temperature (bara \| °C) | 250 \| 100 |
| Liquid product TVP@37.8°C (bara) | 1 |
| Rich gas product (Sm3/d) | 2 953 794 |
| Liquid product (tonne/d) | 1908 |
| Total power consumption (kW) | 8930 |
| Total Cooling load (kW) | 14 075 |
| Total Cooling area (Note 1;Note 2) ($m^2$) | 392 |
| Power consumption compressors (kW) | 4798 |
| Heating utility (kW) | 3736 |
| Outlet temperature Heater (°C) | 100 |
| Glycol circulation rate (99,5%MEG) $m^3$/h | 0.55 |
| Pressure levels HP \| MP \| LP (bara) | 78 \| 23 \| 2 |
| Note 1: Sea water temperature is assumed 5°C constant through the whole cooling process.<br>Note 2: Assumed total heat transfer coefficient, U=800 W/$m^2$K for active subsea coolers. All coolers is assumed active in this calculations<br>Note 3: Temperatures is below 130°C in all compressors<br>Note 4: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications. | |

## 6.2 Process solutions with emphasis on recompression

When stabilising hydrocarbon liquid, light hydrocarbons flash at low pressure. The flash gas needs to be recompressed and mixed into the associated gas stream for further processing. In this section different process designs is presented with emphasis on gas recompression. The design is done regarding to the subsea design philosophy, including use of robust equipment, low power demand, operational flexibility and to use as few components as possible.

The maximum pressure ratio has been set to 4:1 for both compressor and ejector. In **Table 4.1** the pressure ratio from LP to HP is in the range 7:1-8:1, giving need for minimum two units for recompression of flash gas from LP to HP.

In the two-stage subsea process solution shown earlier in **Figure 4.4**, the pressure in the liquid stabilisation process is reduced by choke valves. Since there are valves used for pressure reduction, exergy is lost during throttling. By replacing the valves with equipment able to convert exergy to usable work, power consumption can be reduced. One solution analysed is use of ejectors that utilises pressure energy from a high pressure stream (motive) to increase pressure in a low pressure stream (suction). Potential for use of ejectors is high since there are relatively large flow rates going down in pressure compared to the recompression gas rate. The inlet flow, with 250 bar pressure, is 7 to 250 times larger than the flash gas stream from the LP separator, see **Table 4.1**. Ejectors are robust equipment with no moving parts and will provide robustness and lower power demand for recompression. See **Section 5.1.4** for more details around ejector technology, ejector analysis, and why ejector technology is selected for further evaluation.

Use of screw compressors is analysed to increase operational flexibility. The main reasons for selecting screw compressors is that fits the low flow rate in recompression, and is robust that can handle gas with entrained liquid. More details about screw compressor technology and why they are selected are presented in **Section 5.1.5**. As temperature will increase in compression this also requires one or more coolers to keep the temperature low. Screw compressors are robust and mature technology used in the oil and gas industry.

For gas processing the cricondenbar control is done in a conventional process where the gas is throttled, if needed, and cooled down to 25°C. This process condenses heavy hydrocarbons, and some water, which is separated from the gas in a downstream separator. Gas dehydration is done by use of glycol co-current contactors and separators, where lean glycol is imported and rich glycol is exported to a host. Cricondenbar control is done upstream of the gas dehydration process to reduce water load, avoid foaming from mixing of liquid hydrocarbons and glycol, and to lower the temperature to improve absorption.

As seen in **Section 6.1** and **Section 4.2.2**, use of a heater for stabilisation of liquid is energy intensive. The heater also adds complexity and will need technological qualification for subsea use, see **Section 5.3.1**. For the recompression part it will be more conservative with a low pressure in the LP separator, adding a heater will just add operational flexibility. So for the following process designs, that has focus on the recompression part, there is not used a heater upstream the LP separator, see **Section 6.1** for systems using a heater for stabilisation purposes. A small heater is used on liquid going from the cricondenbar control to the stabilisation process, to keep the temperature above 25°C and prevent hydrate formation. this type of heaters is already used subsea. Glycol for hydrate inhibition could be used here, but due to concerns about glycol going with the produced water this is avoided in this process.

### *6.2.1 Ejector upstream and downstream HP separator*

In **Figure 6.3** a system using two ejectors for recompression of flash gas is presented. Results from simulations are presented in **Table 6.3**, **Figure 6.4** and **Figure 6.5**.

In the two-stage solution presented in **Figure 4.4** liquid going from the HP separator to the LP separator provides potential for an ejector. To utilise some of the exergy lost in the throttling process in an ejector there is need for an additional pressure step. If the ejector is placed between the HP and LP separator without adding another pressure level, it will only lead to recirculation of flash gas. The pressure level is made by adding a separator between the HP and LP separator, represented by the MP separator shown in **Figure 6.3**. By placing an ejector upstream the MP separator this can pressurise flash gas from the LP separator, see **Figure 6.3**. The exit pressure of Ejector 1 and inlet pressure of MP separator depends on motive flow, suction flow and ejector efficiency.

When the flash gas leaves the MP separator the pressure is still too low for the gas processing section. So by adding an ejector to the feed stream, Ejector 2 in **Figure 6.3**, the flash gas can be compressed from MP to HP. To make it possible to control the pressure in the HP separator the feed stream is divided. Part of the stream enters the ejector while the other part goes through a throttling process. There is also possible to move the throttling upstream or downstream of Ejector 2, but then the mass flow going through the ejector is large, making need for a larger ejector.

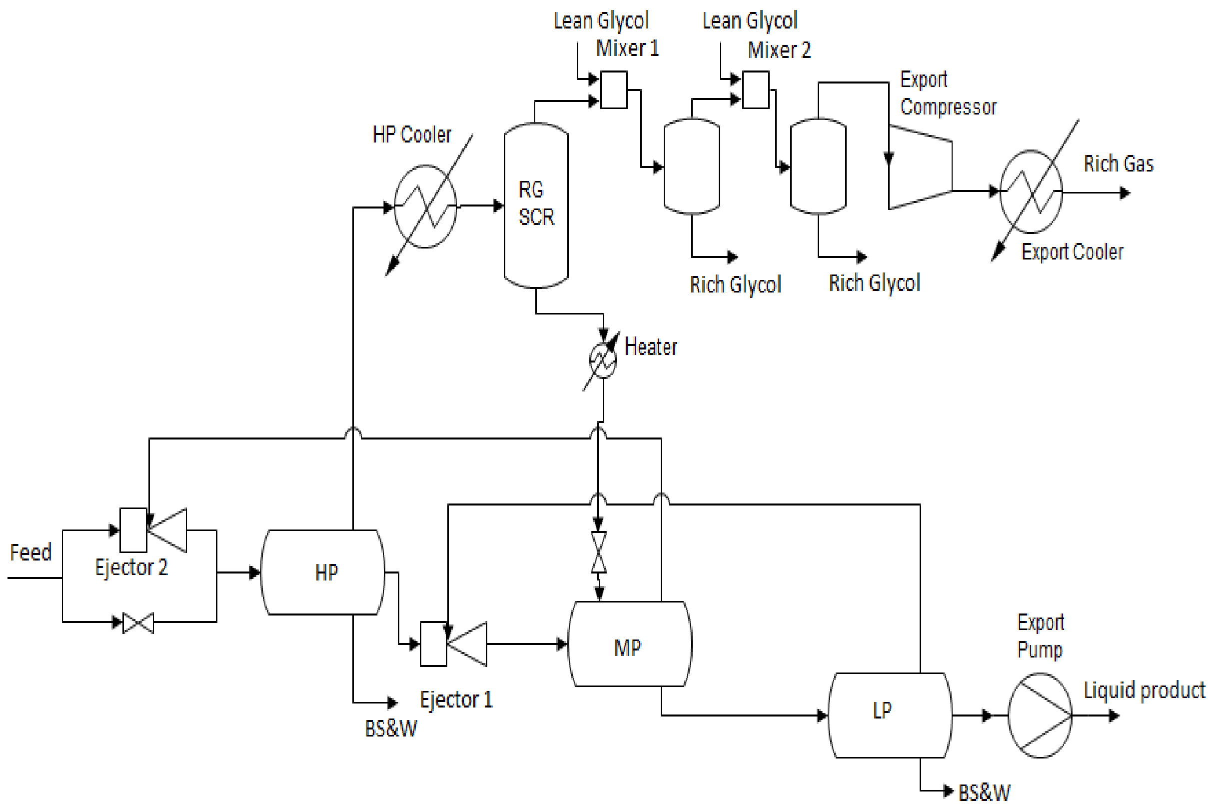



**Figure 6.3 Recompression using ejector upstream and downstream HP separator**

In **Figure 6.4** and **Figure 6.5** ejector performance for rich feed and lean feed is presented respectively. Ejector 1 utilises the total liquid stream from HP separator to increase the pressure as much as possible. In ejector 1 the motive pressure is 70-80 bara giving a large motive flow rate. The mass entrainment ratio, defined in Equation (3), is found to be as low as 0.025.

Ejector 2 is tested for decreasing inlet pressure from the wells to see how this impacts ejector performance. It is clear in both **Figure 6.4** and **Figure 6.5** that decreasing motive pressure gives a steep increase in required motive mass flow. In the simulations there is more motive fluid available, the total mass flow from the well is about 29kg/s and 54kg/s for the lean and rich feed respectively. This makes it possible to run Ejector 2 with even lower pressure, in special for the lean feed case where less than 1kg/s is utilised from the well stream.

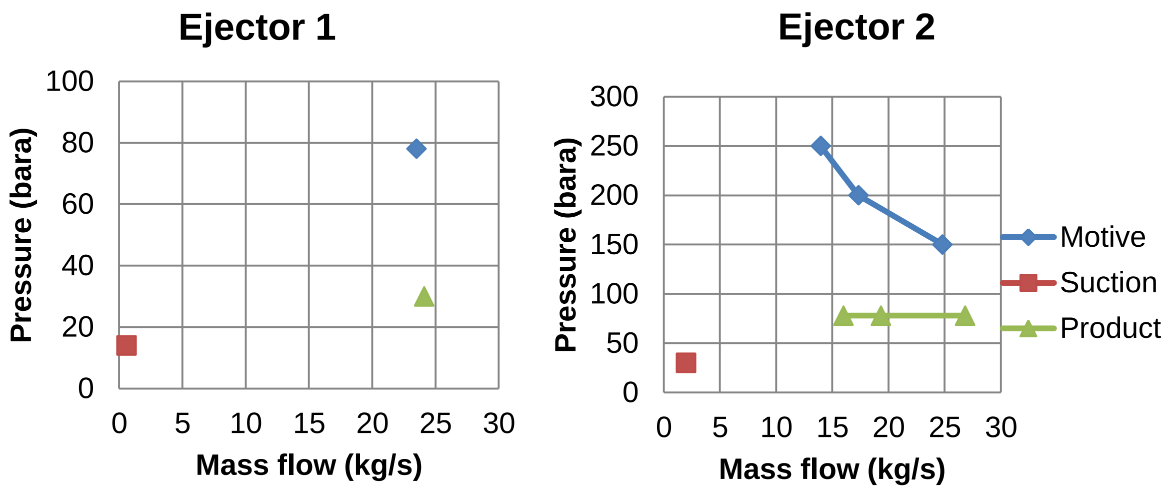



**Figure 6.4 Ejector performance rich feed processed to TVP<10bara (PFD see Figure 6.3)**

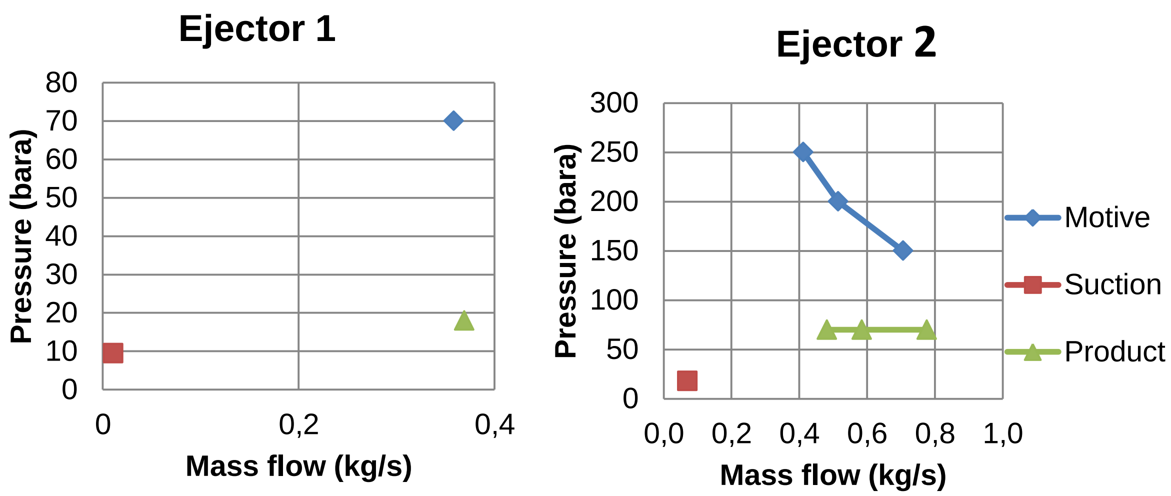



**Figure 6.5 Ejector performance lean feed processed to TVP<10bara (PFD see Figure 6.3)**

Placing ejectors upstream the HP and MP separator inlet yields some increase in light hydrocarbons in the HP and MP separator, compared to taking the flash gas directly to the gas processing system. This is analysed in **Figure 6.6** and **Figure 6.7**, and it is found that mixing flash gas with the well stream has low impact on the composition going into the HP separator. The reason for this behaviour is that the well stream is much larger than the flash gas stream.

Adding flash gas to upstream the MP separator has some impact on composition coming into the MP separator as shown in **Figure 6.7**, in special for the lean feed case with low liquid flow rate. But for the end product this mixing has no real impact, and the liquid product specifications are still reached.

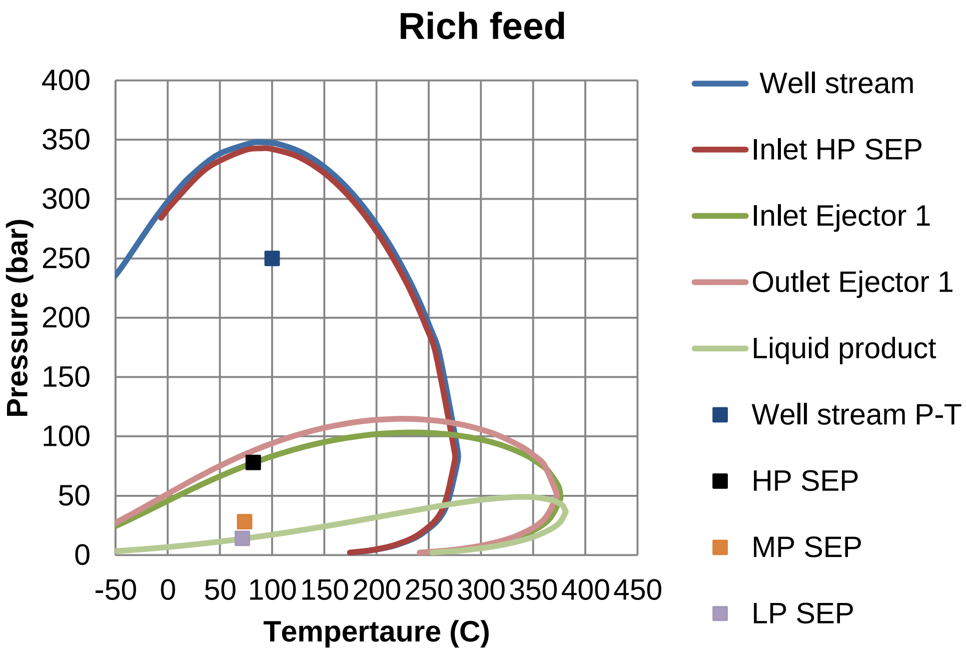



**Figure 6.6 Phase envelopes for Rich feed (PFD see Figure 6.3)**

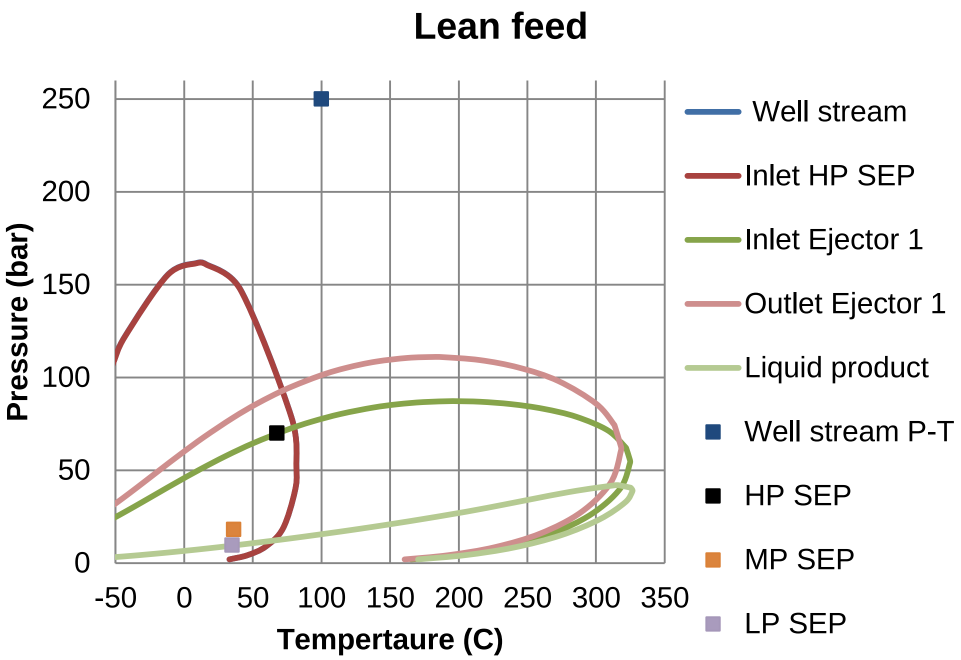



**Figure 6.7 Phase envelopes for Lean feed (PFD see Figure 6.3)**

In **Table 6.3** simulation data from the system shown in **Figure 6.3** is presented. Suction volume flow for the export compressor is found to be within the lower range for centrifugal compressor in **Figure 5.1**. It is also a clear advantage that there is no use of coolers or power for the recompression part lowering the number of units and power consumption.

**Table 6.3 Simulation data (PFD Figure 6.3 ejector upstream and downstream HP Sep)**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Liquid vapour pressure | TVP<10bara | |
| Total number of units | 15 | |
| Total power consumption (kW) | 5145 | 4432 |
| Total Cooling load (kW) | 8789 | 9702 |
| Total Cooling area (Note 2;Note 3) ($m^2$) | 197 | 235 |
| Rich gas product ($Sm^3$/d) | 2 996 685 | 2 882 758 |
| Liquid product (Actual $m^3$/d) | 112 | 2829 |
| Rich gas product (tonne/d) | 2346 | 2536 |
| Liquid product (tonne/d) | 85 | 2038 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3$/h) | $P_2/P_1$=200/70=2.9<br>P=5137<br>V=1549 | $P_2/P_1$=200/78=2.6<br>P=3989<br>V=1218 |
| Recompression power (kW) | 0 | 0 |
| Recompression cooling (kW) | 0 | 0 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 67.5 \| 25 \| 37<br>UA=94<br>Q=3516 | 82 \| 25 \| 42<br>UA=134<br>Q=5630 |
| Export cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=64<br>Q=5273 | 105 \| 60 \| 75<br>UA=54<br>Q=4072 |
| Heating utility (kW) | 8 | 63 |
| Export pump (Note 1) (kW) | 16 | 380 |
| Pressure levels HP \| MP \| LP (bara) | 70 \| 18 \| 9.5 | 78 \| 30 \| 14 |
| Note 1: Discharge pressure of export pump is set to 100bara<br>Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.<br>Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient U=800W/$m^2$K<br>Note 4: Temperatures is below 130°C in all compressors<br>Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.<br>Note 6: Ejector efficiency is set to 20% | | |

### 6.2.2 Two ejectors driven by the well stream

The process solution presentenced in this section is an improvement of the process shown in **Figure 6.3**, **Section 6.2.1**. The improved solution is shown in **Figure 6.8**. The intention of the improved design is to equalise pressure ratios for the two ejectors during recompression, and increase operational flexibility. Equal pressure ratios should provide a better overall efficiency for the ejectors.

In **Figure 6.3**, Ejector 1 is placed on the liquid line downstream of the HP separator. This results in a limited suction pressure ratio of about 2:1 in the first recompression stage, and low operational flexibility. There are different solutions that can be used to improve this. One solution is to add a cooler on the suction flow upstream Ejector 1. But this cooler provides additional equipment to the system. Another possibility is to increase the pressure of the HP separator. With a higher pressure in the HP separator the flash gas rate from the LP separator increases. This flash gas needs to be recompressed, adding extra load on the ejector system. The selected solution is shown in **Figure 6.8**, where some of the high pressure feed stream is used as motive flow for Ejector 1. Then the motive pressure can be held higher without affecting the HP separator.

The feed stream has a mass flow of 29kg/s and 54kg/s for the lean and rich feed respectively. In analysis of the process with only one Ejector 2 upstream HP separator, see **Figure 6.3**, there is still a part of the feed that is lead through a throttling valve. This stream can be used as motive flow in Ejector 1, see **Figure 6.4** and **Figure 6.5**.

The system presented in **Figure 6.8** could theoretically be operated with only two pressure steps for liquid stabilisation. But without the MP separator, the product stream from Ejector 1 would end as suction flow for Ejector 2. Another benefit with the MP separator is that more gas will flash off at higher pressure, reducing the load on Ejector 1. As an example, it where found a product stream from Ejector 1 on 3kg/s, where 2 kg/s was gas, to the MP separator. The two other streams connected with the MP separator where 25kg/s where 3kg/s was gas. Due to the MP separator the suction stream going to Ejector 2 is reduced to just above 3kg/s gas.

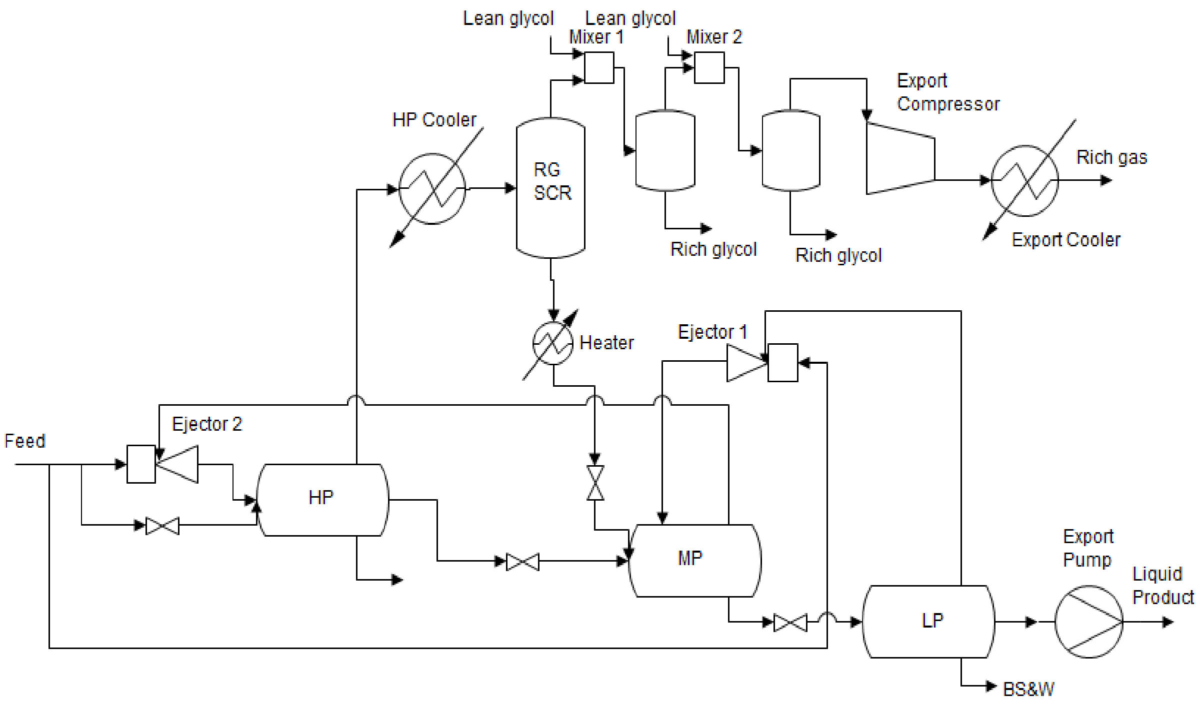



**Figure 6.8 Recompression with two ejectors driven by the well stream**

Ejector performance for rich feed is presented in **Figure 6.9**. This shows that the ejectors can be operated down to 150bara motive pressure. At this low pressure about 43kg/s of the 54kg/s available motive flow rate is utilised. So when there is a rich feed, the reservoir pressure can not go far below 150bara before the ejectors is unable to operate. Utilising the high pressure inlet stream as motive flow for Ejector 1 gives a significant reduction in motive flow rate compared to placing Ejector 1 downstream of the HP separator (PFD **Figure 6.3**), going from 23.5kg/s to 2-3 kg/s, see **Figure 6.4** and **Figure 6.9**.

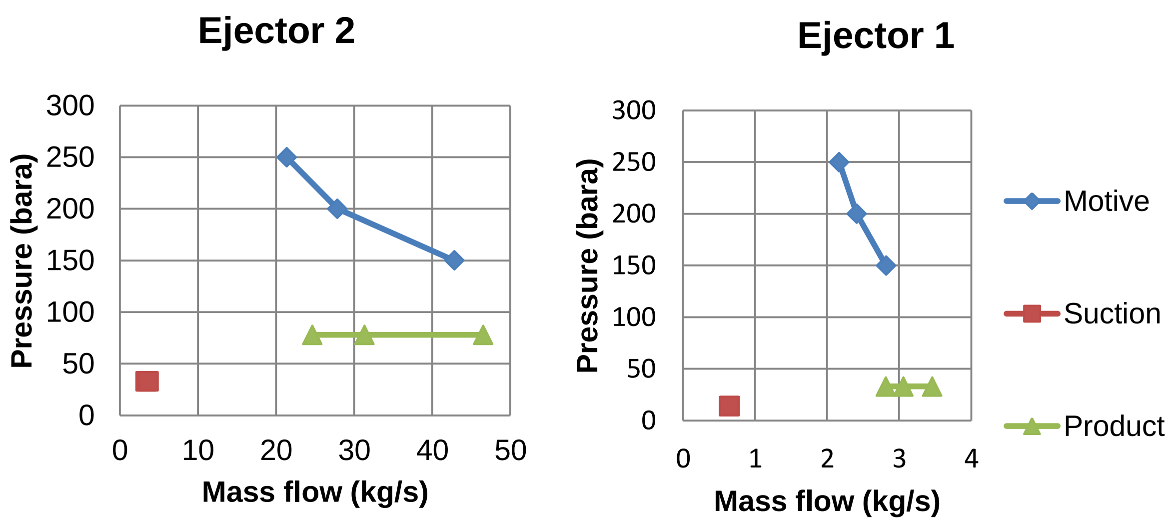



**Figure 6.9 Ejector performance, Rich feed processed to TVP<10bara (PFD see Figure 6.8)**

For the lean feed ejector performance is presented in **Figure 6.10**. Here it is found that available motive flow rate is much larger than what is needed in the ejectors, less than 1kg/s of the 29kg/s feed stream is utilised. This configuration increases flexibility to a large extent for Ejector 1, and it is chosen to equalise the suction pressure ratio for the two ejectors. Lowering the pressure ratio of Ejector 2 and increasing the mass entrainment ratio of Ejector 1 is expected to increase the ejectors efficiency, see **Figure 3.8**, but this is not taken into consideration in this study where 20% efficiency is used for all scenarios. Due to the large flowrate, the pressure from the reservoir can be much lower than 150bar without affecting operation of ejectors. Assuming that there is installed an ejector system with multiple ejectors in parallel to control changes in pressure and mass flow rate.

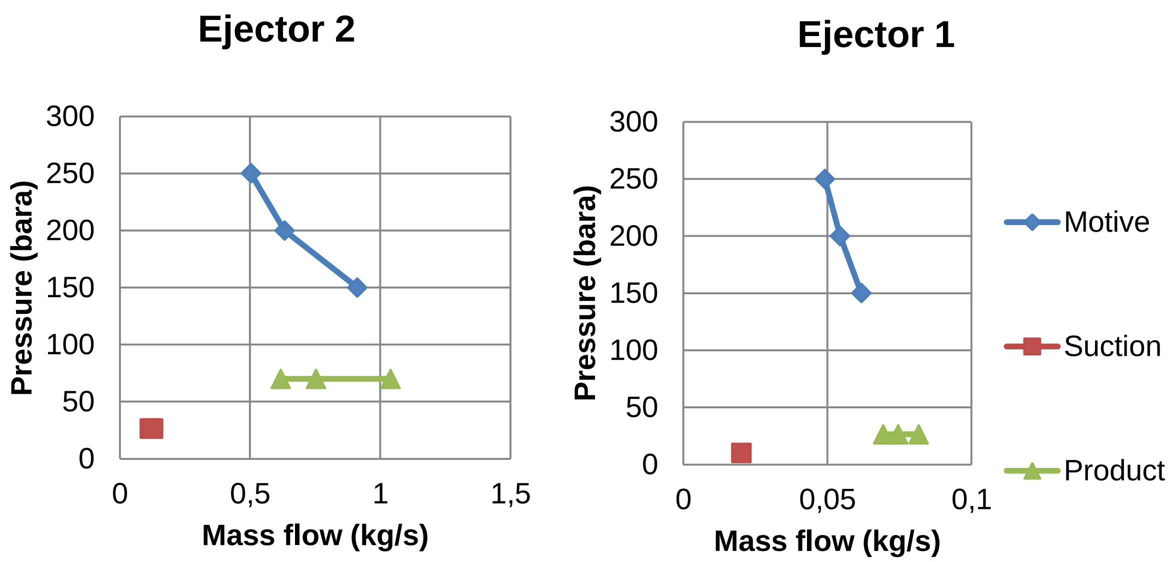



**Figure 6.10 Ejector performance, Lean feed processed to TVP<10bara (PFD see Figure 6.8)**

In **Table 6.4** data from simulations on the process in **Figure 6.8** is presented. There is not found to be a significant difference in this data compared to the results presented in **Table 6.3**. One difference is that pressure in the MP separator is increased from 18bara to 26.5bara and 30 to 33 bara for the lean and rich feed respectively. This provides equal suction pressure ratios over the ejectors. Based on simulation data and discussion in this section, it is preferred to utilise the feed stream as motive flow for both ejectors.

**Table 6.4 Simulation data, (PFD Figure 6.8 two ejectors driven by well stream)**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Liquid vapour pressure | TVP<10bara | |
| Total number of units | 15 | |
| Total power consumption (kW) | 5159 | 4424 |
| Total Cooling load (kW) | 8786 | 9766 |
| Total Cooling area (Note 2; Note 3) ($m^2$) | 197 | 236 |
| Rich gas product ($Sm^3/d$) | 2 996 663 | 2 885 596 |
| Liquid product (Actual $m^3/d$) | 113 | 2814 |
| Rich gas product (tonne/d) | 2346 | 2541 |
| Liquid product (tonne/d) | 86 | 2031 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3/h$) | $P_2/P_1$=200/70=2.9<br>P=5137<br>V=1549 | $P_2/P_1$=200/78=2.6<br>P=3988<br>V=1217 |
| Recompression power (kW) | 0 | 0 |
| Recompression cooling (kW) | 0 | 0 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD (Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 67.5 \| 25 \| 37<br>UA=94<br>Q=3513 | 82 \| 25 \| 42<br>UA=136<br>Q=5692 |
| Export cooler $T_1$ \| $T_2$ \| LMTD (Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=64<br>Q=5273 | 105 \| 60 \| 75<br>UA=54<br>Q=4074 |
| Heating utility (kW) | 6 | 58 |
| Export pump (Note 1) (kW) | 16 | 378 |
| Pressure levels HP \| MP \| LP (bara) | 70 \| 26.5 \| 10 | 78 \| 33 \| 14 |
| Note 1: Discharge pressure of export pump is set to 100bara<br>Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.<br>Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient U=800W/m2K<br>Note 4: Temperatures is below 130°C in all compressors<br>Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.<br>Note 6:Ejecor efficiency is set to 20% | | |

### 6.2.3 Screw compressor and ejector

In the first two designs presented there are placed ejectors on the feed stream coming from the well. The well stream will most likely have some variation in flow rate and flow pattern. If liquid slugs or pure gas columns enters the ejector, operational performance will be influenced. There are also flow assurance issues like sand, scale or other impurities that can affect the ejector operational performance. To increase controllability and avoid flow assurance problems, one alternative could be to place some kind of equipment upstream the ejector to remove sand and secure stabile flow through the ejectors. Another option is to use the HP separator to remove sand and stabilise the flow, and use the downstream gas in an ejector. This is the solution analysed in this section, see **Figure 6.11**. To provide some operational flexibility a throttling valve is used for the excess gas.

To use gas from the HP separator, the pressure must be high enough to drive an ejector with an exit pressure at least as high as the pressure in the cricondenbar control. The HP separator will be a bulky three-phase separator, where increased pressure will lead to increased wall thickness, weight and CAPEX. The HP separator is assumed to be a vessel separator with a maximum operating pressure of 100bara.

As seen for the system in **Figure 6.3**, where the ejector is placed on the liquid line downstream of the HP separator, the low pressure for the motive flow reduces the operational flexibility of the ejector. To increase operational flexibility a screw compressor could be used for the first pressure step, this will also remove the need for an MP separator.

In **Figure 6.11** a screw compressor with an upstream cooler is used for the first pressure step. There is a cooler upstream the screw compressor to lower the gas temperature to 25°C. Gas out of the LP separator is on the dew point, and some gas will condense in the cooler. It is assumed that the screw compressor can handle entrained liquid, so no scrubbers is needed upstream the screw compressor. Liquid in the compressor will help to keep the temperature down, due to the high heat capacity and utilisation of heat for evaporation.

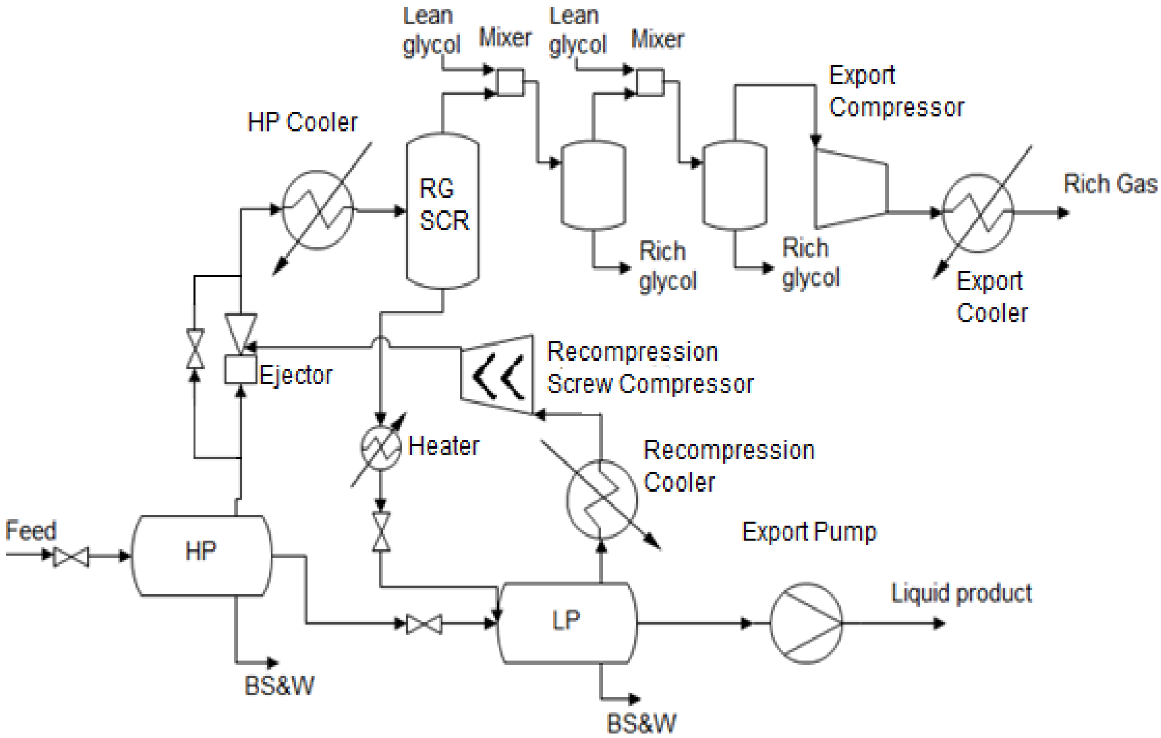



**Figure 6.11 Recompression with screw compressor and ejector**

**Figure 6.12** shows operational performance for the ejector and pressure lift done by the screw compressor. The pressure lift of the screw compressor is limited by pressure ratio and temperature. For lean feed case the screw compressor outlet has a temperature of 111°C with a pressure ratio of 2.8:1. In the rich feed scenario the outlet temperature is 95°C with a pressure ratio of 3.7:1. The high temperature at the compressor outlet decreases the ejector performance. This can be avoided by adding a cooler downstream the screw compressor.

When operating the HP separator at 100bara the low pressure provides a relatively large motive flow rate in the ejector. Mass entrainment ratios are found to be around 0.06 and 0.11 for lean and rich feed respectively. For the rich feed the entire 29.5kg/s gas stream available as motive flow is utilised. In the lean feed scenario operational flexibility is much better, only 1.5kg/s of the total gas flow of 28kg/s is utilised in the ejector.

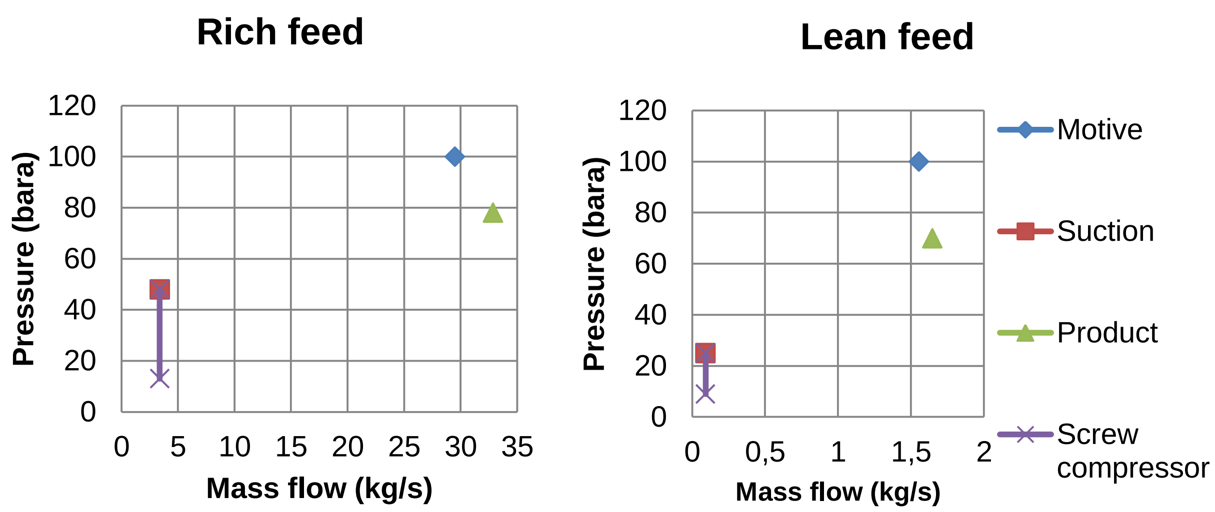



**Figure 6.12 Screw compressor pressure lift, and ejector performance, feed processed to TVP<10bara (PFD see Figure 6.11)**

In **Table 6.5** results from simulations done on the process solution in **Figure 6.11** is presented. Compared to data found for the two previous processes the changes is small, see **Table 6.3** and **Table 6.4** for comparison. The advantage with this solution is increased operational flexibility. By adding a screw compressor in the recompression, there is additional power consumption of 15kW and 529kW for the lean and rich case respectively. For the rich feed scenario the recompression cooling load is on 500kw, in addition load and UA value for the HP cooler is increased with about 5%. Consumed power for the heater used for hydrate prevention, is increased with 250% and 235% for the lean feed and rich feed respectively. This is due to more heavy hydrocarbons is going through the cricondenbar control.

In the lean feed scenario, the HP separator is operated just above the dew-point line, (100bara/75°C) see **Figure 3.3**, meaning that it is all gas and no liquid flow is going to the LP separator. The only flow going to the LP separator is coming from the cricondenbar control (RG SCR in **Figure 6.11**). This gives an operational temperature in the LP separator of 25°C and no need for a cooler upstream the screw compressor. For the HP cooler there is less than 1% increase in load and UA value.

**Table 6.5 Simulation data, (PFD Figure 6.11, Screw compressor and ejector)**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Liquid vapour pressure | TVP<10bara | |
| Total number of units | 15 | |
| Total power consumption (kW) | 5194 | 4564 |
| Total Cooling load (kW) | 8822 | 10 543 |
| Total Cooling area (Note 2;Note 3) ($m^2$) | 198 | 262 |
| Rich gas product ($Sm^3$/d) | 2 995 691 | 2 897 434 |
| Liquid product (Actual $m^3$/d) | 115 | 2765 |
| Rich gas product (tonne/d) | 2344 | 2563 |
| Liquid product (tonne/d) | 87 | 2010 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3$/h) | $P_2/P_1$=200/70=2.9<br>P=5142<br>V=1550 | $P_2/P_1$=200/78=2.6<br>P=3995<br>V=1219 |
| Recompression Screw compressor $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3$/h) | $P_2/P_1$=25/9=2.8<br>P=15<br>V=40 | $P_2/P_1$=48/13=3.7<br>P=529<br>V=766 |
| Recompression cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | No cooler needed, See Note 7. | 68.5 \| 25 \| 38<br>UA=13<br>Q=500 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD (Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 67 \| 25 \| 37<br>UA=95<br>Q=3532 | 81 \| 25 \| 42<br>UA=141<br>Q=5942 |
| Export cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=64<br>Q=5290 | 104.7 \| 60 \| 75<br>UA=55<br>Q=4101 |
| Heating utility (kW) | 21 | 194 |
| Export pump (Note 1) (kW) | 16 | 375 |
| Pressure levels HP \| LP (bara) | 100 \| 9 | 100 \| 13 |

Note 1: Discharge pressure of export pump is set to 100bara
Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.
Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient U=800W/$m^2$K
Note 4: Temperatures is below 130°C in all compressors
Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.
Note 6: Ejector efficiency is set to 20%
Note 7: There is no liquid separated out in HP separator in Lean feed scenario, the only fluid entering LP separator is coming from the cricondenbar control with a temperature of 25°C.

### 6.2.4 T-Separator, ejector and screw compressor

In this section the process solution in **Figure 6.11** is improved to increase operational flexibility. The process is shown in **Figure 6.13** and utilise a screw compressor for the first pressure step, and an ejector for the second pressure step. This process solution is most relevant for the rich case, but it can also operate with lean feed.

The process in **Figure 6.13** uses a high pressure T-Separator to get a high motive pressure for the ejector. When gas from the HP separator was used, see **Figure 6.11**, **Section 6.2.3**, the pressure was limited due to constraint on the HP separator. By adding a separator which can withstand a higher pressure upstream of the HP separator, it is possible to get a high motive pressure to drive the ejector.

One type of high pressure separator is the T-Separator shown in **Figure 5.19**. This is a simple and robust separator where the dimensions can be about the same as the transport pipe. A low diameter gives a low wall thickness compared to a large diameter for the same pressure. This will give an advantage for the T-Separator, as it will be significant less bulky than a conventional vessel separator. The drawback with the T-Separator is that the separation efficiency is limited, and carryover of liquid will be expected. In this analysis a carryover of 5weigth% liquid is used.

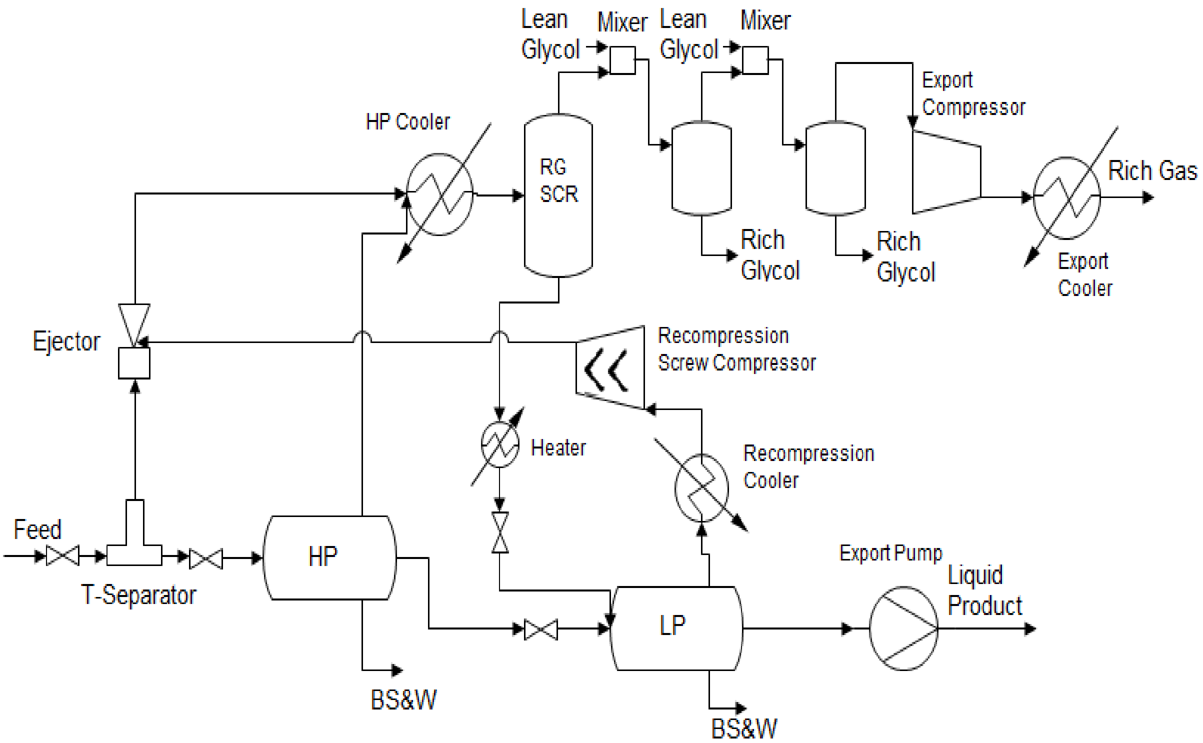



**Figure 6.13 Recompression with T-separator, ejector and screw compressor**

In **Figure 6.14** the screw compressor and ejector performance is presented. For the lean feed case less than 3% of the gas stream from the T-separator is used as motive flow in the ejector, while in the rich case less than 39% motive flow is utilised. Maximum available motive flow rate for the ejectors is 28kg/s for lean feed and in the range 29-31kg/s for the rich feed. This means that there is good flexibility in both scenarios for this configuration.

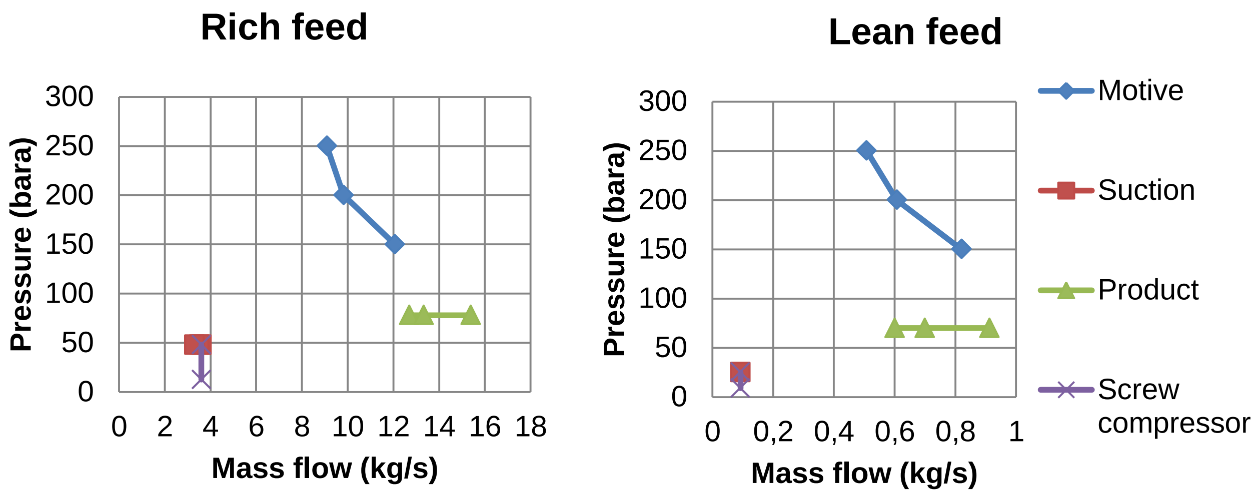



**Figure 6.14 Screw compressor and ejector performance, feed processed to TVP<10bara (PFD see Figure 6.13)**

In **Table 6.6** data from simulations done on the process solution in **Figure 6.13** is presented. For lean feed there are no significant changes in system parameters compared to **Table 6.5** in **Section 6.2.3**. There is still no need for the cooler upstream of the screw compressor since the temperature is 25°C, due to no liquid from the HP separator. Actually the HP separator only separates water and a small amount of gas when the feed is lean, a solution without the HP separator is presented in **Section 6.2.5** and shown in **Figure 6.15**.

With rich feed it is found 1-3% increased liquid production for this system compared to the previous processes solutions in **Section 6.2**. The down side is that when operating the T-Separator at 250bara, the gas phase is heavier than at low pressures. This increases the mass flow going to the cricondenbar control, and increases the HP cooling load with 14-20% and the UA value with 17-24% compared to the previous solutions in **Section 6.2**.

**Table 6.6 Simulation data, (PFD Figure 6.13, T-Separator, ejector, screw compressor)**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Liquid vapour pressure | TVP<10bara | |
| Total number of units | 16 | |
| Total power consumption (kW) | 5190 | 5142 |
| Total Cooling load (kW) | 8704 | 11 280 |
| Total Cooling area (Note 2;Note 3) ($m^2$) | 196 | 289 |
| Rich gas product ($Sm^3$/d) | 2 995 925 | 2 870 421 |
| Liquid product (Actual $m^3$/d) | 115 | 2854 |
| Rich gas product (tonne/d) | 2344 | 2499 |
| Liquid product (tonne/d) | 87 | 2074 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3$/h) | $P_2/P_1$=200/70=2.9<br>P=5138<br>V=1549 | $P_2/P_1$=200/78=2.6<br>P=4002<br>V=1222 |
| Recompression Screw compressor $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3$/h) | $P_2/P_1$=25/9=2.8<br>P=16<br>V=40 | $P_2/P_1$=12.5/48<br>P=573<br>V=832 |
| Recompression cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | No cooler needed, See Note 7. | 61 \| 25 \| 35<br>UA=12<br>Q=408 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD (Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 66 \| 25 \| 37<br>UA=93.5<br>Q=3432 | 78.5 \| 25 \| 41<br>UA=165<br>Q=6799 |
| Export cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=63.5<br>Q=5272 | 106 \| 60 \| 76<br>UA=54<br>Q=4073 |
| Heating utility (kW) | 20 | 177 |
| Export pump (Note 1) (kW) | 16 | 390 |
| Pressure levels HP \| LP (bara) | 250 \| 70 \| 9 | 250 \| 78 \| 13 |

Note 1: Discharge pressure of export pump is set to 100bara
Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.
Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient $U=800W/m^2K$
Note 4: Temperatures is below 130°C in all compressors
Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.
Note 6: Ejector efficiency is set to 20%
Note 7: There is no liquid separated out in HP separator in Lean feed scenario, the only fluid entering LP separator is coming from the cricondenbar control with a temperature of 25°C.

### *6.2.5 T-Separator and two-stage ejector*

In this section a special design based on lean feed is presented, see **Figure 6.15**. Rich feed was tested for this configuration, but the available motive flow was found incapable to drive both ejectors. There is also significant more liquid that need to be stabilised in the rich case, giving need for the HP separator to flash light hydrocarbons at high pressure.

For the solutions in **Section 6.2.3** and **Section 6.2.4**, the inlet separator is operated above the dew-point line for lean feed, see **Figure 3.3**. Therefore it is no significant separation done in the HP separator. In addition only 1/200 of the available motive flow is utilised in the ejector**,** see **Figure 6.13**, **Section 6.2.4**. This gives potential for use of a two-step ejector solution. So in the process presented in **Figure 6.15** the HP separator is removed, and there are used two ejectors for recompression of flash gas.

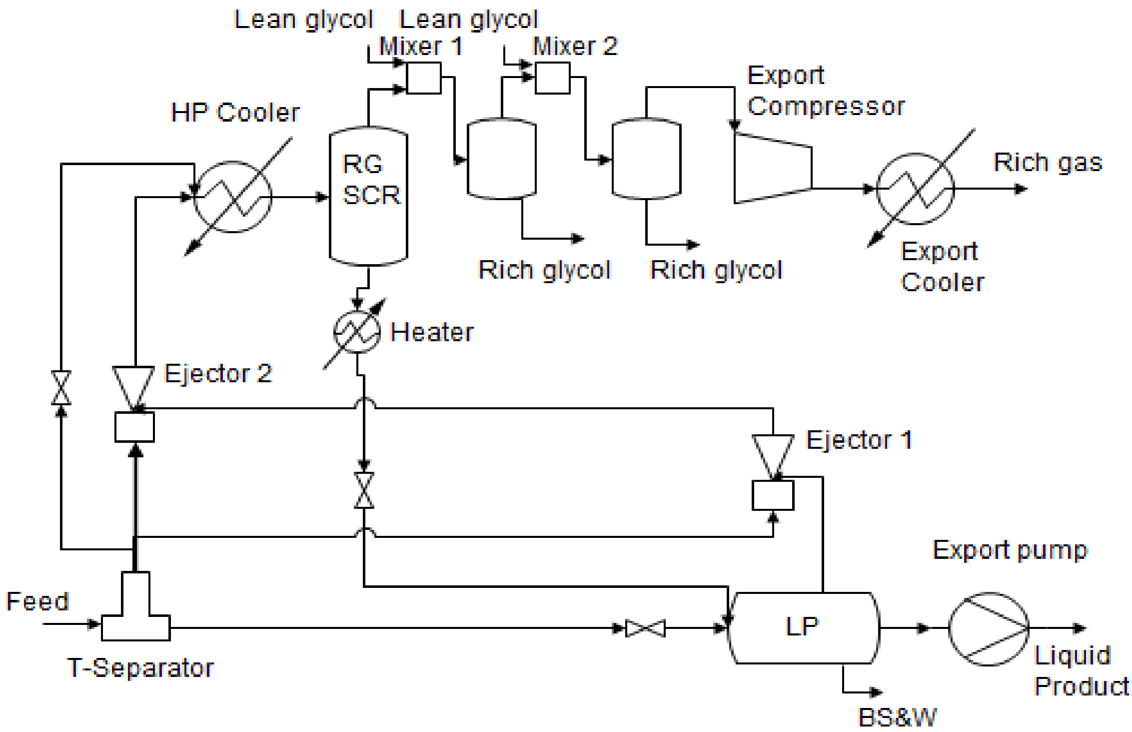



**Figure 6.15 Recompression with T-separator and two-step ejector (based on lean feed)**

Ejector performance is presented in **Figure 6.16**. It is found that suction flow rate in Ejector 1 is relatively low compared to the available motive flow rate. Due to the low suction flow rate and a pressure lift from 9 to 25 bara, the motive mass flow rate is barely increasing for decreasing inlet pressure. The low flow rates in Ejector 1 increases operational flexibility of Ejector 2, as Ejector 2 needs to pressurise the total product flow from Ejector 1. It is in total about 28kg/s separated out as gas in the T-Separator available as motive flow. For an inlet pressure of 150 bar only around 4kg/s or 14% is used as motive flow for the ejectors. This means that this ejector configuration has a large enough flow rate to operate with lower motive pressure than 150 bara. Even for 100bara the motive flow rate used in both ejectors is found to be just around 9kg/s or 32% of the available motive flow utilised.

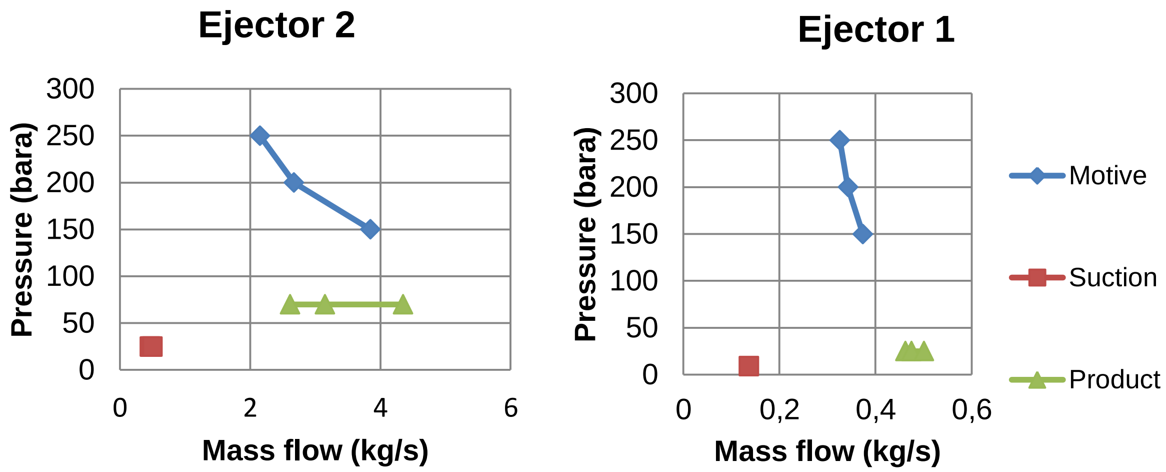



**Figure 6.16 Ejector performance, lean feed processed to TVP<10bara (PFD see Figure 6.15**

When temperature and pressure in the T-Separator is held above lean feed dew point line, only free water is separated out in the T-Separator. Mixing this water into the LP separator gives around 2vol% water in the liquid hydrocarbon product, which is on the limit for pipe transport and above for ship transport. One solution could be to take separated water directly to the water treatment system. But if the T-Separator operates within the phase envelope of the lean feed, liquid hydrocarbons will be mixed with the water. If this occurs the phases should be separated in the LP separator to avoid high content of hydrocarbon liquid entering the water treatment system. The water issue may require special internal equipment for water removal in the LP separator. For example will coalescing internals make it easier to collect water droplets, or a heater could be used upstream of the separator to lower viscosity and ease separation.

In **Table 6.7** results from simulations on the process solution in **Figure 6.15** is presented. This solution removes the need of a screw compressor with the upstream cooler in recompression, saving electrical power and cooling load. In addition the bulky HP separator is removed, which should give a significant CAPEX reduction.

**Table 6.7 Simulation data (PFD Figure 6.15, T-Separator and two ejectors)**

| **Designation and unit** | **Lean feed** |
|---|---|
| Liquid vapour pressure | TVP<10bara |
| Total number of units | 14 |
| Total power consumption (kW) | 5174 |
| Total Cooling load (kW) | 8717 |
| Total Cooling area (Note 2;Note 3) ($m^2$) | 197 |
| Rich gas product ($Sm^3/d$) | 2 997 346 |
| Liquid product (Actual $m^3/d$) | 116 |
| Rich gas product (tonne/d) | 2347 |
| Liquid product (tonne/d) | 85 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3/h$) | $P_2/P_1$=200/70=2.9<br>P=5138<br>V=1549 |
| Recompression power (kW) | 0 |
| Recompression cooling (kW) | 0 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 66 \| 25 \| 37<br>UA=94<br>Q=3443 |
| Export cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=83<br>Q=5274 |
| Heating utility (kW) | 20 |
| Export pump (Note 2) (kW) | 16 |
| Pressure levels HP \| MP \| LP (bara) | 250 \| 9 |
| Note 1: Discharge pressure of export pump is set to 100bara<br>Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.<br>Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient U=800W/$m^2$K<br>Note 4: Temperatures is below 130°C in all compressors<br>Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications.<br>Note 6: Ejector efficiency is set to 20% | |

### 6.2.6 Dual screw compressors

In the previous solutions presented in **Section 6.2**, operational flexibility for ejectors has been limited, especially for the rich feed scenarios. If compressors are used for recompression this would improve flexibility, and make the recompression system able to operate more freely from the rest of the system. In some of the previous process solutions there is already included one compressor for recompression. It can be expected that using the same type of equipment in series instead of two different types, can lower development costs, maintenance cost, and in general reduce CAPEX and OPEX. One of the drawbacks using compressors compared to ejectors, is that a cooler will most likely be needed upstream of each compressor to be within the compressors temperature limit.

Based on flow conditions it is possible to use screw compressors for recompression. The screw compressor will be able to operate with the entrained liquid, and low flow rates that is found in the recompression system. In **Figure 6.17** a solution with two screw compressors in series is presented.

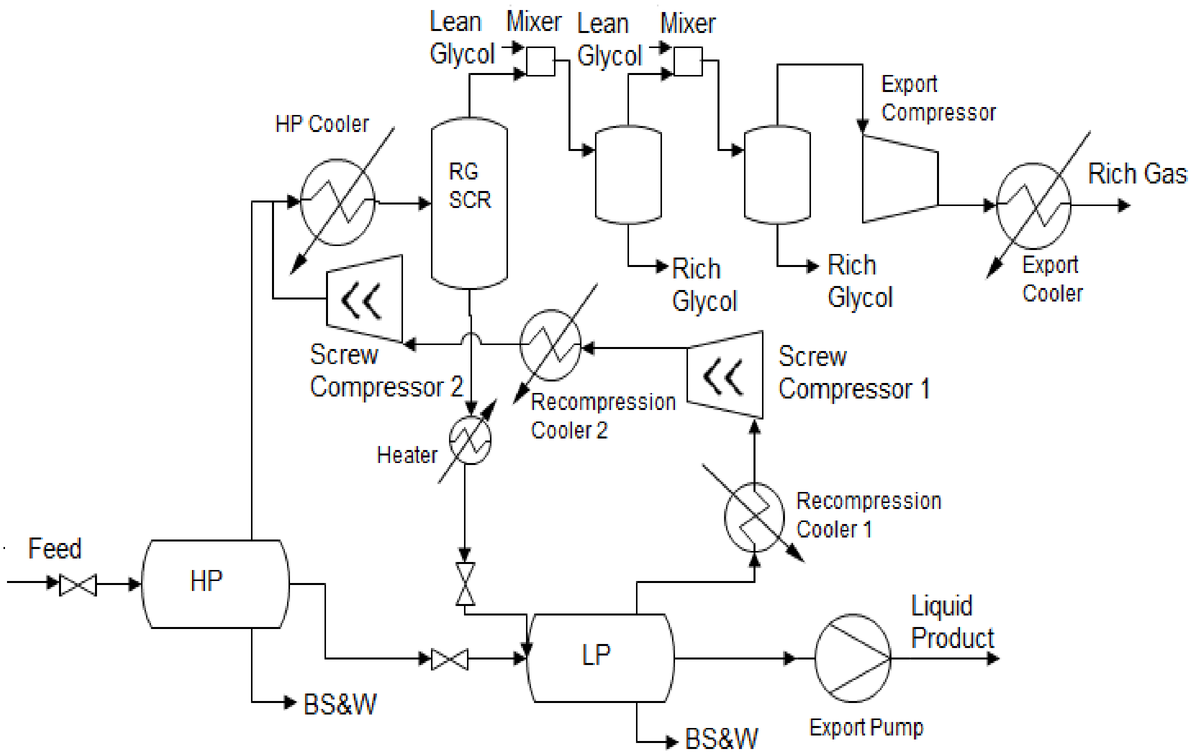



**Figure 6.17 Recompression with dual screw compressors**

In **Table 6.8** data from simulations done on the process solution shown in **Figure 6.17** is presented. For the lean case the screw compressors and recompression coolers are found to be relatively small. For Screw compressor 2 the suction volume flow is just $12m^3/h$, this is in the low end of conventional screw compressors, see **Figure 5.1**.

For the rich feed scenario the power consumption is actually less than for the two other solutions presented with screw compressors in **Section 6.2.3** and **6.2.4**, see **Table 6.5** and **Table 6.6**. This is mainly due to a lower pressure ratio for screw compressor 1 with use of an intercooler at the lower outlet pressure (Recompression cooler 2). When the gas is cooled the volume decreases, and therefore less work is needed for compression. Compared to the screw compressors solutions in **Section 6.2.3** and **6.2.4,** the HP cooler UA value is reduced with 4-18% (from 141-165kW/K to 135kW/K), this is due to the relatively small intercooler with UA value of 11kW/K.

**Table 6.8 Simulation data (PFD Figure 6.17 Dual screw compressor)**

| Designation and unit | Lean feed | Rich feed |
|---|---|---|
| Liquid vapour pressure | TVP<10bara | |
| Total number of units | 16 | |
| Total power consumption (kW) | 5196 | 5002 |
| Total Cooling load (kW) | 8835 | 10 492 |
| Total Cooling area (Note 2;Note 3) ($m^2$) | 198 | 263 |
| Rich gas product ($Sm^3/d$) | 2 996 913 | 2 897 660 |
| Liquid product (Actual $m^3/d$) | 111 | 2763 |
| Rich gas product (tonne/d) | 2347 | 2565 |
| Liquid product (tonne/d) | 84 | 2009 |
| Export Compression $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3/h$) | $P_2/P_1$=200/70=2.9<br>P=5142<br>V=1550 | $P_2/P_1$=200/78=2.6<br>P=3994<br>V=1219 |
| Screw compressor 1 $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3/h$) | $P_2/P_1$=25/9=2.8<br>P=13<br>V=35 | $P_2/P_1$=32/13=2.5<br>P=264<br>V=572 |
| Screw compressor 2 $P_2/P_1$<br>Power - P (kW)<br>Suction volume flow rate - V ($m^3/h$) | $P_2/P_1$=70/25=2.8<br>P=12<br>V=12 | $P_2/P_1$=78/32=2.4<br>P=218<br>V=199 |
| Recompression cooler 1 $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 35 \| 25 \| 25<br>UA=0.1<br>Q=2.2 | 68 \| 25 \| 37.5<br>UA=10<br>Q=369 |
| Recompression cooler 2 $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 106 \| 25 \| 50<br>UA=0.3<br>Q=16 | 70 \| 25 \| 38<br>UA=11<br>Q=428 |
| HP Cooler $T_1$ \| $T_2$ \| LMTD (Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 68 \| 25 \| 37<br>UA=95<br>Q=3526 | 80 \| 25 \| 41.5<br>UA=135<br>Q=5594 |
| Export cooler $T_1$ \| $T_2$ \| LMTD(Note 2) (°C)<br>UA (kW/K)<br>Load - Q (kW) | 124 \| 60 \| 83<br>UA=64<br>Q=5291 | 105 \| 60 \| 75<br>UA=55<br>Q=4101 |
| Heating utility (kW) | 13 | 151 |
| Export pump (Note 1) (kW) | 16 | 375 |
| Pressure levels HP \| LP (bara) | 70 \| 9 | 78 \| 13 |
| Note 1: Discharge pressure of export pump is set to 100bara<br>Note 2: Sea water temperature is assumed 5°C constant through the whole cooling process.<br>Note 3: All coolers is assumed active coolers with an overall heat transfer coefficient U=800W/$m^2$K<br>Note 4: Temperatures is below 130°C in all compressors<br>Note 5: See **Chapter 3** Framework for general assumptions, feed properties, feed composition, and product specifications. | | |

# 7. Assessment and analysis of gas dehydration

In this chapter, systems for gas dehydration is discussed and analysed. The analysis is done with the basis given in **Section 3.3.4.**

Since the gas is saturated with water when it comes from the reservoir, it needs to be dehydrated to given transport specifications, see **Table 3.6** for Rich gas specifications. Without gas dehydration water knockout will occur in the transportation pipeline, due to decreasing temperature. Low temperature, free water and natural gas under high pressure will create an environment for hydrate formation, as shown in **Figure 3.4** and **Figure 4.5**. During screening of process solutions it was discussed dehydration by absorption, adsorption, cooling/expansion, and selective membrane technology.

Absorption using glycol is often used in upstream processing, and is able to reach Rich gas dew point specification. The conventional glycol absorber is a counter-current tower, with packing or equilibrium stages, which provides good contact between gas and glycol. Water rich glycol is regenerated in a stripping column, where the water is boiled out of the glycol. Subsea a conventional system is seen to be too complex, bulky, and power consuming. One large simplification will be to use a host for glycol regeneration, either topside or at shore. This would eliminate power consumption for gas dehydration subsea, and remove most of the complexity. The remaining part on the seafloor is then the absorption system, which can be simplified to use of compact co-current contactors and separators, see **Figure 4.8**. As discussed in **Section 3.3.4** MEG is most likely available subsea for hydrate inhibition. To use the same glycol for dehydration and hydrate inhibition, will reduce facilities and infrastructure. The simplified dehydration process using a host for regeneration, leaving only co-current contactors and separators subsea, is selected for further evaluation in this study.

Adsorption needs at least two large columns filled with adsorbents, one in operation and one in regeneration. These type of processes is able to reach very low water dew points, and can in practise remove almost all the water from the gas. After some time of operation the adsorption column get saturated with water and need to be regenerated. Desorption can be done by adding heat and/or reducing the pressure, water is picked up by a stripping gas, condensed in a cooler and separated from the stripping gas. Subsea the heat will most likely come from an electrical heater with direct impact on the plants power consumption. There are also concerns with fouling gas, liquid and other impurities, which make adsorption less attractive for subsea implementation. Based on this discussion and assessment of the technology described in **Section 4.3.2**, adsorption is not further evaluated in this study.

Cooling and/or expansion can be used for gas dehydration. This is often relatively simple and robust systems. Due to low temperatures hydrate inhibitor will be needed. Since glycol is needed in this process there will most likely be more attractive to use a glycol absorption process. To reach the Rich gas water dew point specification (-18°C at 70bara) it is expected that a low temperature and pressure will be needed for such a system. In the solutions presented in **Chapter 6**, cooling and/or expansion is used for cricondenbar control, which also reduces the water content to some degree. Based on this discussion and assessment of information given in **Section 4.3.3**, in special increased power consumption for the export compressor, this type of systems is not further evaluated for subsea dehydration in this study.

Membrane technology is seen to be in a relatively early stage for gas processing, and issues with entrained liquid, fouling gas and other impurities must be solved before subsea implementation. Some of the benefits with membrane technology is that there is no moving parts, continues operation, and low power consumption. Membrane technology can be an option in the future, but a technological qualification program will be needed.

## 7.1 Dual lean glycol mixer system

In this section a glycol absorption system is presented, see **Figure 7.1**. The solution is based on the co-current contactor system presented in **Figure 4.8**, but the pump is removed to avoid rotating equipment. Two mixers are applied, where lean glycol is added in both mixers. After each mixer gas and water rich glycol is separated. Lean glycol is imported from a host, and rich glycol is sent back to the host for regeneration.

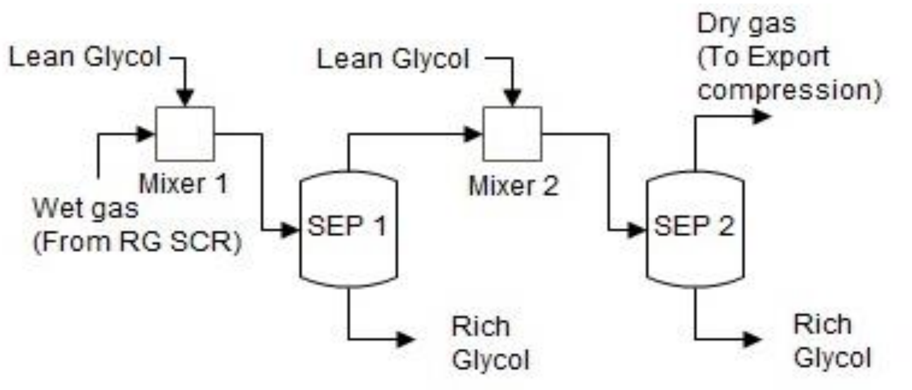


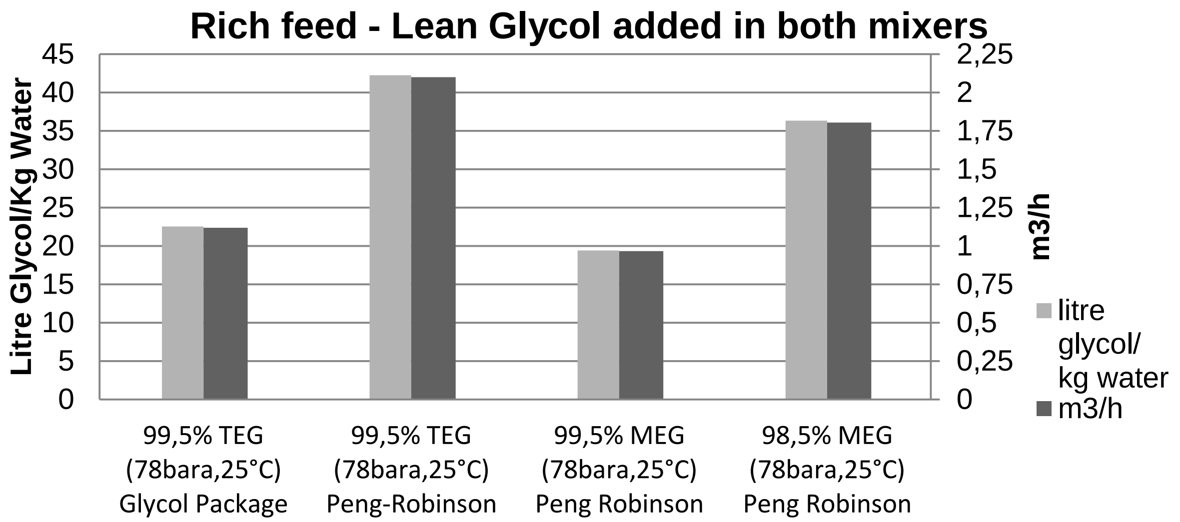

**Figure 7.1 Dual lean glycol mixer system**

In **Figure 7.2** circulation rates for the system in **Figure 7.1** is presented. There is an increase in glycol circulation rate from the rich to the lean feed scenario. The reason for this is mainly due to the pressure difference, 70bara and 78 bara for the lean and rich case respectively. There is also about 8% more water removed in the lean case, due to larger flow rate of gas and higher water content than for the rich feed scenario, see **Table 3.7**.

In **Figure 7.2** it is clear that a good prediction of glycol circulation rate is highly dependent on a good thermodynamic model. When using Peng Robinson it seems very conservative in the rich case, but in the lean case it is the HYSYS Glycol package that gives the largest circulation rate. The Glycol package predicts about 170% increased circulation rate from the rich to the lean case. This seems very high, since there is only about 8% more water removed and the pressure decrease is only 10%. For simulations done with both compositions at the same pressure, the difference is significantly reduced. For Peng Robinson the increased glycol circulation rate from lean to rich case is only about 15%, when using 99,5wt% TEG. Which seems realistic regarding the pressure decrease and the increase in removed water. Based on this results it is not possible to say which of the models that provides the most accurate result for TEG circulation rate, so both are presented in **Figure 7.2**.

Decreasing MEG purity from 99,5wt% to 98,5wt% increases MEG circulation rate with 80-130%. The circulation rate is still within an acceptable range, so reducing the MEG purity should be feasible.

Another issue with MEG is that it has a higher solubility in gas than TEG, so reaching the specification of 8liter MEG/MSm3 for the export gas difficult. In simulations the MEG content in the export gas is found to be in the range 10-20 liter MEG/MSm3, without carryover in the separator.



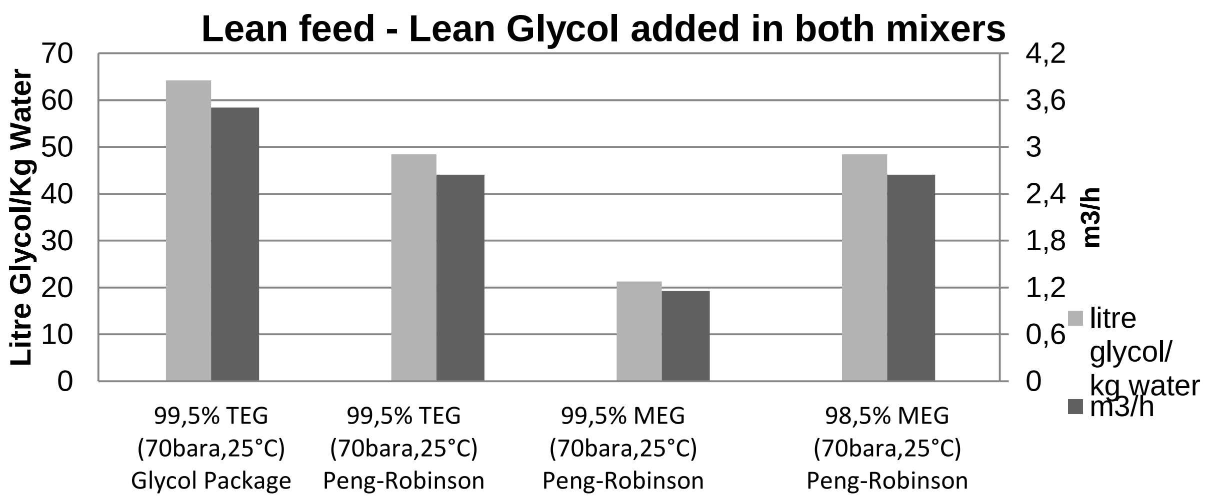




**Figure 7.2 Glycol circulation rate using dual mixer system shown in Figure 7.1**

## 7.2 Two-stage glycol system

The system presented in this section is identical to the two-step co-current system in **Figure 4.8**. This is two-stage system, where lean glycol is added in the second mixer and then separated out as a semi-lean solvent which is pumped to the first mixer, see **Figure 7.3**. This is a system with multiple equilibrium stages, see **Figure 4.7**

According to Henry's law, equation (2), the partial pressure of water in the gas phase is proportional to the molar fraction of water in the liquid phase. This principle is utilised for the system in **Figure 7.3**. In Mixer 2 the lean glycol has a low concentration of water and is therefore able to absorb water from the gas phase. At the exit of Mixer 2 the glycol has gone towards equilibrium with the dry gas. Since the wet gas has a higher partial pressure of water than the dry gas, this semi-lean glycol can also be used in Mixer 1 to absorb more water.

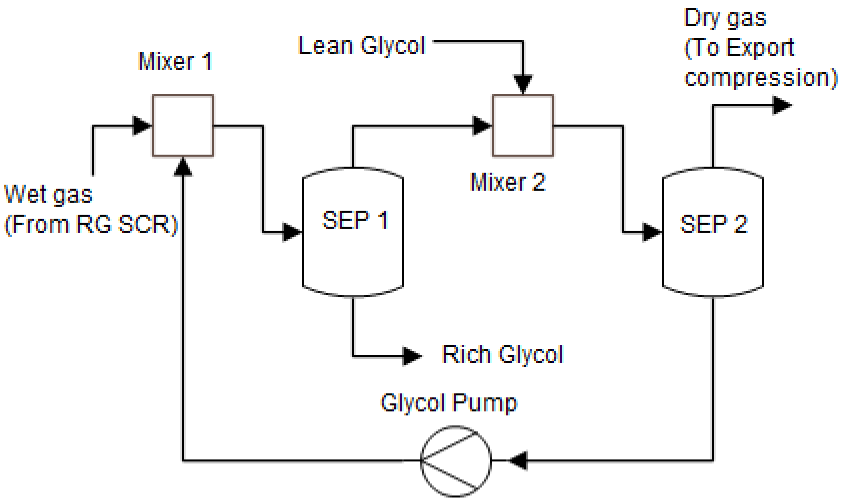
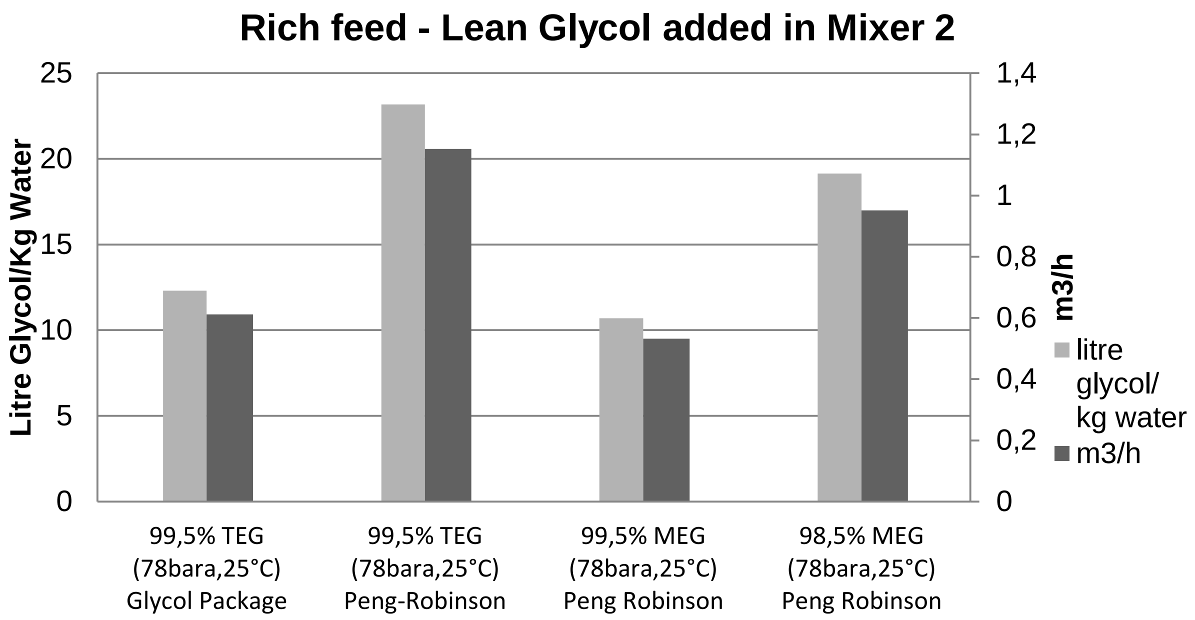



**Figure 7.3 Two-step glycol system**

In **Figure 7.4** glycol circulation rates for the system in **Figure 7.3** is shown. The variation in circulation rate between the different scenarios follows the same pattern as was seen in **Figure 7.2** and discussed in **Section 7.1**. Comparing circulation rates for the systems with two mixers, it is found a reduction of about 50% in glycol circulation rate when adding lean glycol only to the second ejector, see **Figure 8.5** for comparison of the systems. The reason for this behaviour is that the glycol from Mixer 2 is able to pick up much more water when mixed with the wet gas according to Henry's law. Due to the high partial pressure of water in the wet gas, a new equilibrium can be approached, transferring water from the wet gas to the semi-lean glycol.



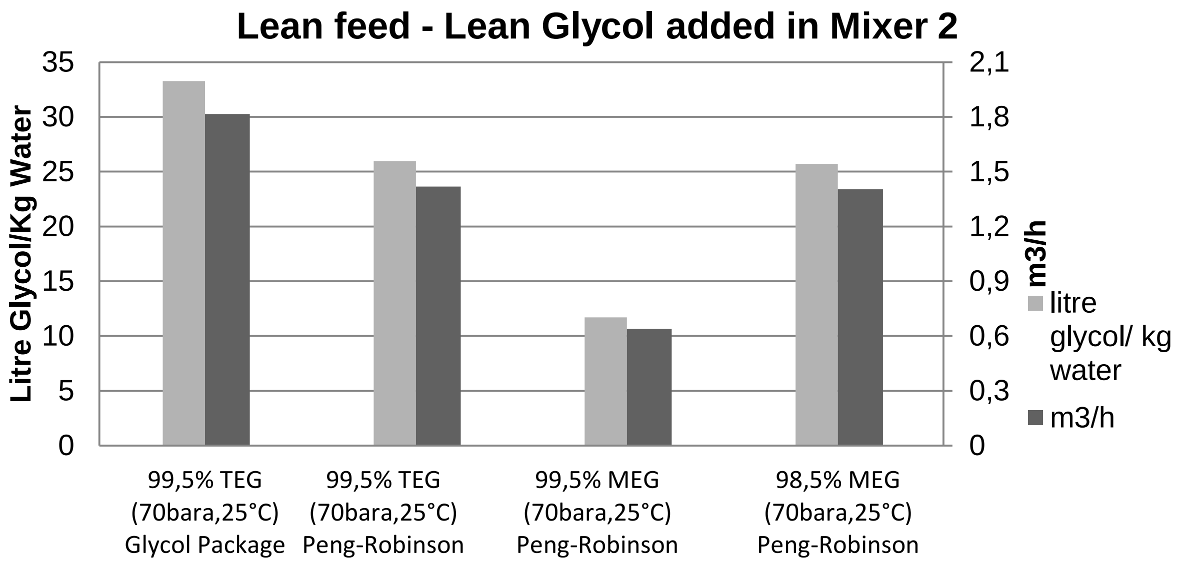




**Figure 7.4 Glycol circulation rates for the Two-stage system shown in Figure 7.3**

## 7.3 Single glycol mixer

Another solution is to use a single co-current contactor, as shown in **Figure 7.5**. This approach would require large circulation rates of glycol as shown in **Figure 7.6** compared to the two other solutions presented, see **Figure 8.5**. The reason for this is that it is limited how much water the glycol can pick up in one equilibrium stage.

As discussed in **Section 3.3.4** absorption is favoured by high pressure and low temperature. This is stated in Equation (1) given in the same section. The pressure can be increased by adding a compressor upstream of the system. This will also require a cooler to reduce the temperature downstream of the compressor. The reduction in glycol circulation rate is found to be up to 50% for the rich case when increasing the pressure from 78bara to 120 bara, and almost 70% for the lean case increasing the pressure from 70 to 120bara, see **Figure 7.6**.

Operating the single stage mixer system seems feasible for the high pressure TEG case, having a circulation rate just above 40liter TEG/kg $H_2O$ or 2m$^3$/h. When using TEG the HYSYS Glycol package is used for simulations, and operating at 120bara, there is only 15% increase in circulation rate from the rich to the lean feed scenario. This seems more realistic than what is seen for the other cases with 70bara and 78 bara, for lean and rich feed respectively.

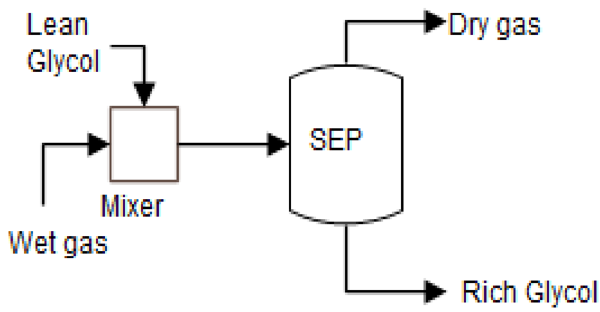
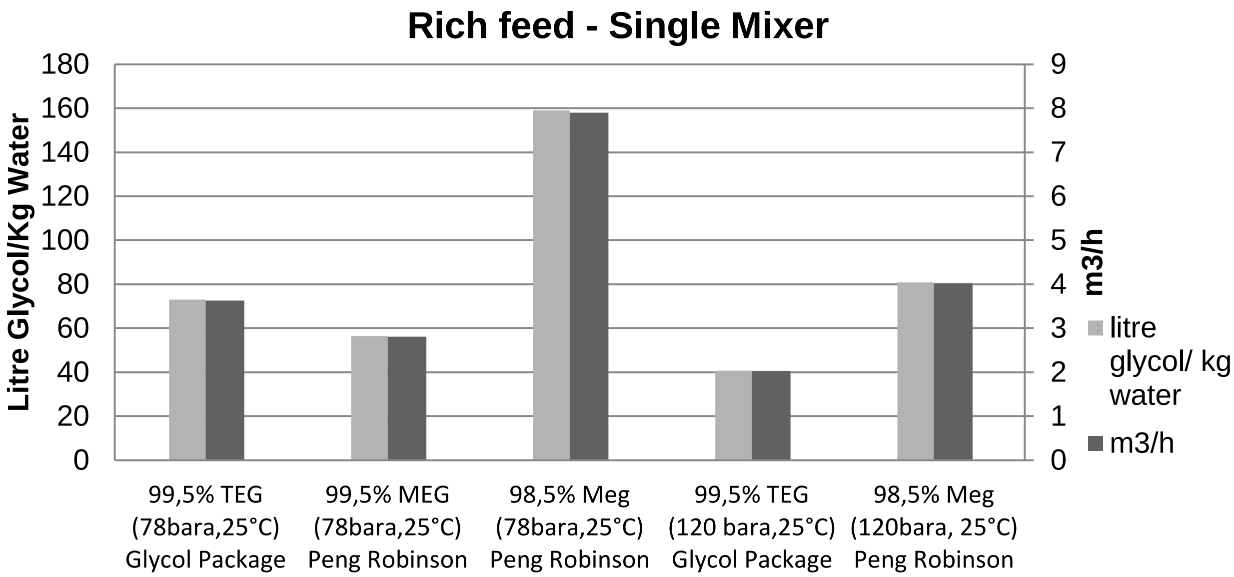



**Figure 7.5 Single glycol mixer**



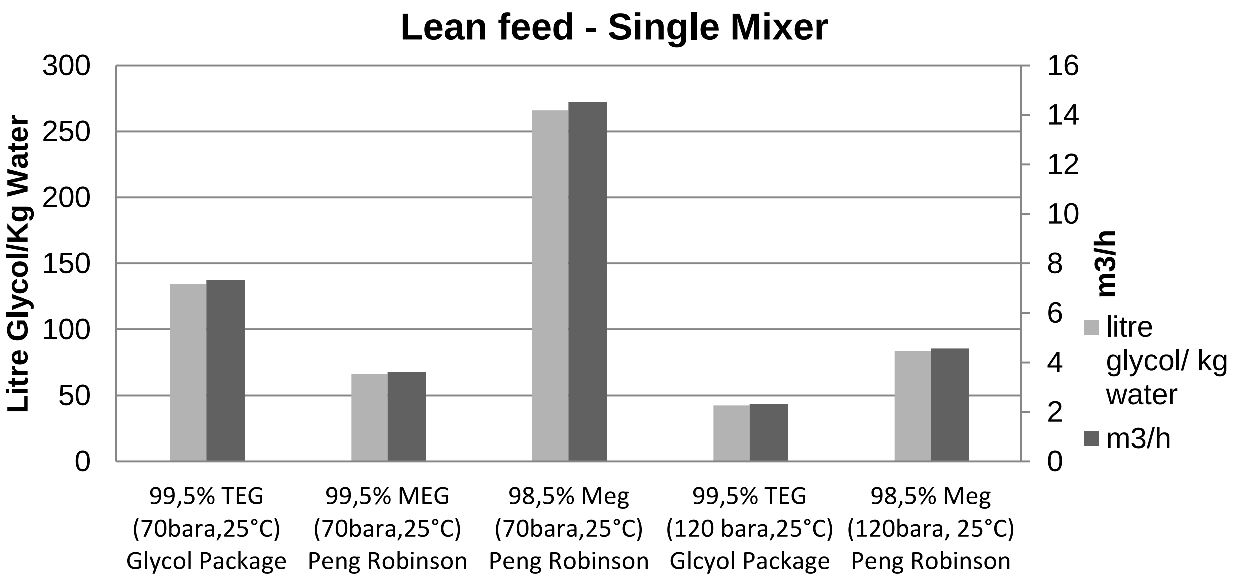




**Figure 7.6 Glycol circulation rate single-single stage mixer shown in Figure 7.5**

# 8. Overall discussion and evaluation

This chapter connects the dots throughout this master thesis with an overall discussion and evaluation, this includes, but is not limited to, evaluation of process solutions, impact of liquid vapour pressure, and subsea heat and power production.

The design philosophy for subsea processing, given in **Section 3.2**, provides additional challenges for a subsea system compared to a conventional topside installation. In a subsea environment there maintenance are difficult and expensive, and possibilities for visual control of the plant is limited to use of cameras. Factors like high reliability, low maintenance, simplicity, robustness, and use of mature/available equipment to avoid expensive qualification programs is some of the most important factors in a subsea development.

In addition to requirements in the subsea design philosophy, the oil and gas industry is known for its conservatism. So converting topside solutions and using known technology to a large extent, provides security and increases the likelihood for subsea implementation of the system.

## 8.1 Impact of liquid product vapour pressure

The process analysis shows that the liquid product vapour pressure spesification has significant impact on process parameter and design. To lower the liquid product vapour pressure, the pressure can be reduced and/or heat can be added.

In a subsea process with high external pressure, low internal pressure will create challenges in equipment design and risk of sea water leaking into the system. Leakages of sea water into the system can ruin process components and the liquid product, and in worst case block pipes and equipment with hydrates and ruin downstream processes.

Reducing the product vapour pressure from TVP<10bara to TVP<5bara, gives an increased power consumption of 26% for the rich case, and less than 1% for the lean case, see **Figure 8.1**. In **Figure 8.1** the dual screw compressor system with a heater upstream LP separator, shown in **Figure 6.1**, **Section 6.1**, is used for all scenarios except for complete stabilisation of rich feed.

The two-stage systems are not capable of complete stabilisation of rich feed. Use of a more complex three-stage system, as the one shown in **Figure 6.2**, must be applied. This will add complexity and has 5-7 units more than the other systems presented in this study. Even with a heat input of 3.7MW the pressure in the LP separator is as low as 2 bara to reach complete stabilisation of the liquid product.

For rich feed the cooling area increases with increased heat input from TVP<10bara to TVP<5bara, and due to additional coolers in the system from TVP<5bara to TVP<1bara.

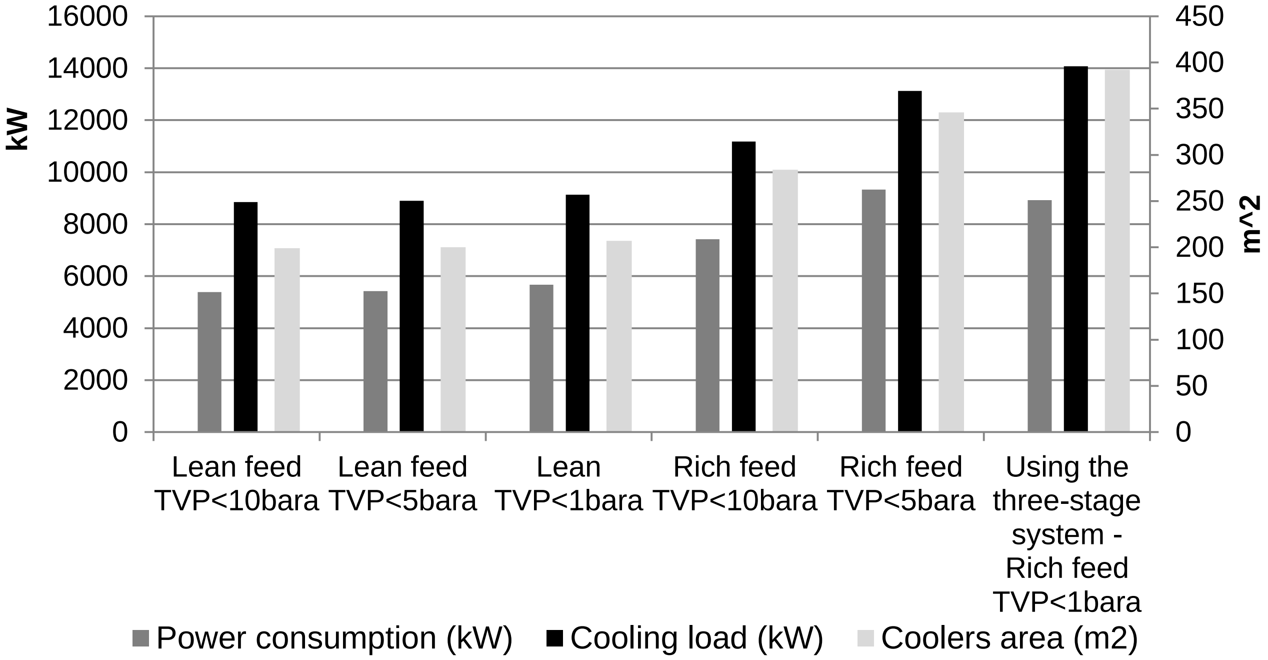



**Figure 8.1 Impact of liquid product vapour pressure on power, cooling load and area. (PFD see Figure 6.1 and Figure 6.2)**

## 8.2 Discussion and evaluation of process solutions

For the overall process design the emphasis is put on the recompression system. The recompression system compresses flash gas from the liquid stabilisation process, to mix it back with the associated gas stream. This is found to have large impact on system complexity, and is the most severe technological gap for realisation of subsea stabilisation of liquid products.

For the recompression system use of ejectors provides a robust solutions. They have no moving parts, which increases reliability and lowers maintenance. The drawback with these solutions is operational flexibility. They are dependent on having a high pressure flow, with a large enough flow rate to pressurise the low pressure stream. Since there is no moving parts, the ejectors will have a maximum mass flow rate (choked flow) when the velocity at the throat gives Mach equal to one. For operational control of the ejector a configuration with multiple ejectors in parallel can be used. There is also a possibility to use an ejector with higher outlet pressure than needed, and then control the pressure in a downstream valve, but this will increase requirements of the motive stream, and reduce flexibility.

As shown in process designs and analysis in **Section 6.2** the inlet flow is the main flow used as motive flow for ejectors. The reservoir pressure will be reduced with time of production, and after some time the mass flow rate will also be reduced, which in the end reduces flexibility. In a worst case scenario the pressure and/or mass flow gets too low to drive the ejector. This can be solved by using another motive stream or adding compressors.

The motive stream may be taken from one of the export streams, but this will increase power consumption. How large this additional power consumption will be depends on the recompression flow rate and the motive flow. Gas can be taken from the rich gas export stream and used directly as motive stream in an ejector. This solution will have very low efficiency, and increase throughput of gas in the export compressor. Another option is to take a part of the liquid product and pressurise this to high pressure. Using liquid as motive flow for the ejector will provide better efficiency compared to a gas ejector, but a downstream separator will be needed to take out liquid. Instead of adding a pump there may be better to add a compressor directly at the recompression flow.

### 8.2.1 Evaluation of process solutions

For the solution with ejectors upstream and downstream of the HP separator, see **Section 6.2.1**, the flexibility is found lower than for the other solutions. Due to the ejector downstream of the HP separator, this solution needs an additional separator stage to make two pressure stages for the ejectors. The benefit of this additional separator is that the liquid production rate is increased with about 1%, compared to the two stage solutions.

To increase flexibility the ejectors in **Section 6.2.2** is connected to the high pressure well stream. The inlet feed is a multiphase flow, so the MP separator is used to remove liquid after the first ejector step. Use of a MP separator decreases the load for both ejectors. The concern with this type of system is that there will be sand and dirt coming from the wells, which may block the ejectors. There can also be problems with different flow patterns, such as slug flow that will affect the ejector performance.

To remove problems with multiphase flow as motive stream for ejectors, the solution in **Section 6.2.3** can be used. In this process the ejector is placed on the gas outlet of the HP separator. Since the HP separator is a bulky three phase separator, there is set a limit at 100bar pressure. This relatively low pressure reduces the operational flexibility. For the rich feed case a two-step ejector solution with this low pressure is not feasible. To solve this there is added a compressor to the system. There could also be used a ejector that is driven by gas from the export stream. But this would give significant increase in power consumption, due to low ejector efficacy and increased throughput in export compressor. Adding a screw compressor that can be used for a part of the pressure increase seems like a better solution, providing a much lower increase in power consumption.

The solution in **Section 6.2.4** utilise the high pressure gas from the feed stream as motive stream, and a screw compressor for the first pressure step. This provides a solution with high flexibility, controllability, and robustness. This requires that a high pressure separator is added upstream of the HP separator. In this study a simple T-Separator is representing the high pressure separator, but this can also be some other type of high pressure separator, for example a pipe separator, see **Section 5.2.3**. For the rich feed scenario this is probably the best solution presented in this paper that includes ejectors, due to the controllability, flexibility and robustness. The drawback in the rich feed scenario is that this system has the largest cooling load of all systems analysed, see **Figure 8.2**. This arrangement could also be turned around, using the ejector first and then a screw compressor. This will reduce the high pressure requirement for the motive stream, but increase the flow through the screw compressor and consumption of power.

In lean feed scenarios operating the inlet separator at pressures above 100bar, no liquid is separated out in the inlet separator. This leads up to a significant simplification for scenarios with lean feed. The simplified system is presented and analysed in **Section 6.2.5**. Here the bulky three phase HP separator is replaced by a high pressure separator, represented by the T-Separator. This system provides a large flexibility for lean feed. Reduction in reservoir pressure will most likely not be a big issue for this system operating with lean feed. The drawback with this solution is that is it not feasible for a rich feed scenario, due to a much larger recompression gas flow rate. Another issue is that all the produced water is removed in one separator. This will make requirement of equipment to make sure that the water content in the liquid product stream is within given transport specifications.

In the rich feed case there will be need for at least one screw compressor for recompression to provide flexibility. So there is no concerns about adding another screw compressor to this system, see **Section 6.2.6**. From the analysis it is clear that adding this additional screw compressor has very low impact on power consumption and cooling load, compared to the other solutions with screw compressors. This system is also able to handle fluids with both lean and rich feed compositions, and provides high flexibility for both cases.

### 8.2.2 Comparison of process solutions

In **Table 8.1** a comparison of the process solution from **Section 6.2** is done. The main units and equipment that will need severe theological qualification are counted. There is also stated operational flexibility for the different solutions based on the analysis in **Section 6.2.**

The count for equipment that needs technological qualification is based on the following. Only equipment that needs a significant qualification program is counted. Oil-free screw compressors are not yet available subsea, and would need a significant technological qualification. The twin-screw multiphase pump is operated subsea, but the flow rate in recompression is found too low for this machine. Ejectors is used subsea, but not in this type of system where the pressure need to be controlled more precisely. So a significant qualification program for ejector system with improved operational controllability will most likely be needed. For example parallel ejectors can be used. The applied heater is a low temperature heater for hydrate prevention. This type of heaters is available subsea and is not taken into the counting. Coolers, separators, centrifugal pumps and compressors are available subsea. Some adaption will be needed, but this should be minor details.

The total number of units is found to be more or less equal for all solutions, the difference lies in type of equipment. For each screw compressor added there is need for an upstream cooler. This gives the dual screw compressor solution a total of 4 coolers, which is the largest number of coolers. The benefit is no use of ejectors, and a high operational flexibility for both lean and rich feed.

For lean feed, the solution with T-Separator and two ejectors seems like a very good option. It has the lowest number of equipment, and rotating equipment is used for export only. In addition it has high flexibility for lean feed. This is also the most compact system with the high pressure inlet separator, and no bulky HP separator. The drawback is that this is not applicable for rich feed, and it is seen a higher water content in the liquid product than for the other solutions, due to use of only on three phase separator.

**Table 8.1 Comparison of processes**

| | **Ejector up- and down-stream of HP Sep.** | **Two ejectors driven by the well stream** | **Screw and ejector** | **T-Sep. ejector, and screw** | **Dual screw** | **T-Sep. and two ejectors** |
|---|---|---|---|---|---|---|
| **Section** | **6.2.1** | **6.2.2** | **6.2.3** | **6.2.4** | **6.2.6** | **6.2.5** |
| **PFD** | **Figure 6.3** | **Figure 6.8** | **Figure 6.11** | **Figure 6.13** | **Figure 6.17** | **Figure 6.15** |
| Number of units | 15 | 15 | 15 | 16 | 16 | 14 |
| Coolers | 2 | 2 | 3 | 3 | 4 | 2 |
| Separators | 6 | 6 | 5 | 6 | 5 | 5 |
| Ejectors | 2 | 2 | 1 | 1 | 0 | 2 |
| Rotating equipment | 2 | 2 | 3 | 3 | 4 | 2 |
| Equipment that need qualification | 1 | 1 | 2 | 2 | 1 | 1 |
| **Rich feed** | | | | | | |
| Flexibility | Low | Medium | Low | Medium | High | Not applicable |
| Additional remarks | Multiphase flow ejector | Multiphase flow ejector | Low motive pressure | Three-stage system for stabilisation | Independent from reservoir pressure | |
| **Lean feed** | | | | | | |
| Flexibility | Medium | High | Low | High | High | High |
| Additional remarks | Multiphase flow ejector | Multiphase flow ejector | No need for HP Sep. | No Need for HP Sep. | Independent from reservoir pressure | Compact system |

In **Figure 8.2** and **Figure 8.3**, the total power consumption, cooling load and coolers area is compared, data is taken from the results in **Section 6.2**. The total power consumption includes heat input, compressors and export pump. Heat input is only used for hydrate prevention, so the heat input is in the range 58-194kW for rich feed, and 6-21KW for lean feed. The pump uses only about 180kW and 16kW in the lean and rich case respectively. The rest of the power consumption is on the compressor, and mainly on the Export compressor, see tables in **Section 6.2**.

In calculation of heat exchanger area, it is assumed 5°C sea water temperature throughout the whole cooling process in calculation of LMTD.  It is assumed use of active subsea coolers with an overall heat transfer coefficient U=800W/m$^2$K.

There are only small deviations between these alternatives. In the lean case recompression flow rate is so small, that it has no real impact on these parameters. For the rich feed some deviation is found. The main cooling loads are found for the HP cooler and export cooler. The main power consumption is in the export compressor, the recompression part has low impact on power consumption.

In the rich feed scenario it is found that adding a screw compressor increases the power consumption with about 10%, compared to only using ejectors. The same 10% increase is also found for the cooler load and area. When adding a T-Separator for the rich feed scenario there is a bit larger increase in cooling load and area, on about 16% and 23% respectively. The reason for this additional increase is that the T-Separator operates at high pressure, giving a heavier gas phase, and larger mass flow rate going to the HP cooler.

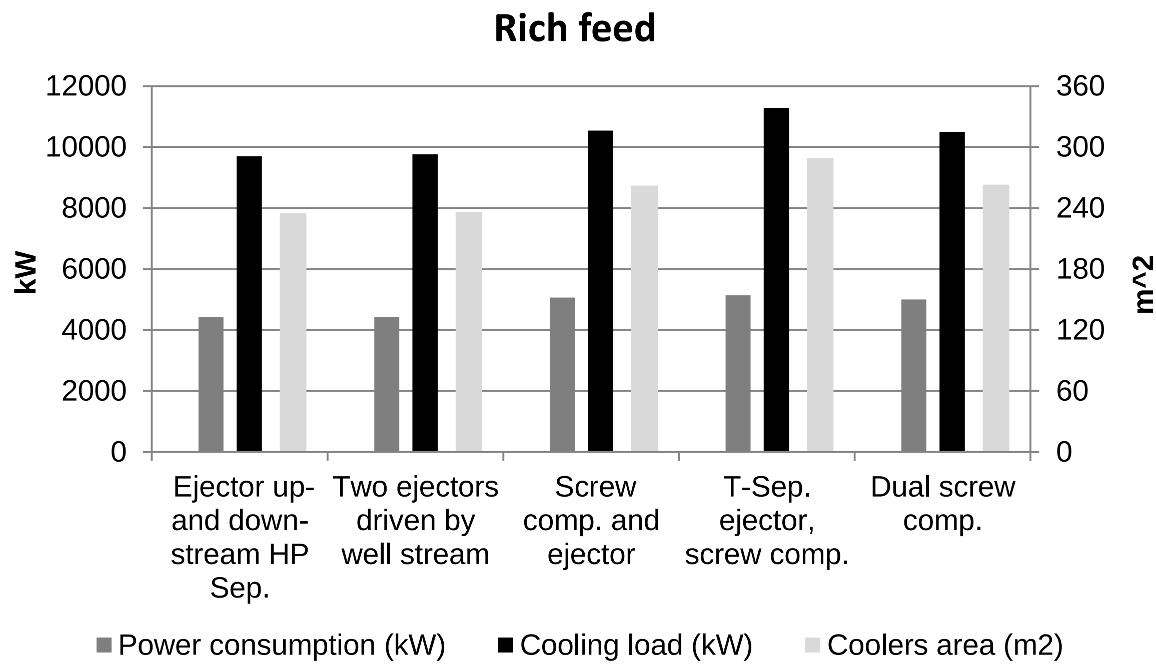



**Figure 8.2 Comparison of power consumption, cooling load and area, Rich Feed**

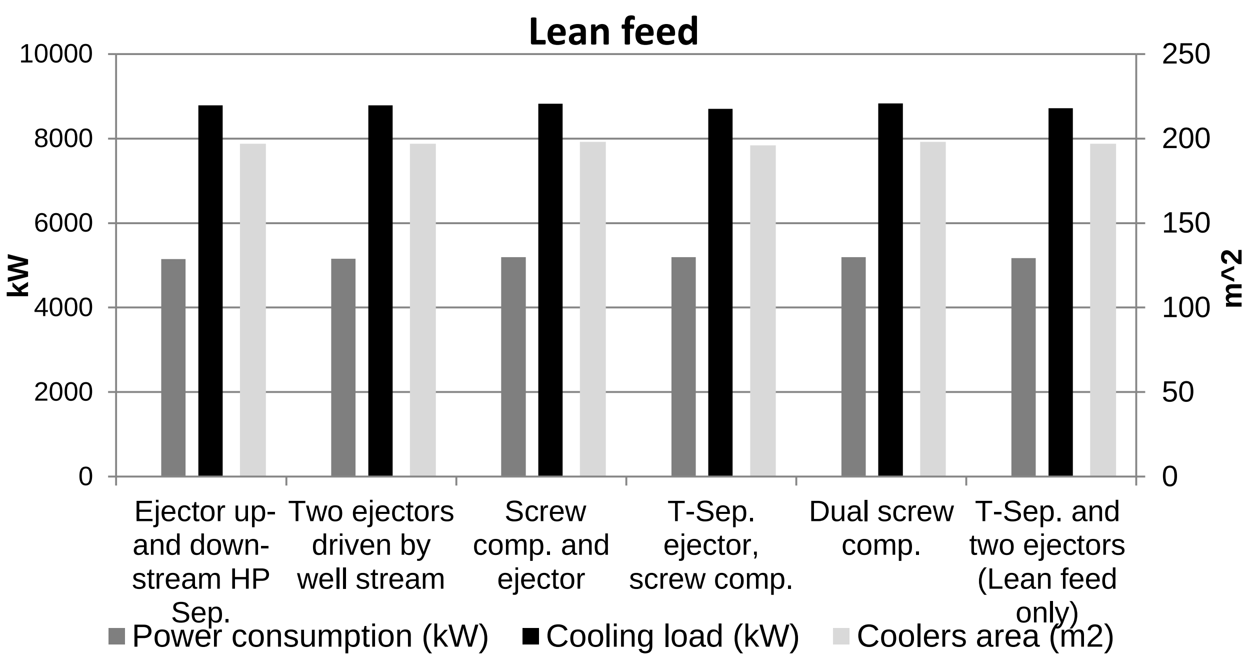



**Figure 8.3 Comparison of power consumption, cooling load and area, Lean Feed**

### *8.2.3 Heater for stabilisation of liquid*

For stabilisation of liquid there is possible to boil of light hydrocarbons by adding heat. Comparing the dual screw compressor system analysed with a heater in **Section 6.1** with the same system without a heater in **Section 6.2.6**, it is clear that using a heater gives a significant increase in power consumption for heating, see **Figure 8.4**.

For the rich feed scenario the power consumption used for heating is 2.6MW, for a pressure increase of only 4bara. This increases the total power consumption with about 50%. For the lean feed scenario the heater has lower impact, due to the low liquid flow. By adding 215kW in the heater the pressure in the LP separator is increased with 6bar. The total increase in power consumption in the lean feed scenario is about 10%.

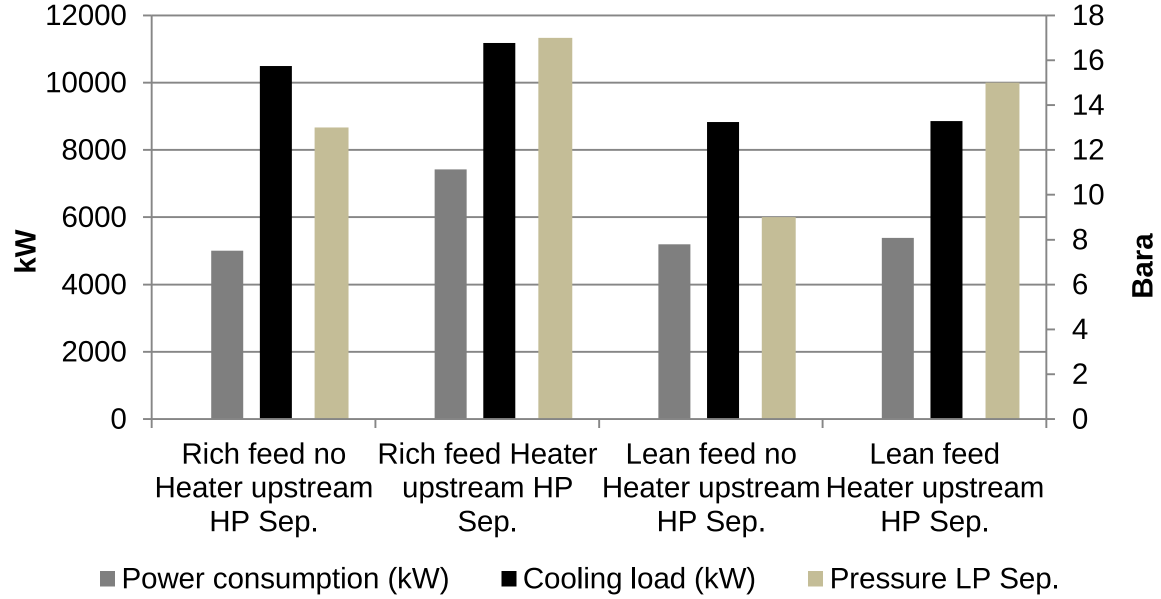



**Figure 8.4 Compassion of a process with, and without a heater upstream LP separator, (PFD see Figure 6.1 and Figure 6.17)**

## 8.3 Gas dehydration

In **Section 4.3** there is given a number of technologies for gas dehydration which is discussed in **Section 7**. Adsorption, selective membrane, and cooling/expansion are discussed in **Section 7**. These technologies were not chosen for further evaluation due to power consumption, need for technological qualification, concerns about handling of gas with entrained liquid or fouling gas, and ability to reach the dew point specification for Rich gas.

The technology chosen for analysis and evaluation is absorption by glycol. This is a well-known process often used topside to dehydrate gas to Rich gas dew point specification (-18°C at 70 bara). The benefit with using glycol is that it can be regenerated at a host. This removes the need for complex and energy intensive systems subsea. The only thing that must be subsea is a system that can provide sufficient contact between the gas and glycol, and then separate the two phases for single phase transport from the plant.

In **Section 7** there is three different glycol absorption solution analysed. These are based on the co-current gas/glycol contactor process in **Section 4.3.1.2**, which is developed as a robust and compact system suitable for gas dehydration subsea. Using mixers as the one shown in **Figure 4.9** should give sufficient contact between the two phases. The co-current contactors are robust equipment that can run without maintenance. Problems with foaming are reduced, compared to conventional counter-current contactors, so some entrained liquid in the gas phase can be handled. Separation of the two phases is then done by a gravitational separator, which already is in operation subsea reducing need for technological qualification.

Two of the systems in **Section 7** use dual co-current contactors with downstream separators. For the process shown in **Figure 7.1**, called the dual lean glycol mixer system, lean glycol is added in both mixers. The other process is the two-stage glycol system shown in **Figure 7.3**, here lean glycol is added in Mixer 2 and then pumped as a semi lean glycol into Mixer 1. From analysis done on these systems the dual lean glycol system use about two times the glycol circulation rate of what the two-stage glycol system, see **Figure 8.5**. The reason for this is that the semi lean glycol from Mixer 2 is in equilibrium with the dry gas, and is therefore able to pick up more water when contacted with the wet gas having higher water content.

The third system is a single mixer system with a downstream separator, see **section 7.3**. This system uses significantly more glycol than the system with two mixers, see **Figure 8.5**. But since absorption is favoured by high pressure, the glycol circulation rate is significantly reduced when increasing the pressure, see **Figure 7.6**. The single mixer system seems feasible for high pressure contacting (120bara).

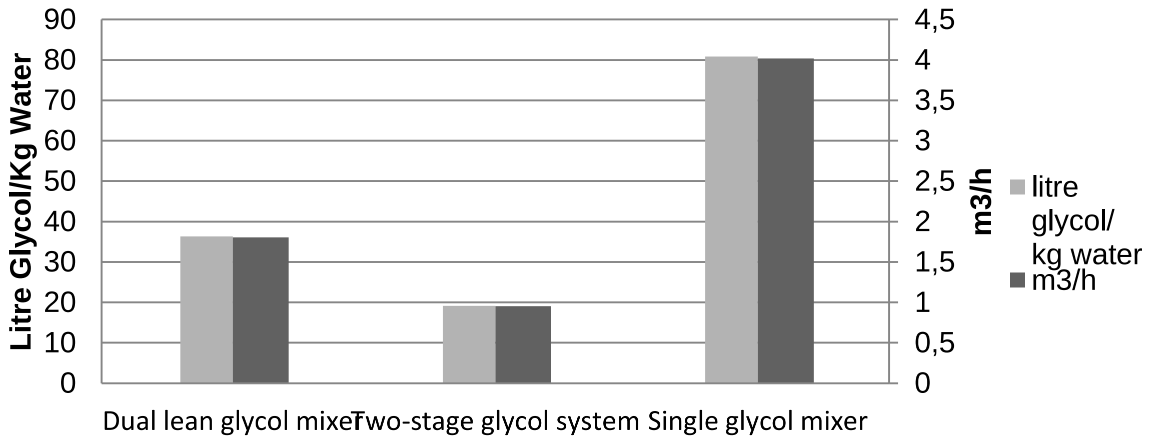



**Figure 8.5 Comparison of dehydration systems, (Rich feed, 98,5% MEG)**

**Figure 8.5** is based on the rich feed case, and using 98,5% lean MEG for dehydration. This is just chosen for comparison of the system. The same pattern is seen for all the other glycol alternatives, and differences between lean and rich feed is small in comparison of the systems.

It should be noted that in HYSYS simulations, there is found deviating results depending on choice of equation of state. Both Glycol package and Peng Robinson were tested, but without data to compare the result with, it is difficult to say which of these that gives the most accurate results. Which equation of state that is used will not have impact on comparison of dehydration systems, only on the accuracy of the glycol circulation rate for individual systems.

Both MEG and TEG is tested in simulations, but it is not clear which one that should be selected. MEG has the advantage that it most likely is available for hydrate inhibition. Use of the same glycol for dehydration and hydrate inhibition will reduce facility and infrastructure needs. The drawback with MEG is that it is difficult to regenerate to needed purity, but there is suppliers that claims to be able to regenerate MEG to 99,5% purity (CAMERON, 2015). TEG has no technological gaps for regeneration to high purity, and regeneration of TEG is less energy intensive than for MEG. Another issue with MEG compared to TEG, is that it has a higher solubility in the gas phase. So a high efficiency separator is needed to avoid breaking the limit of MEG content in the gas.

## 8.4 Discussion on subsea heat and power production

Subsea heat and power production is briefly addressed in **Appendix A**. The focus is on technologies that can be used subsea. There are also a number of topside alternatives available for production of electrical power, for example the sun, wind, waves, fuel cells and conventional combustion machines. But all of this requires some kind of topside installation. Subsea there is much more restricted possibilities for production of power, but there is some alternatives which can use marine current or heat to produce electricity. The use of sea current will need a relative large current to produce enough energy for the entire plant. The turbine called HS1500, needs about 3m/s to produces 1.5MW.

The other solution presented utilise thermo-electrical generators for power production. This can convert heat directly to electricity when there is a temperature difference over the generator. This seems very interesting to use in a subsea system, where there is coolers that can be packed with these generators. The generator has low maintenance without moving parts. The drawback is that it is only able to convert about 10% of the heat to power. So even for the systems with the lowest power consumption analysed in this study, there will be needed 50MW of heat. But the thermal generator can at least be a good option to save imported power consumption.

Heat production is even more restricted. In this study it has been assumed that heat is produced by electrical heaters. Combustion on the sea floor would be difficult since air must be added from topside, and maintenance is also a concern. The only real alternative to direct electrical heating is use of geothermal heat. In production of hydrocarbons there is already need for drilling, so drilling of a geothermal well nearby should not be very expensive. Use of geothermal heat with high temperatures is still in a research stage at this point, but if the DESCRAMBLE research project is successful, the possibilities for utilisation of such high temperature wells will be feasible. Another concern is how deep such a well must be to get a high enough temperature, this must be considered for each area.

# 9. Conclusion

The main objective of the thesis has been to further develop and evaluate solutions for subsea processing of hydrocarbons, with focus on simplicity, utility need(power, heat, glycol), and operational flexibility. The most important has been that the subsea processes must be able to produce gas and liquid products, which have acceptable specifications for transportation and further processing in a downstream process.

This study reviles that it is possible to apply a simple and robust system for subsea processing of a hydrocarbons, which are able to deliver Rich gas specifications and a partial stabilised liquid. Using robust equipment like oil-free screw compressors and/or ejectors, removes the need for scrubbers in recompression of flash gas seen in previous designs, but there will be need for further development and a technology qualification program for the equipment.

The system design and process parameters are clearly depending on transportation method for the liquid product. Transportation method sets the vapour pressure specification for the liquid product, which is found to be a significant parameter. These findings are in accordance with earlier research on subsea stabilisation (Kraabøl, 2015) (Hove, 2013). It will be preferred to use pipe transportation, providing the least constrained scenario with a high vapour pressure (TVP<10bara). The systems presented can also be used to produce liquid for transportation with semi pressurised tankers (TVP<5bara), but this will decrease operational flexibility and reduce internal pressure in the low pressure part of the system. For complete stabilisation (TVP<1bara), a much more complex and energy intensive system is needed. It is not clear that such a complex system in a subsea environment will surpass the benefit of a complete stabilised liquid.

The dual screw compressor solution presented in **Section 6.2.6**, is to recommend for subsea processing with partial stabilisation of the liquid product. There is used a two-stage system for stabilisation, and dual screw compressors for recompression of flash gas. The system has the highest operational flexibility and controllability of all systems considered, and can handle both lean and rich feeds. It is also independent from the reservoir pressure. For rich feed scenarios, there is needed at least one screw compressor for recompression. Using dual screw compressors, instead of one screw compressor and one ejector, will ease development and lower maintenance costs for the process plant. Neither is it seen to increase power consumption. There will be need for a technology qualification program to implement oil-free screw compressors subsea.

For all the solutions with ejectors, presented in **Section 6.2**, lean feed provides much better flexibility than rich feed. With the high flexibility as found in lean feed scenarios, it is possible to operate a process using two ejectors in series for recompression of flash gas. The recommended process for lean feeds is the solution presented in **Section 6.2.5**. This process has a high pressure inlet separator represented by a T-Separator and two ejectors in series for recompression. This is a very robust, compact, and simple system, where the only rotating equipment is the export compressor and export pump. The large operational flexibility for lean feed scenarios reduces the impact of falling reservoir pressure. There is need for development of a controllable ejector system to make this system feasible. The big drawback is that this system is not feasible for rich feed scenarios.

For gas dehydration the two-stage glycol system presented in **Section 7.2** is clearly the preferred gas dehydration system. By adding lean glycol only in the Mixer 2, the glycol circulation rate is halved, compared to using lean glycol in both mixers. Detailed research to find exact amount of glycol and type of glycol needed for this process should be done.

Use of a heater in the stabilisation process would ease separation of water and oil, due to reduced viscosity, and increase internal pressure in the low part of the system. The issue with the heater is that it is energy intensive if not used carefully. In this study it is not found need for a heater for partial stabilisation, but this must be considered for each field development. If a high temperature heater should be used subsea, there would be need for a technology qualification program.

The recommended systems can also be implemented on an unmanned topside installation. Operational flexibility for ejectors that is driven by the well stream will decreases with increasing water depth, since the fluid must be lifted to a higher attitude losing some of the motive pressure. The dual screw compressor solution presented in **Section 6.2.6**, is independent of the inlet pressure, so this would be recommended for a unmanned installation.

Production of heat and power subsea is in an early stage, but has large potential in the right areas. Using geothermal heat has potential for use in a subsea process, and then a heater could be used without influencing the power consumption. Producing electricity from marine current will provide stable and sufficient power production if the current is large enough. Thermal-generators can be used to reduce use of imported power, but has too low efficiency to provide electricity for the whole plant. Combination of geothermal heat and thermal-generators can be an option to produce more power.

# Recommendations for further work

- Life cycle cost analysis should be conducted on the most promising solutions.
- Technology qualification program for subsea screw compressors.
- Technology qualification program for high temperature electrical heaters
- Develop robust and controllable ejector packages.
- Selection and detailed design of subsea separators.
- Detailed study of glycol requirements for dehydration and hydrate inhibition.
- Detailed design of subsea coolers, in special active coolers.
- Analysis of subsea heat and power production.

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# Appendix A Subsea heat and power production

In this appendix a brief presentation of some possibilities for subsea heat and power production is presented. The focus has been to present a solution that is possible to install subsea in conjunction with the subsea process plant. In addition to technologies presented there could be possible to use offshore windmills, wave power, collar cells, fuel cells, etc., but all these technologies will be or need topside installations to operate.

## A.1 Geothermal energy

Geothermal energy is available in various degrees throughout the globe, and can at least be an option for subsea heating. There is also developed subsea equipment for direct conversion of heat to electricity, see **Section A.2.2**, that can be used to produce electricity from geothermal energy. Since it already is needed to drill down to the hydrocarbon well, drilling an geothermal well should be possible without adding large expenses. In some fields there can be wells producing water which can be used for heating in the subsea plant.

Inside the earth there is a large energy storage known as geothermal energy, which can be utilised with the right technology. It is estimated that 99% of the earth body is above 1000°C and less than 0,1% is colder than 100°C. Temperature increases with depth and the normal geothermal gradient is 3°C per 100m, but there are local differences in geothermal gradients. Actual geothermal gradient can be estimated from downhole measurements. (Stober & Bucher, 2013)

Geothermal heating can be divided into shallow and deep geothermal systems. In **Figure A.1** geothermal systems is shown with a characteristic power output for the different depths. Shallow systems are in most cases operating at 150m or at most down to 400m, and utilisation of the energy can be done through heat pumps where the temperature is increased. Deep geothermal systems are operating below 400m, but for low-enthalpy systems the most realistic is operation below 1000m and above 60°C. In areas where there is volcanic activity, like on Iceland, high-enthalpy fields can be found near the surface producing high temperatures. (Stober & Bucher, 2013)

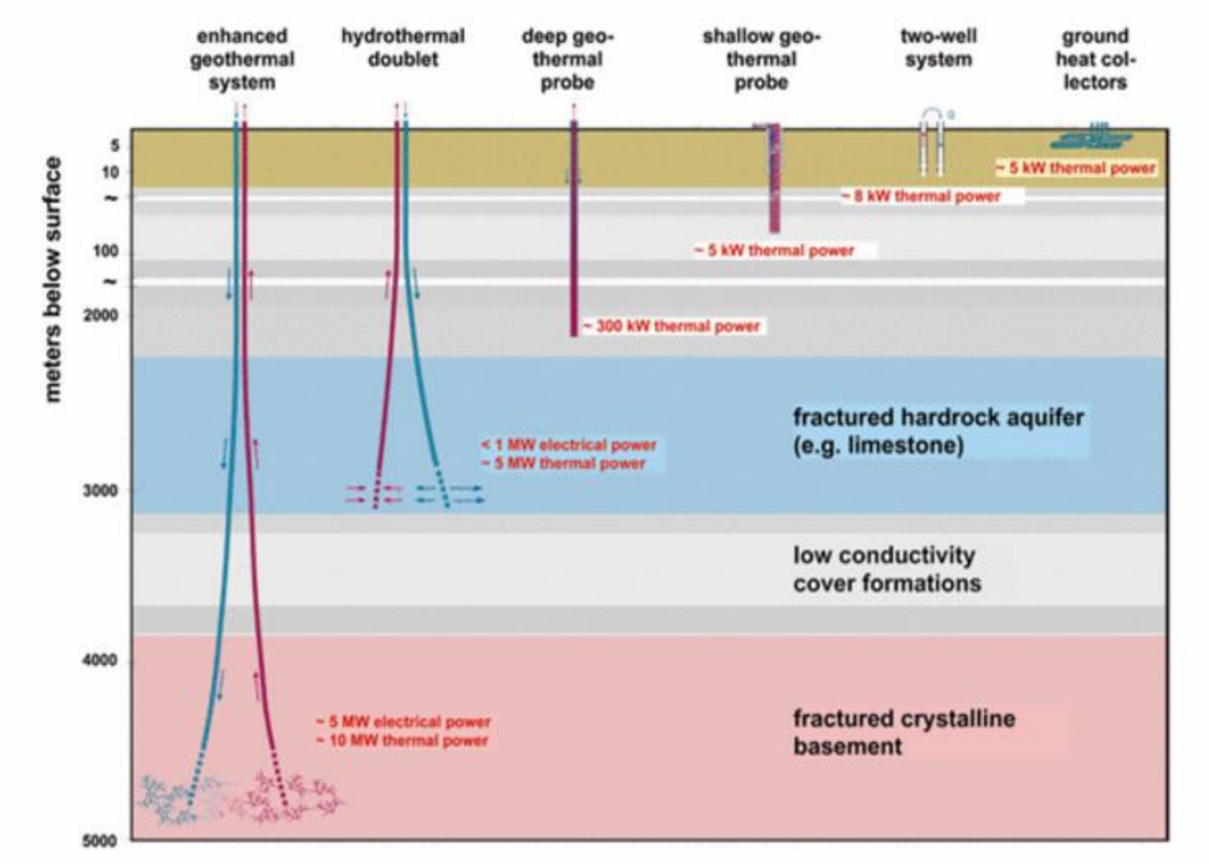


**Figure A.1 Geothermal systems with characteristic power output (Stober & Bucher, 2013)**

In deep geothermal systems it can be possible to use an injection and production well drilled from the same site by inclined drilling, as seen in **Figure A.1** for the deepest wells. It is important that the distance between the wells is large enough to avoid that cold reinjected water is cooling down produced hot water. But the injection well needs to be close enough to provide hydraulic support. (Stober & Bucher, 2013)

In Lardello in Tuscany, Italy, there is an ongoing project called DESCRAMBLE (Drilling in dEep, Super-CRitical AMBients of continentaL Europe). They are going to extend the depth of an existing well from 2,2km to 3-3,5km, to produce supercritical water making the well produce ten times more energy than a standard geothermal well. Supercritical water is found above 374°C and 218atm, with entirely different properties than normal water which demands specially developed equipment. In the bottom of the well there will be about 450°C. Success in this project will be a major technological breakthrough, and it will reduce financial and technical risk in development and operation of deep geothermal wells. (Benjaminsen & Stamnes, 2015) (DESCRAMBLE, 2015)

# A.2 Electrical power production subsea

## A.2.1 Power from marine current

In the ocean there is tide movements and ocean circulation generating currents which can be used for power generating. This current is much more reliable than wind and sun power, which can reduce need for power storage subsea. The drawback is that there will be need for relatively large flow rates passing the turbines, so the current in the subsea location must be considered carefully before choosing this technology.  But if there is a large current the power production can be sufficient for the whole subsea process.

Energy from ocean current can be converted to mechanical energy using a rotating or reciprocating device, which can be further converted to electrical power in a generator. Compared to other renewable sources the marine currents are predictable and so far it seems to have no impact on the environment it is placed in. The drawback is that development is in an early stage making it expensive to build commercial power plants, the development is driven by government support and technological advancements. (Rourke, Boyle, & Reynolds, 2009)

One of the first pilots for marine current power production was a tidal stream turbine, HS300, installed in Kvalsundet at the Norwegian north coast. The Turbine could deliver 300kW and was in opearation from 2003-2012. Based on technology from the pilot(HS300) HS1000 was developed and installed outside the Orkney Islands in 2011, with a power output of 1MW. On the ongoing Meygen project HS1500 and AR1500 is under development and will deliver 1,5MW at a flow rate of 3m/s. In 2016 there will be installed three HS1500 and one AR1500 in the Meygen project. In the second phase of the project 60 turbines will be in production by 2020, and the total goal for the project is to install 269 turbines producing 398MW of electrical power from the marine current. See **Table A.1** and **Figure A.2** for data and schematic drawing of the turbines. (Meygen, 2016) (Nilsen, 2015)

**Table A.1 Marine current turbine data (Meygen, 2016)**

| Designation and unit | HS300 | HS1000 | AR1500/ HS1500 |
|---|---|---|---|
| Power (kW) | 300 | 1000 | 1500 |
| Weight (tonne) | | | 200 |
| Flow speed (m/s) | | | 3 |
| Rotor diameter (m) | | | 18 |

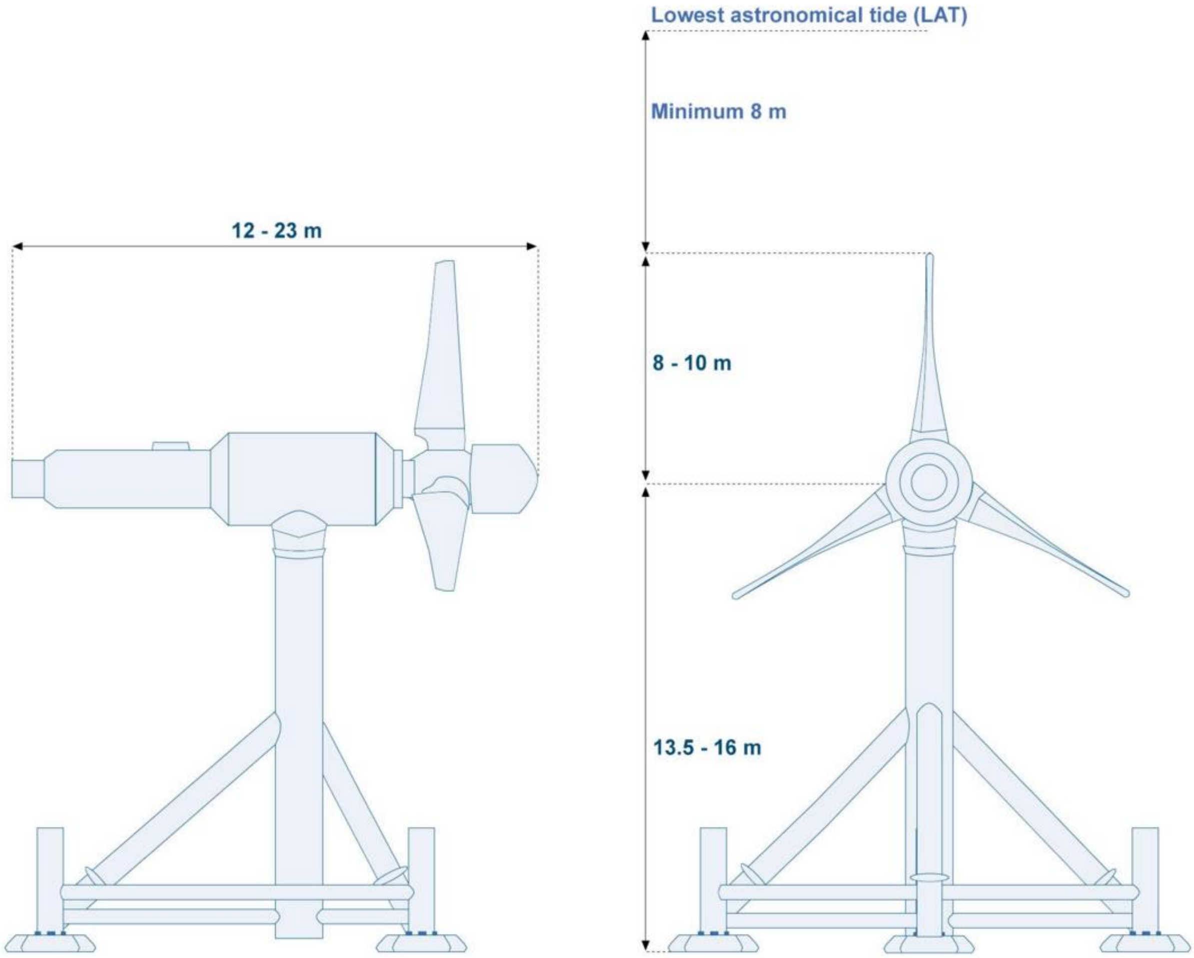


**Figure A.2 Turbine Schematic** (Meygen, 2016)

### *A.2.2 Thermo-electrical generator*

This type of generator can be added to conventional subsea coolers, geothermal heat systems, or other available heat sources. The drawback is that only about 10% of the heat can be converted to electricity so there will be needed a large heat source if the whole subsea process should run on this system. For the process systems tested in this study, that is operating with TVP<10bara for the liquid product and without a large heat input, see **Section 6.2**, there will be needed about 50MW heat to produce enough power for the whole process.

A thermo-electrical generator creates electrical power directly from a temperature difference by utilising the Seeback effect. The generator can be composed by a p-type and n-type semiconductor connected in an electrical circuit as shown in **Figure A.3**. (Rosi, 1968)

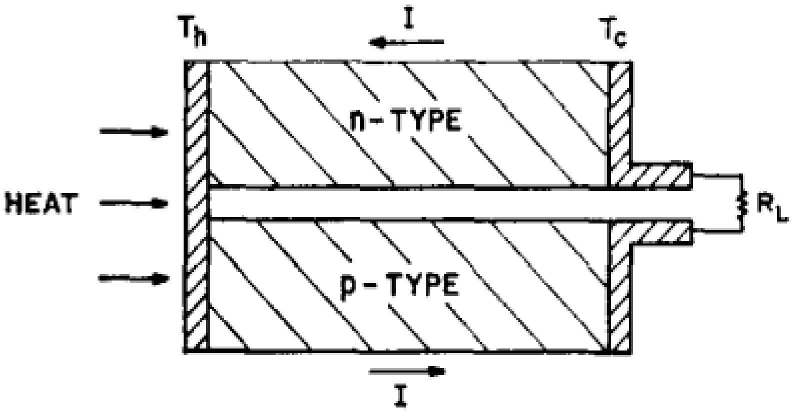


**Figure A.3 Thermocouple as power generator (Rosi, 1968)**

In a subsea processing, temperature difference between process fluids and sea water can be used to create electrical power. Placing multiple thermocouples on a tube or in a subsea module will convert waste heat into electrical power with about 10% efficiency, see **Figure A.4**. (Ellison, 2015)

**Figure A.4 Thermoelectric generators** (EXNICS, 2016)

# Appendix B Process solutions, Pros and cons

In this appendix the key findings is presented, looking at advantages and disadvantages.

**Table B.2 Process comparison**

| Process design | Advantages | Disadvantages |
|---|---|---|
| Ejector upstream and downstream HP separator **Section 6.2.1** **Figure 6.3** | ➢Robust (Only export pump and compressor is rotating equipment) 1% increased liquid production with three stages for rich feed scenario | ➢Dependent on multiphase well stream as motive stream in ejectors ➢Low operational flexibility ➢Three bulky separators for liquid stabilisation |
| Two ejectors driven by the well stream **Section 6.2.2** **Figure 6.8** | ➢Robust (Only export pump and compressor is rotating equipment) ➢Good flexibility, in special for lean feed with low recompression flow rate. | ➢Dependent on multiphase well stream as motive stream in ejectors ➢Three bulky separators for liquid stabilisation |
| Screw compressor and ejector **Section 6.2.3** **Figure 6.11** | ➢More stable operation of ejector (Motive flow is gas flow from HP separator) | ➢Low flexibility (Low pressure in the HP separator compared to the well stream) ➢Decreased robustness and increased power consumption (Screw compressor used) |
| T-separator, Ejector and screw compressor **Section 6.2.4** **Figure 6.13** | ➢Good flexibility for both lean and rich feed. ➢Increased controllability of ejector (The motive stream is high pressure gas separated from the feed stream.) | ➢Too complex for lean feed (No fluid going to HP separator) ➢Decreased robustness and increased power consumption ( ➢Increased cooling load for rich feed |
| T-separator and two-stage ejector **Section 6.2.5** **Figure 6.15** | ➢Simple process for lean feed ➢Robust (Only export pump and compressor is rotating equipment) ➢Compact system (high pressure inlet separator) ➢High flexibility and controllability for lean feed with low recompression flow rate | ➢Not feasible for rich feed with high liquid production and larger recompression flow rate. |
| Dual screw compressors **Section 6.2.6** **Figure 6.17** | ➢High operational flexibility for both lean and rich feed ➢Recompression system is not dependent on available motive flow | ➢Decreased robustness and increased power consumption due to the screw compressors, compared to solutions with only use of ejectors. |
| Use of Heater for liquid stabilisation **Section 6.1** **Figure 6.1** | ➢Increased internal pressure in LP separator. ➢Improves separation of water and hydrocarbon liquid. | ➢High power consumption for low pressure increase.(2.6MW gives 4 bara in the rich feed case, 215kW gives 6bara for the lean feed) |
| Three-stage system for complete stabilisation **Section 6.1.1** **Figure 6.2** | ➢Able to produce a completely stabile liquid product with TVP<1bara | ➢Complex system, with 5-7 extra units compared to the other systems ➢High power consumption (with heat input of 3.7MW the LP separator operates at 2bara internal pressure) |

**Table B.3 Process comparison continues**

| Process design | Advantages | Disadvantages |
|---|---|---|
| | **Gas Dehydration** | |
| Dual lean glycol mixer system **Section 7.1** **Figure 7.1** | ➢Simple and robust system with no moving parts | ➢Relatively high glycol circulation rate |
| Two-stage glycol mixer system **Section 7.2** **Figure 7.3** | ➢Halved glycol circulation rate compared to the dual lean glycol mixer system (two equilibrium stages) | ➢Needs a glycol pump |
| Single glycol mixer **Section 7.3** **Figure 7.5** | ➢The simplest system | ➢High glycol circulation rate<br>➢Only feasible for high pressure and low temperature absorption |
| | **Subsea heat and power production** | |
| Geothermal energy **Section A.1** **Figure A.1** | ➢Large potential for heat production<br>➢Low cost, (drilling is already needed to produce hydrocarbons) | ➢Depends on plant location (how fare down it must be drilled to get high enough temperature)<br>➢Development for high temperature wells is at a research stage |
| Power from marine current **Section A.2.1** **Figure A.2** | ➢Good potential for power production in areas with large marine currents | ➢Needs about 3m/s current to produce 1.5MW<br>➢In the start phase for commercial operation, so it is relatively expensive at this point |
| Thermo-electrical generator **Section A.2.2** **Figure A.4** | ➢Robust equipment for power production, with no moving parts | ➢Large heat flows is needed to produce enough energy to run the whole process, but can be used to save imported power. (with 10% efficiency about 50MW is needed to drive the systems with 5MW power consumption) |

# Appendix C Ejector efficiency

In this appendix development of the ejector efficiency is presented.

Maximum recovered power can be calculated as shown in equation (8). The upper limit of the integral is found by an isenthalpic throttling process from the motive pressure to the exit pressure, point A in **Figure 3.6**. The lower limit is found by isentropic expansion of the fluid from motive pressure to the exit pressure, point B in **Figure 3.6**. (Elbel & Hrnjak, 2007)

$$W_{recmax} = m_M \int_{S_B}^{S_A} T_{out}\, ds \tag{8}$$

$W_{recmax}\,(kW)$ maximum possible expansion power, , $m_M(\frac{kg}{s})$ is motive mass flow, $T_{out}\ (K)$ is the outlett temperature in Kelvin

Basic thermodynamic T dS equation (Moran & Shapiro, 2012):

$$Tds = dh - vdP \tag{9}$$

T(K) temperature, $s\ (\frac{kj}{kgk})$ entropy, $h\left(\frac{kj}{kg}\right)$ enthalpy, $v\ (\frac{m^3}{kg})$ specific volume, $P(kPa)$ pressure.

Since the pressure at the exit is equal for the two expansion processes the dP equals zero in equation (9). Combining equation (8) and (9) simplifies the calculation as shown in equation (10):

$$W_{recmax} = m_M(h_A - h_B) \tag{10}$$

$W_{recmax}\,(kW)$ maximum possible expansion power, , $m_M(\frac{kg}{s})$ is motive mass flow, $h\left(\frac{kj}{kg}\right)$ enthalpy.

Work recovered by the ejector can be calculated from equation (11), (Elbel & Hrnjak, 2007):

$$W_{rec} = m_s \int_{P_D}^{P_C} v(P)dP \tag{11}$$

$W_{rec}(kW)$ Recovered expansion power, $m_s(\frac{kg}{s})$ suction mass flow, $v\ (\frac{m^3}{kg})$ specific volume, $P(kPa)$ pressure.

To perform the integration of equation (11) knowledge of how specific volume changes with pressure is needed. By assuming isentropic compression and combining equation (9) and (11) the recovered power can be calculated according to equation (12). Assuming isentropic compression is the most conservative approach, as it will give the lowest possible amount of work recovered by the ejector.

$$W_{rec} = m_S(h_D - h_C) \tag{12}$$

$W_{rec}\ (kW)$ maximum possible expansion power, , $m_S(\frac{kg}{s})$ is motive mass flow, $h\left(\frac{kj}{kg}\right)$ enthalpy.

Ejector efficiency can now be calculated from equation (6):

$$\eta_{ejec} = \frac{m_S(h_D - h_C)}{m_M(h_A - h_B)} \tag{13}$$

$\eta_{ejec}$ Ejector efficiency, $m_s(\frac{kg}{s})$ is suction mass flow, $m_S(\frac{kg}{s})$ is motive mass flow, $h\left(\frac{kj}{kg}\right)$ enthalpy (Point A-B-C-D is shown in **Figure 3.6**)

# Appendix D Pump and compressor technologies

Some of the technologies available pump technologies are shown in **Figure D.5**, and compressor technologies are shown in **Figure D.6**.

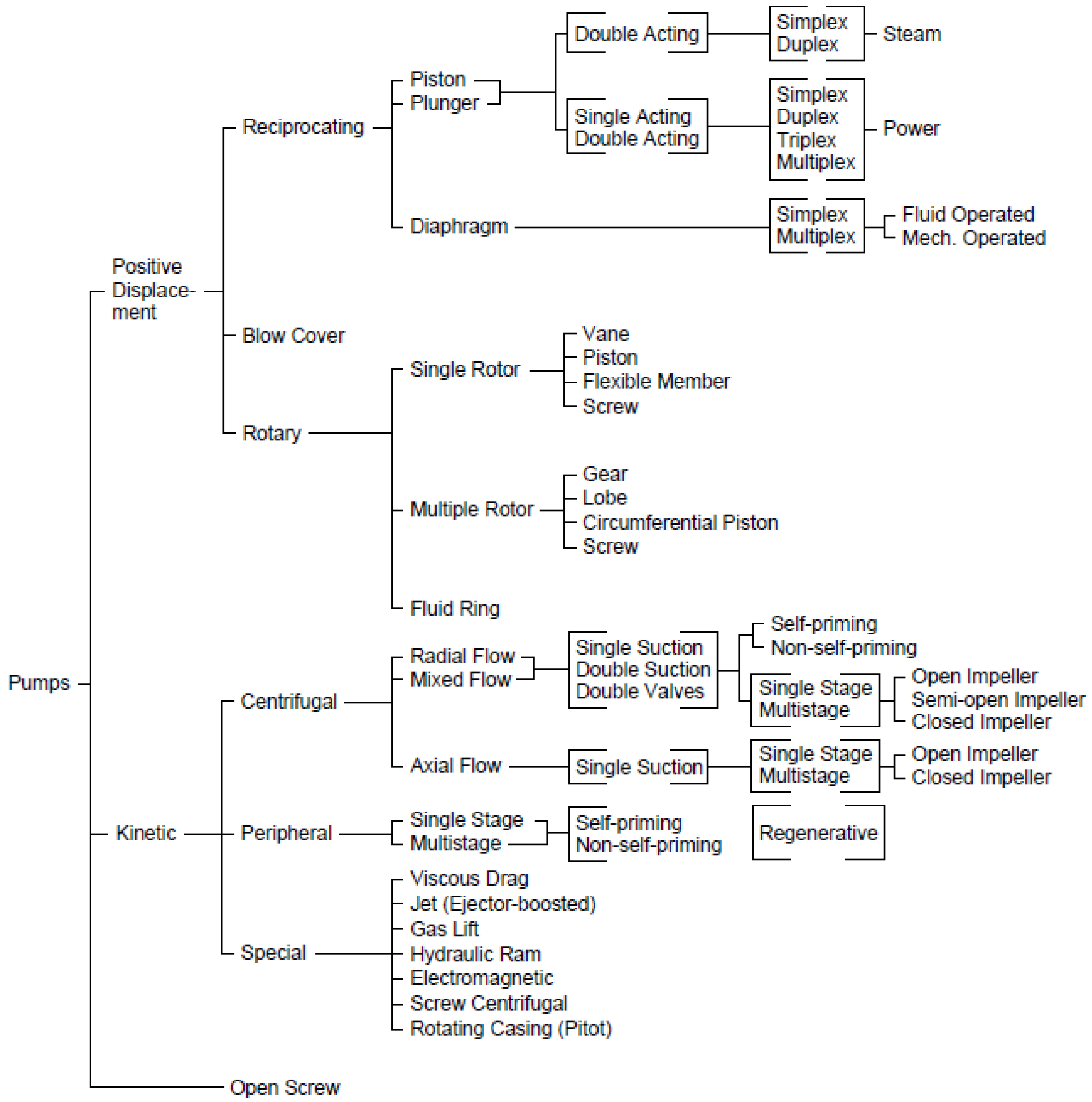


**Figure D.5 Classification of pumps (Perry & Green, 1997)**

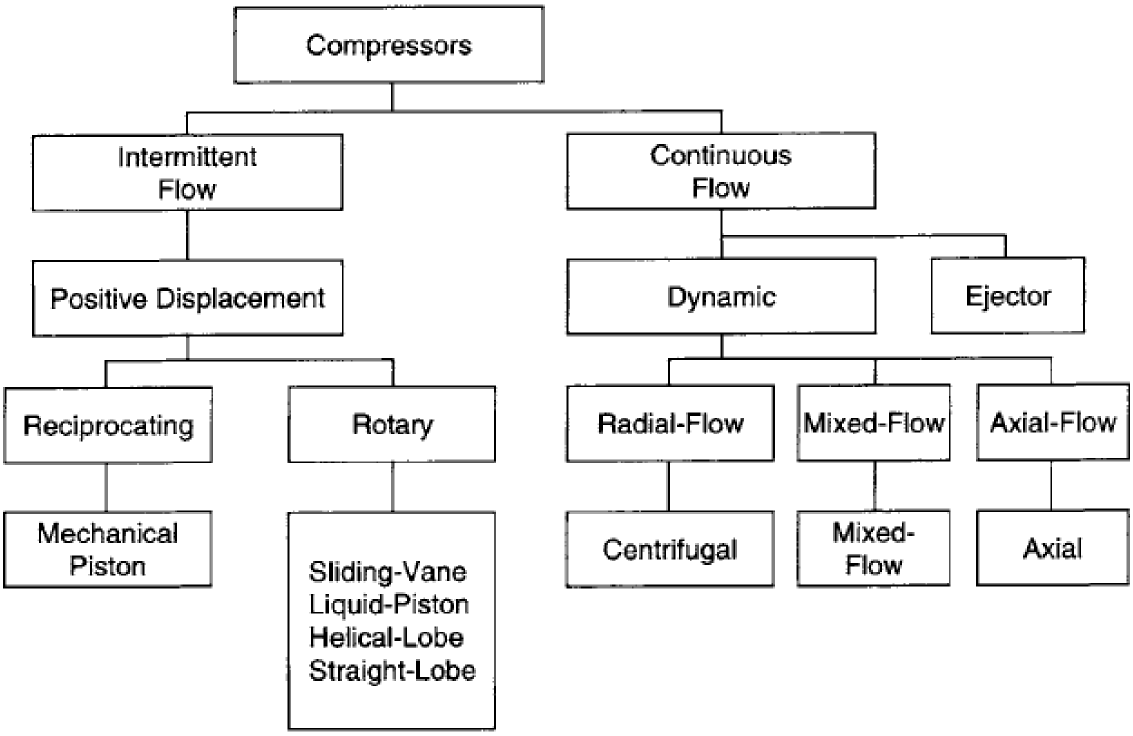


**Figure D.6 Classifications of compressors (Brown, 2005)**

# Appendix E Glycol dehydration

Here is diagrams used for prediction of water content, TEG purity and thermal decomposition temperatures for glycols presented.

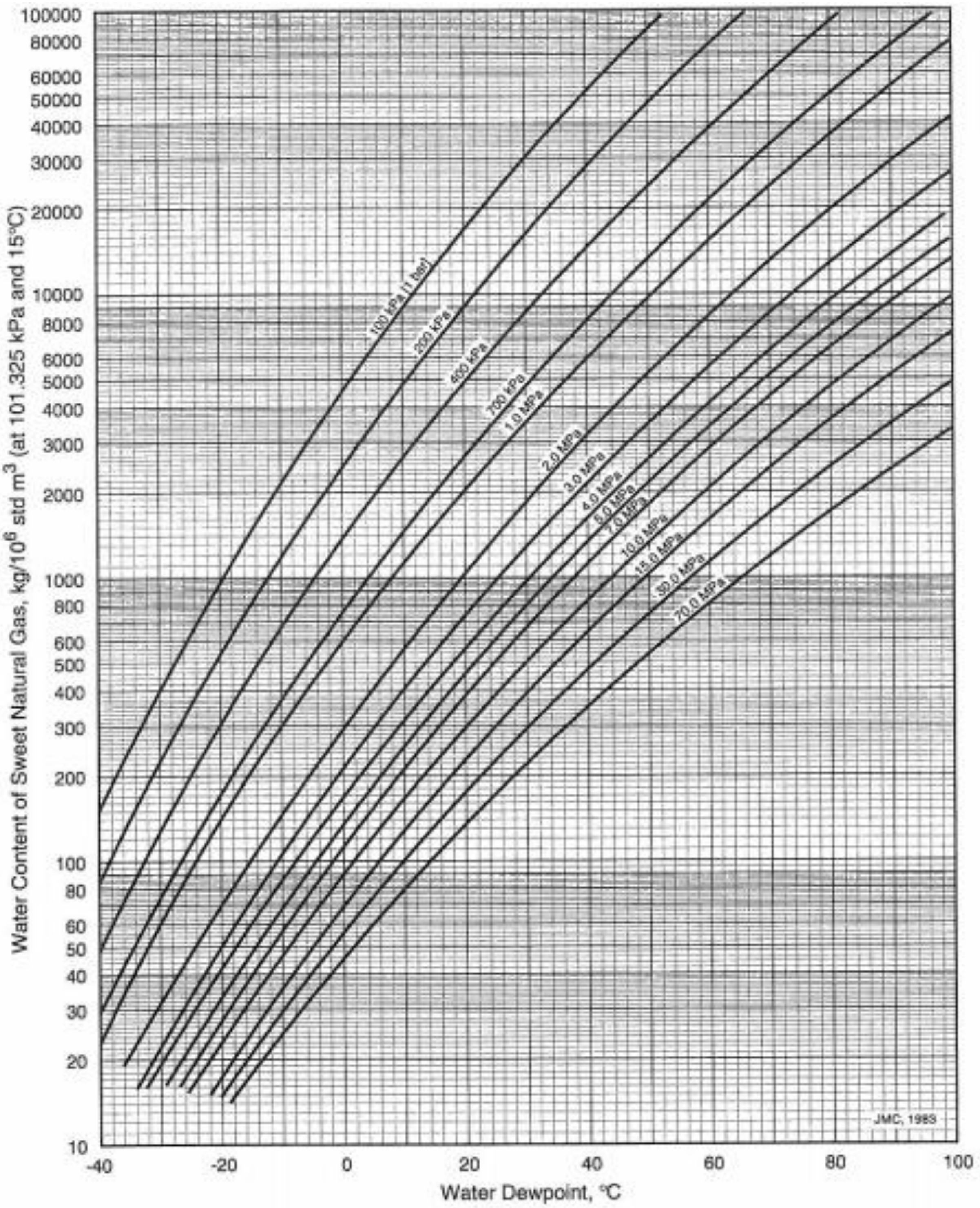


**Figure E.7 Water content of Sweet natural gas**

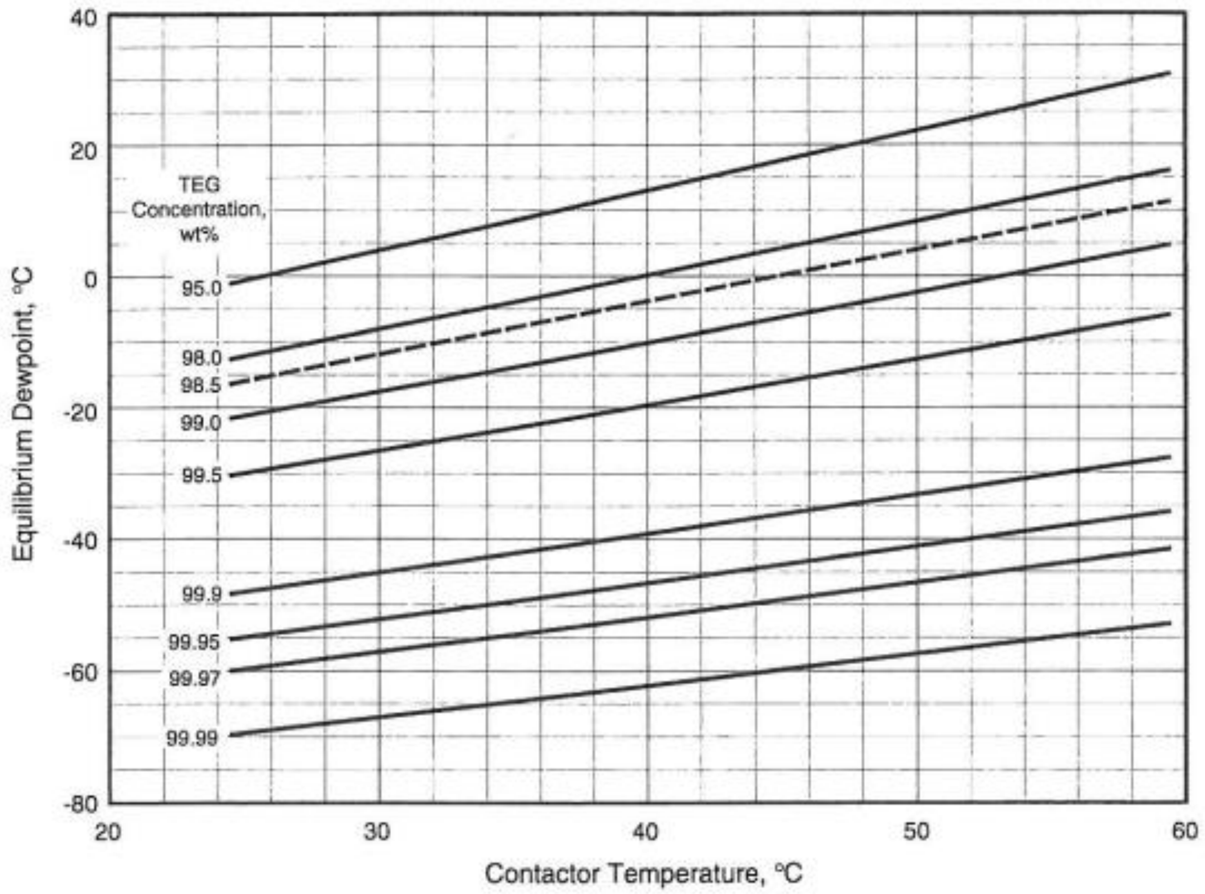


**Figure E.8 TEG Concentration chart with equlibrium dewpoint**

| Glycol | Decomposition Temperature | Lean Glycol* Concentration, wt% | Equilibrium** Water Dewpoint @ 38°C [100°F] |
|---|---|---|---|
| EG | 165°C [329°F] | 96.0 | 3°C [37°F] |
| DEG | 164°C [328°F] | 97.1 | 3°C [37°F] |
| TEG | 206°C [404°F] | 98.7 | -8°C [18°F] |
| TREG | 238°C [460°F] | > 99 | -18°C [0°F] |

* Equilibrium concentration at decomposition temperature and 1 atm

** At concentration in column at left

**Figure E.9 Thermal decomposition temperatures for glycol (Campbell, 1992)**

# Appendix F Status of the subsea technology

INTECSEA, Inc. and Offshore magazine has presented posters for status of subsea technology since 2008. Some parts of the poster from 2016 and 2014 are presented in this appendix.


**Figure F.10 Gravity separation systems, March Status 2016** (INTECSEA & Magazine, INTECSEA.com, 2016)

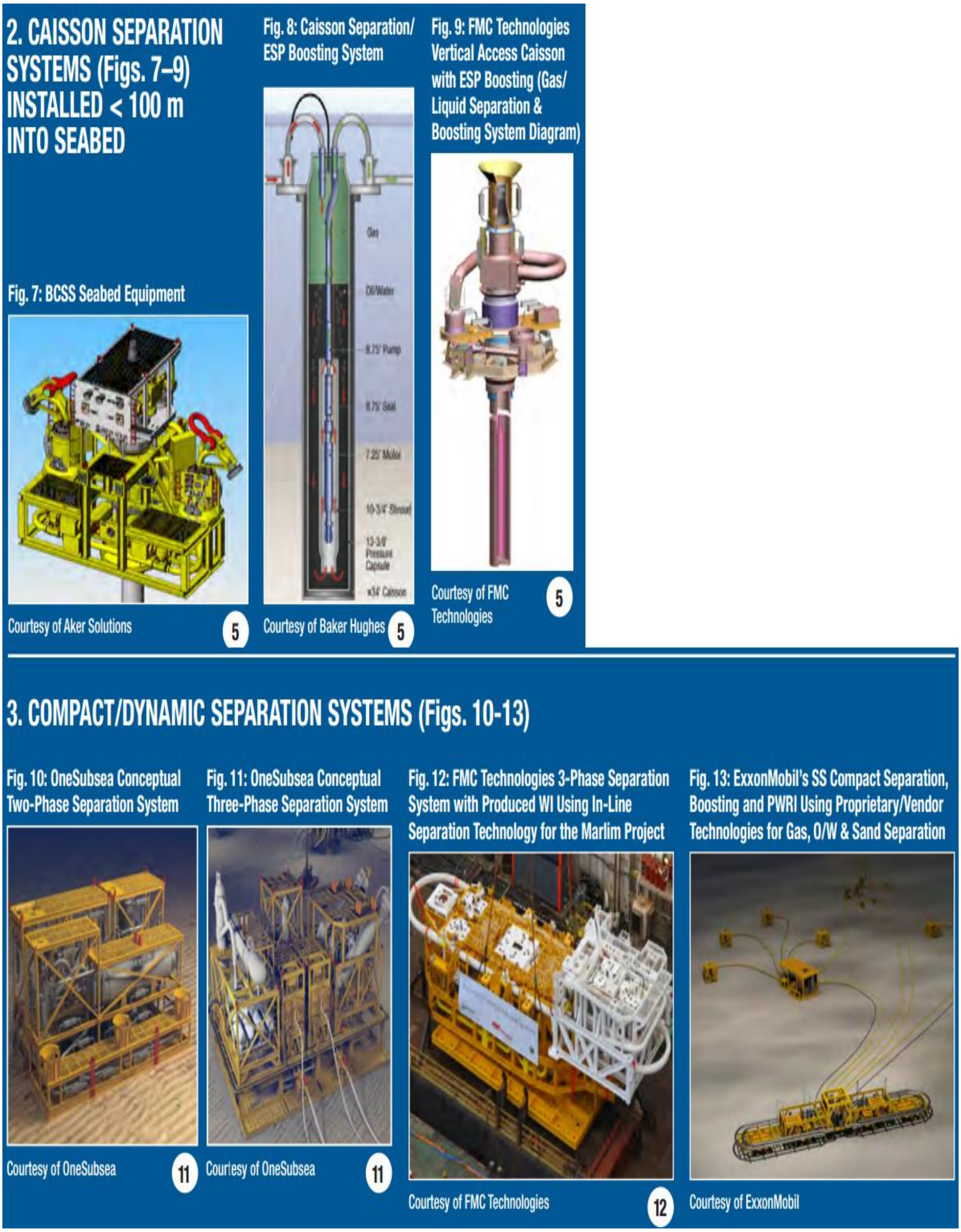


**Figure F.11 Caisson(2) and Compact separation systems (3), Status 2016** (INTECSEA & Magazine, INTECSEA.com, 2016)

## SUBSEA SEAWATER TREATMENT AND INJECTION

**Fig. 1: Aker Solutions' LiquidBooster™ Subsea Raw Seawater Injection System (Photo: System for Statoil Tyrihans Subsea Raw Water Injection Project)**

Courtesy of Aker Solutions

**Fig. 2: Conceptual Illustration of Installation of Tyrihans Subsea Raw Seawater Injection (SRSWI) System**

Courtesy of Aker Solutions

**Fig.3: OneSubsea Raw Seawater Injection System being installed for Columbia E Field**

Courtesy of OneSubsea

**Figs. 4 & 5: Saipem-Veolia –Total Conceptual and Tested**

Westgarth Conceptual and Tested Prototype SS Sulphate Removal Unit capable of treating 60 kbwpd

Courtesy of Saipem

**Fig. 6: GE Subsea Sulphate Removal and Injection System built on a combination of GE ultra-filtration and nano-filtration membrane technologies scalable to any capacity.**

Courtesy of GE

**Fig. 7: SWIT™ Technology**

SWIT™ a fully integrated subsea water treatment facility for 20 kbpd sulfate free and low salinity water. The combination of Seabox™ and Microfiltration modules only delivers water free of suspended solids.



Courtesy of NOV

**Fig.8: OneSubsea's Testing of the Albacore Raw Seawater Injection System during SIT of Pump and Filtration System**

Courtesy of OneSubsea

**Figure F.12 Subsea seawater treatment and injection** (INTECSEA & Magazine, INTECSEA.com, 2016)

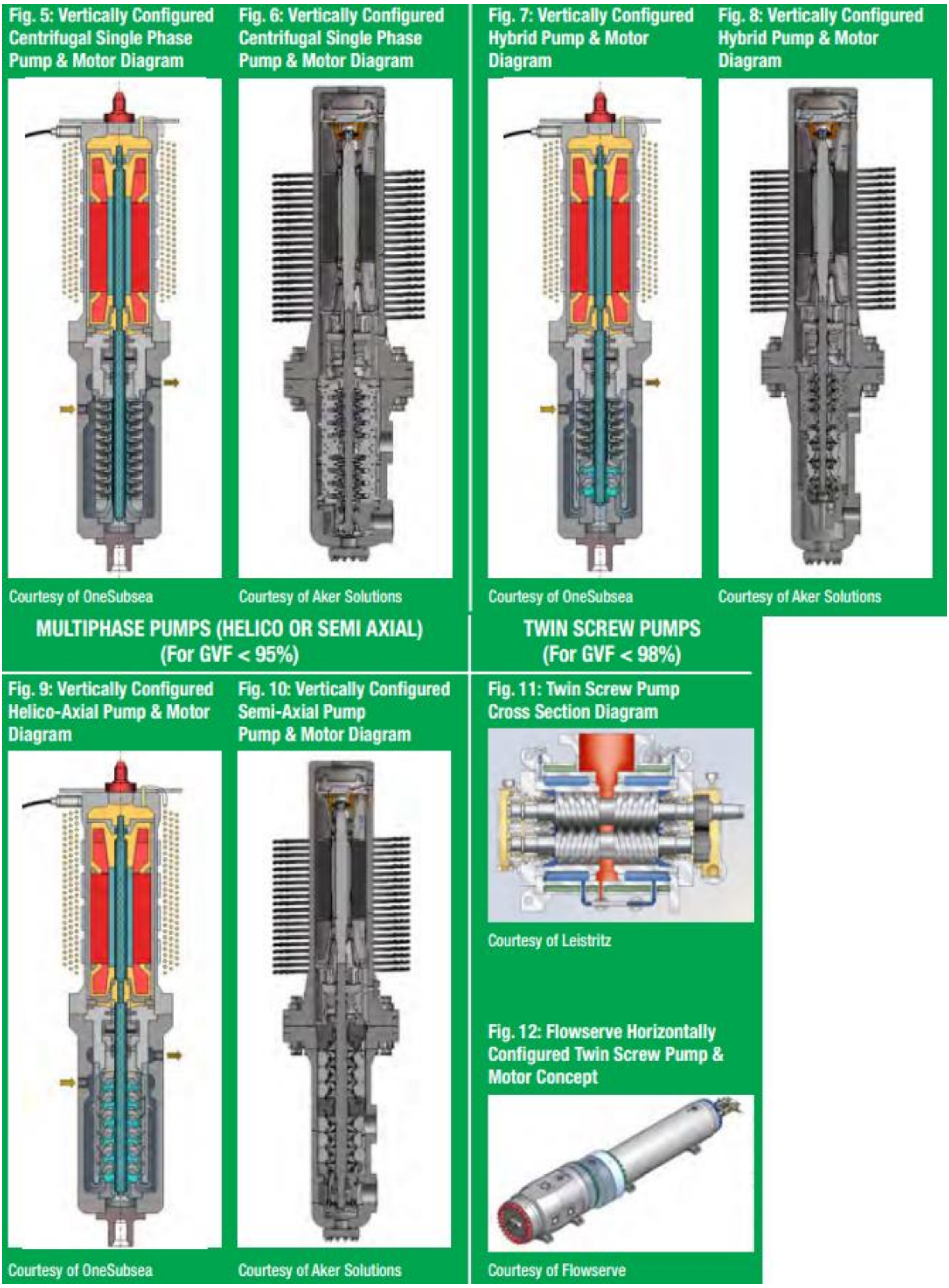


**Figure F.13 Subsea pump types, Status March 2016,**

(INTECSEA & Magazine, INTECSEA.com, 2016)


**Figure F.14 Boosting system examples (Conceptual and delivered), Status 2016** (INTECSEA & Magazine, INTECSEA.com, 2016)

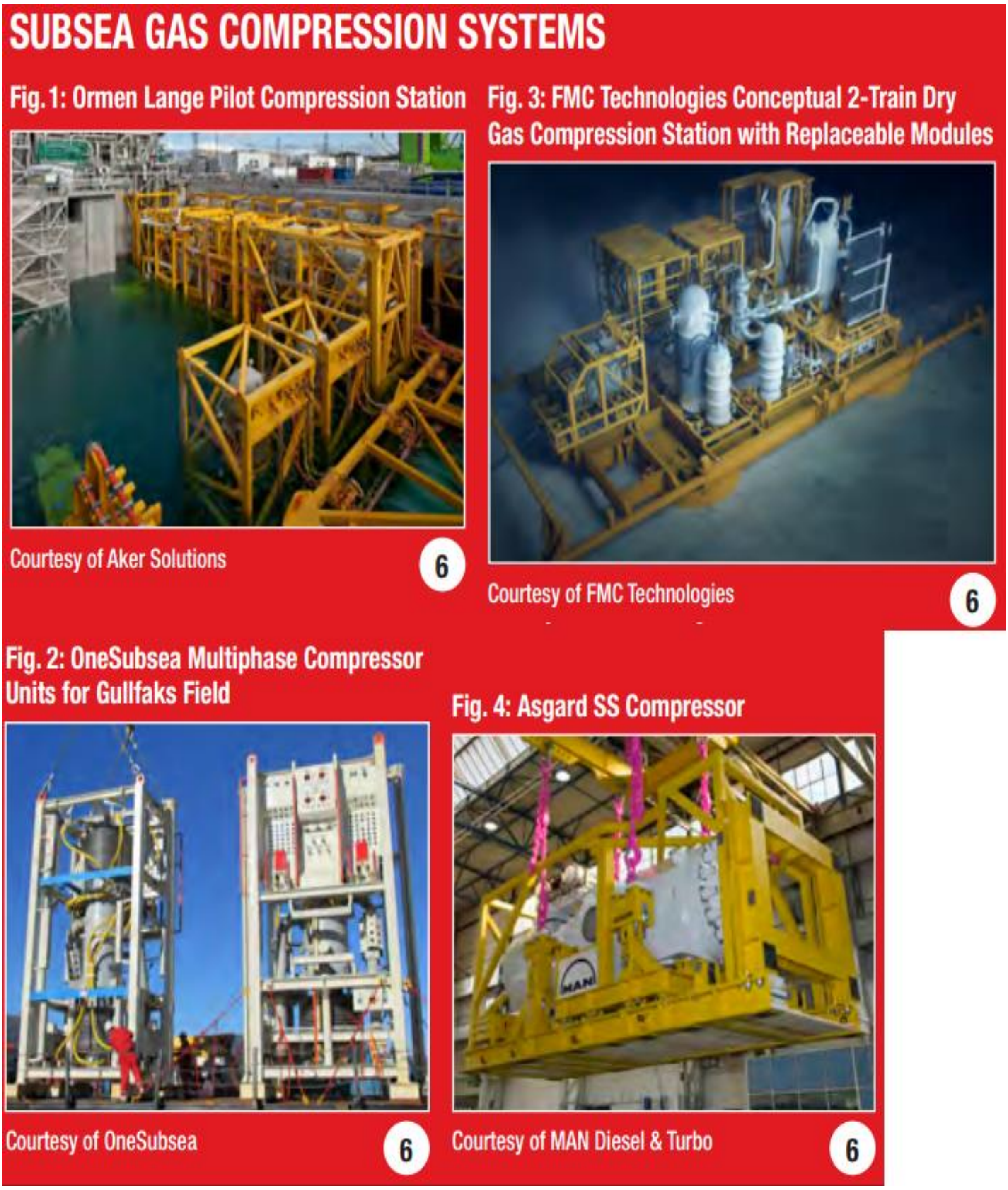


**Figure F.15 Subsea gas compression systems (1), Status March 2016** (INTECSEA & Magazine, INTECSEA.com, 2016)


**Figure F.16 Subsea gas compression systems (2), Status March 2016**

(INTECSEA & Magazine, INTECSEA.com, 2016)

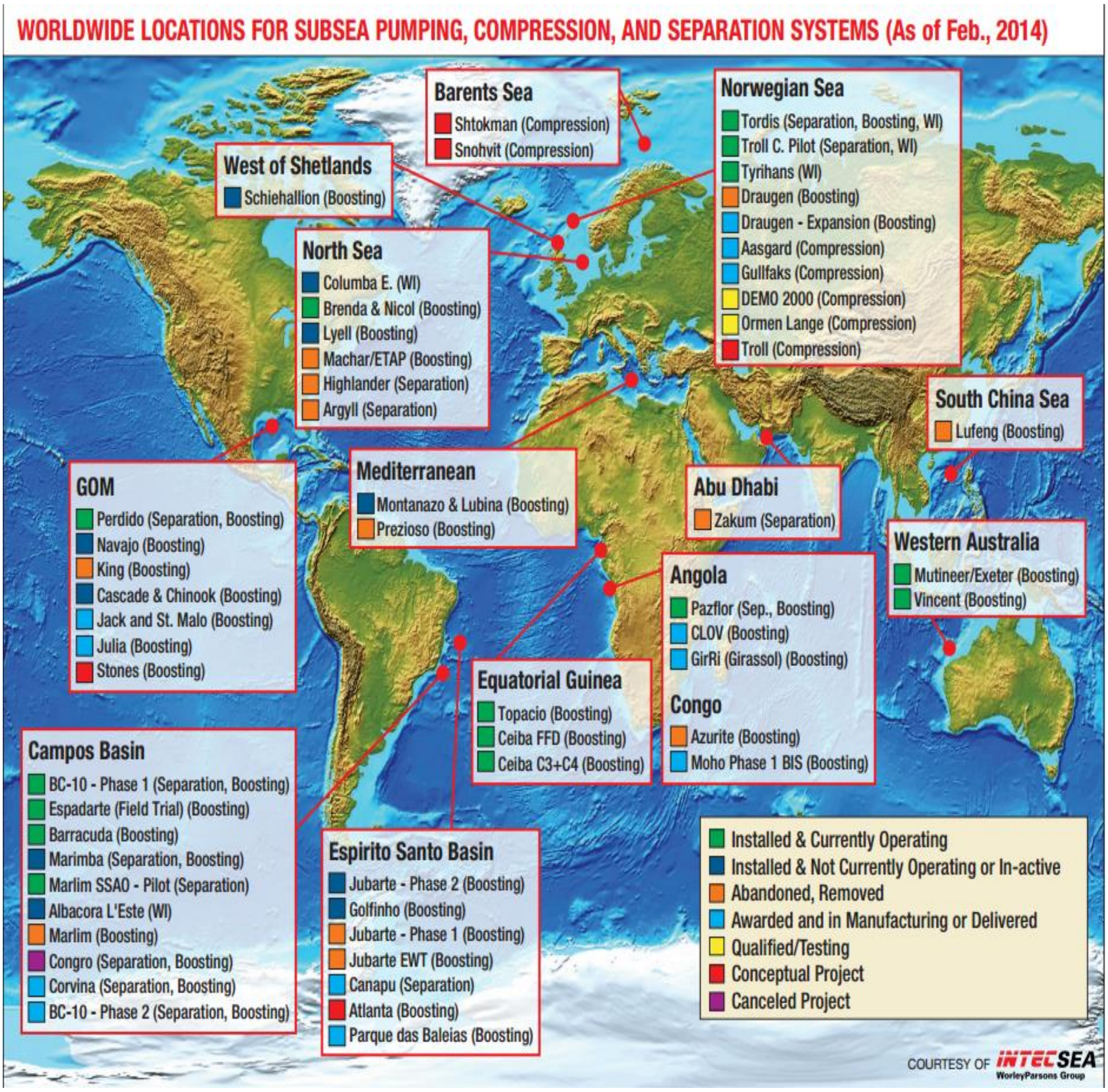


**Figure F.17 Worldwide locations for subsea pumping, compression, and separation system, Status February 2014** (INTECSEA, Intecsea.com, 2014)