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Design and evaluation of gas protraction for an offshore oil field

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Master of Science in Energy and Environment
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Problem Description

Background
Field A is a marginal oil field in the North Sea operated by Det norske. The company is currently working on the maturation of the field development. Three development options have been identified, either a Floating Production Storage and Offloading (FPSO) with a bridge link to a WellHead Platform (WHP), a Jack-Up with production facility (JUDP) and WHP. For the JUDP option, oil will be stored in a Floating Storage Unit (FSU). The final option is a subsea tie-back to a host facility.

Produced oil will be stabilized to export specification and offloaded to a shuttle tanker. The base case for gas export is to export wet gas to Field B, where the gas will be dehydrated and conditioned to meet the Vesterled or Statpipe specifications. There are uncertainties concerning the processing suitability at Field B, and Det norske is therefore investigating the possibilities for installing gas drying and conditioning equipment on the Field A platform.

Objective
The main objective is to build up knowledge and understanding of gas drying and conditioning processes on offshore production facilities. Building competence will be done by literature study and by developing simulation models for gas dehydration and conditioning processes.

The following questions should be considered in the project work:

1. Carry out a literature study about:
    a. Mapping of existing gas infrastructure in the Field A area.
    b. Mapping of processing methods to reach adequate dry gas quality.
2. Build up simulation models in HYSYS:
    a. Gas drying and conditioning at the Field A platform to reach export specifications at Vesterled and Statpipe.
    b. Simulating the Field B process (Vesterled/Statpipe)
3. Evaluate the different concepts with regards to energy efficiency, weight, cost and complexity, and come up with a recommendation for the Field A development project.

Assignment given: 01. February 2010
Supervisor: Truls Gundersen, EPT

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Abstract

Field A is a marginal oil field located in the North Sea, operated by Det norske oljeselskap ASA (hereby Det norske). Det norske are currently working on the maturation of the field development, and there are uncertainties concerning the gas protraction from the field. The base case is wet gas export to Field B, but dry gas export directly from Field A is also investigated.

A mapping of the gas infrastructure in the Field A area is first performed, revealing Statpipe and Vesterled as dry gas export pipelines.

A literature study concerning different methods to dehydrate and condition the gas is performed. The simulations later in the thesis are based on this study.

To get a comparison basis for the different development alternatives in terms of weight, costs and complexity, a simulation of a basic Field A process with no dehydration or conditioning is first simulated. The basic process is then expanded with different equipment for dehydration of the gas, and the different dehydration processes are briefly evaluated. The dehydration processes are then varied and/or expanded to achieve both adequate dehydration and hydrocarbon dew point in the gas, based on the export specifications in Statpipe and Vesterled.

The final processes, achieving both proper water- and hydrocarbon dew point in the export gas, are evaluated in terms of weight, costs and complexity, and compared to the basic process without dehydration and conditioning. Based on the findings in this thesis it seems dry gas export from Field A is difficult to achieve.

Based on the limited information available from the operating company of Field B, the Field B process is simulated in HYSYS to check the process suitability for the Field A gas.

Finally, an alternative solution to achieve dry gas export from Field A, by bleeding off propane in the process, is briefly discussed. The solution of rich gas export in either FUKA or Sage is also briefly discussed.

Norwegian University
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MASTER THESIS

for

Stud.techn. Øystein Lindland
Spring 2010

Design and evaluation of gas protraction for an offshore oil field

Design og evaluering av gassavsetning for et oljefelt offshore

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Department of Energy and Process Engineering, 5 February 2010

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Summary

Field A is a marginal oil field located in the North Sea, operated by Det norske oljeselskap ASA (hereby Det norske). Det norske are currently working on the maturation of the field development, and have identified three development options.

- Floating production, storage and offloading (FPSO), with a bridge link to a well head platform (WHP).
- Subsea tie-back to a host facility.
- Jack-up with production facility (JUDP) and WHP. Produced oil will be stored in a floating storage unit (FSU).

The produced oil will be stabilized to the export specifications and offloaded to a shuttle tanker. The base case for gas export is to export wet gas to a platform on another field, Field B, where it will be dehydrated and conditioned to meet the export specifications in one of the export pipelines connected to the platform. There are however uncertainties concerning the processing suitability at Field B, and Det norske is therefore investigating the possibilities for installing dehydration and conditioning equipment on the Field A platform. In this thesis, different methods for dehydration and conditioning of the Field A gas is tested through process simulations in HYSYS.

A mapping of the gas infrastructure in the Field A area reveals two dry gas export pipelines; Statpipe and Vesterled, and two rich gas export pipelines; FUKA and Sage. This thesis is mainly based on dry gas export, and therefore most of the work is put into export in Statpipe or Vesterled.

A literature study concerning different methods to dehydrate and condition the gas is performed. The simulations later in the thesis are based on this study.

To get a comparison basis for the different development alternatives in terms of weight, costs and complexity, a simulation of a basic Field A process with no dehydration or conditioning is first simulated. This will also be the Field A process if wet gas export to Field B is chosen as the solution for gas protraction at Field A.

The basic process is then expanded with different equipment for dehydration of the gas, and the different dehydration processes are briefly evaluated. The dehydration processes are then varied and/or expanded to achieve both adequate dehydration and hydrocarbon dew point in the gas, based on the export specifications in Statpipe and Vesterled.

The final processes, achieving both proper water- and hydrocarbon dew point in the export gas, are evaluated and compared to the basic process without dehydration and conditioning, in terms of weight, costs and complexity,. Based on the findings in this thesis it seems dry gas export from Field A is difficult to achieve.

Based on the limited information available from the operating company of Field B, the Field B process is simulated in HYSYS to check the process suitability for the Field A gas.

Finally, an alternative solution to achieve dry gas export from Field A, by bleeding off propane in the process, is briefly discussed. The solution of rich gas export in either FUKA or Sage is also briefly discussed.

Sammendrag

Felt A er et marginalt oljefelt I Nordsjøen, med Det norske oljeselskap ASA (heretter Det norske) som operatør. Det norske jobber for tiden med utviklingen av feltet, og har identifisert tre utbyggingsalternativer.

- Flytende produksjon, lagring og lossing, med brokobling til en brønnhodeplattform.
- Undervanns tie-back til en vertsplattform.
- Oppjekkbar plattform med produksjonsutstyr og brønnhodeplattform. Produsert olje vil bli lagret i en flytende lagringstank.

Den produserte oljen vil bli stabilisert i henhold til eksportspesifikasjonene og losset til en skytteltanker. Base case for gasseksport er våtgasseksport til en plattform på et annet felt, Felt B, hvor den vil bli tørket og behandlet for å møte eksportspesifikasjonene til en av rørledningene koblet til plattformen. Men det er usikkert om prosessen på Felt B er egnet for gassen fra Felt A, og Det norske jobber derfor med å kartlegge mulighetene for å installere utstyr til tørking og behandling av gassen på Felt A, slik at tørr gass kan eksporteres direkte. I denne oppgaven testes ulike metoder for tørking og behandling av gassen fra Felt A, gjennom prosessimuleringer i HYSYS.

En kartlegging av gassinfrastrukturen i området rundt Felt A viser to aktuelle tørrgassrørledninger; Statpipe og Vesterled, og to aktuelle rikgassrørledninger; FUKA og Sage. Denne oppgaven baserer seg hovedsakelig på tørrgasseksport, og derfor er Statpipe og Vesterled studert mest nøye.

En litteraturstudie angående ulike metoder for å tørke og behandle naturgass er gjennomført. Simuleringene senere i oppgaven er basert på dette studiet.

Først utvikles en simuleringsmodell for den grunnleggende prosessen på Felt A uten tørking og behandling av gassen. Denne modellen vil være sammenligningsgrunnlaget de senere modellene vil bli sammenlignet med, i forhold til vekt, kostnader og kompleksitet. Det vil også være prosessen på Felt A hvis våtgasseksport til Felt B blir det endelige utbyggingsalternativet for gassavsetningen fra Felt A.

Den grunnleggende prosessen blir så utvidet med ulikt utstyr for tørking av gassen, og de ulike tørkeprosessene blir kort evaluert. Deretter blir de ulike tørkeprosessene variert og/eller utvidet, for å oppnå både tilfredsstillende vann- og hydrokarbonduggpunkt i eksportgassen, basert på spesifikasjonene i Statpipe og Vesterled.

De prosessene som oppnår både tilfredsstillende vann- og hydrokarbonduggpunkt, evalueres og blir sammenlignet med den grunnleggende prosessen uten gasstørking og gassbehandling, i forhold til vekt, kostnader og kompleksitet,. Basert på funnene i denne oppgaven ser det ut til at tørrgasseksport fra Felt A er vanskelig å gjennomføre.

Basert på den begrensede tilgjengelige informasjonen fra operatøren av Felt B, simuleres en modell av prosessen på Felt B, for å sjekke om den er egnet for gassen fra Felt A.

Til slutt foreslås en alternativ løsning for å oppnå tørrgasseksport fra Felt A, ved å ta ut en del propan i prosessen. Mulighetene for rikgasseksport i FUKA eller Sage diskuteres også kort.

Table of contents

List of tables

List of figures

1. Introduction

Field A is a marginal oil field located in the North Sea, operated by Det norske oljeselskap ASA (hereby Det norske). Det norske are currently working on the maturation of the field development, and have identified three development options.

- Floating production, storage and offloading (FPSO), with a bridge link to a well head platform (WHP).
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The produced oil will be stabilized to the export specifications and offloaded to a shuttle tanker. The base case for gas export is to export wet gas to a platform on another field, Field B, where it will be dehydrated and conditioned to meet the export specifications in one of the export pipelines connected to the platform. There are however uncertainties concerning the processing suitability at Field B, and Det norske is therefore investigating the possibilities for installing dehydration and conditioning equipment on the Field A platform. In this thesis, different methods for dehydration and conditioning of the Field A gas is tested through process simulations in HYSYS.

First gas infrastructure in the Field A area is mapped, to get an idea of the gas export opportunities from the field, and the gas specifications the gas needs to fulfill. Then a literature study regarding different methods to dehydrate and condition the gas is performed.

A simulation model of the basic Field A process without any dehydration and conditioning is developed. This will be the process at Field A if wet gas export to Field B is chosen as the development solution. It will also work as a basis to which the different processes with dehydration and conditioning will be compared in terms of weight, costs and complexity.

The basic process is expanded with different equipment for dehydration and conditioning, and simulation models dehydrating and conditioning the gas to reach the export specifications of the pipelines in the area are developed. These models are compared to the basic process in terms of weight, costs and complexity.

Finally the Field B process is simulated, to check the process' suitability for the Field A gas. Some alternative developments for the gas protraction at Field A are also suggested.

1.1 Simulation software and basis

During the work with this thesis, stationary simulations have been implemented in the software *Aspentech HYSYS 2006.5,* hereafter named HYSYS. Peng-Robinson[1] has been used as the equation of state. Standard components have been used directly as they are predefined in HYSYS. Other groups of components have been manually implemented in HYSYS with the physical properties listed in Table 2-2. The other necessary properties for these component groups have been assumed by HYSYS based on the given ones. Sea water is not a selectable component, and the detailed properties of the formation water are not known, so for both of these materials regular water has been chosen.

1.2 Naming of material streams and process equipment

The different streams and equipment have been named following NORSOK Standard P-100. A detailed description of the naming procedure can be found in Appendix B.

[1] Appendix A

2. Field A

Field A is a marginal oil field located in the North Sea, operated by Det norske. Det norske are currently working on the maturation of the field development.

2.1 Design basis

Det norske has established a Design basis for the Field A development. The main parts of this Design basis are listed in the following sub chapters.

2.1.1 Composition of production fluids

Table 2-1 shows the composition of the production fluid.

Table 2-1: Composition of production fluid [2]

Component	Mole%
N_2	1,17
CO_2	2,07
C1	45,08
C2	8,12
C3	7,96
i-C4	1,33
n-C4	3,52
i-C5	1,39
n-C5	1,58
C6	3,48
C7	3,32
C8	2,27
C9	1,92
C10	0,84
C11+	15,95

When using Table 2-1 as a basis for process simulations, the hydrocarbon components lighter than C5 can be used directly as predefined in the simulation software. The hydrocarbon components heavier than C4 have the properties listed in Table 2-2 below.

Table 2-2: Properties of heavier hydrocarbons [2]

Component	Mole%	Mole weight [kg/kmole]	Density [kg/m^3]	Boiling temperature [C]	Critical temperature [C]	Critical pressure [bar]	Acentricity
i-C5	1,39	72,15	625	27,8	187,3	33,8	0,2286
n-C5	1,58	72,15	631	36,1	196,5	33,7	0,2524
C6	3,48	86,18	664	68,7	234,3	30,1	0,2998
C7	3,32	100,20	688	98,4	267,1	27,4	0,3494
C8	2,27	114,23	707	125,7	295,7	24,9	0,3981
C9	1,92	128,26	722	150,8	321,5	22,9	0,4452
C10	0,84	142,30	734	174,2	344,5	21,0	0,4904
C11+	15,95	262,00	884	340,0	514,6	12,8	0,5500

These components have to be implemented as hypothetical components in the simulation software, with the properties listed in the table. The component C11+ represents a group of all the hydrocarbons heavier than C10.

Figure 2-1 shows the phase envelope for the production fluids.

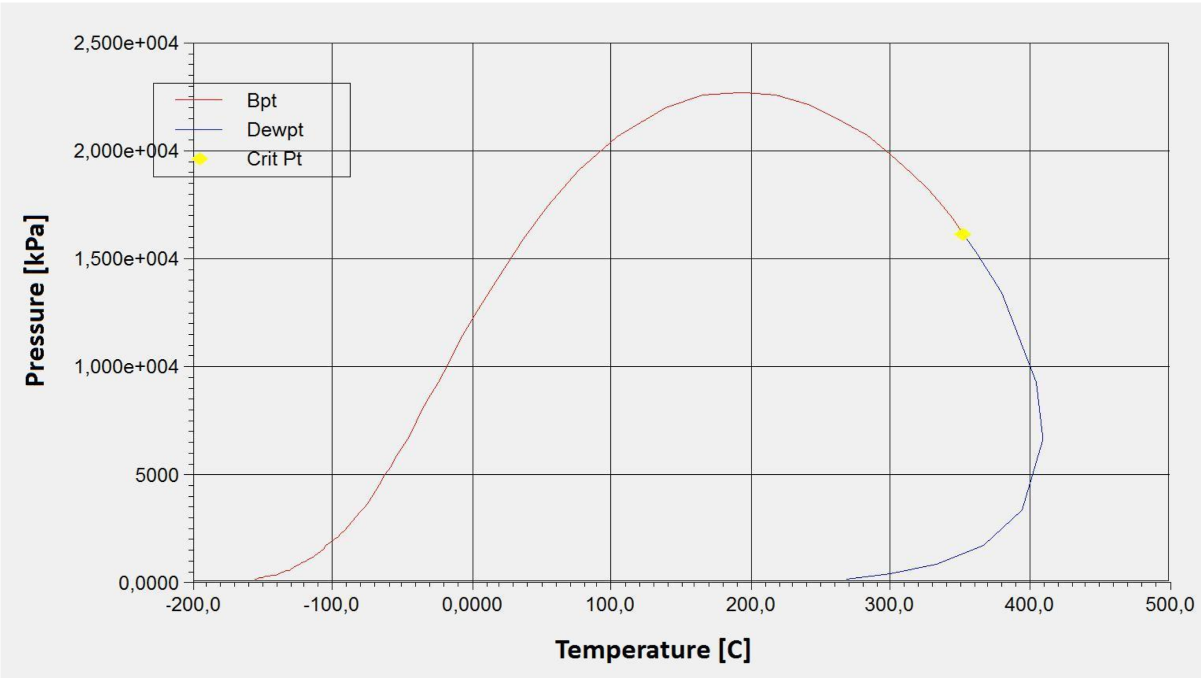

Figure 2-1: Phase envelope for production fluids

The static bottom-hole pressure and temperature are 321 bara and 105˚C respectively [2].

2.1.2 Design capacities

The design capacities for Field A have been identified by Det Norske and are listed in Table 2-3.

Table 2-3: Field A design capacities [2]

Fluid	Capacity
Total liquid rate [Sm^3/d]	18000
Oil rate [Sm^3/d]	7000
Gas export incl. Gas lift [MSm^3/d]	1,5
Water injection rate [Sm^3/d]	20000
Storage capacity [Sm^3]	50000

2.1.3 Process specifications

The most important process specifications are listed in the sub chapters below.

2.1.3.1 Properties of stable crude oil

The processed crude oil should be stabilized so that TVP < 0,965 bar at 37,8˚C [2].

2.1.3.2 Gas export and gas lift

The base case for gas export is wet export to Field B. The export pressure shall be 125 bara [2].

The gas lift manifold shall have capacity to supply gas to 10 production wells. The gas lift rate shall be 400000 Sm^3/d (maximum 150000 Sm^3/d per well), and the gas lift pressure shall be 170 bara [2].

2.1.3.3 Water injection

The produced water shall be supplemented by deoxygenated seawater to make up the total injection water requirement. The maximum design injection rate shall be 13000 Sm^3/d [2].

3. Gas infrastructure in the Field A area

The oil produced from Field A will be stabilized to export specifications and offloaded to a shuttle tanker. The base case for the produced gas is wet gas export to Field B, where the gas will be dehydrated and conditioned to meet the Statpipe or Vesterled export specifications [1], [2]. But there are uncertainties concerning the process suitability at Field B, and therefore Det norske are investigating the possibility to install gas dehydration and conditioning equipment at the Field A production platform. This way the gas can be conditioned to meet the specifications of gas pipelines in the Field A area, and then exported directly into this infrastructure.

3.1 Export pipelines in the area

There are mainly two export pipelines in the area being considered for export of the produced gas, Statpipe and Vesterled. They are both already connected to Field B, so if the wet gas is exported there for dehydration and conditioning, it can be exported in either one of them. If the produced gas is to be exported directly from Field A, it has to be conditioned to meet the export specifications in the chosen pipeline. Field A can be connected to either Statpipe or Vesterled, but the export specifications are different in the two pipelines. Before a pipeline is chosen, the properties of the produced gas have to be evaluated, and also the complexity of the conditioning and dehydration methods needed to reach the different specifications [1], [3].

If wet gas export to Field B, or dry gas export from Field A in either Statpipe or Vesterled are all found to be difficult, a third export alternative is rich gas export. There are two rich gas pipelines in the area, the Frigg UK pipeline (FUKA) and Sage, both in the British sector [1], [4], [5].

3.1.1 Statpipe

Statpipe links northern North Sea gas fields with Norway's gas export system. It transports gas from Statfjord, Gullfaks, Veslefrikk, Snorre, Brage, Tordis and Field B. It is connected to the Kårstø onshore processing facility on the west coast of Norway, and is also directly connected to Norpipe. Norpipe transports natural gas from Ekofisk to Emden in Germany [3].

The most important export specifications the produced gas from Field A will have to meet if it is to be transported in Statpipe are listed in Table 3-1 below.

Table 3-1: Export gas specifications for Statpipe [3]

Specification [unit]	Value
Maximum operating pressure [barg]	151,8
Maximum operating temperature [˚C]	50,0
Minimum operating temperature [˚C]	-10,0
Hydrocarbon dew point [˚C at 50 barg]	< -10,0
Water dew point [˚C at 69 barg]	-18,0
CO_2 [mole %]	2,5
Gross calorific value [MJ/Sm^3]	38,1-43,7
Wobbe index [MJ/Sm^3]	48,3-52,8

3.1.2 Vesterled

Vesterled runs from Field B in the North Sea to St. Fergus in Scotland. The most important export specifications the produced gas from Field A will have to meet if it is to be transported in Vesterled are listed in Table 3-2 below [3].

Table 3-2: Export gas specification for Vesterled [3]

Specification [unit]	Value
Minimum contractual pressure [barg]	41,0
Maximum operating temperature [˚C]	N/A
Minimum operating temperature [˚C]	1,0
Hydrocarbon dew point [˚C at 50 barg]	< -3,0
Water dew point [˚C at 69 barg]	-12,0
CO_2 [mole %]	2,5
Gross calorific value [MJ/Sm^3]	38,1-43,7
Wobbe index [MJ/Sm^3]	48,3-52,8

3.1.3 FUKA

The rich gas export phase of FUKA runs from the Alwyn Area to St. Fergus. The most important specifications for gas entering FUKA are **cricondenbar below 106 bara** and **maximum 24 kg water/MSm^3** [4].

3.1.4 Sage

The Sage pipeline runs from Beryl A to St. Fergus. The most important specifications for export gas in Sage are **water volume of less than 63 ppm** and **10,67-21,82 mole% of C2-C12 components** [5].

4. Processing methods to dehydrate natural gas

Export pipelines for natural gas have different specifications. Among others, the water content of the gas cannot be too high, as this increases the risk for hydrate formation. If the water content of the produced gas is higher than the pipeline specification, the gas has to be dehydrated. It exist different methods that can be used for the gas to reach adequate quality. They are dehydration by cooling/depressurizing, dehydration by absorption and dehydration by adsorption [7].

The most common method to dehydrate natural gas in the industry is dehydration by absorption. The water content specification is seldom so low that adsorption needs to be utilized, and cooling/depressurizing processes often need to be utilized to control the hydrocarbon dew point of the gas (more on this in Chapter 5). Still, all the dehydration methods have been examined in this thesis [8].

4.1 Dehydration by cooling/depressurizing

Under natural conditions, natural gas is normally saturated with water vapor. The amount of water vapor the natural gas can be saturated with increases both with increasing temperature and increasing pressure. Thus, one way to dehydrate natural gas is to lower its temperature and/or pressure before separation. The reduced temperature and/or pressure will cause free water to fall out from the gas, which can be separated out in a gravitational separator. Figure 4-1 shows a schematic representation of the principle behind dehydration by cooling/depressurizing [7].

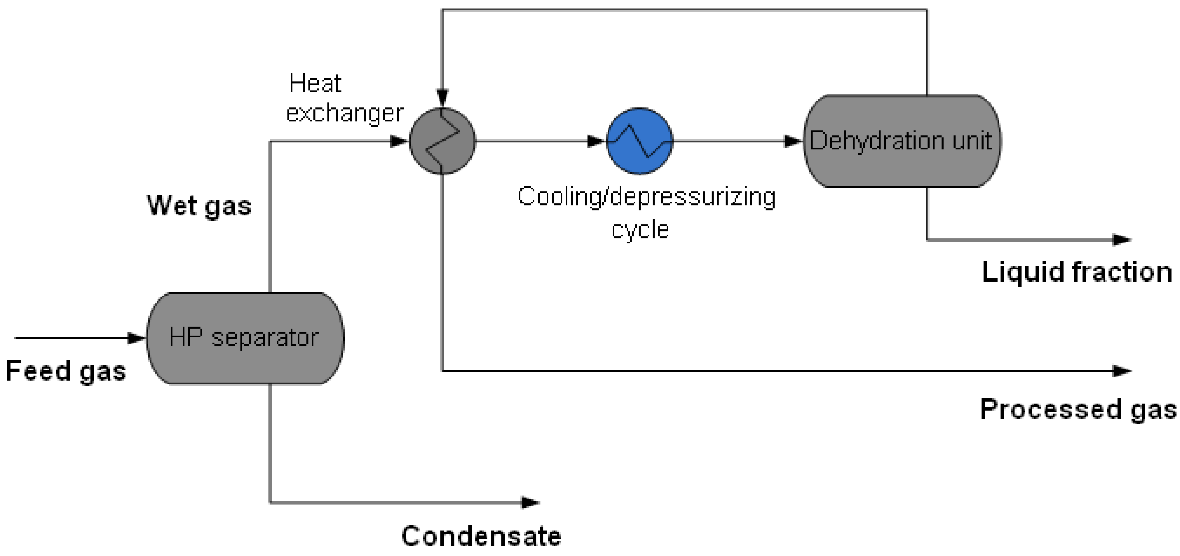

Figure 4-1: Dehydration by cooling/depressurizing

The dehydrated gas can be used to pre-cool the wet gas, as seen in the figure. The method used in the cooling/depressurizing cycle can vary. There are mainly three methods used, these are refrigerant cycle, turbo expander process and Joule-Thomson valve process [7], [9].

Dehydration by cooling/depressurizing is the simplest method to dehydrate natural gas, but it can normally not be used to reach demands for extremely low water content [6].

4.1.1 *Refrigerant cycle*

In a refrigerant cycle, the temperature of the gas is lowered by heat exchanging with a refrigerant. Figure 4-2 shows a schematic representation of a typical refrigerant cycle, which will replace the blue cooling/depressurizing cycle in Figure 4-1 if it is implemented in the process [6].

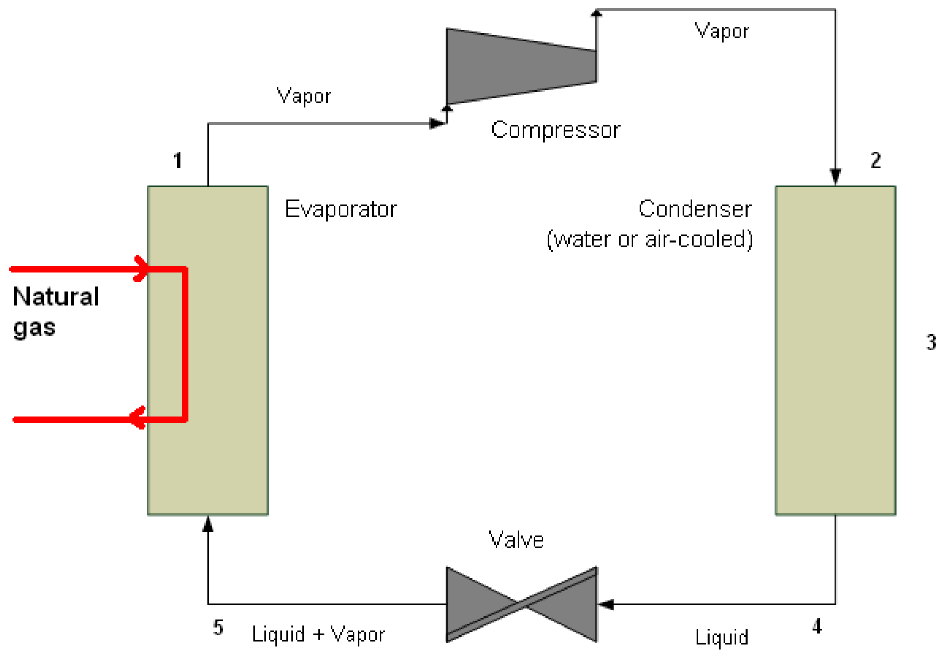

Figure 4-2: Refrigeration cycle

Refrigerant vapor is compressed before being condensed by heat exchanging with air or water. The pressure of the liquid refrigerant is then let down through a valve, so that parts of the refrigerant evaporate. The expansion also causes the temperature of the refrigerant to fall. It can then be heat exchanged with the natural gas initially in need of cooling, causing the rest of it to evaporate. The cycle is then repeated. Figure 4-3 shows the path of the refrigerant in a TS-diagram [6].

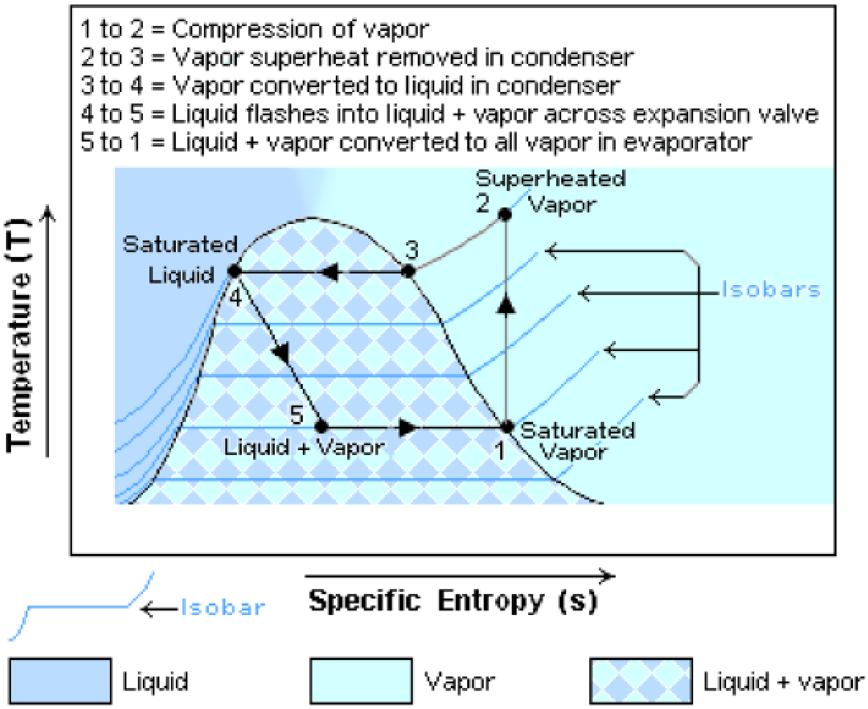

Figure 4-3: Path of refrigerant in TS-diagram[2]

After the natural gas has been cooled in the refrigerant cycle, it can be dehydrated in a separator as shown in Figure 4-1.

Different components can be used as the refrigerant. Measures should be taken to select the right one. The ideal refrigerant is nontoxic, noncorrosive and has physical properties compatible with the system's needs (vaporizes and condenses at temperatures and pressures achievable in the system). It should also have a high latent heat of vaporization. For cooling above -40°C, propane, ammonia and R-22 are common refrigerants. For cryogenic cooling ethylene, nitrogen and methane might be used [6].

Some operational problems might occur when using a refrigerant cycle. The most common ones are loss of refrigerant, contamination of refrigerant (change of properties) and fouling and/or degrading of heat transfer surfaces [6].

4.1.2 Turbo expander process

In a turbo expander process, the pressure and temperature of the gas is lowered through a turbo expander. The blue cooling/depressurizing cycle in Figure 4-1 is then a turbo expander. After the

[2] Picture from *http://www.ipt.ntnu.no/~jsg/undervisning/naturgass/lysark/LysarkFoerde2008.pdf*

turbo expander the gas can be dehydrated in a separator as shown in Figure 4-1. Just like in the refrigerant cycle, the dehydrated gas can be used to pre-cool the feed gas.

The main advantages using a turbo expander process is that work can be extracted from the expanding high pressure gas. The process is isentropic. Today turbo expander isentropic efficiencies are approaching 85 %. The work extracted can be used to compress the export gas. Figure 4-4 shows a schematic representation of the turbo expander dehydration process [6].

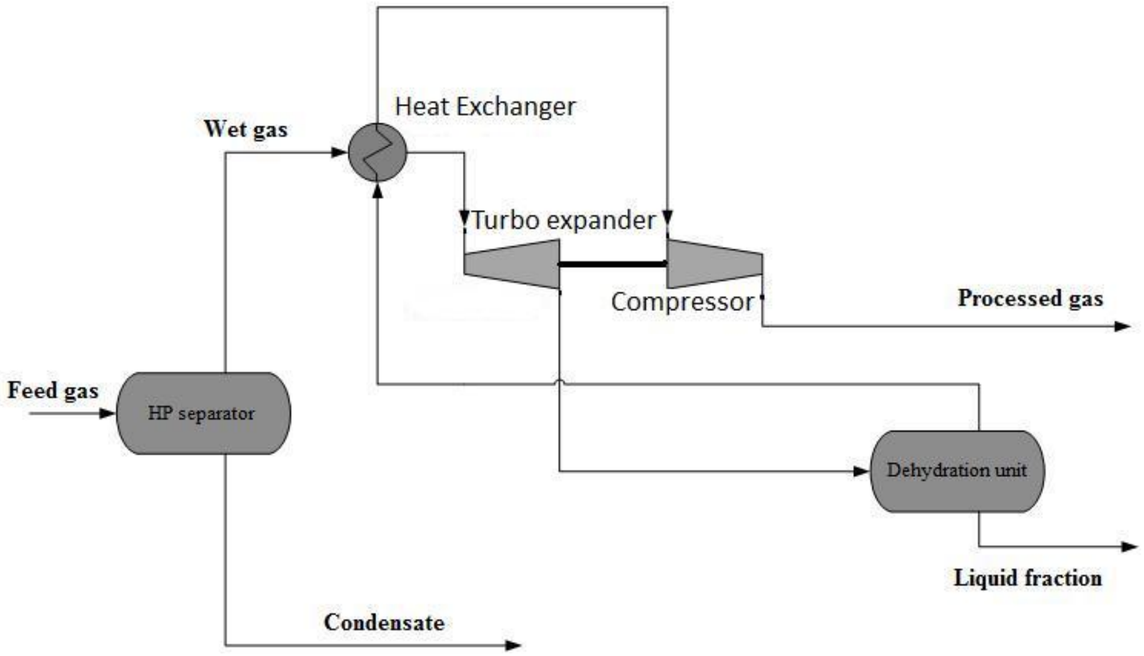

Figure 4-4: Turbo expander dehydration process

Often a Joule-Thomson valve is installed in parallel with the turbo expander. This one can be used during start up and during maintenance of the expander [6].

4.1.3 Joule-Thomson valve process

In a Joule-Thomson valve process the gas is depressurized and cooled through a Joule-Thomson valve. The blue cooling/depressurizing cycle in Figure 4-1 is then a Joule-Thomson valve. Figure 4-5 shows a schematic representation of the process.

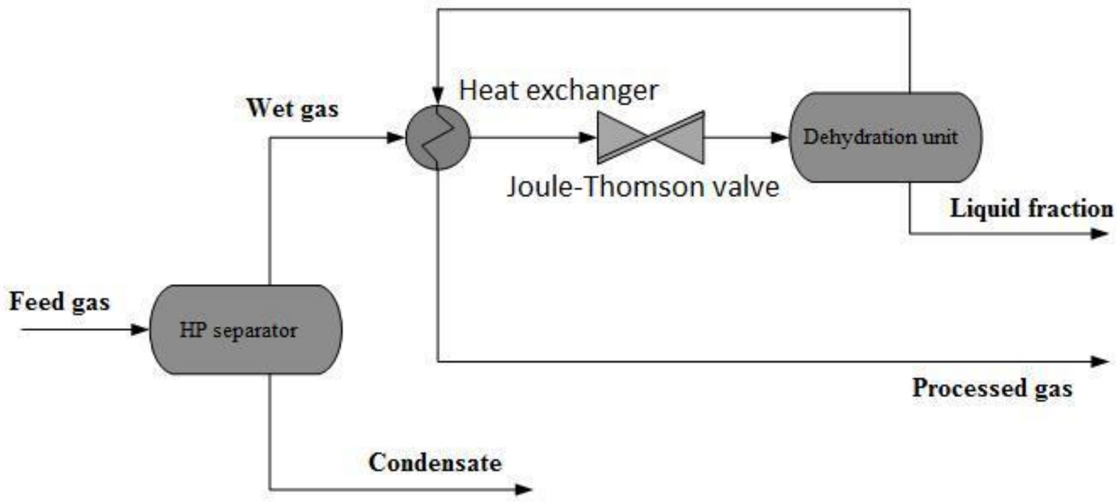

Figure 4-5: Joule-Thomson valve process

The separated gas can be used to pre-cool the feed gas also in this process.

It is both cheaper and easier to install a Joule-Thomson valve then a turbo expander, but the process is less efficient, as no work can be extracted from the high pressure gas [6].

4.2 Dehydration by absorption

If the natural gas contains very large amounts of water vapor it is often better to dehydrate it by absorption, which is more efficient for removing large volumes of water. The absorption process is performed in a counter current scrubbing unit, where the gas is scrubbed by an absorbent with strong affinity for water [6].

The ideal absorbent should have [6]:

- Strong affinity for water
- Low cost
- Non corrosive
- Low affinity for hydrocarbons and acid gases
- Thermal stability
- Easy regeneration
- Low viscosity
- Low vapor pressure at the contact temperature

- Low tendency to foam

Glycol is the most common absorbent, either as diethylene glycol (DEG) or triethylene glycol (TEG) [6].

4.2.1 Glycol dehydration process

Figure 4-6 shows a simplified flow diagram for a glycol dehydration process.

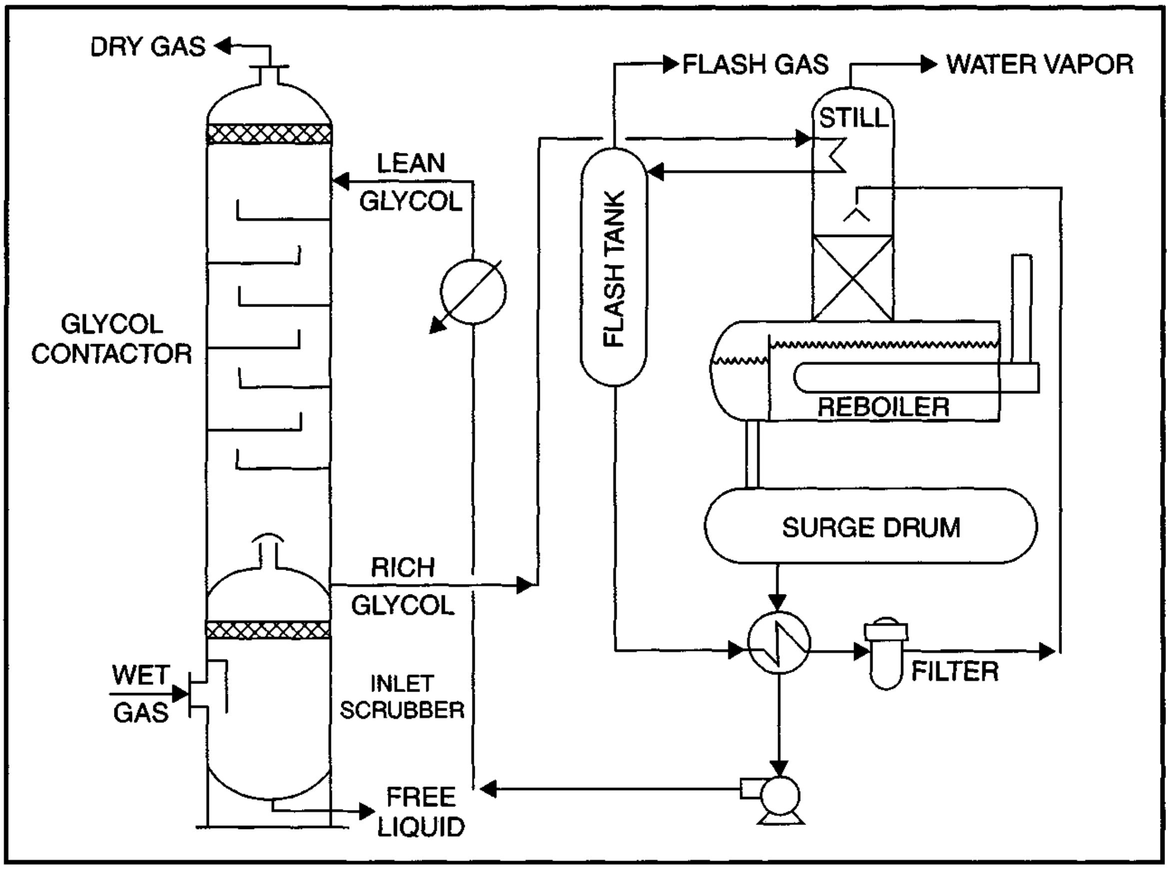

Figure 4-6: Glycol dehydration unit[3]

Wet gas enters at the bottom of the absorption column, where a mesh pad is used to remove any small amount of free liquids. It then enters the absorption zone, where plates or structured packing are installed to make a high contact area. Here it comes in contact with lean glycol (low water content) being fed near the top of the column. The lean glycol absorbs water from the gas, so that dry gas leaves the column at the top. Near the bottom of the column the rich glycol (high water

[3] Picture from GPSA engineering data book, 11th edition

content) leaves and is sent to the regeneration process. The glycol is regenerated by boiling and is then pumped back to the absorption column [6].

When designing an absorption process one has to consider the following [6]:

- The flow rate of glycol has to be high enough, so that the wanted amount of water is removed. The more glycol circulates, the more expensive the pump system will be.
- The diameter of the column has to be big enough to handle the gas rate
- The plates or structured packing has to provide enough equilibrium stages
- The lean glycol has to be lean enough. Leaner glycol means a more expensive regeneration system

4.2.2 Glycol regeneration

The glycol regeneration system contains different equipment [6].

Flash tank: Used to remove light hydrocarbons, CO_2 and H_2S. Operates at a lower pressure then the absorption column.

Filters: Removes solid particles and chemical impurities. Will result in a pressure drop which has to be compensated.

Re-boiler: Used to remove the water from the rich glycol. Due to degeneration of the glycol, the temperature should not exceed 204˚C when TEG is used.

Stripping column (named STILL in Figure 4-6): Stripping gas lowers the partial pressure of H_2O in the gas phase, and more water can be absorbed by the gas. The column is usually trayed or structural packed.

Surge drum: Should be able to hold all the re-boiler glycol, to allow inspection and repair of the re-boiler heating coil.

4.2.3 Advantages and disadvantages using glycol absorption

Table 4-1 shows advantages and disadvantages when using dehydration by glycol absorption.

Table 4-1: Advantages and disadvantages using glycol absorption [6]

Advantages	Disadvantages
Low initial cost	Fouling and polluting particles may contaminate the glycol solution
Low pressure drop across column	Overheating may degenerate the glycol, and create decomposition products
Easy to recharge the columns	When both oxygen and hydrogen sulfide is present, corrosion may occur
	Inadequate separation of the inlet gas will result in liquids in the column, which will degenerate it
	Foaming may occur, resulting in carry-over liquid. Often a small amount of an anti-foam compound is added, to limit this problem

4.3 Dehydration by adsorption

If the natural gas contains small volumes of water, but the process or export specifications demands extremely low water content, i.e. in LNG processes, the gas can be dehydrated by adsorption. Adsorption describes any process where gas molecules are held on the surface of a solid by surface forces [6], [7].

The most commonly used categories of sorbents are [6]:

- **Gel.** A granular amorphous solid (silica gel (SiO_2), alumina gel (Al_2O_3))
- **Alumina.** Hydrated form of alumina oxide (Al_2O_3) activated by drying off part of the adsorbed water on the surface.
- **Molecular sieves.** Alkali metal crystalline alumina silicates, very similar to natural clays.

Figure 4-7 shows a schematic representation of the dehydration by adsorption process.

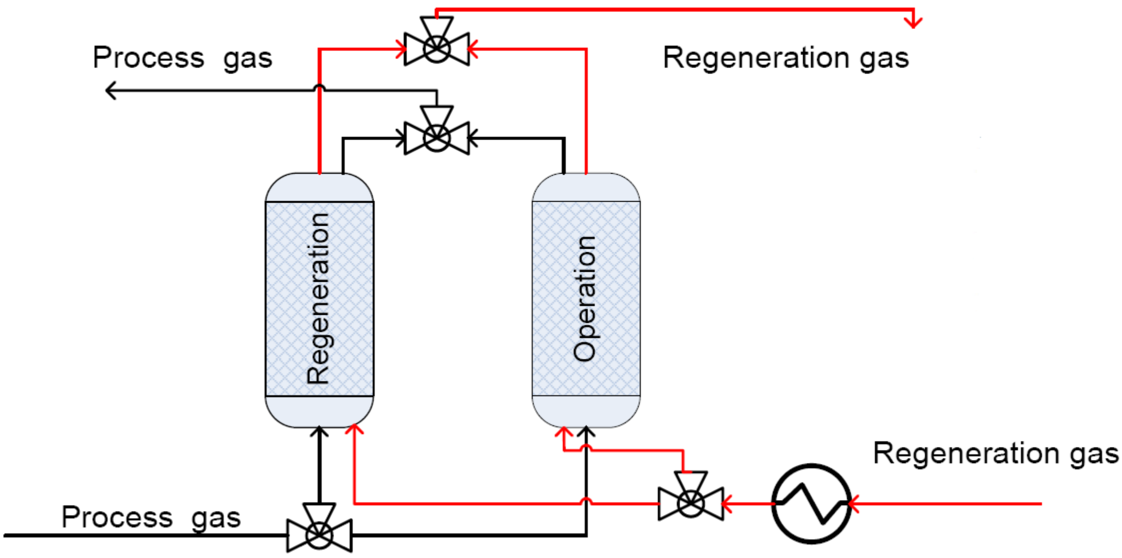

Figure 4-7: Flow diagram for dehydration by adsorption[4]

The unit consists of two columns, one is used for operation and one is regenerated. When the operating one is saturated with water, and the other one is fully regenerated, they switch [6].

The natural gas on Field A contains too much water for adsorption to be used as the dehydration method. Also, the water content export specification is not so low that adsorption will have to be used. Therefore adsorption will not be discussed any further in this thesis [8], [9].

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[4] Pivture from *http://www.ipt.ntnu.no/~jsg/undervisning/naturgass/lysark/LysarkFoerde2008.pdf*

5. Conditioning and dew point control of natural gas

As mentioned in Chapter 4, the gas has to be dehydrated to meet the water content specifications in the export pipeline. The pipelines will also have hydrocarbon dew point specifications. This is to control the heating value of the gas, often demanded by the customers buying the end product, and to avoid a liquid phase occurring during transport, as the gas export pipelines are not designed to handle multiphase flow. As for dehydration of gas, there exist different methods to remove intermediate hydrocarbons (condensate) from the gas, also called conditioning or dew point control of the gas [9].

5.1 Conditioning by cooling/depressurizing

As for dehydration, the natural gas can also be conditioned by cooling and/or depressurizing. In fact, the dehydration and the conditioning may occur simultaneously in the same unit operation. The different hydrocarbon components have different dew points/boiling points, depending on pressure and temperature. The heavier the hydrocarbon component is, the lower the dew point temperature and/or pressure is [8], [9].

Figure 4-1 shows a schematic representation of dehydration by cooling/depressurizing. This is the same process used for conditioning by cooling/depressurizing. The heavier hydrocarbon components will be separated from the gas in the HP separator, and the intermediate hydrocarbon components will be separated from the gas in the dehydration unit, together with the water. One can control which components that will be separated out to a certain degree, by varying the pressure and temperature in the dehydration unit.

The cooling/depressurizing unit used in the process can vary; mainly a Joule-Thomson valve process, a turbo expander process or a refrigerant cycle is used.

5.1.1 Joule-Thomson valve process.

The process is explained in Chapter 4.1.3. A schematic representation of the process can be found in Figure 4-5. The gas is dehydrated and conditioned simultaneously. As the pressure of the gas is lowered over the JT-valve, some of the heavier and intermediate hydrocarbon components will condense, and they can be separated out in the dehydration unit. The pressure drop will also cause the temperature of the gas to decrease.

5.1.2 Turbo expander process

The process is explained in Chapter 4.1.2. A schematic representation of the process can be found in Figure 4-4. The gas is dehydrated and conditioned simultaneously. As the pressure of the gas drops over the turbo expander, some of the heavier and intermediate hydrocarbon components will condense, and they can be separated out in the dehydration unit. The pressure drop will also cause the temperature of the gas to decrease, and as work can be extracted from the turbo expander, the temperature drop will be higher than it would have been for the same pressure drop in a JT-valve [6]. This will make the conditioning more efficient, as slightly lighter hydrocarbon components will condense due to the lower temperature. The work extracted can also be used elsewhere in the process.

5.1.3 Refrigerant cycle

The process is explained in Chapter 4.1.1. A schematic representation of the process is shown in Figure 4-2. The gas is dehydrated and conditioned simultaneously. As the gas is cooled through the refrigerant cooler, the heavier and intermediate hydrocarbon components condense, and can be separated in the dehydration unit. The more the gas is cooled, the lighter the components separated out will be.

5.2 Conditioning in a distillation column

The gas can also be conditioned in a distillation column. In a distillation column it is possible to be very exact in terms of the product specifications, but it is an energy demanding process. Also, a distillation column needs a large area, and has high investment and operational costs [6].

For the conditioning of the gas at Field A, a distillation column is not an alternative, so the process won't be discussed in detail in this thesis [8], [9].

6. The Field A separation and stabilization process

This chapter contains a description of the basic separation and stabilization process at Field A, without any gas dehydration and conditioning. If the produced gas is to be exported to Field B for dehydration and conditioning, the process described in this chapter will be the process used at Field A to separate the gas, oil and water.

If the gas is to be dehydrated and conditioned at Field A, the separation process in this chapter needs to be expanded, but it will still work as a base case.

The HYSYS simulation file used in Chapter 6 is named *FieldA_BasicProcess.*

6.1 General description

The fluids from the production well are depressurized, and enter the process at the 1st stage separator pressure. The produced oil is separated and stabilized in two stages, before an electrostatic coalescer removes the last residues of water. The oil is then pumped and cooled to the storage pressure and temperature.

The produced water is separated from the oil in the separators and the coalescer, and is pumped back into the reservoir together with seawater [2].

The produced gas is taken from the separators to the gas compression train, where it is compressed over two stages to the export pressure of 125 bara. Some of the gas is taken off as lift gas, and is compressed to the gas lift pressure of 170 bara [2].

Figure 6-1 shows a simplified schematic representation of the process. A detailed process flow diagram from HYSYS can be found in Appendix E.

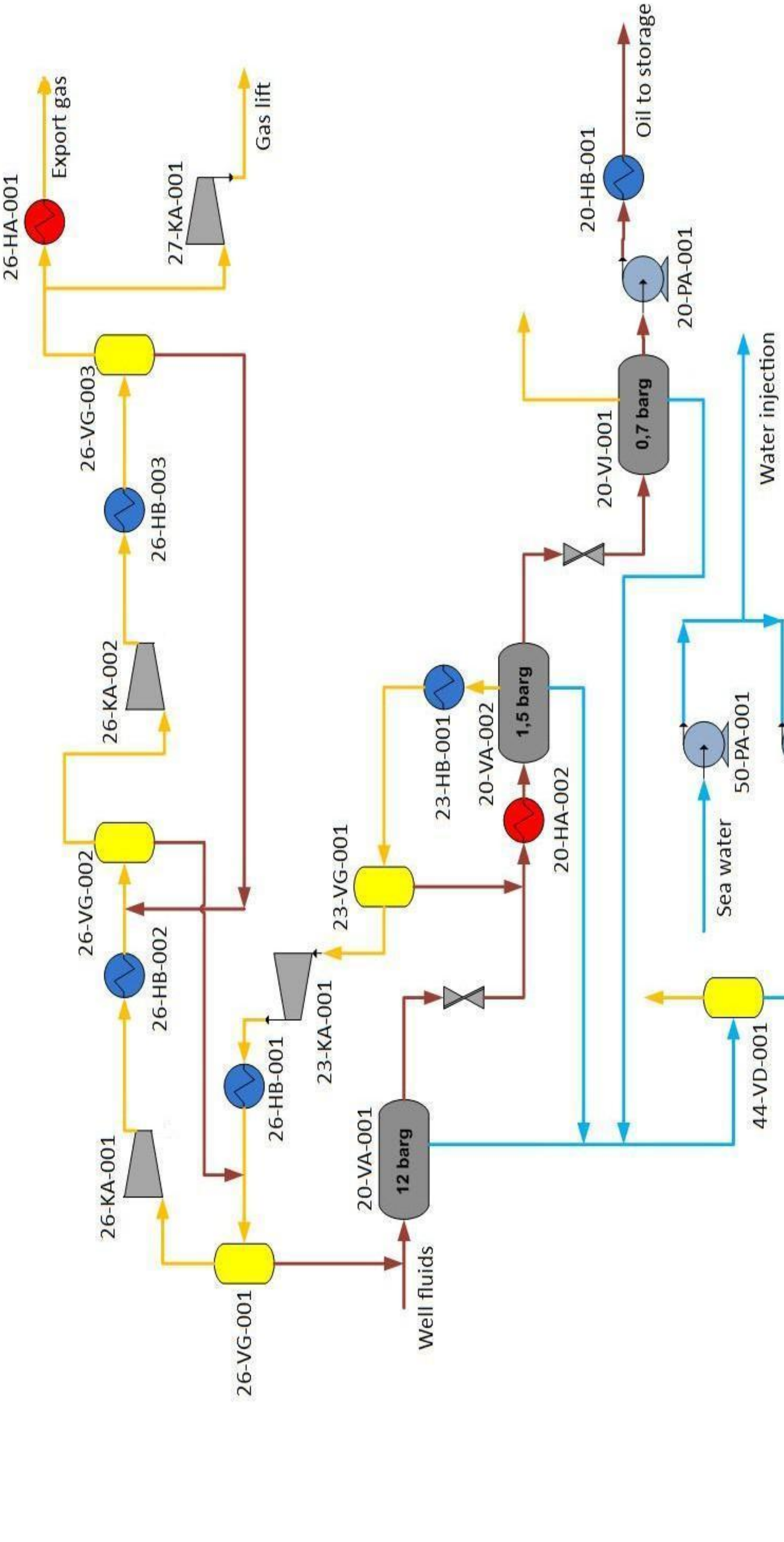

Figure 6-1: Field A separation process

6.2 Detailed description

The following sub chapters contain a detailed description of the Field A separation and stabilization process.

6.2.1 Well fluids

The well fluids consist of 7000 Sm^3/d oil and 11000 Sm^3/d water. The gas associated with the oil will be approximately 1,8 MSm^3/d. This is higher than the design gas handling capacity of 1,5 MSm^3/d, but because this is the gas associated with the design oil capacity, it is used as the gas flow in this thesis [9].

The well fluids enter the process at 105°C and 320 barg. No pipeline losses or subsea choking of the well stream has been accounted for in this thesis. A choke valve lowers the pressure to the operating pressure of the 1st stage separator [9].

6.2.2 Separation and stabilization

The separation consists of two separation stages and an electrostatic coalescer. The 1st stage separator operates at 12 barg, the 2nd stage separator at 1 barg and the coalescer at 0,7 barg. The TVP-spec for the stabilized oil is TVP < = 0,965 bara at 37,8°C. A heater is installed upstream of the 2nd stage separator in order for the oil to reach the desired TVP-spec. The simulation of the process shows that the oil is not in need of heating in order to be stabilized, but the heater should still be installed, as it could be necessary later in the field's lifetime, or during startup or shutdown of the process [8], [9].

The stabilized oil downstream of the coalescer is pumped and cooled to 2,5 barg and 50°C before being sent to storage.

The gas being separated is cooled to 30°C and scrubbed before being sent to the gas compression train. The liquids being scrubbed out are sent back to the separators.

Water carry-over levels of 15 mole% in the 1st stage separator, 2 mole% in the 2nd stage separator and 0,5 mole% in the coalescer are implemented [8].

6.2.3 Produced water

The produced water is separated from the oil and sent to a degassing tank for removal of nitrogen, CO_2 and any remaining hydrocarbon components. It is then pumped to a water injection pressure of 150 barg, and mixed with injection seawater at the same pressure [2], [8].

6.2.4 Gas compression train

The gas leaving the separators is sent to the gas compression train, where it is first compressed to 45 barg, cooled to 30°C and scrubbed. It is then compressed to 125 bara, cooled to 30°C and scrubbed once more. The liquids being scrubbed out are sent back to scrubbers earlier in the process. Before export, the gas is heated to 65°C [2].

400000 Sm^3/d of gas is taken out as lift gas after the 2^{nd} stage of compression, cooling and scrubbing, at a pressure of 125 bara and a temperature of 30°C. This gas is compressed to a gas lift pressure of 170 bara [2].

6.2.5 Heat exchangers

This thesis does not focus on the heating or cooling medium used in the separation process, so the heat exchangers used in the simulations are implemented without them. The exchangers only give out the amount of heat which has to be added to or removed from the fluids in need of heating or cooling.

All heat exchangers with duty have been simulated with a pressure drop of 0,5 bar. Pressure drop over heaters with no duty has been set to 0 bar as a bypass is assumed [9].

6.2.6 Efficiencies

All pumps, compressors and turbines have been implemented with 75 % adiabatic efficiency [10].

6.3 Demands for power, heating and cooling

The demands for power, heating and cooling for the Field A separation and stabilization process follows in the sub chapters below.

6.3.1 Power demand

The pumps and compressors in the process make up the total power demand. Table 6-1 shows the process' power demand.

Table 6-1: Field A process power demand

Unit operation	Power demand [kW]
20-PA-001	28
23-PA-001	0
26-PA-001	2
29-PA-001	2632
44-PA-001	2
44-PA-002	0
50-PA-001	2030
23-KA-001	390
26-KA-001	4294
26-KA-002	2557
27-KA-001	125
Total power demand [kW]	**12060**

6.3.2 Heating demand

Neither the inlet heater (20-HA-001) nor the interstage heater (20-HA-002) have to heat the oil in order for it to be stabilized. Therefore the only heater duty needed in the process is to heat the export gas in the heater 26-HA-001. The heating duty needed is **1953 kW**.

6.3.3 Cooling demand

Table 6-2 shows the cooling demand for the Field A process.

Table 6-2: Field A process cooling demand

Unit operation	Cooling demand [kW]
20-HB-001	4289
23-HB-001	1759
26-HB-001	11003
26-HB-002	7132
26-HB-003	5380
Total cooling demand [kW]	**29563**

The demands for power, heating and cooling will vary from the values found in this chapter when the gas is dehydrated and conditioned. This will be further discussed in Chapter 7 and Chapter 8.

6.4 Export gas

This thesis focuses on the gas protraction for Field A. From the simulation of the process with no dehydration and conditioning, the state of the export gas is examined.

6.4.1 *Water dew point*

A case study has been performed in HYSYS to find the water dew point of the export gas at 69 barg. Figure 6-2 shows the result of the study; the mass flow of liquid water in the export gas at 69 barg and different temperatures.

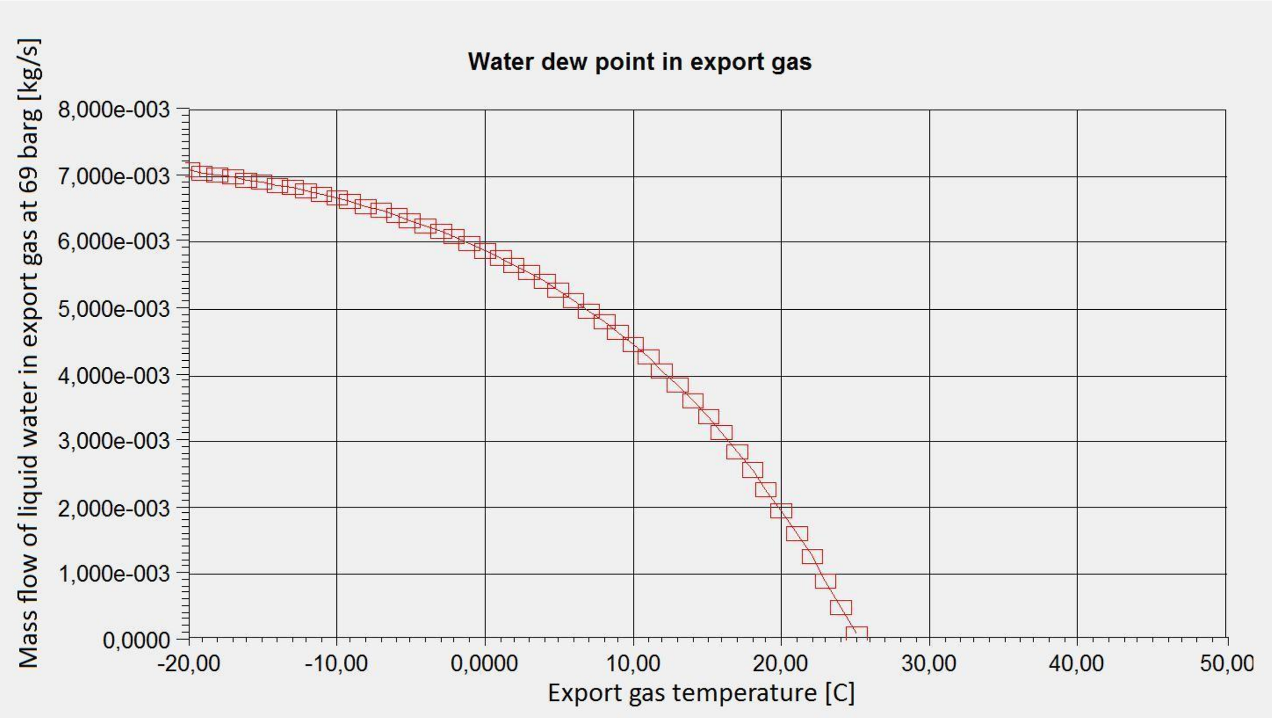

Figure 6-2: Water dew point in export gas

The study shows that the water dew point of the export gas at 69 barg when it is not dehydrated is approximately 25°C. In Table 3-1 and Table 3-2 it can be read that the specifications for the water dew point is -18°C at 69 barg for Statpipe and -12°C at 69 barg for Vesterled. This means the basic separation process will not dehydrate the gas enough for export in neither Statpipe nor Vesterled. To be able to export the gas in one of the pipelines, the gas needs to be additionally dehydrated. This will be further discussed in Chapter 7.

6.4.2 *Hydrocarbon dew point*

A case study has been performed in HYSYS to find the hydrocarbon dew point of the export gas at 50 barg. Figure 6-3 shows the result of the study; the mass flow of liquid hydrocarbons in the export gas at 50 barg and different temperatures.

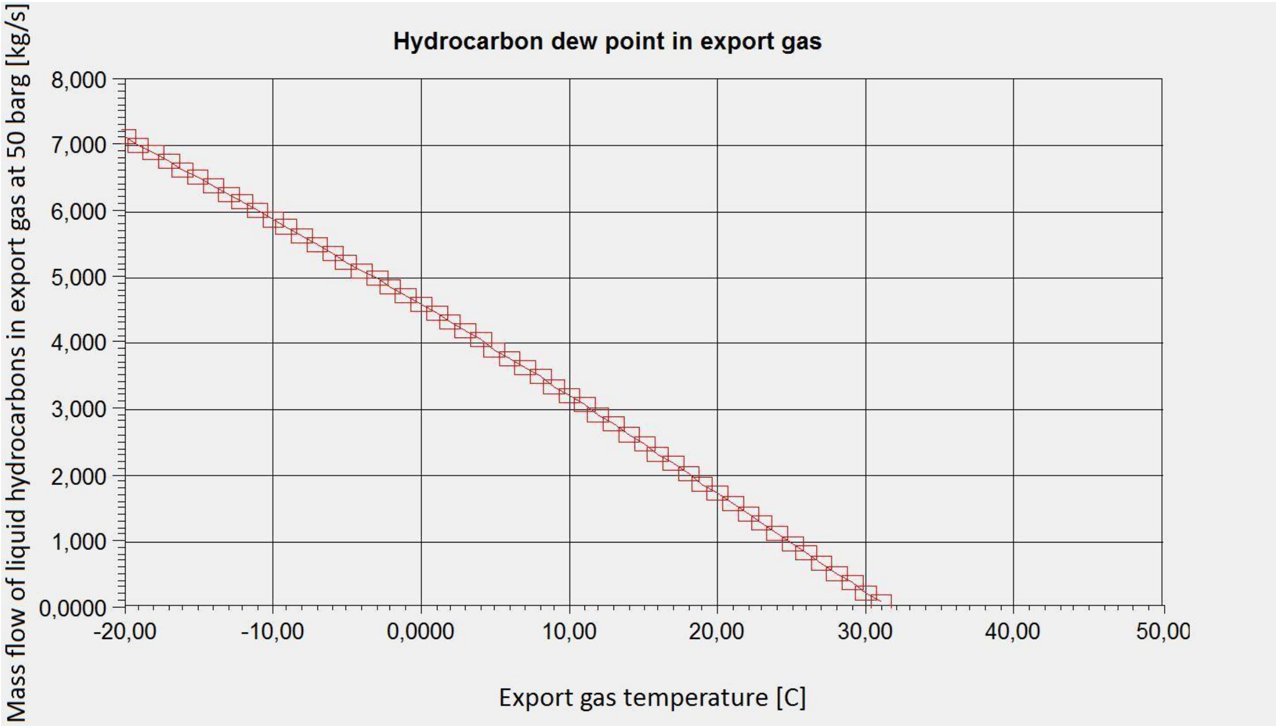

Figure 6-3: Hydrocarbon dew point in export gas

The study shows that the hydrocarbon dew point of the export gas at 50 barg when it is not conditioned is approximately 31°C. Table 3-1 and Table 3-2 show that the specification is -10°C for Statpipe and -3°C for Vesterled. This means the gas has to be conditioned before export in either of the pipelines. Conditioning of the gas will be discussed in Chapter 8.

7. Dehydration of the Field A gas

As mentioned in Chapter 6.4.1, the export gas from Field A has to be dehydrated before export in Statpipe and Vesterled. Different methods to obtain acceptably low water content in the gas have been described in Chapter 4. Simulations have been performed in HYSYS to test these dehydration processes. The basic separation process described in Chapter 6 has been used as a basis, and the simulation model has been expanded with the equipment needed for dehydration.

As mentioned in Chapter 4, dehydration by absorption is the most common method used, but the other methods have also been examined in this chapter [8].

The export specifications for Statpipe and Vesterled are listed in Table 3-1 and Table 3-2 respectively.

7.1 Dehydration by Joule-Thomson valve process

Dehydration by depressurizing using a Joule-Thomson valve has initially been simulated. This is the dehydration method with the lowest investment costs [6].

At the 2nd stage in the gas compression train, instead of being compressed to the export pressure of 125 bara, the gas is compressed to 170 bara. The gas is then depressurized over a Joule-Thomson valve and sent to a knock-out drum for removal of liquids. It is then compressed to the export pressure and cooled to 50°C, the maximum export temperature in Statpipe [9]. Figure 7-1 shows a schematic representation of the dehydration process. A detailed process flow diagram of the whole process as it has been simulated in HYSYS can be found in Appendix E.

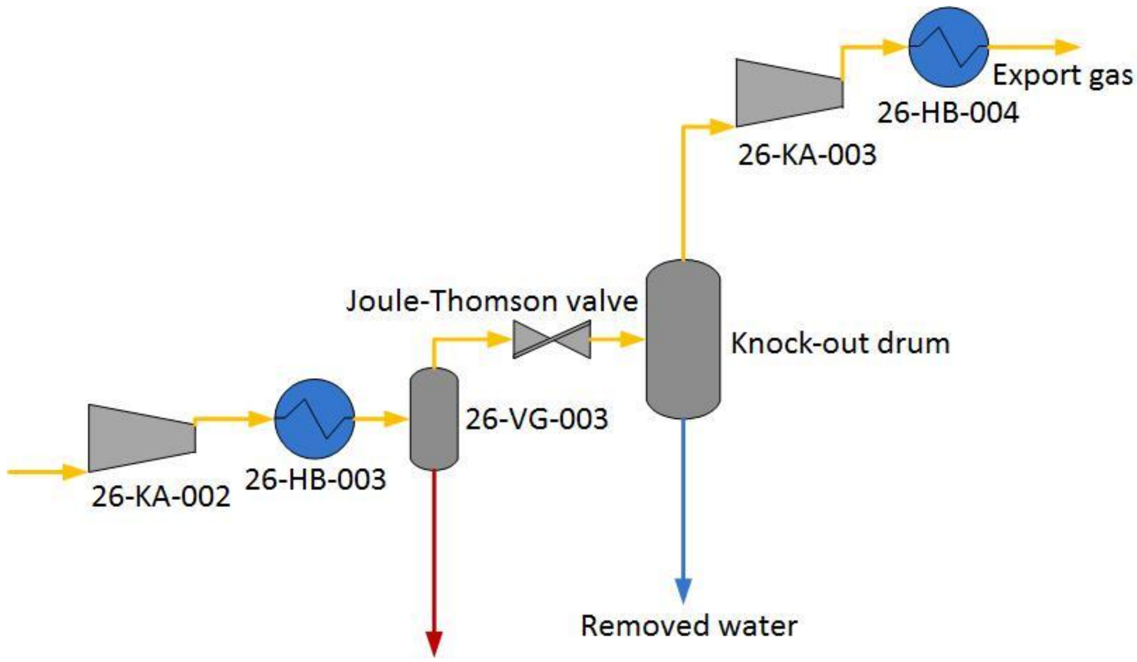

Figure 7-1: Dehydration by Joule Thomson valve process

The pressure drop over the valve can be regarded as a loss, and the higher the drop is, the more work has to be put in compressing the gas to the export pressure afterwards. If the demand for recompression after the dehydration is high, it also has to be done over several steps, and the investment costs for the process increases. It would therefore be preferable to lower the pressure just enough to get out the amount of water that is needed. The 2nd stage compressor in the gas compression train will also need more work, as it compresses the gas to 170 bara instead of 125 bara.

7.1.1 Export in Statpipe

A case study has been performed, where the pressure after the JT-valve is varied and the mass flow of liquid water in the export gas at 69 barg and -18°C is checked. Table 7-1 shows the result of the case study. The HYSYS simulation file used in Chapter 7.1.1 is named *FieldA_JT_Statpipe.*

Table 7-1: Liquid water in export gas after JT-valve dehydration

Pressure after JT-valve [barg]	Liquid water flow in export gas (69 barg, -18°C) [kg/s]
0	0,0129
5	0,0129
10	0,0129
15	0,0129
20	0,0129
25	0,0129
↓	↓
125	0,0129

The table shows that dehydration using a Joule-Thomson valve is not a possible solution. No matter how low the pressure after the valve is, no liquid water will fall out of the gas, and the gas is not dehydrated at all. Since the JT-valve process not will be used, the amount of additional work needed for compression will not be discussed.

7.1.2 Export in Vesterled

In Chapter 7.1.1 it was shown that dehydration of the gas using a Joule-Thomson valve would not work if the gas is to be exported in Statpipe, since no liquid water would form as a result of the pressure drop. Even though the specifications for Vesterled are not as strict as those for Statpipe, the dehydration process itself will be the same, so still no liquid water can be separated from the gas using a Joule-Thomson valve. This means a JT-valve dehydration process can't be used for export in Vesterled either.

7.2 Turbo expander process

Instead of depressurizing the gas over a Joule-Thomson valve, it can be depressurized through a turbo expander. Pressure drop over a turbo expander can be used as work elsewhere in the process, so installing a turbo expander instead of a valve has a positive impact on the process' power demand. Also, by extracting work from the pressure drop, the temperature of the gas will be lowered more, so more free water could fall out of the gas and be separated [6].

The investment and maintenance cost are higher for a turbo expander than for a Joule-Thomson valve [6].

The turbo expander process will be identical to the JT-valve process, apart from the valve being replaced by a turbo expander. The 2nd stage in the gas compression train takes the gas to 170 bara, instead of the export pressure of 125 bara [9]. Figure 7-2 shows a schematic representation of the dehydration process. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

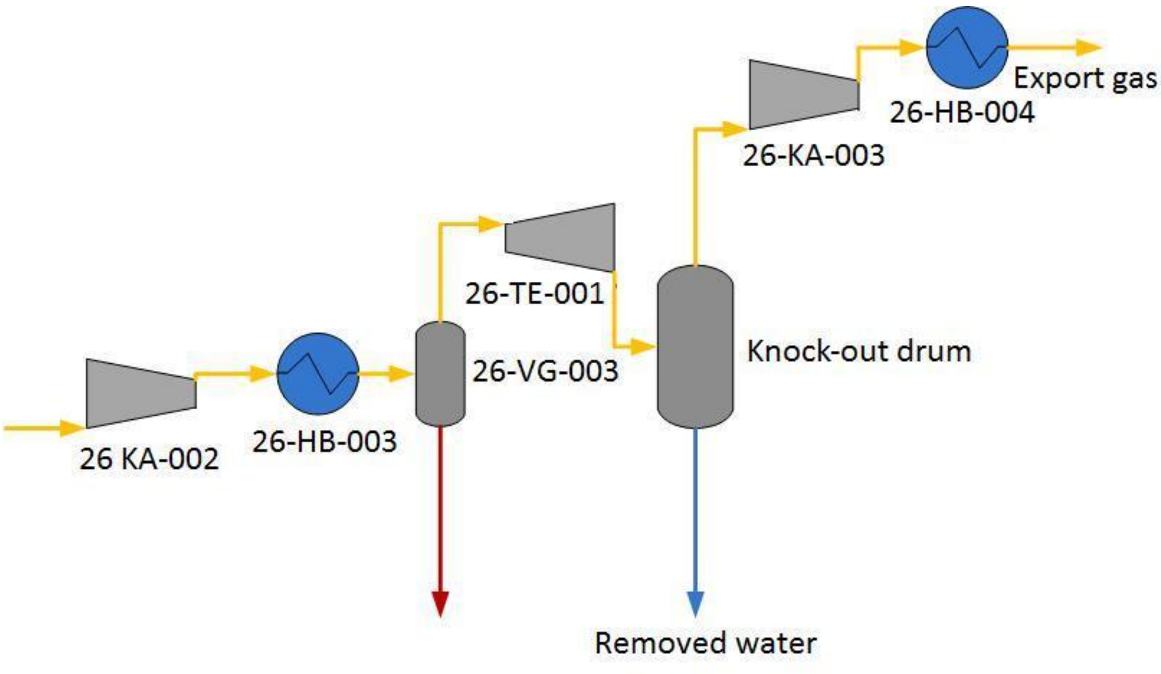

Figure 7-2: Dehydration by turbo expander

7.2.1 Export in Statpipe

A case study has been performed, where the pressure after the turbo expander is varied and the mass flow of liquid water in the export gas at 69 barg and -18°C is checked. Figure 7-3 shows the result of the case study. The HYSYS simulation file used in Chapter 7.2.1 is named *FieldA_TE_Statpipe.*

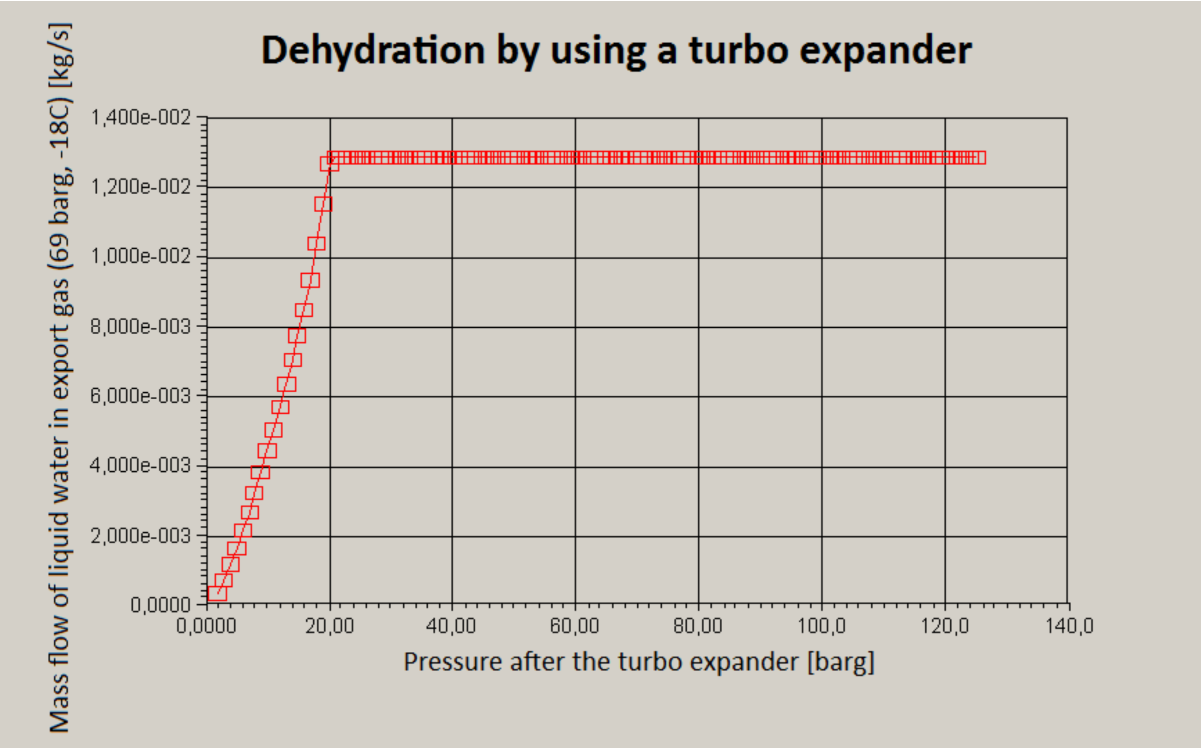

Figure 7-3: Dehydration by using a turbo expander for export in Statpipe

The figure shows that it is possible to dehydrate the gas using a turbo expander, but the pressure drop has to be very high to achieve it. In order to reach the water dew point specification for Statpipe, the pressure after the turbo expander has to be 0 barg. If this is done, the temperature of the gas downstream of the expander will get as low as -62°C. Calculations in HYSYS shows that the hydrate formation temperature for the gas is approximately -30°C downstream of the expander, so unless MEG is injected, hydrates almost certainly will form.

Also, even though work can be extracted from the turbo expander, the total power demand for the process will still increase, as the gas needs to be recompressed to the export pressure after the dehydration. The compression to 170 bara instead of 125 bara at the 2nd stage compressor also causes an increase in the power demand. The demands for heating and cooling may also vary, as the recycle streams vary due to the dehydration. Table 7-2 shows a comparison between the demands for power, heating and cooling for the basic process with no dehydration and conditioning described in Chapter 6, and the process with dehydration by turbo expander for export in Statpipe. The total power demand for the process with dehydration is the sum of all power demands in the process with the extractable work from the turbo expander subtracted.

Table 7-2: Comparison of power demand with turbo expander dehydration for export in Statpipe

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander	19169	0	34567

The table shows that the power demand for the process increases with 7109 kW if a turbo expander dehydration process is used for the gas to reach the export water dew point specification. The heating demand decreases with 1953 kW. The reason the gas is in no need of heating when it is dehydrated, is that after the recompression following the dehydration, the gas is so hot that it needs cooling instead of heating to reach the export temperature. This also makes the cooling demand increase, totally by 5004 kW.

The investment and maintenance costs will also increase by introducing a turbo expander dehydration process. The export recompression after the dehydration will have to be from 0 barg to 125 bara, an increase that probably will demand installation of three compression stages.

Based on the hydrate formation problems, the increased power demand and the increased investment and maintenance costs, it seems dehydration with a turbo expander is not a good solution, though it is possible [9].

7.2.2 Export in Vesterled

A case study has been performed, where the pressure after the turbo expander is varied and the mass flow of liquid water in the export gas at 69 barg and -12°C is checked. Figure 7-4 shows the result of the case study. The HYSYS simulation file used in Chapter 7.2.2 is named *FieldA_TE_onlyDehyd_Vesterled.*

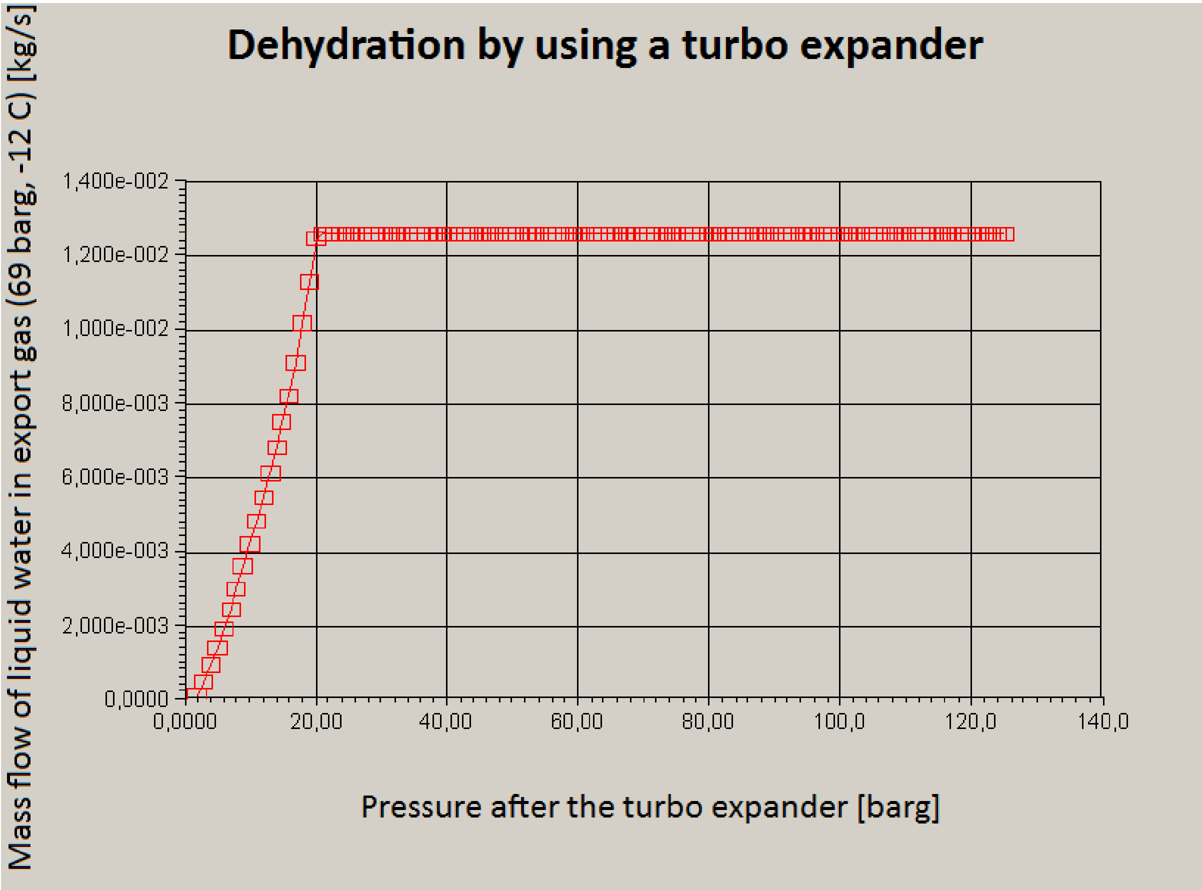

Figure 7-4: Dehydration by using a turbo expander for export in Vesterled

Just as for export in Statpipe, the figure shows that dehydration by a turbo expander for export in Vesterled requires a large pressure drop over the turbo expander. The pressure after the turbo expander has to be 1 barg if the export water dew point specification is to be reached. Even though this is 1 bar higher than the demand for export in Statpipe, the difference is so small that the same problems described in Chapter 7.2.1; substantially increased power demand and hydrate formation after the turbo expander, will be experienced here. So even though it is possible to use a turbo expander for dehydration, it doesn't seem a good solution.

Table 7-3 shows the increase in power demand if a turbo expander lowering the pressure to 1 barg is used for dehydration, compared to the basic process with no dehydration. The extractable power from the turbo expander has been subtracted. The table also shows the changes in demand for heating and cooling.

Table 7-3: Comparison of power demand with turbo expander dehydration for export in Vesterled

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander	17890	0	32469

The table shows that the power demand increases with 5830 kW, the heating demand diminishes for the same reason as described in Chapter 7.2.1 and the cooling demand increases with 2906 kW.

7.3 Propane refrigeration cycle

Another solution to dehydrate the gas could be to cool it without lowering the pressure. A refrigeration cycle using propane as the refrigerant is one way to do this. In the basic separation process the gas is cooled to 30°C after the 1st stage compressor in the gas compression train, before being scrubbed and sent to the 2nd stage compressor. If the temperature at this cooling stage instead is lowered substantially more, the scrubber following can remove much more water [9].

Figure 7-5 shows a schematic representation of the process. A detailed description of the refrigerant cycle can be found in Appendix C. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

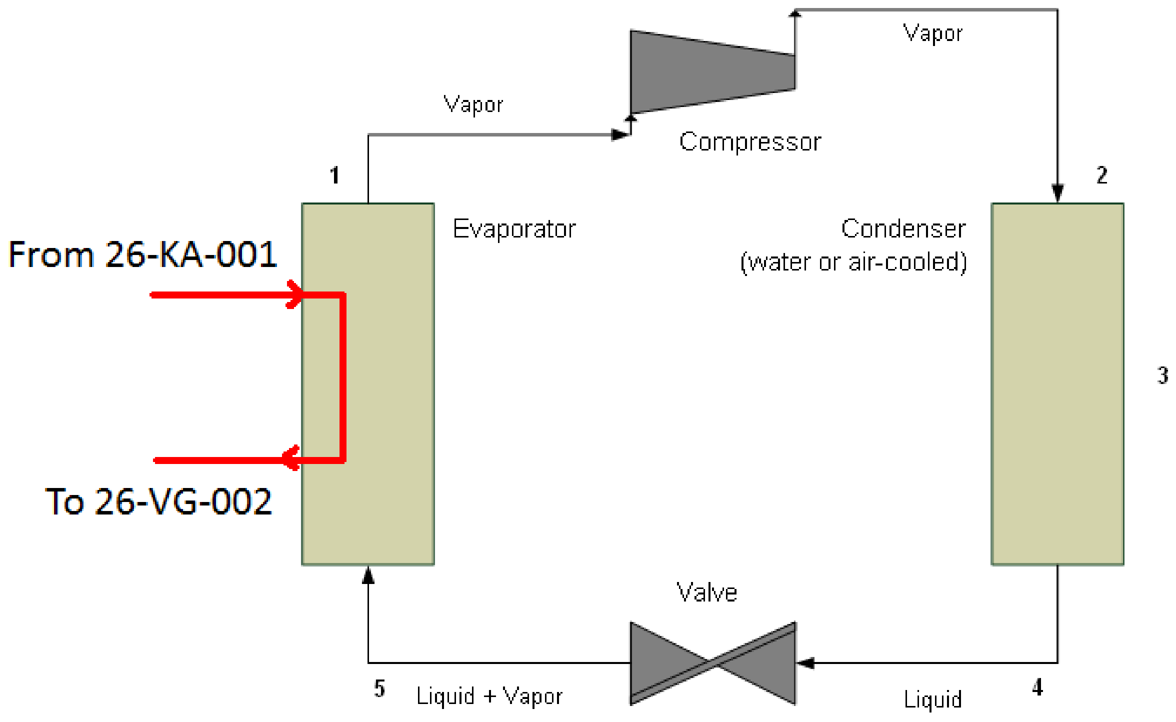

Figure 7-5: Dehydration by propane refrigerant cycle

7.3.1 Export in Statpipe

A case study has been performed, where the temperature of the gas after the propane refrigerated cooler is varied, and the mass flow of liquid water in the export gas at 69 barg and -18°C is checked. Figure 7-6 shows the result of the case study. The HYSYS simulation file used in Chapter 7.3.1 is named *FieldA_refrig_Statpipe.*

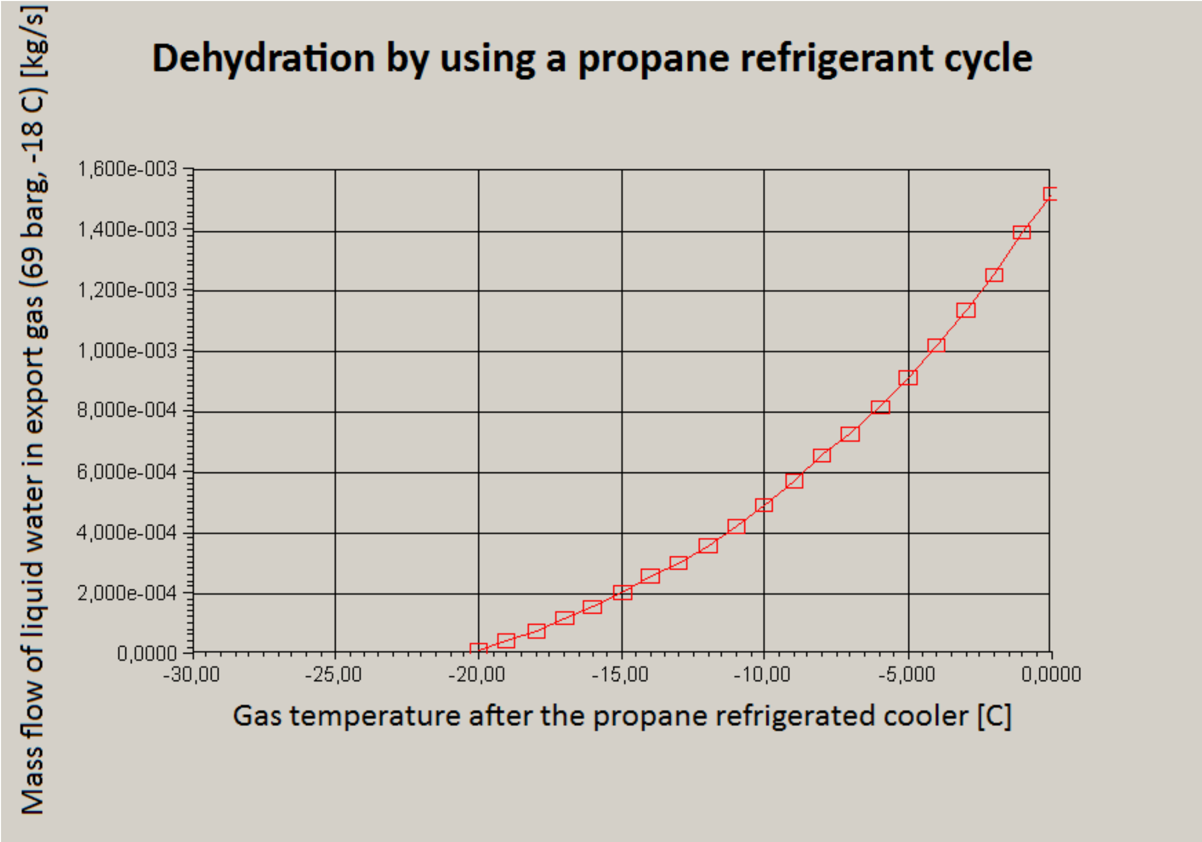

Figure 7-6: Dehydration by using a propane refrigerant cycle for export in Statpipe

The figure shows that the gas needs to be cooled to -21°C in order for it to reach the export water dew point specification for Statpipe. Calculations in HYSYS show that the hydrate formation temperature downstream of the propane cooler is approximately 16°C, so without injection of a hydrate inhibitor, hydrates will form.

Due to the low temperature downstream of the cooler, many of the heavier hydrocarbons in the gas will condense, and be separated out in the next scrubber. This will cause a large recycle stream, which will be carried back to previous stages in the process, and thereby being compressed again by the early compressors in the process, increasing the mass flows in these compressors. The low temperature of the recycle stream will also cause a lower temperature where it is mixed back into the process, eventually causing a large heating demand upstream of the 2nd stage separator to stabilize the export oil. In other words, dehydrating the gas using a propane refrigeration cycle will cause large increases in the demands for power and heating.

Table 7-4 shows a comparison of the demands for power, heating and cooling between the basic process with no dehydration and conditioning, and the process where the gas is dehydrated using a propane refrigeration cycle.

Table 7-4: Comparison of power demand with dehydration by a propane refrigeration cycle for export in Statpipe

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	170360	563895	776851

The table shows that all three demands increase substantially. The power demands increases by approximately 158 MW, to a total of approximately 170 MW, mainly due to the compressors 23-KA-001 and 26-KA-001. This is because these two compressors have to recompress the large recycle streams caused by the low temperature after the propane cooler. A power demand of 170 MW for a marginal field like Field A, will probably cause both the investment costs and the operational cost to be too high for a development of the field to be economically profitable.

As for dehydration using a turbo expander, dehydration using a propane refrigerant cycle is possible, but this alternative also will lead to problems. The necessary temperature after the propane refrigerated cooler will be much lower than the hydrate formation temperature, so a hydrate inhibitor will have to be injected, which again leads to the problem of removing this inhibitor from the gas again before export. But the main problem, which probably makes this dehydration method impossible to implement, is a massive power demand of 170 MW [9].

7.3.2 Export in Vesterled

A case study has been performed, where the temperature of the gas after the propane refrigerated cooler is varied, and the mass flow of liquid water in the export gas at 69 barg and -12°C is checked. Figure 7-7 shows the result of the case study. The HYSYS simulation file used in Chapter 7.3.2 is named *FieldA_refrig_Vesterled.*

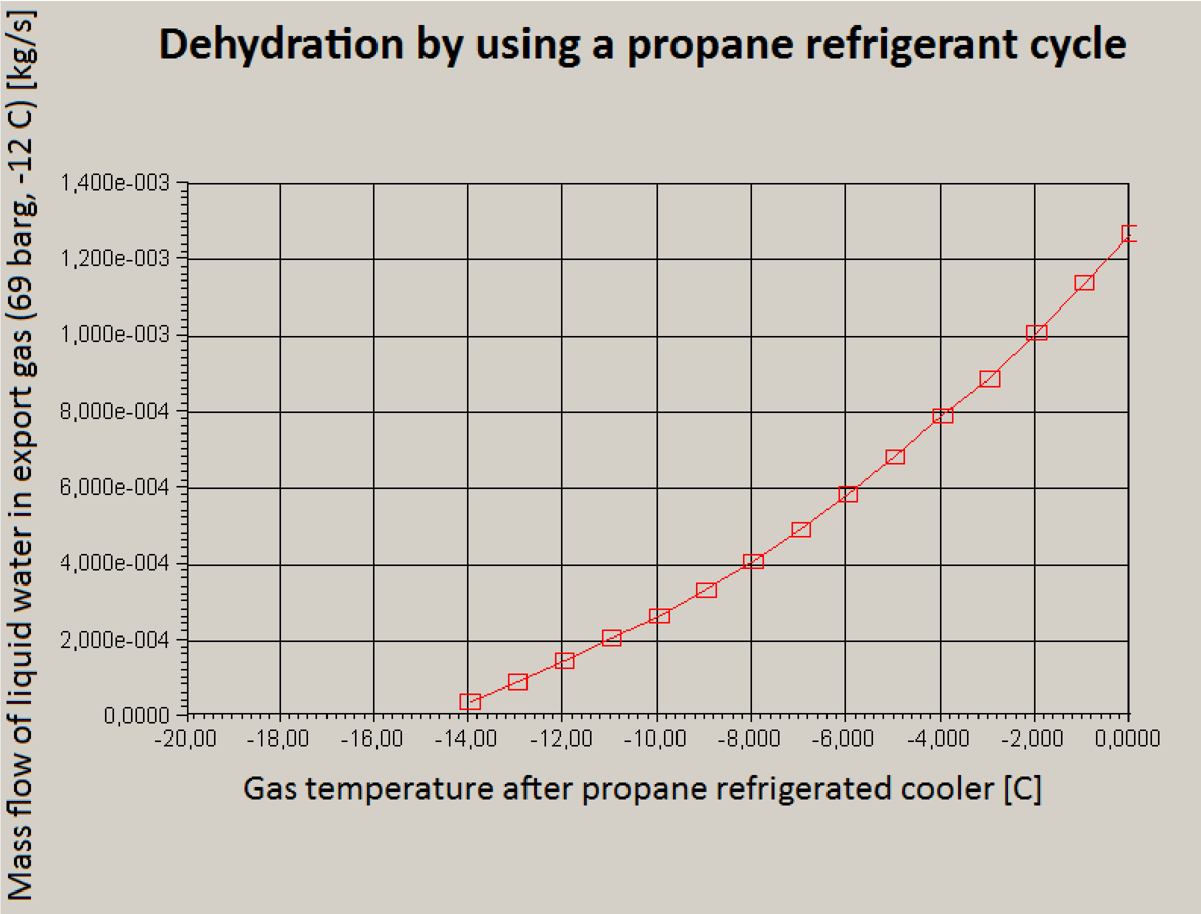

Figure 7-7: Dehydration by using a propane refrigerant cycle for export in Vesterled

The study shows that the gas needs to be cooled to -15°C to reach the export specification for Vesterled. Since the hydrate formation temperature after the propane refrigerated cooler is 16°C, cooling the gas this much will cause hydrate problems.

As explained in Chapter 7.3.1, the demands for power, heating and cooling will increase due to the large recycle stream caused by the low temperature downstream of the propane cooler. For export in Vesterled the gas only has to be cooled to -15°C, compared to -21°C for export in Statpipe. This means the recycle mass flows will be somewhat smaller, and some power could be saved. Table 7-5 shows a comparison of the demands for power, heating and cooling between the basic process without dehydration and conditioning, and the process using a propane refrigeration cooler to dehydrate the gas for export in Vesterled.

Table 7-5: Comparison of power demand when using a propane refrigeration cycle to dehydrate the gas for export in Vesterled

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	136685	443153	622352

As for the process described in Chapter 7.3.1, all three demands increase substantially. The total power demand is approximately 137 MW, somewhat smaller than the demand in Chapter 7.3.1, but still most likely too high to make the development alternative profitable for a marginal field like Field A. Most of the increase in power demand is in the compressors 23-KA-001 and 26-KA-001, due to recompression of large recycle streams.

As for export in Statpipe, dehydration using a propane refrigerant cycle for export in Vesterled is possible in principle, but a massive power demand of 137 MW makes it not profitable to implement. Hydrate problems will also occur [9].

7.4 Combination of turbo expander process and refrigerant cycle

When using a turbo expander cycle to dehydrate the gas, the main problem was the increased demands for power, as the pressure had to be lowered very much. When using a propane refrigerant cycle, the main problems were a very high power demand and hydrate formation due to very low temperatures. A simulation model has been developed where the two methods have been combined, in an attempt to remove them both.

In this model, after the 1^{st} stage in the gas compression train, the gas is cooled in a propane refrigerated cooler, and then scrubbed, as described in Chapter 7.3. It is then compressed to 170 bara, scrubbed, and sent trough a turbo expander, as described in Chapter 7.2. After the pressure drop, free water is removed in a knock-out drum, and the gas is recompressed for export. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

7.4.1 Export in Statpipe

A case study was performed on the simulation, where the flow of liquid water in the export gas at 69 barg and -18°C was checked when the pressure after the turbo expander was varied. The temperature after the propane cooler was set at 17°C, as the hydrate formation temperature is approximately 16°C. Table 7-6 shows the result of the case study. The mass flows of liquid water listed in the table are in the export gas at 69 barg and -18°C. The HYSYS simulation file used in Chapter 7.4.1 is named *FieldA_refrig_TE_onlyDehyd_Statpipe.*

Table 7-6: Dehydration by combining turbo expander and refrigerant cycle for export in Statpipe

Pressure after turbo expander [barg]	Mass flow of liquid water [kg/s]
0	0,00000
1	0,00000
2	0,00000
3	0,00000
4	0,00006
5	0,00022
6	0,00039
7	0,00058
8	0,00079
9	0,00101
10	0,00123

The case study shows that a combination of 17°C downstream of the propane cooler and a pressure of 3 barg downstream of the turbo expander will dehydrate the gas sufficiently. This will eliminate the hydrate problem downstream of the propane cooler, but there will still be hydrate problems downstream of the turbo expander, as the temperature here will be approximately -44°C, with a hydrate formation temperature of approximately -9°C.

Some power will be saved by using this dehydration method, both because of smaller recycle streams due to higher temperature after the propane cooler, and because of a lower pressure drop over the turbo expander. Table 7-7 shows the demands for power, heating and cooling when the gas is dehydrated using a turbo expander and a propane cooler compared to the basic process with no dehydration. The extractable work from the turbo expander has been subtracted.

Table 7-7: Comparison of power demand with dehydration by turbo expander and propane cooler for export in Statpipe

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	16952	0	35233

The table shows that the power demand increases with 4892 kW, the heating demand is 0 kW due to the recompression of the export gas after the turbo expander dehydration, and the cooling demand increases with 5670 kW.

This solution is a possible method to dehydrate the gas, but it will increase the demands for power and cooling, and hydrate formation problems after the turbo expander. A hydrate inhibitor, like MEG, will have to be injected here, and then be removed later in the process [9].

7.4.2 *Export in Vesterled*

A case study was performed on the simulation, where the flow of liquid water in the export gas at 69 barg and -12°C was checked when the pressure after the turbo expander was varied. The temperature of the gas after the propane cooler was set at 17°C, thereby eliminating the hydrate formation problems there. Table 7-8 shows the results of the case study. The HYSYS simulation file used in Chapter 7.4.2 is named *FieldA_refrig_TE_onlyDehyd_Vesterled.*

Table 7-8: Dehydration by combining turbo expander and refrigerant cooler for export in Vesterled

Pressure after turbo expander [barg]	Mass flow of water [kg/s]
0	0,00000
1	0,00000
2	0,00000
3	0,00000
4	0,00000
5	0,00000
6	0,00016
7	0,00035
8	0,00055
9	0,00077
10	0,00100

The table shows that a combination of 17°C downstream of the propane cooler and 5 barg downstream of the turbo expander will dehydrate the gas enough for export in Vesterled. The hydrate problems downstream of the propane cooler will be eliminated, but there will still be hydrate formation problems downstream of the turbo expander, as the temperature here will be approximately -34°C, with a hydrate formation temperature of approximately -5°C.

Table 7-9 shows the power demand if dehydration by combining a propane refrigerated cooler cooling the gas to 17°C and a turbo expander lowering the pressure to 5 barg is used, compared to the basic process with no dehydration. The extractable power from the turbo expander has been subtracted. The table also shows the demands for heating and cooling.

Table 7-9: Comparison of power demand with combined dehydration for export in Vesterled

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander	16431	0	34266

The table shows an increase in the power demand of 4371 kW, the heating demand is 0 kW due to recompression of the export gas after the turbo expander dehydration, and the cooling demand increases with 4703 kW.

7.5 Dehydration by glycol absorption

The final dehydration alternative for Field A is glycol absorption. A simulation model has been developed where the gas is dehydrated in an absorption column using TEG as the absorbent. A TEG regeneration process has also been implemented.

The absorber is placed between the 1st and 2nd stage in the gas compression train [9]. An amount of approximately 25 liters TEG at 65˚C and 44 barg per kilogram water needed to be removed from the gas is fed at the top of the column, and the wet natural gas is fed at the bottom of the column [6]. Some specifications are implemented in the absorber. Then the number of theoretical contact plates inside the absorber is increased stepwise, until adequate water content in the export gas is reached. The rich TEG containing the removed water is taken out at the bottom of the absorber, and is then sent to the TEG regeneration unit. Some TEG will be lost in the cycle, either to the natural gas or to the removed water. Therefore a makeup TEG stream is added to the regenerated TEG entering the absorber again [6], [9], 10].

A detailed description and schematic representation of the absorption and TEG regeneration process can be found in Appendix D. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

7.5.1 Export in Statpipe

The HYSYS simulation file used in Chapter 7.5.1 is named *FieldA_TEG_onlyDehyd_Statpipe.*

The simulation show that if the absorber is fed with 119 m^3/d of TEG at 65˚C and 44 barg and contains 8 theoretical contact plates, the gas will be sufficiently dehydrated. Approximately 0,23 m^3/d of TEG will be lost, and will have to be added in the makeup TEG stream.

By dehydrating the gas in a TEG absorber, no hydrate problems will occur. Also, the gas won't have to be recompressed before export, as it will have to be if it is dehydrated using a turbo expander process. The power demand for the process is 12,3 MW, which is approximately the same as the basic process with no dehydration or conditioning.

Some extra heat flow is needed for a re-boiler in the TEG regeneration cycle. The heating demand increases from 1953 kW with no dehydration to 3183 kW when using absorption. The cooling demand for the process decreases slightly, and is 28361 kW.

7.5.2 Export in Vesterled

The HYSYS simulation file used in Chapter 7.5.2 is named *FieldA_TEG_onlyDehyd_Vesterled.*

The simulation gives almost the same results as the one in Chapter 7.5.1. If the absorber is fed with 119 m^3/d of TEG at 65˚C and 44 barg and contains 5 theoretical contact plates, the gas will be sufficiently dehydrated. The reduction in theoretical plates needed is a result of the less strict specifications for Vesterled. Approximately 0,22 m^3/d of TEG will be lost, and will have to be added in the makeup TEG stream. The demands for power, heating and cooling are the same as in Chapter 7.5.1.

7.6 Summary of dehydration methods at Field A

Five different dehydration alternatives have been tested for the Field A process.

Dehydration by using a **Joule-Thomson valve process** can't be used, as the gas won't be dehydrated no matter how high the pressure drop over the valve is.

Dehydration by using a **turbo expander process** is possible, though it will cause some problems. The pressure of the gas has to be lowered to 0 barg (Statpipe) or 1 barg (Vesterled) in the turbo expander in order for the gas to be dehydrated enough. This will cause large increase in power demand, compared to the process with no dehydration. There will be no need for heating in the process, but the cooling demand will increase. Also a hydrate inhibitor, like MEG, will have to be injected after the turbo expander. The inhibitor will have to be removed again later in the process.

Dehydration by using a **propane refrigeration cycle** is also possible, but very high demands for power, heating and cooling makes it very unlikely to be profitable.

A simulation where the gas was dehydrated by **combining a propane refrigeration cycle and a turbo expander process** was implemented, in an attempt to remove both the hydrate formation and the increased power demand problems. By cooling the gas to 17°C in the propane cooler and letting the pressure down to 3 barg (Statpipe) or 5 barg (Vesterled) in the turbo expander, the hydrate problems downstream of the propane cooler are removed. But there will still be hydrate problems downstream of the turbo expander and the demands for power and cooling will increase quite much.

Dehydration in a **TEG absorption column** was tested, and the simulation showed that this is a possible solution. The main problem with this dehydration alternative is increased heat flow demands.

Based on the results from the different simulations, it seems dehydration by using a **TEG absorption column** is the best alternative. This is also the dehydration method most commonly used in the industry. A final decision can however not be made until conditioning of the gas to reach adequate hydrocarbon dew point also is considered [8], [9].

8. Dew point control of the Field A gas

As mentioned in Chapter 6.4.2, the export gas has to be conditioned before export in either Statpipe or Vesterled. Different methods to condition the gas has been described in Chapter 5. Simulations have been made in HYSYS to test these conditioning methods. The different simulations described in Chapter 7 have worked as bases, and if necessary, they have been expanded with equipment to condition the gas. This means the simulations in this chapter are performed so that the gas is both dehydrated and conditioned adequately.

8.1 Joule-Thomson valve process

In Chapter 7.1 it was found that the gas could not be dehydrated using a Joule-Thomson valve process. This means that even if the gas may be conditioned using this process, it cannot be exported in Statpipe or Vesterled anyway, due to its water content. Therefore this conditioning method will not be further studied in this chapter.

8.2 Turbo expander process

The process is identical to the one described in Chapter 7.2. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

8.2.1 Export in Statpipe

The HYSYS simulation file used in Chapter 8.2.1 is named *FieldA_TE_Statpipe.*

In chapter 7.2.1 it was found that the pressure had to be lowered to 0 barg through the turbo expander in order to dehydrate the gas adequately. As the gas needs a certain driving force further in the process, the pressure cannot be lowered any further.

The simulation also shows that even though lowering the pressure to 0 barg dehydrates the gas adequately, it will not lower the hydrocarbon dew point of the export gas enough to meet the specifications in Statpipe. This means a turbo expander process cannot be used to both dehydrate and condition the gas properly for export there.

8.2.2 Export in Vesterled

The HYSYS simulation file used in Chapter 6 is named *FieldA_TE_Vesterled.*

In chapter 7.2.2 it was found that the pressure had to be lowered to 1 barg through the turbo expander in order to dehydrate the gas adequately. Further investigation of the simulation shows that this will not lower the hydrocarbon dew point of the gas enough.

The pressure cannot be lowered to less than 0 barg through the turbo expander, as the gas needs a certain driving force further in the process. The pressure after the turbo expander was therefore set to 0 barg, to see if the gas then would be properly conditioned to meet the Vesterled hydrocarbon dew point specification. But investigation of the simulation show that the gas still won't be adequately conditioned, so a turbo expander process is not applicable for both dehydration and conditioning of the gas for export in Vesterled.

8.3 Propane refrigeration cycle

The process is identical to the one described in Chapter 7.3. A detailed description of the refrigerant cycle can be found in Appendix C. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

8.3.1 Export in Statpipe

The HYSYS simulation file used in Chapter 8.3.1 is named *FieldA_refrig_Statpipe.*

In Chapter 7.3.1 it was found that the temperature after the propane cooler had to be -21°C in order to dehydrate the gas properly. Further investigations of the simulation show that this also will lower the hydrocarbon dew point in the export gas enough for export in Statpipe. This means the process described in Chapter 7.3 can be used if the gas is to be exported in Statpipe. But as mentioned in Chapter 7.3.1, the power demand for this process, approximately 170 MW, is too high for it to be a realistic development alternative.

8.3.2 Export in Vesterled

The HYSYS simulation file used in Chapter 8.3.2 is named *FieldA_refrig_Vesterled.*

Chapter 7.3.2 showed that the temperature after the propane cooler had to be -15°C in order to dehydrate the gas properly. Further investigations of the simulation show that this also will lower the hydrocarbon dew point in the export gas enough for export in Vesterled. This means the process described in Chapter 7.3 can be used if the gas is to be exported in Vesterled. But as mentioned in Chapter 7.3.2, the power demand for this process, approximately 137 MW, is too high for it to be a realistic development alternative.

8.4 Combination of turbo expander process and refrigeration cycle

The process is identical to the one described in Chapter 7.4. A detailed description of the refrigerant cycle can be found in Appendix C. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

8.4.1 *Export in Statpipe*

As mentioned in Chapter 8.2.1, a turbo expander lowering the pressure to 0 barg will dehydrate the gas adequately for export in Statpipe, but it will need more conditioning. This can be achieved by combining a turbo expander process with a propane refrigeration cycle. Since the turbo expander will condition the gas at least some, the temperature after the propane cooler can be higher than -21°C, as it had to be when only a refrigeration cycle was used to dehydrate and condition the gas. This will cause lower mass flows in the recycle streams, resulting in a smaller increase in the demands for power, heating and cooling.

A case study has been performed, where the mass flow of liquid hydrocarbons in the export gas at 50 barg and -10°C is checked with varying temperature downstream of the propane cooler. The pressure after the turbo expander has been locked at 0 barg. Figure 8-1 shows the result of the case study. The HYSYS simulation file used in Chapter 8.4.1 is named *FieldA_refrig_TE_Statpipe*.

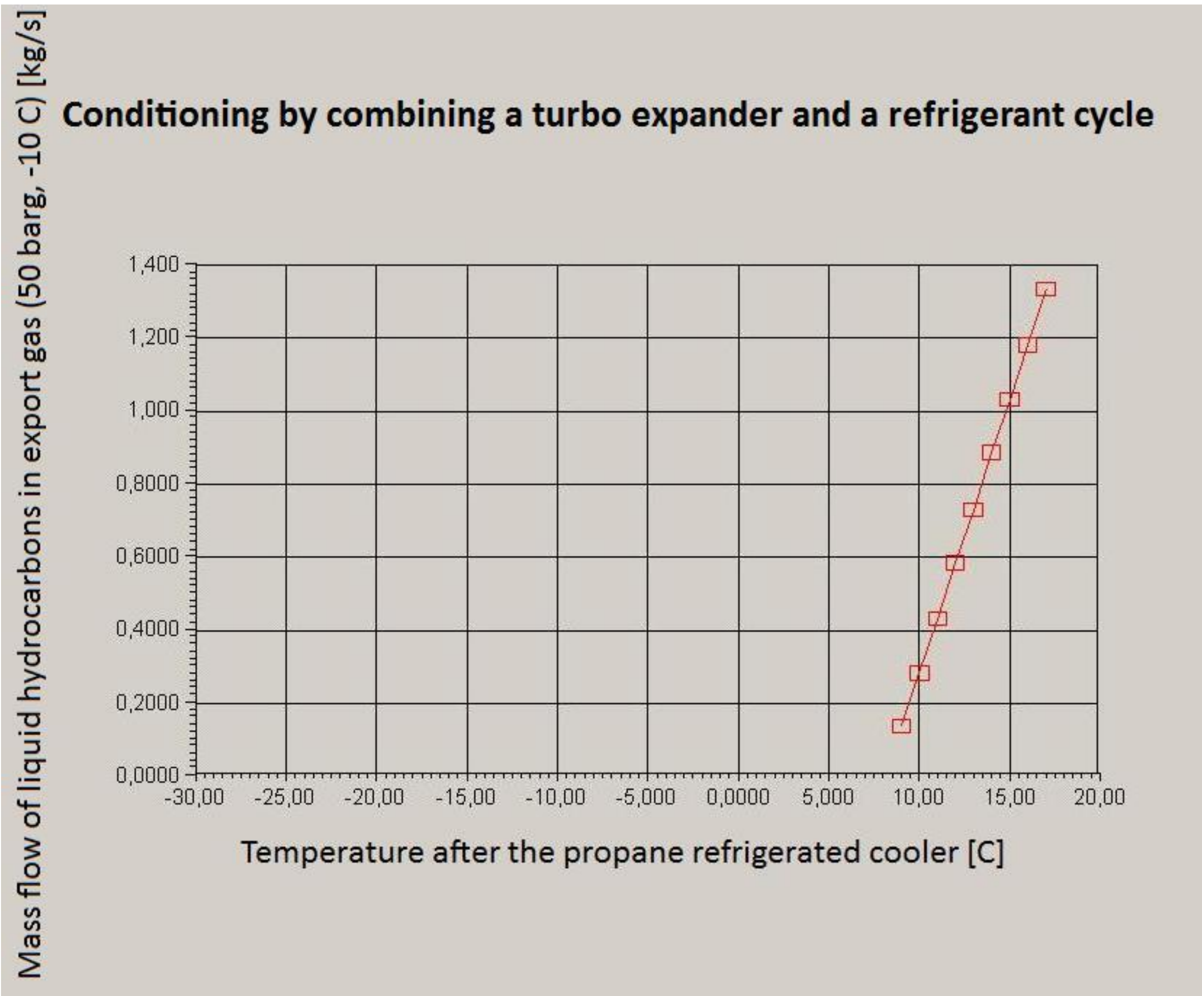

Figure 8-1: Conditioning by combining a turbo expander and a propane cooler for export in Statpipe

The figure shows that the gas has to be cooled to 8°C in the propane cooler in order for it to be adequately dehydrated and conditioned for export in Statpipe.

Table 8-1 shows a comparison of the demands for power, heating and cooling between the basic process with no dehydration or conditioning, and a process both dehydrating and conditioning the gas adequately for export in Statpipe by combining a turbo expander and a propane refrigerated cooler. The extractable power from the turbo expander has been subtracted.

Table 8-1: Comparison of power demand when combing a turbo expander and a propane cooler to dehydrate and condition the gas for export in Statpipe.

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	19774	0	41464

The demands for power and cooling increase with 7714 kW and 11901 kW respectively. A total power demand of approximately 20 MW could make it a profitable development alternative for Field A. The heating demand for the process will be 0 kW, due to recompression of the export gas after the turbo expander.

Hydrate formation problems will occur downstream of both the propane cooler and the turbo expander. As mentioned, the temperature has to be lowered to 8°C in the propane cooler, and the hydrate formation temperature downstream of the cooler is approximately 17°C. After the turbo expander the temperature will be approximately -80°C, with a hydrate formation temperature of -31°C. To avoid hydrates forming, MEG will have to be injected both places, and then be removed later in the process.

8.4.2 Export in Vesterled

A case study has been performed, where the mass flow of liquid hydrocarbons in the export gas at 50 barg and -3°C is checked with varying temperature downstream of the propane cooler. The pressure after the turbo expander has been locked at 0 barg. Figure 8-2 shows the result of the case study. The HYSYS simulation file used in Chapter 8.4.2 is named *FieldA_refig_TE_Vesterled.*

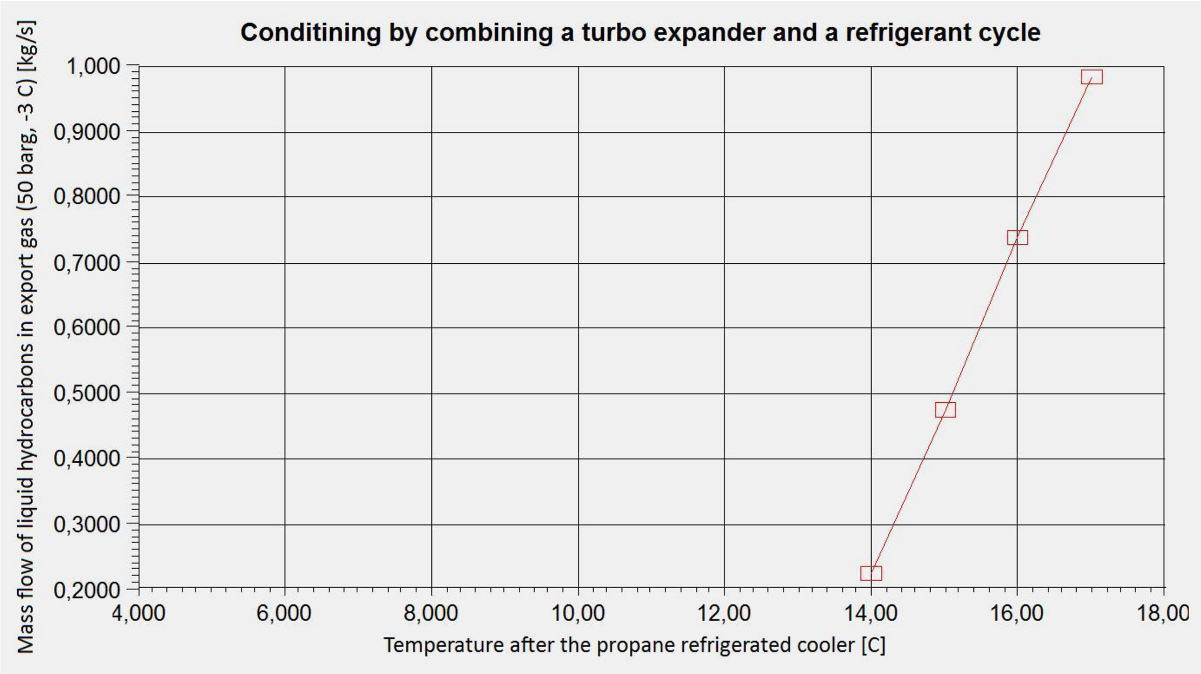

Figure 8-2: Conditioning by combining a turbo expander and a propane cooler for export in Vesterled

The figure shows that the gas has to be cooled to 13°C in the propane cooler in order for it to be adequately dehydrated and conditioned for export in Vesterled.

Table 8-2 shows a comparison of the demands for power, heating and cooling between the basic process with no dehydration or conditioning, and a process both dehydrating and conditioning the gas adequately for export in Vesterled by combining a turbo expander and a propane refrigerated cooler. The extractable power from the turbo expander has been subtracted.

Table 8-2: Comparison of power demand when combing a turbo expander and a propane cooler to dehydrate and condition the gas for export in Vesterled.

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	19530	0	38951

The demands for power and cooling increase with 7470 kW and 9388 kW respectively. A total power demand of approximately 19,5 MW could make it a profitable development alternative for Field A. The heating demand for the process will be 0 kW, due to recompression of the export gas after the turbo expander.

Hydrate formation problems will occur downstream of both the propane cooler and the turbo expander. As mentioned, the temperature has to be lowered to 13°C in the propane cooler, and the hydrate formation temperature downstream of the cooler is approximately 16°C. After the turbo expander the temperature will be approximately -76°C, with a hydrate formation temperature of -31°C. To avoid hydrates forming, MEG will have to be injected both places, and then be removed later in the process.

8.5 Dehydration by glycol absorption and conditioning with a refrigeration cycle

In Chapter 7.5 the gas was dehydrated using a glycol absorption column. To condition the gas after this, one can use a Joule-Thomson valve process, a turbo expander process or a refrigeration cycle. If a Joule-Thomson valve or a turbo expander is used, the power demand will increase due to recompression of the gas after the conditioning. Also, Chapter 8.2 indicates that a turbo expander cannot condition the gas adequately. Therefore a combination of dehydration by glycol absorption and conditioning by a refrigeration cycle has been simulated in this thesis. Since the gas is already dehydrated before entering the propane cooler, the problems with hydrate formation will not occur here [8], [9].

The process described in Chapter 7.5 was used as a basis for the simulation case. A propane refrigerated cooler was installed downstream of the absorption column. Even after the absorption, the gas still contains small amounts of water. Some of this water condenses after and propane cooler, and is separated out with the liquid hydrocarbons in the following scrubber. This water will be recycled back in the process, and eventually some of it will end up upstream of the absorption column. This means the column will have to remove more water from the gas in this simulation, than it had to in the simulation described in Chapter 7.5. To do this, the amount of TEG will have to be increased. As in the simulations in Chapter 7.5, an amount of approximately 25 liters TEG per kilogram water needed to be removed have been used in the simulations in this chapter too [6], [9].

A detailed description of the refrigerant cycle can be found in Appendix C. A detailed description and schematic representation of the absorption and TEG regeneration process can be found in Appendix D. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

8.5.1 Export in Statpipe

The HYSYS simulation file used in Chapter 8.5.1 is named *FieldA_TEG_refrig_Statpipe.*

The simulation show that if the absorber is fed with 200 m^3/d of TEG at 65°C and 44,5 barg and contains 12 theoretical contact plates, the gas will be sufficiently dehydrated. Approximately 0,60 m^3/d of TEG will be lost, and will have to be added in a makeup TEG stream. To condition the gas sufficiently, the propane cooler downstream of the absorber will have to cool the gas to -12°C. Since the gas has already been dehydrated, this will not cause any hydrate formation problems. But it will cause a relatively large recycle stream, and some of the lighter hydrocarbons in this stream will eventually end up being recompressed and sent to the absorber again (it is with these hydrocarbons the recycled water follows), so the power demand for the process will increase. Table 8-3 shows a comparison of the demands for power, heating and cooling between the basic process with no dehydration or conditioning, and the process using glycol absorption for dehydrating and a propane cooler for conditioning for export in Statpipe.

Table 8-3: Comparison of power demand by using glycol dehydration and propane cooler conditioning for export in Statpipe

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	57942	149137	251339

The table shows a relatively large increase in the power demand to a total demand of approximately 58 MW. This is probably too high for this to be a profitable development alternative for a marginal field like Field A. The demands for heating and cooling also increase substantially.

Due to the high demands for power, heating and cooling, it is very unlikely this will be a possible development alternative for Field A, even though all hydrate problems have been removed [9].

8.5.2 Export in Vesterled

The HYSYS simulation file used in Chapter 8.5.2 is named *FieldA_TEG_refrig_Vesterled.*

The simulation show that if the absorber is fed with 200 m^3/d of TEG at 65°C and 44,5 barg and contains 10 theoretical contact plates, the gas will be sufficiently dehydrated. Since the specifications for export in Vesterled are less strict than the specifications for Statpipe, two theoretical plates are saved. Approximately 0,60 m^3/d of TEG will be lost, and will have to be added in a makeup TEG stream. To condition the gas sufficiently, the propane cooler downstream of the absorber will have to cool the gas to -11°C. No hydrate formation problems will occur, but the recycle streams will be relatively high. Table 8-4 shows a comparison of the demands for power, heating and cooling between the basic process with no dehydration or conditioning, and the process using glycol absorption for dehydrating and a propane cooler for conditioning for export in Vesterled.

Table 8-4: Comparison of power demand by using glycol dehydration and propane cooler conditioning for export in Vesterled

	Power demand [kW]	Heating demand [kW]	Cooling demand [kW]
No dehydration/conditioning	12060	1953	29563
Dehydration by turbo expander and propane cooler	58986	152661	255640

The table shows a relatively large increase in the power demand to a total demand of approximately 59 MW. This is probably too high for this to be a profitable development alternative for a marginal field like Field A. The demands for heating and cooling also increase substantially [9].

9. Evaluation of solutions in terms of weight, costs and complexity

Both for export in Statpipe and Vesterled, five different development alternatives have been simulated for dehydration and conditioning of the produced gas from Field A. These alternatives will be evaluated in the following sub chapters.

Neither a **Joule-Thomson valve process** nor a **turbo expander process** is able to both dehydrate and condition the gas adequately, so these two alternatives cannot be implemented for the development of Field A. Therefore, they will not be evaluated any further.

9.1 Basis for calculating weight, area and costs

When evaluating the development alternatives, the basic process with no dehydration and conditioning described in Chapter 6 is used as a base case. The increases in weight, costs and area needed for the process plant for the different dehydration and conditioning alternatives are evaluated based on the extra equipment needed. The values in Table 9-1 are used in the evaluation.

Table 9-1: Weight, footprint and investment costs for process equipment [9], [11], [12]

Equipment	Weight [kg]	Total footprint [m2]	Cost [MUSD]
Separator	23000	10	0,28
Pump	25000	N/A	1,40
Heater	220	5	0,03
Cooler	15000	20	0,45
Compressor	85000	110	12,70
Scrubber	7000	9	0,11
Condenser	42000	21	0,43
Re-boiler	3000	10	0,12
Turbo expander	80000	105	13,50
TEG contactor	15000	8	0,20
TEG regeneration cycle	40000	64	2,00

These values are assumed to be fair estimates, based on previous projects and developments in the North Sea, and information from different suppliers [9], [11], [12].

Increase in the number of gas turbines needed for power production will also result in increased weight, costs and area needed, but there are still uncertainties of how power will be supplied to the Field A process. This thesis does not focus on the supply of power, so no weight, investment costs or footprint values for gas turbines have been implemented in the calculations. However, an estimate of the number of turbines needed to supply the power is evaluated, based on a GE LM2500 gas turbine, able to supply 25 MW [13].

Costs based on the increase in fuel gas needed to supply the power are evaluated. It is assumed 0,0276 Sm^3 of fuel gas is needed for 1 MJ of power, which is based on the fuel consumption for a GE LM2500 gas turbine [13]. The costs of increased power demand are evaluated as loss in gas sales, with a gas price of 0,3 USD/Sm^3 assumed [1]. A CO_2 tax of 0,125 USD/Sm^3 of burned gas in Norway is also implemented in the calculations [14].

9.1.1 Basic process with no dehydration or conditioning

The weight, area and investment costs for the basic process described in Chapter 6 are shown in Table 9-2.

Table 9-2: Weight, area and costs for the basic Field A process

Number of unit oprations	Increase in weight [kg]	Increase in footprint [m2]	Increased investment costs [MUSD]
3	69000	30	0,84
7	175000	N/A	9,80
3	660	15	0,09
5	75000	100	2,25
4	340000	440	50,80
7	49000	63	0,77
0	0	0	0,00
0	0	0	0,00
0	0	0	0,00
0	0	0	0,00
0	0	0	0,00
Total	708660	648	64,55

Table 9-3 shows the costs due to loss in gas sale and the CO_2-tax in Norway.

Table 9-3: Costs realted to power demand for the basic Field A process

Power demand [kW]	12060
Power demand [MJ/d]	1041984
Compressor gas consumption [Sm3/MJ]	0,0276
Fuel gas demand [Sm3/d]	28759
CO2 Tax [usd/d]	**3595**
Loss of export gas [usd/d]	**8628**

The power demand for the process can be supplied by one gas turbine.

The values in Table 9-2 and Table 9-3 will be used as a basis to which the other development solutions will be compared.

9.1.2 *Propane refrigeration cycle*

The propane refrigeration cycle process is almost identical to the basic process in terms of equipment, apart from one sea water cooler being replaced by a propane cooler. This propane cooler is part of a propane refrigerant cycle, which needs a condenser and a compressor. So the extra equipment needed is one condenser and one compressor. Table 9-4 shows the weight, area needed and investment costs for this development.

Table 9-4: Weight, area and investment costs for a propane refrigeration cycle process

Number of unit oprations	Increase in weight [kg]	Increase in footprint [m2]	Increased investment costs [MUSD]
3	69000	30	0,84
7	175000	N/A	9,80
3	660	15	0,09
5	75000	100	2,25
5	425000	550	63,5
7	49000	63	0,77
1	42000	21	0,43
0	0	0	0,00
0	0	0	0,00
0	0	0	0,00
0	0	0	0,00
Total	835660	779	77,68

The increases in weight and area are not very high, but an increase of more than 13 MUSD, more than 20 %, in investment costs is quite high for a marginal field like Field A.

Also, it was found in Chapter 7.3.2 that if this development is used to dehydrate and condition the gas for export in Vesterled, the total power demand will be approximately 137 MW. Table 9-5 shows the daily costs due to the fuel gas needed.

Table 9-5: Costs related to power demand for export in Vesterled using a refrigerant cycle

Power demand [kW]	136685
Power demand [MJ/d]	11809584
Compressor gas consumption [Sm3/MJ]	0,0276
Fuel gas demand [Sm3/d]	325945
CO2 Tax [USD/d]	**40743**
Loss of export gas [USD/d]	**97783**

An increase in the daily costs of approximately 126000 USD is very high. That is an increase of over 1000 %. For export in Statpipe, the power needed is even higher, resulting in higher daily costs.

The large power demand also means that weight, area needed and investment costs for the gas turbines needed to be installed will be high. Six gas turbines will have to be installed to supply the power.

Also, as mentioned in Chapter 7.3.2, there will be hydrate formation problems downstream of the propane cooler in this process. This will have to be solved by injecting MEG, causing the weight, area needed, investment costs and daily costs to increase even more.

Propane to flow in the refrigerant cycle will also be needed. The cooling demand is high, and the simulation show that approximately 280 kg/s of refrigerant propane is needed. Some refrigerant will be lost, but most of it will be recycled through the cycle constantly.

Based on the findings in this chapter, it is concluded that dehydration and conditioning using a propane refrigeration cycle at Field A for export in Statpipe or Vesterled, is not a possible development alternative [8], [9].

9.1.3 Combination of turbo expander process and propane refrigeration cycle

As for the propane refrigeration cycle process, this process needs one extra compressor and one extra condenser for the refrigerant cycle. Also, it needs one turbo expander, and one extra scrubber downstream of the expander. Taking the dehydrated and conditioned gas from 0 barg to the export pressure of 125 bara will probably take three compressor stages, and two interstage coolers. Table 9-6 shows the weight, area needed and investment costs for this development.

Table 9-6: Weight, area and investment costs when combining a turbo expander process and a propane refrigeration cycle

Number of unit oprations	Increase in weight [kg]	Increase in footprint [m2]	Increased investment costs [MUSD]
3	69000	30	0,84
7	175000	N/A	9,80
3	660	15	0,09
7	105000	140	3,15
8	680000	880	101,6
8	56000	72	0,88
1	42000	21	0,43
0	0	0	0,00
1	80000	105	13,50
0	0	0	0,00
0	0	0	0,00
Total	1207660	1263	130,29

These values are all very high. The investment costs are approximately doubled, which is high for a marginal field.

The total power demand for the process if the gas is dehydrated and conditioned for export in Vesterled was found to be 19,5 MW in Chapter 8.4.2. This gives the values in Table 9-7 for daily costs related to the power demand.

Table 9-7: Costs related to power demand for export in Vesterled when combining a refrigerant cycle and a turbo expander.

Power demand [kW]	19530
Power demand [MJ/d]	1687392
Compressor gas consumption [Sm3/MJ]	0,0276
Fuel gas demand [Sm3/d]	46572
CO2 Tax [USD/d]	**5822**
Loss of export gas [USD/d]	**13972**

The power needed for export in Statpipe is 20 MW, so the daily costs will be slightly higher for that alternative.

The daily costs for this development alternative increase by approximately 62 %. This is quite high, but may not be enough to make the solution not profitable. But combined with the increases in weight, area needed and investment costs they probably are, even though the power can be supplied by only one gas turbine. Also, as mentioned in Chapter 8.4.2, equipment for MEG injection have to be installed both downstream of the propane cooler and downstream of the turbo expander in order to avoid hydrate formation, further increasing the weight, area needed and investment costs.

Approximately 57 kg/s of refrigerant propane is needed in the refrigerant cycle. Some refrigerant will be lost, but most of it will be recycled through the cycle constantly.

It is therefore concluded that dehydration and conditioning by combining a turbo expander process and a propane refrigeration cycle at Field A for export in Statpipe or Vesterled, is not a possible development alternative [8], [9].

9.1.4 Dehydration by glycol absorption and conditioning by a refrigerant cycle

The refrigeration cycle in this process needs one extra compressor and one extra condenser. The process also needs one TEG contactor and one TEG regeneration cycle. In the TEG regeneration cycle, one re-boiler and one condenser need to be installed. Table 9-8 shows the weight, area needed and investment costs for the development.

Table 9-8: Weight, area and investment costs when using glycol absorption for dehydration and a refrigerant cycle for conditioning

Number of unit oprations	Increase in weight [kg]	Increase in footprint [m2]	Increased investment costs [MUSD]
3	69000	30	0,84
7	175000	N/A	9,80
3	660	15	0,09
5	75000	100	2,25
5	425000	550	63,50
7	49000	63	0,77
2	84000	42	0,86
1	3000	10	0,12
0	0	0	0,00
1	15000	8	0,20
1	40000	64	2,00
Total	935660	882	80,43

The increases in weight and area are not very high, but an increase of approximately 25 %, almost 16 MUSD, in the investment costs is quite high for a marginal field like Field A.

The total power demand when using this development for export in Statpipe was found to be approximately 58 MW in Chapter 8.5.1. This gives the values in Table 9-9 daily cosst related to the power demand.

Table 9-9: Increases in costs related to power demand for export in Statpipe

Power demand [kW]	57942
Power demand [MJ/d]	5006189
Compressor gas consumption [Sm3/MJ]	0,0276
Fuel gas demand [Sm3/d]	138171
CO2 Tax [USD/d]	**17271**
Loss of export gas [USD/d]	**41451**

The power demand for export in Vesterled is slightly higher, but approximately the same.

An increase in daily costs of over 46000 USD, approximately 380 %, is high for a marginal field, probably erasing the profitability of the development alternative. Three gas turbines will have to be installed to supply the power.

Approximately 170 kg/s of refrigerant propane is needed in the refrigeration cycle. Some refrigerant will be lost, but most of it will be recycled through the cycle constantly.

The increase in investment costs, daily costs and demands for weight and area leads to the conclusion that dehydration using glycol absorption and conditioning using a refrigerant cycle at Field A for export in Statpipe or Vesterled, is not a possible development alternative [8], [9].

9.2 Dry gas export from Field A

Based on the results in this thesis, it seems dry gas export from Field A is economically impossible, and that wet gas export to Field B is the only solution. The Field B process will be investigated in Chapter 10.

The problem seems to be the high propane content in the production fluids (C3 in Table 2-1), combined with the fact that Field A does not have an export line for condensate products. If both the stabilization criteria of the crude oil and the hydrocarbon dew point specification of the dry export gas shall be fulfilled, there will be a lot of excess propane in the process, which will be recycled multiple times, causing large demands for power, heating and cooling [8].

One solution to this problem can be to bleed off some propane in the process, to stop it from recycling over and over again. For this to be a possible development alternative, the propane being bled off have got to be usable for something [8], [12]. This development alternative will be briefly discussed in Chapter 11.1.

Another solution may be to export the gas in one of the rich gas export pipelines in the area [9]. This alternative will be briefly discussed in Chapter 11.2.

10. Wet gas export to Field B

The findings in this thesis so far indicate that dehydration and conditioning of the produced natural gas at Field A for export in Statpipe or Vesterled may be theoretically possible, but is impossible in practice, for economical reasons. Therefore, the base case of wet gas export to Field B seems to be the solution. But there are uncertainties concerning the processing suitability at Field B. A simulation of the Field B process has been performed in HYSYS, to check if the gas will be adequately dehydrated and conditioned.

10.1 The Field B process

Field B is operated by a different company, hereby named the operating company. The field receives wet gas from many different satellite fields. These gas streams are comingled, and then dehydrated and conditioned in the Field B process. Information supplied by the operating company regarding the different gas streams and details about the process is limited. The only information available is a simplified process flow diagram, revealing that Field B uses a Joule-Thomson valve process for dehydration and conditioning. Also, the inlet temperature and pressure of the Field A gas can be assumed quite accurate based on the distance between the two fields [15].

The solution in this thesis has been to build a simulation model based on the process flow diagram supplied by the operating company. Only the part of the process that handles the export gas from Field A has been simulated. That means that processing of the condensate separated from the gas is not simulated. The feed gas in the simulation consists only of the wet export gas from Field A. This stream has been retrieved from the simulation of the basic process described in Chapter 6. Temperatures and pressures in the process have been assumed, based on normal values in the industry, and some of them have been varied in an attempt for the gas to reach the export specifications. Dry gas export from Field B can be done in Statpipe or Vesterled, so the specifications listed in Table 3-1 and Table 3-2 yields also for the Field B process [9], [15].

Figure 10-1 shows a schematic representation of the Field B process as it has been simulated in this thesis. The pressures and temperatures indicated in the figure have been assumed. The real values at Field B may differ somewhat from these [9], [15]. A detailed process flow diagram of the simulation from HYSYS can be found in Appendix E. The HYSYS simulation file used in Chapter 10 is named *FieldB.*

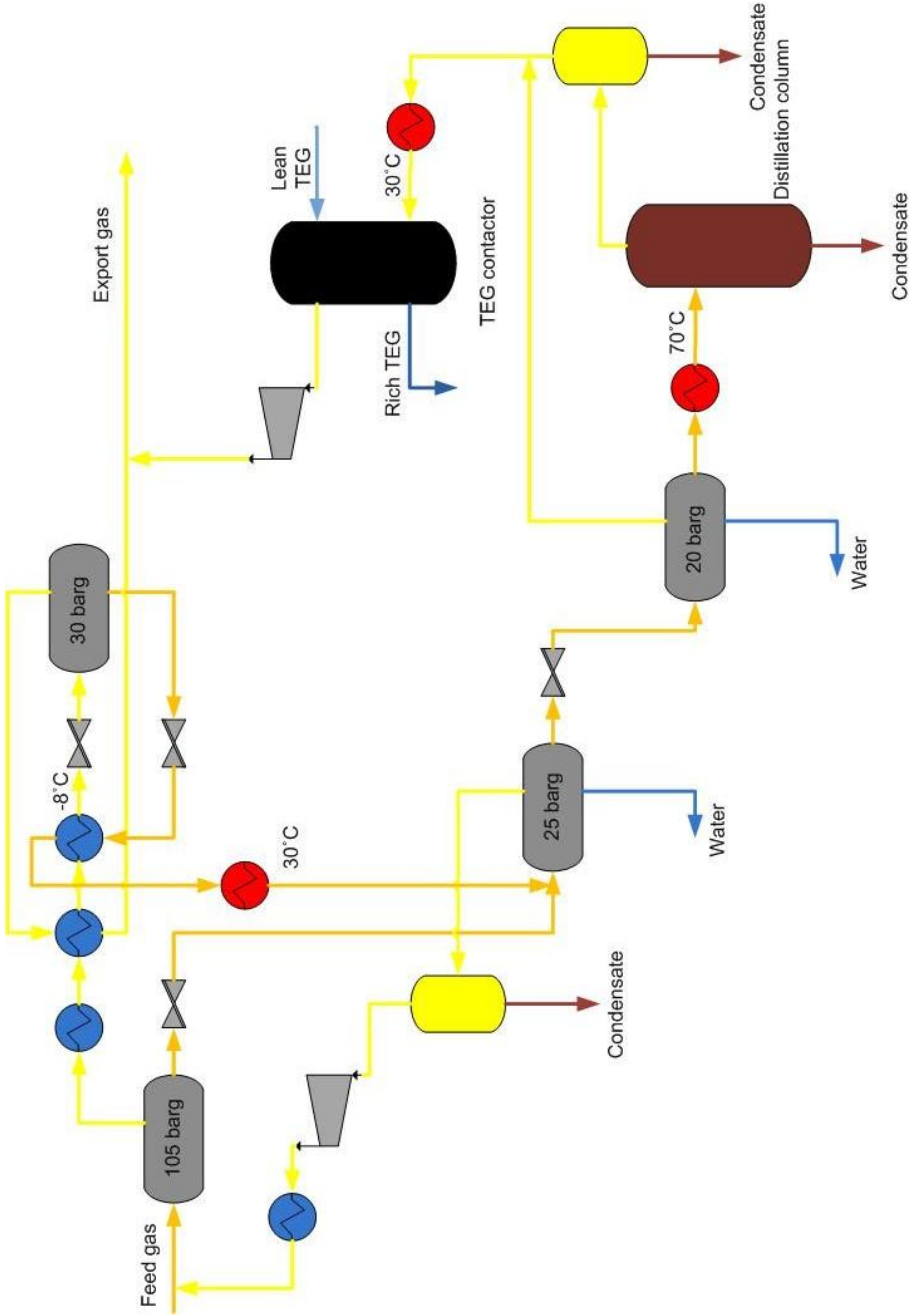

Figure 10-1: Field B dehydration and conditioning process

The simulation shows that the gas will reach the export specifications, both for Statpipe and Vesterled. So it is theoretically possible to dehydrate and condition the export gas from Field A at Field B. However, if the actual operating temperatures and pressures at Field B differ much from the values that has been assumed in the simulation, the process may not be able to dehydrate and condition the gas adequately [8], [9].

Also, because the simulation performed in this thesis only focuses on the dry gas, it is difficult to evaluate the process suitability. Specifications for the condensate export from Field B have not been available information during the work with this thesis. Therefore, the condensate distillation column has been specified only so that the gas will be properly conditioned, without any restrictions based on condensate specifications. Due to the high propane content of the Field A gas, it is very likely that it will be difficult to reach both the hydrocarbon dew point specification of the dry gas and the specifications of stable condensate. This will have to be evaluated by the operating company, who has all the information needed for a proper evaluation [8].

If wet gas export from Field A to Field B is going to work, it is probably depending on the propane content of the other feed gas streams at Field B. If the propane content in these streams is low enough, the Field A gas can be thinned out when comingled with them, making the final inlet stream at Field B suited for the process [8].

11. Alternative solutions for the Field A gas protraction

Two alternative solutions for the gas protraction from Field A have been briefly studied.

11.1 Propane bleed off

As mentioned in Chapter 9.2, one solution to the problems with high demands for power when exporting dry gas from Field A in Statpipe or Vesterled could be to bleed off some propane in the process, to avoid it from recycling multiple times in the process. A simulation has been performed, to check the impact of this [8], [12].

The process described in Chapter 8.5.2, with dehydration by glycol absorption and conditioning by a propane refrigerated cooler, has been used as the basis. Propane is bled off after the scrubber downstream of the propane refrigerated cooler [12]. A detailed process flow diagram of the process as it has been simulated in HYSYS can be found in Appendix E.

In Chapter 9.1.4 it was found that both the investment costs and the daily costs for a process combining absorption and a refrigerant cycle were high. The investment costs will not decrease if propane is bled off. They will rather increase, as some equipment for the bleeding will have to be installed. But the daily costs could decrease quite much, as the power demand goes down.

The HYSYS simulation file used in Chapter 11.1 is named *FieldA_TEG_refrig_C3Bleed_Vesterled.*

By bleeding off 2 kg/s of propane, the power demand is decreased to approximately 15 MW. The gas will still be adequately dehydrated and conditioned for export in Vesterled. With such a low power demand, this could be a possible development alternative, but the propane being bleed off has to be useful for something, i.e. fuel or heating/cooling. More work should be put into this, to check if the solution is satisfactory [8].

11.2 Rich gas export in FUKA or Sage

As mentioned in Chapter 9.2, one alternative for the gas protraction at Field A is rich gas export. This could be done in FUKA or Sage, which are rich gas pipelines in the area [4], [5], [9].

The basic process described in Chapter 6 has been used as a basis. The condition of the export gas from this process has been checked, and compared to the export specifications in FUKA and Sage (Chapters 3.1.3and 3.1.4).

The simulation show that the cricondenbar of the gas is 104,8 barg, which is below the specification of 106 bara for FUKA. The water content of the gas however, is approximately 460 kg/Sm^3, which is much higher than the FUKA specification of maximum 24 kg/Sm^3. So if the gas is dehydrated, i.e. by glycol absorption, it could be exported in FUKA. More work should be put into this possible solution.

The specification of 10,67-21,82 mole% of C2-C12 components in Sage, is not met. The export gas at Field A has almost 29 mole% of C2-C12 components. The water content specification in Sage is less than 63 ppm. The Field A gas has a water content of 630 ppm. This means the gas both has to be dehydrated and conditioned if it is to be exported in Sage, making it a less possible solution than export in FUKA.

It seems export in FUKA is the best option for rich gas export from Field A. Then the gas will not have to be conditioned, although it will have to be dehydrated. Further studies should be put into this development alternative, before it is chosen as the solution for the gas protraction at Field A.

12. Discussion

The main objective of this thesis has been to build up knowledge and understanding of gas drying and conditioning processes on offshore production facilities. The base case for the gas produced at Field A is wet gas export to Field B, but due to uncertainties concerning the processing suitability at Field B, simulations has been performed in HYSYS to check if dehydration and conditioning can be implemented in the Field A process. If this is possible, dry gas export directly from Field A can be achieved.

In the Field A area, Statpipe and Vesterled are the existing dry gas export pipelines. Different methods to dehydrate and condition the gas have been simulated in an attempt for the gas to reach the specifications in the pipelines.

Dehydration of the gas does not cause any serious problems. It was found that the best way to dehydrate the gas is to use a TEG absorption column. This does not cause any problems with hydrate formation, and does not increase the power demand for the process very much, although an increase in weight, area needed and investment costs will follow.

It was found that the main problem concerning dry gas export from Field A is to condition the gas. The production fluids contain relatively large amounts of propane, which makes it difficult to reach both the crude oil stabilization criteria and the hydrocarbon dew point specification of the export gas. If too much propane is present in the export gas, it will not be lean enough for export in Statpipe or Vesterled. To obtain the right split of propane the crude oil and the export gas by using conventional conditioning methods, the gas will have to be recycled multiple times in the process, causing very high demands for power. As Field A is a marginal field, it will not be economically profitable to develop it if the power demand is too high.

No possible development alternatives for Field A that will both dehydrate and condition the gas, and still be economically profitable, was found in this thesis. It seems wet gas export to Field B is the only profitable solution. But Field B is operated by another company, and information about the Field B process is limited. Although a simulation of the Field B process in this thesis indicates that it is theoretically possible to dehydrate and condition the Field A gas there, it depends on the actual operating conditions and specifications in the process. Information about this has not been available from the operating company of Field B.

If it is found that the Field B process cannot dehydrate and condition the Field A gas adequately for dry gas export in Statpipe or Vesterled, another solution for the gas protraction at Field A has to be identified.

One solution may be to dehydrate and condition the gas at Field A, while bleeding of propane in the process. If the hydrocarbons being bled off can be used for something useful, i.e. as fuel to generate power to the process, this could be a good solution.

Another possible solution may be to wet gas export from Field A in a wet gas export pipeline. Possible wet gas export pipelines in the Field A area are FUKA and Sage. It seems FUKA is the best option, as the Field A gas only needs to be dehydrated before export in this pipeline.

12.1 Uncertainty

The main bases of this thesis are simulations performed in HYSYS. Even though HYSYS is considered to be one of the leading process simulation software on the market, a simulation can never replace real life. Simplifications and assumptions have to be made, which will lead to uncertainty. For example the composition and thermodynamic parameters of the production fluids can vary, the efficiencies of process equipment may differ, etc.

Peng-Robinson has been used as the equation of state in this thesis. This is considered to be the best one available for oil and gas processes. But in some of the simulations TEG absorption is implemented, and for these situations Peng-Robinson may give some less accurate results. There exists a glycol-package equation of state for HYSYS, which is recommended to use for glycol absorption. But during the work with this thesis, some problems were experienced concerning the license of this package, so simulations with it could not be performed properly. Therefore, Peng-Robinson was used also in these simulations, which may lead to increased uncertainty.

Table 9-1 contains a list of weight, area and investment costs for process equipment. The actual values for this is dependent on the specific project the equipment is used in, as duties, volume- and mass flows, compositions etc. may vary from project to project. The values selected in this thesis are considered to be good estimates, based on previous developments in the North Sea and available information from different suppliers.

13. Conclusions

In this thesis process simulations have been implemented in HYSYS, to check if the produced gas from Field A can be dehydrated and conditioned at Field A for dry gas export in Statpipe or Vesterled. The main conclusion of the thesis is that this cannot be done economically profitable.

A **Joule-Thomson valve process** or a **turbo expander process** will not dehydrate and/or condition the gas adequately for dry gas export.

A **refrigerant cycle process**, using propane as the refrigerant, can theoretically dehydrate and condition the gas adequately. But this process will have a very high power demand, approximately 137 MW for export in Vesterled and approximately 170 MW for export in Statpipe. This will result in very high daily costs related to fuel gas consumption, erasing the profitability of the development alternative. There will also be problems with relatively high investment costs and hydrate formation in the process.

It is also possible to dehydrate and condition the gas properly by **combining a refrigerant cycle process and a turbo expander process**, but this too will not be economically profitable. Although the power demand for the process only will be approximately 19,5 MW for export in Vesterled and 20 MW for export in Statpipe, the investment costs will be doubled compared to the basic process without dehydration and conditioning. There will also be hydrate formation problems in the process.

The most common way to dehydrate natural gas in the industry is by glycol absorption. This will remove water from the gas at high temperatures, thereby removing problems with hydrate formation. It was shown that the gas could be adequately dehydrated and conditioned when **dehydrated by glycol absorption and conditioned in a refrigerant cycle process**, without experiencing any hydrate formation problems in the process. But the power demand for this process is approximately 58 MW, resulting in high daily costs related to fuel gas consumption. In combination with relatively high investment costs, this causes the process not to be economically profitable.

This thesis concludes with a recommendation that wet gas export to Field B is the best development alternative for the gas protraction at Field A. However, the process suitability at Field B for the Field A gas should be verified by the operating company of Field B, as limited information about Field B has been available during the work with this thesis.

14. Suggestions for further studies

The conclusion of this thesis is that wet gas export to Field B is the best solution for the gas protraction from Field A. However, due to limited information about the Field B process, work should be put into verifying this. Either the operating company of Field B should receive all the information needed about Field A to verify this, or Det norske could receive all the information needed about Field B.

In Chapter 11.1, an alternative solution for the Field A process resulting in adequately dehydrated and conditioned gas was suggested. By bleeding off propane in the process, the power demand was lowered to an acceptable level. More work should be put into investigating the complexity and profitability of this process. The main challenge is to find out if the propane being bled off can be useful for something, like for power production or heating/cooling.

As mentioned in Chapter 3.1, there exist rich gas export pipelines in the Field A area. If dehydration and conditioning of the gas at Field A is found impossible, and the process Field B does not suit the Field A gas, it should be investigated if rich gas export in one of these pipelines is a possible solution. This is briefly discussed in Chapter 11.2. It seems export in FUKA suits the Field A gas best. The gas has to be dehydrated before export in FUKA, but it looks like it will not have to be conditioned.

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Appendices

Appendix A - Equation of state [16]

In physics and thermodynamics, an equation of state is a mathematical relation between state variables, for example temperature, pressure or specific volume. The equation describes the state of matter under certain given physical conditions. It exist several equations of state, the most famous one presumably being *the ideal gas law.* This equation of state is roughly accurate for gases at low pressures and moderate temperatures, but will not be accurate enough for more complex systems. The cause of this is mainly that it neglects the size of the molecules, and interactions between molecules.

An offshore process plant handling oil, gas and water at a wide range of temperatures and pressures is an example of a complex system that needs a more accurate equation of state. The two equations of state currently seen as the best in the petroleum industry are *Peng-Robinson* and *Soave-Redlich-Kwong*.

In this thesis Peng-Robinson is used as the equation of state. It is formulated like this

$$p = \frac{RT}{V_m - b} - \frac{a\alpha}{V_m^2 + 2bV_m - b^2}$$

$$a = \frac{0.45724R^2T_c^2}{p_c}$$

$$b = \frac{0.07780RT_c}{p_c}$$

$$\alpha = (1 + (0.37464 + 1.54226\omega - 0.26992\omega^2)(1 - T_r^{0.5}))^2$$

$$\omega = -log_{10}(p_r^{sat}) \text{ at } T_r = 0.7$$

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

$$p_r^{sat} = \frac{p}{p_c}$$

Here **p** is pressure, **R** is the universal gas constant, **T** is absolute temperature, $\mathbf{V_m}$ is molar volume, $\mathbf{T_c}$ is absolute temperature at the critical point, $\mathbf{p_c}$ is pressure at the critical point, **ω** is the acentric factor, $\mathbf{T_r}$ is the reduced temperature and $\mathbf{p_r^{sat}}$ is the reduced pressure.

Peng-Robinson equation of state was developed in 1976 in order to satisfy the following goals:

- The parameters should be expressible in terms of critical properties and the acentric factor.
- The equation should provide reasonable accuracy near the critical point, particularly for calculations of the compressibility factor and liquid densities
- The mixing rules should not employ more than a single binary interaction parameter; which should be independent of temperature, pressure and composition.
- The equation should be applicable in all calculations of all fluid properties in natural gas processes.

Peng-Robinson's performance is for the most similar to Soave-Redlich-Kwong, but it is superior in calculating liquid densities for many materials, especially non-polar ones. This is the main reason why it has been chosen over Soave-Redlich-Kwong in this thesis.

Appendix B - Procedure for naming process equipment and material streams [18]

The process equipment have been named and tagged based on NORSOK standard P-100. The process systems used in this thesis and their corresponding NORSOK system number are:

- Separation and stabilization, system 20
- Low pressure gas compression, system 23
- High pressure gas compression, system 26
- Gas lift, system 27
- Water injection, system 29
- Oily water treatment, system 44
- Sea water, system 50

Different types of equipment have different tags:

- Separator – VA
- Electrostatic coalescer – VJ
- Scrubber/Knock-out drum – VG
- Degassing tank – VD
- Pump – PA
- Compressor – KA
- Turbo expander - TE
- Heater – HA
- Cooler – HB

One piece of equipment is named with a tag; xx-yy-zzz, where xx represents the system it is a part of, yy represents what type of equipment it is and zzz is a number to separate different equipment of the same type in the same system. For example 20-VA-002 is one of the separators in the separation and stabilization system.

The material streams are named after the equipment they are connected to. They are named with a tag; xx-yy-zzz-Nw, where xx-yy-zzz points to a piece of equipment, and Nw tells if it is an inlet or outlet stream. N1 is an inlet stream of the corresponding piece of equipment and N2 is an outlet stream. N3 and N4 are used if the specific piece of equipment has more than one outlet stream.

Some streams and pieces of equipment are named with a single number or a written name. These are either not part of any of the mentioned systems, or not connected to a specific piece of equipment.

Appendix C - Propane refrigeration cycle [10]

In this thesis, when material streams are being cooled to below 30°C, it is assumed sea water cannot be used as cooling medium. Instead the streams are cooled by a refrigerant cycle, using propane as the refrigerant.

A simulation of a propane refrigeration cycle has been developed in HYSYS, and saved as a template. This template can then be connected to other simulation files. Figure C-1 shows a process flow diagram of the refrigerant cycle.

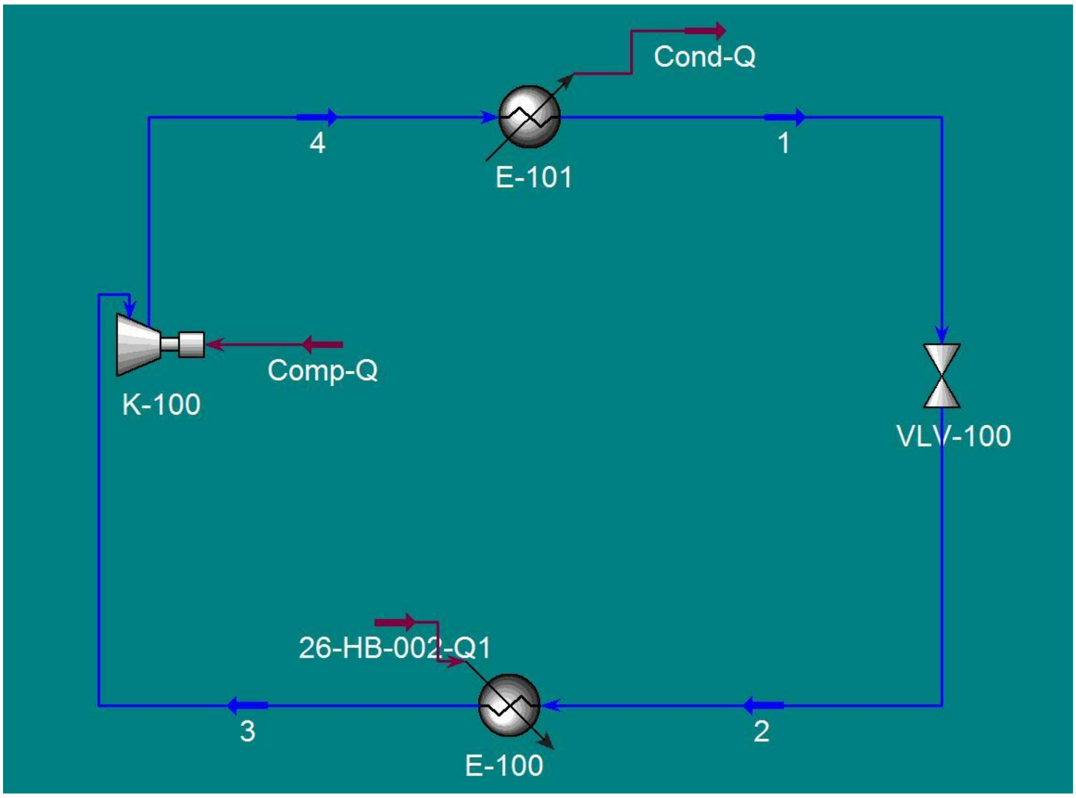


Figure C-1: Refrigerant cycle

For the stream 1, the propane is set at the boiling point at 50°C. This means the pressure is 16,2 barg. For the stream 3, the propane is set at the dew point at -20°C. It is in the cooler E-100 (shown as a heater in Figure C-1, as it heats the propane) the material stream in the original process is cooled by heat exchanging with the propane. The pressure drop on the propane side in the cooler is set at 0,5 bar. This pressure drop, combined with the temperature and dew point specification in stream 3, decides the pressure drop in the valve VLV-100, and thereby the pressure of stream 2. The pressure in stream 2 is 1,9 barg, and the pressure in stream 3 is 1,4 barg. The compressor K-100 compresses the propane, so that the pressure of stream 4 is 16,7 barg. The vapor fraction in stream 4 is 1,0. In the cooler E-101, the propane is cooled to 50°C. The pressure drop in E-101 is set at 0,5 barg.

The mass flow of propane in the cycle is not fixed, and will be the parameter that varies based on the cooling demand in the cooler from the original simulation the template is connected to.

Appendix D - TEG absorption and regeneration system [10]

Figure D-1 shows a schematic representation of the TEG absorption and regeneration cycle used in this thesis.

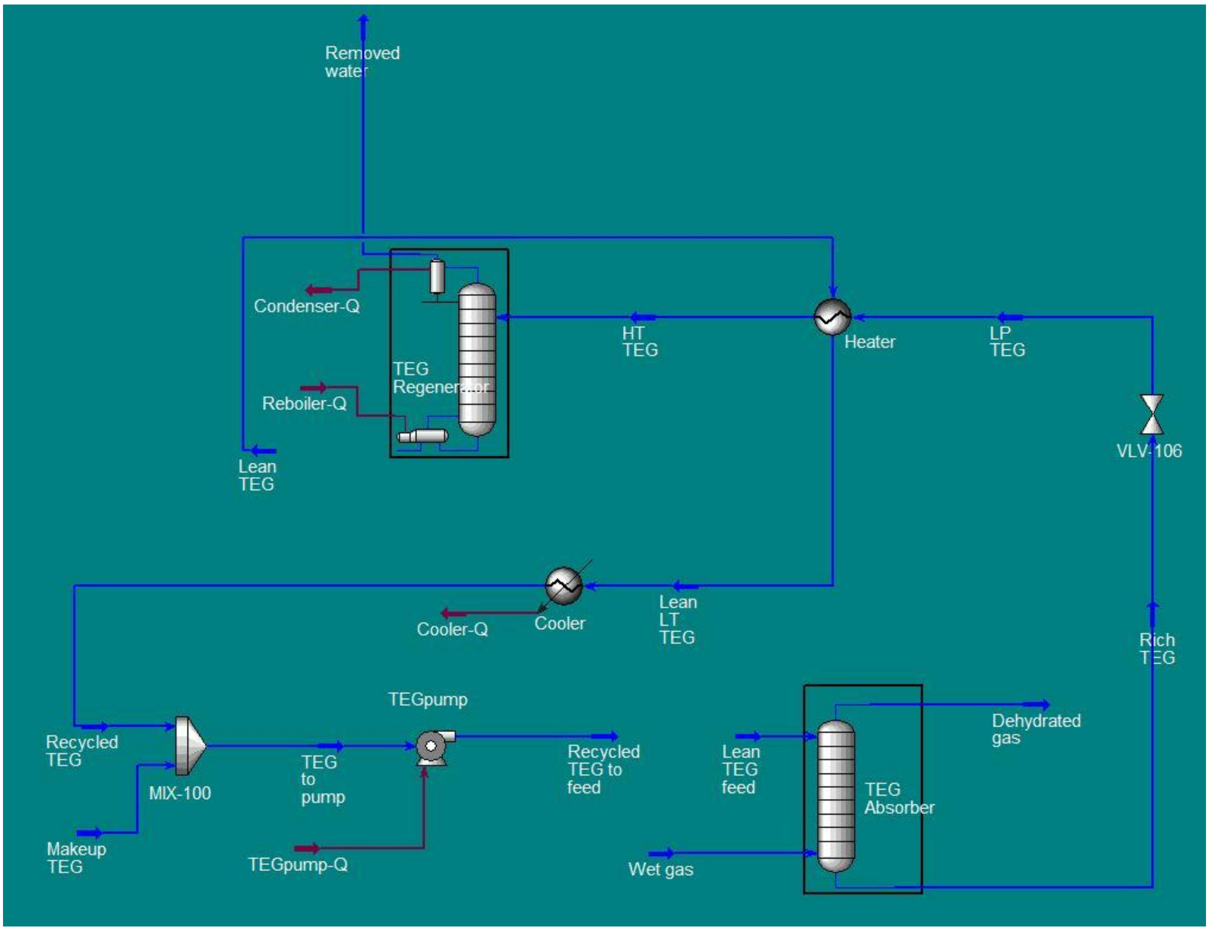


Figure D-1: TEG absorption and regeneration system

TEG Absorber: The HYSYS unit operation *Absorber* is used to model the TEG absorber. The wet gas is fed at the bottom and lean TEG is fed at the top. The dehydrated gas exiting the column carries on in the main process, and the rich TEG exiting the column is sent to the regeneration cycle. The only specification put into the absorber is a pressure drop of 0,5 bar from the bottom to the top. The amount of TEG entering the absorber is varied in different cases, based on the amount of water needed to be removed from the wet gas.

VLV-106: The valve reduces the pressure of the rich TEG to 1,5 barg before the stripper.

Heater: The TEG is heated to 105°C before the stripping, by heat exchanging with the lean TEG exiting the stripper.

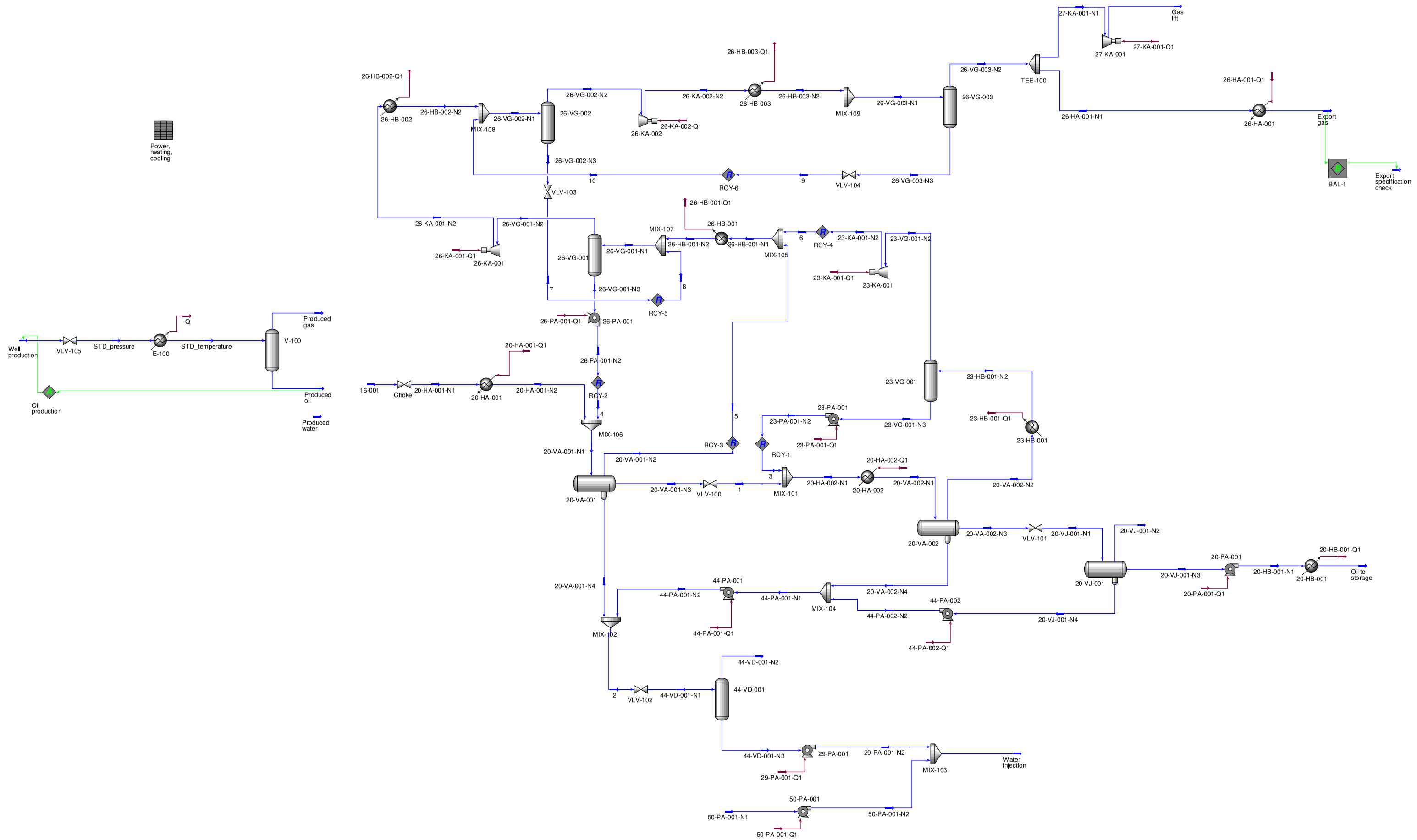
TEG Regenerator: The HYSYS unit operation *Distillation Column* is used to model the TEG regenerator (stripper). The specifications put into the stripper are 100°C at the top, 200°C at the bottom, three theoretical contact plates, a reflux ratio of 1,0 and an overhead vapor flow rate (different from case to case, based on the rich TEG stream entering the stripper, to make the regenerated TEG as lean as possible).

Cooler: The regenerated TEG is cooled to 65°C.

TEGpump: Makeup TEG is added to the regenerated TEG, based on the amount lost in the regeneration. The TEG is pumped to the same pressure as the wet gas entering the absorber, and is then sent back to the absorber. In the simulations the cycle is cut here. This is because HYSYS had problems with reaching a stable solution if the regeneration was a connected cycle. It is solved by adjusting the mass flow of the makeup TEG stream so that the streams *Lean TEG feed* and *Recycled TEG to feed* have the same mass flow of TEG.

Appendix E - Process flow diagrams

1. **Field A – Basic process**
2. **Field A – Joule-Thomson valve process**
3. **Field A – Turbo expander process**
4. **Field A – Refrigerant cycle process**
5. **Field A – Combination of refrigerant cycle and turbo expander process**
6. **Field A – Glycol absorption process**
7. **Field A – Combination of glycol absorption process and refrigerant cycle process**
8. **Field B process**
9. **Field B – Combination of glycol absorption process and refrigerant cycle process, with propane bleed off**



Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
Power, heating, cooling
26-HB-002-Q1
26-HB-002
26-HB-002-N2
MIX-108
26-VG-002-N1
26-VG-002
26-VG-002-N2
26-VG-002-N3
VLV-103
26-KA-002
26-KA-002-Q1
26-KA-002-N2
26-HB-003-Q1
26-HB-003
26-HB-003-N2
MIX-109
26-VG-003-N1
26-VG-003
26-VG-003-N2
26-VG-003-N3
VLV-104
9
10
RCY-6
TEE-100
27-KA-001-N1
27-KA-001
27-KA-001-Q1
Gas lift
26-HA-001-N1
26-HA-001-Q1
26-HA-001
Export gas
BAL-1
Export specification check
26-KA-001-N2
26-KA-001-Q1
26-KA-001
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
26-HB-001-N2
26-HB-001-Q1
26-HB-001
26-HB-001-N1
MIX-105
6
RCY-4
23-KA-001-N2
23-KA-001-Q1
23-KA-001
23-VG-001-N2
7
8
RCY-5
26-PA-001-Q1
26-PA-001
26-PA-001-N2
RCY-2
4
MIX-106
16-001
Choke
20-HA-001-N1
20-HA-001-Q1
20-HA-001
20-HA-001-N2
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
1
5
RCY-3
RCY-1
3
MIX-101
20-HA-002-N1
20-HA-002-Q1
20-HA-002
20-VA-002-N1
23-PA-001
23-PA-001-N2
23-PA-001-Q1
23-VG-001-N3
23-VG-001
23-HB-001-N2
23-HB-001-Q1
23-HB-001
20-VA-002-N2
20-VA-002
20-VA-002-N3
VLV-101
20-VJ-001-N1
20-VJ-001-N2
20-VJ-001
20-VJ-001-N3
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001-Q1
20-HB-001
Oil to storage
20-VA-002-N4
20-VJ-001-N4
44-PA-002
44-PA-002-N2
44-PA-002-Q1
MIX-104
44-PA-001-N1
44-PA-001
44-PA-001-N2
44-PA-001-Q1
20-VA-001-N4
MIX-102
2
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
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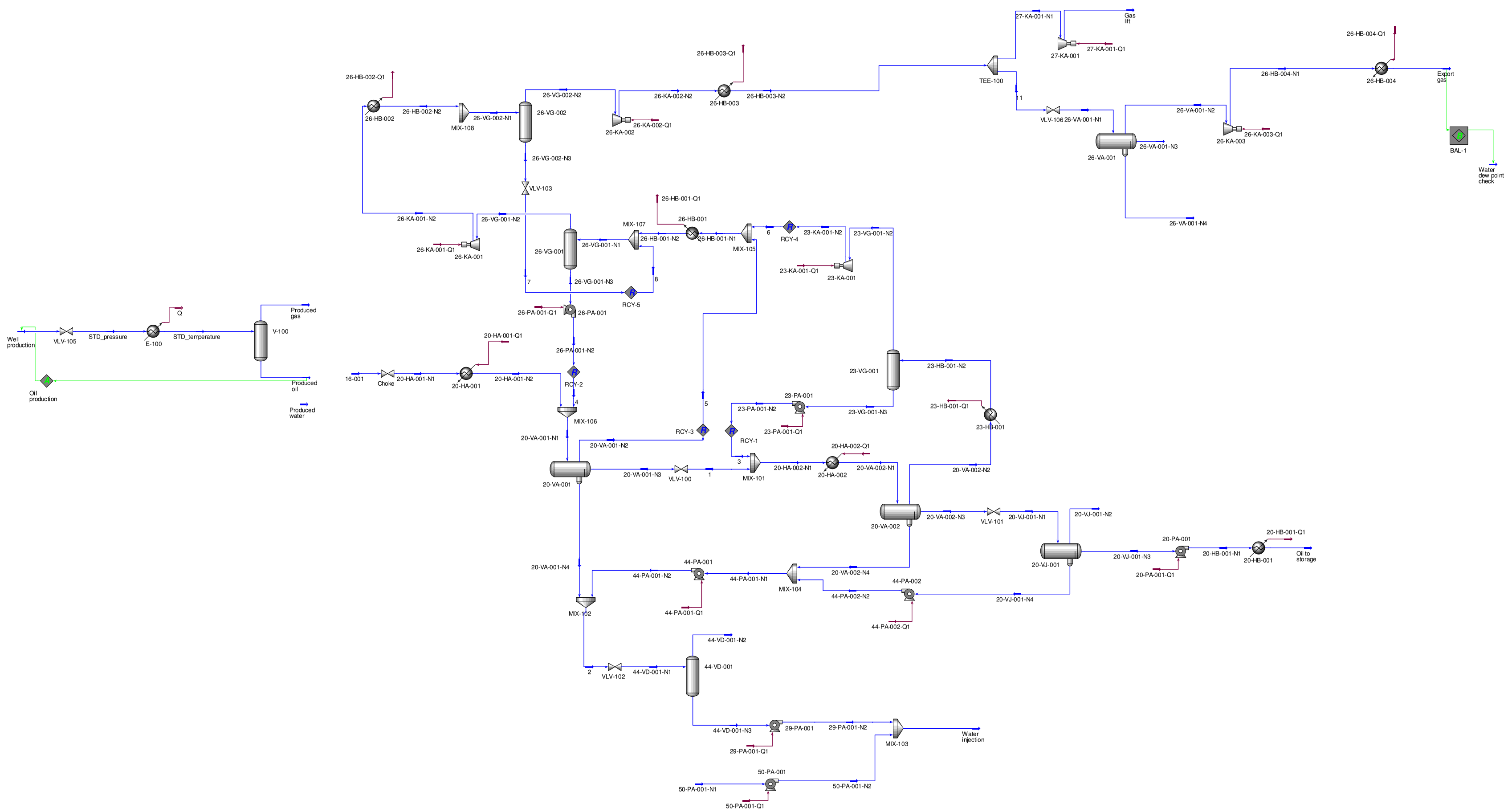
50-PA-001-Q1
50-PA-001-N2


Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
26-HB-002-Q1
26-HB-002
26-HB-002-N2
MIX-108
26-VG-002-N1
26-VG-002
26-VG-002-N2
26-VG-002-N3
VLV-103
26-KA-001-N2
26-KA-001-Q1
26-KA-001
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
26-HB-001-N2
26-HB-001
26-HB-001-Q1
26-HB-001-N1
MIX-105
RCY-5
RCY-4
RCY-3
RCY-2
RCY-1
26-PA-001
26-PA-001-Q1
26-PA-001-N2
MIX-106
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
MIX-101
20-HA-002-N1
20-HA-002
20-HA-002-Q1
20-VA-002-N1
20-VA-002-N2
20-VA-002
20-VA-002-N3
VLV-101
20-VJ-001-N1
20-VJ-001-N2
20-VJ-001
20-VJ-001-N3
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
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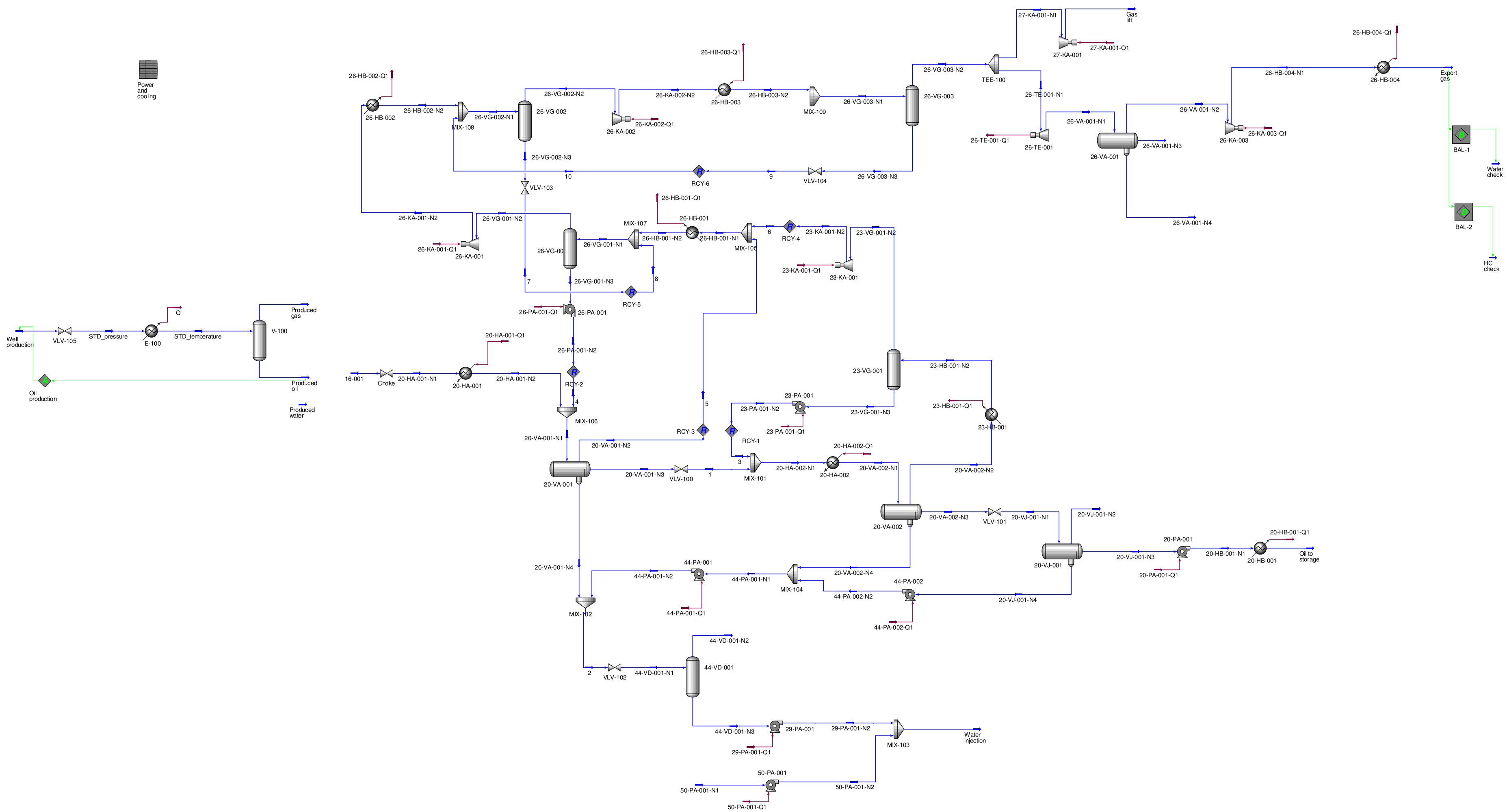
44-PA-002
44-PA-002-Q1
44-PA-002-N2
20-VA-002-N4
MIX-104
44-PA-001-N1
44-PA-001
44-PA-001-Q1
44-PA-001-N2
20-VA-001-N4
MIX-102
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001-N1
50-PA-001
50-PA-001-Q1
50-PA-001-N2
23-VG-001
23-VG-001-N2
23-VG-001-N3
23-KA-001-N2
23-KA-001
23-KA-001-Q1
23-PA-001
23-PA-001-N2
23-PA-001-Q1
23-HB-001-N2
23-HB-001
23-HB-001-Q1
26-KA-002
26-KA-002-Q1
26-KA-002-N2
26-HB-003
26-HB-003-Q1
26-HB-003-N2
TEE-100
27-KA-001-N1
27-KA-001
27-KA-001-Q1
Gas lift
VLV-106
26-VA-001-N1
26-VA-001
26-VA-001-N2
26-VA-001-N3
26-VA-001-N4
26-KA-003
26-KA-003-Q1
26-HB-004-N1
26-HB-004
26-HB-004-Q1
Export gas
BAL-1
Water dew point check


Power and cooling
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STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
26-HB-002-Q1
26-HB-002
26-HB-002-N2
MIX-108
26-VG-002-N1
26-VG-002
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26-VG-002-N3
VLV-103
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26-KA-002-Q1
26-KA-002-N2
26-HB-003
26-HB-003-Q1
26-HB-003-N2
MIX-109
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26-VG-003
26-VG-003-N2
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VLV-104
RCY-6
TEE-100
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Gas lift
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26-TE-001
26-TE-001-Q1
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26-VA-001
26-VA-001-N2
26-VA-001-N3
26-VA-001-N4
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26-KA-003-Q1
26-HB-004-N1
26-HB-004
26-HB-004-Q1
Export gas
BAL-1
BAL-2
Water check
HC check
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26-KA-001
26-KA-001-Q1
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
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MIX-105
RCY-4
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RCY-5
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26-PA-001-N2
RCY-2
MIX-106
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20-VA-001-N4
VLV-100
RCY-3
RCY-1
MIX-101
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VLV-101
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20-VJ-001-N2
20-VJ-001-N3
20-VJ-001-N4
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20-HB-001-N1
20-HB-001
20-HB-001-Q1
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44-PA-001-Q1
MIX-104
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44-PA-002-Q1
MIX-102
VLV-102
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44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001
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50-PA-001-N2
50-PA-001-Q1
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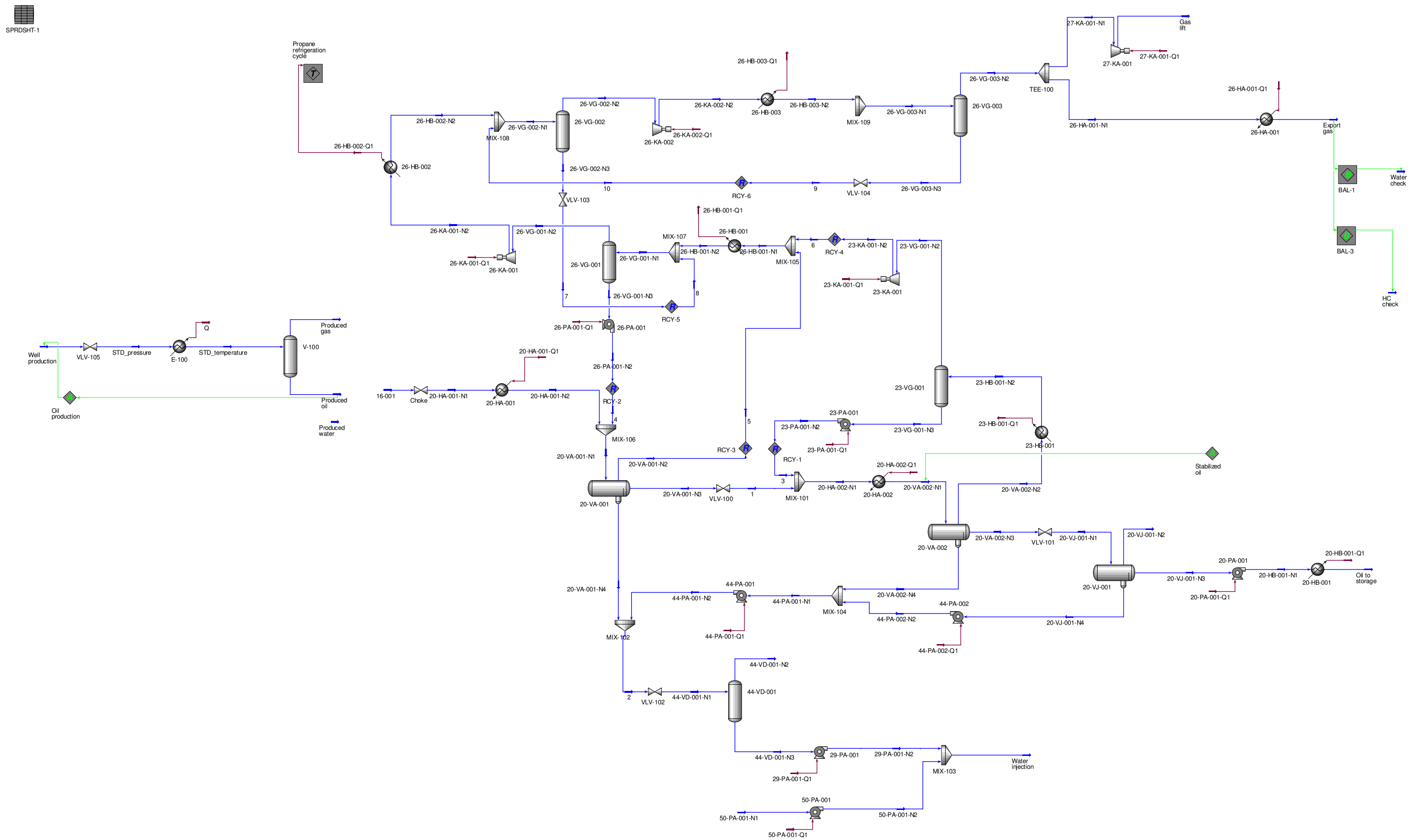
2
3
4
5
6
7
8
9
10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Propane refrigeration cycle
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VLV-103
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26-KA-002
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VLV-104
RCY-6
TEE-100
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27-KA-001-Q1
Gas lift
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Export gas
BAL-1
Water check
BAL-3
HC check
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MIX-107
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MIX-105
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23-VG-001-N2
RCY-5
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26-PA-001-Q1
26-PA-001-N2
RCY-2
MIX-106
Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
RCY-3
RCY-1
MIX-101
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20-VJ-001-N3
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20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
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20-VJ-001-N4
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44-PA-002-N2
44-PA-002-Q1
MIX-104
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44-PA-001
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44-PA-001-Q1
20-VA-001-N4
MIX-102
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
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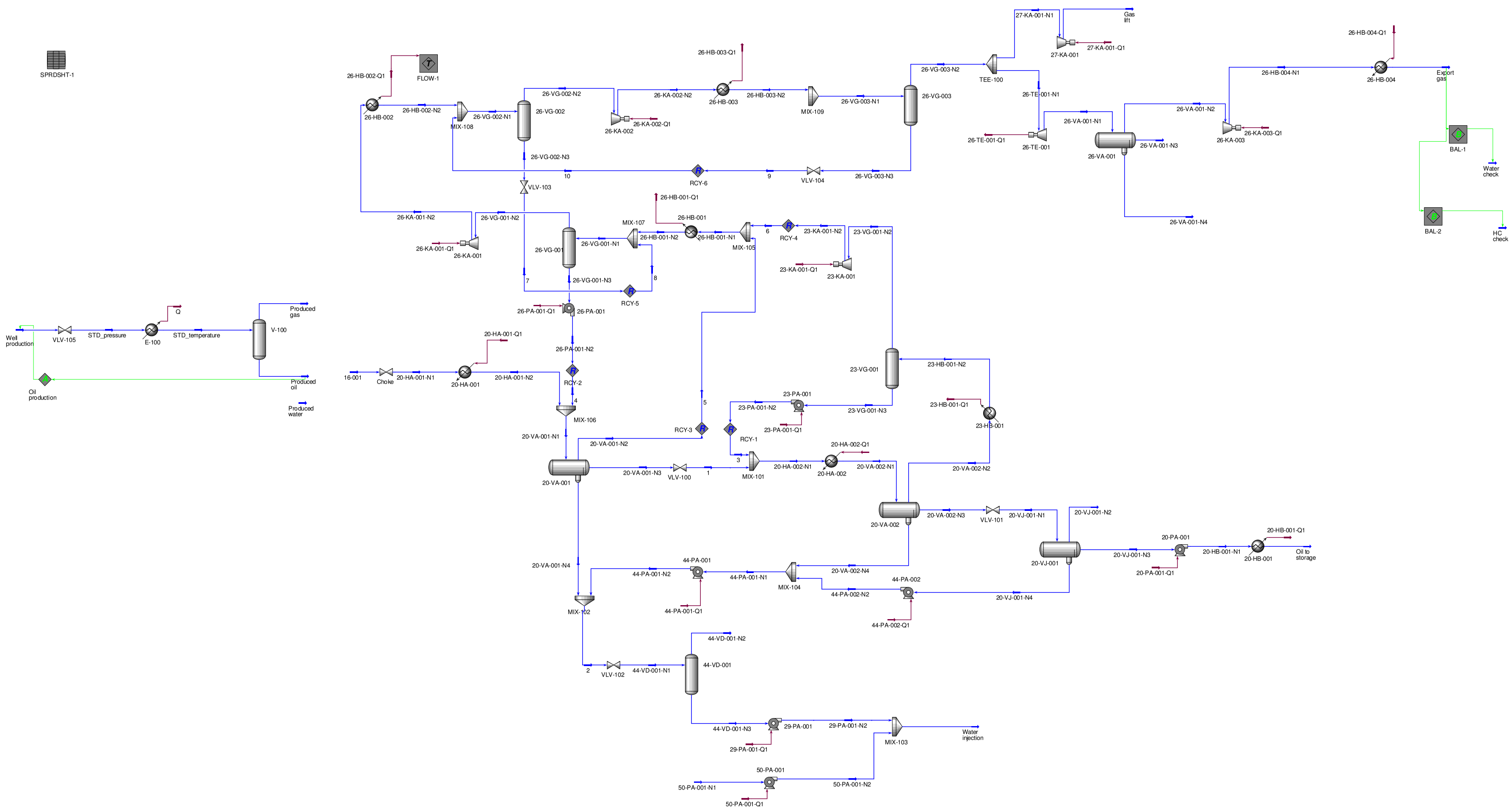
50-PA-001-N1
50-PA-001-N2
50-PA-001-Q1


SPRDSHT-1
Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
26-HB-002-Q1
FLOW-1
26-HB-002
26-HB-002-N2
MIX-108
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26-VG-002
26-VG-002-N2
26-VG-002-N3
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MIX-109
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26-VG-003-N3
VLV-104
RCY-6
VLV-103
TEE-100
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27-KA-001
27-KA-001-Q1
Gas lift
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26-TE-001
26-TE-001-Q1
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26-VA-001
26-VA-001-N2
26-VA-001-N3
26-VA-001-N4
26-KA-003
26-KA-003-Q1
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26-HB-004
26-HB-004-Q1
Export gas
BAL-1
BAL-2
Water check
HC check
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26-KA-001
26-KA-001-Q1
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
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26-HB-001
26-HB-001-Q1
26-HB-001-N1
MIX-105
RCY-4
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23-KA-001
23-KA-001-Q1
23-VG-001-N2
RCY-5
26-PA-001
26-PA-001-Q1
26-PA-001-N2
RCY-2
MIX-106
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
20-VA-001-N1
20-VA-001
20-VA-001-N2
20-VA-001-N3
20-VA-001-N4
VLV-100
RCY-3
RCY-1
MIX-101
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20-HA-002
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20-VA-002-N1
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20-VA-002
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20-VA-002-N4
VLV-101
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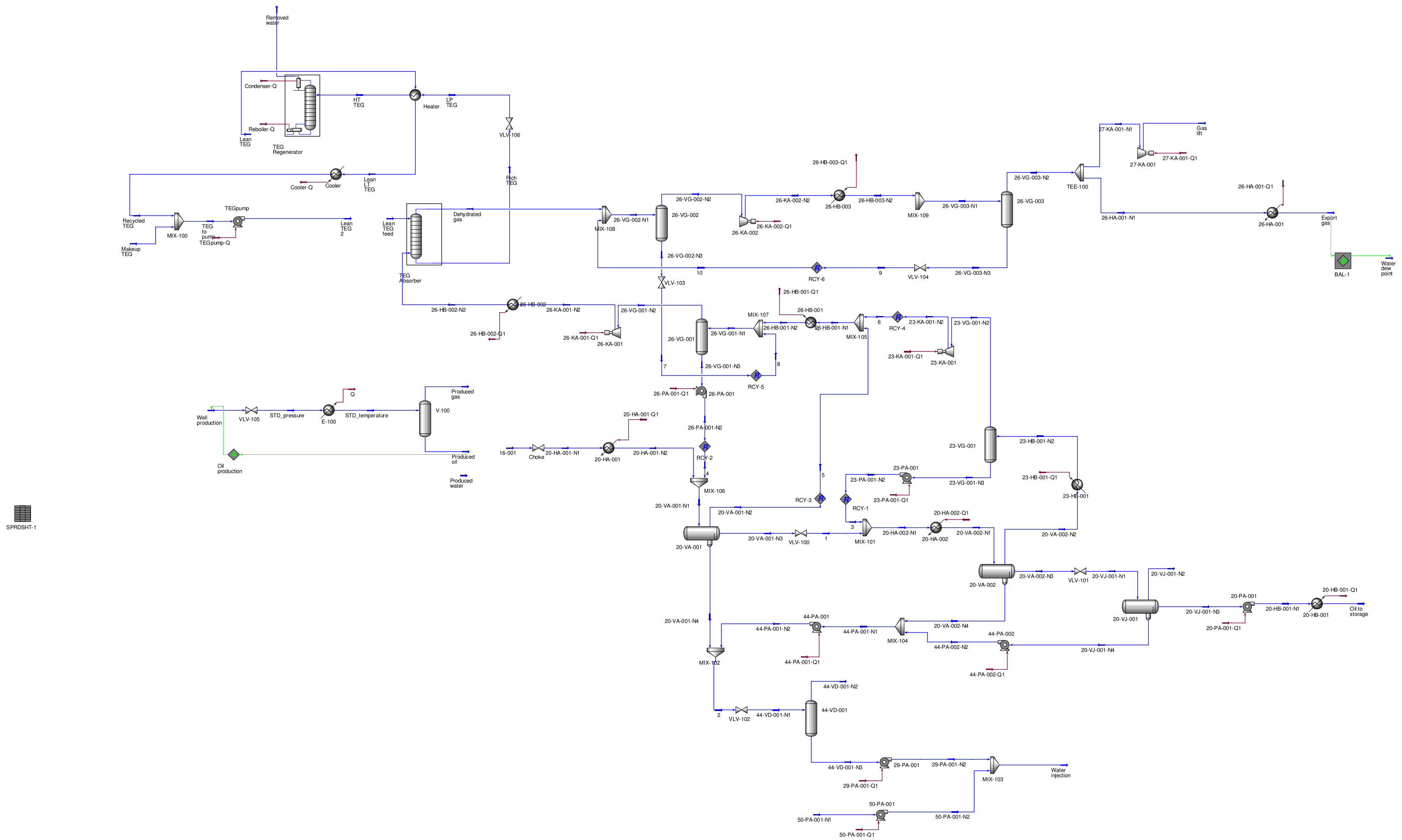
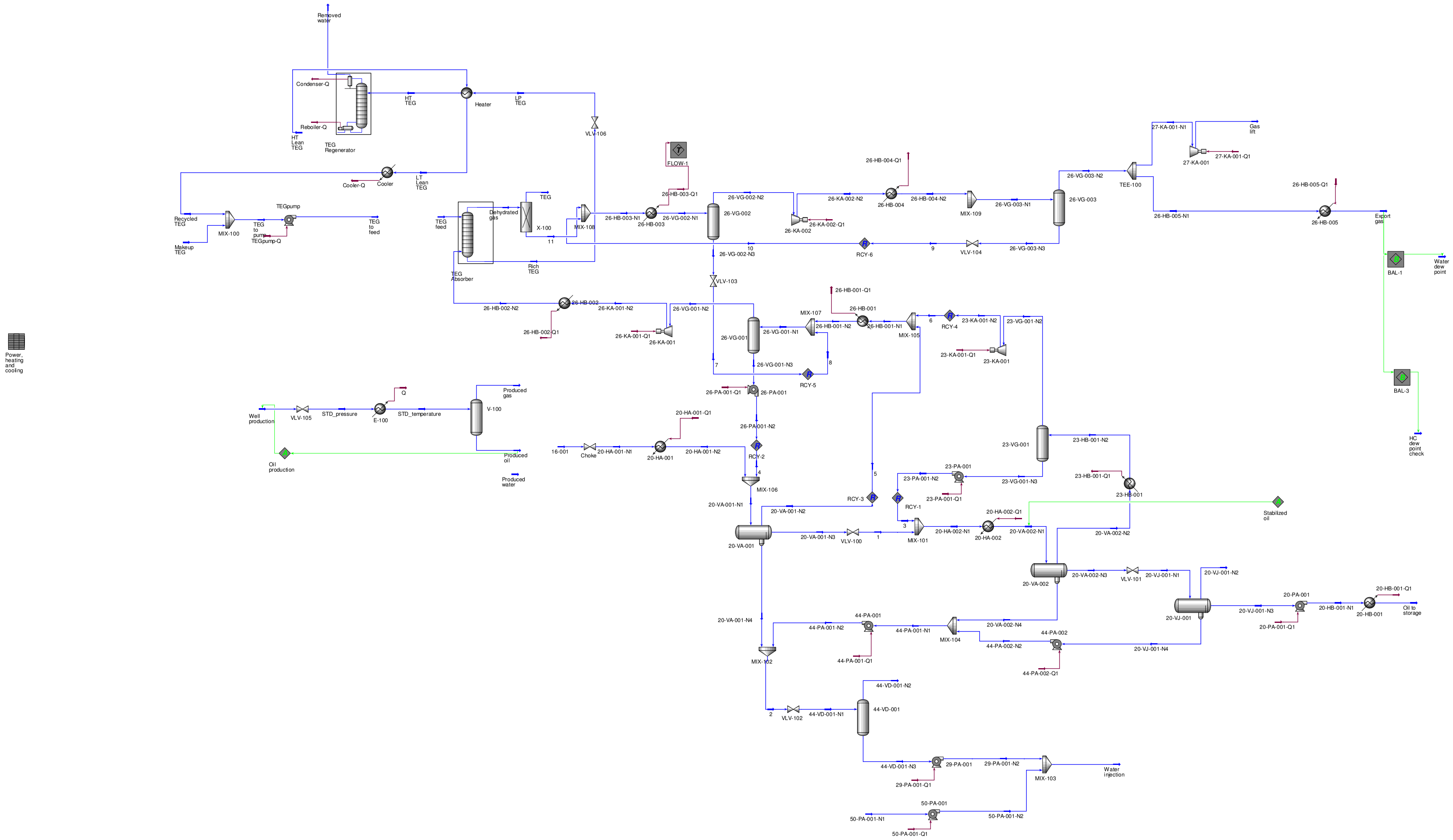
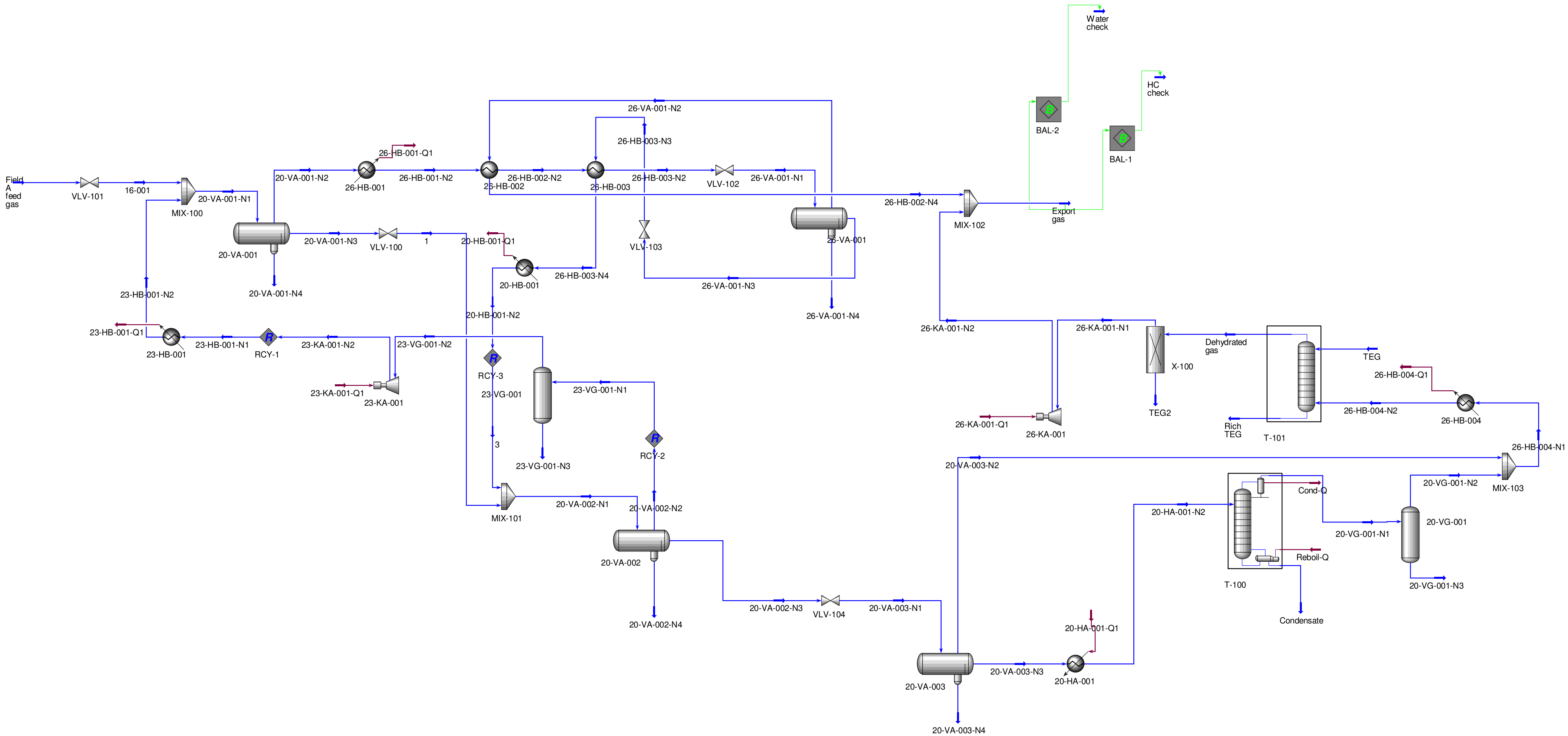
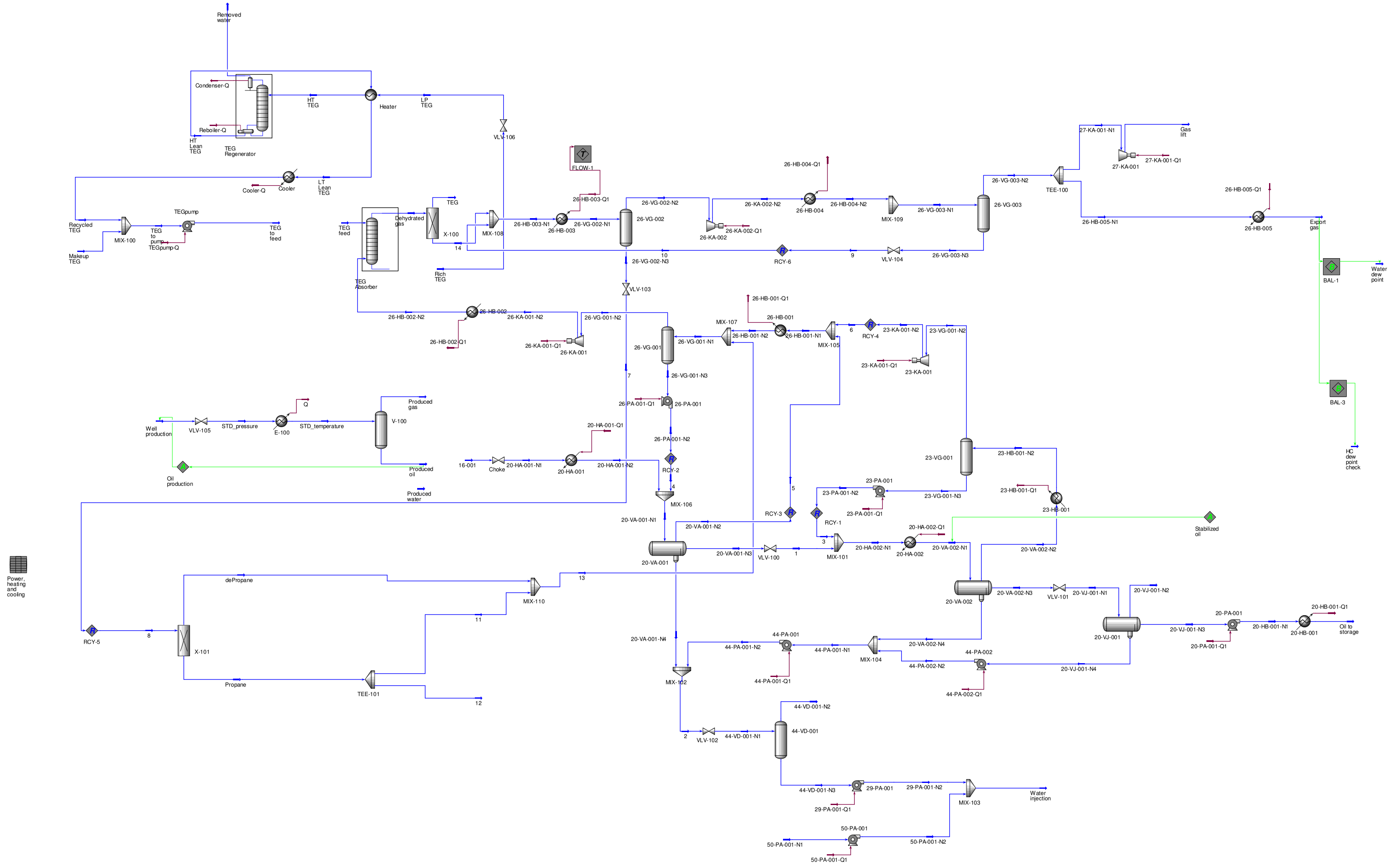
20-VJ-001
20-VJ-001-N2
20-VJ-001-N3
20-VJ-001-N4
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
MIX-104
44-PA-001-N1
44-PA-001
44-PA-001-N2
44-PA-001-Q1
44-PA-002
44-PA-002-N2
44-PA-002-Q1
MIX-102
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001
50-PA-001-N1
50-PA-001-N2
50-PA-001-Q1


Removed water
Condenser-Q
Reboiler-Q
Lean TEG
TEG Regenerator
HT TEG
Heater
LP TEG
VLV-106
Rich TEG
Cooler-Q
Cooler
Lean LT TEG
Recycled TEG
MIX-100
TEG to pump
TEGpump
TEGpump-Q
Makeup TEG
Lean TEG 2
Lean TEG feed
Dehydrated gas
TEG Absorber
MIX-108
26-VG-002-N1
26-VG-002
26-VG-002-N2
26-KA-002
26-KA-002-N2
26-KA-002-Q1
26-HB-003
26-HB-003-Q1
26-HB-003-N2
MIX-109
26-VG-003-N1
26-VG-003
26-VG-003-N2
26-VG-003-N3
TEE-100
27-KA-001-N1
27-KA-001
27-KA-001-Q1
Gas lift
26-HA-001-N1
26-HA-001
26-HA-001-Q1
Export gas
BAL-1
Water dew point
26-VG-002-N3
VLV-103
RCY-6
VLV-104
26-HB-002
26-HB-002-N2
26-HB-002-Q1
26-KA-001-N2
26-KA-001
26-KA-001-Q1
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
26-HB-001-N2
26-HB-001
26-HB-001-Q1
26-HB-001-N1
MIX-105
RCY-4
RCY-5
23-KA-001-N2
23-KA-001
23-KA-001-Q1
23-VG-001-N2
26-PA-001
26-PA-001-Q1
26-PA-001-N2
RCY-2
MIX-106
Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
RCY-3
RCY-1
MIX-101
20-HA-002-N1
20-HA-002
20-HA-002-Q1
20-VA-002-N1
23-VG-001
23-VG-001-N3
23-PA-001
23-PA-001-N2
23-PA-001-Q1
23-HB-001-N2
23-HB-001
23-HB-001-Q1
20-VA-002-N2
20-VA-002
20-VA-002-N3
VLV-101
20-VJ-001-N1
20-VJ-001-N2
20-VJ-001
20-VJ-001-N3
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
20-VJ-001-N4
20-VA-001-N4
20-VA-002-N4
MIX-104
44-PA-001
44-PA-001-N1
44-PA-001-N2
44-PA-001-Q1
44-PA-002
44-PA-002-N2
44-PA-002-Q1
MIX-102
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001
50-PA-001-N1
50-PA-001-N2
50-PA-001-Q1
SPRDSHT-1


Removed water
Condenser-Q
Reboiler-Q
HT Lean TEG
TEG Regenerator
HT TEG
Heater
LP TEG
VLV-106
Cooler-Q
Cooler
LT Lean TEG
Recycled TEG
Makeup TEG
MIX-100
TEGpump
TEG to pump
TEGpump-Q
TEG to feed
TEG feed
Dehydrated gas
TEG Absorber
X-100
TEG
Rich TEG
MIX-108
11
FLOW-1
26-HB-003-Q1
26-HB-003-N1
26-HB-003
26-VG-002-N1
26-VG-002-N2
26-VG-002
26-KA-002-N2
26-KA-002
26-KA-002-Q1
26-HB-004-Q1
26-HB-004
26-HB-004-N2
MIX-109
26-VG-003-N1
26-VG-003
26-VG-003-N2
26-VG-003-N3
TEE-100
27-KA-001-N1
27-KA-001
27-KA-001-Q1
Gas lift
26-HB-005-Q1
26-HB-005-N1
26-HB-005
Export gas
BAL-1
Water dew point
BAL-3
HC dew point check
10
26-VG-002-N3
RCY-6
9
VLV-104
VLV-103
26-HB-002
26-HB-002-N2
26-HB-002-Q1
26-KA-001-N2
26-KA-001-Q1
26-KA-001
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
26-HB-001-Q1
26-HB-001
26-HB-001-N2
26-HB-001-N1
MIX-105
RCY-4
23-KA-001-N2
23-KA-001-Q1
23-KA-001
23-VG-001-N2
RCY-5
7
8
6
26-PA-001-Q1
26-PA-001
26-PA-001-N2
RCY-2
4
MIX-106
Power, heating and cooling
Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
1
RCY-3
5
RCY-1
3
MIX-101
20-HA-002-N1
20-HA-002
20-HA-002-Q1
20-VA-002-N1
20-VA-002-N2
23-VG-001
23-HB-001-N2
23-HB-001-Q1
23-HB-001
23-VG-001-N3
23-PA-001
23-PA-001-N2
23-PA-001-Q1
Stabilized oil
20-VA-002
20-VA-002-N3
VLV-101
20-VJ-001-N1
20-VJ-001-N2
20-VJ-001
20-VJ-001-N3
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
20-VA-001-N4
44-PA-001
44-PA-001-N2
44-PA-001-N1
44-PA-001-Q1
MIX-104
20-VA-002-N4
44-PA-002
44-PA-002-N2
44-PA-002-Q1
20-VJ-001-N4
MIX-102
2
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001
50-PA-001-N1
50-PA-001-Q1
50-PA-001-N2



Field A feed gas
VLV-101
16-001
MIX-100
20-VA-001-N1
20-VA-001
20-VA-001-N2
20-VA-001-N3
20-VA-001-N4
VLV-100
1
26-HB-001
26-HB-001-Q1
26-HB-001-N2
26-HB-002
26-HB-002-N2
26-HB-003
26-HB-003-N2
26-HB-003-N3
26-HB-003-N4
VLV-102
VLV-103
26-VA-001
26-VA-001-N1
26-VA-001-N2
26-VA-001-N3
26-VA-001-N4
26-HB-002-N4
MIX-102
Export gas
BAL-1
BAL-2
Water check
HC check
20-HB-001
20-HB-001-Q1
20-HB-001-N2
23-HB-001
23-HB-001-Q1
23-HB-001-N1
23-HB-001-N2
RCY-1
RCY-2
RCY-3
23-KA-001
23-KA-001-Q1
23-KA-001-N2
23-VG-001
23-VG-001-N1
23-VG-001-N2
23-VG-001-N3
3
MIX-101
20-VA-002
20-VA-002-N1
20-VA-002-N2
20-VA-002-N3
20-VA-002-N4
VLV-104
20-VA-003
20-VA-003-N1
20-VA-003-N2
20-VA-003-N3
20-VA-003-N4
20-HA-001
20-HA-001-Q1
20-HA-001-N2
26-KA-001
26-KA-001-Q1
26-KA-001-N1
26-KA-001-N2
X-100
TEG2
Dehydrated gas
T-101
TEG
Rich TEG
26-HB-004
26-HB-004-Q1
26-HB-004-N1
26-HB-004-N2
MIX-103
T-100
Cond-Q
Reboil-Q
Condensate
20-VG-001
20-VG-001-N1
20-VG-001-N2
20-VG-001-N3


Removed water
Condenser-Q
Reboiler-Q
HT Lean TEG
TEG Regenerator
HT TEG
Heater
LP TEG
VLV-106
Cooler-Q
Cooler
LT Lean TEG
Recycled TEG
MIX-100
Makeup TEG
TEG to pump
TEGpump
TEGpump-Q
TEG to feed
TEG feed
TEG Absorber
Dehydrated gas
X-100
TEG
Rich TEG
MIX-108
FLOW-1
26-HB-003-Q1
26-HB-003-N1
26-HB-003
26-VG-002-N1
26-VG-002
26-VG-002-N2
26-VG-002-N3
26-KA-002
26-KA-002-N2
26-KA-002-Q1
26-HB-004
26-HB-004-Q1
26-HB-004-N2
MIX-109
26-VG-003-N1
26-VG-003
26-VG-003-N2
26-VG-003-N3
RCY-6
VLV-104
TEE-100
27-KA-001-N1
27-KA-001
27-KA-001-Q1
Gas lift
26-HB-005-N1
26-HB-005
26-HB-005-Q1
Export gas
BAL-1
Water dew point
BAL-3
HC dew point check
VLV-103
26-HB-002
26-HB-002-N2
26-HB-002-Q1
26-KA-001-N2
26-KA-001-Q1
26-KA-001
26-VG-001-N2
26-VG-001
26-VG-001-N1
26-VG-001-N3
MIX-107
26-HB-001-N2
26-HB-001
26-HB-001-Q1
26-HB-001-N1
MIX-105
RCY-4
23-KA-001-N2
23-KA-001
23-KA-001-Q1
23-VG-001-N2
26-PA-001
26-PA-001-Q1
26-PA-001-N2
RCY-2
MIX-106
Well production
VLV-105
STD_pressure
E-100
Q
STD_temperature
V-100
Produced gas
Produced oil
Produced water
Oil production
16-001
Choke
20-HA-001-N1
20-HA-001
20-HA-001-Q1
20-HA-001-N2
23-VG-001
23-HB-001-N2
23-HB-001-Q1
23-HB-001
23-VG-001-N3
23-PA-001
23-PA-001-N2
23-PA-001-Q1
RCY-3
RCY-1
MIX-101
20-HA-002-N1
20-HA-002
20-HA-002-Q1
20-VA-002-N1
20-VA-002-N2
Stabilized oil
20-VA-001-N1
20-VA-001-N2
20-VA-001
20-VA-001-N3
VLV-100
20-VA-002
20-VA-002-N3
VLV-101
20-VJ-001-N1
20-VJ-001-N2
20-VJ-001
20-VJ-001-N3
20-PA-001
20-PA-001-Q1
20-HB-001-N1
20-HB-001
20-HB-001-Q1
Oil to storage
20-VJ-001-N4
Power, heating and cooling
dePropane
MIX-110
RCY-5
X-101
Propane
TEE-101
20-VA-001-N4
44-PA-001-N2
44-PA-001
44-PA-001-Q1
44-PA-001-N1
MIX-104
20-VA-002-N4
44-PA-002
44-PA-002-N2
44-PA-002-Q1
MIX-102
VLV-102
44-VD-001-N1
44-VD-001
44-VD-001-N2
44-VD-001-N3
29-PA-001
29-PA-001-Q1
29-PA-001-N2
MIX-103
Water injection
50-PA-001
50-PA-001-N1
50-PA-001-Q1
50-PA-001-N2