NTNU
Norges teknisk-naturvitenskapelige
universitet

Fakultet for naturvitenskap og teknologi
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Process design project, 2011 - Fall

Title: Process design and economical investigation of natural gas dehydration plants	**Subject (3-4):** Natural gas dehydration, cooling, absorption
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Supervisor: Sigurd Skogestad **Co-supervisors:** Mehdi Panahi and Maryam Ghadrdan	**Total number of pages:** **Main report:** 44 **Appendix:** 16

SUMMARY OF THE WORK AND CONCLUTIONS

Assumptions and performance:

Simulation of two processes for dehydration of natural gas. Optimazation, economical study and comparison. The dehydration process is part of larger natural gas processing plants.

Conclustions and recommendations:

The economical analysis showed that the absorption was the most economical process for dehydration of natural gas.

Date and signatures:

Abstract

Dehydration is an essential part of the processing of natural gas. By dehydrating the gas, hydrate formation is avoided at low temperatures, and subsequently the risk of pipe blockage is reduced. As of 2006, dry natural gas was ranked as the third largest primary energy source. Statistics from 2010 ranks Norway as the fifth largest producer of natural gas in the world.

Removal of water from natural gas can be achieved in several different ways. Among them is the process for dehydration by absorption using glycols and the process of dehydration by cooling with a refrigeration cycle. The two alternative processes were simulated using UniSim and subjected to economical analysis, including a sensitivity study.

The economical analysis showed that the absorption was the most economical process for dehydration of natural gas, compared to the cooling process. Both capital investment and manufacturing costs were considerably larger for the cooling process. The cooling process was also the most sensitive to variations in direct manufacturing costs.

Preface

This report was written as a part of the course TKP4170, “Process Design Project”, during the fall semester 2011. The project title is “Process design and economical investigation of natural gas dehydration plants”.

We would like to thank Ph. D Mehdi Panahi and Ph. D Maryam Ghadrdan for helpful guidance throughout the process.

We would also like to thank Professor Sigurd Skogestad for being available and answering our questions.

Trondheim, 18.11.11

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1 Introduction

Natural gas is found together with crude oil in reservoir rocks where it has been formed by accumulation of organic matter in millions of years. Natural gas is used as a fuel and also as raw material in manufacturing (Guo & Ghalambor, 2005).

Norway is the second largest exporter of natural gas after Russia as of 2010, and is the fifth largest producer of natural gas in the world. In 2010, production of crude oil, natural gas and pipeline services accounted for 50 % of the export value of Norway and 21 % of the GDP (gross domestic product). All of the oil and natural gas fields in Norway are located subsea on the Norwegian Continental Shelf, and Troll is the single largest field, representing one-third of the country's natural gas production. As of 2011 there are several process plants in Norway where dehydration of natural gas takes place. Among them, Melkøya outside Hammerfest, Kolsnes outside Bergen and Kårstø north of Stavanger, all these are run by Statoil.

1.1 Properties and composition

Methane is the principal constituent of natural gas. Paraffinic hydrocarbons such as ethane, propane and butane are other. Often natural gas also contains carbon dioxide, nitrogen and hydrogen sulfide. Small amounts of compounds like argon, hydrogen, helium, as well as C_5+-hydrocarbons and aromatics can be present. Typical composition of natural gas is shown in Table 1.1.1.

Table1.1.1: Typical Composition of natural gas (Guo & Ghalambor, 2005).

Name	**Volume (%)**
Methane	>85
Propane	3-8
Butane	1-2
Carbon dioxide	<1
Hydrogen sulfide	<1
Nitrogen	1-5
Helium	<0.5

Natural gas has neither colour, nor odour and is lighter than air. In Table 1.1.2 some properties of natural gas are listed.

Table 1.1.2: Properties of natural gas (Guo & Ghalambor, 2005).

Properties	**Value**
Relative molar mass	17-20
Carbon content, wt %	73.3
Hydrogen content, wt %	23.9
Oxygen content, wt %	0.4
Relative density, 15 °C	0.72-0.81
Boiling point, °C	-162

1.2 Applications of natural gas

As of 2006, dry natural gas was ranked as the third largest primary energy source accounting for 22.8 % of the world primary energy production (I1).

As an example the different uses of natural gas in the US in 2010 are listed under (I2):

- 34 % Industrial
- 29 % Electric power
- 21 % Residential
- 13 % Commercial
- 3 % Transportation

Figure 1.2.1 shows that the production of natural gas has increased in recent years. Natural gas has various uses in industry, and is the base ingredient in products like plastic, fertilizer, fabrics and anti-freeze (I3). Natural gas can serve as a fuel for vehicles in the form of liquified natural gas (LNG) or compressed natural gas (CNG), heating, cooking and electricity generation. In addition it is used as a feedstock for petrochemical plants and methanol production. The inorganic components of natural gas have no value, rather they may corrode or destroy the equipment used for gas processing and therefore has to be removed (Caroll, 2009).

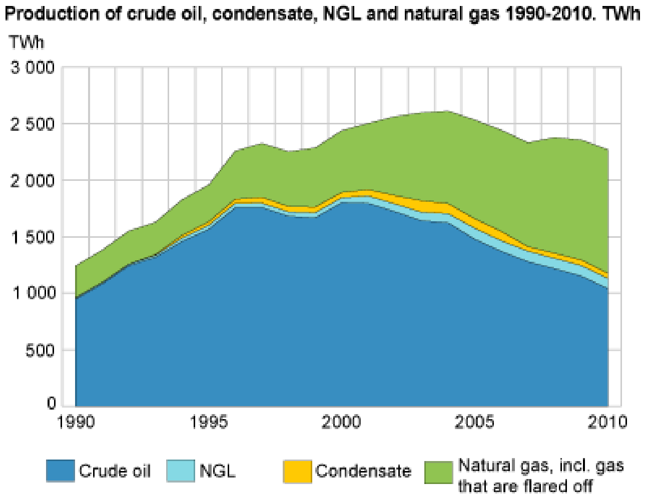

Figure 1.2.1: Development of the production of crude oil, condensate, liquefied natural gas in the period from 1990 to 2010 (I4).

1.3 Processing of natural gas

Dehydration is a necessary part of the processing of natural gas. Dehydration of the gas is a way to avoid hydrate formation at low temperatures, and subsequently reduce the risk of pipe blockage. Figure 1.3.1 shows a simplified liquid natural gas (LNG) plant diagram. Further discussion on the need of dehydration is given in section 1.3.1.

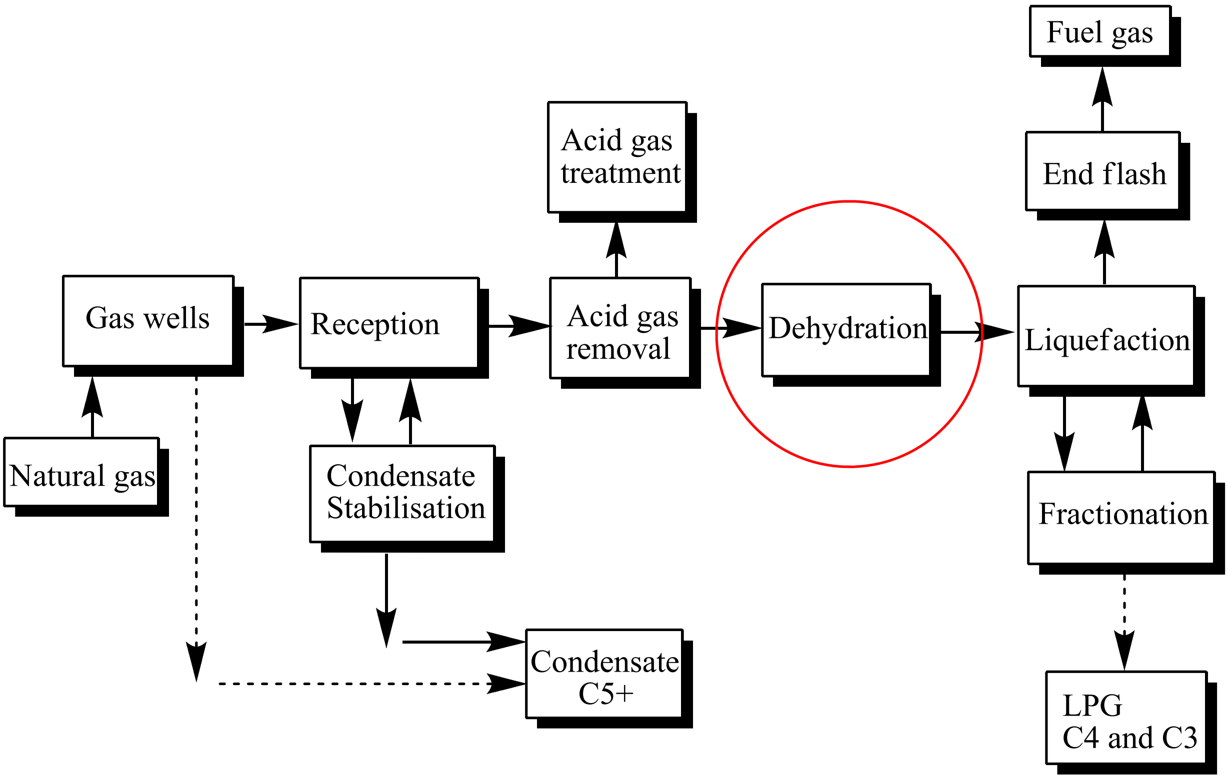

Figure 1.3.1: Simplified liquid natural gas plant diagram (I5).

1.3.1 Dehydration of natural gas

In addition to crude oil and natural gas, the reservoir rock contains water. The presence of water in natural gas at source or as a result of sweetening with an aqueous solution presents no danger as long as the water is in the vapour phase. However, several processing and transmission problems arise as soon as liquid water is formed (Mokhatab et al., 2006, Brathen, 2008). Operation experience has proved that it is necessary to reduce and control the water content of gas to ensure safe processing and transmission. Dehydration is the process used to remove water from natural gas, and it is necessary due to the following reasons:

- Natural gas in the right conditions can combine with liquid water to form ice like material called hydrates that can plug valve fittings, pipelines as well as other equipment.
- Liquid water can accelerate the corrosion of transmission lines by combining with CO_2 and/or H_2S present in natural gas.

- Liquid water in a natural gas pipeline potentially causes slugging flow conditions resulting in lower flow efficiency of the pipeline.
- Water vapour increases the volume and decreases the heating value of the gas.
- A minimum water content of 112 mg/m^3 of gas or less has to be met (Mokhatab et al., 2006, Guo & Ghalambor , 2005, Kirk-Othmer, 2001).

Removal of water from natural gas can be achieved by employing absorption using solid desiccants, liquid desiccants or refrigeration (direct cooling). The concept of dehydration with solid desiccants is to adsorb the water on a bed, which requires a given time, and then heating will remove the adsorbed water. The amount of heating depends on the bed in use. Solid desiccants used in natural gas dehydration are regenerated and used in many cycles. Among the most common solid desiccants, are alumina and silica gel. Alumina will adsorb heavy hydrocarbons, which are not easily removed. This is to a less extent a problem for silica gel dehydration. This method is most commonly used for feed streams to cryogenic processing systems. The processes of liquid dehydration and dehydration by cooling will be investigated in this project. Anhydrous calcium chloride can also be used to dehydrate natural gas, but this technique is best suited for remote application where modest dew point depressions are required and gas processing volumes are small. Other dehydration methods include membrane permeation, gas stripping and distillation (Kirk-Othmer, 2001, GPSA SI Engineering Data Book, 2004).

Economically, the absorption method is often considered the most favourable method of dehydration. It is by far the most common process, partly because the process is easily operatable and automated. The main components of the process include an absorber in which the natural gas is dehydrated by glycol, and a regeneration cycle where the glycol is stripped of water. To increase the concentration of glycol in the regeneration cycle, a stripping gas can be added to the stripper. Glycols are mostly used as solvent, because of their desirable properties that meet commercial properties. Glycols are described later (SI Engineering Data Book, 1980).

Dehydration by cooling is an alternative to absorption. When decreasing the temperature of the natural gas, the saturated water content will decrease (Katz et al., 1959). TEG will be added to the natural gas to prevent formation of hydrates during cooling. When the excess water is removed, the temperature of the dry natural gas must be increased or the pressure decreased. This is because the natural gas will be at its water dew point after the water is removed (Guo & Ghalambor, 2005).

1.4 Project basis

Water is to be removed from natural gas and two different alternatives are to be evaluated. The first one is to use an absorption column to separate the water, and the other one is to use a cooling cycle. These processes are simulated using UniSim. Profitability of these alternatives is compared by doing a feasibility study and a sensitivity analysis.

1.4.1 Feed conditions and capacity

The molar flow rate of the inlet stream is 20000 kmol/h, the temperature is 50 °C, and the pressure is 70 bar. The mole fractions in the inlet stream are listed in Table 1.4.1.1.

Table 1.4.1.1: Mole fractions for the inlet stream.

Component	**Molar ratio**
CH_4	0.8860
C_2H_6	0.0370
C_3H_8	0.0150
i- C_4H_{10}	0.0030
n-C_4H_{10}	0.0050
N_2	0.0500
H_2O	0.0040

The number of operating days is assumed to be 90 % of one year, 328 days, and the life time for the plant is assumed to be 10 years.

1.4.2 Product specifications

The requirements for the dry natural gas are

- Maximum water content of 20 ppm.
- Maximum content of dehydration liquid of 2 ppm.

1.4.3 Assumptions

- The plant will be built in Norway, and by the sea.
- Sea water will be used as cooling agent instead of electric power, and is assumed to be 10 °C.
- The dehydration plant is part of a larger natural gas-recovery plant, so the steam used for heating comes from an already existing steam generation plant. The duty needed is then only the duty for generating of steam.

2. Process description

In this project two alternative processes for dehydration of natural gas are simulated. The two processes have the same feed conditions and product specification. In this section the processes, including optimization, will be described.

Peng-Robinson is a fluid package suitable for hydrocarbons. The Peng-Robinson fluid package was used for the absorption process in this project. When using this fluid package for the cooling process, the regeneration of the glycol became difficult to simulate. For this process the fluid package was changed to Peng-Robinson Stryjek Vera (PRSV). The PRSV package is a modified Peng-Robinson model, which gives a better representation of vapour pressures, and makes it possible to use the Peng-Robinson model for non-ideal systems. The most appropriate fluid package for natural gas dehydration with glycols is the Glycol package. Peng-Robinson can still be used for dehydration of natural gas under normal dehydration conditions (I6). The Glycol package was tried without significant changes.

The simulation flow sheets from UniSim and selected parameters from the important streams for both processes are given in Appendix A.

2.1 Dehydration of natural gas by absorption

The most common process for removing water from natural gas is dehydration with glycols. The process consists of an absorption column, where the natural gas and glycol come into contact, and a regeneration cycle where the absorbed water and hydrocarbons are removed from the glycol stream, so that it can be recycled back to the absorber, as shown in Figure 2.1.1.

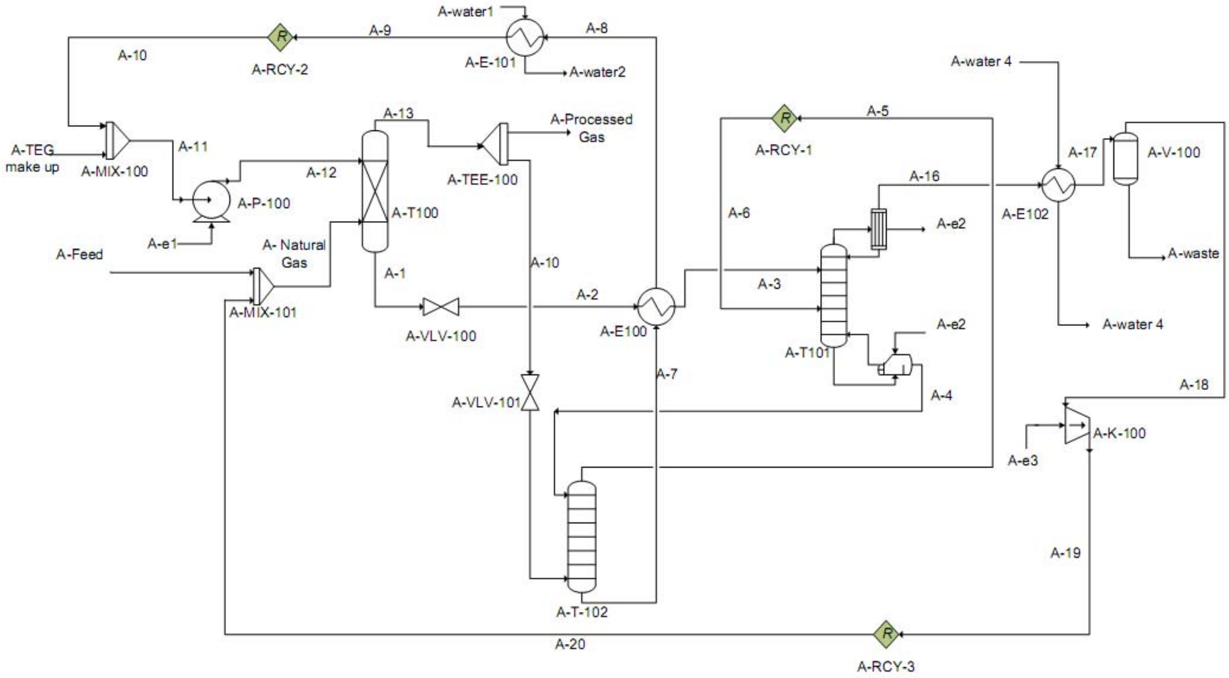

Figure 2.1.1: Process used for dehydration of natural gas by absorption.

Triethylene glycol (TEG) is the most commonly used glycol in the dehydration process, and the glycol chosen for this project. TEG can be regenerated to high purity for recycling at temperatures of 170 °C to 205 °C, at higher temperatures the TEG will thermally decompose. If the temperature in the absorption column exceeds 50 °C, one can expect a higher loss of glycol in the natural gas than what is necessary. There are several other glycols that can substitute TEG in the dehydration process. Monoethylene glycol is one of the alternatives, but high vapour equilibrium with water, will give high glycol losses to the natural gas. Diethylene glycol is another alternative, but high vapour pressures will cause high losses of glycol to the natural gas. In addition, DEG will decompose at relatively low temperatures, and the regeneration process might not reach the required purity. Tetraethylene glycol is a more expensive alternative, but will give less glycol loss at high pressures (Katz et al., 1959, Mokhatab et al., 2006).

In the absorption column, operated at 70 bar, the natural gas rich in water is contacted counter currently with the hygroscopic TEG. Pressure variation in the absorption column has little effect on the process for pressures under 207 bar. A high pressure reduces the water content in the inlet gas stream. This means less water has to be removed by absorption. At high pressures the gas velocity will be reduced, resulting in a smaller vessel needed. The pressure will also have an effect on the wall thickness in the column. When the rich natural gas comes into contact with the TEG, the water vapour becomes dissolved in the TEG stream (Mokhatab et al., 2006).The product stream, the “dry” natural gas, leaves the top of the absorber, while the glycol rich in water leaves in the bottom stream. The processed gas leaves the absorber with a water content of 19 ppm and a TEG content of 1.3 ppm. The temperature of the natural gas entering the absorber is 50.63 °C, and the temperature of the TEG entering is 52.47 °C. To get the largest effect of dew-point depression, the temperature of TEG should be low. A low TEG temperature will decrease the TEG flow needed through the absorber. On the other side, the temperature of the TEG entering the absorber should be held slightly above the inlet gas temperature so hydrocarbons will not condensate (Arnold & Stewart, 1999). The temperature in the absorber exceeds the 50 °C limit for TEG loss mentioned above, but the requirement of less than 2 ppm TEG in the processed gas allows the absorber temperature of 52.94 °C. To reach the given requirement of 20 ppm water in the processed gas stream, an absorber with 8 theoretical stages was used. Assuming 70 % plate efficiency, this gives 12 actual stages.

The simplest kind of regeneration cycle for glycol consists of a distillation column, A-T-101, where the rich glycol is stripped of water and hydrocarbons by boiling, and the bottom stream is recycled back to the absorber. The stripper was converged by using mole fraction of water in the lean TEG and the temperature in the condenser as specifications. These specifications were set to 0.081 and 96 ºC respectively. The regeneration cycle was operated at atmospheric pressure. The number of theoretical stages was found to be 3 by using a shortcut distillation column in UniSim. This gives 5 actual stages. Since TEG will thermally decompose at temperatures above 204 °C, the rich stream can not be heated enough to get the needed concentration to reach the water content specification of the processed natural gas (Hernandez-Valencia et al, 2006).

To reach the required 20 ppm of water in the natural gas, another contactor column with 5 actual stages, A-T-102, was added to the regeneration cycle. In the column, the TEG from the stripper is counter currently contacted with processed natural gas to reach a higher concentration of TEG than the boiling will allow. The addition of the extra contactor column and the use of stripping gas resulted in a recycled TEG concentration of 99 mole%. The high concentration of TEG makes it possible to operate the absorption column with a TEG flow of 71.99 kmol/h. The “wet” gas flow from the stripper is cooled and separated to reduce the water content from 85 mole% to 20 mole%. The gas stream with 20 mole% water is recycled back to the natural gas feed. The waste stream from the process consists of over 99 mole% water, and a small amount of nitrogen and TEG.

2.1.1 Optimization of the absorption process

The parameters that should be considered when optimizing the TEG absorption cycle is the number of trays in the absorption column, the glycol circulation rate through the contactor, temperature in the reboiler, amount of stripping gas used and operating pressure in the regenerator (Hernandez-Valencia et al., 2006).

Number of stages in the contactor influences the recycle stream of TEG. By reducing the number of stages the recycle flow has to be increased in order to reach the desired specification of 20 ppm for water in the processed natural gas stream.

The recycle stream was optimized by finding the lowest stream of TEG in the recycle needed to reach the water specification of 20 ppm in the processed gas stream. A graphical display of the relationship between the molar flow in the recirculation stream and the water content in the processed gas is shown in Figure 2.1.1.1.

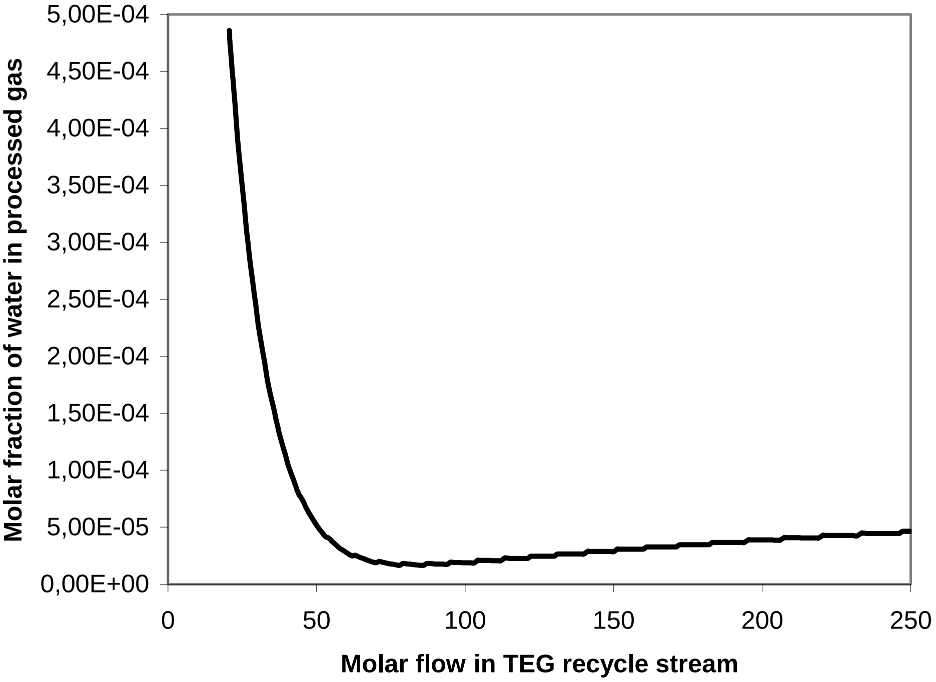

Figure 2.1.1.1: Molar flow in the recycle stream plotted against the water content in the processed gas.

The temperature in the reboiler is limited by the thermal decomposition of TEG, as mentioned above. The temperature in the reboiler was chosen to be 199.9 °C. This reboiler temperature gives a concentration of 92 mole% out of the regeneration column. Lower temperatures will give lower concentration of TEG out of the regeneration column, and the process will then require more stripping gas. A larger amount of stripping gas will result in more expensive equipment for the gas recycle.

The regenerator is assumed to be vented to the atmosphere, and because of this it is not an optimization parameter in this case.

2.2 Dehydration of natural gas by cooling

The process of dehydration of natural gas by cooling mainly consists of a refrigeration cycle and two separators, one of them separating water from the feed, and the other one separating the TEG rich in water, from the natural gas product. The cooling process is shown in Figure 2.2.1. The specification of water content in the natural gas is 20 ppm. Hydrocarbons are well suited as refrigerants because of their low expenses, low toxicity, and their low ozone depletion potential and global warming potential. Propane is widely used as a refrigerant in the petrochemical industry, and will be used in the cooling cycle. Propane is a stable substance, and satisfies the requirements for COP and pressure ratio, but there is a small explosion and fire hazard to take into account (Dinçer & Kanoglu, 2010).

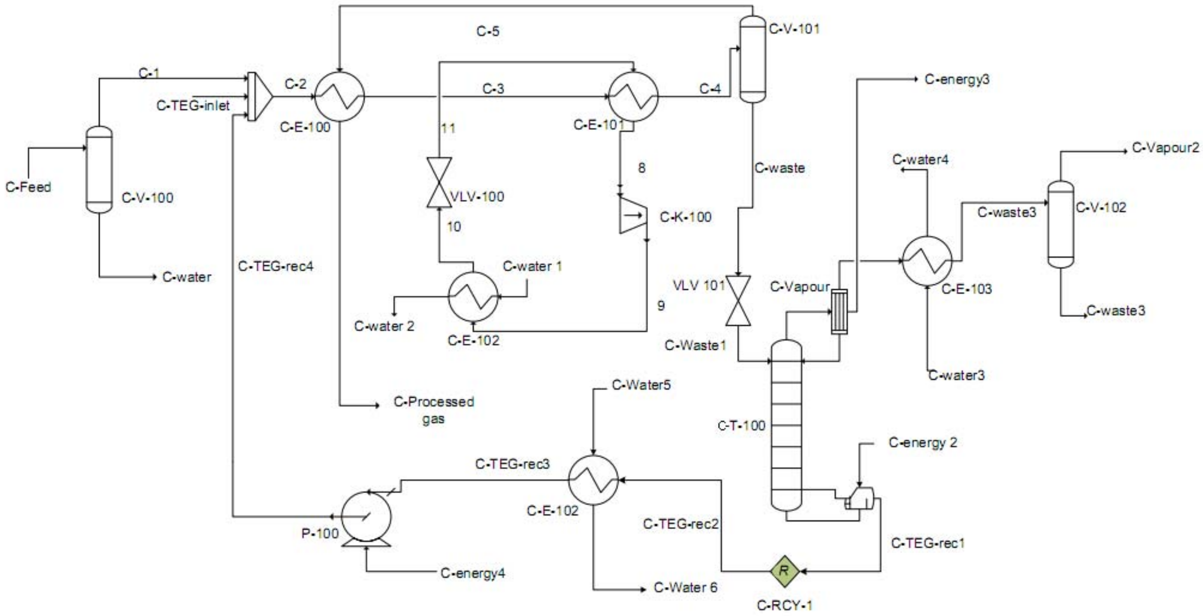

Figure 2.2.1: Process for dehydration of natural gas by cooling.

The first step in the process is to separate the removable water from the natural gas stream, C-feed, in a flash tank. The pressure in the flash tank is 70 bar. Due to gravitational forces, the water in liquid phase will fall to the bottom of the column. The water content in the natural gas stream is reduced from 0.40 mole% to 0.24 mole%. This makes the rest of the process more economical.

The natural gas stream is then mixed with TEG to lower the dew point. The TEG content is 0.35 mole%. This makes it possible to remove the water without any hydrate formation. With a pressure of 70 bar, hydrates will be formed at 15.6 °C (Arnold & Stewart, 1999). The cooling is done in two steps, the first in a heat exchanger where the natural gas is cooled by the processed gas stream, and the second one in a refrigeration cycle.

The natural gas with TEG, C-2, is cooled from 52 °C to 1 °C by C-5 in heat exchanger C-E-100. There is no pressure drop in this heat exchanger. Further cooling of the wet natural gas is done by a refrigeration cycle. The propane used as a refrigerant in the cycle will be reused. It is assumed that the same propane is used throughout the plant life.

The temperature of the natural gas stream C-4 is -3 °C. This is the highest temperature possible with the given specification of the processed gas. At this temperature and a pressure of 70 bar, the stream is separated in a flash tank. The liquid outlet, C-waste1, mainly consists of water and TEG.

The TEG in the liquid outlet is recycled by removing it from the stream. This is done by reducing the pressure to 1 bar in a valve, so that the stream is partly liquid and partly vapours. The stream is then stripped in a column, and the content of TEG in the liquid outlet is 92.0 mole%. The stripper was converged by using mole fraction of water in the lean TEG and the condenser reflux ratio as specifications. The values were set to 0.08 and 1 respectively. The temperature in this recycle stream is reduced to 49 °C, and the pressure is increased to 70 bar.

The vapour out of the stripping column is cooled to 60 °C, and separated. The liquid from the separator contains more than 99 % water, and can safely be returned to the environment. The vapour stream hydrocarbons and TEG is used as fuel.

2.2.1 Optimization of the cooling process

The simulation was optimized to get the water content in the processed gas to 20 ppm, and to lower the costs of equipment and operation costs. The optimal molar flow of propane in the refrigeration cycle was found by plotting the sum of the major variable costs, which is the heat exchanger C-E-100 and the compressor costs, against the molar flow. The most cost effective molar flow was found to be 210 kmol/h, as shown in Figure 2.2.1.1.

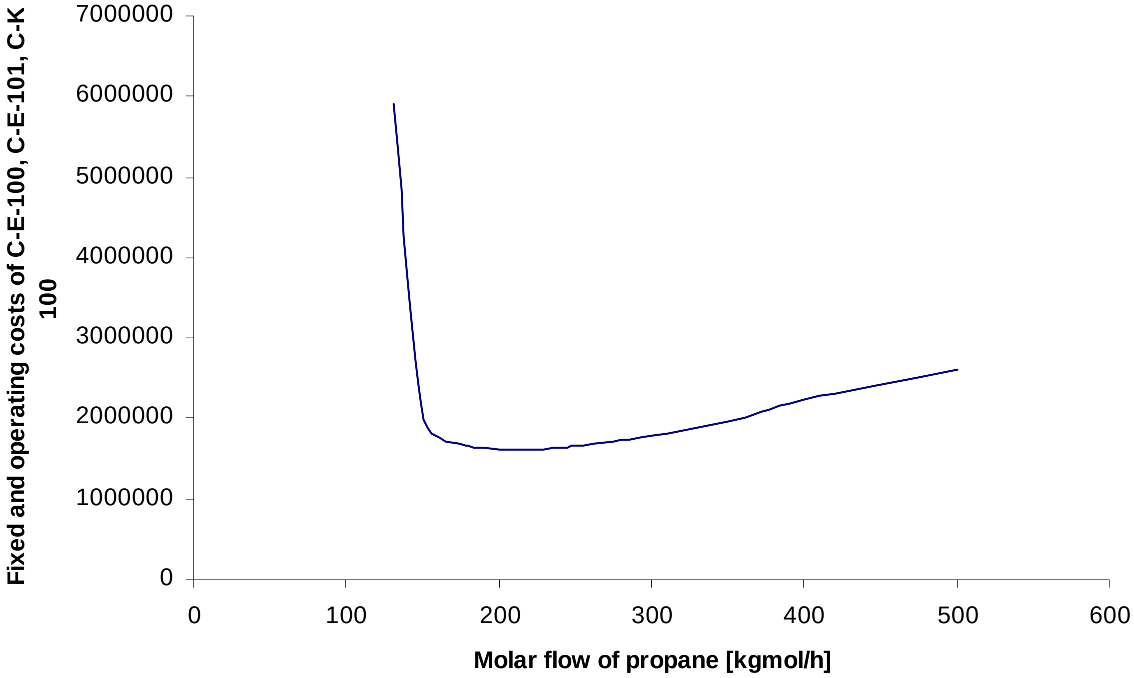

Figure 2.2.1.1 : Cost of heat exchangers C-E-100 and C-E-101, cost of compressor C-K-100, and operating costs of the compressor for 10 years plotted against the molar flow of propane in the refrigeration cycle.

The pressures in the cooling cycle were chosen to get the volumetric efficiency of the compressor to 80 %. The temperatures were chosen to match the pressure-enthalpy diagram of propane for these pressures, see Appendix B. The amount of TEG needed in the natural gas to reach the specification was found to be 76 kmol/h by plotting the molar fraction of water in the processed gas against the molar flow in the recycle stream, as shown in Figure 2.2.1.2

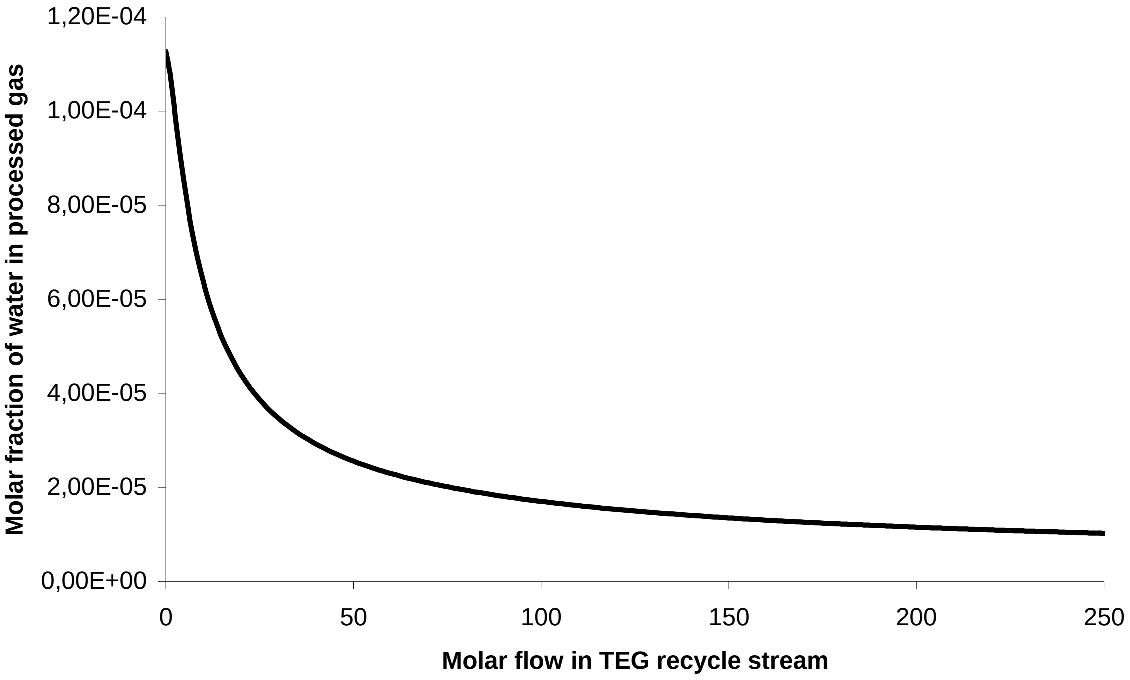

Figure 2.2.1.2: Molar flow in the recycle stream plotted against the water content in the processed gas.

2.2.2 Refrigeration cycle

The refrigeration system is based on four thermal processes: evaporation, compression, condensation, and expansion.

Evaporation occurs when heat is added to a liquid, and the amount of molecules in the gaseous phase increases. Enthalpy and entropy increases with constant pressure and temperature. This process is shown as step 4-1 in Figure 2.2.2.1. A detailed diagram of a pressure-enthalpy diagram for propane is given in Appendix B. In the process of natural gas dehydration, this will be the step that is cooling the natural gas stream. By using a heat exchanger, the refrigerator will be evaporated by the natural gas.

The next step in the cycle is compressing the vapour by using a compressor, which is an energy requiring process. This is shown as point 1-2 in the figure. The pressure, temperature, and enthalpy increases while the entropy in the ideal case is constant.

After the compressor, the high-pressure vapour will be changed into high-pressure saturated liquid by a condensator, shown as step 2-3 in the figure. In the refrigeration cycle in the natural gas process, the vapour is condensated by water in a heat exchanger. Because the refrigerant is condensated in this step, it is possible to reuse it in the cycle.

A valve is then used to reduce the pressure, and the refrigerant becomes partly liquid, shown in step 3-4 in the figure. This is an isenthalpic process, and the temperature is decreased to a temperature below the heat source, which in our case is the natural gas (Dinçer & Kanoglu, 2010).

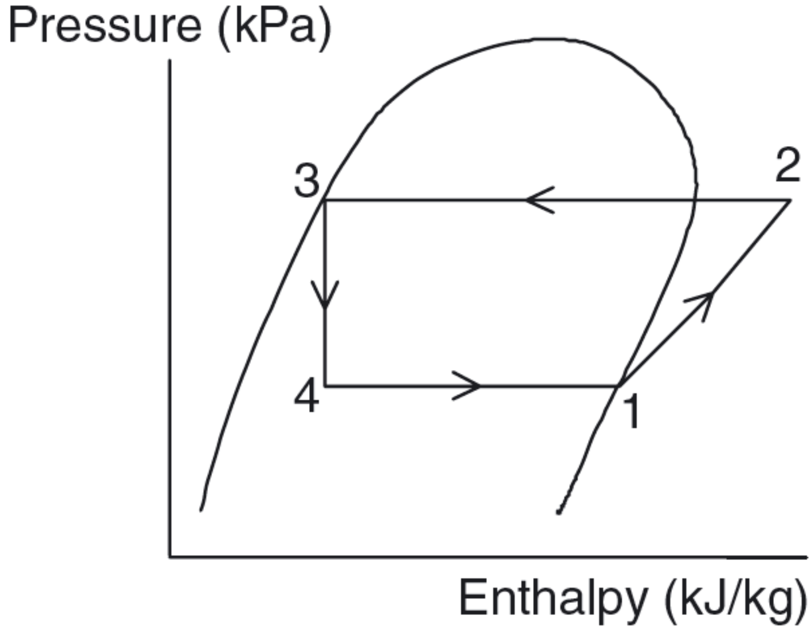

Figure 2.2.2.1: Principal sketch of a pressure-enthalpy diagram for a refrigeration cycle (Dinçer & Mehmet, 2010).

3. Mass and energy balances

For both processes the overall mass and energy balances were made. This was done to check that mass and energy was conserved in the simulations. Section 3.1 and 3.2 gives the results for the absorption process and the cooling process.

3.1. The absorption process

Section 3.1.1 and 3.1.2 shows the mass and energy balances for the absorption process.

3.1.1 Mass balance

From the numbers in Table 3.1.1.1 the total mass balance error was found to be 0.0 %.

Table 3.1.1.1: Total mass balance for the absorption process.

Mass flow in		**Mass flow out**	
Stream	**Mass flow [kg/h]**	**Stream**	**Mass flow [kg/h]**
A-feed	358515	A-waste	1434
A-TEG makeup	4	A-processed gas	357084
Sum of mass flow in	358519	Sum of mass flow out	358519

3.1.2 Energy balance

The values in Table 3.1.2.1 give an error of 0.0 %.

Table 3.1.2.1: Heat flows in and out of the absorption process.

Heat flow in		**Heat flow out**	
Stream	**Heat flow [kJ/kg]**	**Stream**	**Heat flow [kJ/kg]**
A-feed	-1467683429	A-waste	-22573568
A-TEG makeup	-22327	A-processed gas	-1444278724
A-water1	-87160894	A-water2	-86807232
A-water3	-870279472	A-water4	-866748250
A-e1	87864	A-e3	105643
A-e2	4306425		
A-e4	452775		
Sum heat flow out	-2420299058	Sum heat flow out	-2420302133

3.2 The cooling process

Section 3.2.1 and 3.2.2 shows the mass and energy balances for the absorption process.

3.2.1 Mass balance

The values in Table 3.2.1.1 give an error of 0.0 %.

Table 3.2.1.1: Total mass balance for the cooling process.

Mass flow in		**Mass flow out**	
Stream	**Mass flow [kg/h]**	**Stream**	**Mass flow [kg/h]**
C-feed	358515	C-water	594
C-TEG	0.00001	C-processed gas	357052
		C-vapor 2	34
		C-waste3	834
Sum of mass flow in	358515	Sum of mass flow out	358515

3.2.2 Energy balance

The values in Table 3.2.2.1 give an error of 0.2 %.

Table 3.2.2.1 Heat flows in and out of the cooling process.

Heat flow in		**Heat flow out**	
Stream	**Heat flow [kJ/h]**	**Stream**	**Heat flow [kJ/h]**
C-feed	-1464545724	C-water	-9359322
C-TEG inlet	-0.4	C-processed gas	-1444728554
C-energy 1	728782	C-energy 3	1952059
C-energy 2	10582434	C-water 2	-1242709854
C-energy 4	88880	C-water 4	-511599993
C-water 1	-1247627600	C-water 6	-1339953875
C-water 3	-513624536	C-vapor 2	-135577
C-water 5	-1345256442	C-waste 3	-13118254
Sum heat flow in	-4559654206	Sum heat flow out	-4559653368

4. Equipment

All the calculations are based on the equations given in Sinnott & Towler (2009). For calculation details, see Appendix C.

4.1 Flash tanks

Separators (flash tanks) are essentially required to separate liquid droplets or mist in a gas stream. In this project, they are required to separate water droplets from natural gas stream prior to cooling (C-V-100) and to separate water and TEG from the cooled gas stream (C-V-101). They are also used to separate water from the waste stream in both processes. Vertical separators are selected based on the fact that they require gravity for separation; low floor space and can handle large fluid volumes.

Sizing of the separators are based on the design procedure described by R. K. Sinnott and G. Towler. A demister pad is used to ensure high level of separation, and 10 minutes liquid hold up time is assumed. The heights and diameters for the flash tanks in the processes are shown in Table 4.1.1.

Table 4.1.1: Diameters and heights for the vessels.

Size parameters	**C-V-100**	**C-V-101**	**C-V-102**	**A-V-100**
Diameter [m]	2.86	2.54	0.08	0.26
Height [m]	4.70	4.99	29.47	5.35

The material of construction for all the separators shall be carbon steel, due to the low corrosiveness of the steams.

4.2 Absorption and stripping columns

Absorption columns are used to remove a component from a gas by contacting the gas with a solvent that will selectively dissolve the component. In the absorption process, water is the component to be removed with TEG. The absorbed solute can be removed from the solvent in a stripping column. This is done by counter current contact with vapour. In the absorption process the stripping column is used to regenerate the solvent from the absorption column.

The size of the column is calculated based on the number of actual trays and the column diameter. The heights and diameters for the absorption and stripping columns in the processes are shown in Table 4.1.1.

Table 4.2.1: Diameters and heights for strippers and absorbers.

Size parameters	**A-T-100**	**A-T-101**	**A-T-102**	**C-T-100**
Diameter [m]	3.5	0.8	0.3	0.5
Height [m]	10.8	3.0	3.0	3.0

The absorption and stripping columns are constructed of carbon steel.

4.2 Heat exchangers

A heat exchanger is equipment required to transfer heat from one fluid to another across a solid boundary (made of appropriate metals) between the two fluids. Coolers and heaters are specific types of heat exchangers. The equation:

$$Q = UA\Delta T_{lm} \tag{2.3.3.1}$$

can generally be applied to heat exchangers. The overall heat transfer coefficient is a function of several factors, among others the thermal conductivity of the heat transfer material and the fouling factor. Generally, the most corrosive fluid should be on the tube side to minimize the costs of expensive material (Sinnott & Towler 2009).

For the heat exchangers, the values of UA are found in UniSim, and are assumed to be correct. The values of U found in UniSim are too high compared to literature values. Because of this, more realistic values of the heat transfer coefficient are found in the literature for the different process streams (Skogestad, 2003, Sinnott & Towler, 2009). The areas of the reboilers and the condensers in the strippers are calculated by using the Equation 2.3.3.1 by taking the duty given in UniSim.

The heat exchangers with non-corrosive streams are constructed with carbon steel. For the heaters and coolers, the see water will be in the tubes, and thus they are constructed with a shell made of carbon steel and tubes made of stainless steel. The values of the overall heat transfer coefficients and the areas for the heat exchangers are given in Table 4.2.1.

Table 4.2.1: Values for the overall heat transfer coefficients and the calculated area of the heat exchangers for the two processes.

Heat exchanger	**U [W/m^2°C]**	**A [m^2]**
A-E-100	250	107.7
A-E-101	500	4.5
A-E-102	500	33.6
A-condenser	500	0.6
A-reboiler	1000	18.4
C-E-100	450	16670.4
C-E-101	500	384.2
C-E-102	730	190.8
C-E-103	730	13.0
C-E-104	730	21.2
C-condenser	500	13.4
C-reboiler	1000	69.0

4.2.1 Heat integration

To reduce the utility consumption of the heat exchangers, a heat integration network can be made, which combines the streams to be heated and cooled in the most economical way. In both processes, the only heater is the kettle reboiler in the stripper. Because of the large amount of cooling compared to heating in the processes, they are not suited for heat integration.

4.3 Compressors

A compressor is used to increase the pressure of a gas. In the two processes, a reciprocal compressor is used. This type of compressor is commonly used for a wide range of pressures and flow rates, as shown in Figure 2.3.4.1. The volumetric efficiency should be approximately 80 % (Vandenbrink et al., 2003).

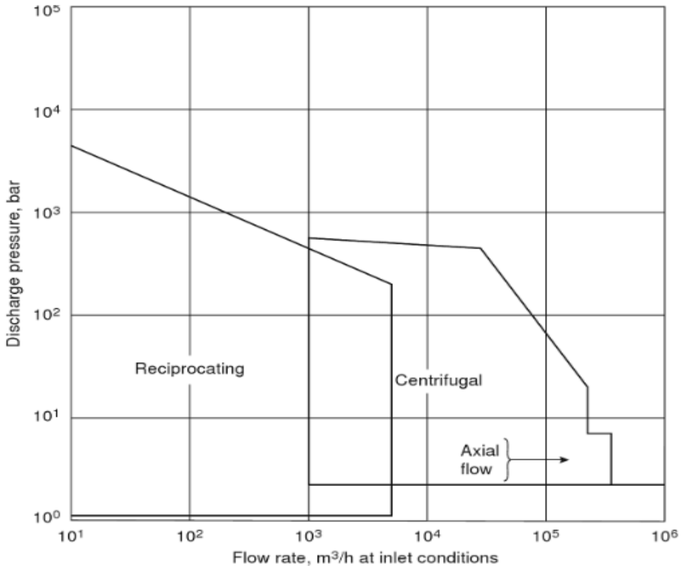

Figure 2.3.4.1: Operating ranges for reciprocating and centrifugal compressors (Sinnott & Towler, 2009).

The gases that are compressed in the processes are non-corrosive, and therefore the material used is carbon steel.

4.4 Pumps

The pumps are used to increase the pressure in a liquid flow. The pump used in the processes is a centrifugal pump, which is the most common type of pump.

The material used for pumps are carbon steel.

5. Economic estimation

The main reason for starting up a chemical plant, is to make money and an investment analysis has to be done to see if it is profitable before it can be assessed. To run an investment analysis, all the details of the economy has to be detected, basically the investment cost, the money required to run the plant and the revenues. As for this project, the dehydration process unit is a part of a bigger natural gas processing plant so the price of the gas coming into to the dehydration unit and going out of it, cannot be accounted for. The main purpose is then to look at the capital investment and the manufacturing costs, to see which is more profitable. The fixed capital investment and the working capital makes out the necessary total working capital needed to build the facility, start the plant up and run it until it starts earning income (Sinnott and Towler, 2009).

5.1 Fixed capital investment

The fixed capital investment, the investment needed to build the chemical plant, is a function of different factors. One has to take into account the direct project expenses, such as the cost and installation of equipment, and indirect project expenses consisting of transport, insurance, taxes, contractor engineering expenses and construction overhead. To find the grass root cost (C_{GR}) of the plant, which is the cost of a completely new facility developed on undeveloped land, two additional groups of costs are taken into account. The grass root represent the same as the fixed capital investment, FCI . These are contingency and fees, which covers unforeseen circumstances, and auxiliary facilities, such as the expenses to buy land, build infrastructure on the land, additional buildings needed and off sites such as storage space.

Many different methods for deriving the fixed capital costs are described in the literature (Sinnott & Towler, 2009, Turton et al., 1998). The equipment module costing technique is accepted as the best method for making preliminary cost estimates. This method relates costs back to the purchased cost of the equipment. For deviations from the purchased cost, a multiplying factor is used. This factor is based on equipment type, materials and operating pressure.

The purchased cost (C_p) of the equipment in carbon steel and at ambient pressures is given by Equation (5.1.1) and data in Appendix D. A is the capacity or sizing parameter of the equipment (Turton et al., 1998).

$$\log_{10} C_p = K_1 + K_2 \log_{10} A + K_3 (\log_{10} A)^2 \tag{5.1.1}$$

All cost estimating methods use historical data, and because of this a correction for inflation is needed. The correction is done based on published composite cost indices. The indices relate present cost to historical costs based on published data for labour, material and energy costs. By using the Chemical engineering plant cost index (CEPCI), Equation 5.1.2 can be used to correct for inflation.

$$\text{Cost in year A} = \text{Cost in year B} \cdot \frac{\text{Cost in year A}}{\text{Cost in year B}} \tag{5.1.2}$$

The bare module cost, C_{BM}^0, which is the direct and indirect costs for each piece of equipment, can be calculated by multiplying the purchased cost at the base conditions[*], C_P, with a bare module cost factor, F_{BM}^0, that takes into account the factors mentioned above, as shown in Equation 5.1.3.

$$C_{BM}^0 = C_p F_{BM}^0 \tag{5.1.3}$$

The bare module cost is a function of several factors, among them a factor correcting for deviations from ambient pressures and a factor correcting for material choices other than carbon steel. The bare module cost factor can be calculated by Equation 5.1.4.

$$F_{BM}^0 = C_p (B_1 + B_2 F_M F_P) \tag{5.1.4}$$

The pressure factors for heat exchangers are based on Equation 5.1.5 and data in Appendix D. P is the units of pressure given in barg.

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[*] Equipment made of the most common material (carbon steal) and operating at ambient pressures.

$$\log_{10} F_p = C_1 + C_2 \log_{10} P + C_3(\log_{10} P)^2 \qquad (5.1.5)$$

For process vessels a modified equation for the pressure factor is used, shown in Equation 5.1.6.

$$\begin{aligned} F_p &= 0{,}5146 + 0{,}6838 \log_{10} P + 0{,}2970(\log_{10} P)^2 \\ &+ 0{,}0235 * (\log_{10} P)^6 + 0{,}0020(\log_{10} P)^8 \end{aligned} \qquad (5.1.6)$$

For pumps the pressure factor can be written as shown in Equation 5.1.7.

$$F_p = C_1 + C_2 \log_{10} P + C_3(\log_{10} P)^2 \qquad (5.1.7)$$

In the two processes the cost of equipment is distributed differently. For the absorption process, the process vessels represent the highest equipment costs, while the heat exchangers represent the highest equipment costs for the cooling process. This is shown for the in Figure 5.1.1 and Figure 5.1.2, respectively.

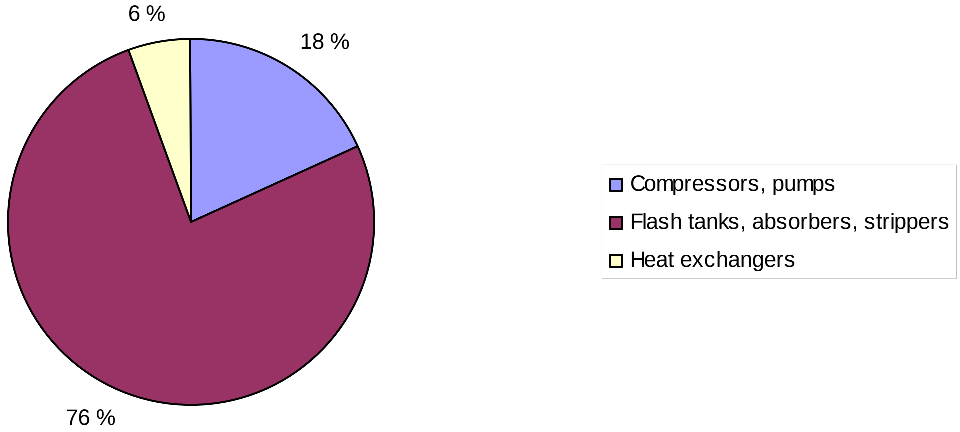

Figure 5.1.2: Distribution of material costs, C_{BM}^0 *, for the process of dehydration by absorption.*

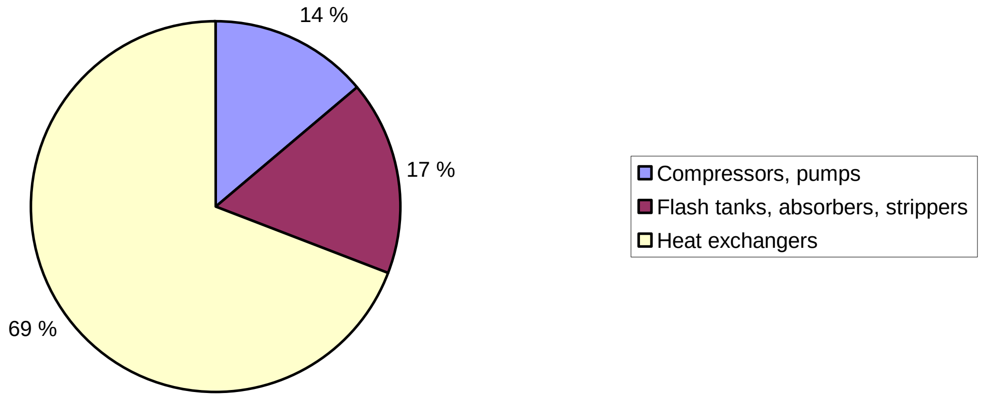

Figure 5.1.1: Distribution of material costs, C_{BM}^{0} *, for the process of dehydration by cooling.*

When the bare module costs for conditions deviating from the base conditions are found, the grass root cost can be calculated by adding two other groups of costs, besides the direct and indirect costs. By adding the contingency and fees required to build the plant and the costs of auxiliary facilities, the grass root cost mentioned above is calculated. The contingency and fees are assumed to be 15 % and 3 % respectively, and the auxiliary costs are assumed to be 35 %. The auxiliary costs are unaffected by the pressure and material used. Equation 5.1.8 shows the calculation of the grass root costs.

$$C_{GR} = 1,18 \cdot \sum_{i=1}^{n} C_{BM,i}^{\circ} + 0,35 \cdot \sum_{i=1}^{n} C_{BM,i} \qquad (5.1.8)$$

Calculations done to find the grass root cost are shown in Appendix D. The results in Table 5.1.1 show that the grass root for the cooling process is 41 % larger than for the absorption process.

Table 5.1.1: Grass root cost for both processes

Process	C_{GR} **[USD]**
Absorption	4 835 074
Cooling	8 161 828

5.2 Working capital

Working capital is the capital required to start up the plant and finance the first few months of operation before revenues from the process starts. Typically the working capital is 15-20 % of the fixed capital investment and for these plants 15 % is chosen because of the small size. The capital investment will be covered in the end of the project. Working capital for both plants is shown in Table 5.2.1.

Table 5.2.1: Working capital for both plants

Process	**FCI [USD]**	**Working capital [USD]**
Absorption	4835074	5560335
Cooling	8161828	9386103

5.3 Manufacturing costs

Before a proposed process can be assessed, costs associated with day-to-day must be estimated. To estimate the manufacturing costs we need the information from the flow diagram, the fixed capital investment and an estimate of the numbers of operators required. In contrast to the capital costs that are expressed in USD, the manufacturing costs are expressed in USD per unit time.

To estimate the operating costs a technique based on an approach given by Ulrich (Turton et al., 2009) is used: It is consistent with the module approach used to evaluate the capital plant cost.

Manufacturing costs can be divided into three categories:

1. Direct operating costs: vary with production rate

    - Raw materials
    - Utilities
    - Labour

- Direct supervisory
- Maintenance and repairs
- Operating supplies
- Laboratory charges
- Patents and royalties

2. Fixed manufacturing costs: independent of changes

    - Depreciation
    - Local taxes and insurance
    - Plant overhead costs

3. General expenses: represents an overhead burden that is necessary to carry out business.

    - Administration costs
    - Distribution and selling
    - Research and development

Cost of manufacture can be found Equation 5.3.1:

$$COM = 0{,}304FCI + 2{,}73C_{OL} + 1{,}23(C_{UT} + C_{RM}) \tag{5.3.1}$$

Where FCI is the fixed capital investment, C_{RM} is the raw material cost, C_{OL} is the operating labour cost, and C_{UT} is the utility cost. The total manufacturing cost is calculated from Equation 5.3.1. The cost for each of the categories listed above is found in the next section.

Table 5.3.1: Cost of manufacture for both processes

Process	***COM* [USD/year]**
Absorption	6372066
Cooling	11785151

5.3.1 Direct manufacturing costs

Raw materials, C_{RM}

In both processes the feed is 20000 kmol/h of natural gas from pipelines. Since this is only a part of a bigger plant, it is not possible to estimate a reasonable price in and out of the plant. Anyhow, the product streams are not the same for both plants. This will be discussed under section 5.4.

In both plants, TEG is used and regenerated. It is assumed that the amount of TEG in the regeneration cycles is small compared to the make-up stream. Propane is not taken into account in the cooling cycle, because it is a closed cycle and the amount inside the cycle is considered low compared to total operating costs. The price of 98.5 wt% TEG is 1780 USD/ton (I7).

Cost of operating labour, C_{OL}

The number of operators, N, for each process is estimated by Equation 5.3.1.1 found in Turton et al., 1998,

$$N = \sqrt{6,29 + 31,7p^2 + 0,23U} \qquad (5.3.1.1)$$

where U is the number of main units in the process and p is the number of steps including solids. P is zero for both processes.

The absorption process has 9 main equipments and requires 3 operators. The cooling process has 10 main equipments and requires 3 operators.

A number of 5 shifts is assumed, and the wage for one operator is assumed to be 330000 NOK/year that equals 57300 USD/year.

Cost of utilities, C_{UT}

- Electric power
- Cooling water
- Steam for heating

The electricity price for manufacturing was found by using the average price of 2010. This was found to be 0.34 NOK/kWh, which equals 0.06 USD/kWh (I8). This is assumed throughout the estimation. Electric power is used for compressors and pumps.

Cooling water systems usually use power for pumping the cooling water through the system. The power used in a recirculating cooling water system is usually around 0.39 kWh/m^3 (Sinnott & Towler, 2009). In both of the processes it is assumed that, since the plants are by the sea, the sea water are pumped up from the sea at 10 °C and then pumped out to the sea afterwards at maximum 25 °C.

Steam is used in the reboilers in the strippers for regeneration of TEG. It is assumed that the dehydration unit is supplied with high pressure (HP) steam from another part of the NG-plant or another industrial producer so the steam production unit is not considered. The price was 8.65 USD/1000 kg in 1998, so the price in 2011 will be approximately 13.01 USD/1000 kg when using the Chemical Engineering Cost Plant Index (CEPCI). Costs of utility per year for both processes are listed in Table 5.3.1.1.

Table 5.3.1.1: Utility costs.

Utility	**Absorption [USD/year]**	**Cooling [USD/year]**
Electric power	70814	78998
Cooling water	12619	46860
HP steam	2007454	5574893
Sum utility	2090887	5700750

The rest of the direct manufacturing cost is dependent on the operating labour cost, fixed capital investment, raw material cost, or utility price. The correlations are taken from Turton et al., and shown in Table 5.3.1.2.

Table 5.3.1.2: Equations for calculating direct manufacturing cost.

Type of cost	**Correlation**
Direct supervisory	$0{,}18\,C_{OL}$
Maintenance and repairs	$0{,}06\,FCI$
Laboratory charges	$0{,}15\,C_{OL}$
Operating supplies	$0{,}09\,FCI$
Patents and royalties	$0{,}03COM$

5.3.2 Fixed manufacturing costs

Total fixed operating costs include depreciation, local taxes and plant overhead costs. Overhead costs can also be referred to as factory expenses and includes employee benefits, like medical services, cafeteria etc. Equation 5.3.2.1 is used for finding the total fixed manufacturing costs. It is taken into account a depreciation of 10 %.

$$0{,}708C_{OL} + 0{,}168FCI \tag{5.3.2.1}$$

5.3.3 General manufacturing costs

General manufacturing costs are costs that are not directly related to the manufacturing process, but associated with management level and other administrative activities. The equations for calculating these are given in Table 5.3.3.1.

Table 5.3.3.1: Equations for calculating general manufacturing costs.

Type of cost	**Correlation**
Administration	$0{,}177C_{OL} + 0{,}009FCI$
Distribution and selling cost	$0{,}11COM$
Research and development	$0{,}05COM$

5.3.4 Total manufacturing costs

The summary of total manufacturing costs for each category is listed in Table 5.3.4.1. Note that all of these values are corrected for inflation. The CEPCI was 382 in 1998 when Turton et al. was written, and is 574.6 in February 2011 (Chemical Engineering, June 2011).

Table 5.3.4.1: Summary of total manufacturing costs.

Manufacturing costs	**Absoption cycle [USD/year]**	**Cooling cycle [USD/year]**
Raw materials	57 144	0
Utilities	2090063	5700750
Labour	828 256	839572
Direct supervisory	149086	151123
Maintenance and repairs	290104	489710
Operating supplies	43516	73456
Laboratory charges	124238	125936
Patents and royalties	191162	353555
Fixed manufacturing costs	1398698	1965604
Administration cost	190117	222061
Distribution and selling	700927	1296367
Research and development	318603	589258

The total manufacturing cost is distributed as shown in Figure 5.3.4.1 for the absorption process and Figure 5.3.4.2 for the cooling process. In the absorption process, the manufacturing costs are approximately evently distributed, while the direct manufacturing costs accounts for 65 % of the total manufacturing costs.

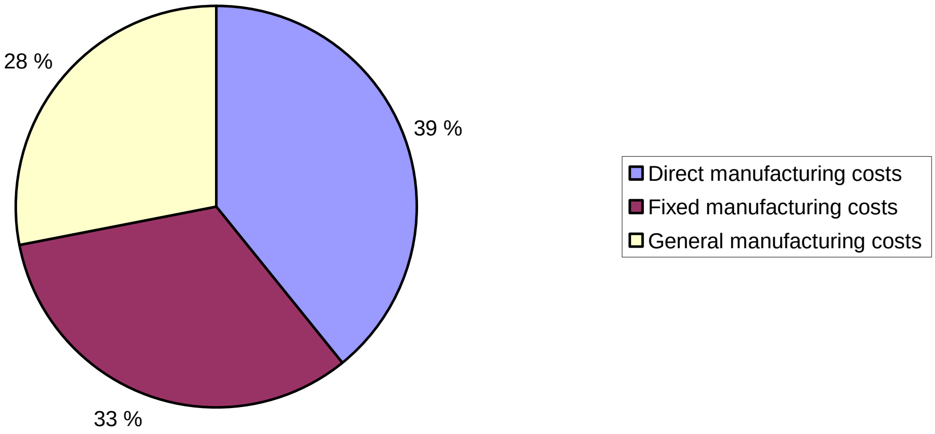

Figure: Distribution of the direct, fixed and general manufacturing costs for the absorption process.

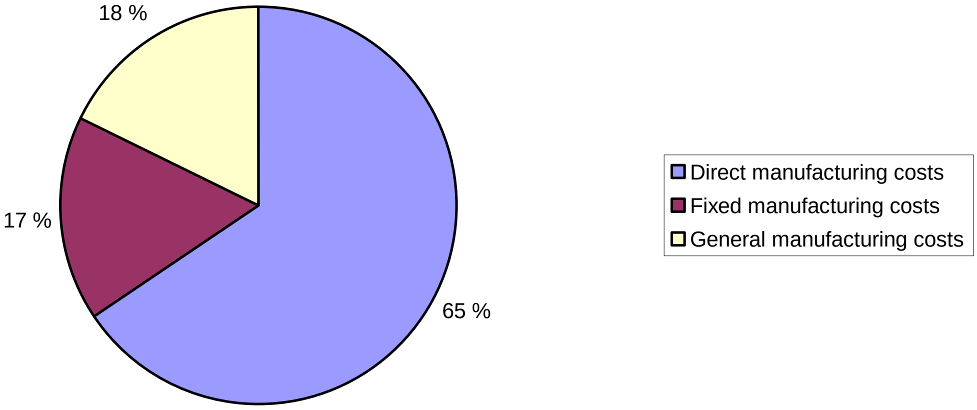

Figure: Distribution of the direct, fixed and general manufacturing costs for the cooling process.

5.4 Revenues

For the cooling process, the waste gas contains natural gas and can be sold or used as a fuel in the plant. The value of this is 13400 USD/year when a price of 5 USD/MMBtu is used, which is 50 % of natural gas price (P1, P2), Appendix D. The absorption has no waste besides water.

The process gases have different properties and also different economical values. If we look at the lower heating value of the two processes and convert it to MMBtu/year, the process gas from the absorption has a value of 278000 USD/year more than the cooling process, see Appendix E.

6. Sensitivity Analysis

Sensitivity analysis is usually conducted to study the impact of variations in the projections on the feasibility of a project (Sinnott & Towler, 2009). The sensitivity analysis shows how deviations in electricity price (including HP steam-price) and in TEG-price affect the total costs of manufacturing. As for the TEG-price, it is only done for the absorption process, since all the raw material costs for the cooling cycle are neglected. The prices are increased and decreased with 10 %. The results of the sensitivity analysis of electricity prices are presented in the Table 6.1.

Table 6.1: Sensitivity for electricity prices.

Table: Utility prices affect on the manufacturing costs.

Process	**100%**	**Deviation +10 % (USD/year)**	**Deviation, -10 % (USD/year)**	**Impact, %**
Absorption	6373080	6630259	6115901	+/- 4.0
Cooling	11785151	12486343	11083959	+/- 5.9

The results of the sensitivity analysis of TEG prices are presented in the Table 6.2.

Table 6.2: Sensitivity for TEG-prices.

Process	**100%**	**Deviation +10 % (USD/year)**	**Deviation, -10 % (USD/year)**	**Impact, %**
Absorption	6373080	6380108	6366051	+/- 0.11

In the first quarter of 2003 the electricity price was as high as 0.624 NOK/kWh, which corresponds to 0.108 USD/kWh (I8). This is 80 % higher than today's value of 0.34 NOK/kWh. That would have an impact on the total manufacturing costs for absorption and cooling of 30 % and 48 % increase, respectively. In Figure 6.1 the development in the electricity prices over the last 13 years are shown.

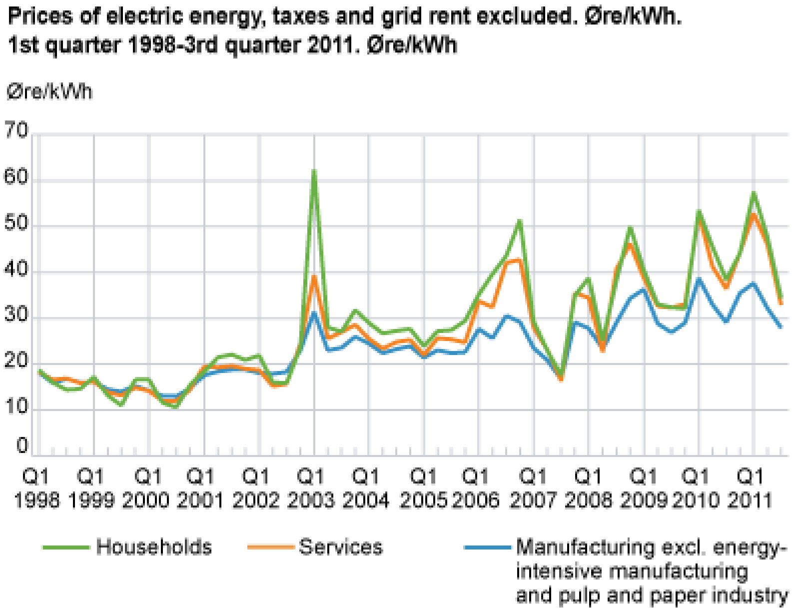

Figure 6.1: Electricity prices in Norway for the last 13 years (I9).

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7. Discussion

There are different ways of removing water from natural gas. Two of such processes are absorption with TEG and cooling with a refrigeration cycle. The main purpose of this project is to compare these two methods by simulating in UniSim, optimizing and economical study. The water content of the processed gas has to be 20 ppm or lower.

By looking at the simulations, one can see that the absorption process has no loss of natural gas, while the cooling cycle has some natural gas in the waste gas. About 48 wt% of the waste gas contains hydrocarbons, and was used as an energy resource. In the absorption process, there was no need for any waste treatment due to the fact that all of the potential waste streams were recycled.

Both processes have a regeneration of TEG, however the cooling cycle has a larger molar flow that requires more cooling water and HP steam are very energy demanding. The purity of TEG in the recycle in the absorption process needs to be high, and for this reason, an additional stripping column is added in the simulation.

In this project, the recycle rate of TEG was considered the most important optimization variable for the absorption process. The recycle of TEG was minimized to give the lowest flow possible, while still reaching the given specification. By recycling the TEG, the purity in the stream increased from 88.7 mole% to 99.4 mole%. Because of this, the required molar flow of the TEG in the recycle was lowered. In addition, low TEG losses in the recycle resulted in a small make up stream of 0.03 kmol/h. In the cooling process, the optimization of the TEG recycle was done in the same way as in the absorption process. The loss of TEG in the recycle was in this case so small that the TEG needed in the make up was $4.9 \cdot 10^{-7}$ kmol/h. The molar flow of propane in the refrigeration cycle was also optimized.

Based on the economical analysis, the absorption process was considered the most economical, with regards to both total capital investment and the total manufacturing costs. The cooling process has a high number of heat exchangers, because of large temperature differences. Heat exchangers accounted for 69 % of the total equipment cost of the process. For the absorption process it accounts for 6 % of the total equipment cost. Also the sensitivity analysis shows that the cooling process was more affected by increase of utility prices. An increase of 10 % in the utility price will yield an increment of 5.9 % in the manufacturing cost for the cooling process and 4.0 % for the absorption process.

The direct manufacturing costs for the cooling process accounts for 65 % of the manufacturing costs as opposed to 39 % for the absorption process. This is an indication that a potential future scale up will be more profitable with the absorption process than the cooling process.

8. Conclusion and recommendation

From the economic analysis it is evident that dehydration by absorption is the most economical method of dehydration both in terms of the total capital investment and manufacturing costs.

From the sensitivity analysis, the cooling process is more sensitive variations in direct manufacturing cost.

The impact of the choice of fluid package in the simulation of the processes on the overall economics of the two processes should be investigated.

9. List of symbols

Symbol	**Unit**	**Description**
A	m^2	Area
c_p	kJ/kg°C	Spesific heat capacity
D_v	m	Minimum allowable diameter
D_c	m	Diameter of stripping/absorption column
h	m	height
J	Nm	Energy
l_t	m	Tray spacing
P	bar	Pressure
Q	J/s	Heat energy
T	°C	Temperature in celsius
U	W/m^2K	Overall heat transfer coefficient
u_t	m/s	Settling velocity
V	m^3	Volume
V_l	m^3/s	Liquid volumetric flow
V_v	m^3/s	Vapour volumetric flow
W	J/s	Watt
ΔT_{lm}	°C	Logarithmic mean value
ρ_v	kg/m^3	Vapour density
ρ_l	kg/m^3	Liquid density

10. References

Arnold, K., Stewart, M., “Design of Gas-Handling Systems and Facilities”, 2nd edition, 1999, Elsevier

Benítez, J., “Principles of Modern Applications of Mass transfer operations”, 2nd edition, John Wiley & Sons, Inc., Hobroken, New Jersey, 2009.

Brathen A. Development of Processes for natural gas drying: further exploring the TEG injection concept, MSc. Thesis, NTNU, Trondheim, Norway, 2008

Caroll J., “Natural Gas Hydrates; A Guide for Engineers”, Second Edition, Elsevire, 2009

Chemical Engineering, June 2011, Vol. 118, No. 6, page 64

Dinçer, İ., Kanoglu, M., “Refrigeration systems and applications”, 2nd edition, 2010, U.K, Wiley

“GPSA SI Engineering Data Book”, Gas Processors Suppliers Association, Tulsa, USA, 2004.

Guo B., Ghalambor A., Natural gas engineering handbook, Gulf publishing company, 2005

Katz, D. L., Cornell, D., Kobayashi, R., Poettmann, F. H., Vary, J. A., Elenbaas, J. R., Weinaug, C. F., ”Handbook of Natural Gas Engineering”, 1959, McGraw-Hill Book Company, USA

Kirk-Othmer, “Kirk-OthmerEncyclopeadia of chemical technology”, vol. 12, John Wiley & Sons, 4th Edition, 2001, pg. 171.

Mokhatab S., Poe W. A., Speight J. G., Handbook of natural gas transmission and processing, Elsevier

“SI Engineering Data Book”, Gas Processors Suppliers Association, Tulsa, USA, 1980.

Sinnott, R. and Towler, G., “Chemical engineering design”, 5th edition, 2009, Elsevier,USA.

Skogestad, S., “Chemical and Energy Process Engineering”, 2009, CRC press, USA.

Turton, R., Bailie, R. C., Whiting, W. B. and Shaeiwitz, J. A., “Analysis, Synthesis and Design of Chemical Processes”, 1998, Prentice Hall PTR, USA.

Turton, R., Bailie, R. C., Whiting, W. B. and Shaeiwitz, J. A., “Analysis, Synthesis and Design of Chemical Processes”, 2009, Prentice Hall PTR, USA.

Vandenbrink, J. W., Potter, R. C, Huhn, D. M., “Design of a Refrigeration Cycle Evaporator Unit”, 2003

Internet references:
I1: http://205.254.135.24/iea/overview.html
I2: http://www.naturalgas.org/overview/uses.asp
I3: http://www.naturalgas.org/overview/uses_industry.asp
I4: http://www.instituteforenergyresearch.org/energy-overview/fossil-fuels/
I5: http://www.ipt.ntnu.no/~jsg/undervisning/naturgass/lysark/LysarkPettersen2011A.pdf
I6: http://sites.poli.usp.br/d/pqi2408/BestPracticesOptimumSimulationsHYSYSPropertyPackages.pdf
I7: http://chemicalland21.com/petrochemical/teg.htm
I8: http://www.ssb.no/english/subjects/10/08/10/elkraftpris_en/arkiv/tab-2011-04-01-04-en.html
I9: http://www.ssb.no/elkraftpris

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Appendix A: UniSim flow sheets with stream properties

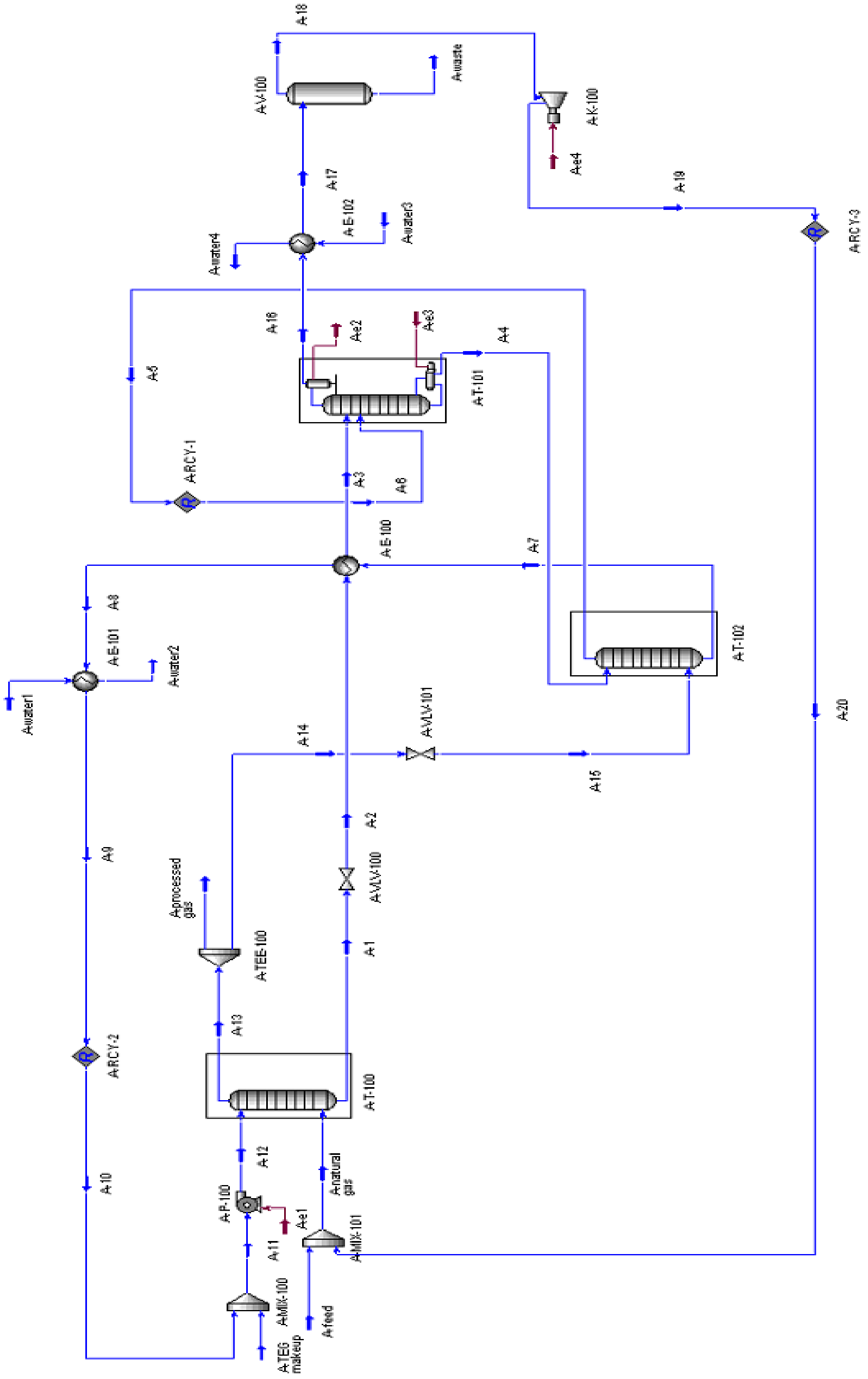

Figure A.1: UniSim flow sheet for the absorption process.

Table A.1: Properties of feed streams, product streams, waste streams and recycle streams for the absorption process.

Name	A-feed	A-TEG makeup	A-processed gas	A-waste	A-10	A-20
Vapour fraction	1.0000	0.0000	1.0000	0.0000	0.0000	1.0000
Temperature [C]	50.00	55.00	52.93	60.00	55.00	549.3
Pressure [kPa]	7000	200.0	7000	100.0	200.0	7000
Molar flow [kmole/h]	20000	0.02780	19920.4	79.62	70.99	19.67
Heat flow [kJ/h]	-1.468e+9	-2.175e+4	-1.444e+9	-2.257e+7	-5.583e+7	-1.477e+6
Mole fractions						
Methane	0.8860	0.0000	0.8895	0.0000	0.0009	0.6772
Ethane	0.0370	0.0000	0.0371	0.0000	0.0001	0.0332
Propane	0.0150	0.0000	0.0151	0.0000	0.0000	0.0145
i-Butane	0.0030	0.0000	0.0030	0.0000	0.0000	0.0023
n-Butane	0.0050	0.0000	0.0050	0.0000	0.0000	0.0039
Nitrogen	0.0500	0.0000	0.0502	0.0000	0.0001	0.0688
Water	0.0040	0.1126	0.0000	1.0000	0.0051	0.1992
TEG	0.0000	0.8874	0.0000	0.0000	0.9937	0.0009

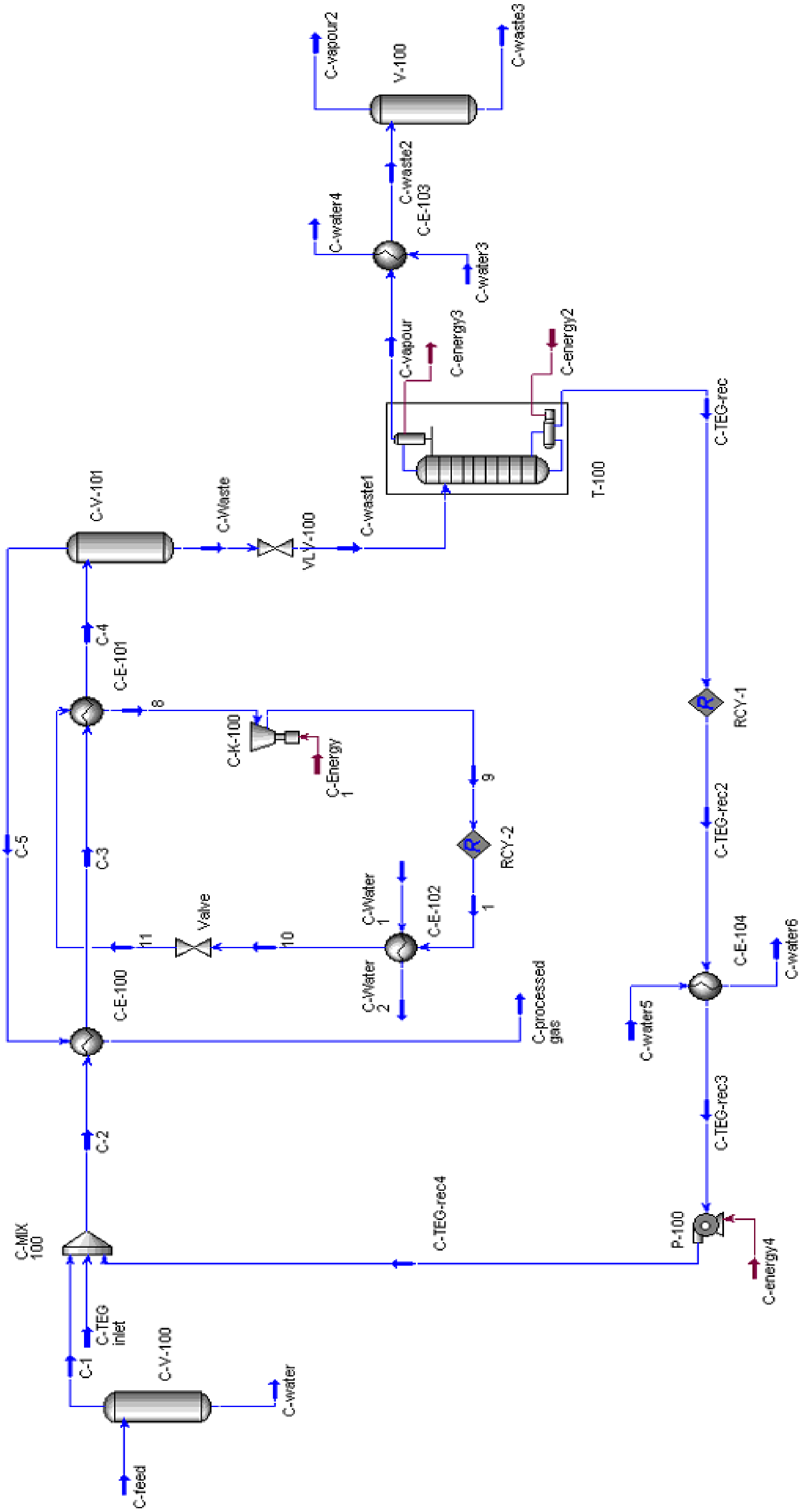

Figure A.2: UniSim flow sheet for the cooling process.

Table A.3: Properties of feed streams, product streams, waste streams and recycle streams for the cooling process.

Name	C-feed	C-TEG inlet	C-processed gas	C-waste3	C-vapour2	C-rec4
Vapour fraction	0.9984	0.0000	1.0000	0.0000	1.0000	0.0000
Temperature [C]	50.00	50.00	49.07	-3.00	60.00	48.82
Pressure [kPa]	7000	7000	7000	7000	100	7000
Molar flow [kmole/h]	20000	4.89e-7	19920	123.9	1.576	76.00
Heat flow [kJ/h]	-1.465e+9	-0.3590	-1.445e+9	-7.173e+7	-1.356e+4	-5.709e+7
Mole fractions						
Methane	0.8860	0.0000	0.8896	0.0051	0.4033	0.0000
Ethane	0.0150	0.0000	0.0151	0.0003	0.0247	0.0000
Propane	0.0370	0.0000	0.0371	0.0007	0.0579	0.0000
i-Butane	0.0030	0.0000	0.0030	0.0000	0.0011	0.0000
n-Butane	0.0050	0.0000	0.0050	0.0000	0.0021	0.0000
Nitrogen	0.0500	0.0000	0.0502	0.0039	0.3074	0.0000
Water	0.0040	0.1126	0.0000	0.4255	0.2033	0.0000
TEG	0.0000	0.8874	0.0000	0.5643	0.0000	0.9200

Appendix B: Entalpy-Pressure diagram

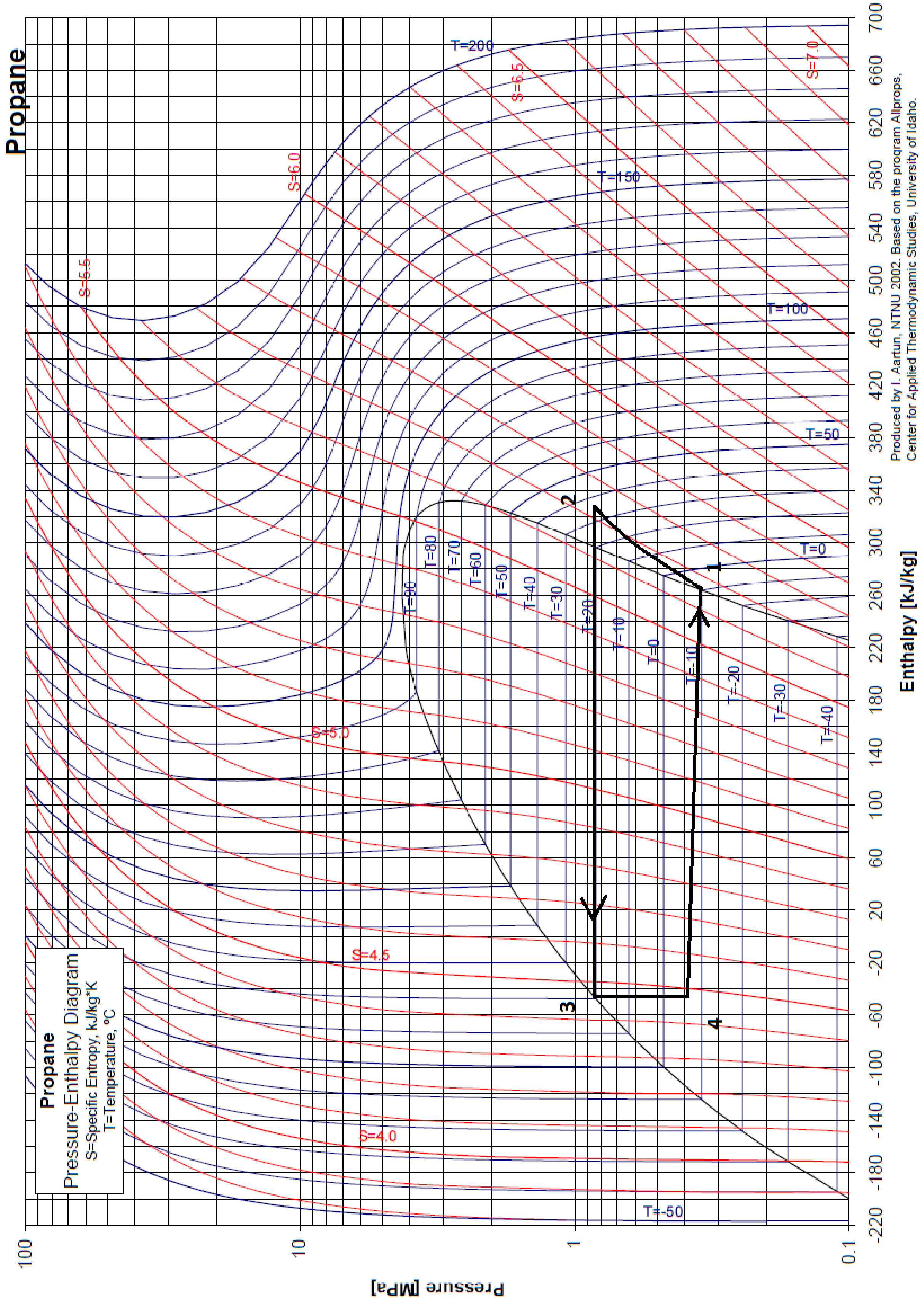

Figure B.1: Entalpy- Pressure diagram for propane.

Appendix C: Size calculations

Flash tanks

All the flash tanks are assumed to be vertical, and with demister pads.

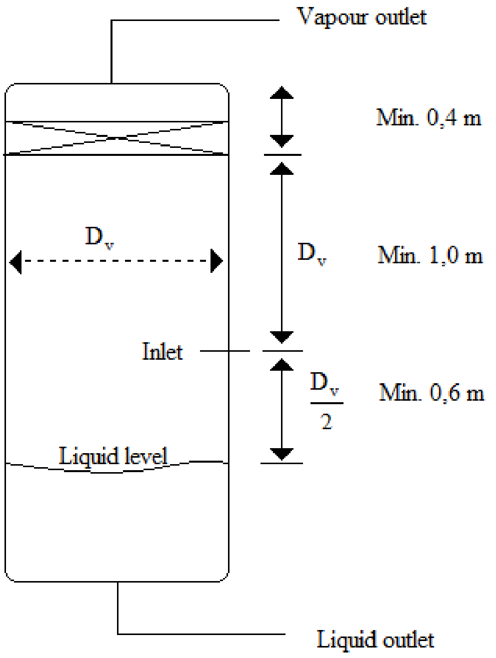

Figure C.1: Sketch of how to calculate the height of a flash tank

The settling velocity [m/s] of the liquid droplets is given by

$$u_t = 0,07\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

where ρ_L and ρ_V is liquid and vapour density [kg/m^3], respectively.

Minimum allowable diameter [m] is calculated by

$$D_v = \sqrt{\left(\frac{4}{\pi}\frac{V_v}{u_s}\right)}$$

where V_V is the volumetric flowrate [m^3/s] of the vapour, and u_s is equal to u_t if a demister pad is used. The liquid level depends on the hold-up time, which in this case is assumed to be 10 minutes. It is then possible to calculate the volume held in vessel by the equation

$$\text{Volume held in vessel} = V_L \cdot \text{Hold-up time}$$

where V_L is the liquid volumetric flow rate [m^3/s]. The liquid depth required, h_v [m], can be found by

$$h_v = \frac{\text{Volume held-up}}{\text{Vessel cross-sectional area}}$$

The height of the flash tank is calculated by adding all the parts as showed in Figure C.1.

Table C.1: Physical values for the flash tanks from the UniSIM-simulations.

Physical values	**C-V-100**	**C-V-101**	**C-V-102**	**A-V-100**
Liquid density, ρ_L [kg/m^3]	990.5097	1145.862	980.4253	980.4573
Vapour density, ρ_V [kg/m^3]	52.46504	72.4565	0.787625	0.6698831
Liquid volumetric flow, V_L [m^3/s]	0.000174	0.00654	0.000236	0.00406
Vapour volumetric flow, V_V [m^3/s]	1.89	1.37	0.0121	0.143

Absorption and stripping columns

The diameter of the distillation columns (which includes absorbers and strippers), are calculated based on maximum allowable vapour velocity, u_V [m/s]:

$$u_v = (-0{,}171 l_t^2 + 0{,}27 l_t - 0{,}047)\left[\frac{\rho_L - \rho_V}{\rho_V}\right]^{1/2}$$

where l_t [m] is the plate spacing. The tray spacing depends on the diameter of the column (Benítez, 2009). The diameter of the column can then be calculated from the equation:

$$D_C = \sqrt{\frac{4V_w}{\pi \rho_V u_v}}$$

where V_w is the maximum vapour rate [kg/s]. The overall column efficiency is defined as:

$$E_O = \frac{\text{number of ideal stages}}{\text{number of real stages}}$$

and is assumed to be 70 %. The number of ideal stages is found in UniSim, and from the number of real stages, the height of all the stages in the column can be found from:

$$H_{trays} = N_{\text{real trays}} \cdot l_t$$

The overall height of the column is the height of the trays and the head. The height of the head of the column is assumed to be 20 % of the height of trays.

$$H_{total} = 1,20 \cdot H_{trays}$$

Table C.2: Values required to calculate the column height of the strippers and absorbers.

Physical values	**A-T-100**	**A-T-101**	**A-T-102**	**C-T-100**
Liquid density, ρ_L [kg/m^3]	1096.79	967.131	981.961	937.604
Vapour density, ρ_V [kg/m^3]	51.7591	0.594131	0.743946	0.591619
Tray spacing, l_t [m]	0.75	0.50	0.50	0.50
Maximum vapour rate, V_w [kg/s]	99.25	0.4945	0.1558	0.2413
Number of real trays, $N_{\text{real trays}}$	12	5	5	5

Heat exchangers

The area of the heat exchangers are calculated from the value of UA found in UniSim.

$$A = \frac{UA}{U}$$

Table C.3: The values of UA for the heat exchangers.

Heat exchanger	**UA [W/K]**
A-E-100	26925.72
A-E-101	2231.84
A-E-102	16812.35
C-E-100	27006047
C-E-101	691527
C-E-102	501394
C-E-103	34060
C-E-104	55740

Reboilers and condersers

$$m = \frac{Q}{c_p \Delta T}$$

Table C.4: Values needed to calculate the molar flow of water in the reboilers and condensers

Equipment	**Q [kJ/h]**	**c_p [kJ/kmolC]**	**ΔT [C]**	**m [kmol/h]**
A-Reboiler	4302498	98.85	40	22095.82
A-Condenser	105624	77.79	10	135.78
C-Reboiler	10582434	98.85	40	2676.39
C-Condenser	1952059	77.79	10	1672.93

Appendix D: Equipment costs

The purchased costs of the different equipment C_p are calculated from the following equation if nothing else is given specifically.

$$\log_{10} C_p = K_1 + K_2 \log_{10} A + K_3 (\log_{10} A)^2$$

where A is the capacity or size parameter. It is corrected for pressure using the following equation

$$\log_{10} F_p = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2$$

where P is the pressure given in barg (1 bar = 0.0 bar). The bare module cost is calculated by the equation

$$C_{BM}^{0} = C_p F_{BM}^{0} = C_p (B_1 + B_2 F_M F_P)$$

Heat exchangers

The material factor F_M for carbon steel (CS) is 1.0.

Table D.1: Correlation factors for calculating the bare module cost corrected for deviation from ambient pressure and material choice for heat exchangers.

Type of exchangers	K_1	K_2	K_3	C_1	C_2	C_3	B_1	B_2
Shell and tube	3.2138	0.2688	0.07961	-0.06499^2	0.05025	0.01474	1.8	1.5
Kettle Reboiler	3.5638	0.1906	0.11070	-0.06499^2	0.05025	0.01474	1.8	1.5

Example: C-E-100:

Area= 17686 m^2

$$C_p = 10^{K_1 + K_2 \log_{10} A + K_3 (\log_{10} A)^2}$$

$$C_p = 10^{3.2138 + 0.2688 \log_{10} 16670 + 0.07961 (\log_{10} 16670)^2} = 874175$$

$$F_p = 10^{C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2}$$

$$F_p = 10^{-0.06499^2 + 0.05025 \log_{10} 69 + 0.01474 (\log_{10} 69)^2} = 1.4011$$

$$C_{BM}^0 = C_p F_{BM}^0 = C_p (B_1 + B_2 F_M F_P)$$

$$C_{BM}^0 = C_p F_{BM}^0 = 874175 \cdot (1.8 + 1.5 \cdot 1 \cdot 1.4) = 78018$$

Process vessels:

$$\begin{aligned} F_p = {} & 0.5146 + 0{,}6838 \log_{10} P + 0.2970 (\log_{10} P)^2 \\ & + 0.0235 (\log_{10} P)^6 + 0.0020 (\log_{10} P)^8 \end{aligned} \quad \text{for } 3.8 \langle P \langle 400 \text{ barg}$$

$$F_p = 1 \text{ for } -0.5 \langle P \langle 3.7 \text{ barg}$$

$$F_M = 1 \text{ for CS}$$

Table D.2: Correlation factors for calculating the bare module cost corrected for deviation from ambient pressure and material choice for process vessels.

Diameter	K_1	K_2	K_3	B_1	B_2
0.3	3.3392	0.5538	0.2851	1.62	1.62
0.5	3.4746	0.5893	0.2053	1.62	1.47
2.0	3.9484	0.4623	0.1717	1.62	1.47
2.5	4.0547	0.4620	0.1558	1.62	1.47

Height is the size parameter.

Process Vessel Internals:

Regarding sieve trays for absorber and stripper and demister pads for the separators.

$$C_p = 235 + 19.80D + 75.07D^2$$

where D is the diameter of the vessel.

$$C_{BM}^0 = C_p F_{BM} N F_q$$

Material factor F_M for trays and demister pads is 1.2 for CS.

Where F_q is a factor depending on numbers of trays. N .

Table D.3: Correction factor for number of trays

N	F_q
1	3.0
4	2.5
7	2.0
10	1.5

Pumps with electric drives

Table D.4: Correlation factors for calculating the bare module cost corrected for deviation from ambient pressure and material choice for pumps.

Type of pump	K_1	K_2	K_3	C_1	C_2	C_3	B_1	B_2
Centrifugal	3.5793	0.3590	0.05577	0.1682	0.3477	0.4841	1.8	1.51

The pressure factor is given as. The capacity parameter is the shaft work (kW)

$$F_p = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2$$

Compressors

Table D.5: Correlation factors for calculating the bare module cost for compressors.

Type of compressor	K_1	K_2	K_3
Reciprocating	2.9945	0.8542	0

Material factor F_M for reciprocating compressor in CS is 2.9.

$$C_{BM}^0 = C_p F_{BM}$$

Drives for compressors

Table D.6: Correlation factors for calculating the bare module cost for drives.

Type of drive	K_1	K_2	K_3
Gas Turbine	3.4171	0.6112	0

$$C_{BM}^0 = C_p F_{BM}$$

Table D.7: Equipment cost for the absorption process.

Equipment	**Size parameter**	C_p **(USD)**	C_{BM}^0 **(USD)**	C_{BM} **(USD)**
A-E-100	102.6 m^2	17935	78018	59186
A-E-101	6.7 m^2	4646	20212	15333
A-E-102	32.1 m^2	9482	41248	31292
A-P-100	24 kW	18015	59628	98921
A-K-100	126 kW	149667	434036	434036
A-K-100 drive	126 kW	50154	175538	175538
A-V-100	5.3 m	12370	52200	52200
A-V-100 (demister)	0.3 m	245	883	883
A-T-100	10.8 m	93412	2613789	394197
A-T-100 (trays)	3.5 m	1813	39413	16422
A-T-101	3.0 m	12616	53239	53239
A-T-101 (trays)	0.8 m	294	4859	3230
A-T-101 condenser	0.6 m^2	1542	4313	3007
A-T-101 reboiler	18.7 m^2	14115	63488	46581

A-T-102	3.0 m	7009	29576	29576
A-T-102 (trays)	0.3 m	375	6000	6000
SUM			**3676440**	**1419641**

Table D.8: Equipment cost for the cooling process.

Equipment	**Size parameter**	C_p **(USD)**	C_{BM}^0 **(USD)**	C_{BM} **(USD)**
C-E-100	16670 m^2	874175	3410718	2884779
C-E-101	384m^2	41135	160493	135745
C-E-102	191 m^2	25987	122695	85759
C-E-103	13 m^2	6100	26535	20130
C-E-104	21 m^2	7662	33332	25286
C-V-100 (demister)	2.86 m	1351	4862	4862
C-V-100 (column)	4.70 m	40713	302134	125804
C-V-101 (demister)	2.6 m	1159	4172	4172
C-V-101 (column)	5.0 m	42400	314650	131016
C-V-102 (demister)	0.1 m	354	1274	1274
C-V-102 (column)	29.5 m	87565	271850	270577
C-T-100 (trays)	0.5 m	399	6383	6383
C-T-100 (column)	3.0 m	9471	39969	29266
C-T-100 (reboiler)	69 m	14167	61626	46751
C-T-100 (condenser)	13 m	6184	26902	20409
C-K-100	142 kW	156681	391703	391703
C-K-100 drive	142 kW	80526	281841	281841
C-P-100	25 kW	22975	101092	101092
Sum			**5562231**	**4566846**

The grass root cost is given by the following equation.

$$C_{GR} = 1{,}18 \cdot \sum_{i=1}^{n} C_{BM,i}^{\circ} + 0{,}35 \cdot \sum_{i=1}^{n} C_{BM,i}$$

Table D.9: Grass root costs

Process	C_{GR} **USD**
Absorption	4 835 074
Cooling	8 161 828

Appendix E: Product and waste

Table E.1 shows the difference in income per year for the two processes. The calculations are based on the lower heating value and a natural gas price of 10 USD/Mbtu given by statoil.

Table E.1: Difference in income for the two processes.

Process	**[Btu/year]**	**[USD/year]**
Absorption	121954631616000	1219546316
Cooling	121926864168000	1219268641

The absorption process will have a larger income than the cooling process. The difference is 277674 USD/year.

Only the cooling cycle has waste in the form of hydrocarbons. These hydrocarbons can be used as fuel. The value of the waste is assumed to be 50 % of the natural gas. The waste value is shown in Table E.2.

Table E2: Value of the wate that can be used as fuel in the cooling process.

Process	**[Btu/year]**	**[USD/year]**
Cooling	2683656941	13418