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PROJECT TITLE:

Simulation, Design and Optimal Operation of LNG Process for Arctic Conditions

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ABSTRACT

In this study a precooling model of the LNG process has been built in UniSim. The C3MR process has been used as reference when building the model. In this paper it is studied what effect large variation in the ambient temperature, which is the case in the Arctic region, has on the precooling part of the LNG process. The Shtokman field in the north western part of Russia has been used as base case and temperature variations of $\pm 30^{\circ}\text{C}$ have been assumed. MATLAB has been used for the optimization part of the study.

In the study ten different cases have been simulated and optimized; five cases where different mixtures of ethane and propane (mole fractions depending on the ambient temperature) are used as precooling medium and five cases where propane is used, for comparison. The optimization using MATLAB turned out to be difficult due to noise and uncertainties in the UniSim model. However, the main focus in this report is to show the trends and the overall picture, not the exact optimal operating point for ten different cases. There has been showed that it would be beneficial to use a mixture of ethane and propane when the ambient temperature gets below 0°C .

A degree of freedom analysis has been performed on the system. The analysis shows that we have six degrees of freedom in the optimization; two pressures, three flows and one DOF related to the ethane content in the precooling refrigerant.

A self-optimizing control structure has not been proposed for the system, but central issues have been discussed.

Key Words: LNG, precooling, Arctic region, Shtokman, varying ambient temperature, DOF analysis

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1 INTRODUCTION

1.1 MOTIVATION

The LNG technology has been extensively studied in many papers, but there is less work done on LNG production in the Arctic region, at least in the open literature. When energy companies are looking for new areas to find hydrocarbons, the Arctic region looks very promising; about 22% of the undiscovered (recoverable) petroleum resources in the world are located in the Arctic region [4].

For most of the LNG plants in the world air is used as coolant to condense the precooling medium [1]. LNG production using ambient air as coolant in the Arctic region imposes many new aspects that must be taken into consideration because of the large temperature swings during a production year. Most LNG plants in the world use the C3MR technology where propane is used as precooling medium [2], but it could be beneficial to use a mixed refrigerant also for the precooling section when there are large ambient temperature variations.

The basis for the study is the Shtokman field in the Russian sector of the Barents Sea. Ambient temperature variations of $\pm 30^{\circ}\text{C}$ are used in the study.

1.2 GOAL

The goal of the study is to see if there could be savings in energy when the ambient temperature is getting low by using mixed refrigerant in the precooling section instead of only using pure propane. UniSim is used for design and simulation, while MATLAB is used for optimization. To limit the scope only mixtures of ethane and propane has been considered. Existing models of the liquefaction part of the process has been used to be able to do a complete comparison.

2 BACKGROUND

2.1 THE ARCTIC REGION AND THE SHTOKMAN FIELD

The Arctic region is the area north of the Arctic Circle. This area amount to about 6% of the Earth's surface area or about the size of the African continent. The Arctic consists of continental shelves under less than 500 meters of water, ocean deeper than 500 meters, and land, each contributing with about one-third of the total area. About 61 large oil and natural gas fields (> 500 MMBOE) have been discovered (October 2009) in the Arctic region. 43 of these are located in Russia, while the remaining are shared in between Alaska (6), Canada's Northwest Territories (11) and Norway (1). 15 of the 61 discovered fields have not gone into production yet. [3]

However, most of the petroleum resources in the Arctic region are undiscovered. According to [4], these resources account for about 22% of the undiscovered (recoverable) resources in

the world. The Arctic accounts for about 30% of the undiscovered natural gas. In other words, the potential for future LNG production in the Arctic region is high.

The Shtokman field was discovered in 1988 and located in the central part of the Russian sector of the Barents Sea. The field is about 600 kilometres northeast of Murmansk, and the sea depths vary in between 320 and 340 metres. The methane content of the field is 3.9 trillion cubic metres and 56 million tons of gas condensate, making it one of the largest natural gas deposits in the world [4]. The annual gas production rate is estimated to be 70 billion cubic metres of natural gas, which is comparable to the annual gas production in Norway. The produced gas will be used for LNG production at the coast of the Teriberka Bay and also distributed via pipelines into the Unified Gas Supply System of Russia [5]. The target market for the LNG are consumers in the USA and Europe [6].

2.1 REFRIGERATION CYCLES

A refrigeration cycle is a type of heat pumping system where, as for air conditioning, heat removal is the desired effect. On the contrary, a heat pump has heat supply as the desired effect. A common feature of the three systems is that they absorb heat at low temperature and reject heat at a higher temperature.

The vapour compression cycle is the most common refrigeration cycle [7]. The cycle is shown in figure 2.1.

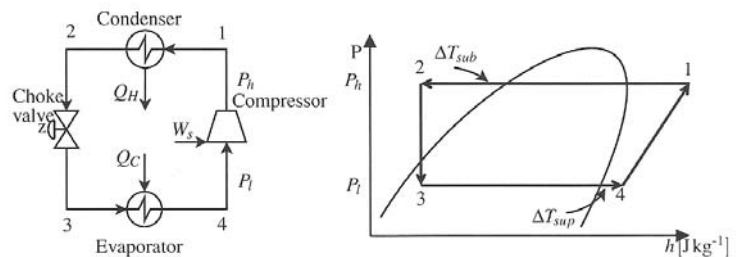


Figure 2.1: Vapour compression cycle [8]

As indicated in the figure the refrigeration cycle has four states. From state 3 to state 4 the refrigerant is evaporated and superheated in the evaporator. This is done by exchanging heat with the fluid to be cooled. The reason why the refrigerant is superheated is because there are practical issues associated with compressing a gas which contains liquid droplets. After superheating the refrigerant is compressed in a compressor from state 4 to state 1. The pressure is increased in the compressor and the refrigerant is then condensed in the condenser from state 1 to state 2 by exchanging heat with the hot source (e.g. air or sea water). Finally, from state 2 to state 3, the liquid is expanded in a Joule-Thompson valve to get a two-phase mixture at a low temperature. The refrigerant is further described in Section 2.2.3.

Coefficient of performance (COP) is often used as a measure of efficiency of a vapour compression cycle and is given by

$$COP = \frac{h_4 - h_3}{h_1 - h_4} = \frac{Q_C}{W_S} \quad (1.1)$$

where Q_C is the cooling duty or heat removed from the system by the refrigerant and W_s is the compressor work.

2.2 LIQUEFIED NATURAL GAS

Liquefied Natural Gas (LNG) is, as the name indicates, cooled and condensed natural gas at approximate atmospheric pressure and about -160°C . Depending on the natural gas feed the mixture consists of mainly methane (85-95 mole%) and heavier hydrocarbons. The mixture often contains some nitrogen as well.

When the distance from the gas field to existing piping infrastructure is too large, building a complete new pipeline is often not economical feasible. LNG is produced to make the transportation of natural gas more convenient; liquefying the gas reduces its volume by a factor of approximately 600 and the product can be shipped by LNG carriers or transport trucks. Another benefit is the greater flexibility to choose customers when the natural gas is not transported in a pipeline. The initial investment costs are high for LNG, but the transport costs are lower compared to pipelines [2].

2.2.1 Production Principles

The natural gas must be pretreated to remove impurities (nonhydrocarbons and water) before it can be cooled and liquefied. Heavier hydrocarbons are also removed. A block diagram of a typical LNG process is presented in figure 2.2.

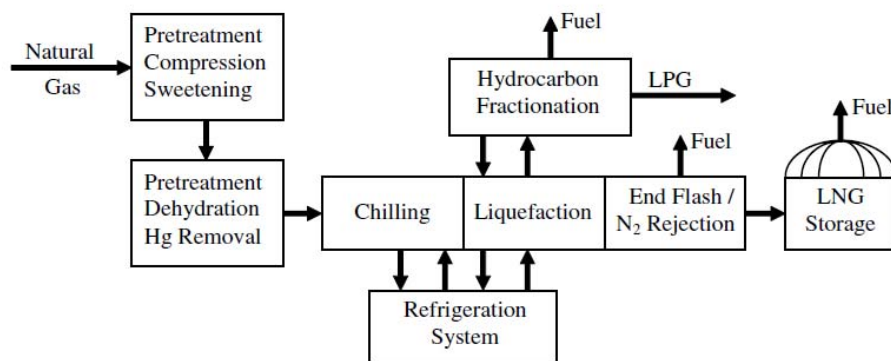


Figure 2.2: Typical LNG process [9]

The refrigeration and liquefaction of the natural gas are the key elements of the LNG technology and consume about 35% of the capital expenditures and 50% of the operational costs [9]. The liquefaction of natural gas can be divided in three stages: precooling, liquefaction and subcooling. Figure 2.3 shows the three stages where the red arrows indicate the heat transfer.

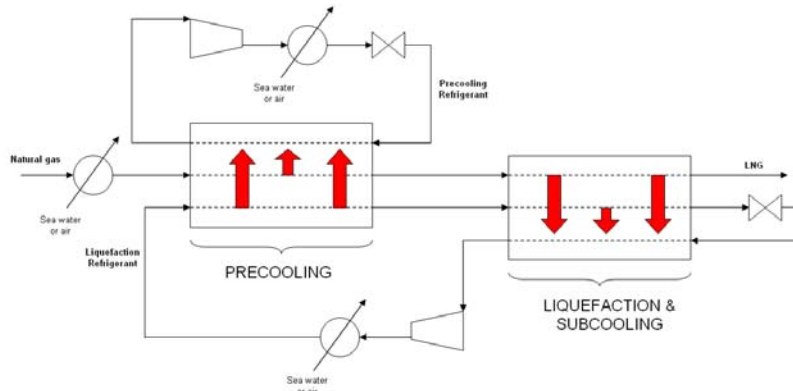


Figure 2.3: Simplified LNG process

The process could also be presented in a pressure-enthalpy diagram, as shown in figure 2.4. Assuming inlet conditions for the natural gas of 60 bara and 20°C, the gas is pre-cooled to about -50°C, condensed, liquefied and subcooled to about -150°C. During this cooling process there is only a small pressure drop related to transport in pipes and heat exchangers, so the pressure must be let down to about atmospheric pressure after subcooling. This is done in the end flash process with a turbine and a Joule-Thomson valve. The turbine lets down the pressure to almost the boiling point to ensure that there is not a two-phase flow within the turbine. The end flash is important to remove nitrogen from the natural gas and it is often used as fuel gas. [10]

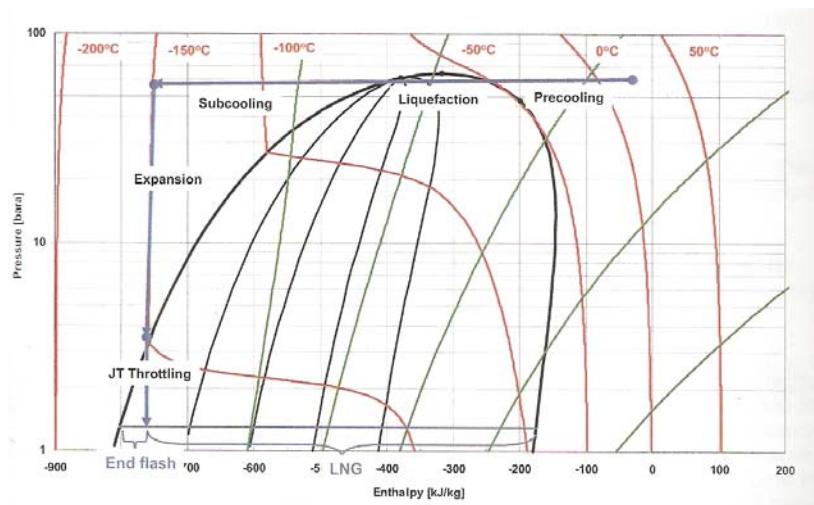


Figure 2.4: Pressure vs Enthalpy diagram [2]

2.2.2 Production Technologies

There are several different production technologies for liquefying natural gas, ranging from the simple PRICO process to complicated cascade processes using mixed refrigerants like Statoil-Linde's mixed-sluid cascade (MFC) [11]. In this study the C3MR process has been used as reference since the goal is to study how large the energy savings are when an ethane/propane mixture is used compared to using only propane as precooling medium. The C3MR process is the most used LNG technology in the world and has been described in many

papers, for instance [2] and [13] give a good overview of the process. Therefore, only a brief introduction is given here.

In figure 2.5 a simplified flow scheme of the C3MR process is shown. The precooling part consists of three (can also be two or four) heat exchangers in series; E1A, E1B and E1C. The precooling medium is superheated in the heat exchangers to ensure no liquid is introduced to the compressors. After compression to the saturation pressure in C1 the precooling medium is condensed by heat exchanging with sea water or air in CW1. The temperature is decreased in a J-T valve, the two-phase flow is separated and the liquid is introduced to the heat exchanger. Both the natural gas and the mixed refrigerant (MR) for the liquefaction part are pre-cooled. After precooling the mixed refrigerant is a two-phase flow. The MR is separated in D1 and the heavier liquid fraction is subcooled in E2A and used to liquefy the natural gas, while the lighter vapour fraction is condensed and used to subcool the natural gas. The two heat exchangers E2A and E2B shown in the scheme are actually one spiral wound heat exchanger. [2]

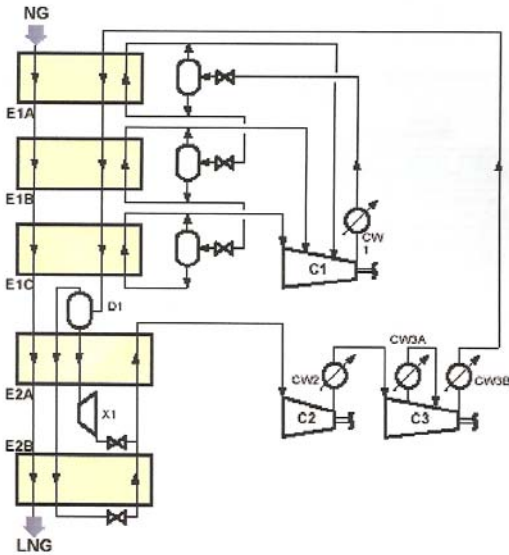


Figure 2.5: Simplified C3MR process [2]

Shell developed the Dual Mixed Refrigerant (DMR) process where mixed refrigerant is also used in the precooling section, instead of only propane as in C3MR. Other differences are only two (or one) pressure levels and two SWHEs in series in the precooling part. The liquefaction part is very similar to C3MR [2].

2.2.3 Cooling Medium and Refrigerants

Air is normally used as cooling medium, but if available, sea water is also considered in the design phase. Also, a combination could be possible, where air is used for de-superheating and water as the final precooling medium [1]. It should also be taken into consideration that sea water is heated in the precooling section and this could lead to some environmental issues.

The refrigerant can be selected among many fluids, but the fluid should have some desired properties. The refrigerant should have a high latent heat of vaporization to limit the refrigerant flow and the density of the gas introduced to the compressors should be high to

limit the volumes. The working pressures should not be below atmospheric pressure and not too high. The refrigerant should also be environmental friendly, have a low cost and good availability. [10]

In this study different mixtures of propane and ethane are considered as precooling medium. Since the precooling part would not require cooling down to lower temperatures than -50°C to -60°C propane/ethane mixtures are sufficient. Propane and ethane have good availability as well since they both can be separated from the feed.

Why should we use mixed refrigerants? If a single component refrigerant is used this is evaporating at a constant temperature, while a mixture will evaporate at gliding temperatures. Since the temperature is increasing during heat exchange with natural gas the average temperature lift will be lower for a mixed refrigerant compared to a single component refrigerant. This can be explained through the *thermodynamic average temperature*, T_m , which is defined as

$$T_m = \frac{h_2 - h_1}{s_2 - s_1} \quad (1.2)$$

where the h is the enthalpy at state 1 and 2 and s is the entropy at state 1 and 2. It can further be shown that

$$COP = \frac{T_{mL}}{T_{mH} - T_{mL}} \quad (1.3)$$

where T_{mL} and T_{mH} is the thermodynamic average temperature of heat rejection (refrigerant) and heat supply (natural gas/mixed refrigerant), respectively. Thus, when T_{mL} increases due to gliding temperatures, the temperature lift will decrease and the COP will increase. For further details about the thermodynamic average temperature it is referred to [10].

When using only propane in the precooling section the split temperature, i.e. the natural gas temperature *out* of the precooling section and *in* to the liquefaction section, is limited by the normal boiling point (NBP) of propane, which is -42.1°C . The minimum temperature split is typically limited to -35°C for a process using propane as precooling medium. This is to make sure the pressure is not expanded below atmospheric and to avoid vacuum suction on the compressor [10]. The minimum temperature approach must also be taken into consideration ($\approx 5^{\circ}\text{C}$). Also, the flow rate might get too low and lead to surge problems. This would require recycling of the refrigerant and further lead to an efficiency loss.

When the ambient temperature is low, e.g. -30°C , the load on the precooling section becomes low with a resulting high *relative* work load on the liquefaction/subcooling section. By relative work load it is meant the work in the liquefaction (or precooling) part compared to the *total* work in the process. Having a balanced work load allow for better machinery selection between the precooling and the liquefaction part [13].

If the LNG plant was water cooled, pure ethane could be considered for a plant in the Arctic region. Assuming a sea water temperature of 10°C the condensing pressure for ethane is approximately 30 bar, according to figure 2.6. Taking a ΔT_{\min} into consideration the condensing pressure would be a few bars higher. For air cooled plants propylene could be

considered [14]. One issue with propylene is that it would have to be imported to the plant site (for make-up purposes) since it does not exist in the natural gas feed.

To achieve a balanced work load a mixture of refrigerants should be considered. Ethane has a NBP of -88.7°C , which means the split temperature could be lowered if an ethane/propane mixture is used. Reducing the split temperature means we are “transferring” work from the liquefaction part to the precooling part. Also, the split temperature is important for extraction purposes as well. The colder the split temperature is, the higher can the recovery of heavy hydrocarbons be [11].

Figure 2.6 shows the vapour pressure as a function of temperature for different fluids. It is seen that both propane and ethane are natural choices for the precooling part, while methane and/or nitrogen should be included for the liquefaction/subcooling part.

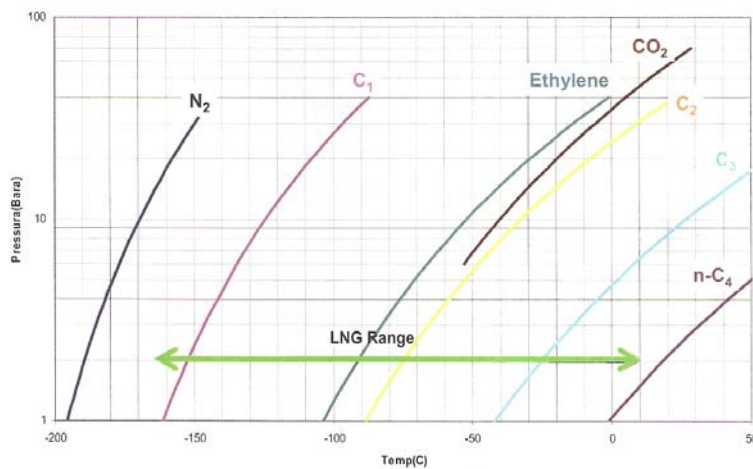


Figure 2.6: Pressure vs temperature for relevant refrigerants [2]

In table 2.1 ethane and propane properties are listed. From the table we can see that ethane has a critical temperature of 32.2°C which is much lower compared to propane. Using pure ethane would therefore be more practical for water-cooled plants where the seasonal variations are smaller.

Table 2.1: Ethane and Propane Properties

Property	Ethane	Propane
Molar mass [g/mol]	30.07	44.1
Normal boiling point [$^{\circ}\text{C}$]	-88.7	-42.2
Critical temperature [$^{\circ}\text{C}$]	32.2	96.6
Critical pressure [bar]	48.8	42.5

It should also be mentioned that other types of heat exchangers are used when mixed refrigerants are used instead of pure refrigerants. In C3MR, shell-and-tube, or kettle, heat exchangers are commonly used because of their low cost. However, when mixed refrigerants are used a special design is required to fully utilize the potential of the refrigerant effect of the multicomponent fluid. Plate-fin or spiral wound heat exchanges can then be used. [13]

2.2.4 LNG Production in the Arctic Region

When producing LNG in the Arctic region there are several new aspects that must be taken into consideration. When air is used as cooling medium the temperature variations can be very large. In extreme cases the differences can be above 60°C from summer to winter compared to a few degrees in difference when sea water is used. Figure 2.7 shows typical temperature variations for the Arctic region, both for air and sea water. These large variations in ambient air temperature affect the precooling section of the LNG plant since this is where most of the heat is rejected to ambient air. Other factors that should be taken into consideration are the harsh conditions in the Arctic region making the working environment less comfortable. An ice-free port is a requirement for the LNG carrier. Icing prevention in air cooled heat exchangers and gas turbines are also issues that must be taken into consideration [1].

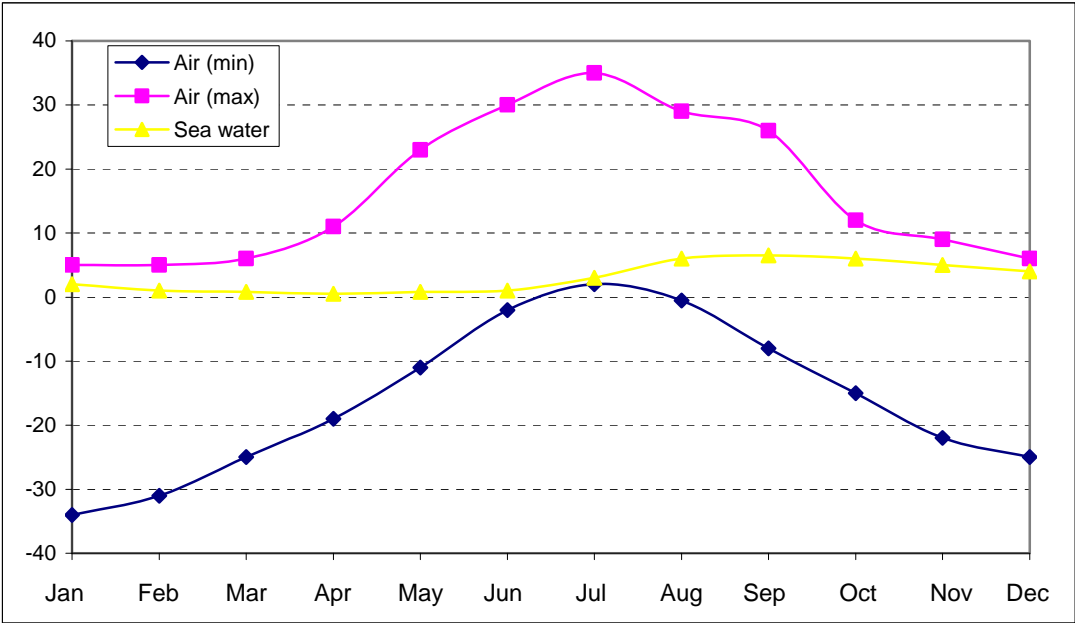


Figure 2.7: Typical air and sea water temperatures in the Arctic region [10 (modified)]

3 MODEL DESCRIPTION

A precooling model has been built in UniSim and is presented in figure 3.1.

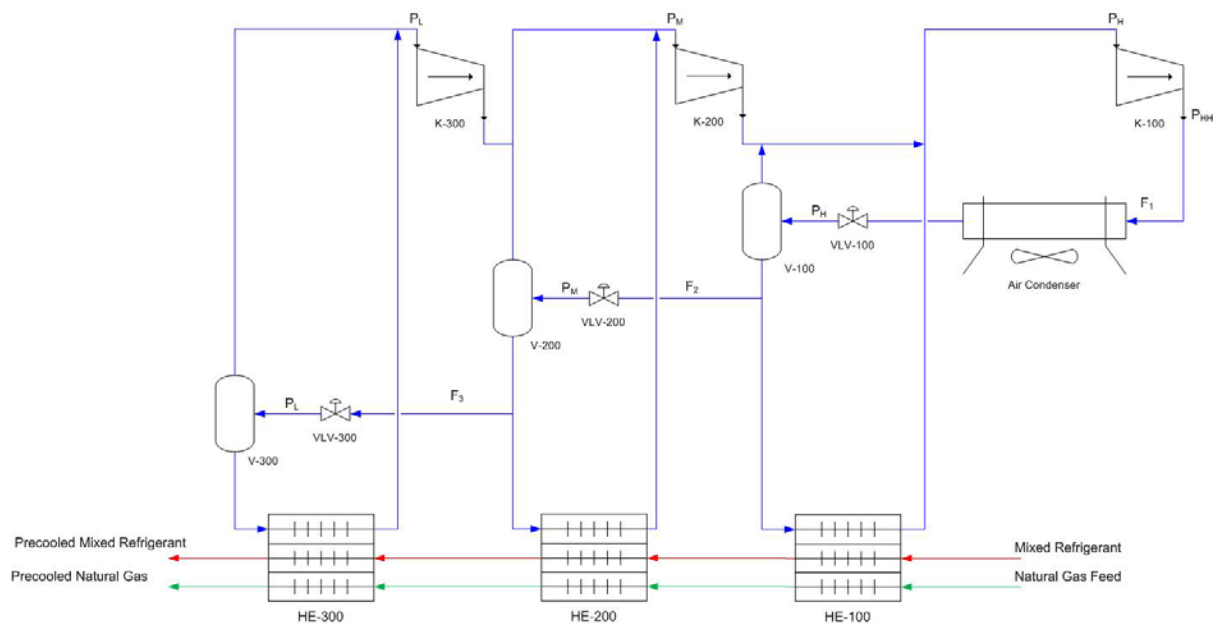


Figure 3.1: Precooling model

When building the model there some assumptions were made. The natural gas feed conditions are presented in table 3.1, along with the MR conditions.

Table 3.1: NG and MR conditions

Parameter	Natural Gas	Mixed Refrigerant
Flow rate [MTPA]	5	10
Flow rate [kmole/s]	9.4678	18.9336
Max. inlet temperature [°C]	50	50
Min. inlet temperature [°C]	-10	-10
Pressure [bar]	60	45

The model is similar to the precooling part of the C3MR process, but with the possibility to vary the precooling medium, i.e. vary the ethane/propane ratio. Introducing, for instance, methane in the precooling medium was not considered. To limit the scope of the study ten different cases has been studied. Five cases with various amount of ethane and five cases with pure propane for comparison. The ethane content varies in between 40mole% and 80mole%, where the lowest ethane content is for the highest ambient temperature (+30°C) and the highest ethane content is for the lowest ambient temperature (-30°C). This is about the same as what is used on Sakhalin-II in east Russia [20].

The natural gas flow rate is 5 MTPA or approximately 9.5 kmole/s. The mixed refrigerant from the liquefaction part has been set to be twice as large as the natural gas stream. This is the assumed amount needed to liquefy the natural gas in the liquefaction part of the process. When the split temperature (the NG/MR temperature out of the precooling section) is kept constant it is reasonable to keep the flow rate ratio of NG and MR constant. The feed gas pressure is 60 bar and the The MR pressure is 45 bar. The approximate natural gas composition for the Shtokman field is given in table 3.2 together with the MR composition.

Table 3.2: Mole fraction for NG and MR

Component	Natural Gas	Mixed Refrigerant
Methane	0.95	0.475
Ethane	0.012	0.425
Propane	0.008	0.02
n-Butane	0.002	0
n-Pentane	0.001	0
Nitrogen	0.024	0.08
CO ₂	0.003	0

Both the natural gas and MR inlet temperature are set to be 20°C above ambient temperature. This is because a ΔT_{\min} of 15°C for the condenser is used and taking into consideration that the air is heated during heat exchange with the precooling medium (cross flow), a total temperature difference of 20°C is used as a basis. The highest pressure (P_{HH}) is set to be the condensing pressure. In practice it will be higher to make sure total condensation is achieved also when disturbances occur. A small degree of subcooling could be optimal [8], but this is not considered in this study.

For all the three heat exchangers a ΔT_{\min} of 5°C is used for the design case. The design case is at $T_{\text{amb}} = 0^\circ\text{C}$. The design temperature is selected to make sure the annual plant capacity is achieved [1]. The idea is that the production gain during low ambient temperatures should compensate for the production loss during high ambient temperature, if the natural gas feed rate could be adjusted. However, because the UA-values are not large enough and the pressure drop is increased when the flow rate exceeds the design flow rate, the gap between production loss and gain can not be fully compensated. One solution is to increase the design temperature, but this has not been considered in this study. In this case the natural gas feed is kept constant, which means the design case could be for the worst case scenario, i.e. at $T_{\text{amb}} = 30^\circ\text{C}$.

The LNG heat exchangers were added to allow for multiple streams. Another option is to split the precooling medium and use six heat exchangers; three for the natural gas stream and three for the mixed refrigerant stream. A flow sheet of this setup, made by co-Supervisor Magnus G. Jacobsen, is presented in Appendix D. It turns out that the two models are close to equivalent when comparing the COP; both models operating with a COP close to three when using similar conditions ($T_{\text{cond}} = 30^\circ\text{C}$).

The pressure drop through each of the three heat exchangers in figure 2.8 is assumed to be 0.5 bar for the natural gas. There is assumed to be no pressure drop for both MR and PMR through the heat exchangers.

The weighted rating method is used to calculate the LMTD (log mean temperature difference) and the UA values. This is the only option when using more than two LNG sides [15]. With this method the LMTD and UA are calculated for each heat curve interval and then summed to get the overall UA. The step type is set to 'Equal Enthalpy' where each interval has an equal enthalpy change and the pressure profile is set to have a constant dP/dH .

4 DEGREES OF FREEDOM AND OBJECTIVE FUNCTION

According to [17], the first step in a systematic approach to plantwide control (further described in Chapter 6) is to formulate the operational objectives. The cost function J must be defined and minimized with respect to the N_{ss} optimization degrees of freedom, subject to some constraints. N_{ss} equals the number of steady-state degrees of freedom. This number is the same as the number of dynamic manipulated variables N_{MV} minus the number of degrees of freedom with no steady-state effect, N_0 . N_0 are typically number of liquid levels that need to be controlled, but also valves that are used to improve the dynamic response affecting the steady-state (e.g. heat exchanger bypass). Thus, $N_{ss} = N_{MV} - N_0$.

It is important to find the number of degrees of freedom (DOF) in a process. This number tells us how many degrees of freedom that are available for us to solve the optimization problem. It also determines how many steady-state controlled variables that must be selected and kept constant.

The potential degrees of freedom in the model are presented in table 4.1.

Table 4.1: Potential degrees of freedom

Unit/stream	Variable	DOF
NG/MR Feed	Flow	2
Splitter	Flow	2
Compressor speeds	rpm	3
Choke valves, outlet pressure	Pressure	3
Active charge	-	1
Heat exchangers	Temperature	6
Air flow in condenser	Flow	1
Composition	-	1
Total		19

The potential degrees of freedom add up to 19. It might not be obvious that the active charge is a degree of freedom. The active charge in the cycle can be modified by introducing a tank with variable holdup, and the pressure level in the system depends on the charge [8]. The compressor duties have three degrees of freedom; one for each compressor and the choke valves has three pressure drops as degrees of freedom. The heat exchangers (with no bypasses) have six degrees of freedom; two for each exchanger (two UA values), when considering the design case. When considering the operational case there are no degrees of freedom for the heat exchangers. The condenser has one degree of freedom and this will always be set to max duty. The precooling MR composition has two components, i.e. one degree of freedom.

In table 4.2 the number of specifications in UniSim are presented.

Table 4.2: Specifications in UniSim

Unit	Specification
Pressure	4
Temperature	7
Mixer (Equalize all)	2
Precooling MR flows	3
NG/MR Feed	2
Composition	1
Total	19

There are in total set 4 pressures in the model. One in the high pressure end of the process, one after the first choke valve, one after the second choke valve and one after the third choke valve. The 7 temperatures comes from setting three temperatures because we want the NG temperature and MR temperature to be equal out of each heat exchanger and three temperatures because we are defining ΔT_{\min} for each heat exchanger for the design case. For the operational case there are no degrees of freedom in choosing temperatures out of the heat exchangers. In addition we are setting the lowest temperature on the precooling MR. The Mixer Equalize all sets two pressures; one after the low pressure compressor and one after the intermediate pressure compressor. There are three flows set for the precooling MR; one at the high pressure side, one after the first split and one after the second split.

The total degrees of freedom are 19, but we loose 13 of them:

- 6 temperatures related to the heat exchangers because during operation these are not DOFs
- 1 temperature because the condenser duty is set to max
- 2 compressor speeds because the mixers are set to 'equalize all'
 - The outlet pressure of K-300 and K-200 will then be set by the pressure out of VLV-200 and VLV-100, respectively.
- 2 related to the NG and MR flows which are set constant (disturbances)
- 1 pressure (P_{HH}) because saturated liquid is assumed out of the condenser
- 1 pressure (P_L) because the lowest temperature is set to -55°C and thus the pressure as well

We are then left with 6 degrees of freedom; 2 pressures, 3 flows and 1 DOF related to amount of ethane. To simplify the optimization procedure the ethane content was not included as a degree of freedom in the optimization. Thus, only the two pressures (P_H and P_M) and the three flow rates (F_1 , F_2 and F_3) were used as DOFs in the optimization.

The objective function to be minimized in the optimization is given by equation 1.4.

$$\begin{aligned} \min_u W_s \\ \text{subject to } c \leq 0 \end{aligned} \quad (1.4)$$

where W_s is the total compressor work in the precooling section and c are the constraints. Here it is assumed that there is no cost for air cooling, no income from turbine work and the feed and production rates are equal.

The constraints are further discussed in the next section. The MATLAB code is presented in appendix E.

5 RESULTS

5.1 Design

Using a $\Delta T_{\min} = 15^{\circ}\text{C}$ for the air cooled condenser and a $\Delta T_{\min} = 5^{\circ}\text{C}$ for all the other three heat exchangers, a design has been made in UniSim. The design has been based on an ambient air temperature of 0°C since this is approximately the average temperature. Taking into consideration that the air will also be heated during heat exchange with the precooling medium, a total temperature difference of 20°C in between air and precooling medium is used as a basis. In the design case the precooling medium (PMR) is a mixture of 60mole% ethane and 40mole% propane. For comparison a design was also made with pure propane at the same ambient temperature.

In figure 5.1 the UniSim flow sheet model is presented.

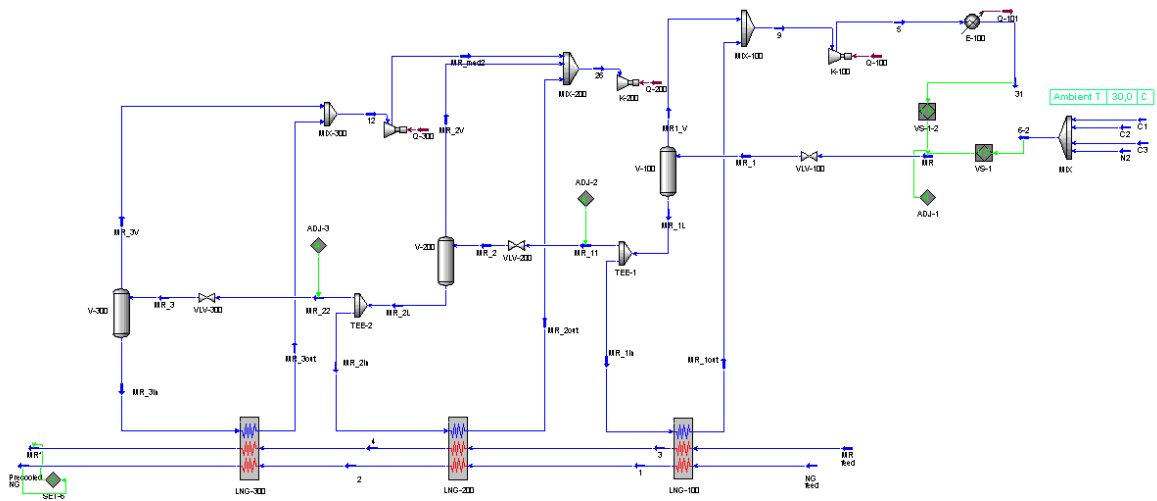


Figure 5.1: UniSim flow sheet model

In the design case there are 5 degrees of freedom since the ethane content is set; two pressures and three flow rates. First the UniSim flow sheet was manually adjusted to make the flow sheet converge. Initially, the pressures were first optimized “manually” in UniSim using Case Studies. This was to be able to generate a curve which relates the pressures to the objective function (showed in figure 5.2) and to compare this method with optimization using MATLAB. The two methods turned out to yield similar results. The pressures were considered separately first to make sure we were in the approximate optimal region before considering the flow rates as well.

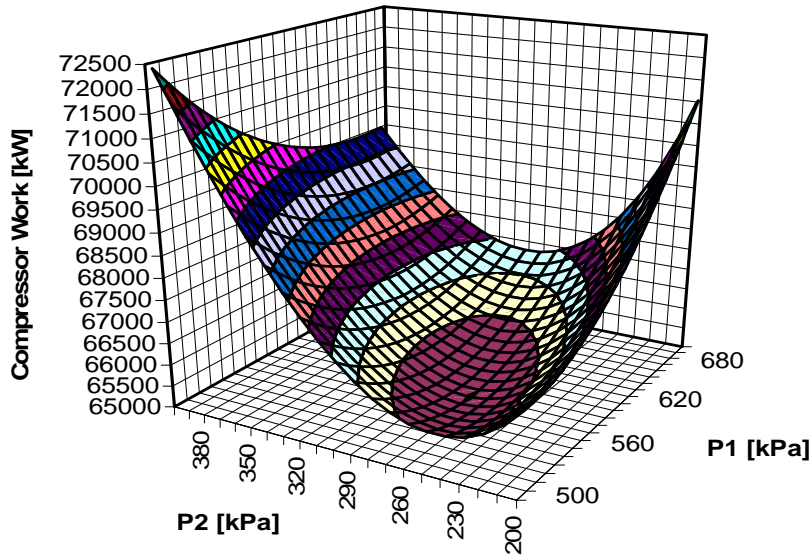


Figure 5.2: Work as a function of pressures.
 The case shown is just for illustration (100mole% propane, $T_{\text{cond}} = 30^{\circ}\text{C}$)

After the pressures were optimized the flow rates was also taken into consideration, using MATLAB's `fmincon` function. The rest of the cases were optimized using all the five degrees of freedom at the same time.

One of the constraints is that we should have superheated vapour out of each heat exchanger to make sure no vapour is introduced into the compressor. Enthalpies instead of temperatures were used for this purpose to make sure the optimizer was not running into problems. For a single-component fluid the superheating, defined as $T - T_s$, becomes zero when entering the two phase region. More importantly, also the gradient becomes zero in this region since the fluid is boiling under constant temperature. This will make it difficult for the optimization algorithm to identify which step direction to take to retain feasibility. By replacing the temperature constraint with an enthalpy constraint we avoid this problem since the gradient is no longer zero inside the two phase region. For a two component fluid a temperature constraint could be used since the fluid temperature is increasing as the fluid is boiling. However, enthalpies are used for both cases to be consistent.

$\Delta T_{\text{min}} = 5^{\circ}\text{C}$ was also implemented as a constraint and the natural gas and MR temperature out of each heat exchanger should be equal for the design case. The results are presented in table 5.3 and table 5.4.

The UA-values of each heat exchanger is presented in table 5.1 and table 5.2.

Table 5.1: Ethane/propane UA

HE	UA	
	MR [10^6 kJ/C*h]	NG [10^6 kJ/C*h]
LNG-100	13.9	2.8
LNG-200	18.7	2.8
LNG-300	18.7	3.0

Table 5.2: Propane UA

HE	UA	
	MR [10 ⁶ kJ/C*h]	NG [10 ⁶ kJ/C*h]
LNG-100	8.5	1.3
LNG-200	17.6	2.6
LNG-300	18.9	3.1

5.2 Operation

By adding additional equality constraints for the UA-values in MATLAB we are moving from a design case to an operation case. The tolerance on the UA-values must be relatively wide to allow the UniSim model to converge, a relative tolerance of approximately 10^{-3} was used. This tolerance is higher than the calculation tolerance in UniSim, which is set by default to 10^{-4} . Thus, the UA-constraint could have been kept in UniSim. When the heat exchanger is designed with a specified UA-value the ΔT_{\min} constraint is no longer needed. Thus, the ΔT_{\min} constraint is modified to be larger than 0°C instead of equal to 5°C . The temperatures out of each heat exchanger were also used as free variables in the optimization to make MATLAB able to vary these. It is necessary to vary these temperatures in order for UniSim to get the same UA-values as specified in MATLAB.

As discussed in Section 2.2.4 the variations in the air temperature in the Shtokman area is typically $\pm 30^{\circ}\text{C}$. This means that the minimum condensation temperatures are in the range -10°C to 50°C when using a temperature difference of 20°C . To be able to do a systematic approach on the problem, some requirements must be set upon the system. These requirements should be met by all the different cases making it easy to do comparison. The requirements and assumptions were mentioned in the model description.

The precooled natural gas and mixed refrigerant from the liquefaction part should be precooled to at least -35°C when using pure propane and to -50°C when using an ethane/propane mixture.

Since the air varies in between $\pm 30^{\circ}\text{C}$ the propane/ethane mixture must necessarily vary throughout the year to efficiently utilize the precooling part of the LNG process. The problem is: How can we find this optimal composition? In other words; how can we find the optimal amount of ethane in the refrigerant for a given ambient temperature? Using the ethane content in the precooling refrigerant as a free variable in the optimization gave no feasible results. The optimizer always ended in no feasible solution without fulfilling the constraints.

To investigate the different composition possibilities a bubble point-composition plot for different pressures was made, see figure 5.3.

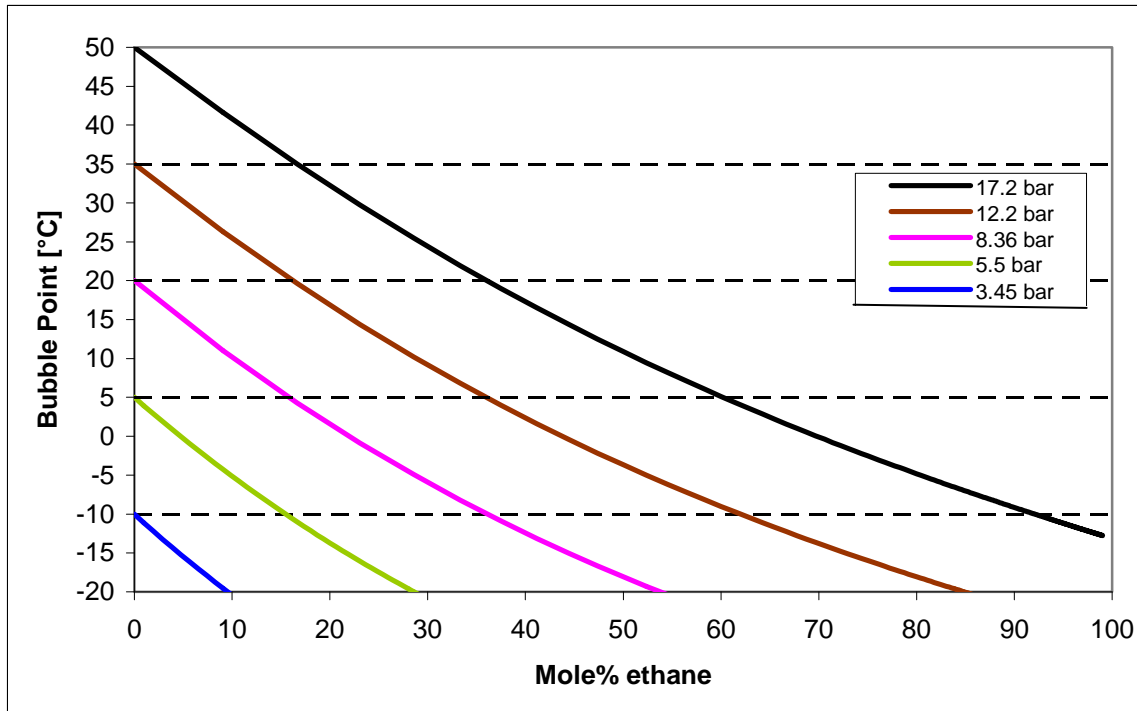


Figure 5.5: Bubble points vs mole% ethane at different pressures

The plot is only for P_{HH} up to 17.2 bar, which is the saturation pressure for propane at 50°C. If the highest pressure is limited to 17.2 bar there is only one choice for the highest ambient temperature (30°C), and that is to use pure propane as refrigerant. When the ambient temperature is decreasing, more ethane should be used in the precooling medium. If a maximum pressure ($P_{HH,max}$) of 17.2 bar is used, the *maximum* ethane content at various ambient temperatures can be found by finding the composition that corresponds to the bubble point at $P_{HH,max}$. The bubble point should be 20°C above T_{amb} .

However, it is not realistic to say that the maximum pressure is 17.2 bar (which is a relatively low pressure), but rather use the condensation pressure as a free variable. At the same time the range of composition (i.e. ethane content) should be limited to make sure the molecular weight is not varying too much. A higher molecular weight gas would be compressed more than a lighter molecular weight gas. Surge will be reached earlier and consequently the range is more limited for a heavier gas [18].

As mentioned in Chapter 3, five cases with various amount of ethane were investigated and five cases with pure propane were investigated for comparison. In all cases the models were manually optimized first and then optimized using the `fmincon`-function in MATLAB. The ethane content was varying from 40mole% to 80mole%.

From the equilibrium chart in figure 5.4 we see that the minimum amount of ethane content is about 12 mole% when we require that the split temperature should be -55°C.

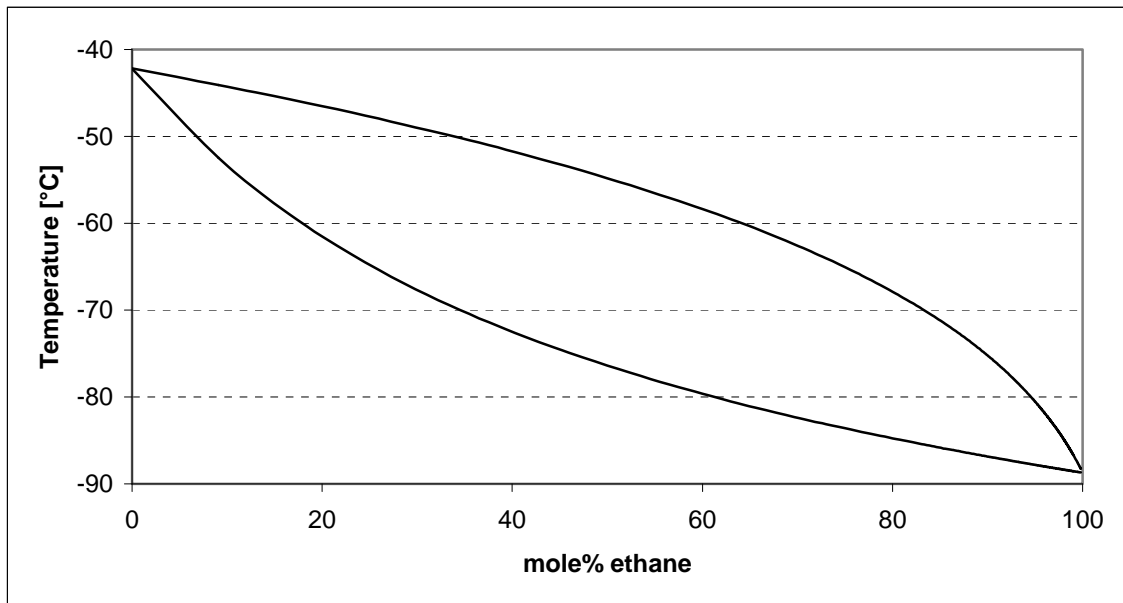


Figure 5.4: Equilibrium Chart for Propane and Ethane at $P = 1$ atm

The optimization turned out to require a lot of effort to make it work sufficient. Adjusting the tolerances on the adjusted variables (TolVar), the objective function (TolFun) and constraints (TolCon) were necessary in order to find a feasible solution. Because the objective function is subject to noise, also the minimum size of the steps (DiffMinChange) had to be altered. This was necessary to make sure the flow sheet was able to converge. The steps are the perturbations performed to by the optimizer algorithm to calculate the two gradients $\partial J / \partial x$ and $\partial c / \partial x$, where J is the objective function, c the constraint function and x is the states. Finally, the interior-point algorithm was used to require the state variables to be kept within the specified boundaries.

The final results are presented in table 5.3 for the ethane/propane case and in table 5.4 for the propane case.

Table 5.3: Optimization results when using PMR (C2/C3)

	Case I	Case II	Case III (DESIGN CASE)	Case IV	Case V
T_{amb} [°C]	-30	-15	0	15	30
T_{PMR} [°C]	-10	5	20	35	50
C2 [mole%]	80	70	60	50	40
C3 [mole%]	20	30	40	50	60
Mole weight [g/mole]	32.88	34.28	35.68	37.08	38.49
T_{low} [°C]	-55	-55	-55	-55	-55
RESULTS					
P_{HH} [bar]	15.10	19.38	23.83	28.30	32.70
P_H [bar]	10.03	11.94	15.65	15.24	16.08
P_M [bar]	8.03	7.82	8.00	10.16	6.08
P_L [bar]	3.50	2.83	2.12	1.48	1.03
F_1 [kmole/s]	12.55	18.0	26.5	28.9	36.0
F_2 [kmole/s]	8.43	10.6	15.7	16.9	19.9
F_3 [kmole/s]	5.39	6.9	8.0	11.9	9.9
$T_{out(NG)}$ [°C]	-50.08	-46.08	-50	-49.16	-48.89
$T_{out(MR)}$ [°C]	-53.27	-50.41	-50	-49.98	-50.63
$\Delta T_{min(HE-100)}$ [°C]	2.23	2.80	5.0	4.48	3.30
$\Delta T_{min(HE-200)}$ [°C]	1.10	0.68	5.0	2.66	1.84
$\Delta T_{min(HE-300)}$ [°C]	0.50	4.59	5.0	5.02	4.37
Compressor work [MW]	27.83	49.31	82.96	139.061	179.77
COP	4.905	3.643	3.212	1.928	1.640

Table 5.4: Optimization results when using pure propane (C3)

	Case VI	Case VII	Case VIII (DESIGN CASE)	Case IX	Case X
T_{amb} [°C]	-30	-15	0	15	30
T_{PMR} [°C]	-10	5	20	35	50
C2 [mole%]	0	0	0	0	0
C3 [mole%]	100	100	100	100	100
T_{low} [°C]	-40	-40	-40	-40	-40
RESULTS					
P_{HH} [bar]	3.45	5.50	8.36	12.20	17.20
P_H [bar]	2.80	4.00	5.99	8.10	5.63
P_M [bar]	2.00	2.00	3.00	3.07	2.64
P_L [bar]	1.11	1.11	1.11	1.11	1.11
F_1	5.70	9.00	13.47	19.98	22.5
F_2	4.80	7.00	10.93	14.97	11.59
F_3	2.50	3.00	5.43	5.47	4.70
$T_{out(NG)}$ [°C]	-36.62	-35.85	-35	-34.53	-35.22
$T_{out(MR)}$ [°C]	-36.77	-35.94	-35	-34.45	-35.10
$\Delta T_{min(HE-100)}$ [°C]	2.57	4.38	5	5.81	15.40
$\Delta T_{min(HE-200)}$ [°C]	2.24	0.18	5	6.78	7.33
$\Delta T_{min(HE-300)}$ [°C]	0.68	0.29	5	5.47	4.78
Compressor work [MW]	10.78	26.05	45.94	73.79	109.56
COP	8.310	5.253	3.936	2.980	2.257

From the two tables it can be seen that the compressor work is decreasing significantly with decreasing ambient temperature, as expected. The compressor work is plotted against ambient temperature for the ten different cases in figure 5.5.

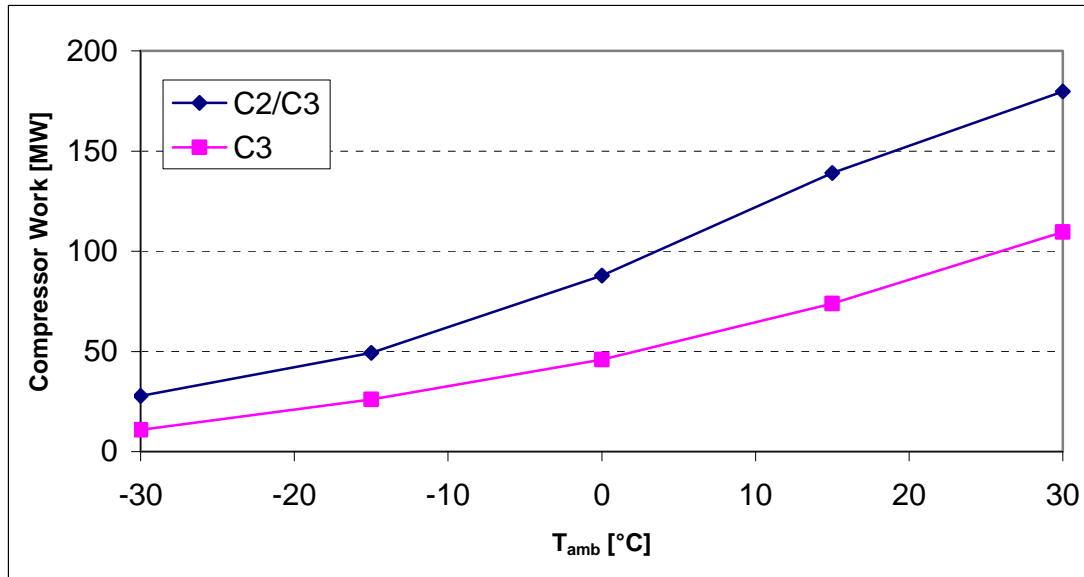


Figure 5.5: Compressor Work vs Ambient Temperature

It can be seen from the figure that the compressor work when using pure propane is lower than when using a mixture for all the different ambient temperatures. This is due to the fact that propane only precools down to about -35°C while the ethane/propane mixture precools down to about -50°C . This leaves less work for the liquefaction and subcooling part of the process, which is further discussed in the next section.

In figure 5.6 and figure 5.7 the pressure ratios for each compressor is plotted against the flow rate through the compressor. The corresponding data is presented in table 5.5 and table 5.6.

Table 5.5: Pressure ratios and flows for the compressors (C2/C3)

	Case I	Case II	Case III	Case IV	Case V
K-300	2,29	2,76	3,77	6,86	5,90
F₃ [kmole/s]	5,39	6,90	8,00	11,90	9,90
K-200	1,25	1,53	1,94	1,50	2,64
F₂ [kmole/s]	8,43	10,60	15,70	16,90	19,90
K-100	1,51	1,62	1,54	1,86	2,03
F₁ [kmole/s]	12,55	18,00	26,50	28,90	36,00

Table 5.6: Pressure ratios and flows for the compressors (C3)

	Case VI	Case VII	Case VIII	Case IX	Case X
K-300	1,80	1,80	2,70	2,77	2,38
F₃ [kmole/s]	2,50	3,00	5,43	5,47	4,70
K-200	1,40	2,00	2,00	2,64	2,13
F₂ [kmole/s]	4,80	7,00	10,93	14,97	11,59
K-100	1,23	1,38	1,40	1,51	3,06
F₁ [kmole/s]	5,70	9,00	13,47	19,98	22,50

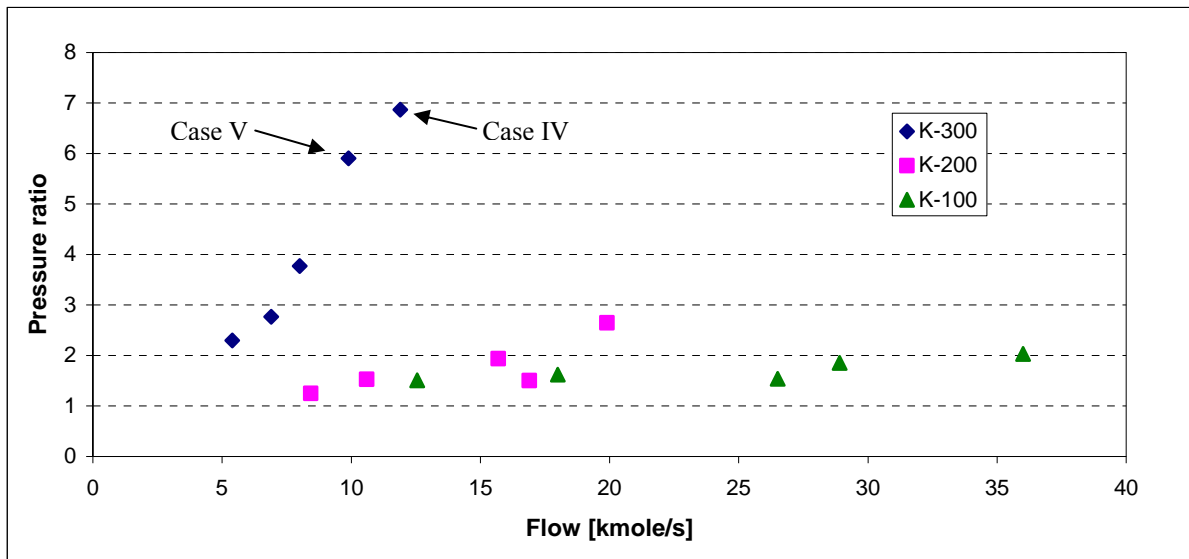


Figure 5.6: Pressure ratios when using C₂/C₃

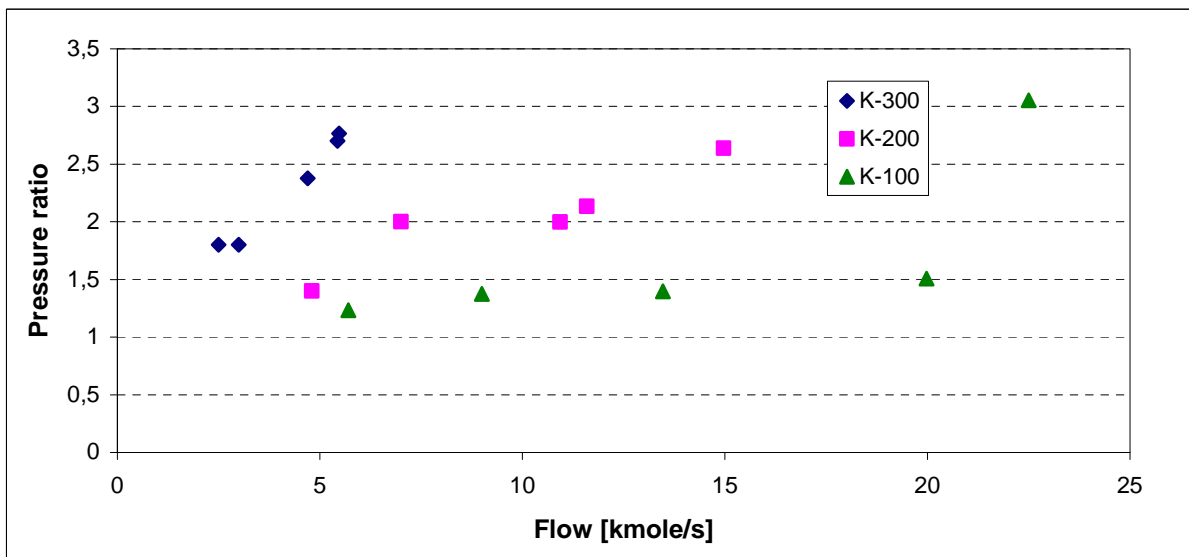


Figure 5.7: Pressure ratio when using only C₃

The different points (from left to right) indicate which case the pressure ratios and flows belong to, where the leftmost point correspond to Case I ($T_{amb} = -30^{\circ}\text{C}$), and so on. It is clear from the two figures that the pressure ratios for the three compressors are much lower for the propane cases compared to the PMR cases. This is of course because the condensing pressure (P_{HH}) is higher when introducing a lighter component to propane.

There are some points that should be discussed further. When using PMR, in Case V there is actually *less* vapour going through K-300 than in Case IV and the pressure ratio is also lower. Such a high pressure ratio for K-300 does not provide a good work distribution among the compressors. It would therefore be better to have a lower flow through this compressor with resulting lower pressure ratio for this compressor and thus higher flow and higher pressure ratio for the two other compressors. From the figure it can be seen that the pressure ratio for K-200 is deviating from the trend for Case IV which indicates that this pressure ratio should be higher.

Besides of Case IV and Case V for K-300 all the other pressure ratios are below 4 when using PMR, while the pressure ratios are all below 3 when using pure propane, except for K-100, Case V. This pressure ratio is significantly higher, about the double, of the other pressure ratios for this compressor. This also indicated a potential for better optimization.

From the table 5.5 and table 5.6 it is seen that introducing ethane in the precooling medium significantly raises the condensation pressure. The lowest condensation pressure for the ethane/propane mixture is 15.10 bar (Case I) and the highest condensation pressure for pure propane is 17.20 bar (Case X). The differences in compressor work are given in table 5.7.

Table 5.7: Compressor work, differences

	$T_{\text{amb}} = -30^{\circ}\text{C}$	$T_{\text{amb}} = -15^{\circ}\text{C}$	$T_{\text{amb}} = 0^{\circ}\text{C}$	$T_{\text{amb}} = +15^{\circ}\text{C}$	$T_{\text{amb}} = +30^{\circ}\text{C}$
ΔW_s [MW]	17.1	23.3	41.9	65.3	70.2

From the table it can be seen that the difference is increasing with increasing temperature. This means that the compression work in the liquefaction section must be *at least* 17.1 MW smaller when using an ethane/propane mixture compared to using pure propane for this option to be considered.

Temperature-enthalpy plots for Case III and Case VIII (both design cases) are presented in appendix B and C.

5.3 Liquefaction Compressor Work

To be able to do a complete evaluation of the performance when the ambient temperature is decreasing, also the work required in the liquefaction/subcooling cycle must be taken into consideration. An existing model made by co-supervisor Magnus G. Jacobsen has been scaled up and used for this purpose. The model is shown in figure 5.8.

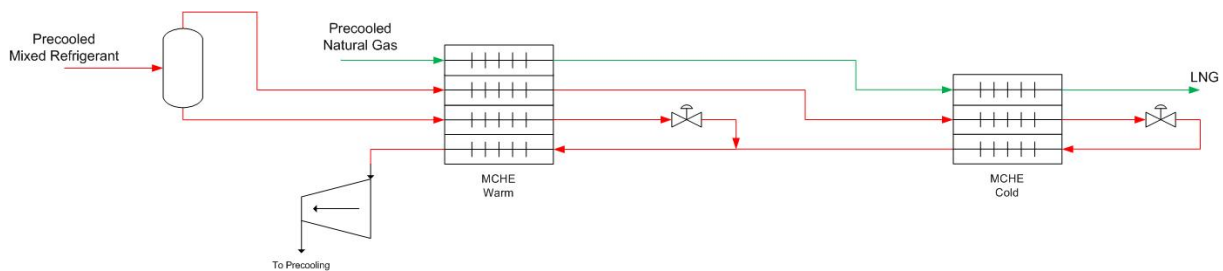


Figure 5.8: Liquefaction and subcooling of pre-cooled natural gas

When decreasing the split temperature the MR inlet pressure and the MR flow rate was also decreased. The pressure was decreased to yield the same initial vapour/liquid fraction in the inlet stream to get the proper flow distribution in the heat exchangers. The flow rate was decreased until the model failed to converge. The model was unsuccessful in converging when the split temperature got lower than -41°C . The result was that the cold part main cryogenic heat exchanger could not be solved. Decreasing the split temperature even further made it insensitive. The reason might be that small ΔT_{min} causes problems for the solver. Increasing the intervals in the heat exchanger helped to a certain degree, but not all the way down to $T_{\text{split}} = -50^{\circ}\text{C}$. However, the compression work and the split temperature turned out to

have an approximately linear relationship, at least for this small temperature window, see figure 5.9.

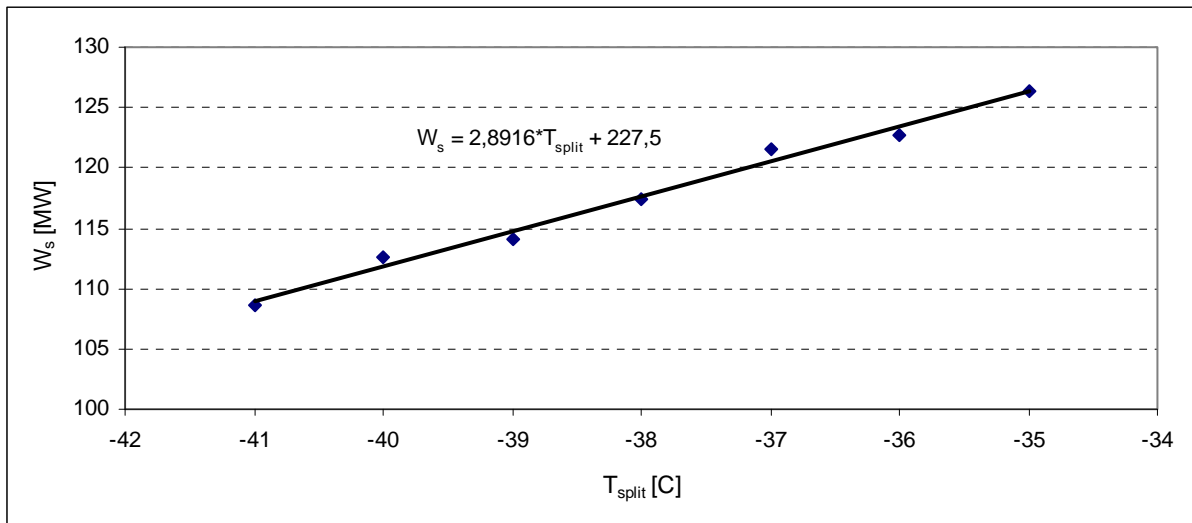


Figure 5.9: W_s vs T_{split}

Using this linear relationship, the compressor work with a temperature split of -50°C is found to be 82.92 MW. Compared to the compressor work with a temperature split of -35°C which is 126.35 MW the difference is approximately 43 MW. Using the compressor work from the precooling section and adding the corresponding compressor work from the liquefaction part, the total work (W_T) is found and a new graph is generated and presented in figure 5.10. The data are presented in table 5.8.

Table 5.8: Total Compressor Work

T_{amb} [°C]	$W_T, C_2/ C_3$ [MW]	W_T, C_3 [MW]
-30	110.75	137.13
-15	132.23	152.40
0	170.74	172.29
15	221.98	200.14
30	262.69	235.91
Total	898.39	897.87

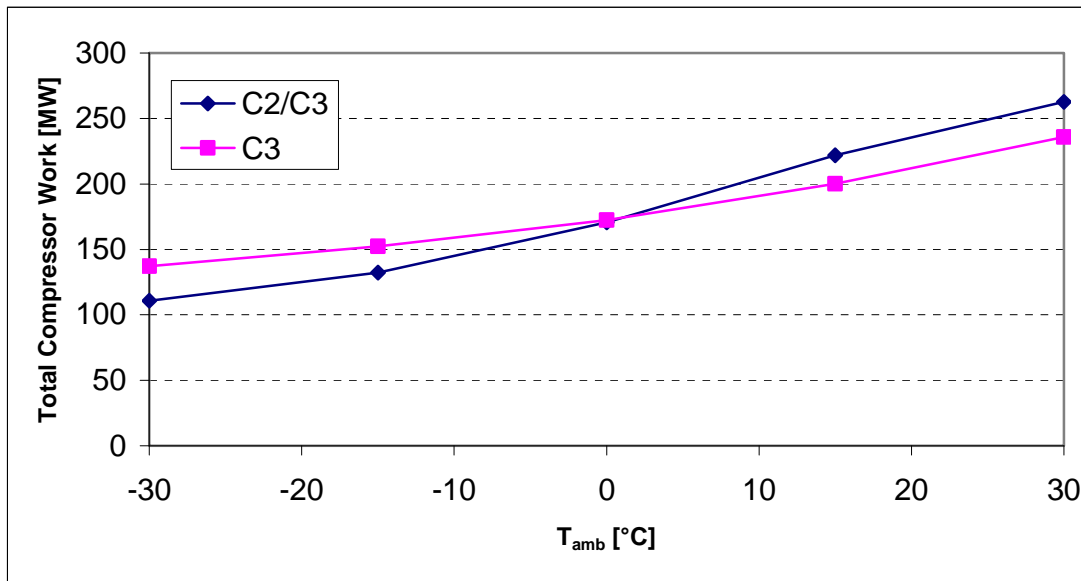


Figure 5.10: Total Compressor Work

From the figure above it is easy to see that there is a cross-over in compressor work about ambient temperature of 0°C. This tells us that for ambient temperatures below 0°C it would not be beneficial to use pure propane since the total compressor work is higher compared to the mixed refrigerant case. However, it is seen that for the mixed refrigerant case the total compressor work is higher for ambient temperatures above 0°C, which indicates that there might be a too high fraction of ethane in the composition. Adding the total compressor work for both cases they are actually almost equal and close to 898 MW. If the ethane content had been reduced at the higher temperature levels (15°C and 30°C) the total work would necessarily be decreased. This would just be a matter of how large changes in concentrations the compressor could deal with.

6 PLANTWIDE CONTROL

Plantwide control is control structure design for complete chemical plants. The control system can be divided into five different layers; scheduling, site-wide optimization, local optimization, supervisory control and regulatory control, which all have different time scales (typically weeks, days, hours, minutes and seconds, respectively), see figure 6.1. The link in between the layers is the controlled variables (c), and these variables can be difficult to select without a systematic procedure. In [17] Skogestad summarizes the procedure in seven steps, where the four first steps are related to the “Top-down analysis” and the three last steps are related to the “Bottom-up design”. For details about the procedure it is referred to [17], only some key points will be enlightened here.

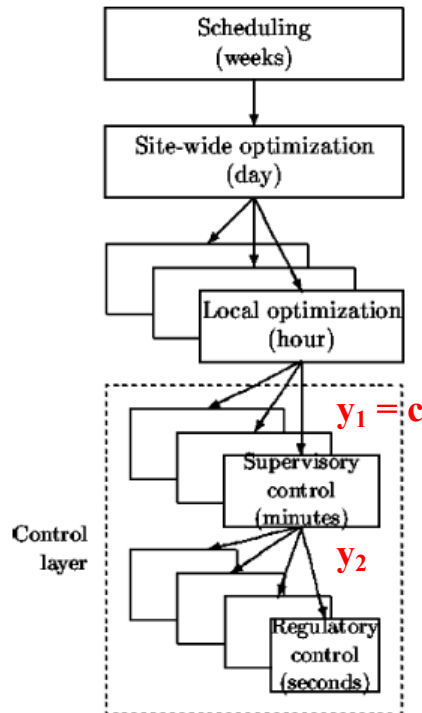


Figure 6.1: Typical control hierarchy in a chemical plant [17]

After the operational constraints, objective function and DOF have been identified (Step 1 and 2) we must decide which variables we should control (Step 3). The active constraints should always be controlled, but the question is which other variables to control with the remaining (unconstrained) DOF. This is the concept of “Self-optimizing control”, which was formulated in 1980 by Morari et al.: “*We want to find a function c of the process variables which when held constant, leads automatically to the optimal adjustment of the manipulated variables.*” Further, in [16], Skogestad more precisely says “*Self-optimizing control is achieved if a constant setpoint policy results in an acceptable loss L (without the need to reoptimize when disturbances occur).*” To identify good controlled variables the “minimum singular value rule” could be used: “*Look for variables that maximize the minimum singular value of the appropriately scaled steady-state gain matrix G from u to c* ”, where u is the steady-state degrees of freedom [17]. This method is good for small disturbances and for eliminating poor candidate variables, but for the final selection one should evaluate the loss for larger disturbances. This is also summarized in seven steps. After the important disturbances are identified the idea is to do optimization; nominally and with the disturbances applied. The controlled variable resulting in the smallest loss would be the preferred selection.

After the controlled variables are selected, the TPM (throughput manipulator) is identified and the plant is stabilized using PID loops (Step 4 and 5). Then, in step 6, we keep the controlled outputs of the supervisory level at constant optimal setpoints using the setpoints for the layer below (regulatory layer). In Step 7 the optimal setpoints (c_s) for the controlled variables are computed, either using RTO (Real Time Optimization) or “manual” optimization (operators).

To investigate if there are any potential self-optimizing controlled variables in the established precooling model, the main disturbances would first have to be identified. Some important disturbances are:

- Natural gas flow rate
- Mixed refrigerant flow rate
- Ambient temperature
- UA-value
- NG inlet pressure
- NG composition

Performing a step in the each of the disturbances (e.g. 5% increase) and then reoptimize the model would enable us to calculate the scaled gain for candidate controlled variables. Selecting controlled variables with maximum scaled gain are preferred.

From figure 5.5 and figure 5.6 it can be seen that the K-100 curve is very flat (except for the last point in figure 5.6), which means that the pressure ratio is close to constant. This indicates that P_{HH}/P_H could be a potential self-optimizing variable. However, self-optimizing control has not been considered in this project since it is not the focus of the study.

7 DISCUSSION

Using these approximated calculations it is seen that there are energy savings related to using ethane/propane mixture when the ambient air temperature drops below 0°C, but not for temperatures above 0°C. The ethane/propane mixtures were selected using existing knowledge from the Sakhalin-II field, and there could have been put more effort in adjusting the ethane/propane ratio. A lower ethane content would reduce P_{HH} and further reduce the compressor work. As long as the requirement $T_{split} < -50^\circ\text{C}$ is fulfilled the ethane content could be decreased.

When decreasing the split temperature from -35°C to -50°C the MR pressure could also be decreased accordingly. This would further lead to a reduced compressor work for the liquefaction part.

The reason why the pressure ratios are barely changing for K-300 is because the bounds are set relatively tight for both the pressures and the flow in order to allow the optimization algorithm to converge. It has been mentioned that finding the absolute right optimal operating point for all the different cases is not the purpose of the study. If this is the purpose, some other software should be used, for instance gProms (general PROcess Modeling System). In UniSim, the step calculations are affected by noise in the system; too small steps will give wrong gradients giving meaningless results. Having a too large step may result in an actual minimum in the objective function is being missed by the optimizer algorithm. See figure 7.1 for an illustration

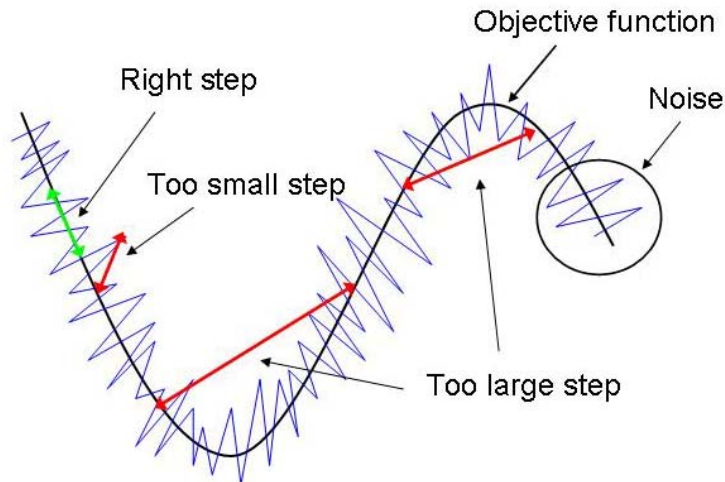


Figure 7.1: Consequence of choosing wrong step size during optimization

There was only assumed a pressure drop for the NG stream, but at a later point it is realized that there would be realistic to assume a small pressure drop for the MR stream as well. It has been investigated and it turns out it will not affect the overall trends of the results if a pressure drop for the MR stream was included as well; it would result in a small increase in the compressor work for all the cases.

To simplify the model there has not been put constraints on the compressor suction volumes or the pressure ratios for the compressors. For Case V with a mixture of 40mole% ethane and 60mole% propane the pressure ratio is above two for K-100 and the outlet pressure is about 32 bars. This is a very high pressure and would require specialized components.

Another option, that has not been taken into consideration in this study, is to include P_L as a degree of freedom and not let it be set by the lowest temperature (-55°C). Also, the ethane content should optimally be included as a degree of freedom. For MATLAB to be able to access this variable one easy option is to access it indirectly by using the flow rate of a pure ethane stream into a mixer. This setup is shown in figure 5.1. Since the optimizer failed in finding a solution to such a problem five different cases was considered instead. There might be some benefits in using other softwares, such as gPROMS for this purpose.

It is realized that the pressure in the MR stream could be set lower than 45 bar. This pressure was taken from a case where the condensation temperature of the precooling medium was 30°C and the split temperature was about -35°C . When decreasing the split temperature the pressure could be set lower to get the same vapour/liquid ratio for the MR flow out of the precooling section. However, if it is assumed that the heat capacity is constant for the MR stream entering the precooling section, this would not affect the performance.

In some cases (Case I, II, VI and VII) the optimizer was not able to fulfil the superheating constraint. When lowering the flow rate of the precooling medium either a temperature cross in the heat exchanger was observed or the flow sheet was not able to converge. Increasing or decreasing the pressure out of the choke valve did not seem to help either. The solution to this problem was to move the constraint from the outlet of the heat exchanger to the inlet of the compressor, which is really the place where superheating is necessary. If the preceding compressed stream is sufficient superheated to also vaporize the entering two-phase stream,

the inlet stream to the compressor would be superheated. This would not be the case for the last heat exchanger (HE-300) since this stream is not mixed with any superheated streams prior to entering the low pressure compressor (K-300). Thus, this stream must always be superheated. The heat exchanger outlet streams were *almost* superheated in all the cases, so the trends and comparison of the cases are still valid.

8 CONCLUSION

Based on the results in this study it can be concluded that introducing a mixed refrigerant also in the precooling part of the LNG process will lower the total compressor work at low ambient temperatures, when air is used as coolant. It was assumed before the study was started that it would be beneficial to use a PMR, and this study has confirmed the hypothesis.

However, there has been experienced quite a few problems during optimization of the system. Building the model in UniSim introduces noise and loose tolerances are needed. Thus, the accuracy of the optimization results is lowered. However, since the main focus of this study was not to find the exact optimal operating point for all the different cases, the results are useful.

A self-optimizing control structure has not been proposed for the system, but central issues have been discussed.

9 FURTHER WORK

A complete model should be built, including also the liquefaction part of the process. This would enable us to get a complete picture of the affect ambient temperature has on the process. Previous work by [19] has shown that finding an optimal operating point for the whole process is difficult using UniSim and MATLAB due to many of the issues discussed in this report. Therefore it is advised that alternative software is used, for instance gPROMS. As a start only the precooling model could be build in gPROMS, and at a later point the liquefaction part could be integrated.

Also, there should be investigated if it is possible to find a self-optimizing structure of the process. The procedure described in chapter 6 could be used for this purpose. A more accurate model than what the UniSim software can provide is probably necessary then.

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A UniSim Design Workbook – CASE III

Name	NG3	MR3	MR_1out	NG1	NG2	MR1
Vapour Fraction	1,0000	0,0000	0,6840	1,0000	1,0000	0,7912
Temperature [C]	-50,00	-50,00	14,93	8,446	-14,52	8,446
Pressure [kPa]	5850	4500	1565	5950	5900	4500
Molar Flow [kgmole/s]	9,467	18,93	3,999	9,467	9,467	18,93
Mass Flow [kg/h]	5,776e+005	1,768e+006	5,224e+005	5,776e+005	5,776e+005	1,768e+006
Std Ideal Liq Vol Flow [m3/h]	1838	4846	1233	1838	1838	4846
Heat Flow [kJ/h]	-2,706e+009	-6,653e+009	-1,431e+009	-2,605e+009	-2,641e+009	-6,061e+009
Molar Enthalpy [kJ/kgmole]	-7,941e+004	-9,761e+004	-9,942e+004	-7,645e+004	-7,750e+004	-8,892e+004
Name	MR2	MR_3out	MR_1in	MR1_V	NG feed	MR feed
Vapour Fraction	0,5038	1,0000	0,0000	1,0000	1,0000	0,9032
Temperature [C]	-14,52	-21,26	3,446	3,446	20,00	20,00
Pressure [kPa]	4500	212,3	1565	1565	6000	4500
Molar Flow [kgmole/s]	18,93	6,211	3,999	4,201	9,467	18,93
Mass Flow [kg/h]	1,768e+006	8,601e+005	5,224e+005	4,964e+005	5,776e+005	1,768e+006
Std Ideal Liq Vol Flow [m3/h]	4846	1924	1233	1286	1838	4846
Heat Flow [kJ/h]	-6,300e+009	-2,221e+009	-1,561e+009	-1,378e+009	-2,588e+009	-5,949e+009
Molar Enthalpy [kJ/kgmole]	-9,242e+004	-9,933e+004	-1,084e+005	-9,111e+004	-7,594e+004	-8,727e+004
Name	9	MR_high	MR	C1	C3	C2
Vapour Fraction	1,0000	1,0000	0,0000	1,0000	1,0000	1,0000
Temperature [C]	29,75	55,02	20,00	30,00	30,00	30,00
Pressure [kPa]	1565	2383	2383	30,00	30,00	30,00
Molar Flow [kgmole/s]	24,00	24,00	24,00	0,0000	2,778e-004	0,0000
Mass Flow [kg/h]	3,083e+006	3,083e+006	3,083e+006	0,0000	44,10	0,0000
Std Ideal Liq Vol Flow [m3/h]	7390	7390	7390	0,0000	8,703e-002	0,0000
Heat Flow [kJ/h]	-8,077e+009	-7,984e+009	-9,105e+009	0,0000	-1,035e+005	0,0000
Molar Enthalpy [kJ/kgmole]	-9,348e+004	-9,240e+004	-1,054e+005	-7,472e+004	-1,035e+005	-8,449e+004
Name	N2	6-2	MR_2out	12	MR_med2	MR_3
Vapour Fraction	1,0000	1,0000	1,0000	1,0000	1,0000	0,2236
Temperature [C]	30,00	30,00	2,429	-27,82	36,61	-55,00
Pressure [kPa]	30,00	30,00	800,0	212,3	800,0	212,3
Molar Flow [kgmole/s]	0,0000	2,778e-004	4,856	8,000	8,000	8,000
Mass Flow [kg/h]	0,0000	44,10	6,488e+005	1,069e+006	1,069e+006	1,069e+006
Std Ideal Liq Vol Flow [m3/h]	0,0000	8,703e-002	1500	2471	2471	2471
Heat Flow [kJ/h]	0,0000	-1,035e+005	-1,688e+009	-2,816e+009	-2,715e+009	-3,234e+009
Molar Enthalpy [kJ/kgmole]	143,5	-1,035e+005	-9,658e+004	-9,776e+004	-9,428e+004	-1,123e+005
Name	MR_2*	MR_2	MR_1L	MR_2L	MR_2V	MR_med1
Vapour Fraction	0,0000	0,1864	0,0000	0,0000	1,0000	1,0000
Temperature [C]	3,446	-19,52	3,446	-19,52	-19,52	53,40
Pressure [kPa]	1565	800,0	1565	800,0	800,0	1565
Molar Flow [kgmole/s]	15,80	15,80	19,80	12,86	2,944	15,80
Mass Flow [kg/h]	2,064e+006	2,064e+006	2,586e+006	1,718e+006	3,462e+005	2,064e+006
Std Ideal Liq Vol Flow [m3/h]	4871	4871	6104	3970	901,0	4871
Heat Flow [kJ/h]	-6,166e+009	-6,166e+009	-7,727e+009	-5,197e+009	-9,692e+008	-5,268e+009
Molar Enthalpy [kJ/kgmole]	-1,084e+005	-1,084e+005	-1,084e+005	-1,123e+005	-9,143e+004	-9,261e+004
Name	MR_3in	MR_2in	MR_3*	MR_3V	MR_1	31
Vapour Fraction	0,0000	0,0000	0,0000	1,0000	0,1750	0,0000
Temperature [C]	-55,00	-19,52	-19,52	-55,00	3,446	20,00
Pressure [kPa]	212,3	800,0	800,0	212,3	1565	2383
Molar Flow [kgmole/s]	6,211	4,856	8,000	1,789	24,00	24,00
Mass Flow [kg/h]	8,601e+005	6,488e+005	1,069e+006	2,088e+005	3,083e+006	3,083e+006
Std Ideal Liq Vol Flow [m3/h]	1924	1500	2471	547,2	7390	7390
Heat Flow [kJ/h]	-2,640e+009	-1,963e+009	-3,234e+009	-5,945e+008	-9,105e+009	-9,105e+009
Molar Enthalpy [kJ/kgmole]	-1,181e+005	-1,123e+005	-1,123e+005	-9,231e+004	-1,054e+005	-1,054e+005
Name	33	26	32	34	Feed Study	Ref Stream 3
Vapour Fraction	1,0000	1,0000	1,0000	1,0000	0,0000	1,0000
Temperature [C]	-38,21	16,76	-3,529	17,52	-168,1	-34,85
Pressure [kPa]	212,3	800,0	800,0	1565	101,3	212,3
Molar Flow [kgmole/s]	8,000	15,80	15,80	24,00	42,98	6,211
Mass Flow [kg/h]	1,069e+006	2,064e+006	2,064e+006	3,083e+006	2,614e+006	8,601e+005
Std Ideal Liq Vol Flow [m3/h]	2471	4871	4871	7390	8334	1924
Heat Flow [kJ/h]	-2,833e+009	-5,373e+009	-5,447e+009	-8,153e+009	-1,388e+010	-2,239e+009
Molar Enthalpy [kJ/kgmole]	-9,835e+004	-9,446e+004	-9,577e+004	-9,437e+004	-8,970e+004	-1,001e+005

Name	Ref Stream 2	Ref Stream 1	36	Propane	36-2	Ethane
Vapour Fraction	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000
Temperature [C]	-0,6433	20,06	89,57	100,0	93,03	100,0
Pressure [kPa]	800,0	1565	750,0	1750	750,0	1750
Molar Flow [kgmole/s]	4,856	3,999	39,64	39,64	39,64	39,64
Mass Flow [kg/h]	6,488e+005	5,224e+005	6,293e+006	6,293e+006	4,291e+006	4,291e+006
Std Ideal Liq Vol Flow [m3/h]	1500	1233	1,242e+004	1,242e+004	1,206e+004	1,206e+004
Heat Flow [kJ/h]	-1,692e+009	-1,369e+009	-1,417e+010	-1,417e+010	-1,159e+010	-1,159e+010
Molar Enthalpy [kJ/kgmole]	-9,679e+004	-9,507e+004	-9,932e+004	-9,932e+004	-8,120e+004	-8,120e+004
Name	16	29	37	Ref Stream 3*	Ref Stream 2*	Ref Stream 1*
Vapour Fraction	0,0000	0,0000	0,0000	1,0000	1,0000	1,0000
Temperature [C]	0,0000	5,684e-014	45,00	-31,85	2,357	23,06
Pressure [kPa]	500,0	500,0	2604	212,3	800,0	1565
Molar Flow [kgmole/s]	39,64	39,64	6,964e-006	6,211	4,856	3,999
Mass Flow [kg/h]	6,293e+006	6,293e+006	1,000	8,601e+005	6,488e+005	5,224e+005
Std Ideal Liq Vol Flow [m3/h]	1,242e+004	1,242e+004	2,163e-003	1924	1500	1233
Heat Flow [kJ/h]	-1,755e+010	-1,755e+010	-2753	-2,235e+009	-1,688e+009	-1,365e+009
Molar Enthalpy [kJ/kgmole]	-1,230e+005	-1,230e+005	-1,098e+005	-9,996e+004	-9,659e+004	-9,485e+004
Name	7	nrNG	LNG	2	3	1
Vapour Fraction	0,0000	1,0000	0,0000	0,0000	0,0000	1,0000
Temperature [C]	20,00	-30,00	-160,0	-161,0	-74,42	50,00
Pressure [kPa]	2383	5850	5850	4500	4500	4500
Molar Flow [kgmole/s]	24,00	9,467	9,467	18,93	18,93	18,93
Mass Flow [kg/h]	3,083e+006	5,776e+005	5,776e+005	1,693e+006	1,693e+006	1,693e+006
Std Ideal Liq Vol Flow [m3/h]	7390	1838	1838	4759	4759	4759
Heat Flow [kJ/h]	-9,105e+009	-2,667e+009	-3,041e+009	-6,855e+009	-6,480e+009	-5,499e+009
Molar Enthalpy [kJ/kgmole]	-1,054e+005	-7,825e+004	-8,924e+004	-1,006e+005	-9,508e+004	-8,067e+004
Name	Q-101	Q-100	Q-300	Q-200	Q-102	*** New ***
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/s]	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	1,121e+009	9,313e+007	1,004e+008	1,051e+008	9,819e+008	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	

B T-H DIAGRAMS, CASE III

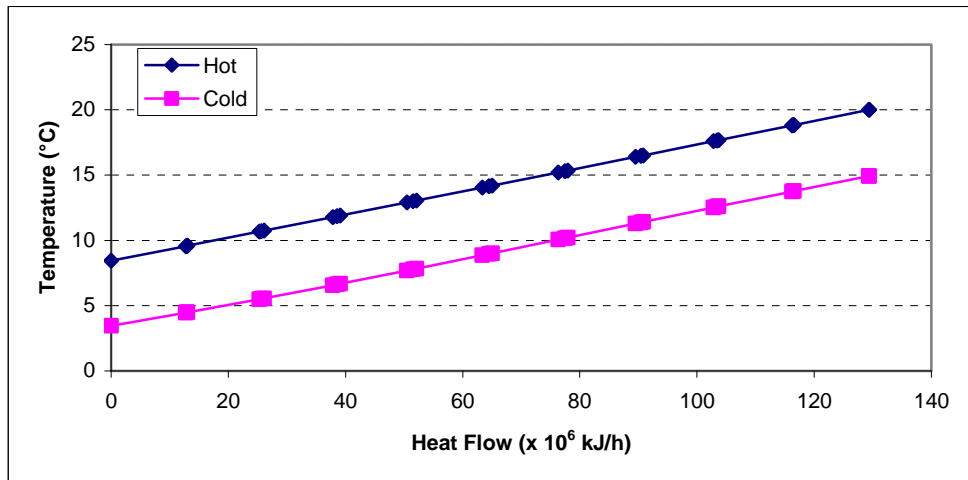


Figure B1: Composite Curve, LNG-100

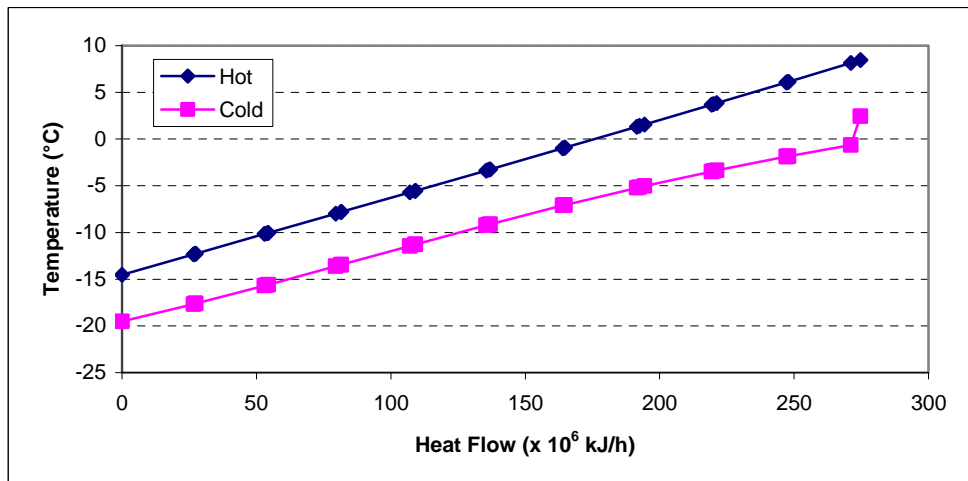


Figure B2: Composite Curve, LNG-200

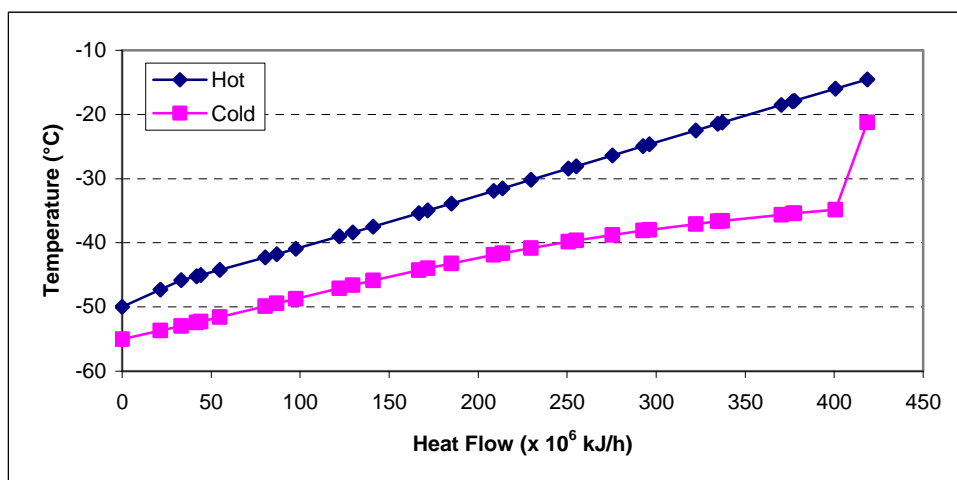


Figure B3: Composite Curve, LNG-300

C T-H DIAGRAMS, CASE VIII

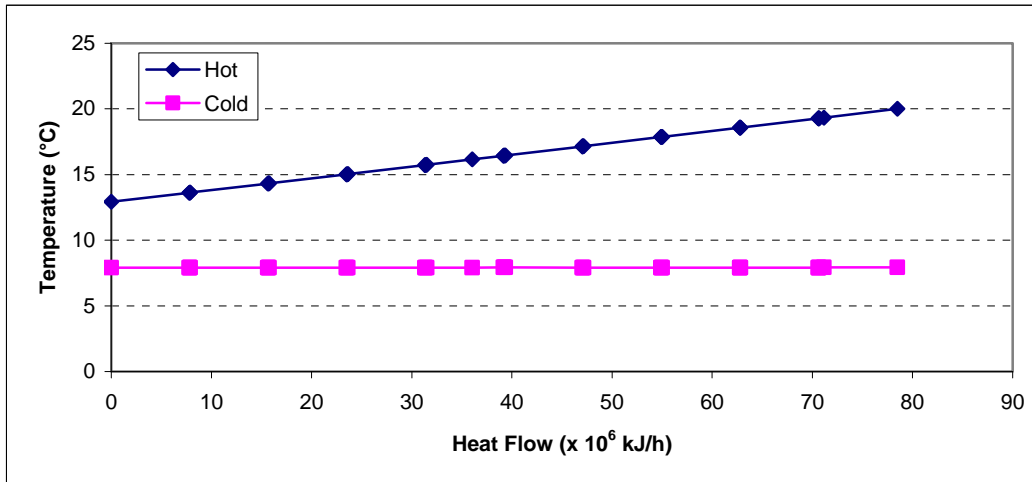


Figure C1: Composite Curve, LNG-100

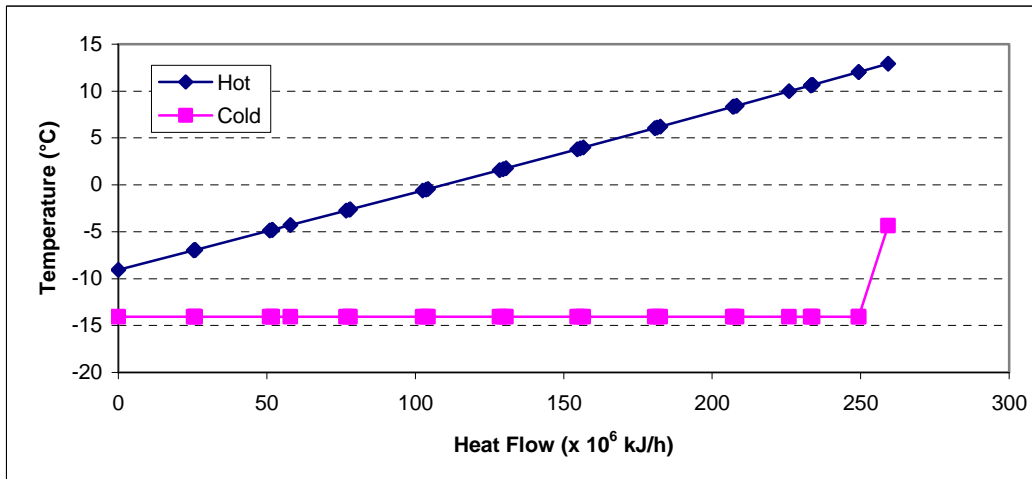


Figure C2: Composite Curve, LNG-200

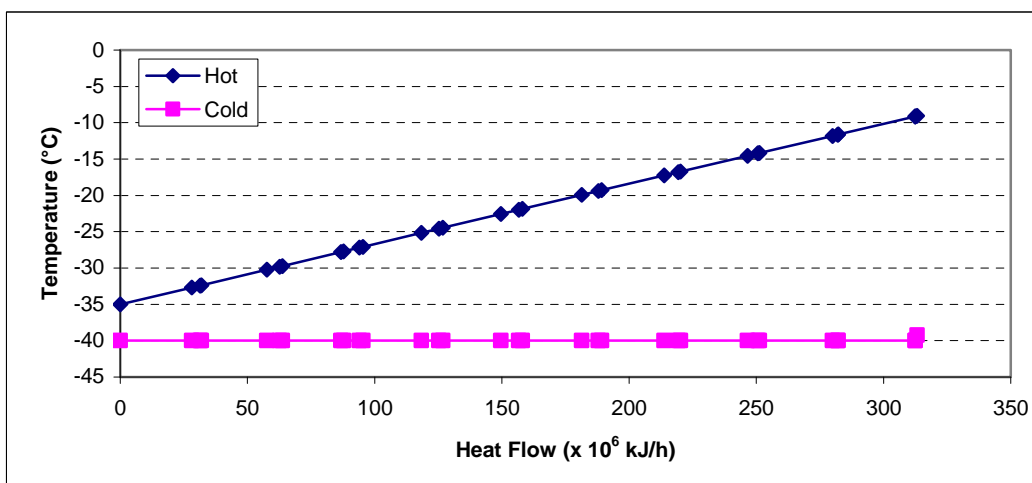


Figure C3: Composite Curve, LNG-300

E MATLAB Documentation

main.m

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%  
%  
%   UNISIM COMMUNICATION                                     %  
%   AUTHOR: DANIEL GREINER EDVARDBSEN                       %  
%   DATE: DECEMBER 5, 2010                                  %  
%   PLACE: DEPARTMENT OF CHEMICAL ENGINEERING, NTNU, TRONDHEIM %  
%  
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%  
  
% This m-file accesses the mymodel.usc model in UniSim Design and provides  
% all communication in between MATLAB and the UniSim software.  
  
clc;  
clear all;  
  
% Establishing access to the UniSim Design Server and mymodel.usc  
h = actxserver('UnisimDesign.Application');  
hyCase = h.ActiveDocument ;  
solv = hyCase.Solver ;  
f = hyCase.Flowsheet ;  
m = f.MaterialStreams ;  
  
%Initial values  
x0 = 1.0e+003 * [  
  
    1.5650  
    0.8000  
    0.0240  
    0.0158  
    0.0080  
    0.0081  
    0.0081  
   -0.0145  
   -0.0145  
   -0.0500  
   -0.0500  
    ];  
  
%Scaling  
global scale1;  
scale1 = [500 500 2 2 2 ] 5 5 5 5 5 5]';  
  
u0 = x0./scale1;  
u0 = u0(:,1);  
  
N = size(x0);  
  
%Lower bounds  
lb = [1500  
    700  
    23.8  
    15.7  
    7.9
```

```

7
7
-15
-15
-51
-51] ;

%Upper bounds
ub = [1600
      900
      24.1
      15.9
      8.1
      9
      9
      -14
      -14
      -49
      -49] ;

lb=lb./scale1;
ub=ub./scale1;

options = optimset('Display','iter', 'Diagnostics', 'on', 'TolFun', 1e-12,
'PlotFcns', @optimplotfirstorderopt, 'DiffMinChange', 1e-5, 'TolX', 1e-4,
'MaxFunEvals', 100*N(1), 'Algorithm', 'interior-point');

%Optimization
[u,fval,exitflag,output,lambda,grad,hessian] =
fmincon(@myfun,u0,[],[],[],[],lb,ub,@mycon,options);

```

myfun.m

```

function J = myfun(u)

% Establishing access to the UniSim Design Server and mymodel.usc
h = actxserver('UnisimDesign.Application');
hyCase = h.Activedocument ;
solv = hyCase.Solver ;
f = hyCase.Flowsheet ;
m = f.MaterialStreams ;

global scale1;
x = u.*scale1;

solv.CanSolve = 0 ;

m.Item('MR_1').PressureValue = x(1); %kPa
m.Item('MR_2').PressureValue = x(2);
m.Item('MR').MolarFlowValue = x(3); %kmole/s
m.Item('MR_2*').MolarFlowValue = x(4);
m.Item('MR_3*').MolarFlowValue = x(5);
m.Item('NG1').TemperatureValue = x(6);
m.Item('MR1').TemperatureValue = x(7);
m.Item('NG2').TemperatureValue = x(8);
m.Item('MR2').TemperatureValue = x(9);
m.Item('NG3').TemperatureValue = x(10);
m.Item('MR3').TemperatureValue = x(11);

```

```

solv.CanSolve = 1;

J = f.Operations.Item('Efficiency').Cell('B5').CellValue;

```

mycon.m

```

%The constraints in mymodel.usc are established. There are constraints in all
%the heat exchangers, superheating, temperatures after air cooling,
%precooled NG and MR temperatures
function [c,ceq] = mycon(u)

% Establishing access to the UniSim Design Server and mymodel.usc
h = actxserver('UnisimDesign.Application');
hyCase = h.Activedocument ;
solv = hyCase.Solver ;
f = hyCase.Flowsheet ;
m = f.MaterialStreams ;
global scale1;
x = u.*scale1;

solv.CanSolve = 0 ;

m.Item('MR_1').PressureValue = x(1); %kPa
m.Item('MR_2').PressureValue = x(2);
m.Item('MR').MolarFlowValue = x(3); %kmole/s
m.Item('MR_2*').MolarFlowValue = x(4);
m.Item('MR_3*').MolarFlowValue = x(5);
m.Item('NG1').TemperatureValue = x(6);
m.Item('MR1').TemperatureValue = x(7);
m.Item('NG2').TemperatureValue = x(8);
m.Item('MR2').TemperatureValue = x(9);
m.Item('NG3').TemperatureValue = x(10);
m.Item('MR3').TemperatureValue = x(11);

solv.CanSolve = 1;

%UA-values in heat exchangers
ceq = 1e-13*[

    f.Operations.Item('LNG').Cell('D5').CellValue - 8497454 %LNG-100 MR
    f.Operations.Item('LNG').Cell('D6').CellValue - 17550335 %LNG-200 MR
    f.Operations.Item('LNG').Cell('D7').CellValue - 18873920 %LNG-300 MR
    f.Operations.Item('LNG').Cell('E5').CellValue - 1293992 %LNG-100 NG
    f.Operations.Item('LNG').Cell('E6').CellValue - 2623097 %LNG-200 NG
    f.Operations.Item('LNG').Cell('E7').CellValue - 3126706 %LNG-300 NG
];

a=1e-3;
c = [
    a*f.Operations.Item('SUPCON').Cell('B4').CellValue %K-100
    a*f.Operations.Item('SUPCON').Cell('C4').CellValue %K-200
    a*f.Operations.Item('SUPCON').Cell('D4').CellValue %K-300

    -f.Operations.Item('LNG-100').MinApproachValue
    -f.Operations.Item('LNG-200').MinApproachValue
    -f.Operations.Item('LNG-300').MinApproachValue ];

```

F Risk Assessment