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Evaluation of Natural Gas Liquefaction Processes for Floating Applications Offshore

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Master of Science in Product Design and Manufacturing
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Problem Description

The main objective of this Master thesis is to carry out a thorough thermodynamic analysis of a few suitable liquefaction processes for natural gas in offshore (floating) applications, however, main emphasis should be placed on the NicheLNG process. The objective also includes considering ways to improve the energy efficiency of the NicheLNG process.

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Department of Energy and Process Engineering

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MASTER THESIS

for

Student Øyvind Eckhardt

Autumn 2009

Evaluation of Natural Gas Liquefaction Processes for Floating Applications Offshore

Evaluering av flytendegjøringsprosesser for naturgass ved FPSO løsninger offshore

Background

LNG is the fastest growing energy carrier in the world, and ship based transport of LNG is expected to increasingly become an important alternative to pipeline transport. In this market, Höegh LNG is operating traditional LNG ships while the company at the same time looks at new and innovative solutions in floating value chains for LNG. These solutions start with floating production (FPSO = Floating Production, Storage and Offloading), continues with the ship based transport and ends with regasification, either in the form of SRV ships (Shuttle and Regasification Vessel) targeting small to medium gas volumes and short to medium transport distances, or in the form of FSRU ships (Floating Storage Regasification Unit) for medium to large gas volumes and medium to large transport distances.

A Master thesis in the spring of 2009 discussed briefly alternative processes for offshore liquefaction of natural gas in an FPSO setting, and focused on energy and environmental aspects for one of these processes; the NicheLNG concept. While the Master thesis mentioned above had a broad perspective, this Master thesis should focus more deeply on the thermodynamic qualities (energy efficiency) of a few alternative liquefaction processes, while also considering a potential expansion in capacity of 25-30%.

Objective

The main objective of this Master thesis is to carry out a thorough thermodynamic analysis of a few suitable liquefaction processes for natural gas in offshore (floating) applications, however, main emphasis should be placed on the NicheLNG process. The objective also includes considering ways to improve the energy efficiency of the NicheLNG process.

The Master thesis should address the following Tasks:

1. A literature survey of alternative natural gas liquefaction processes should be performed with emphasis on those that are suitable for offshore applications when considering weight and space requirements, complexity and safety. The process efficiency expressed for example in MWh energy per ton LNG is of course a key parameter in this evaluation.

2. The NicheLNG process and possibly one alternative (if there is a promising one) process from the previous subtask should be analyzed using thermodynamics to evaluate their energy efficiencies. The improvement potential for the same processes should also be addressed with respect to energy savings, added investment and added complexity of the processes. The following activities are envisaged:

    a. Adjust and/or expand existing simulation models that are needed to study liquefaction processes for an FPSO application.

    b. Apply thermodynamic tools (numerical and/or graphical) to evaluate the energetic quality of the NicheLNG process and possibly one promising alternative process.

    c. Suggest improvements to the NicheLNG process, while presenting figures for energy savings, qualitatively indicating need for added investment, and discussing the effects on plant complexity in design and operation.

3. Consider the possibility to expand the NicheLNG process with one or two expander/compressor stages, to give an indication about improvement potentials with increased capacity (typically 25-30% expansion of the liquefaction process).

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

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

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

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Two – 2 – copies of the thesis shall be submitted to the Department. Upon request, additional copies shall be submitted directly to research advisors/companies. A CD-ROM (Word format or corresponding) containing the thesis, and including the short summary, must also be submitted to the Department of Energy and Process Engineering.

Department of Energy and Process Engineering, 10.09.2009

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Preface

This master thesis is a result of the study during my graduation at NTNU. It has been carried out at the university and at Höegh LNG between September 2009 and February 2010. The focus has been on thermodynamic qualities and potentials of the NicheLNG process.

I specially want to thank my supervisor at NTNU, Professor Truls Gundersen for his support and feedback during my work. I would also like thank my industrial contact at Höegh LNG, Vegard Hellekleiv for given me the opportunity to work on this project. Lastly, I would like to express my gratitude to Lars Petter Revheim and Thomas Larsen at Höegh LNG for valuable help and supplementary feedback.

Trondheim, 15.02.2010

Øyvind Eckhardt

Sammendrag

Litteraturstudie av LNG prosesser egnet for offshore har blitt evaluert. Undersøkelsen er utført med vekt på plassbehov, kompleksitet, effektivitet og sikkerhet. Ved bruk av termodynamikk er NicheLNG prosessen beskrevet. Flytendegjøringsprosessen på HLNG FPSO-1 har blitt vurdert med hensyn til energiforbruk, forbedringer og muligheten for å utvide prosessen for å gi en indikasjon på dens potensial. I tillegg har en alternativ LNG prosess blitt sammenlignet med NicheLNG prosessen.

Prosesser med dobbel ekspander kjølekretser basert på nitrogen som kjølemedium er den mest foreslåtte løsningen for offshore produksjon. Derfor ble den valgt som en alternativ prosess til NicheLNG. I vurderingen av de to prosessene ble simuleringer gjort med likeverdige vilkår. NicheLNG prosessen, basert på en åpen metan krets og en nitrogen krets, hadde en lavere massestrøm som resulterte i 10% lavere energiforbruk. Valg av kjølemediet (metan eller nitrogen) har forskjellig spesifikk varmekapasitet og dermed en innflytelse på massestrømmen. For en gitt kjøleytelse krever metan som kjølemiddel mindre massestrøm enn nitrogen. I tillegg vil høyere trykk nivåer bidra til økt effektivitet og redusere størrelser på utstyr.

I den åpne kjølekretsen til NicheLNG prosessen blir metan kjølt ned til -1,5°C før den blir ekspandert. Hvis nedkjølingen blir utvidet til -10°C før ekspansjon er det mulig å oppnå høyere effektivitet for metan kretsen.

Kriteriene ved sammenligning av LNG prosesser er viktig når kvaliteten skal fastsettes. Fødegassen og produkt spesifikasjonene gir restriksjoner på oppnåelig effektivitet. Med økende fødegasstrykk kreves det mindre arbeid (fra fødegass til LNG), men på grunn av den høye virkningsgraden til kompressoren og den lave virkningsgraden for selve flytendegjøringen vil den totale virkningsgraden reduseres. Arbeidet flytendegjøringen krever blir mer dominerende enn kompressorarbeidet med høyere fødegasstrykk. Derfor bør virkningsgraden av LNG prosesser ikke beregnes fra sitt fødegasstrykk men fra tilstanden etter fødegasskompressor. For NicheLNG prosessen ble eksergi virkningsgraden av flytendegjøringsdelen beregnet til 26,6%, ved et inngangstrykk på 75 bar.

En vurdering av økt LNG produksjon med vekt på forbedringer for å holde energiforbruket nede ble undersøkt. De fire undersøkte løsningene var; End Flash Gas, væskefaseturbin, en ekstra kompressor og økning av varmeveksler areal. Væskefaseturbin var forbedringen som skiller seg ut som den med høyest bidrag til effektiviteten. Med en 25% økning i LNG produksjon og med de nye enhetene og modifikasjonene reduserte det spesifikke arbeidet fra 0,5502 kWh / kg_{LNG} til 0,4791 kWh / kg_{LNG}. Disse forbedringene kan rettferdiggjøre økte investeringskostnader ved 25% høyere LNG produksjon siden energiforbruket reduserte med 12,9% enn for den opprinnelige utformingen av NicheLNG prosessen. Grunnet begrenset med plass og vektkapasitet på en FPSO må dette tas hensyn til ved en forandring av prosessen.

Summary

A literature survey of LNG processes suitable for offshore environment has been evaluated. The survey has been performed with emphasis on space requirements, complexity, efficiency and safety. Thermodynamics theory is described and used in the investigation of the NicheLNG process. The liquefaction part of the HLNG FPSO-1 has been evaluated with respect to its energy consumption, improvements and the possibility to expand the process to give an indication about improvement potentials. In addition, one alternative liquefaction process has been compared with the NicheLNG process.

Dual expander processes based on nitrogen as refrigerant are the most proposed solution suitable for offshore applications. Therefore it was chosen as an alternative process to NicheLNG. In the investigation of the two processes the processes simulated were with equal conditions. The NicheLNG process, based on an open methane cycle and a nitrogen cycle, had a significantly lower mass flow rate resulting in 10% lower power consumption. Decision of chosen refrigerant gas (methane or nitrogen) has different specific heat capacity and hence an influence on the flow rate. Methane as refrigerant requires less mass flow rate than nitrogen for a given duty. In addition, higher pressure levels will contribute to increased efficiency and reduced unit sizes.

In the open refrigeration cycle of the NicheLNG process, methane is cooled down to -1,5°C before it is expanded. If the internal heat exchange is extended to -10°C before expansion is it possible to achieve some efficiency increase for the methane cycle.

Comparison criteria are important when the quality of liquefaction processes is to be determined. Feed and product specifications provide some restrictions on obtainable efficiency. With an increasing feed gas pressure, the whole liquefaction process (from feed to LNG) demands less work, but the overall process efficiency is reduced due to the high efficiency of the feed gas compressor and the low efficiency of the liquefaction part. Hence, the efficiency of the liquefaction process should not be calculated from its feed gas pressure but rather the liquefaction pressure. For the NicheLNG process, the exergy efficiency of the liquefaction part was calculated to 26,6%, with a liquefaction pressure at 75 bar.

An increase of LNG production with emphasis on improvements to keep work consumption down was also discussed. The four evaluated solutions were utilization of End Flash Gas, liquid expander, additional compressor and increase of heat exchanger area. The liquid expander was the improvement that stands out as the highest contribution to the efficiency. With a 25% increase in LNG production and with new units and modifications of the design resulted in a reduction in the specific work consumption from 0,5502 kWh/kg_{LNG} to 0,4791 kWh/kg_{LNG}. These efficiency improvements can justify higher investment costs since the work consumption, with 25% higher LNG production, was 12,9% lower than for the original design of the NicheLNG process. Never the less, space and weight on a FPSO are limited and has to be considered when a more efficient process is desired.

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Nomenclature

Abbreviations

BTU	British Thermal Unit
CB&I	Chicago Bridge & Iron
CH4/C1	Methane
C2H6/C2	Ethane
C3H8/C3	Propane
CO_2	Carbon dioxide
Comp.	Compressor
COP	Coefficient of Performance
C3MR	Propane Mixed Refrigerant
DMR	Dual Mixed Refrigerant
EFG	End Flash Gas
FEED	Front End Engineering Design
FPSO	Floating, Production, Storage and Offloading vessel
FPSO-1	The first planned FPSO
HC	Hydrocarbon
HLNG	Höegh LNG
HX	Heat Exchanger
LMTD	Log Mean Temperature Difference
LNG	Liquefied Natural Gas
LPG	Liquefied Petroleum Gas
NG	Natural Gas
N2	Nitrogen
N2-Exp	Nitrogen Expander
O_2	Oxygen
Scf	Standard Cubic Foot
SMR	Single Mixed Refrigerant
SRV	Shuttle and Regasification Vessel
T-s diagram	Temperature-entropy diagram
2DLE	NicheLNG original design

Letters

A	Area
C_p	Specific Heat Capacity
e	Specific Exergy
F	Correction factor
h	Enthalpy
m	Mass rate
n	Mole rate
P	Pressure
s	Entropy
T	Temperature
T_0	Ambient temperature
Q	Heat duty/rate
U	Heat Transfer Coefficient
W	Work
μ_J	Joule-Thomson Coefficient
η	Efficiency

Prefixes

k	kilo	10^3
M	Mega	10^6
G	Giga	10^9

Introduction

For many years, several companies have been working on realization of offshore LNG production. A proposed production facility is an LNG Floating, Production, Storage and Offloading vessel (FPSO) as foundation for the liquefaction application. Destinations of an FPSO are isolated gas field. In gas fields remote from land may it be uneconomic to build up an infrastructure to exploit the reserves. In addition is associated gas from offshore oil production that is flared or re-injected into reservoir a possible placement.

At the present time, Höegh LNG is operating LNG ships for transport and is about to hand over two SRV ships (Shuttle and Regasification Vessel). The company wants to have a solution in a floating value chain for LNG with integration of a LNG production facility. The development of an FPSO for LNG production has reached the end of the FEED (Front End Engineering Design) phase. The design is not yet set since final destination is unknown and an increase in production rate may be desired. A possibility of higher LNG production or better energy efficiency can give an advantage in the future customer negotiations.

This master thesis takes the original design of the NicheLNG process and compares it with other liquefaction processes suitable for offshore applications. Restricted to offshore production, some aspects have higher importance than for land based facilities. Potentials and improvements are analyzed both from a practical viewpoint but also some hypothetical situations are discussed. Since detailed information on the processes is restricted, some considerations have been made to get equal assumptions to compare processes. This will be expressed later in the thesis.

The theory of the underlying thermodynamic calculations is expressed followed by a literature survey of liquefaction processes suitable in an offshore environment. These processes are then compared in energy efficiency, chosen design and refrigerant medium. Then the improvement potential with emphasis on the most promising process is analyzed. An indication of the potential expansion in capacity is done by manipulating refrigerants and with a higher equipment count. In contrast to the original design analysis, the evaluation of increased capacity has a more practical view.

Two energy saving sources have in an earlier master thesis been identified and evaluated, and will therefore not be treated in this work. It covered the benefit in terms of energy consumption with lower cooling water temperature and removal of the NGL extraction process.

1 Theory

Calculations performed in this thesis are based on well known thermodynamics and simulations in AspenTech HYSYS. This chapter provides an overview of the thermodynamic principles with emphasis on cryogenic processes.

1.1 Compression and expansion

Compressors used in the liquefaction section are the main consumer of energy. The amount of consumed work depends on the inlet and outlet state. From the first law of thermodynamics [1]:

$$\frac{dE_{cv}}{dt} = \dot{Q} - \dot{W} + \sum_i \dot{m}_i \left(h_i + \frac{\dot{V}_i}{2} + gz_i \right) - \sum_e \dot{m}_e \left(h_e + \frac{\dot{V}_e}{2} + gz_e \right) \tag{1.1}$$

For a steady state system with one inlet, one exit and neglecting change in potential energy, the energy balance is reduced to:

$$0 = \dot{Q} - \dot{W} + \dot{m} \left(h_i + \frac{\dot{V}_i}{2} \right) - \dot{m} \left(h_e + \frac{\dot{V}_e}{2} \right) \tag{1.2}$$

When accounting for entropy, enthalpy and pressure changes, work can be expressed by the magnitude of the specific volume of the fluid (Δke = Δpe = 0) [1]:

$$\left(\frac{\dot{W}_{cv}}{\dot{m}} \right)_{\text{int}\,rev} = -\int_i^e v\,dp \tag{1.3}$$

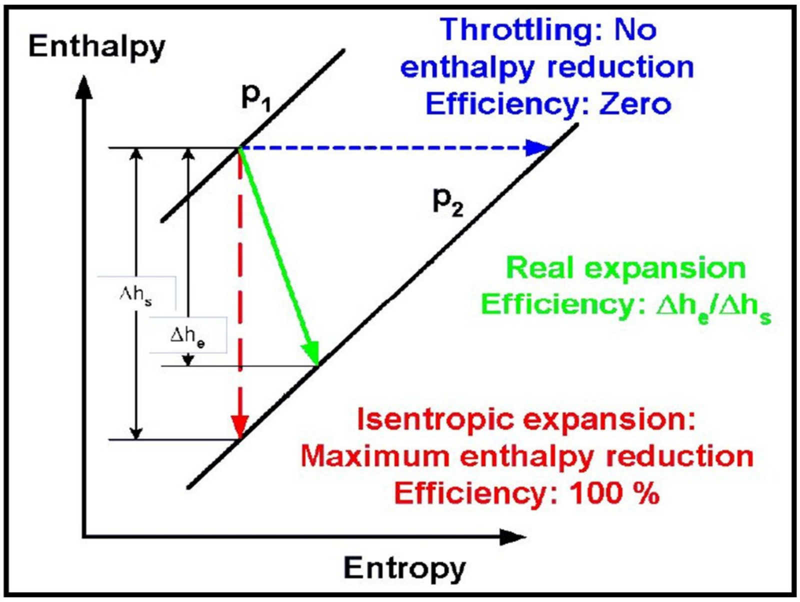


Figure 1.1 Enthalpy – entropy diagram. Isenthalpic and isentropic expansion illustrated with arrows [12]

Figure 1.1 illustrates throttling and expansion of a gas. From the relation between the enthalpy change for real and isentropic expansion, $\Delta h_e/\Delta h_s$, the efficiency can be found. In cryogenics the efficiency is defined as the reduction of the enthalpy and not the gained mechanical power [12]. An expansion without an enthalpy reduction is achieved by throttling. The expansion absorbs no heat and does no work. An isenthalpic expansion is defined by the Joule-Thomson coefficient:

$$\mu_J = \left(\frac{\partial T}{\partial p}\right)_h \tag{1.4}$$

A gas that is cooled through a valve has a positive μ_J. When negative μ_J occur, the temperature increases by an expansion. The phenomenon of increased temperature is of particular importance when handling gases at very low temperatures such as helium.

An isentropic expansion is also illustrated in figure 1.1. The isentropic expansion corresponds to a process with no internal irreversibilities. This is an ideal expansion where maximum achievable work is developed. In a real expander, entropy is produced resulting in a higher outlet enthalpy value. Thus, less work is produced.

1.2 Coefficient of Performance

A refrigeration process withdraws heat at rate Q_C from a cold source at temperature T_C. This is then delivered to a warmer reservoir at rate Q_H and temperature T_H. To accomplish this, work input is necessary. As figure 1.2 illustrates, required work to perform this cooling is $W=Q_H-Q_C$. Higher T_C results in higher Q_C, and thus lower work consumption. With increasing T_C the efficiency of a refrigeration process is increased, as eq. 1.5 and 1.6 express.

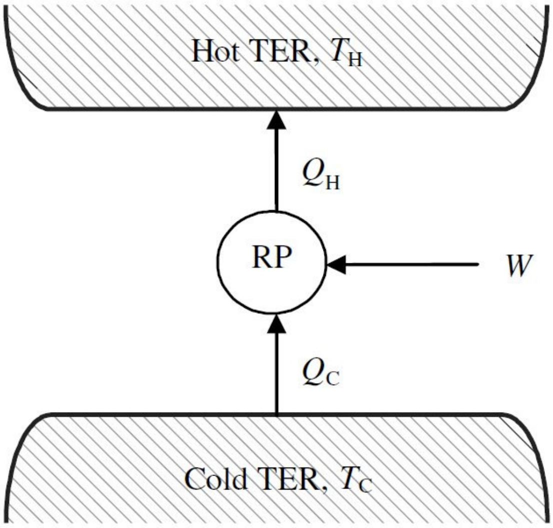


Figure 1.2 A refrigeration process [13]

The efficiency of a refrigeration process is commonly defined by a Coefficient of Performance (COP), and is defined [1]:

$$COP = \frac{Q_C}{W} \tag{1.5}$$

and theoretically maximum [1]:

$$COP_{th,\max} = \frac{1}{\left(\frac{T_H}{T_C}\right) - 1} \tag{1.6}$$

From Eq. 1.6 and with hot reservoir temperature at 38°C as $T_H=T_0$, the maximum COP as a function of cold temperature T_C is plotted in Figure 1.3. The cold temperature T_C is the temperature the cooling duty is delivered.

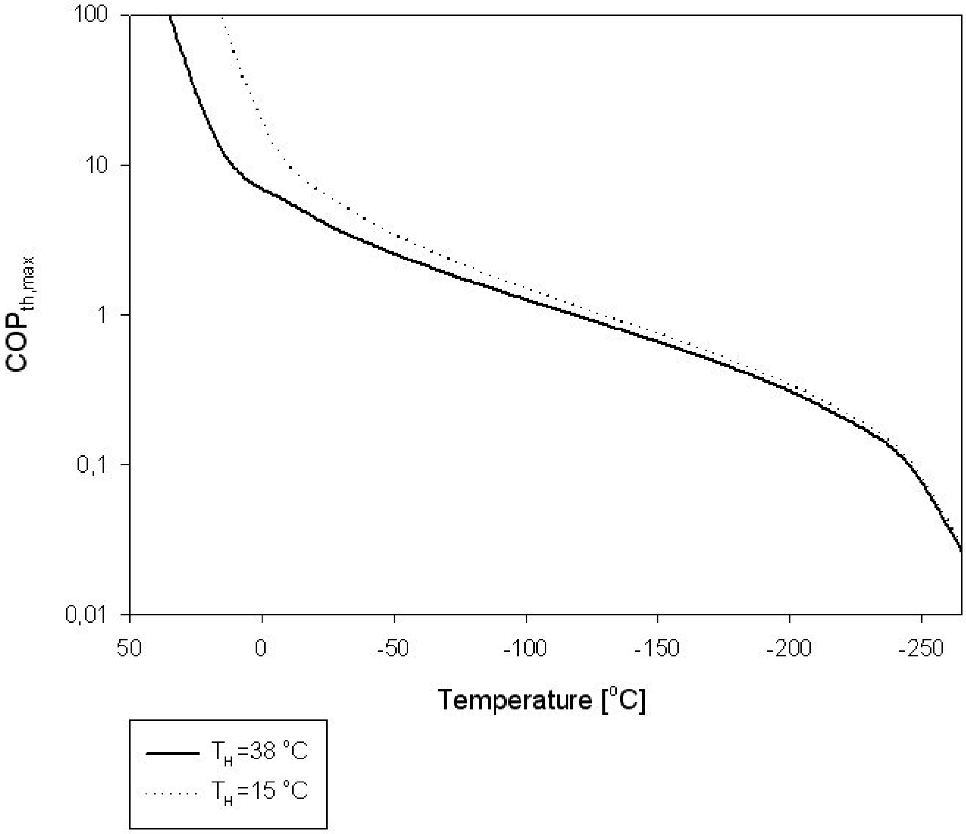


Figure 1.3 Coefficient of Performance as a function of temperature

The dotted line in figure 1.3 expresses the $COP_{th,max}$ at T_H=15 °C. A lower sea water temperature results in a higher COP, hence a more efficient refrigeration process.

In refrigeration processes, the reversed Carnot cycle can be used as an illustration of the theory. An ideal gas with heat absorption and rejection at constant temperatures is illustrated in the T-s diagram in Figure 1.4. The reversible process in Figure 1.4 is stated through isentropic compression (1-2), isothermal compression (2-3), isentropic expansion (3-4) and isothermal expansion (4-1). Work can then be represented as W and heat extracted as Q_C in Figure 1.4.

The COP for the Carnot cycle is expressed as in Eq. 1.6.

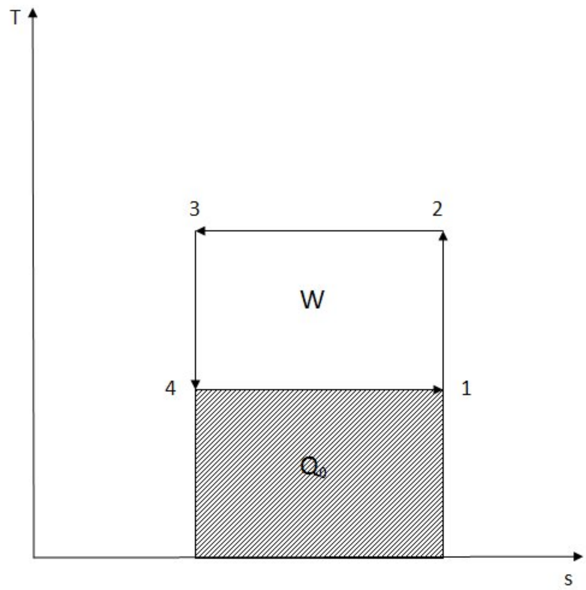


Figure 1.4 Carnot reversed cycle

1.3 Brayton Refrigeration Cycle

A Brayton refrigeration cycle is a reversed Brayton cycle. Heat is transported from a cold reservoir, where the temperature after an expansion is below that of the cold reservoir. Refrigeration is then achieved by attracting heat from a cold region and then later released. Different from other refrigeration cycles this cycle involves no phase change. The working fluid remains as a gas throughout.

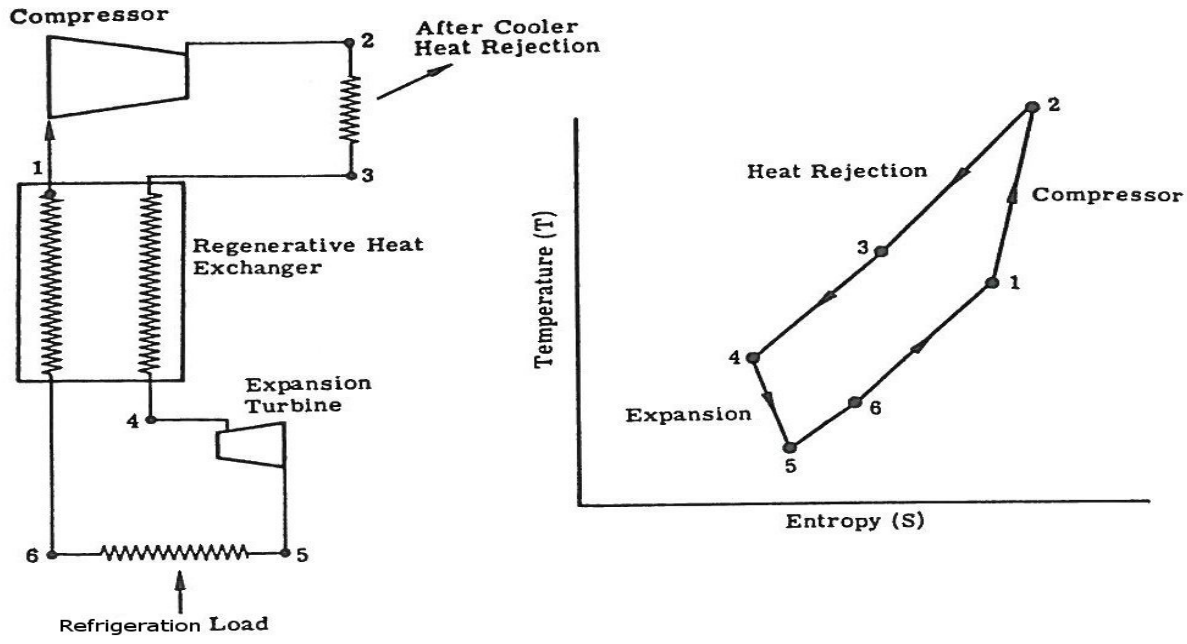


Figure 1.5 Flow sheet and T-s diagram of a reversed Brayton cycle [18]

Figure 1.5 illustrates a refrigeration process based on the reversed Brayton cycle. The T-s diagram represents a real cycle. An ideal cycle operates with isentropic turbine and compressor. As can be seen, the expansion and compression is not isentropic so a real process operates with some losses. Heat transfer from the cold region is from 5-6 and then the heat is released after a compression from 2-3.

1.3 Exergy

Exergy is a measure of the maximum amount of work that can be extracted from a process stream when it is brought to equilibrium with is surroundings in a hypothetical reversible process [1]. When neglecting changes in composition and chemical exergy, this is a measure of the potential in thermo-mechanical exergy and thus defined only in terms of the stream enthalpy and entropy relative to the surroundings. The exergy, e, expressed at steady-state conditions and neglecting kinetic and potential energy [2]:

$$e = (h - T_0 s)_{T,P} - (h - T_0 s)_{T_0 P_0} \tag{1.7}$$

where T_0 and P_0 are at ambient conditions. When taken from one state to another, the change in exergy is given by:

$$\Delta e = (h - T_0 s)_{T_2,P_2} - (h - T_0 s)_{T_1 P_1} \tag{1.8}$$

In a real process irreversibilities exist. So actual work required to bring a process to a state is more than in an ideal case. Given by the second law of thermodynamics, over an actual system, lost work from compression can be defined as the difference between actual work required and the change in exergy:

$$W_{lost} = W_{actual} - \Delta e \tag{1.9}$$

and lost work from expansion:

$$W_{lost} = \Delta e - W_{actual} \tag{1.10}$$

When exergy production and losses are known the exergy efficiency can be decided. Exergy efficiency is defined as the relation between the exergy change of natural gas to be liquefied and the power consumed. Exergy efficiency is defined as:

$$\eta_{ex} = \frac{Minimum\ power\ for\ liquefaction}{Power\ Consumption} \tag{1.11}$$

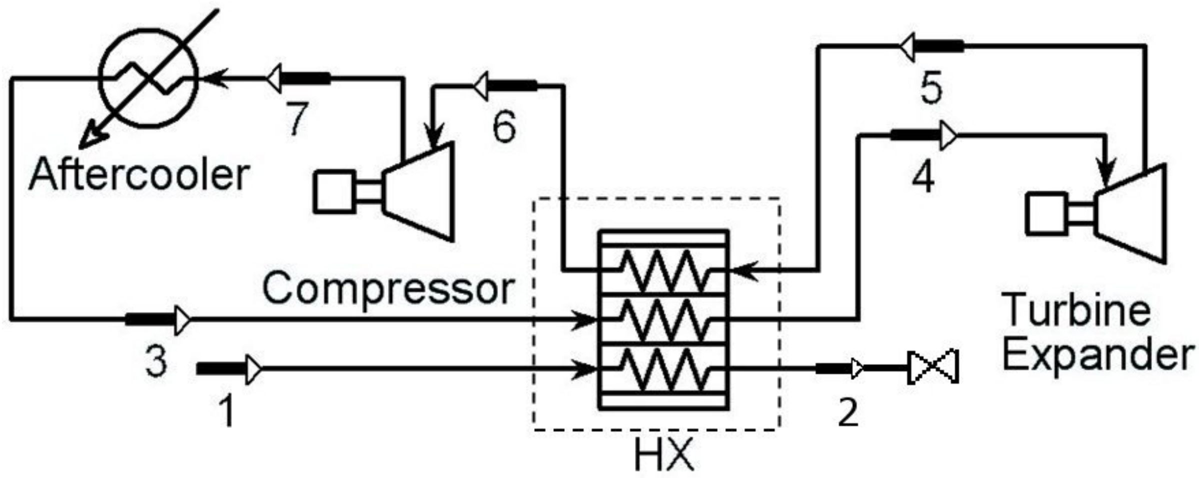


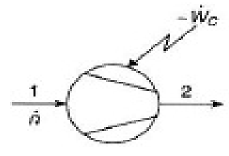
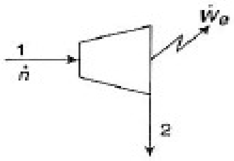
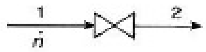
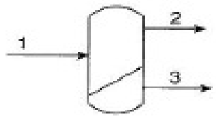
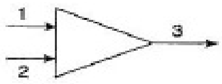
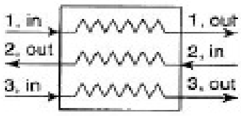
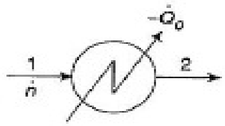
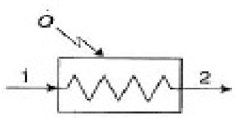
Figure 1.6 Simple illustration of an expander refrigeration process

Figure 1.6 shows a simple a refrigeration process, consisting of a compressor, aftercooler, heat exchanger and expander. The compressor consumes work by increasing the pressure of a refrigerant to a desired level. Necessary cooling is obtained by expanding the refrigerant through a turbine and this will also produce work. From consumed and produced work the exergy efficiency can be expressed as:

$$\eta_{ex} = \frac{\dot{n}_1 (e_2 - e_1)}{-\dot{W}_{comp} - \dot{W}_{\exp}} \tag{1.12}$$

According to [3], the placement of state 2 is in front of the valve when exergy change from feed to after heat exchanger is to be calculated.

Calculations of exergy losses through the components are done as expressed in Table 1.1:

Equipment	Symbol	Exergy loss (kW)
Compressor		$\Delta ex_{loss} = \dot{n}(ex_1 - ex_2) - \dot{W}_c$
Expander		$\Delta ex_{loss} = \dot{n}(ex_1 - ex_2) - \dot{W}_e$
Throttle valve		$\Delta ex_{loss} = \dot{n}(ex_1 - ex_2)$
Phase separator or stream splitter		$\Delta ex_{loss} = \dot{n}_1 ex_1 - \dot{n}_2 ex_2 - \dot{n}_3 ex_3$
Stream mixer		$\Delta ex_{loss} = \dot{n}_1 ex_1 + \dot{n}_2 ex_2 - \dot{n}_3 ex_3$
Heat exchanger		$\Delta ex_{loss} = \sum_{i=1}^{n} \dot{n}_i (ex_{i,in} - ex_{i,out})$
Condenser or aftercooler exchanging heat with ambient		$\Delta ex_{loss} = \dot{n}(ex_1 - ex_2)$
Evaporator operating at low temperature		$\Delta ex_{loss} = \dot{n}(ex_1 - ex_2) + \dot{Q}(1 - T_o/T)$

$\dot{W}_c < 0, \dot{W}_e > 0, \dot{Q} > 0, \dot{Q}_o < 0.$

Table 1.1 Exergy losses in different components [3]

From Table 1.1 the exergy losses of components in a liquefaction system can be found, and thereby the exergy efficiency of a component. Comparing different liquefaction processes the exergy losses of components in each process may differ and a more detailed overview of the processes may give advantages in optimizing them. With an overview of the exergy losses from each component may an optimization be easier to accomplish.

1.4 Heat Exchanger Duty

Transfer of heat through a heat exchanger is typically done for three different reasons. Either a stream needs to be heated or cooled, or a liquid stream needs to be vaporized, or a vapor stream needs to be condensed. To transfer heat, one rule has to be satisfied from the Second Law of Thermodynamics; heat can only be transferred from a higher temperature to a lower one. This means that the higher temperature cooling curve and the lower temperature heating curve cannot intersect.

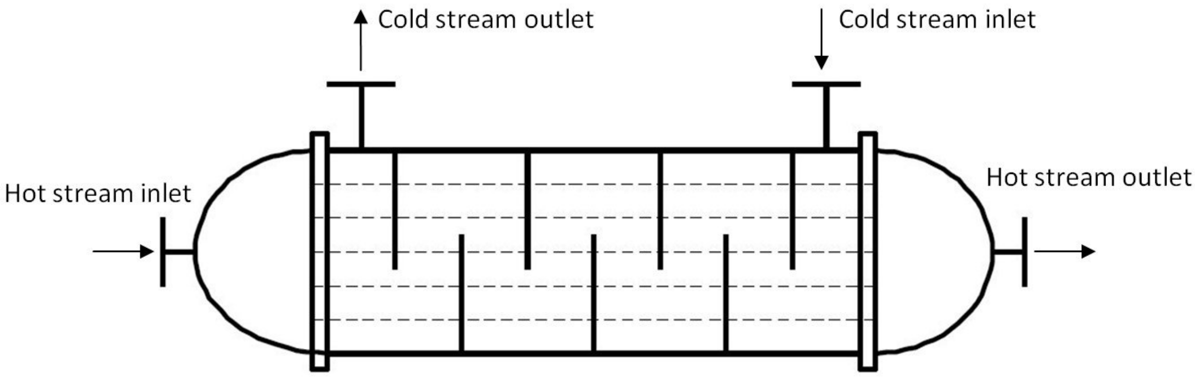


Figure 1.7 Counter current heat exchanger

The duty of a heat exchanger as in Figure 1.7 can be provided from an energy balance. If inlet and outlet conditions of a hot stream are known and the specific heat capacity assumed to be constant, the duty can be expressed as [6]:

$$\dot{Q}_H = \dot{m}\, Cp_H \left(T_{H,in} - T_{H,out}\right) \tag{1.13}$$

or similarly for a cold stream:

$$\dot{Q}_C = \dot{m}\, Cp_C \left(T_{C,out} - T_{C,in}\right) \tag{1.14}$$

To determine the temperature driving force for heat transfer, the log mean temperature difference, LMTD, is used. The use of LMTD is valid both for co-current and counter-current flow [6] as long as the specific heat capacities and the heat transfer coefficient are constant. By defining the temperature difference for each side of the heat exchanger, the LMTD is defined as follows [6]:

$$LMTD = \frac{\Delta T_L - \Delta T_R}{\ln\left(\frac{\Delta T_L}{\Delta T_R}\right)} \tag{1.15}$$

where ΔT_L is the temperature difference on the left side and ΔT_R on the right side of the heat exchanger.

When looking at a cross-flow heat exchanger a common principle is to introduce a correction factor, F. The correction factor is in the $0 < F \leq 1$ region.

With conduction and convection coefficients, the overall heat transfer coefficient, U, can be introduced. For a given heat transfer area, A, the duty of a heat exchanger can then be expressed as [6]:

$$\dot{Q} = U \times A \times LMTD \times F \tag{1.16}$$

1.5 Refrigerant medium

A traditional large onshore liquefaction plant normally contains three refrigeration cycles. Each cycle is supposed to cover different temperature regions of the natural gas to be liquefied. The purpose of the cycles is precooling, liquefaction and subcooling. When it is desirable to have two cycles the first operates as a precooler followed by a second cycle for the liquefaction and subcooling.

1.5.1 Gaseous refrigerants

If a gas is chosen as refrigerant without any phase change, operation in the cold end differs often from a mixed refrigerant. A gaseous refrigerant has the possibility of expanding through a turbine and thereby produce some work in addition to lower its temperature. Placement of expansion is complex and depends on the overall process and the chosen refrigerant gas. Decision of refrigerant gas and placement of expander are covered in Chapter 4.

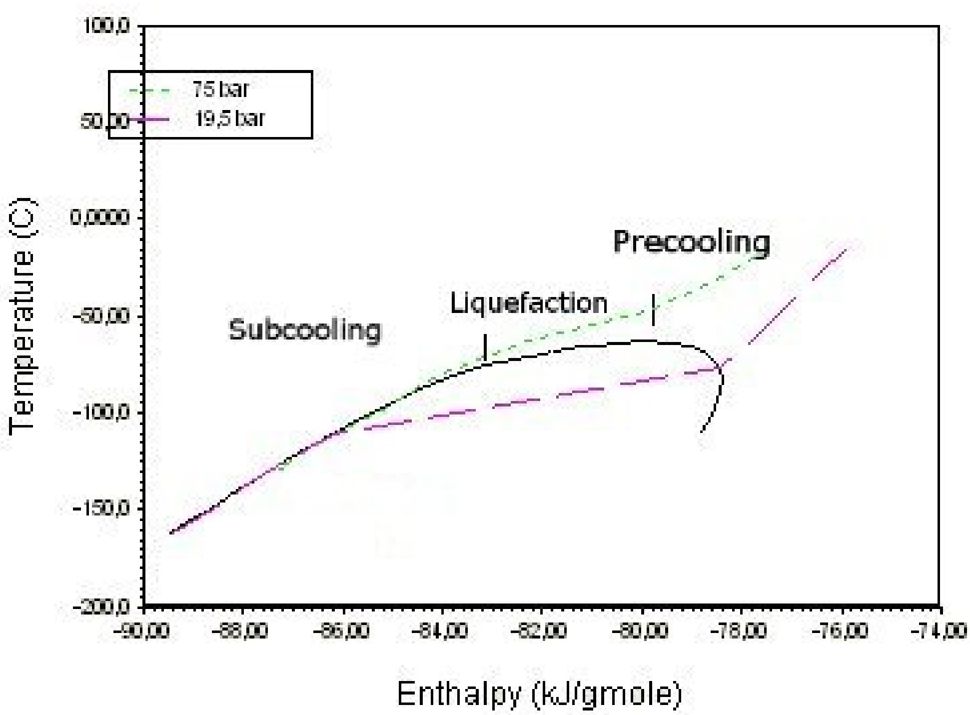


Figure 1.8 Temperature-enthalpy diagram of natural gas with cooling sequences indicated

Figure 1.8 shows the enthalpy-temperature variation for natural gas at a pressure of 19,5 bar and 75 bar. The composition of the natural gas is a typical feed gas before a liquefaction process. The isobar lines are chosen from the NicheLNG operating pressures. The 75 isobar line is the liquefaction pressure and the high pressure level of the methane cycle.

The main objective is to obtain composite enthalpy-temperature variation of hot and cold streams as close to parallel as possible. The 75 bar slope shows that the specific heat, c_p or $(\Delta h/\Delta T)_p$, differ with reducing temperature. As can be seen, the natural gas at 75 bar has three different gradients. Hence, utilization of three refrigerant cycles with different composition will be the best option.

Figure 1.8 also shows that the feed gas pressure influence the slope with reducing temperature. The chosen pressure depends on the composition of the feed gas. Higher fraction of heavier hydrocarbons results in higher feed gas pressure to avoid early entrance in the two phase area. Even though feed gas may differ with different liquefaction processes the variation cannot be large since the end product specifications (composition and higher heating value) have to be almost equal.

The discussion above is done with pretreatment in mind before natural gas is fed into the liquefier. Some liquefaction plants have the removal of impurities, as CO_2 and water, and/or heavy hydrocarbons integrated in the liquefaction process. Integration will influence the chosen feed pressure.

1.5.2 Mixed refrigerants

A liquid containing a pure refrigerant, as propane, evaporates at constant temperature. With single component evaporation the temperature profile will be horizontal as long as some liquid remains. If a fluid has a mixed composition is the temperature profile depending on the chosen fluids and the mixture composition. The mixed refrigerant must contain fluids with boiling temperatures which cover the whole temperature range. Natural gas to be liquefied needs to be cooled from an ambient temperature to -162°C. To cover this range, a composition of different hydrocarbons and nitrogen is often used. Concepts of different mixed refrigerants processes are either as a single mixed cycle or as a mixed cascade cycle. A mixed refrigerant enables the temperature profile of the cold and hot streams to be as close as possible. Closer temperature profile results in reduced compressor power and higher exergy efficiency.

1.6 Simulation specifications

Except for some energy balances and exergy calculations most of the results in this thesis are based on simulations in Aspentech HYSYS. The data for the NicheLNG process was provided by Höegh LNG and included in the HYSYS file 2DLE. This file is a model of the NicheLNG process after the LPG fractionation, where lean natural gas is cooled and ends as LNG. The 2DLE file is defined as the original case and will be referred to as 2DLE.

Efficiencies, ambient conditions, outlet aftercooler temperature and gas compositions are in 2DLE defined and used as basis in the other investigated cases. The compressor efficiencies are defined with vendor curves and the expanders have an adiabatic of 87%. In order to simplify the analysis was polytropic efficiency at 82% chosen instead of vendor curves. Cooling water is able to cool down the streams to 38°C and the streams undergo a pressure drop of 30 kPa through the aftercoolers. The feed gas to be liquefied has the composition expressed in Table 1.2. As Table 1.2 shows, the nitrogen content is close to 2,3 mole%. The desired content of LNG is below 1 mole% and a higher heating value <11,074 kWh/m^3 (<1070 BTU/scf). Due to the volatility of nitrogen its content can be reduced by production of flash gas. To be able to meet the LNG specifications, the natural gas leaving the cold box is cooled down to the same temperature in all investigated cases. One exception is when a liquid expander is introduced in Chapter 5. The nitrogen refrigeration cycles have a composition of 98 mole% nitrogen and 2 mole% oxygen. The FPSO-1 is supposed to have two identical liquefaction trains. In this thesis, the evaluations and analysis will be on a single train.

Methane	0,900409
Ethane	0,073198
Propane	0,003459
i-Butane	0,000008
n-Butane	0,000001
Nitrogen	0,022876
CO2	0,000049

Table 1.2 Natural gas mole% composition

2 Description of liquefaction technologies

Natural gas from a reservoir may have to undergo cleaning and scrubbing before liquefaction. This is necessary if the natural gas contains impurities and do not satisfy the product specifications of LNG. Heavy hydrocarbons, nitrogen, mercury, water and energy content are important specifications for the liquefaction process and the customer. Once the natural gas specification has fulfilled the requirements it is ready for the final liquefaction stage. The liquefaction process is based on the gas being cooled to its condensation temperature -162°C at atmospheric pressure. This temperature is defined to lie within cryogenic temperatures. By converting natural gas to liquid state, the gas volume is reduced to almost 1/600 [14]. This enables efficient storage and transportation.

Liquefaction processes in operation have a wide range of complexity. They differ in efficiency and size. By adding units or cycles, the efficiency may increase, but the size and weight will also increase. Onshore facilities do not have the strict constraints of low weight and small size so their efficiency is higher and the production rate too. They can have higher equipment count and an opportunity of large amount of hydrocarbon storage. These constraints are of importance when designing an offshore facility. Space is limited and the use of hydrocarbons should be inherently safe. It must also offer a high degree of modularity, low equipment count, quick start-up, available and be robust to vessel motion.

Liquefaction processes are either based on cascade, mixed or pure refrigerants cycles. The number of cycles differs from one to three and is of importance in the success of an efficient liquefaction. Proposed processes for offshore applications often involve one or two cycles.

Category	Technology	Cascade	C3MR, DMR	SMR	N2-Exp	Niche LNG
Suitability to LNG FPSO	Equipment count for Liquefaction	50-65	45-65	40-55	12	*11*
	Process sensitive to motion	Yes	Yes	Yes	No	*No*
	Ease of Start-up/operation	Low	Low	Low	High	*High*
	Flexible to feed gas changes	Medium	Low	Low	High	*High*
Safety Issues	Storage of HC Refrigerants	Yes	Yes	Yes	No	*No*
	Cryogenic Equipment count	High	High	Medium	Low	*Low*
	Space requirement	High	High	Medium	Low	*Low*
Efficiency	Thermal Efficiency (% of HHV)	91%	92%	89%	84%	*89%*
	Availability	Medium	Medium	Medium	High	*High*
	Specific Investment	High	High	Medium	Medium	*Low*

Table 2.1 Liquefaction processes suitability for FPSO [9]

Table 2.1 illustrates the most important selection criteria for a liquefaction process for natural gas. This is only a rough indication of the challenges each process face. This thesis will not cover cascade cycles and mixed refrigerant cycles with three stages, since these most likely are not suitable for offshore LNG production. Table 2.1 indicates expander processes (N2-Exp and NicheLNG) as the most suitable ones for an offshore environment.

Several liquefaction processes have been proposed for an FPSO. These processes range from one to two mixed refrigerant cycles or expansion cycles involving pure refrigerants. A typical single mixed refrigerant (SMR) process is the well known PRICO process from Black & Veatch. This refrigerant cycle has a composition of several gases and is carried out at different pressure levels.

The two-cycle C3MR is the dominant liquefaction process for natural gas. It involves a propane cycle as precooling and then a second cycle of mixed refrigerants. From this principle, Shell has developed the Shell Dual Mixed Refrigerant (DMR) liquefaction process. This process uses mixed refrigerants in both cycles and is proposed as a good alternative LNG process on a FPSO [15]. The DMR process has been selected for the Sakhalin Energy LNG project and is currently under construction. Developments of the DMR process have further improved process efficiency, from the C3MR process [8]. In addition and of importance for locations such as the Sakhalin, the DMR process is flexible to various operating conditions [16]. The site of the Sakhalin plant experience temperatures down to -35 °C in the winter and 20 °C in the summer [16].

Plant	Liquefaction Process	Status	Licensor	Efficiency kW*day/ton	Relative Power
OMAN LNG, Trains 1,2 (*1*)	**C3 Precooled MR**	**Operational**	**APCI**	**12.2**	**100%**
Wildwood LNG Plant (2)	Single N_2 expander, closed	Operational	CFS	40.5	332%
LNG Jamal BOG Reliquefier (3)	Single Expander, N_2	Built, des. Cap.	-	37.8	310%
Kryopak EXP-Typical (4)	Single expander, process fluid	Operational??	KryoPak, Inc.	20.4	167%
Predicted / Patented (5)	Dual Expander C1 / N_2	Simulated	ABB	16.5	135%
Predicted / Patented (5)	C3 - Dual Expander C1 / N_2	Simulated	ABB	**13.5**	**111%**

Table 2.2 Efficiency table of expander natural gas liquefiers [4]

The Oman LNG plant is based on propane precooling and mixed refrigerant. It is known as one of the most efficient liquefiers under operation [4], but the DMR at Sakhalin is expected to have even better efficiency. Table 2.2 compares the Oman LNG plant efficiency with different expander liquefiers. Due to its size and complexity it is not preferable for a FPSO but it gives a picture of what is feasible. For expander processes based on pure refrigerants, the number of cycles has a significant influence on the efficiency. It has to be noticed that the two dual expander processes have a methane cycle followed by a nitrogen cycle. Using methane or natural gas cycle instead of nitrogen may give some benefits on efficiency and will later be explained in Chapter 4.2.

Expander processes have been in focus for floating production with refrigerants in gas phase. They vary from single to dual cycles involving nitrogen and/or natural gas as refrigerant. Two promising processes have been developed. Höegh LNG has chosen the NicheLNG process from CB&I and several companies have proposed a nitrogen dual cycle process.

2.1 Niche LNG

The NicheLNG liquefaction process is based on a methane (natural gas) and nitrogen refrigerant cycle, with one open and one closed cycle respectively. They are independent expansion-compression cycles but they do overlap each other by heat exchange.

The process has the benefit of operating with high pressures, resulting in smaller pipes and valves than processes at close to ambient pressure. This gives an advantage related to space constraints on topside of a FPSO.

Neither of the refrigerants experience a phase change. Both remain in gaseous phase, so there are no problems with two phase flow distribution. A cycle remaining in gaseous phase reduces the risks of leakage and eliminates the need for liquid refrigerant storages, drums and separators.

Both advantages of high pressure and no liquid content in refrigerants reduce necessary space and the equipment count. The process is also more robust with respect to hull movements, due to refrigerants operating in gas phase. A non-flammable refrigerant, short start-up time, no venting or flaring of refrigerants after shutdown and smaller footprint increases the benefits of this process. The methane cycle is a flammable refrigerant and has to be included in the safety evaluation. Nevertheless, the proposed FPSO-1 already has large amount of liquid hydrocarbons as LPG so this refrigerant cycle should not have a sufficient involvement in the safety.

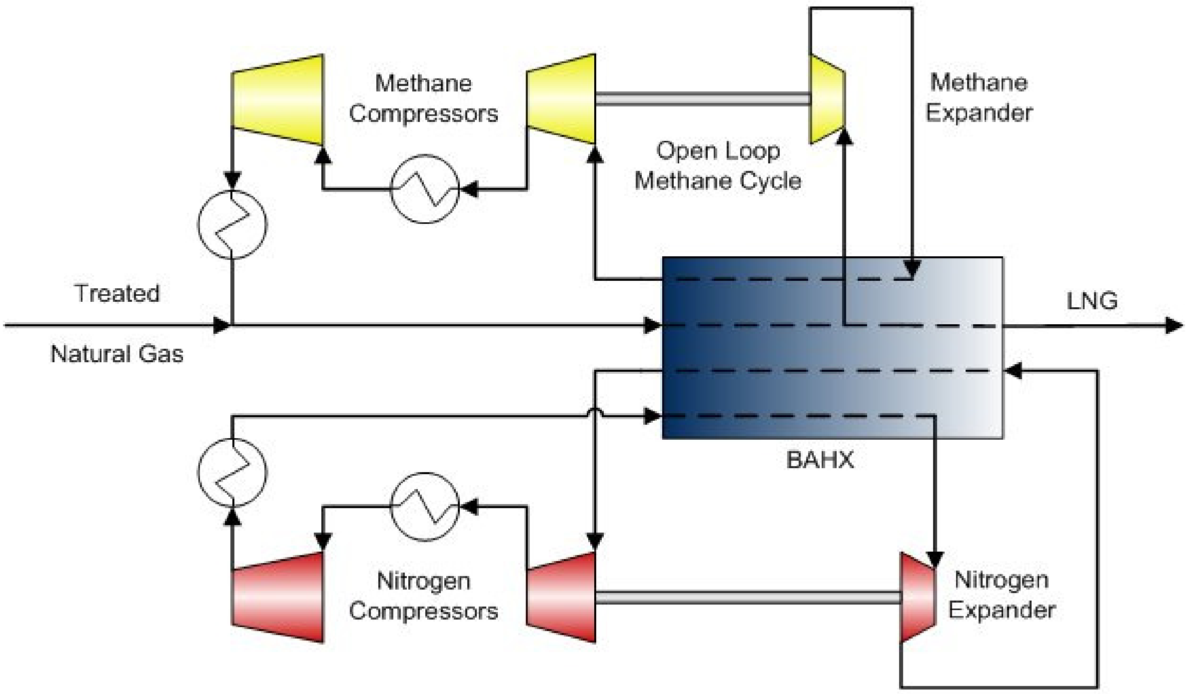


Figure 2.1 Flow sheet of the NicheLNG process [5]

Figure 2.1 is a basic principle of the NicheLNG process with two expander-driven compressors, two compressors and one heat exchanger. It illustrates how the natural gas and nitrogen are utilized as refrigerants. The open cycle is extracted natural gas to be expanded and then re-enters the heat exchanger. This cycle is known as a Claude cycle. In addition, a closed nitrogen cycle cools in the same temperature range but takes first care of the subcooling. The flash gas from boil-off is not shown in Figure 2.1.

The specific energy consumption is estimated to 16.5 kW/ton_{LNG}/day (0,396 kWh/kg_{LNG}) [4]. A relatively low energy consumption when compared with other processes based on pure refrigerants, as expressed in Table 2.2. The Dual Expander C1/N2 corresponds to the NicheLNG process.

2.2 PRICO - Single Mixed Refrigerant

This is one of the most basic processes based on mixed refrigerant and has a low equipment count. The setup is one heat exchanger network with a mixed refrigerant consisting of methane, ethane, propane, pentane and nitrogen. The composition is chosen based on the respective boiling points of the components to match the mixed refrigerant curve with the cooling curve of natural gas. Closer curves will increase the efficiency.

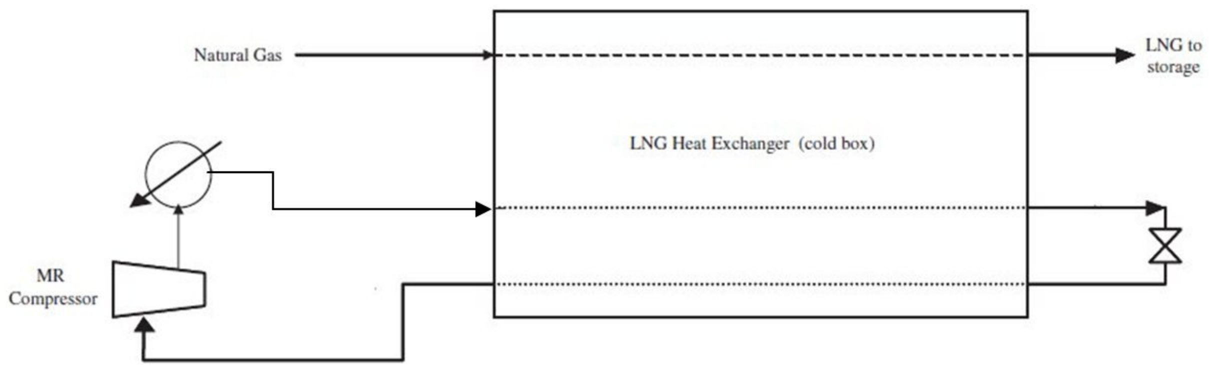


Figure 2.2 Basic principle of a single mixed refrigerant process [7]

2.2.1 Principle

A mixed refrigerant containing different gases is pressurized through a compressor. The discharge pressure has to be sufficiently high to give enough cooling duty after a later expansion. Mixed refrigerant flows through the heat exchanger as a hot stream. After leaving the heat exchanger in the cold end, the refrigerant undergoes a pressure decrease through an isenthalpic expansion valve. The reduction in pressure and temperature, by heat exchange and expansion, is necessary to achieve enough cooling duty. The stream contains now liquids. Then it reenters the heat exchanger as a cold stream. Heat transfer from the two hot streams, natural gas and mixed refrigerant, evaporates the liquid over a wide temperature range. The pressure is then recovered by compression.

Natural gas to be liquefied has initially a higher pressure level than the ambient condition. Not shown in figure 2.2 is a valve in the cold end. After heat is released in the heat exchanger an expansion brings the natural gas to the specifications required of LNG.

2.2.2 Extensions of PRICO

This process has a considerably large flow rate of refrigerant which leads to high compression work. On the other hand, the necessary pressure ratio is lower than for an expander liquefaction process as the NicheLNG. An improvement is compressing over two stages with inter-cooling reduces the consumption of work.

Also, the pressurized natural gas to be liquefied has a potential of work recovery. By utilizing the pressure to lower the temperature, through a turbine, work and cooling duty will be produced. This concept is the same as the open methane cycle in the NicheLNG process. The work recovered from the turbine can be utilized with a generator or a directly driven compressor.

2.3 Dual nitrogen refrigerant

Some proposed solutions for offshore LNG production have been with nitrogen as refrigerant. A nitrogen based liquefaction application has some advantages over the other compact LNG processes intended for offshore environment. It is easier to model, and the equipment is easier to operate, because the nitrogen refrigerant is always in gas phase. Due to the gas phase and the fact that nitrogen is an inert gas, the process is safer because of reduced hydrocarbon inventory compared to other processes.

The setup of the different nitrogen liquefaction processes share specifications as operating at high pressure levels and normally two refrigerant cycles. They do have some differences as the outlet pressure of the expanders. This will affect the size due to suction volume and complexity of the process. Illustrations and discussion of the possibilities will be covered later in this thesis.

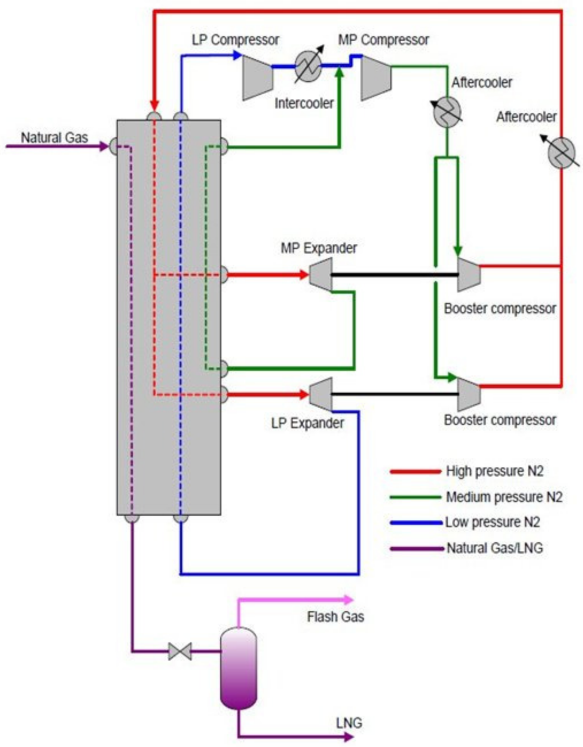


Figure 2.3a Statoil proposed solution

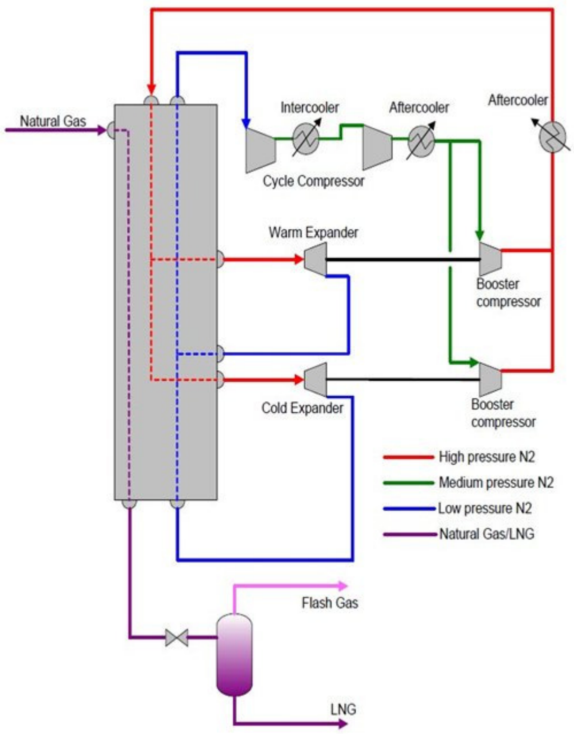


Figure 2.3b BHP Billiton proposed solution

Figure 2.3a illustrates the Statoil solution with pressure reduction to two levels. A large amount of the refrigerant stream flows through the middle pressure expander. This stream is supposed to cover the refrigeration of precooling and liquefaction. The low pressure stream covers the subcooling and the rest of the liquefaction and precooling. The flexibility of the process is rather small and is limited to the temperature splits between the cooling stages and the compressor pressure level.

In figure 2.3b the expanders have the same pressure reduction, and the two cold streams meets and flow as one cold stream through the heat exchanger from the subcooling liquefaction split. The most important factors from the two examples are the mc_p variations through the heat exchangers and the suction volumes of the compressors.

2.4 Comparing conventional with expander liquefaction processes

Focusing on efficiency there is no doubt that mixed refrigerant processes have a better efficiency than pure refrigerant expander processes [4]. Expander processes have lower efficiency but many benefits when production is in an offshore environment. The following sections compares selection criteria of different processes.

Compact:

Mixed refrigerant cycles require large storage capacity [4]. The large flow rate takes up area and increases the weight. Heat exchangers and equipments have to be able to operate with two phase flow.

Gaseous refrigerants, such as nitrogen, have the potential of being compact because there is no refrigerant storage and the refrigerants are operating with high pressures. Although the refrigerant flow rates are decreased, the required heat transfer area may not decrease because of the refrigerant heat transfer coefficient is also much lower. Non-flammable refrigerant will also reduce necessary area for safety.

Safety:

Operating with flammable refrigerants is well known from earlier LNG plants. Even though these plants have good safety records, operating on a ship with restricted area gives stricter safety concerns. Mixed refrigerant and cascade processes have large flammable refrigerant inventories, high circulation rate and flare requirements.

An expander process with nitrogen as refrigerant has higher inherent safety because nitrogen is an inert gas. As for the NicheLNG operating with natural gas as refrigerant, some stricter safety issues are introduced.

Operation:

Mixed refrigerant processes have a more complex operation due to refrigerant composition and high equipment count. It has also a longer start-up time and flare requirements.

Expander processes have an advantage in all three of the following process selection criteria: Ease of operation, quick start-up time and low equipment count.

Efficiency:

Fluids going through vaporization have to attract heat under almost constant pressure. This change of state characterizes a typical single composite refrigerant process. A mix of fluids with different boiling points flowing through a heat exchanger results in an evaporation of the cold stream. As Figure 1.8 shows, the natural gas curve does not have a linear profile. With the right composition of gases in the refrigerant, a gliding temperature profile is possible.

Expander processes operate in gas phase. The heat transfer cannot benefit from evaporation. For pure gases the specific heat is almost constant so a variation in the refrigeration flow rate is necessary to cover the non-linear temperature profile of natural gas.

3 Promising liquefaction processes for FPSO applications

In the proposed FPSO-1 after the pretreatment of the natural gas and after the LPG separation lean gas enters the liquefaction section. This section consists of two identical trains which have a total production rate of 4670 ton/day of LNG. The feed stream is divided into two equal mass flows entering the two trains.

3.1 The NicheLNG process

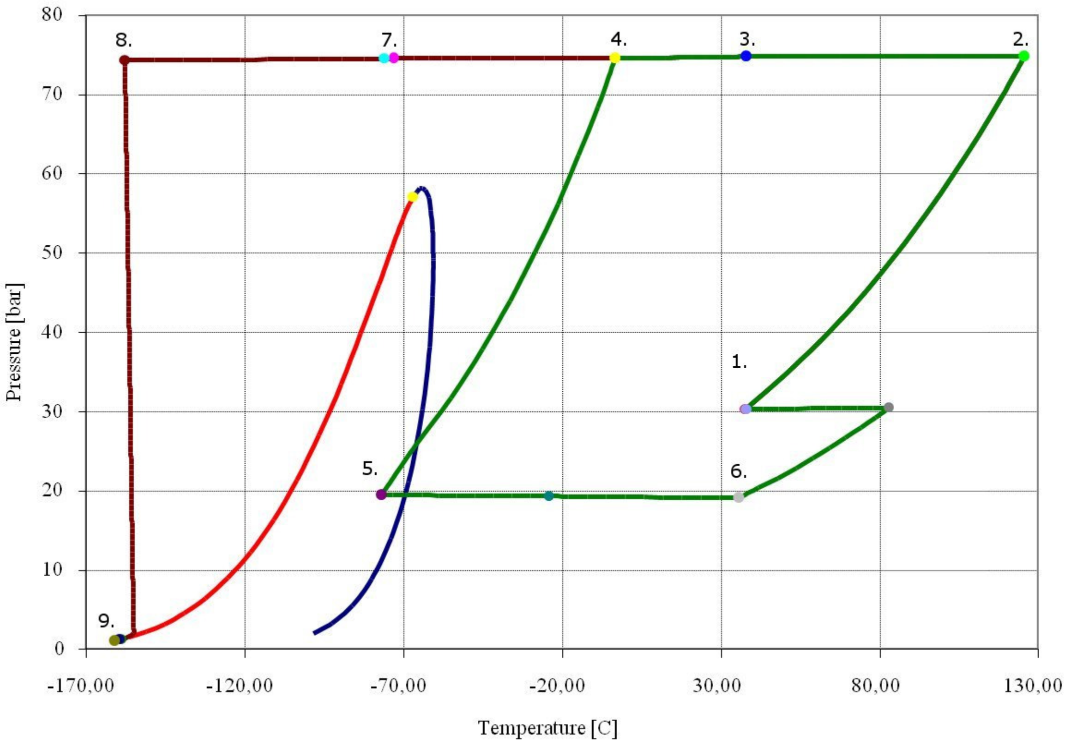


Figure 3.1 Pressure – temperature diagram illustrating the natural gas path [5]

As shown in the pressure-temperature diagram in Figure 3.1, recycled natural gas mixed with the feed gas is compressed to a pressure above cricondenbar (1-2). An aftercooler, utilizing water as refrigerant, will then lower the temperature (2-3). Then natural gas is entering the main LNG heat exchanger as a hot stream, and is cooled against cold low pressure natural gas and a nitrogen stream (3-4). Before further cooling, an amount of the hot natural gas stream is extracted from the heat exchanger and sent to an expander. The pressure of the extracted natural gas is reduced by a turbine (4-5). This extracted natural gas stream now acts as a cold stream. Together with the nitrogen stream, heat is now removed from the pressurized natural gas stream (5-6). Exiting the heat exchanger natural gas flows to a compressor which is mounted on the same shaft as the expander. Energy generated from the expander is utilized by compressing the natural gas. The discharge from the compressor is cooled by an after cooler (6-1).

The remaining natural gas to be liquefied, which has the same flow rate as the feed, is further cooled (4-8). Pressure is reduced across a valve which results in entering the two phase region and produces some flash gas and LNG (8-9).

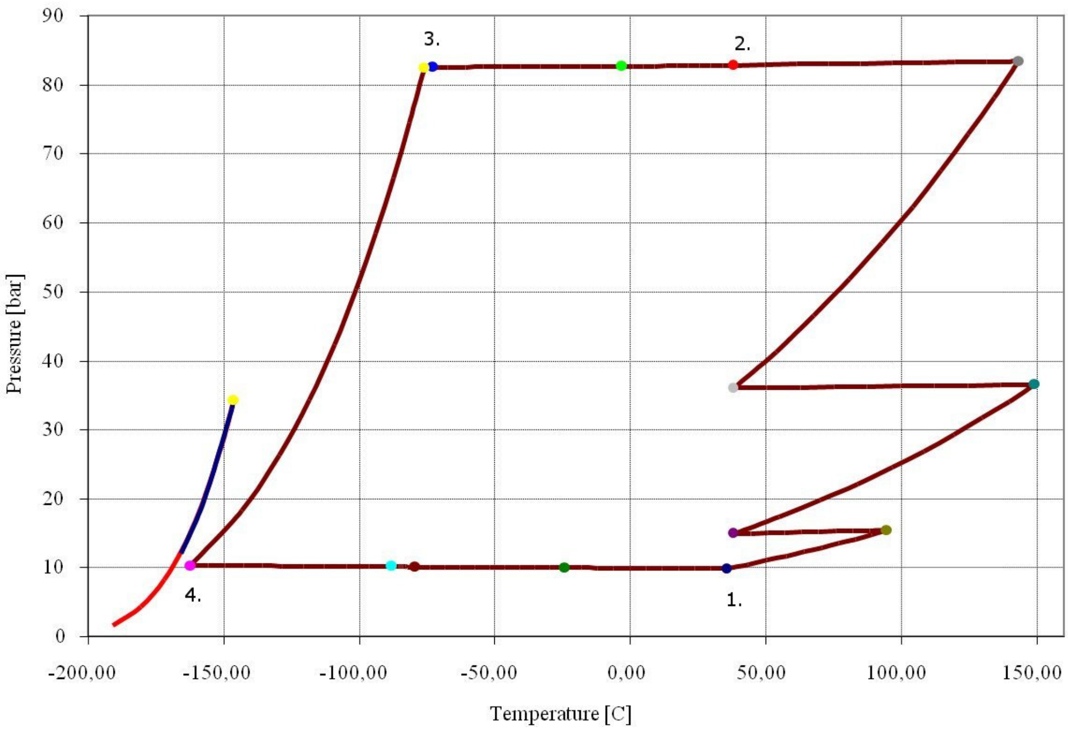


Figure 3.2 Pressure – temperature diagram illustrating the nitrogen path [5]

To bring the natural gas to the required temperature (state 8), nitrogen is used as a second refrigerant in a closed loop. The warm nitrogen stream leaving the heat exchanger is compressed followed by an aftercooler and then further compressed by an interstage compressor with an after cooler (1-2). It then flows as a warm stream through the heat exchanger and is cooled by the cold streams of natural gas and nitrogen (2-3). Then the high pressured nitrogen stream is expanded through a turbine and produces cooling and work (3-4). The expanded nitrogen provides a cooling potential at low temperature that is utilized in the heat exchanger (nitrogen is heated from state 4-1).

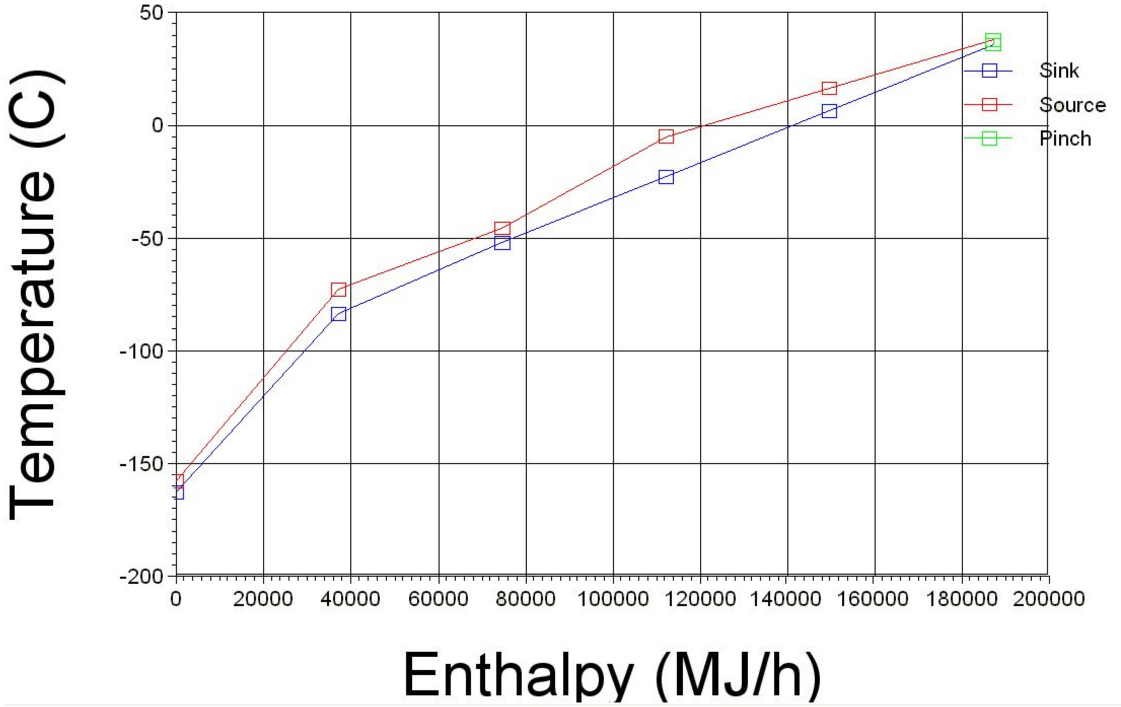


Figure 3.3 Temperature profile in the heat exchanger

The temperature-enthalpy diagram in Figure 3.3 shows the temperature difference in the heat exchanger. As illustrated in Figure 3.3, the pinch point is in the warm end. The cold composite curve, with temperature split at -85°C, has an almost linear profile in two intervals.

The natural gas composition in the 2DLE case is chosen after the pretreatment and the LPG separation to meet the LNG specifications. Some processes have an integrated LPG extraction. A partial or full integration depends on feed gas and onshore/offshore production. Since the focus is on the NicheLNG process the natural gas composition from the 2DLE case applies for all simulated cases in this thesis.

3.1.1 Exergy analysis of the NicheLNG process

One way of analyze the quality of the NicheLNG process is by determining the exergy efficiency. From the ratio of minimum work of liquefaction and the actual compressor work, the exergy efficiency can be expressed. The exergy calculations are done with enthalpy and entropy values obtained from simulations in HYSYS.

Since exergy is dependent on pressure, temperature and ambient conditions, the initial state will affect the efficiency. The feed gas alone has a relatively high exergy value. This affects the exergy efficiency and the specific power consumption. Hence, comparison of processes based on overall exergy efficiency and specific work will favor the one with highest inlet feed pressure.

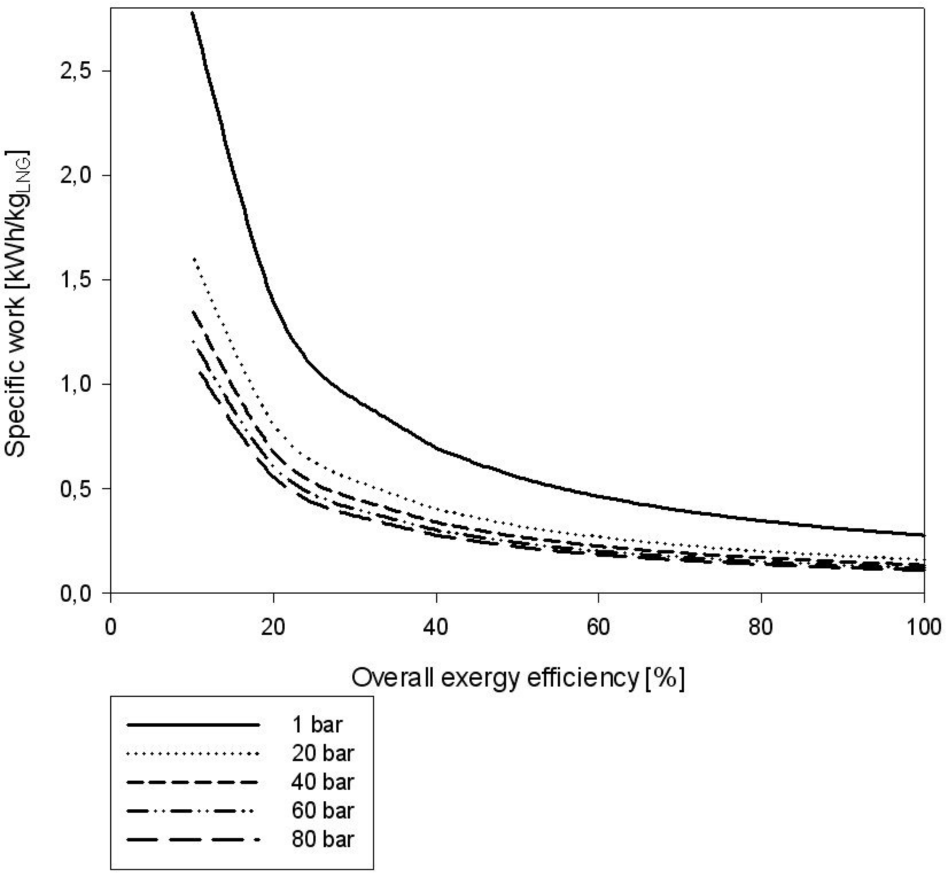


Figure 3.4 Specific work as function of overall exergy efficiency for a LNG process from gaseous feed at different pressures to saturated liquid at 1 bar

As Figure 3.4 illustrates, the minimum work (exergy efficiency of 100%) for liquefaction of natural gas at atmospheric pressure (1 bar) is 0,278 kWh/kg$_{LNG}$. The calculations are done with enthalpy and entropy values from a feed gas simulated in HYSYS, and are attached in Appendix A. Included in

Figure 3.4 is feed gas at higher pressures. The exergy content increases with higher pressure and reduces necessary specific work at a given overall exergy efficiency. So when efficiency of the NicheLNG process is to be compared with other liquefaction processes, the operating conditions have to be uniform. The efficiencies given in Table 2.2 are not specified with operating conditions so they may not be comparable. The efficiency of Dual Expander C1/N2 (NicheLNG) in Table 2.2, do not match with the efficiency 0,503 kWh/kg_{LNG}, simulated in HYSYS by the 2DLE file. An explanation of the variation can be the composition of feed gas, cooling water temperature, chosen equipment and the condition of feed gas after pretreatment. To be able to compare the NicheLNG process with another liquefaction process, a simulation of a dual nitrogen expander process has been done at equal conditions. The dual nitrogen expander process will be investigated later.

In the matter of comparing the NicheLNG process with another process, some design specifications have to be set in the evaluation. The main compressor use vendor curves in the HYSYS 2DLE file. So to achieve equal conditions for the two concepts to be investigated, all compressors were defined with polytropic efficiency of 82% and expanders with adiabatic efficiency of 87%. Every compressor has an aftercooler which lowers the temperature of the compressed gas to 38°C. The minimum temperature approach in the heat exchanger is specified to 3°C. Since phase change occurs in the heat exchanger, the 'Weighted model' was chosen as heat exchanger parameter for UA-value calculation. The background for the chosen model is attached in Appendix B. Exergy calculations are done with ambient conditions at 1 bar and 25°C.

Compressors	
Feed Comp.	23866,7 kW
NG Comp.	1313,5 kW
N2 Comp.	22654,4 kW
Total	47834,7 kW
Min. power for liquefaction	53658122 kJ/h 14905,0 kW
Exergy efficincy	31,2 %
Specific work	0,490 kWh/kg_LNG

Table 3.1 Efficiency and work consumption for the NicheLNG process

From the specifications above the NicheLNG process is simulated in HYSYS. The transition from compressor vendor curves resulted in a lower outlet pressure of the expander driven compressor. To obtain feed gas pressure of the open methane cycle was an additional compressor (NG Comp.) installed to increase the pressure to the same pressure as the feed gas.

The minimum power for liquefaction was calculated with values from feed gas conditions to after the heat exchanger, as discussed in Section 1.3. Feed gas composition from Table 1.2 has been used for all cases.

With feed pressure at 30 bar, the exergy efficiency was calculated from Eq. 1.11 to be 31,2%. As can be seen from Figure 3.4 the calculated exergy efficiency matches with the calculated specific work in Table 3.1.

Feed gas pressure [bar]	Min. liq. work [kWh/kg LNG]
1	0,278
20	0,161
40	0,136
60	0,121
75	0,113
80	0,111

Table 3.2 Minimum liquefaction work to feed gas pressure

A way of characterizing the quality of a liquefaction process is by comparing it with the theoretical minimum liquefaction work. Table 3.2 expresses necessary minimum work for given pressures. The calculations are done with data from HYSYS. The calculations resulting in Table 3.2

are found in Appendix A and the composition of the natural gas is found in Table 1.2. Table 3.2 illustrates how feed gas pressure affects the efficiency and will be further investigated in Chapter 4.

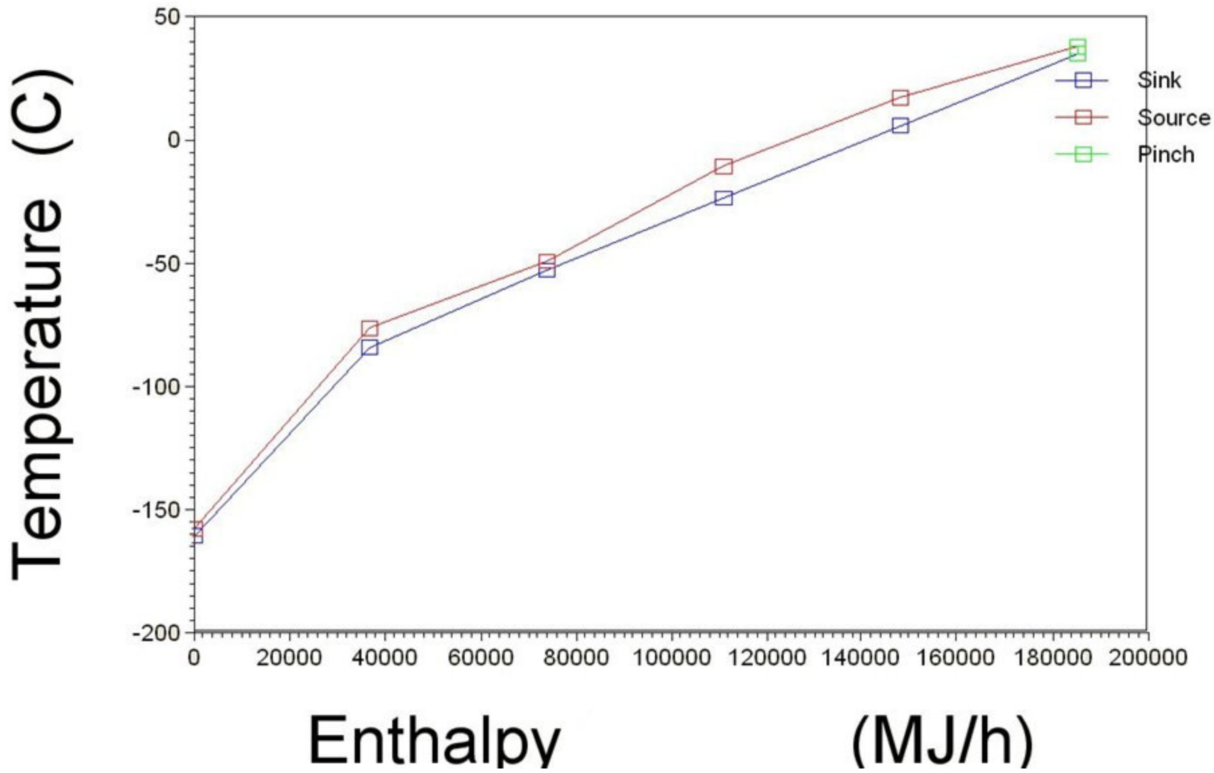


Figure 3.5 The composite curves for the NicheLNG process

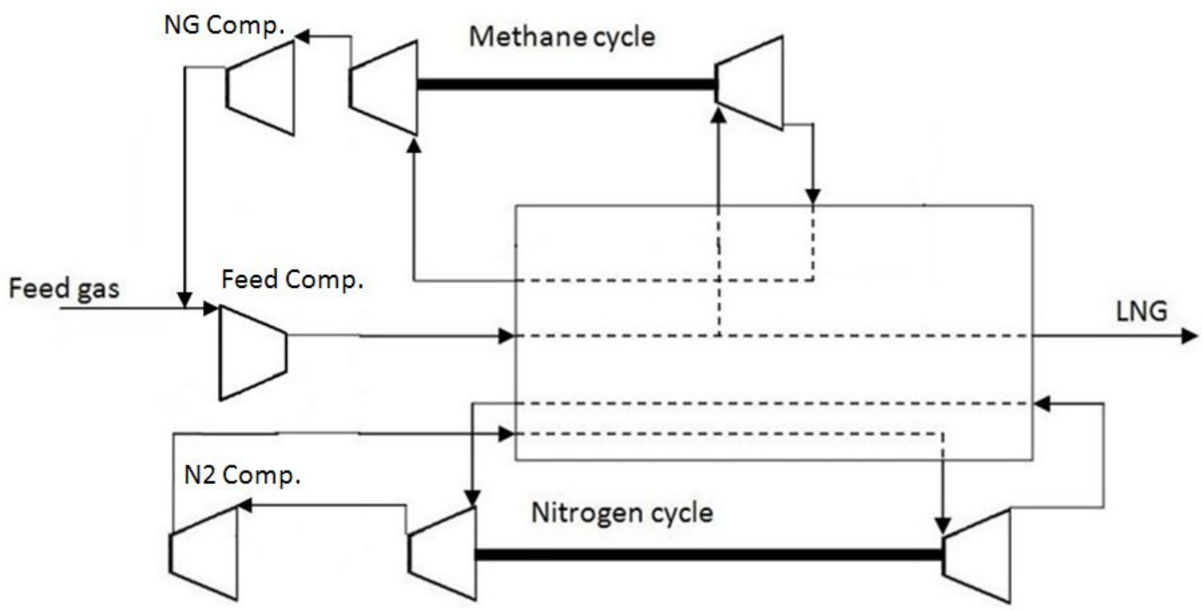


Figure 3.6 Flow sheet of the simulated NicheLNG process

The N2 Comp. in Figure 3.6 represents two compressors with interstage cooling.

3.2 Dual nitrogen process

The most frequently proposed process for offshore liquefaction of natural gas is with two nitrogen refrigerant cycles. One cycle covers the precooling and the second is used for liquefaction and subcooling. Nitrogen as refrigerant medium is very flexible. The relative low dew point gives the opportunity of the refrigerant cycles to operate at a wide temperature range. Chapter 2.3 describes different concepts. The choice of concept in this work was with emphasis on equipment count that matches the NicheLNG process.

3.2.1 Exergy analysis of a dual nitrogen process

To compare a dual nitrogen process with the NicheLNG process equal conditions are necessary. The same efficiencies for compressors and expanders are being used. Feed gas and the produced LNG have the compositions from Table 1.2 and the same conditions as in the exergy analysis for the NicheLNG process. The natural gas in the open methane cycle has been replaced with 98 mole% N_2 and 2 mole% O_2 and configured to a closed cycle. With these specifications, flows and pressures in the two nitrogen cycles have been optimized to minimize work consumption. Figure 3.8 illustrates the design of the dual nitrogen process.

Table 3.3 expresses the quality of the simulated dual nitrogen process. When the natural gas leaves the cold box, a pressure reduction through a valve will result in flash gas formed by evaporation. The evaporation is necessary in order to reduce the nitrogen content to 1 mole%. In both processes, the natural gas leaving the cold box has the same temperature at -157,2 °C. With temperatures at the same level the flash gas production will be at the same rate and thereby a similar LNG composition. Both processes produce LNG with a higher heating value of 10,95 kWh/m^3 (1058 BTU/scf).

Compressors	
Feed Comp. N2 Comp.-1 N2 Comp.-2 Total	5418,2 kW 24333,3 kW 23503,6 kW 53255,1 kW
Min. power for liquefaction	53658122 kJ/h 14905,0 kW
Exergy efficincy	28,0 %
Specific work	0,545 kWh/kg_LNG

Table 3.3 Efficiency and work consumption for the dual nitrogen process

It was decided to operate the refrigerant with only one high pressure and one low pressure level, respectively at 90 bar and at 17 bar. Then only adjustment of the flow rate was necessary to provide the required cooling.

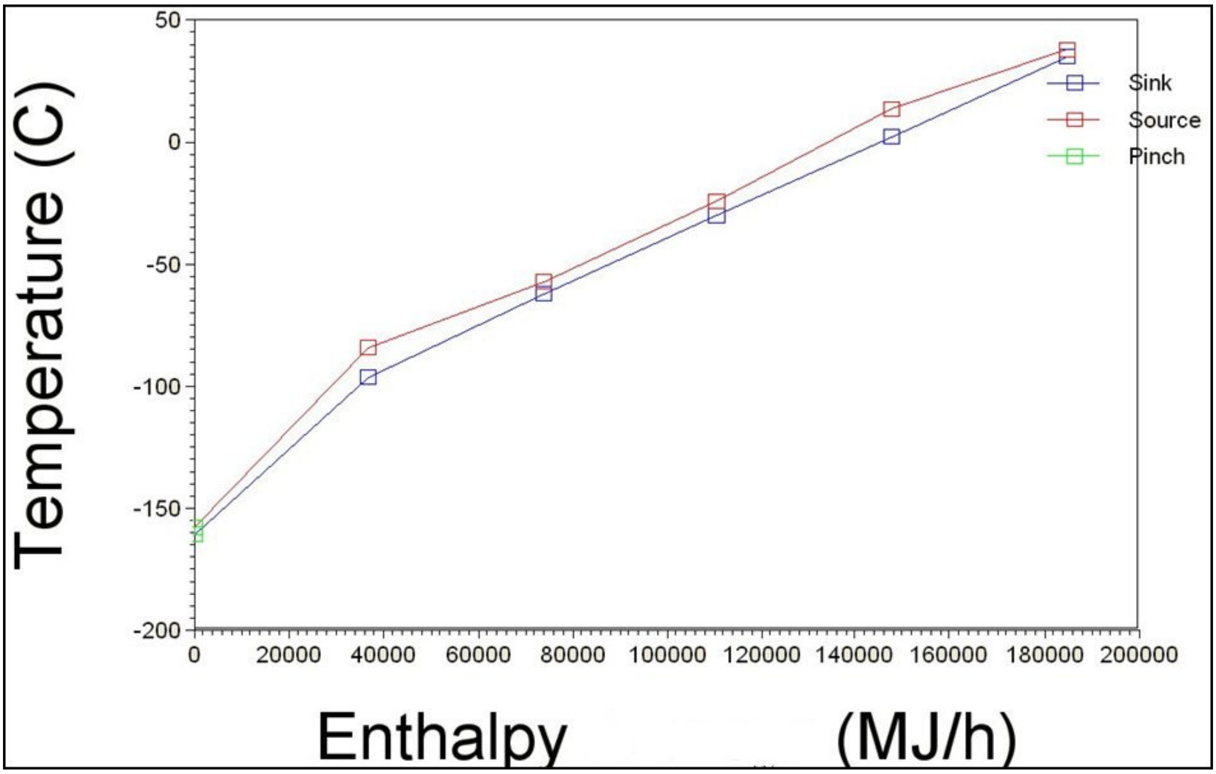


Figure 3.7 The composite curves for the dual nitrogen process

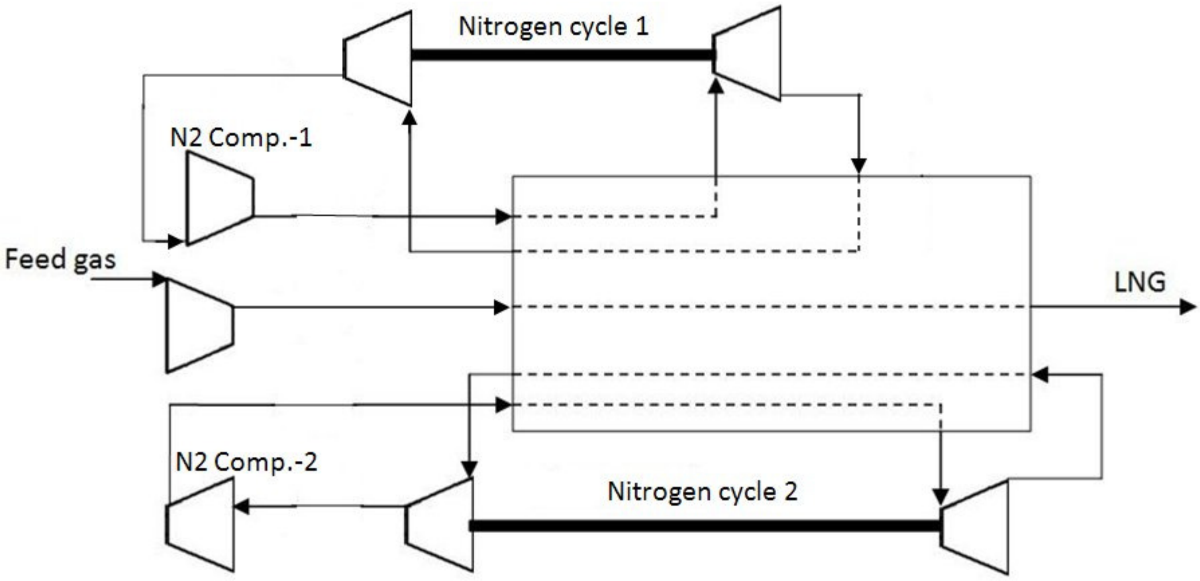


Figure 3.8 Flow sheet of the simulated dual nitrogen process

The N2 Comp-1 in figure 3.8 represents two compressors with interstage cooling.

3.3 Discussion of NicheLNG versus dual N_2 process

Both processes are almost similar in design but are distinguished by the different refrigerant. They have the same amount of compressors and expanders. Equal equipment count is one criterion to achieve a fair comparison. In addition to the two expander driven compressors had both processes four compressors.

The simulated NicheLNG process in Section 3.1.1 distinguished from the original 2DLE by adding one compressor in the open methane cycle. This had to be done after efficiency adjustment of compressors and heat exchanger. It looks like that 2DLE low pressure of the open methane cycle was defined by the feed gas pressure or vice versa. A design with an open methane cycle is influenced by the feed gas pressure. If the feed gas pressure changes it will affect the cooling duty of the open methane cycle, since feed gas and refrigeration gas are pressurized by the same compressor. By closing the loop the NicheLNG process will be more available to different feed pressures. It may also be an advantage to have a possibility of higher pressures in the methane cycle.

The NicheLNG process has about 10 % lower work consumption. This can be explained by the chosen refrigerant and the heat distribution in the heat exchanger. Natural gas as a refrigerant has higher c_p than nitrogen, resulting in a significantly smaller refrigerant mass flow rate. Higher mass flow rate causes more irreversibilities in the compressors and expanders, so the lower mass flow rate explains why NicheLNG is the most energy efficient process.

		NicheLNG	Dual Nitrogen
Specific work	[kWh/kg_LNG]	0,49	0,545
Compressor power	[kW]	47 835	53 255
Refrigerant flow rate	[kg/s]	105+112*	298
Refrigerant high pressure	[bar]	75/83**	90
Refrigerant low pressure	[bar]	17,5/12,5**	17
Max. ΔT in HX	[°C]	12,6	12,1
HX duty	[kJ/s]	51 429	51 308
UA value	[kJ/°C*h]	29 014 561	27 924 137

*105 kg/s NG + 112 kg/s N2

** 75 and 83 bar for NG and N2 cycle respectively

Table 3.4 Results of both processes

Figure 3.5 and 3.7 illustrates that the processes do not differ much in temperature difference and they have almost the same UA value. A larger temperature difference produces more entropy. With closer composite curves (smaller ΔT), the process efficiency will be improved, however, at the expense of a larger heat exchanger.

The NicheLNG uses a flammable refrigerant, so it has a disadvantage from a safety point of view. An evaluation of this issue must be considered with respect to the safety requirements.

4 Adjustments and analysis of NicheLNG

4.1 Precooling

To keep refrigerant cycles at relatively small sizes, even at high LNG production rate, the number of refrigerant cycles can be extended. A precooling helps the more energy demanding liquefaction and subcooling by reduced flow rate of refrigerants. In addition, the specific heat of natural gas varies with temperature so refrigerants with different mixtures can benefit from closer temperature difference in the heat exchangers.

The most common precooling refrigerants are propane or ethane/propane mixtures. These are flammable hydrocarbons and have to be included in the evaluation of the process safety. From Table 2.2 a precooling integration can provide an 18% higher efficiency to the NicheLNG process. Another possibility is CO_2 as precooling refrigerant. CO_2 is not flammable and may be preferable in offshore liquefaction processes. A disadvantage is the risk of forming solids.

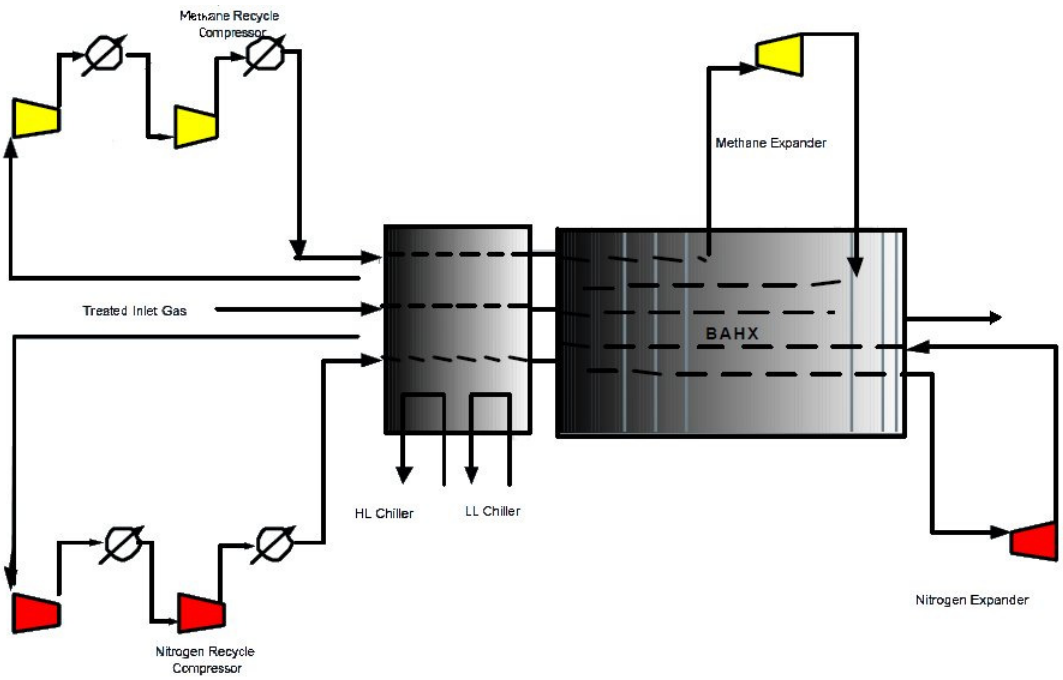


Figure 4.1 The NicheLNG process with a precooler in front [17]

Figure 4.1 is an illustration of a precooler installed for a single train. The FPSO-1 is planned to have two liquefaction trains, so a larger precooler to cover both will keep equipment count down.

4.2 Refrigerant medium

All of the expander processes described in Chapter 3 operates with two refrigerant cycles based on nitrogen or natural gas and nitrogen. They operate in gaseous phase through the cycles and have therefore some constraints. An expander process has to take into the account the boiling point of the gas, due to problems with liquids in a turbine. Hence, a natural gas refrigerant cycle has a stricter constraint on operability than nitrogen. Since the boiling point of nitrogen is -196 °C at atmospheric pressure, entering the two phase region is at a lower risk than operating with natural gas. The relative low boiling point results in a higher degree of freedom than for a refrigerant of natural gas.

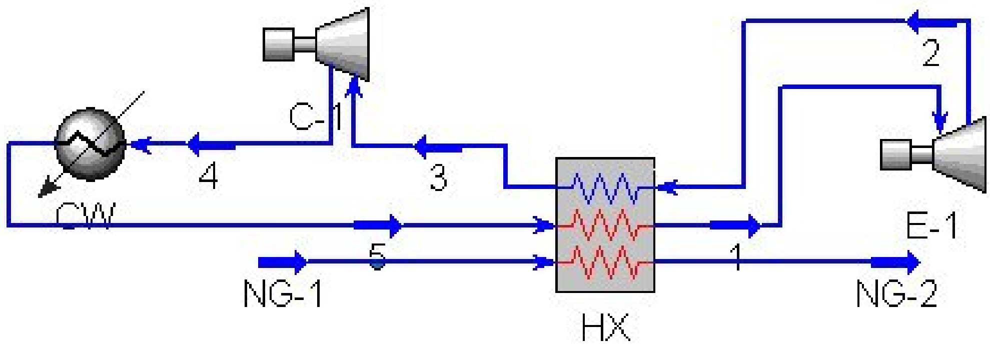


Figure 4.2 Simple illustration of expander precooling

Figure 4.2 is an example of a closed precooling cycle. It contains compressor, expander, cooler and heat exchanger. Simulation is done with an adiabatic efficiency of 80% in the compressor and expander and no pressure loss the through aftercooler and heat exchanger. The heat exchanger has a minimum temperature approach of 3°C.

The calculations shown in Table 4.1 are done with a constant pressure ratio of refrigerant and with a desired outlet temperature of the natural gas. Mass flow is adjusted to cool down the natural gas to -65°C. The chosen refrigerants are pure methane and nitrogen.

This example shows the importance of the refrigeration gas ability to extract heat. Methane as a refrigerant has a higher specific heat capacity, hence lower necessary mass flow resulting in higher efficiency. A nitrogen refrigeration cycle has, with the specifications from Table 4.1, over twice the energy consumption as a methane refrigeration cycle.

Natural gas					
	NG-1	NG-2			
Temperature [C]	26,85	-65			
Mass Flow [kg/h]	65,79	65,79			
Pressure [bar]	65	65			
Refrigerant cycle					
Methane	1	2	3	4	5
Temperature [C]	-22,59	-105,2	23,85	171	26,85
Mass Flow [kg/h]	160,2	160,2	160,2	160,2	160,2
Pressure [bar]	100	21,9	21,9	100	100
Nitrogen	1	2	3	4	5
Temperature [C]	-95,87	-155,9	23,85	222,8	26,85
Mass Flow [kg/h]	558,1	558,1	558,1	558,1	558,1
Pressure [bar]	100	21,9	21,9	100	100
Specific Work					
Methane [kWh/kg]	0,172				
Nitrogen [kWh/kg]	0,424				

Table 4.1 Specifications and results of a precooling example

Ethane has an even higher specific heat capacity. So for the NicheLNG process, operating with an open refrigeration cycle of natural gas, small amounts of heavier hydrocarbons such as ethane will give a small increase in efficiency.

4.2.1 Change in gas characteristics for pressure variations

The composition of a stream flowing through a heat exchanger affects the size. The amount of heat a fluid is able to hold depends on its specific heat capacity and flow rate. If a fluid is supposed to attract a given heat duty at a given temperature range the flow rate will be dependent on the specific heat capacity. From this a higher specific heat capacity will result in a lower necessary flow rate, and thereby influence the heat exchanger size and the compressor work.

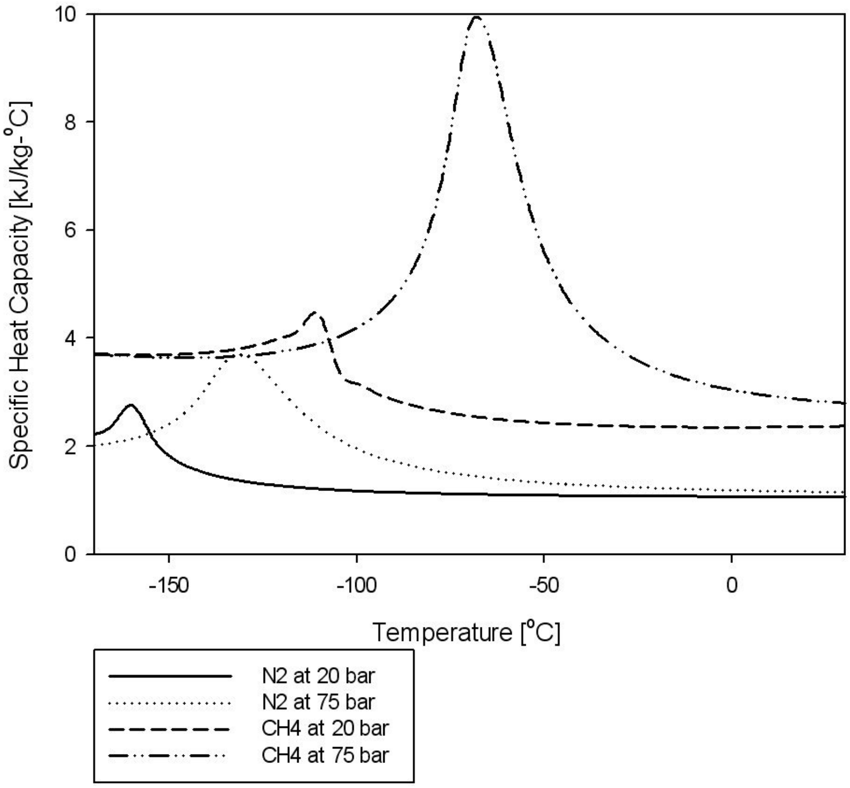
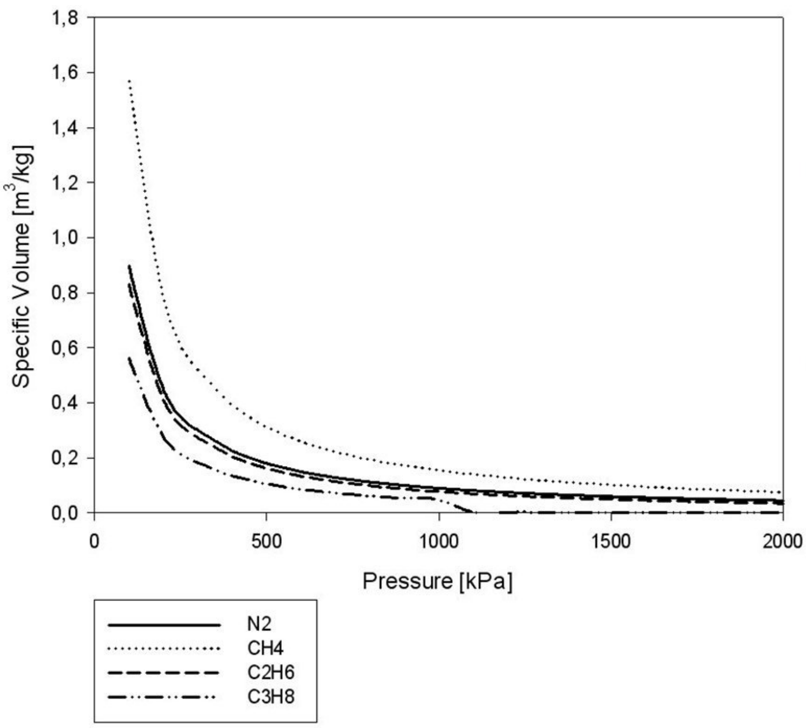


Figure 4.3 Specific heat capacities of N2 and CH4 at pressure levels from the NicheLNG process

Figure 4.3 illustrates the variation of specific heat capacity at constant pressure levels between two gases in a certain temperature range. The most important information from Figure 4.3 is the large c_p of methane. Even though methane is able to attract more heat than nitrogen is its range of operation in gaseous phase restricted to a smaller temperature region. The peaks for each curve in figure 4.2 are the critical points of the respective gases.

It is easy to conclude that CH_4 is able to hold more heat than N_2 at the same flow rate. Hence, it is believed that the choice of refrigerant will affect the heat exchanger size. It has to be noticed that this is not fully true since the size of heat exchangers are also dependent on the heat transfer coefficient. At a fixed conduction through the wall of a heat exchanger the heat transfer coefficient is relative to convection. Speed will vary with the chosen gas to satisfy necessary heat transfer, and thereby influence the turbulence with change in speed and viscosity.

The overall specific efficiency of a liquefaction process is the ratio between the work consumed and the LNG produced. To get necessary cooling duty, a gas has to be expanded and later re-compressed. The compression work of a single compressor is depending on inlet temperature, pressure ratio and specific volume of the gas. Compression in liquefaction processes is done at close to ambient temperature. The pressure ratio depends on desired cooling duty which is also influenced by the chosen refrigerant gas.



Gas	Molecular Weight
N2	28,01
CH4	16,04
C2H6	30,07
C3H8	44,10

Figure 4.4 Specific volume variations with pressure at a temperature of 30 °C

Figure 4.4 illustrates how the specific volume varies with pressure for different gases. The sudden decrease of the propane (C_3H_8) above 1000 kPa is due to condensation.

The molecular weight of gases influences its specific volume. For natural gas containing mostly methane a small amount of heavier hydrocarbons will give a reduction in specific volume when comparing with pure methane. Nevertheless, it will not decrease below the specific volume of nitrogen.

Maybe the most important observation from Figure 4.4 is the specific volume value at higher pressures. The specific volume difference between certain gases reduces with higher pressure. So operating at high pressure levels will not give significant difference in the specific volume. The specific heat capacity will be more effective. A high specific heat capacity demands lower flow rate for a given heat exchange duty.

4.3 Placement of expansion

As explained earlier an expansion instead of a throttling is favorable. In sub-ambient processes can a pressurized stream provide cooling and produce work. A typical expander liquefaction process is normally based on two refrigeration cycles. Each cycle is supposed to cover a given temperature region.

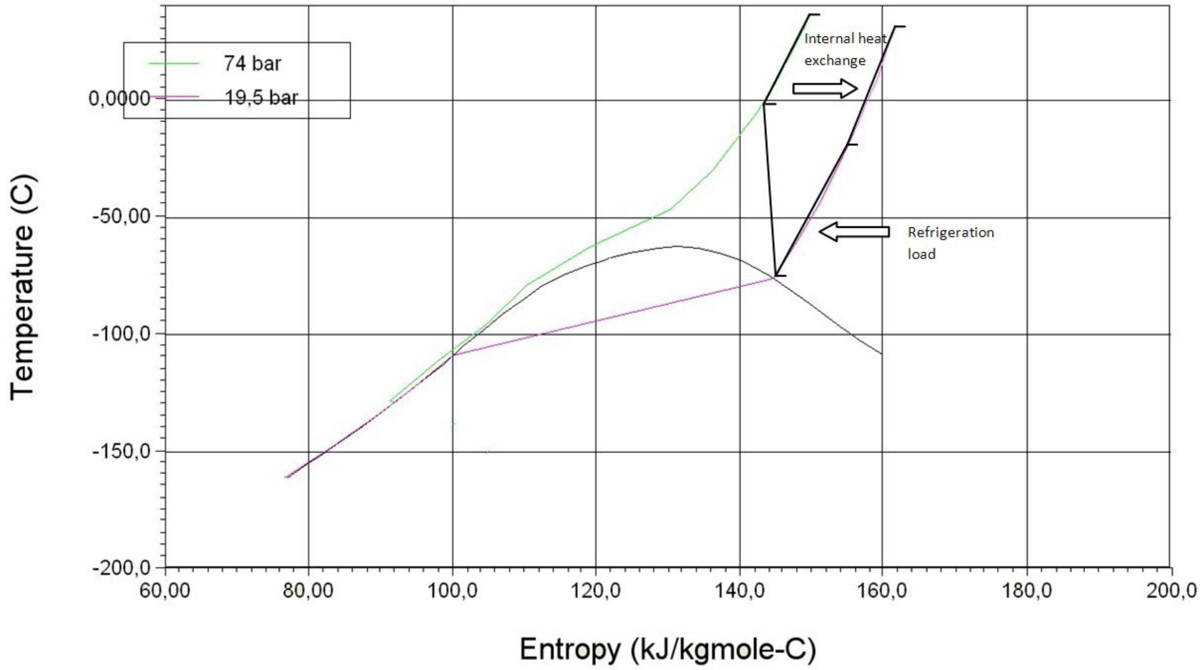


Figure 4.5 The NicheLNG open methane refrigeration cycle with refrigeration regions indicated

Figure 4.5 illustrates the open methane cycle in the NicheLNG liquefaction process. It has the same principle as a typical Brayton cycle. The placement of the expander is at -1,5°C after an internal heat exchange. After the expansion to 19,5 bar the temperature is reduced to -76,8°C which results in a refrigeration load of 2568 kJ/kgmole and a necessary compressor work of 2845 kJ/kgmole (17084 kW). The calculations are done with the natural gas refrigerant flow rate from the 2DLE case at 21618,85 kgmole/h.

4.3.1 Expander placement for the open methane cycle

A way of expressing the efficiency of the open methane cycle is by the ratio of compression work and refrigeration load. By comparing the original design with other hypothetical inlet temperatures the quality can be found. It has to be noticed that the results from Figure 4.6 is restricted to the open natural gas cycle in the NicheLNG process. How the overall efficiency varies with different inlet temperature of the open cycle expander is not included.

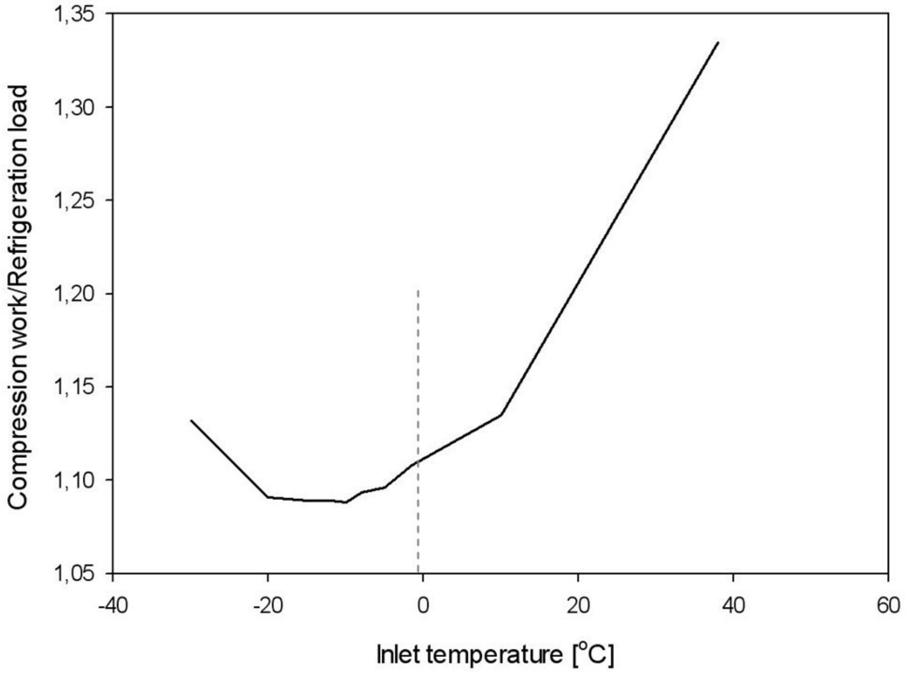


Figure 4.6 Efficiency of the open natural gas cycle with different inlet temperatures

Figure 4.6 illustrates how the efficiency of the open methane cycle varies with inlet temperature from the original indicated with a dotted line. The calculations are attached in Appendix C. All calculations are done with an ideal heat exchanger, turbine with 87% adiabatic efficiency and compressors with 82% polytropic efficiency. The ΔT in the warm end is set to 2°C.

Inlet temperature [°C]	-1,5	-10
Internal heat exchange [kJ/kgmole]	1949,0	2413,0
Refrigeration load [kJ/kgmole]	2568,0	2094,4
Cooling duty [kJ/kgmole]	4517,0	4507,4
Compressor work [kW]	17084,4	13688,1
Work/ refrigeration load [-]	1,108	1,088

Table 4.2 Results of inlet temperature as original at -1,5°C and at -10°C

From Figure 4.6 and Table 4.2 less work per refrigeration load will be the result with cooling down to -10°C before an expansion. As earlier mentioned, these calculations are restricted to the open natural gas cycle. More internal heat exchange, with inlet expansion temperature at -10°C, will give less refrigeration duty to the overall process.

4.3.2 Consequences of expander placement

Inlet expansion of the open methane cycle at -10°C results in almost the same cooling duty (internal heat exchange and refrigeration duty) as with an inlet expansion temperature at -1,5°C. Hence, less work is needed for the same cooling duty. Even though the total cooling produced does not change, the natural gas has been further cooled by internal heat exchange and this will affect the refrigeration load. The reduced refrigeration load is 473,6 kJ/kgmole or in heat flow $10{,}239 \cdot 10^6$ kJ/h. This reduced refrigeration load has to be covered by the closed nitrogen cycle or by higher compressor work in the open methane cycle.

4.4 Liquefaction pressure

Natural gas to be liquefied is always under high pressure. The advantage of liquefying pressurized natural gas can easily be seen from a pressure - enthalpy diagram. Natural gas at constant temperature has lower enthalpy with increasing pressure. Hence, the amount of heat to be removed is reduced with increased pressure at a constant temperature. After subcooling, the natural gas is still pressurized. An expansion, in addition to recovering work and contribute to the cooling, will bring the natural gas to the given specifications.

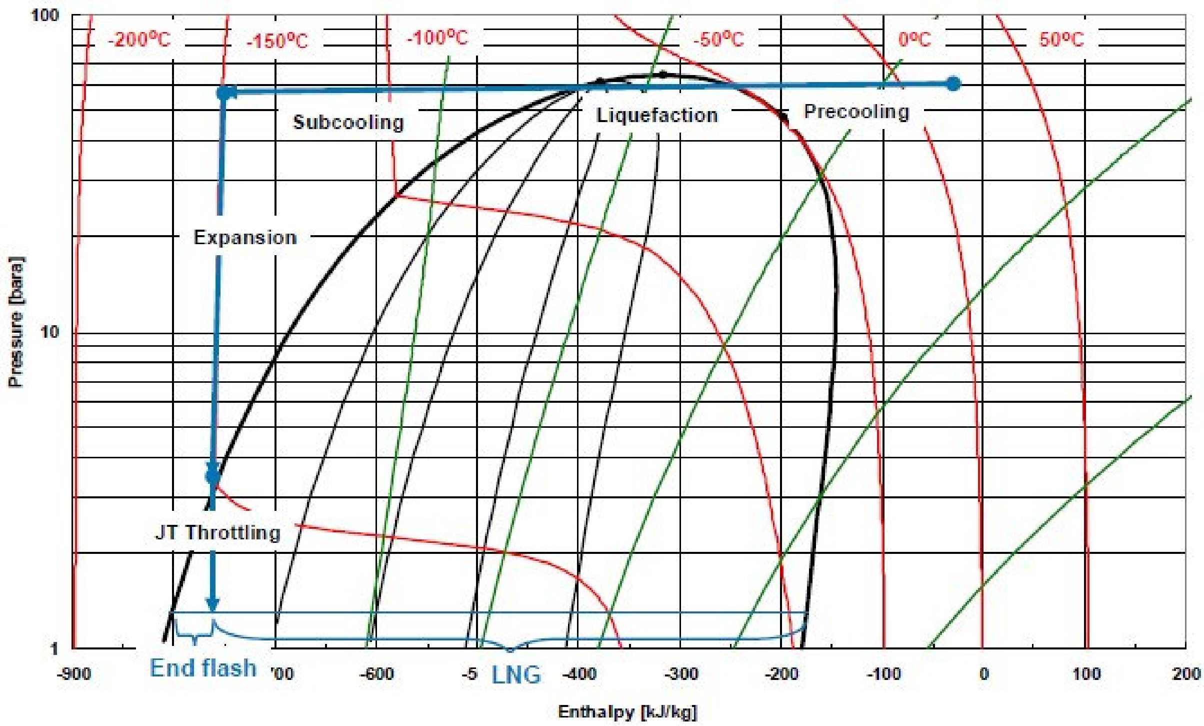


Figure 4.7 Natural gas path through liquefaction for a typically onshore facility [11]

Figure 4.7 illustrates the natural gas path through liquefaction for a typical onshore based LNG production. The composition is C1 89,7%, C2 5,5%, C3 1,8% and N2 2,8% [11]. The natural gas enters with a pressure of 60 bar. As figure 4.7 illustrates is the cooling done below the critical point. By entering the two phase region some liquid can be extracted with a phase separator before further cooling. The NicheLNG process avoids the utilization of a phase separator by higher liquefaction pressure. How the liquefaction pressure affects the liquefaction will be further investigated in the following sections.

4.4.1 Relationship between feed gas and liquefaction pressure

When production of LNG is done offshore some constraints will affect the process. By excluding a phase separator, the pressure can be increased above the critical point. Without a phase separator, the natural gas entering has the same composition as the one leaving the cold box. A lean composition is therefore required to meet the LNG specifications. This will affect the comparison of efficiency between different LNG processes. An integration of a phase separator in the liquefaction section will affect the power consumption due to a different liquefaction pressure. For the NicheLNG process, all LPG fractionation is done in front of the liquefaction. For the processes in Table 2.2 a detailed description on the utilization of phase separator in the liquefaction part could not be found. So a direct comparison of a traditional onshore process as Oman LNG cannot be made without more data.

As cooling water is relatively cheap there are several advantages by using this as a cooling medium. When it comes to saving work, two areas in particular are outstanding. Multiple compressions with interstage cooling in order to achieve close to isothermal compression will save some energy. Second the cooling water temperature is important for the amount of necessary heat to be removed. Figure 4.8 illustrates an ideal liquefaction process for natural gas. The ideal work is shown in the *W* area with heat rejection at constant temperature. The T-s diagram in figure 4.8 shows how the influence of the heat rejection temperature has to work consumption.

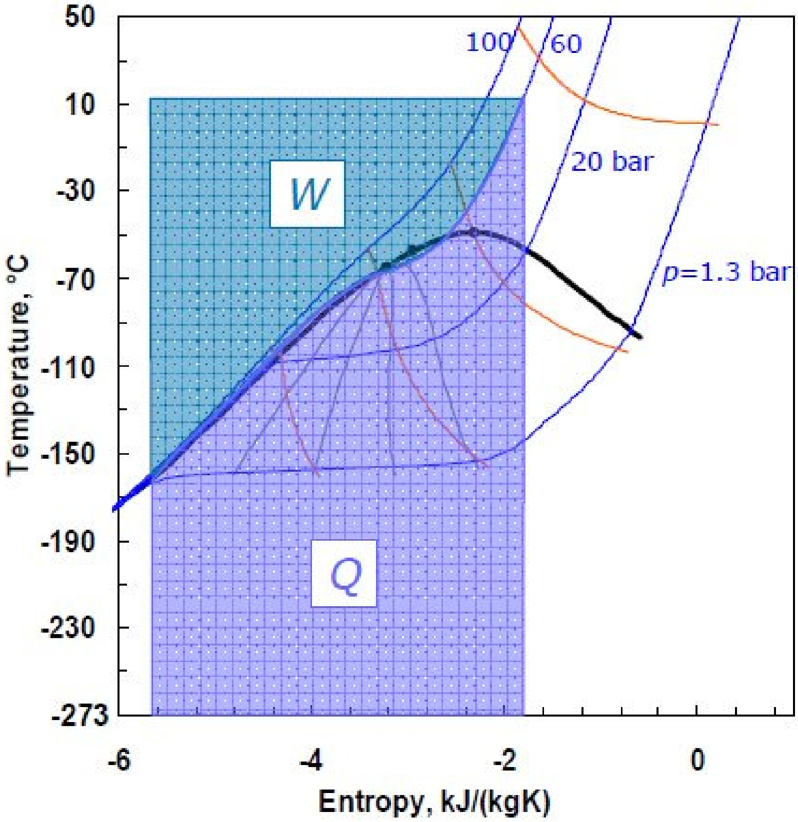


Figure 4.8 Ideal liquefaction process of natural gas [11]

Heat is removed as the gas is cooled at gliding temperature and constant pressure. The isobar lines show two important factors with increased pressure. As pressure increases the gliding temperature gets more linear. A more linear cooling of the natural gas is an advantage when the number of refrigeration cycles is restricted to one or two. The more interesting factor is the changing properties of the feed gas with increased pressure. Figure 4.7 illustrates the enthalpy reduction with increased pressure at a constant heat rejection temperature. A lower enthalpy of the natural gas entering the cold box results in less heat to be removed, and thus lower necessary work consumption for the refrigeration cycles. Even though compression of the feed gas consumes more work with higher pressure, it may be favorable due to the low refrigeration efficiency for liquefaction processes.

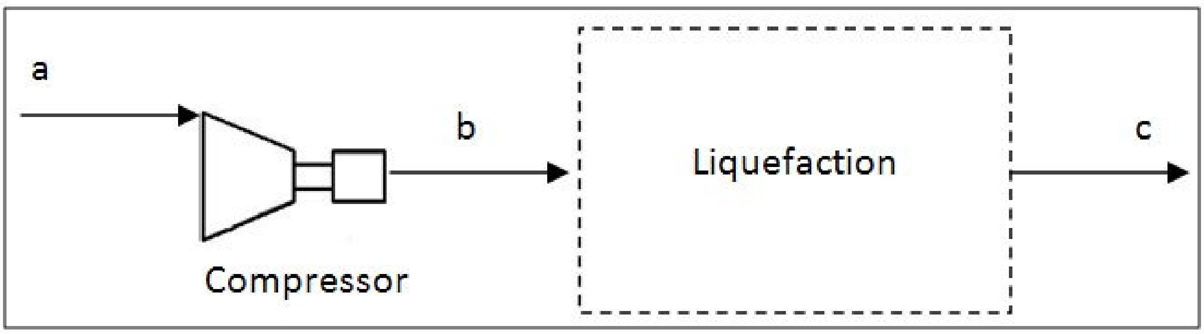


Figure 4.9 Simple flow sheet of the liquefaction path from feed gas to LNG

Figure 4.9 illustrates the liquefaction path from feed gas (a) to LNG (c). A compression of the feed gas (a-b) will change its properties which will influence the liquefaction. In order to illustrate how the feed gas pressure affects the liquefaction some adjustments to the 2DLE process have been made. The original process is the 2DLE with one open methane refrigeration cycle. Closing the loop and using a feed gas compressor will simplify the analysis by focusing only on the natural gas to be liquefied. This adjustment will also give the possibility of limiting the boundaries to the liquefaction part only (b-c). The adjusted process will be referred to as 2DLE-2. Values given in Table 4.3 are from the simulated 2DLE-2 and are based on the principle from Figure 4.9. Stream b is after a compression and aftercooler respectively at 75 bar and 38°C. The feed gas goes through liquefaction and ends up as stream c after the cold box at 74 bar and -157,3°C. The pressure drop of 1 bar is losses through the heat exchanger.

		Stream	Exergy	
Mole flow	1,632 [kgmole/s]	b	10422 [kJ/kgmole]	17008 [kJ/s]
Mass flow	27,31 [kg/s]	c	17684 [kJ/kgmole]	28858 [kJ/s]
		Min. liquefaction power	7262 [kJ/kgmole]	11851 [kJ/s]
Power liquefaction	44498 [kJ/s]			
			Specific work	0,505 kWh/kg_LNG
			Liquefaction efficiency	26,6 %

Table 4.3 Exergy calculations of the liquefaction part (without feed gas compressor and for a single train)

Liquefaction efficiency is calculated after the feed gas compressor. In Table 3.1 the feed gas compressor is included when efficiency is calculated. A reduction in efficiency will therefore occur for the liquefaction part, b-c in Figure 4.9, due to the relative high efficiency of compressors.

The high efficiency of compression and the low efficiency of liquefaction make the change in properties interesting when pressure is increased. Figure 4.7 illustrates that when pressure is increased at constant temperature the enthalpy reduces. A lower enthalpy results in lower refrigeration work.

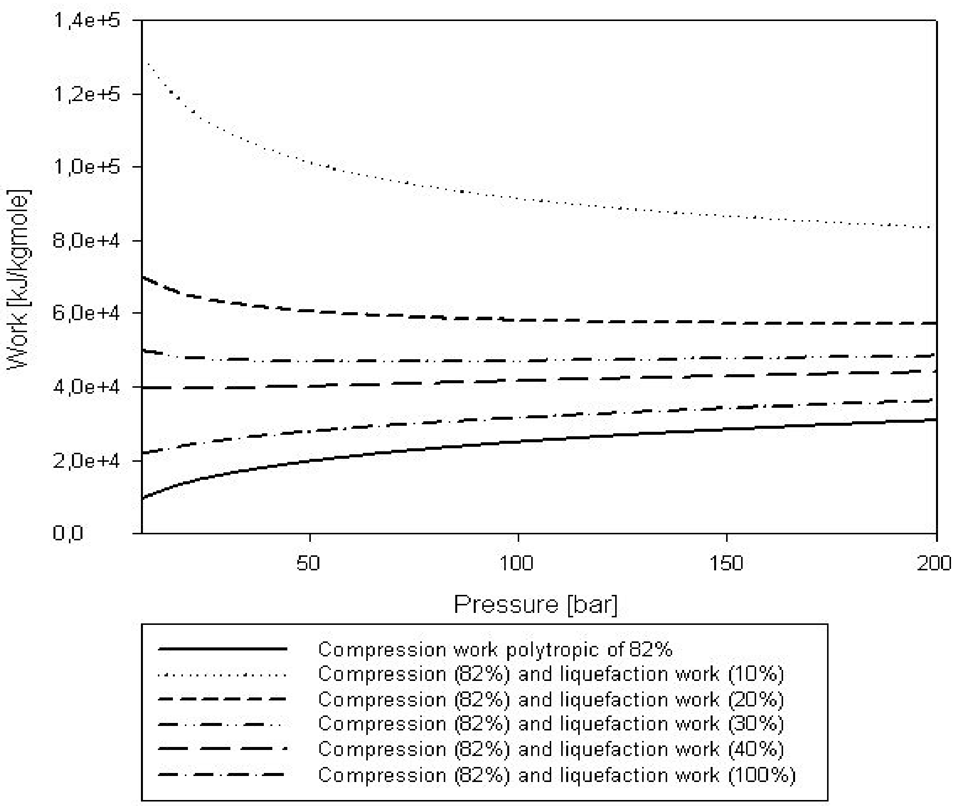


Figure 4.10 Specific work of compression from 1 bar to a certain liquefaction pressure and specific liquefaction work

To benefit from high liquefaction pressure, a compressor is necessary to reach the desired pressure. Figure 4.10 illustrates work consumption for compression alone and compression included liquefaction work. Natural gas is compressed from 1 bar with composition from Table 1.2 and then refrigerated from a certain pressure at 38°C to -157,3°C. The feed gas compressor has a polytropic efficiency of 82% and the liquefaction part is illustrated with different efficiencies. The data are given in Appendix D. Figure 4.10 illustrates that only liquefiers with low efficiencies will benefit of higher liquefaction pressures when the natural gas enters with a pressure of 1 bar. If natural gas, after a pretreatment of LPG, has a pressure higher than 1 bar a liquefier with higher efficiency will benefit from higher liquefaction pressures. Hence, the influence of the feed gas compressor work will be smaller on the whole process.

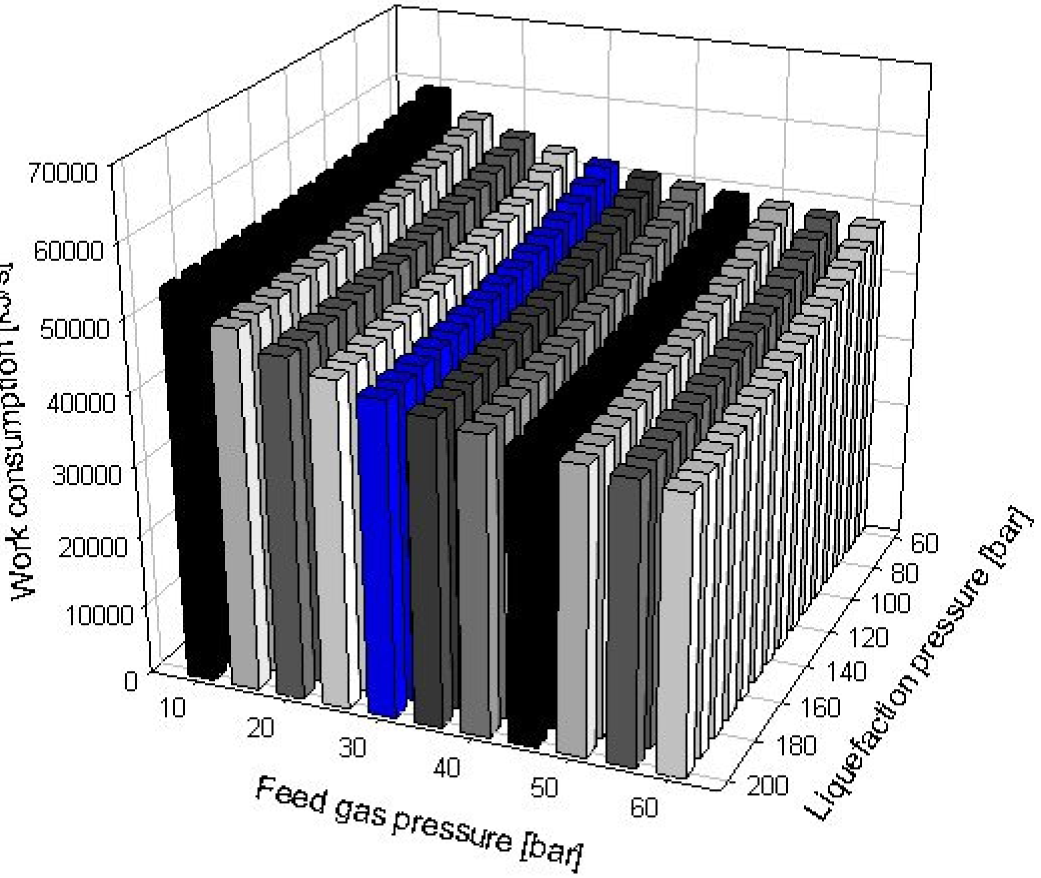


Figure 4.11 Work consumption for compression from a feed gas pressure to a liquefaction pressure and a 26,6% efficient liquefier

In Figure 4.11 feed gas at different pressures is illustrated. As illustrated a 26,6% efficient liquefier will benefit from higher liquefaction pressure for feed gas pressures at a range from 10 to 60 bar. The NicheLNG, expressed in blue, has a feed gas pressure after pretreatment of 30 bar. The total work consumption in NicheLNG is reduced with higher liquefaction pressure. Figure 4.11 is based on calculations attached in Appendix E-2.

If some adjustment to the liquefier is done to improve its efficiency new calculations are necessary. A higher efficiency for the liquefier will influence the benefits of higher liquefaction pressure. The compressor work will dominate more of the total consumption with higher liquefaction pressure. Appendix E-2 shows how an increase in efficiency of the liquefier will give a negative influence to the overall process, when liquefaction pressure is increased.

4.5 Discussion on the analysis

A decision of operating with high pressures in the process keeps the specific volume down. Low specific volume is beneficial to equipment sizes. Smaller equipments can be chosen and therefore lower necessary space for the plant. When operating at high pressures, the specific volume is less important when selecting the refrigerant gas. The difference in specific volume between nitrogen and hydrocarbon gases is very small with higher pressures. More important is their respective heat capacity. It is shown that methane has a relatively high specific heat capacity. In addition, the specific heat capacity of methane is more influenced by temperature variation. At high pressures, the specific heat capacity of methane increases with a reduction in temperature. The benefits of high heat capacity relates to lower necessary mass flow and thereby lower compressor work. So when expanders are used in a refrigeration process, methane rather than nitrogen, is more effective. It has to be noticed that methane has some restrictions on operability due to higher dew point.

The open methane cycle in the NicheLNG process enter the cold box at 38°C and cools down to -1,5°C before an expansion. By cooling it further down to -10°C before expanding, less work is necessary for the same cooling duty. How it will affect the whole process is not investigated, so this is not something one can conclude.

An increase in pressure of natural gas results in higher exergy. Cooling water is a relatively cheap refrigeration resource and is therefore seen as free. By cooling a pressurized natural gas to be liquefied with cooling water an increase in pressure will result in lower enthalpy. The stream leaving the cold box has relative high exergy and an increase in exergy of the natural gas to be liquefied results in a smaller necessary exergy change. Hence, less work is required to remove heat.

When evaluating LNG processes with emphasis on how the liquefaction pressure influence the process the control boundary is of importance. The state of the feed gas affects the benefits of higher liquefaction pressure. A low feed gas pressure with a high efficiency liquefier (40 – 50% or higher) may have a negative effect on the total work consumption if the liquefaction pressure is increased. For the NicheLNG process and other offshore based liquefiers, the efficiencies are low. As shown in Chapter 4 and Appendix D-2 liquefiers with low efficiency will take advantage of higher liquefaction pressure. The NicheLNG has a liquefaction pressure of 75 bar and a liquefier efficiency of 26,6%. With these characteristics an increase in liquefaction pressure has positive effect for any entering feed gas pressure.

5 Increased capacity of the NicheLNG process

Destination and customer for the FPSO is not yet decided. In order to have a more flexible design to meet future demands on production rate, Höegh LNG wants to look at the possibilities to expand LNG production rate of the original design. An increased capacity with implementation of additional units will be covered in this chapter. Focusing on the efficiency for each improvement, an indication of increased capacity can be shown.

The selected improvements to be investigated:

- Utilization of end flash gas (EFG)
- Liquid expander
- Two stage compression
- Higher UA value

The inert gas nitrogen affects the higher heating value of the LNG. Higher nitrogen content in LNG results in reduced heating value. Nitrogen is a more volatile gas than hydrocarbons so the EFG leaving the LNG receiver contains more nitrogen than the LNG. The EFG production is therefore necessary to obtain desired LNG specifications. All simulations in Chapter 5 are done with same efficiencies and UA-values as in the 2DLE case.

5.1 Utilization of end flash gas (EFG)

The cold duty from EFG leaving the LNG receiver is in the original design not utilized. It constitutes about 7% of the feed mass flow rate to be liquefied but has only available 3% cold duty of the necessary cooling of the feed gas. A certain amount of EFG has to be produced in order to meet the specification of LNG. Work consumed increases with the rate of EFG production. Due to the low temperature of EFG (-162°C) the work to produce this cold duty is relative high, referred to figure 1.3. Hence, the EFG has a relative high quality. From Eq. 1.6 the $COP_{th,max}$ is calculated to 0,555. A COP at 0,555 and EFG cold duty at 739 kW results in a work input of 1331 kW. The calculated COP in this case is when operating reversibly and adiabatically. To achieve the same cold duty in a real process higher work consumption is necessary.

2DLE				
Work	49393	kW	COP,th max	0,593
Specific work	0,5029	kWh/kg_LNG	COP	0,471
Necessary cooling	23254	kW	η=COP/COP,th max	0,794
EFG cold duty	739	kW		
COP,th max	0,555	(-162 °C)		
W,ideal	1332	kW		
W,real	1677	kW		
2DLE with EFG				
Work	47785	kW		
Specific work	0,486	kWh/kg_LNG		
	-3,3	*% less power consumption*		

Table 5.1 COP of 2DLE and the EFG

With the real COP for the 2DLE case a calculated efficiency increase of including EFG could be controlled. When including EFG as a cold stream in the cold box, the overall work consumed is simulated in HYSYS to be 47785 kW with the same production rate of LNG. From Table 5.1 the required real work to obtain the cold duty of EFG is 1677 kW. Adding the calculated work of EFG cold duty to the simulated work of the 2DLE with EFG, results in a power consumption of 49462 kW which is close to the power consumption in the 2DLE case. Some losses in the cold box may explain the small difference between the calculated and the simulated power saved with integration of EFG.

5.2 Liquid expander

As the natural gas to be liquefied is under a relatively high pressure and in liquid state when leaving the cold box, a liquid expander may give some improvements to the process. In the 2DLE case, pressure reduction is done through an isenthalpic valve. An isenthalpic expansion results in no work recovery and has a smaller temperature drop than for an isentropic expansion. By installing a liquid expander at the cold side of the cold box, work can be extracted and required cooling duty of the heat exchanger may be reduced.

In order to avoid destruction of the pressure exergy a liquid expander was introduced. This expander is essentially a pump run backwards that allow a subcooled liquid to be isentropically expanded almost to its bubble point. The most important benefit is the temperature reduction at very low temperature as power recovery is small at very low temperatures. Hence, isentropic expansion is an efficient way of rejecting heat, and not necessary as work recovery.

Two-phase expanders are now available and will contribute to fulfill an isentropic expansion into the two-phase region [10]. These expanders are not proven at large scale and will therefore not be covered in this thesis.

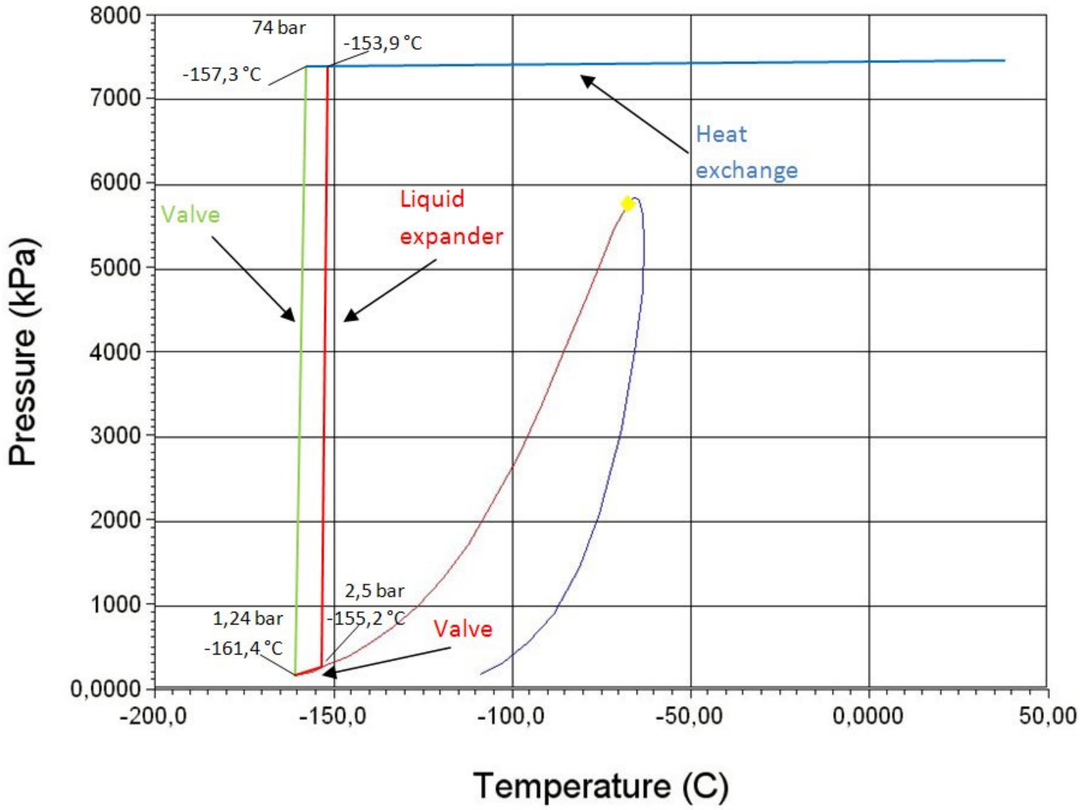


Figure 5.1 Pressure-temperature diagram for pressure reduction with valve and liquid expander followed by a valve

Figure 5.1 illustrates the cooling path of natural gas to be liquefied. The green line indicates the 2DLE case and the red is after installation of a liquid expander. Simulated in HYSYS both processes produce the same amount of LNG, 2357 ton/day for one train. As the temperature out of the heat exchanger is higher with a liquid expander, less refrigerant duty is required. Compressor power can then be reduced either in the methane or nitrogen cycle. To express the gained efficiency by introducing a liquid expander, the nitrogen refrigeration flow rate was reduced. This resulted in 2,155 MW reduced total power consumption and represents 4,3% of the 2DLE case.

Another graphical illustration of how a liquid expander contributes to the efficiency is by a temperature-entropy diagram. In Figure 5.2, a natural gas at 75 bar and 38°C is cooled down to -157,2°C.

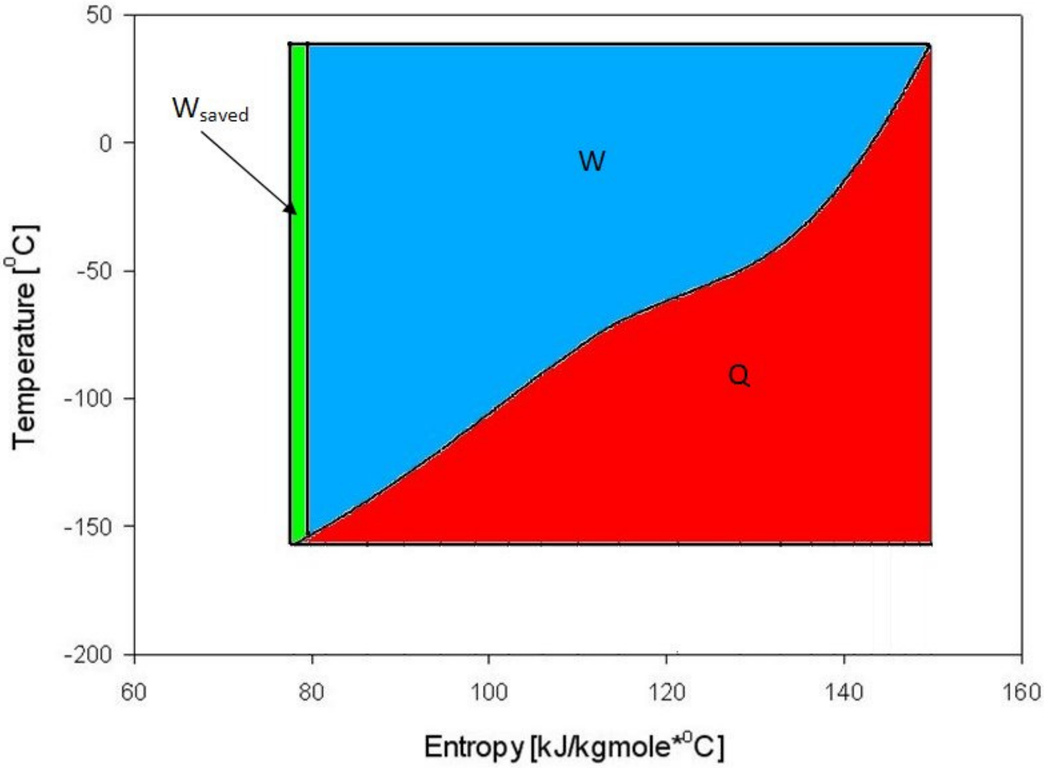


Figure 5.2 A graphical overview work saved by integration of a liquid expander

In Figure 5.2 the heat Q is extracted from the natural gas as a heat source and rejected to the surroundings as a heat sink. The necessary work W and W_{saved} are represented respectively in the blue and green area above Q. With a liquid expander the amount of work is reduced to involve only the W area.

Figure 5.2 illustrates that at low temperatures, higher outlet temperature of the heat exchanger result in a considerable work reduction. By introducing a liquid expander the COP of the reduced duty can be calculated and be an indication of saved work with increased LNG production.

	Temperature [°C]	Enthalpy [kJ/kgmole]	Flow rate [kgmole/h]	Heat flow [kJ/s]		Work [kJ/s]
2DLE	-157,2	-89022,2	5874,66	-145270,8316		49620,9
2DLE w/liquid expander	-153,8	-88819,7	5874,66	-144940,3649		47466,3
			Reduced duty	330,5	Reduced work	2154,7
					COP	0,153

Increased LNG production	Flow rate [kgmole/h]	Reduced duty [kJ/s]	Reduced work [kJ/s]
5 %	6168,4	347,0	2262,4
10 %	6462,1	363,5	2370,1
15 %	6755,9	380,0	2477,9
20 %	7049,6	396,6	2585,6
25 %	7343,3	413,1	2693,4
30 %	7637,1	429,6	2801,1

Table 5.2 Results from the integration of a liquid expander

In Table 5.2 the values for both 2DLE cases are simulated in HYSYS. Based on the two cases the COP is calculated and can then be used as an indication of reduced work with liquid expander at higher LNG production. Even though reduced work increases with LNG rate, reduced work in percentage will stay constant.

5.3 Two stage compression

The compression of feed gas and methane in the refrigeration cycle is done with one compressor. The flow rates of these streams are 63% of the total flow rate (natural gas and nitrogen) through the cold box and require a 22,3 MW compressor. By including a second compressor in series with an intercooler, power consumption will be reduced.

The new compressor is assumed to have the same efficiency as the one already in place. To find the optimum pressure increase for the first compressor, a case was simulated in HYSYS. With a pressure variation from inlet pressure to liquefaction pressure the optimum middle pressure where found to be 48 bar. This resulted in a power reduction of 1,56 MW and represent a reduction of 6,8%. For the 2DLE process, a two stage compression reduced the overall power consumption by 3,2% at the same LNG production.

5.4 The improvements influence by higher LNG production

Höegh LNG wants to look at how an increase in production affects the process and how new improvements will contribute to keep the power consumption down. The three proposed solutions to increase the efficiency of the NicheLNG are all promising without huge changes to its original design. Additional units increase the equipment count and more space is needed. By higher LNG rate the size of each unit will also increase. A higher feed gas rate results in a higher refrigerant rate, and hence an increase in unit size. The heat exchanger will be particularly influenced by higher flow rates. Later it will be illustrated how each improvement affects the process with a 25 % higher LNG production. The 2DLE case was extended with liquid expander, two stage compression and EFG. The compressors in the 2DLE case had polytropic efficiencies decided by vendor curves. They all had efficiencies around 80%, so in the simulations with higher LNG rate, all compressors were defined with polytropic efficiencies at 82%. To meet the LNG specifications, higher heating value was hold constant at 10,95 kWh/m^3 (1058 BTU/scf) for both cases. To accomplish a constant higher heating value the refrigeration duty was varied to keep the nitrogen content below 1 mole%. In the two following cases the refrigeration duty for the methane cycle was held constant and in the nitrogen cycle the flow rate was increased.

An increase in LNG production demands a higher duty of the heat exchanger. In the 2DLE the UA value is constant at 23 860 MJ/°C*h so a higher duty results in a larger ΔT in the heat exchanger. If it is desired to keep the power consumption down and without additional equipment it is necessary to increase the heat exchanger size. A higher UA value will allow reduced ΔT for the same duty, thus reduced exergy losses and reduced power consumption.

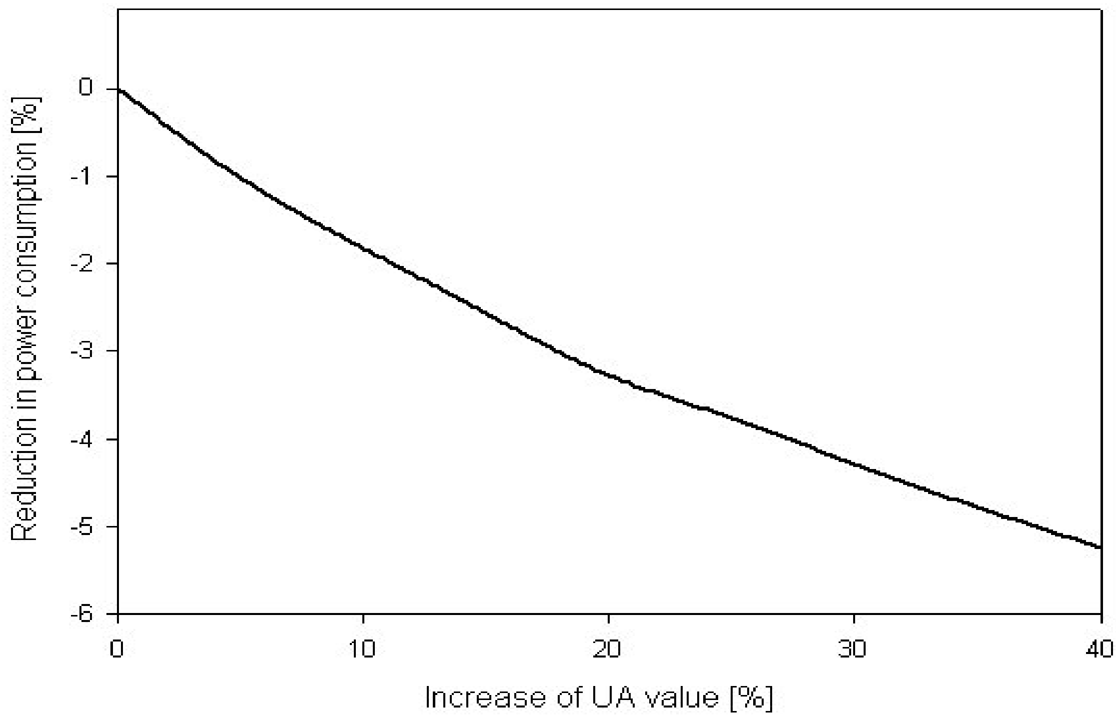


Figure 5.3 25% higher LNG production with increase of UA-value

The results represented in Figure 5.3 are based on simulations of the 2DLE with 25% higher LNG production. The process demanded 0,5502 kWh/kg_{LNG} with the original UA value at 23 860 MJ/°C*h. Illustrated is the power savings with an increase of the UA value.

LNG production of 2946,25 ton/day (25% increase)				
	2DLE	2DLE with EFG	2DLE with Liquid Expander	2DLE with compression
Specific work [kWh/kg_LNG]	0,5502	0,5316	0,5225	0,5361
% reduction		-3,4	-5,0	-2,6

Table 5.3 Individual improvements in efficiency with new a unit or change in the design (EFG)

In Table 5.3 each improvement is represented alone with 25% increase LNG production. The UA value is constant at 23 860 MJ/°C*h and only the nitrogen flow rate is adjusted to meet the LNG specifications. The improvement with a liquid expander stands out as the most efficient solution.

In order to see how the improvements together affected the process they were all simulated in two cases, with constant UA value and constant LMTD. It was chosen to see how the efficiency of the heat exchanger affected an extended process with all three improvements. The two cases were based on the heat exchanger specifications from the 2DLE case with emphasis on its UA value and LMTD. Both extended processes had the same LNG production and specifications.

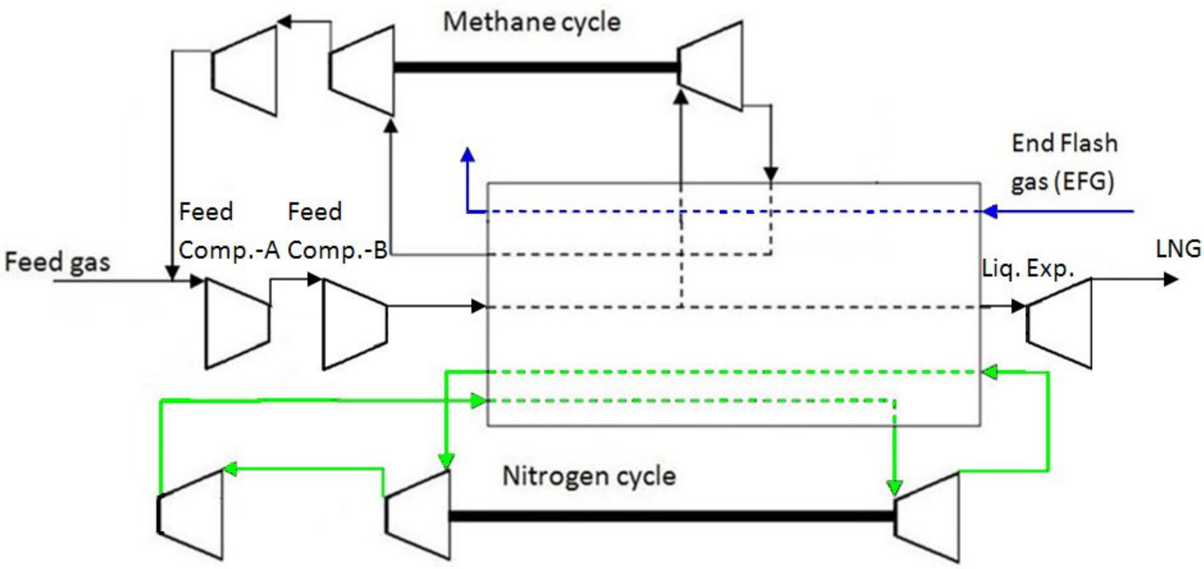


Figure 5.4 The extended NicheLNG process

Figure 5.4 illustrates the extended NicheLNG process simulated in HYSYS. Black represent the natural gas path, green the nitrogen path and blue the EFG. In HYSYS, the process was simulated with the heat exchanger divided in four parts, HX-1, HX-2, HX-3 and HX-4. Each HX has its own UA value and LMTD.

		LNG production of 2946,25 ton/day (25% increase)		
	2DLE original	2DLE	Extended 2DLE with constant UA	Extended 2DLE with constant LMTD
Specific work	0,5049 kWh/kg_LNG	0,5502 kWh/kg_LNG	0,4913 kWh/kg_LNG *-10,7 % of the 2DLE*	0,4791 kWh/kg_LNG *-12,9 % of the 2DLE*
Overall exergy efficiency	30,0 %	27,8 %	30,0 %	30,8 %
LMTD:				
HX-1	6,85 °C	7,04 °C	6,85 °C	6,85 °C
HX-2	7,81 °C	11,31 °C	10,00 °C	7,81 °C
HX-3	13,74 °C	30,35 °C	24,82 °C	13,74 °C
HX-4	10,88 °C	14,15 °C	14,30 °C	10,88 °C
UA	23 860 MJ/°C*h	23 860 MJ/°C*h	23 860 MJ/°C*h	26528 MJ/°C*h *11,2 %higher UA value*
LNG production	2357,25 ton/day	2946,25 ton/day	2946,25 ton/day	2946,25 ton/day

Table 5.4 Results of 25 % increase in LNG production

Table 5.4 represents four simulated cases. Feed gas enters with 30 bar and 44°C. In all cases the compressors and expanders have polytropic efficiency of 82% and adiabatic efficiency of 87% respectively. The *2DLE original* is without increase in LNG production and changes in specifications. In the *2DLE* the capacity is increased with 25% and with the same design as in *2DLE original*. For both of the *Extended 2DLE* the design is illustrated in Figure 5.4. Both cases are either based on the same UA value or LMTD as in the *2DLE original*. Equal LMTD, based on the *2DLE original*, was chosen to see how the UA value affected the efficiency of the process.

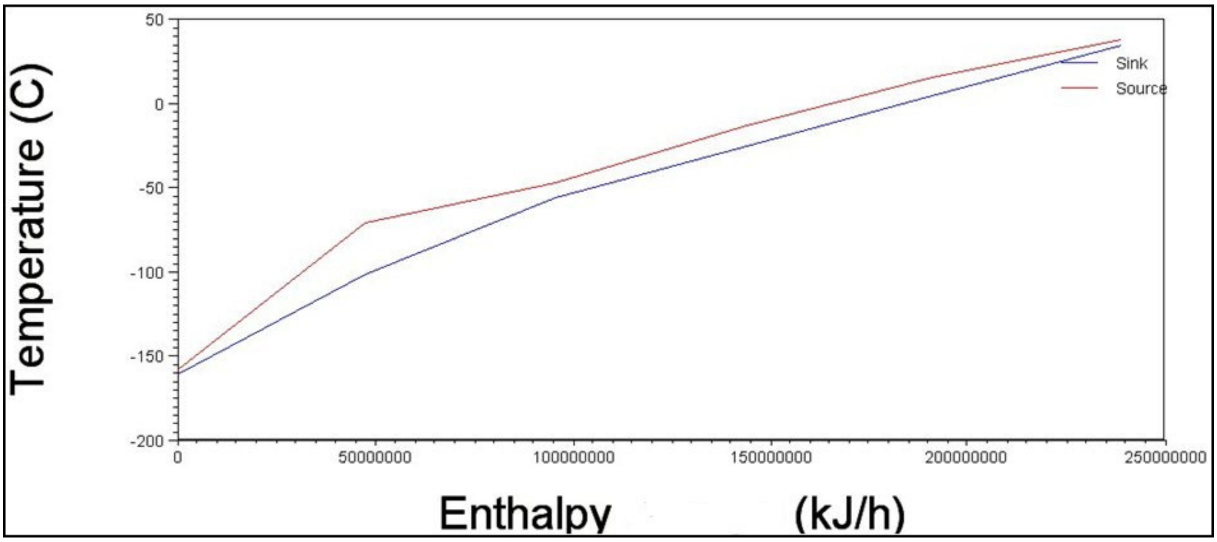


Figure 5.5 Temperature-enthalpy diagram of the 2DLE with 25% increased capacity

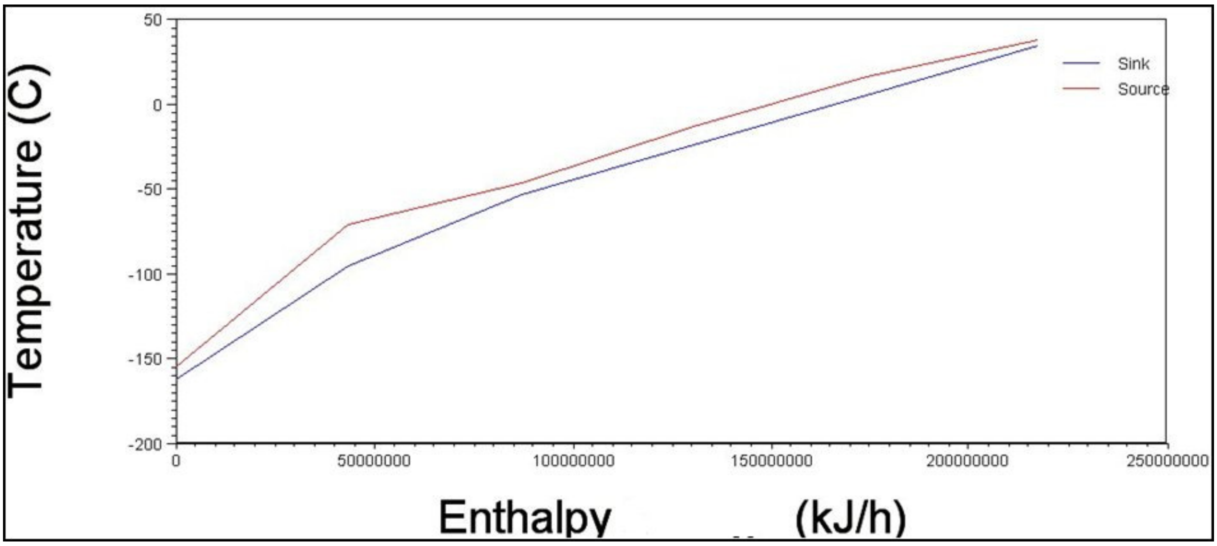


Figure 5.6 Temperature-enthalpy diagram of the Extended 2DLE with constant LMTD and 25% increased capacity

Figure 5.5 and 5.6 illustrate the composite curves for the *2DLE* with 25% increased capacity and the *Extended 2DLE with constant LMTD* also with 25% increased capacity. From the figures it can be observed that the *Extended 2DLE* with constant LMTD demands less duty and the composite curves are closer in the temperature region below -70°C.

5.5 Discussion on increased capacity

Three solutions to make the liquefaction process more efficient have been described. A more efficient process, and especially with higher production rate, can defend higher investment costs.

After LPG extraction the feed gas entering the liquefaction process has too high nitrogen content to meet the LNG specifications. Some end flash gas has to be produced and it has a relative high refrigerant quality due to its low temperature. The refrigerant quality of a gas depends on its capacity to attract heat and the efficiency of the liquefaction process. With less efficient liquefaction process any produced EFG will be more valuable. In the 2DLE case with normal LNG production an EFG utilization would affect the process with 3,3% lower power consumption. Destination of the FPSO is not decided and the nitrogen content in the feed may therefore vary. The nitrogen content has an influence on the process and how utilization of EFG will affect power consumption.

The natural gas stream leaving the cold box is already in liquid phase, and still under pressure. By expansion in a turbine, instead of a valve, power can be recovered. In this case the power recovery is very small (394 kW), but more important is the temperature drop. When fluid does work by expansion, at very low temperatures, heat rejection is more valuable. With a liquid expander, the temperature of LNG leaving the heat exchanger could be increased by 3,4°C. This reduced the refrigeration load and resulted in a 4,3% lower power consumption.

Feed gas and the methane cycle are compressed by one compressor. This compressor is responsible for almost half of the total power consumption. With an additional compressor so the compression is done over two stages with an intercooler, the total power consumption is reduced by 3,2%.

When LNG production was increased by 25% the liquid expander did stand out as the improvement with highest contribution to efficiency. The liquid expander alone reduced the power consumption with 5%. The efficiency of the liquefaction part was found to be 26,6% in Chapter 4.4.1. Due to its low efficiency an improvement that contributes to lower duty of the liquefaction part will have an increasing influence with higher LNG production. In contrast is the improvement with interstage compression. Compressors have a relative high efficiency. When LNG production increases the power consumption of the liquefaction part will grow more rapidly than for the compression. This explains the decrease from 3,2% to 2,6% when LNG production is increased by 25%. The improvement on utilizing the EFG production is not influenced by the higher LNG production. With higher LNG production the EFG production increases with the same rate and the 3,3-3,4% reduction will stay constant.

An alternative or another contribution to higher efficiency is to operate with a more efficient heat exchanger. A heat exchanger with higher UA value reduced the power consumption by an average of 0,16% per percent of increased UA value.

The three improvements together were simulated with higher LNG production. Even though production was increased by 25% it was more efficient than at normal production rate. It applies for both simulated cases. The two extended NicheLNG processes with 25% higher LNG production will also affect the heat exchanger. An equally efficient heat exchanger as in 2DLE results in higher UA value and thereby more necessary space. Depending on available space, a compromise on efficiency and units or size will decide the final design.

6 Conclusions and further work

6.1 Conclusions

The NicheLNG process, the chosen liquefaction process for the HLNG FPSO-1, has been described and evaluated with respect to energy efficiency. The improvement potentials and energy savings have been presented by thermodynamic analysis and simulations in HYSYS. A comparison with a promising alternative process has also been presented. Possibilities to expand the NicheLNG process is considered with increased LNG capacity.

In the evaluation of suitable natural gas liquefaction processes for offshore applications, the expander processes did stand out as the most promising when emphasis is on; compactness, safety, operation and equipment count. The most proposed expander process, a dual nitrogen process, was compared with the NicheLNG process with emphasis on power consumption and energy efficiency. With equal conditions and LNG production, the NicheLNG process with an exergy efficiency of 31,2% had 10% lower work consumption. Natural gas as refrigerant has higher c_p than nitrogen, resulting in a significantly lower mass flow rate. Hence, lower flow rate contributes to lower compression work.

The benefit in terms of energy consumption with higher liquefaction pressure depends on the feed gas pressure and the efficiency of the liquefier. A constant heat rejection temperature and with increasing liquefaction pressure, the necessary heat to be removed is reduced. The NicheLNG liquefier has an exergy efficiency of 26,6%. With this efficiency and feed gas pressure in the range of 10 bar to 60 bar, a higher liquefaction pressure will have a positive influence on the work consumption.

The improvements; utilization of EFG, liquid expander, two stage compression and higher UA-value, were individually discussed and evaluated with LNG production as normal and 25% higher capacity. The utilization of End Flash Gas and a new compressor reduced the work consumption by 3,4% and 2,6%, respectively. These improvements reduced the work consumption by a few percent but the liquid expander, at 25% higher LNG production, stands out alone as the improvement with the highest contribution. With a liquid expander, the work consumption is reduced by 5%.

The NicheLNG process with 25% higher LNG production demands, in the terms of specific work, 0,5502 kWh/kg_{LNG}. By extending the process with the proposed improvements, the work consumption is reduced but on the cost of space. The two extended process resulted a specific work of 0,4913 kWh/kg_{LNG} and 0,4791 kWh/kg_{LNG}. They had the same improvements except that the one with the lowest energy demand differ by its heat exchanger with a 11,2% higher UA value.

6.2 Suggestions on further work

In the next phase of the NicheLNG analysis, a more practical view should be evaluated on how the adjustments and the expansions suggested in this thesis will affect the topside of the FPSO-1. It has to be taken account of available space and how the changes, on operability, will be influenced. It should also be evaluated whether the improvements can be justified with respect to investment costs.

A change of the design by closing the open methane cycle, the impact an additional compressor has on the process should be evaluated. With a closed methane cycle, higher pressures in the cycle can be chosen. The influence, a higher pressure level has on the unit sizes and the work consumption, should be investigated.

Since there already are heavy hydrocarbons in liquid phase (LPG) on the topside of the FPSO-1, a precooler based on propane or a mixture of heavy hydrocarbons can be justified. The benefits, in terms of reduced work consumption, of a precooler in front of the NicheLNG process must also be evaluated with respect to investment costs. On the other hand, if it is decided to operate with liquid refrigerants, a replace of the NicheLNG process with the Single Mixed Refrigerant process could be a promising alternative.

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Appendix A

Exergy change and liquefaction work calculated with data from HYSYS simulations.

Ambient temperature: 298K

Stream	NG_ambient 3101	LNG To tank	After cold box	After feed gas comp.
Phase	1	0	0	1
Temperature [C]	25	-163,4471176	-157,1901697	38
Pressure [kPa]	100	104	7377,3	7500
Molar Flow [kgmole/h]	5874,66	11341,11324	5874,66	5874,66
Mass Flow [kg/h]	102467,3748	196437,1925	102467,3748	102467,3748
Std Ideal Liq Flow [m3/h]	326,2525717	633,6976784	326,2525717	326,2525717
Molar Enthalpy [kJ/kgmole]	-74040,18449	-90696,56503	-89022,17211	-74781,50596
Molar Entropy [kJ/kgmole-C]	186,85275	75,46587186	77,96521789	149,6119422
Heat Flow [kJ/h]	-434960910,2	-1028600014	-522974993,6	-439315921,8

Data and calculation for: 1 bar (NG_ambient 3101)

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
17466,5	5874,66	102609731	28503	102467,3748	0,278	100 %
19407,2	5874,66	114010812	31670	102467,3748	0,309	90 %
21833,1	5874,66	128262164	35628	102467,3748	0,348	80 %
24952,1	5874,66	146585330	40718	102467,3748	0,397	70 %
29110,8	5874,66	171016218	47505	102467,3748	0,464	60 %
34933,0	5874,66	205219462	57005	102467,3748	0,556	50 %
43666,2	5874,66	256524327	71257	102467,3748	0,695	40 %
58221,7	5874,66	342032436	95009	102467,3748	0,927	30 %
87332,5	5874,66	513048654	142514	102467,3748	1,391	20 %
174665,0	5874,66	1026097309	285027	102467,3748	2,782	10 %

Data and calculation for: 10 bar

Stream	1
Phase	1
Temperature [C]	25
Pressure [kPa]	1000
Molar Flow [kgmole/h]	5874,66
Mass Flow [kg/h]	102467,375
Std Ideal Liq Flow [m3/h]	326,252572
Molar Enthalpy [kJ/kgmole]	-74195,3555
Molar Entropy [kJ/kgmole-C]	167,3435
Heat Flow [kJ/h]	-435872487

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
11807,9	5874,66	69367466	19269	102467,375	0,188	100 %
13119,9	5874,66	77074962	21410	102467,375	0,209	90 %
14759,9	5874,66	86709332	24086	102467,375	0,235	80 %
16868,4	5874,66	99096380	27527	102467,375	0,269	70 %
19679,9	5874,66	115612443	32115	102467,375	0,313	60 %
23615,8	5874,66	138734931	38537	102467,375	0,376	50 %
29519,8	5874,66	173418664	48172	102467,375	0,470	40 %
39359,7	5874,66	231224886	64229	102467,375	0,627	30 %
59039,6	5874,66	346837328	96344	102467,375	0,940	20 %
118079,1	5874,66	693674657	192687	102467,375	1,880	10 %

Data and calculation for: 20 bar

Stream	1
Phase	1
Temperature [C]	25
Pressure [kPa]	2000
Molar Flow [kgmole/h]	5874,66
Mass Flow [kg/h]	102467,375
Std Ideal Liq Flow [m3/h]	326,252572
Molar Enthalpy [kJ/kgmole]	-74371,5146
Molar Entropy [kJ/kgmole-C]	161,152765
Heat Flow [kJ/h]	-436907362

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
10139,2	5874,66	59564538	16546	102467,375	0,161	100 %
11265,8	5874,66	66182820	18384	102467,375	0,179	90 %
12674,0	5874,66	74455673	20682	102467,375	0,202	80 %
14484,6	5874,66	85092198	23637	102467,375	0,231	70 %
16898,7	5874,66	99274230	27576	102467,375	0,269	60 %
20278,5	5874,66	119129077	33091	102467,375	0,323	50 %
25348,1	5874,66	148911346	41364	102467,375	0,404	40 %
33797,4	5874,66	198548461	55152	102467,375	0,538	30 %
50696,2	5874,66	297822691	82729	102467,375	0,807	20 %
101392,3	5874,66	595645383	165457	102467,375	1,615	10 %

Data and calculation for: 40 bar

Stream	1
Phase	1
Temperature [C]	25
Pressure [kPa]	4000
Molar Flow [kgmole/h]	5874,66
Mass Flow [kg/h]	102467,375
Std Ideal Liq Flow [m3/h]	326,252572
Molar Enthalpy [kJ/kgmole]	-74735,3138
Molar Entropy [kJ/kgmole-C]	154,490483
Heat Flow [kJ/h]	-439044559

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
8517,7	5874,66	50038420	13900	102467,375	0,136	100 %
9464,1	5874,66	55598245	15444	102467,375	0,151	90 %
10647,1	5874,66	62548025	17374	102467,375	0,170	80 %
12168,1	5874,66	71483457	19857	102467,375	0,194	70 %
14196,1	5874,66	83397367	23166	102467,375	0,226	60 %
17035,3	5874,66	100076840	27799	102467,375	0,271	50 %
21294,2	5874,66	125096050	34749	102467,375	0,339	40 %
28392,2	5874,66	166794734	46332	102467,375	0,452	30 %
42588,4	5874,66	250192101	69498	102467,375	0,678	20 %
85176,7	5874,66	500384201	138996	102467,375	1,356	10 %

Data and calculation for: 60 bar

Stream	1
Phase	1
Temperature [C]	25
Pressure [kPa]	6000
Molar Flow [kgmole/h]	5874,66
Mass Flow [kg/h]	102467,375
Std Ideal Liq Flow [m3/h]	326,252572
Molar Enthalpy [kJ/kgmole]	-75112,4249
Molar Entropy [kJ/kgmole-C]	150,169707
Heat Flow [kJ/h]	-441259958

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
7607,2	5874,66	44689659	12414	102467,375	0,121	100 %
8452,4	5874,66	49655176	13793	102467,375	0,135	90 %
9509,0	5874,66	55862073	15517	102467,375	0,151	80 %
10867,4	5874,66	63842370	17734	102467,375	0,173	70 %
12678,7	5874,66	74482764	20690	102467,375	0,202	60 %
15214,4	5874,66	89379317	24828	102467,375	0,242	50 %
19018,0	5874,66	111724147	31034	102467,375	0,303	40 %
25357,3	5874,66	148965529	41379	102467,375	0,404	30 %
38036,0	5874,66	223448293	62069	102467,375	0,606	20 %
76071,9	5874,66	446896587	124138	102467,375	1,211	10 %

Data and calculation for: 80 bar

Stream	1
Phase	1
Temperature [C]	25
Pressure [kPa]	8000
Molar Flow [kgmole/h]	5874,66
Mass Flow [kg/h]	102467,375
Std Ideal Liq Flow [m3/h]	326,252572
Molar Enthalpy [kJ/kgmole]	-75497,4279
Molar Entropy [kJ/kgmole-C]	146,79389
Heat Flow [kJ/h]	-443521720

Spec. exergy change [kJ/kgmole]	Molar flow [kgmole/h]	Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
6986,2	5874,66	41041550	11400	102467,375	0,111	100 %
7762,4	5874,66	45601723	12667	102467,375	0,124	90 %
8732,8	5874,66	51301938	14251	102467,375	0,139	80 %
9980,3	5874,66	58630786	16286	102467,375	0,159	70 %
11643,7	5874,66	68402584	19001	102467,375	0,185	60 %
13972,4	5874,66	82083101	22801	102467,375	0,223	50 %
17465,5	5874,66	102603876	28501	102467,375	0,278	40 %
23287,3	5874,66	136805168	38001	102467,375	0,371	30 %
34931,0	5874,66	205207751	57002	102467,375	0,556	20 %
69862,0	5874,66	410415503	114004	102467,375	1,113	10 %

75 bar:

Spec. exergy change [kJ/kgmole]		Exergy change [kJ/h]	Exergy change [kJ/s]	Mass flow [kg/h]	Work/ mass [kWh/kg LNG]	Efficiency
7110,1		41769171	11603	102467,375	0,113	100 %

Appendix B

From Aspentech HYSYS Support was 'Weighted Model' chosen as heat exchanger parameter. Bellow is an explanation of the difference between the various heat exchanger models, given by Aspentech.

What are the differences between the various heat exchanger models?	
Solution ID:	109410
Product(s):	Aspen HYSYS
Version(s):	2.0
Primary subject:	Unit Operations, Heat Transfer Equipment, Heat Exchanger
Last Modified:	12-Aug-2005

Applicable Version(s):

Applies to HYSYS Versions 2.0 - current

Problem Statement:

What are the differences between the various heat exchanger models?

Solution:

There are five shell and tube heat exchanger models available in HYSYS. The End Point and Weighted models can be used for material and energy balance for any two-sided heat exchangers. They can also be used for shell and tube exchanger"s material and energy balance. Steady-State Rating model is used for rating of shell and tube exchangers. The Dynamic Basic and Detailed Models are used for rating in steady-state mode as well as in dynamic simulation.

The basics of each model:

1. End Point Model

This model is based on Q = U A Ft LMTD. The main assumptions behind this model are that the overall heat transfer coefficient U is constant the specific heats of the streams at both exchanger sides are constant. The heat curves of both shell and tube side are linear. The heat exchanger geometry is not taken into account in this model.

2. Weighted Model

This model is particularly powerful in dealing with non-linear heat curve problems such as phase change of pure components in one or both heat exchanger sides. The heat curves are divided into a number of intervals and an energy balance is performed in each interval. This model can only be used for energy and material balance. The heat exchanger geometry is not taken into account in this model.

3. Steady State Rating Model

This model makes the same assumptions as the End Point Model. It"s simply an extension of the End Point model which incorporates a rating calculation. If detailed geometry information is provided, the exchanger can be rated using this model. For linear or nearly linear heat curve problems, this model is a good choice because it is much faster than the dynamic rating-detailed model.

4. Dynamic Basic Model

The Basic Model is based on Q = U A Ft LMTD and makes the same assumptions as the End Point model. This model was originally developed for dynamic mode but was extended for rating in steady state. This model is somewhat oversimplified in that geometry configurations are not taken into account. Therefore, this model has limited functionality. When using this model, both pressure drops and the overall UA must be specified.

5. Dynamic Detailed Model

The Detailed Model divides the entire heat exchanger into a number of heat zones. In each heat zone there is a shell hold-up and one or more tube hold-ups, according to the number of tube passes per shell pass. It is a good counterpart to the Weighted Model. The Dynamic Detailed Model is used both in steady state and in dynamic operation and is designed to solve any linear and non-linear heat curve problems.

Keywords:

heat exchanger, model

Appendix C

Calculations of the methane refrigeration cycle with different inlet expander temperatures.

Inlet temperature [°C]	Internal heat exchange [kJ/kgmole]	Refrigeration load [kJ/kgmole]	Cooling duty [kJ/kgmole]	Compressor work [kW]	Work/ refrigeration load [-]
38	0,0	5024,4	5024,4	40275,1	1,335
10	1354,9	3190,0	4544,9	21740,5	1,135
-1,5	1949,0	2568,0	4517,0	17084,4	1,108
-5	2137,1	2360,5	4497,6	15538,1	1,096
-8	2301,5	2211,7	4513,2	14524,7	1,094
-10	2413,0	2094,4	4507,4	13688,1	1,088
-12	2526,2	1999,4	4525,6	13076,4	1,089
-15	2699,4	1846,2	4545,7	12074,8	1,089
-20	2999,1	1588,6	4587,7	10407,0	1,091
-30	3654,8	1128,8	4783,6	7671,4	1,132

Flow rate of 21618,85 kgmole/h

Appendix D

Feed gas compressed to a liquefaction pressure. Compressor has a polytropic efficiency of 82% and the liquefiers efficiencies varies.

Compression from 1 bar (82% polytropic eff.) and liquefiers with different efficiencies

	Compression and liquefaction work				
Pressure [bar]	10 % [kJ/kgmole]	20 % [kJ/kgmole]	30 % [kJ/kgmole]	40 % [kJ/kgmole]	100 % [kJ/kgmole]
10	130231	69963	49874	39829	21749
20	116995	65325	48102	39490	23989
30	109785	62999	47404	39607	25571
40	104942	61543	47077	39843	26824
50	101365	60537	46928	40123	27875
60	98575	59806	46883	40421	28790
70	96316	59253	46899	40722	29603
80	94439	58827	46956	41021	30337
90	92849	58494	47042	41316	31010
100	91479	58230	47147	41606	31631
110	90283	58020	47266	41889	32210
120	89226	57852	47394	42165	32753
130	88281	57716	47528	42433	33264
140	87428	57605	47664	42694	33747
150	86649	57514	47802	42946	34205
160	85932	57438	47939	43190	34642
170	85267	57373	48075	43426	35058
180	84644	57317	48208	43654	35456
190	84056	57268	48338	43874	35837
200	83499	57224	48465	44086	36203

Appendix E-1

Feed gas compressed to a liquefaction pressure. Compressor has a polytropic efficiency of 82%. Exergy values are exergy change from compressed natural gas to stream out of cold box (Liquefier 100% efficiency).

Mole flow	1,63185 [kgmole/s]	Ambient condition	1 bar 27 C = 300K	Molar Enthalpy	Molar Entropy
				-73965,525	187,102323

Pressure [bar]	Compression power from 1 bar 38 C [kJ/s]	[kJ/kgmole]	Molar Enthalpy [kJ/kgmole]	Molar Entropy [kJ/kgmole]	Exergy [kJ/kgmole]	[kJ/s]
10	15821	9695	-73689,9369	169,251367	5631	9189
20	22282	13655	-73851,2731	162,981829	7350	11995
30	26458	16214	-74015,2325	159,17874	8327	13589
40	29608	18144	-74181,5822	156,366588	9005	14694
50	32162	19709	-74349,9773	154,091118	9519	15533
60	34328	21036	-74519,9375	152,152323	9931	16205
70	36211	22190	-74690,8266	150,445022	10272	16762
80	37883	23215	-74861,8394	148,90782	10562	17236
90	39391	24139	-75032,0035	147,502499	10813	17646
100	40766	24982	-75200,2024	146,204323	11035	18007
110	42033	25758	-75365,2226	144,996888	11232	18329
120	43208	26478	-75525,8258	143,869093	11410	18619
130	44306	27151	-75680,8366	142,813199	11571	18883
140	45337	27782	-75829,2304	141,823503	11720	19125
150	46309	28378	-75970,2061	140,895435	11857	19349
160	47230	28943	-76103,2278	140,024967	11986	19559
170	48105	29479	-76228,0303	139,208272	12106	19755
180	48940	29991	-76344,2241	138,442765	12219	19940
190	49738	30480	-76452,2599	137,723827	12327	20116
200	50503	30948	-76552,2357	137,048507	12429	20283

Stream out of coldbox -157,3 C	Molar enthalpy [kJ/kgmole]	Molar entropy [kJ/kgmole]	Exergy [kJ/kgmole]
74 bar	-89022,1721	77,9652179	17684,4843

Appendix E-2

Compressions for all calculations are done with a polytropic efficiency of 82%. *Feed* is pressure entering compressor after pretreatment and *Liq*. is outlet compressor pressure/liquefaction pressure.

Compression from a feed gas pressure (38 C) and to liquefaction pressure [bar]

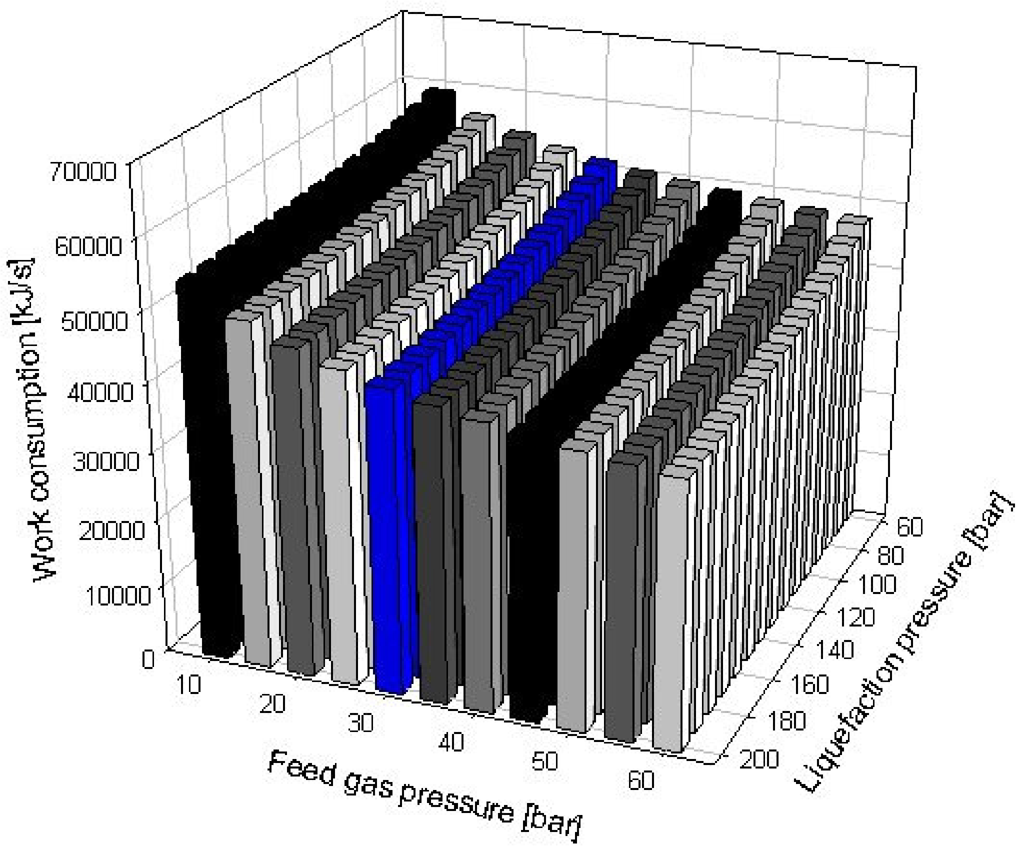
Feed	10	15	20	25	30	35	40	45	50	55	60
Liq.	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]
60	11498	8395	6363	4885	3744	2828	2071	1434	888	414	0
65	12136	8975	6902	5392	4225	3286	2510	1855	1293	805	378
70	12738	9523	7412	5872	4680	3720	2925	2253	1677	1176	736
75	13307	10043	7893	6326	5111	4131	3320	2633	2042	1529	1077
80	13848	10536	8354	6759	5522	4524	3696	2994	2391	1866	1403
85	14364	11007	8792	7174	5915	4900	4056	3341	2725	2188	1715
90	14856	11458	9212	7569	6292	5260	4401	3673	3046	2498	2015
95	15328	11890	9616	7950	6654	5606	4733	3993	3354	2796	2304
100	15784	12305	10004	8316	7002	5939	5053	4301	3651	3084	2583
105	16220	12704	10377	8669	7339	6260	5362	4598	3939	3362	2852
110	16641	13090	10738	9009	7662	6571	5661	4886	4217	3631	3113
115	17047	13463	11087	9339	7977	6872	5950	5165	4486	3892	3366
120	17440	13824	11425	9660	8282	7164	6231	5436	4748	4145	3611
125	17821	14173	11752	9970	8577	7447	6503	5699	5002	4391	3850
130	18190	14513	12070	10271	8865	7722	6768	5955	5250	4631	4083
135	18549	14842	12379	10564	9144	7991	7027	6204	5491	4865	4310
140	18897	15163	12680	10850	9416	8252	7278	6447	5726	5093	4531
145	19237	15476	12973	11128	9683	8507	7523	6684	5955	5315	4747
150	19567	15782	13259	11399	9942	8756	7763	6915	6179	5533	4959
155	19889	16079	13538	11664	10196	8998	7997	7142	6399	5746	5165
160	20204	16369	13811	11923	10443	9236	8226	7363	6613	5954	5368
165	20511	16653	14077	12176	10685	9469	8451	7580	6824	6158	5566
170	20811	16930	14338	12424	10922	9698	8670	7792	7030	6358	5761
175	21105	17201	14593	12666	11154	9921	8886	8001	7232	6554	5952
180	21392	17467	14843	12904	11382	10140	9097	8205	7430	6747	6139
185	21674	17727	15088	13137	11605	10355	9304	8406	7624	6936	6323
190	21949	17982	15329	13366	11824	10565	9508	8603	7815	7122	6504
195	22220	18233	15564	13591	12040	10773	9708	8797	8004	7305	6682
200	22485	18479	15796	13811	12251	10976	9906	8987	8189	7485	6857

Total work consumption of compression and a liquefier with an efficiency of 26,6%. *Feed* is pressure entering compressor after pretreatment and *Liq.* is outlet compressor pressure/liquefaction pressure.

Compression and liquefaction 26,6%

Feed [bar]	10	15	20	25	30	35	40	45	50	55	60
Liq. [bar]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]
60	59010	55907	53875	52397	51256	50340	49584	48946	48400	47926	47512
70	58159	54943	52832	51292	50100	49140	48346	47674	47098	46597	46157
80	57491	54179	51996	50402	49165	48167	47339	46637	46034	45508	45046
90	56959	53560	51314	49671	48394	47362	46503	45775	45148	44600	44117
100	56530	53051	50750	49062	47749	46685	45800	45047	44398	43830	43329
110	56179	52628	50276	48548	47200	46109	45199	44424	43755	43169	42651
120	55889	52273	49874	48109	46731	45613	44680	43885	43197	42594	42060
130	55648	51971	49528	47729	46323	45180	44226	43412	42707	42089	41541
140	55445	51711	49228	47398	45964	44800	43826	42995	42274	41640	41079
150	55273	51488	48965	47105	45648	44461	43469	42621	41885	41238	40664
160	55124	51290	48732	46843	45364	44157	43147	42284	41534	40874	40288
170	54995	51114	48522	46608	45106	43882	42854	41976	41214	40542	39945
180	54881	50956	48332	46393	44871	43629	42586	41694	40919	40236	39628
190	54779	50812	48158	46195	44654	43395	42337	41432	40644	39951	39333
200	54686	50679	47996	46011	44451	43177	42106	41188	40389	39685	39058

Graph for the calculations above:

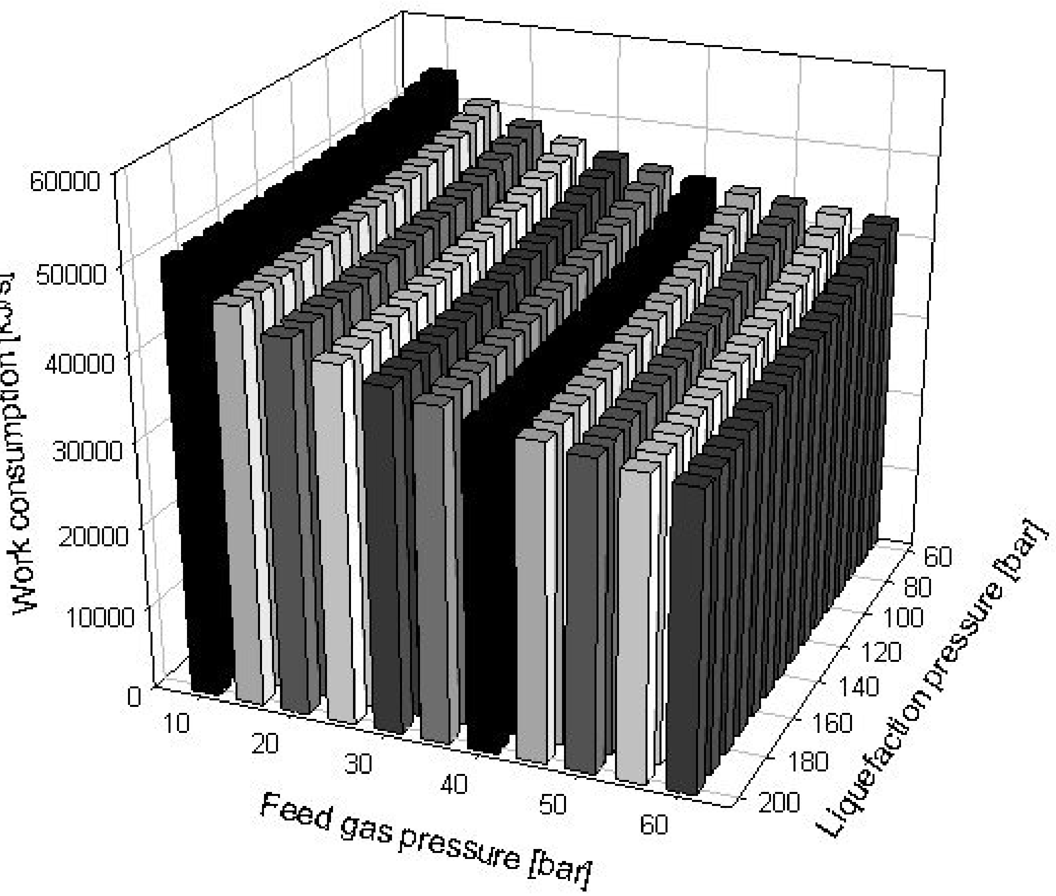


Total work consumption of compression and a liquefier with an efficiency of 30%. *Feed* is pressure entering compressor after pretreatment and *Liq*. is outlet compressor pressure/liquefaction pressure.

Compression and liquefaction 30%

Feed [bar]	10	15	20	25	30	35	40	45	50	55	60
Liq. [bar]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]
60	53675	50572	48540	47063	45922	45005	44249	43611	43065	42592	42177
70	53059	49843	47732	46192	45000	44040	43246	42574	41998	41497	41057
80	52591	49279	47096	45502	44265	43267	42438	41737	41134	40608	40146
90	52231	48833	46587	44944	43667	42634	41776	41048	40420	39873	39390
100	51955	48476	46175	44487	43174	42110	41225	40472	39823	39255	38754
110	51739	48189	45837	44108	42761	41670	40759	39985	39315	38729	38211
120	51572	47956	45556	43792	42414	41296	40363	39568	38880	38277	37743
130	51442	47765	45322	43523	42117	40974	40020	39207	38502	37883	37335
140	51341	47607	45124	43294	41861	40696	39723	38891	38170	37537	36975
150	51264	47479	44956	43096	41639	40452	39460	38612	37876	37229	36655
160	51203	47369	44811	42923	41443	40236	39226	38363	37613	36953	36367
170	51157	47276	44684	42770	41268	40043	39016	38137	37375	36704	36107
180	51121	47196	44572	42633	41111	39868	38826	37934	37159	36475	35868
190	51093	47126	44472	42509	40967	39708	38651	37746	36958	36265	35647
200	51070	47063	44381	42396	40836	39561	38490	37572	36773	36070	35442

Graph for the calculations above:



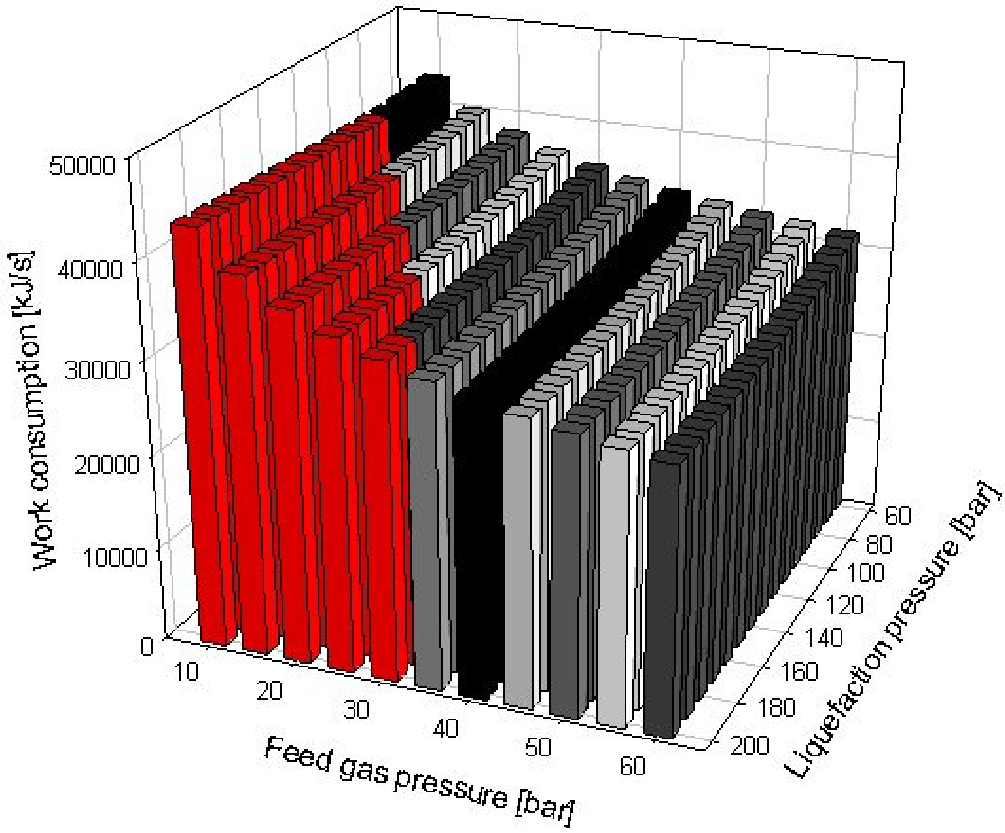
Total work consumption of compression and a liquefier with an efficiency of 40%. *Feed* is pressure entering compressor after pretreatment and *Liq.* is outlet compressor pressure/liquefaction pressure.

Compression and liquefaction 40%

Feed [bar]	10	15	20	25	30	35	40	45	50	55	60
Liq. [bar]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]
60	43131	40028	37996	36518	35377	34461	33704	33067	32521	32047	31633
70	42978	39763	37652	36112	34920	33960	33166	32494	31918	31417	30977
80	42905	39593	37410	35816	34579	33581	32753	32051	31448	30922	30460
90	42888	39489	37243	35600	34323	33291	32432	31704	31077	30529	30046
100	42912	39433	37132	35444	34131	33067	32182	31429	30780	30212	29711
110	42965	39414	37062	35333	33986	32895	31985	31210	30541	29955	29437
120	43039	39423	37023	35259	33881	32763	31830	31035	30347	29744	29210
130	43129	39452	37009	35210	33804	32661	31707	30894	30189	29570	29022
140	43230	39496	37013	35183	33750	32585	31611	30780	30059	29426	28864
150	43339	39554	37032	35172	33715	32528	31536	30688	29952	29305	28731
160	43453	39619	37061	35173	33693	32486	31476	30613	29863	29204	28618
170	43570	39689	37097	35183	33681	32457	31430	30551	29789	29117	28520
180	43689	39763	37140	35201	33678	32436	31393	30502	29726	29043	28436
190	43807	39840	37186	35223	33682	32423	31365	30460	29672	28979	28361
200	43924	39917	37235	35250	33690	32415	31344	30426	29627	28923	28296

Red indicates increase in total work consumption

Graph for the calculations above:



Total work consumption of compression and a liquefier with an efficiency of 50%. *Feed* is pressure entering compressor after pretreatment and *Liq.* is outlet compressor pressure/liquefaction pressure.

Compression and liquefaction 50%

Feed [bar]	10	15	20	25	30	35	40	45	50	55	60
Liq. [bar]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]	[kJ/s]
60	36804	33701	31669	30192	29051	28134	27378	26740	26194	25721	25306
70	36930	33715	31604	30064	28872	27912	27117	26446	25869	25368	24929
80	37094	33782	31599	30005	28768	27770	26941	26240	25637	25111	24649
90	37281	33883	31637	29994	28717	27684	26826	26098	25470	24923	24440
100	37486	34007	31707	30019	28705	27642	26756	26004	25354	24787	24285
110	37700	34149	31797	30069	28722	27630	26720	25945	25276	24690	24172
120	37919	34303	31904	30139	28761	27644	26710	25915	25227	24624	24091
130	38141	34464	32021	30223	28816	27673	26720	25906	25201	24582	24034
140	38364	34630	32147	30316	28883	27719	26745	25913	25192	24559	23998
150	38585	34800	32277	30417	28960	27773	26781	25933	25197	24551	23977
160	38804	34969	32411	30523	29043	27836	26826	25963	25213	24554	23968
170	39019	35137	32545	30631	29130	27905	26878	25999	25237	24565	23968
180	39229	35304	32680	30741	29219	27977	26934	26042	25267	24584	23976
190	39435	35468	32814	30852	29310	28051	26994	26089	25301	24608	23990
200	39636	35629	32947	30962	29402	28127	27056	26138	25340	24636	24008

Red indicates increase in total work consumption

Graph for the calculations above: