NTNU
Norges teknisk-naturvitenskapelige universitet

Fakultet for naturvitenskap og teknologi
Institutt for kjemisk prosessteknologi

SPECIALIZATION PROJECT SPRING 2009

TKP 4555

PROSJECT TITLE:

SIMULATION, OPTIMAL OPERATION AND SELF OPTIMISATION OF TEALARC LNG PLANT

by

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Date: 09-12-2009

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ACKNOWLEDGEMENT

I am so grateful to my supervisor Prof Sigurd Skogestad and my co-supervisor Magnus Glosli Jacobsen for they support and encouragement they gave to me during this project. Your support, patience and wiliness to listen to me at all times is legendary. I will not forget. TUSSEN TAKK

I wish to thank above all the almighty God for the wisdom and good health He gave to me throughout this period of project work. BABA I THANK YOU.

For everyone that contributed in one way or the other to make this project a success, i am very grateful and I pray that God will reward you all.

ABSTRACT

This project report deals with the simulation, optimal operation and self optimisation of LNG process plant. TEALARC LNG process plant was simulated in UNISIM simulator and is used to demonstrate the systematic procedure for control structure design application. A systematic procedure for control structure design starts with carefully defining the operational and economic objective of the TEALRAC plant and the degrees of freedom available to fulfil them. The optimisation results and optimal operation results are discussed and tabulated in this project report. Also the specifications for modelling and simulation of the process plant are tabulated along with some of the problems encountered during the process design.

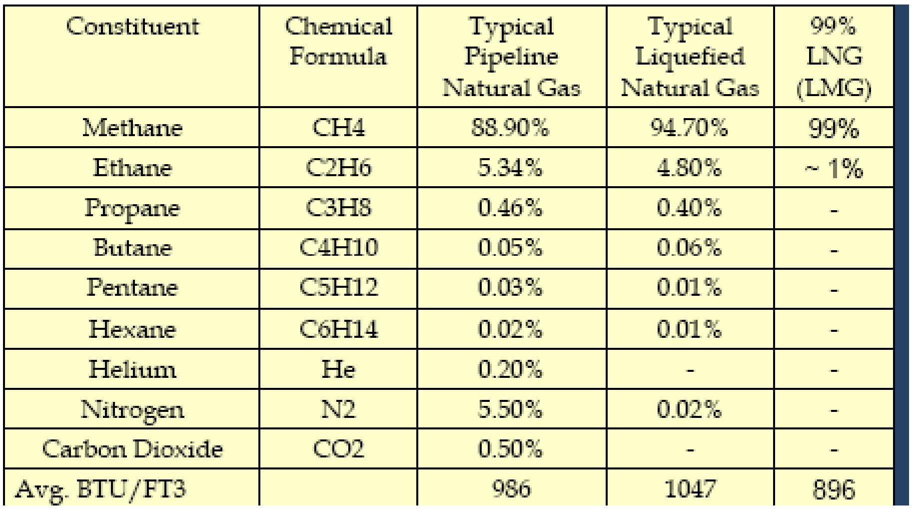
Other issues like types liquefaction processes, thermodynamics of refrigeration cycle, exergy analysis, and etc are discussed in this report.

1.0. INTRODUCTION

1.1 LIQUEFACTION OF NATURAL GAS

Natural gas consists almost entirely of methane (CH4), the simplest hydrocarbon compound. Typically, LNG is 85 to 95-plus percent methane, along with a few percent ethane, even less propane and butane, and trace amounts of nitrogen. The exact composition of natural gas (and the LNG formed from it) varies according to its source and processing history. And, like methane, LNG is odourless, colourless, noncorrosive, and nontoxic.[7]

Table 1 TYPICAL LNG COMPOSITION TABLE [7]



Constituent	Chemical Formula	Typical Pipeline Natural Gas	Typical Liquefied Natural Gas	99% LNG (LMG)
Methane	CH4	88.90%	94.70%	99%
Ethane	C2H6	5.34%	4.80%	~ 1%
Propane	C3H8	0.46%	0.40%	-
Butane	C4H10	0.05%	0.06%	-
Pentane	C5H12	0.03%	0.01%	-
Hexane	C6H14	0.02%	0.01%	-
Helium	He	0.20%	-	-
Nitrogen	N2	5.50%	0.02%	-
Carbon Dioxide	CO2	0.50%	-	-
Avg. BTU/FT3		986	1047	896

One important issue in natural gas utilisation is transportation and storage because of its low density. Natural gas is found at locations that are not economical to transport it in gaseous form to the customers. The most economical way of transporting and storing natural gas is to first liquefy the gas (Liquid Natural Gas) and then transport the LNG by ship. [5,8]

The refrigeration and liquefaction sections of an LNG plant are very important as they account for almost 40% of the capital investment of the overall plant. LNG is natural gas that has been cooled to the point it condenses to liquid, which occurs at a temperature approximately -161℃ at atmospheric pressure. Liquefaction reduces the volume of natural gas to approximately 600 times, thus, making it more economical to transport natural gas over a long distance for which pipelines are more expensive to use or which other constraints exist.

1.2 LNG VALUE CHAIN

LNG value chain as shown below consists of four components:[8]

- Gas production: this is the exploration activity done to bring natural gas from the reservoir to the industry.

- Liquefaction plant: this is the plant where the produced and treated natural gas is liquefied for storage and further use.
- Shipping: this is one major way of bringing natural gas from far distance to customers; the ships for transporting LNG are specially built to keep the liquid gas at its normal condition till the time it gets to the customers.
- Storage and regasification: natural gas is stored inside cryogenic tanks as liquid and regasified to return it to gaseous state.
- It is then delivered to customers for various uses through pipeline

Figure 1 LNG VALUE CHAIN

1.3 LIQUEFACTION PROCESSES

There are many commercial processes available for the liquefaction of natural gas, for example, single mixed refrigeration (SMR), and cascade refrigeration. In the SMR process, a mixture of hydrocarbons is used as the refrigerant rather than a pure refrigerant. The composition of the refrigerant is selected in such a way that the refrigerant evaporates over a temperature range to match the process being cooled. On the other hand, in the cascade refrigeration system, natural gas is cooled down using a cascade of refrigeration cycles. Each cycle uses a different pure refrigerant. However, the mixed refrigerant systems require careful selection of refrigerant compositions; whereas, the cascade systems are expensive to build and maintain.

In this project am concerned with the TEALARC liquefaction process, a type of single mixed refrigeration process (SMR). This project is a follow up to the work previously done at SINTEF by Finn Are. Tealarc liquefaction process flowsheet is shown in figure 2 below and is as described by paradowski and Dufresne [2].

This plant has two cooling circuits, named the liquefaction and pre-cooling gas cycles both containing refrigerants which are mixtures of components. The lower liquefaction cycle cools the natural gas in three heat exchangers. The liquefaction cooling cycle contains two compressors (C1 and C2) and a flash tank. In the flash tank, the mixed refrigerant, consisting of methane and ethane, is split in a gas and liquid fraction in order to utilize the different boiling points of the two components. The liquid fraction goes to the liquefaction heat exchanger (HE2) where the flow's pressure and thus temperature are lowered in order provide cooling to both the coolant and the natural gas flow. The gas fraction from the flash tank goes to the sub-cooling heat exchanger (HE3) where its pressure and temperature is lowered in order for it to cool itself and sub-cool the natural gas flow down to the required LNG temperature.

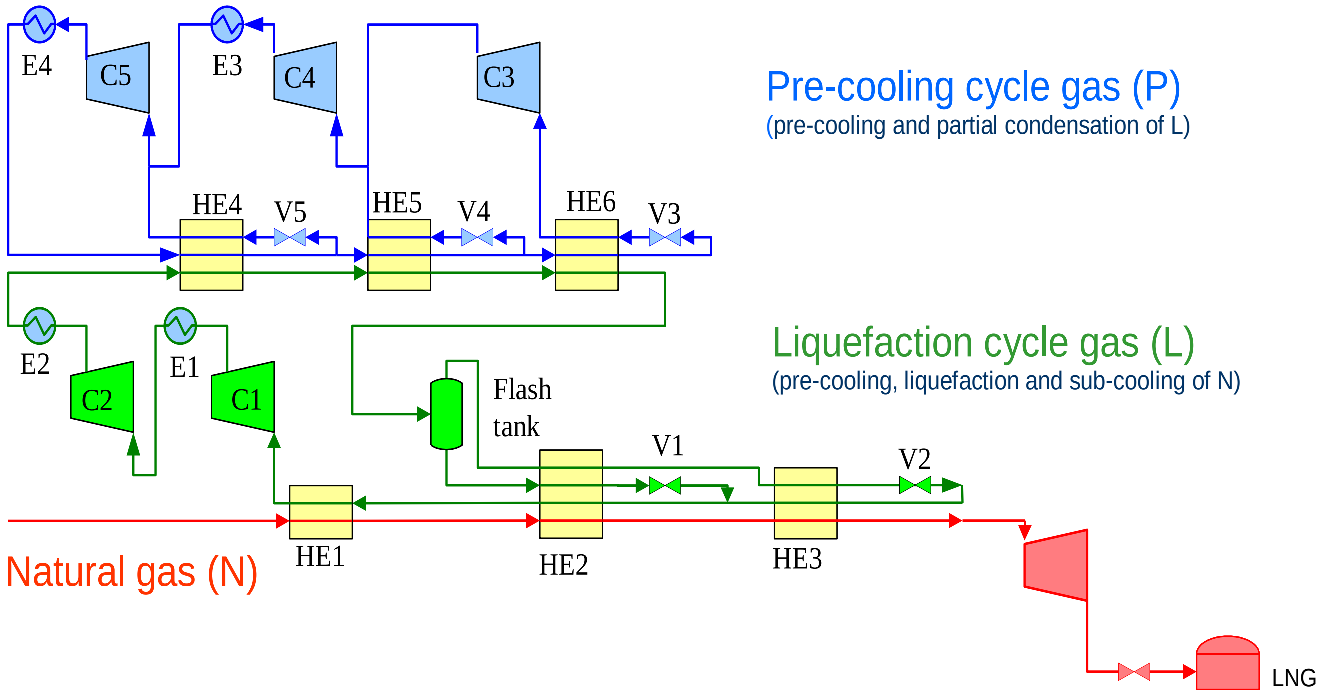


Figure 2 TEALARC LIQUEFACTION PLANT FLOW SHEET

1.4 OBJECTIVE

This project is on the simulation, optimal operation and self optimisation of LNG plant with emphasis on TEALARC liquefaction process. Hence, I will in this report present the modelling and simulation of this TEALARC process using UNISIM simulator. Also my duty will be to optimise the entire process plant applying the plant wide control method as explained in skogestad 2004[1]

However, some other issues like thermodynamics of refrigeration cycle, exergy analysis, and classification of liquefaction processes will be discussed in this report.

2. LIQUEFICATION CYCLE

LNG production for base-load consumption now has over 40 years of history starting with permanent operations of the Camel plant in Algeria in 1964. The earliest plants consisted of fairly simple liquefaction processes based either on cascaded refrigeration or single mixed refrigerant (SMR) processes with train capacities less than one million tonnes per annum (MTPA). These were quickly replaced by the two-cycle propane pre-cooled mixed refrigerant (C3MR) process developed by Air Products and Chemicals Inc. (APCI). This process became the dominant liquefaction process technology by the late 1970s and remains competitive in many cases today. [8]

The number of cycles is a key factor in the efficiency of liquefaction process. A cycle is shown in Figure 3. This cycle takes warm, pre-treated feed natural gas and cools and condenses it into an LNG product. To make the cold temperatures required for the LNG, work must be put into the cycle through a refrigerant compressor, and heat must be rejected from the cycle through air or water coolers. The amount of work (size of refrigerant compressors, drivers and refrigerant flowrate) is a strong function of liquefaction process, feed gas conditions (liquefaction temperature), and cooler temperature. In the single cycle process, there is a single working fluid that can be compressed in a single set of compressors driven by a single driver. An example of a single cycle process is a propane refrigeration system.[,7,8]

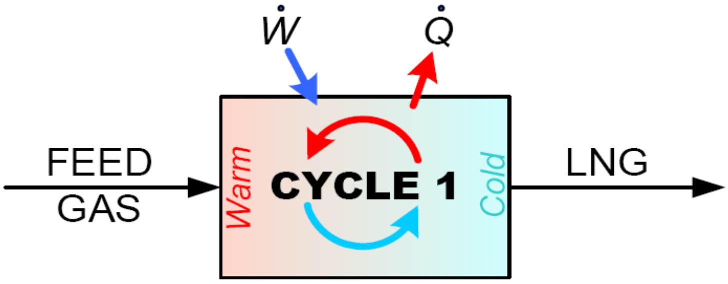


Figure 3 A SINGLE CYCLE LIQUEFACTION PROCESS

All modern base-load liquefaction facilities use either two or three cycles. The C3MR liquefaction process is a typical example of a two-cycle system. The first cycle is the propane cooling that pre-cools the mix refrigerant and feed gas process. The second cycle is the mixed refrigerant that condenses and sub-cools the natural gas to very low temperatures. Because it is a two cycle process, it requires two separate refrigerants each with their own dedicated compressors, drivers, inter and after coolers, heat exchanger, etc. Many of the liquefaction trains currently under development including RasGas, NLNG, Snøhvit, and Darwin feature three-cycle processes. Three-cycle processes include AP-X™, Shell PMR, Linde Mixed Fluid Cascade, and the ConocoPhillips Optimized Cascade. The third cycle on the AP-X™ process allows onshore train capacities to increase to approximately 7.5-10+ MTPA and have thus circumvented the typical C3MR process bottlenecks, namely the main cryogenic heat exchanger diameter and propane refrigerant compressor capacity.

A high level representation of the number of cycles in various liquefaction processes is shown below:

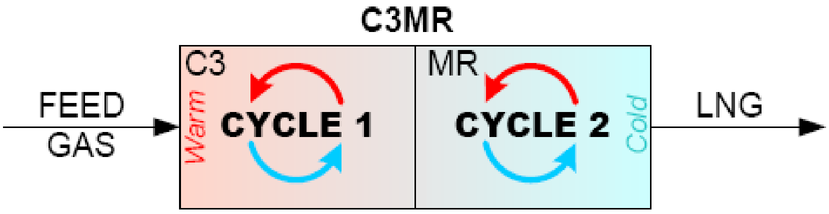


Figure 4 A TWO CYCLE LIQUEFACTION PROCESS

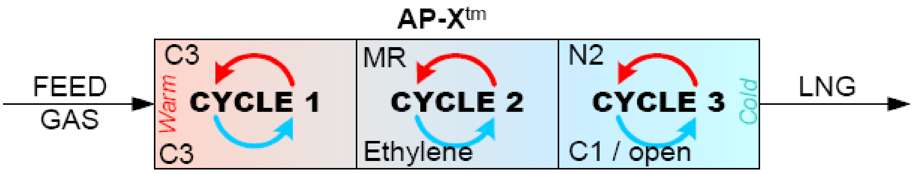


Figure 5 A THREE CYCLE LIQUEFACTION PROCESS

2.1. Classification of natural gas liquefaction processes

LNG processes can be broadly classified into three groups based on the liquefaction Process used as described in figure 3 below:

1. Cascade liquefaction processes,
2. Mixed refrigerant processes,
3. Turbine-based processes.

The first few natural gas liquefaction plants and a few current plants are based on the classical cascade processes operating with pure fluids such as methane, ethylene, and propane. Cascade processes operating with mixtures have also been recently developed. Some cascade processes are; simple cascade and enhanced cascade process by Phillip.[8]

Most existing base-load natural gas liquefaction plants operate on the mixed refrigerant processes, with the propane pre-cooled mixed refrigerant process being the most widely use. Mixed refrigerant processes can be further classified into those that use phase separators and those that do not. They can also be classified into processes that use pre-cooling and those that do not. Pre-cooling may involve refrigerant evaporation at single pressure or refrigerant evaporation at multiple pressures. Some types of mixed refrigerant processes are: LINDE liquefaction process, PRICO liquefaction process, Dual mixed refrigerant process, TECHNO-TEALARC PROCESS etc [8]

Turbine based processes have a number of advantages over both cascade and mixed refrigerant cycles. They enable rapid and simple start-up and shut-down which is important when frequent shut-downs are anticipated, such as on peak-shave plants. Because the refrigerant is always gaseous and the heat exchangers operate with relatively wide temperature differences, the process tolerates changes in feed gas composition with minimal requirements for change of the refrigerant circuit. Temperature control is not as crucial as for mixed refrigerant plants and cycle performance is more stable. Because the cycle fluid is maintained in the gaseous phase, any problems of distributing vapour and liquid phases uniformly into the heat exchanger are eliminated. Two-phase distributors are thus avoided and this, along with the small heat exchangers, results in a relatively small cold box.[7,8]

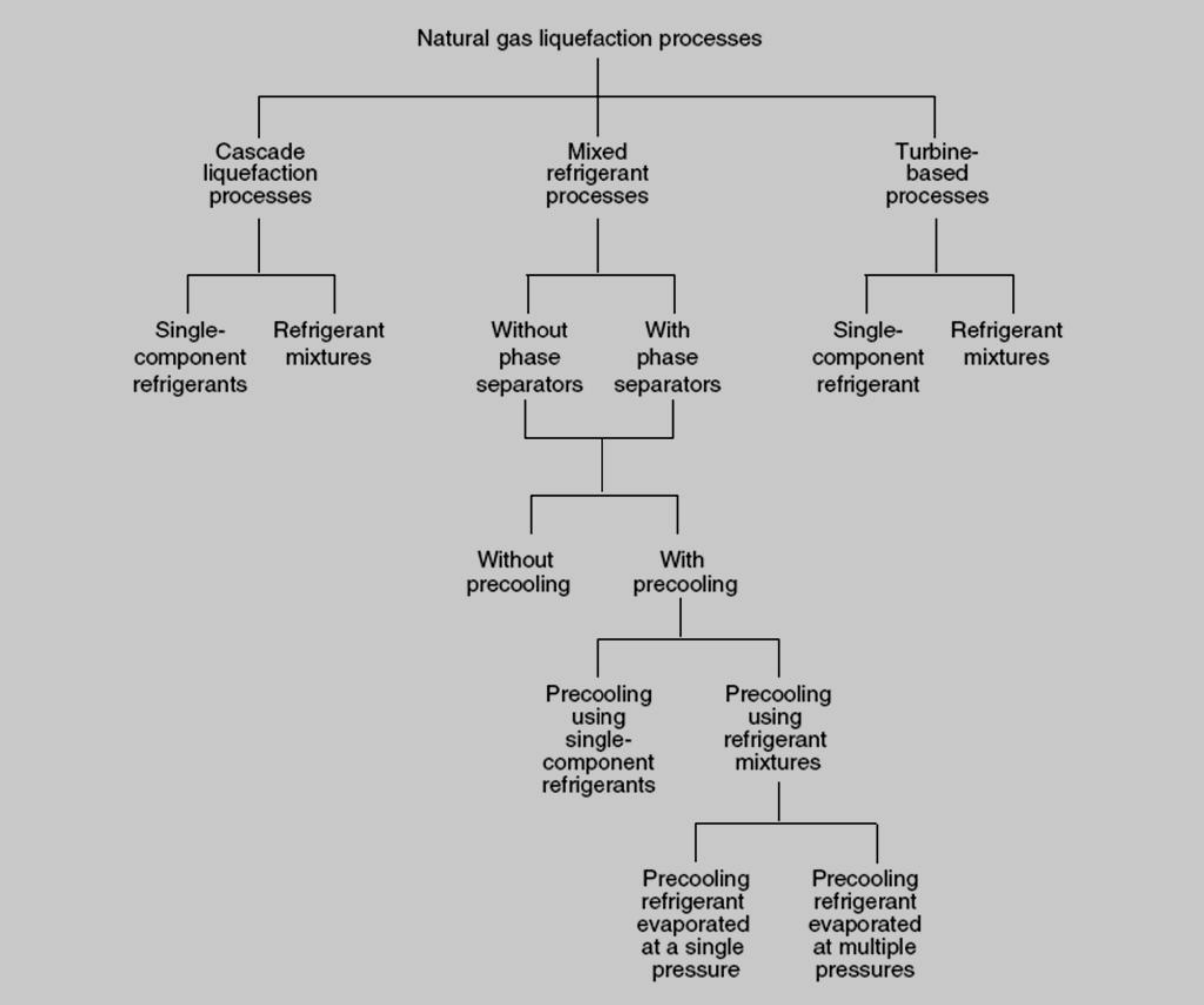


Figure 6 NATURAL GAS LIQUEFACTION PROCESS STRUCTURE

2.2 THERMODYNAMICS OF LIQUEFACTION CYCLE

The liquefaction cycle is a typical example of the refrigerator cycle, which is made up of four major components: compressor, evaporator, expansion valve and condenser. Refrigeration system removes thermal energy from a low-temperature region and transfers heat to a high-temperature region.[9] The first law of thermodynamics tells us that heat flow occurs from a hot source to a cooler sink; therefore, energy in the form of work must be added to the process to get heat to flow from a low temperature region to a hot temperature region. Refrigeration cycles may be classified as:

- vapour compression cycle
- gas compression cycle

We will examine only the vapour compression cycle. The vapour-compression cycle is used in most household refrigerators as well as in many large commercial and industrial refrigeration systems like in natural gas processing. Figure 7 provides a schematic diagram of the components of a typical vapour-compression refrigeration system.

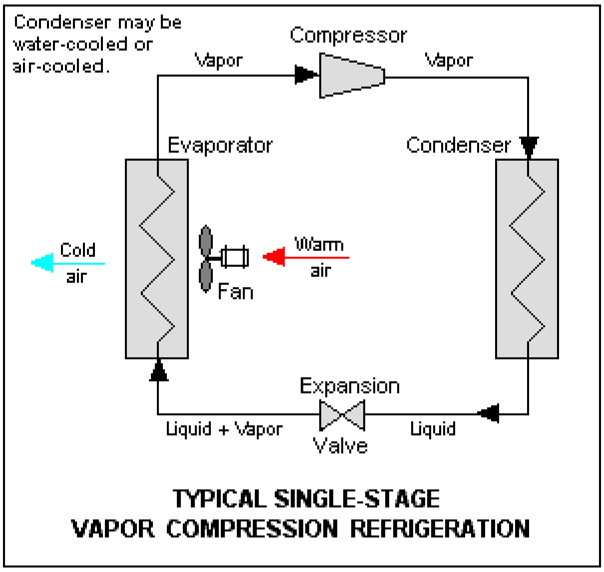


Figure 7 TYPICAL SINGLE STAGE VAPOUR COMPRESSION REFRIGERATION

Circulating refrigerant enters the compressor in the thermodynamic state known as a saturated vapour and is compressed to a higher pressure, resulting in a higher temperature as well. The hot, compressed vapour is then in the thermodynamic state known as a superheated vapour and it is at a temperature and pressure at which it can be condensed with typically available cooling water or cooling air. That hot vapour is routed through a condenser where it is cooled and condensed into a liquid by flowing through a coil or tubes with cool water or cool air flowing across the coil or tubes. This is where the circulating refrigerant rejects heat from the system and the rejected heat is carried away by either the water or the air (whichever may be the case) [9]

The condensed liquid refrigerant, in the thermodynamic state known as a saturated liquid, is next routed through an expansion valve where it undergoes an abrupt reduction in pressure. That pressure reduction results in the adiabatic flash evaporation of a part of the liquid refrigerant. The auto-refrigeration effect of the adiabatic flash evaporation lowers the temperature of the liquid and vapour refrigerant mixture to where it is colder than the temperature of the enclosed space to be refrigerated.

The cold mixture is then routed through the coil or tubes in the evaporator. A fan circulates the warm air in the enclosed space across the coil or tubes carrying the cold refrigerant liquid and vapour mixture. That warm air evaporates the liquid part of the cold refrigerant mixture. At the same time, the circulating air is cooled and thus lowers the temperature of the enclosed space to the desired temperature. The evaporator is where the circulating refrigerant absorbs and removes heat which is subsequently rejected in the condenser and transferred elsewhere by the water or air used in the condenser. To complete the refrigeration cycle, the refrigerant vapour from the evaporator is again a saturated vapour and is routed back into the compressor.[16]

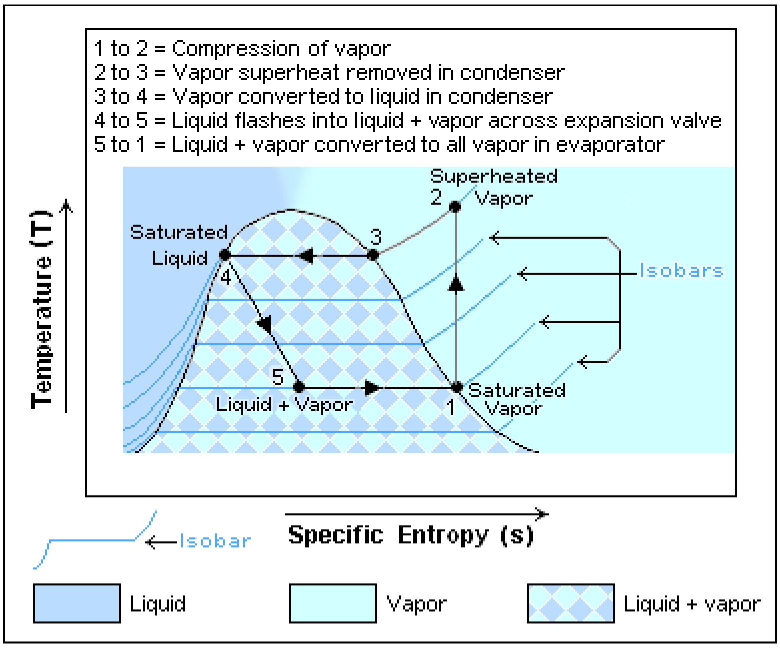


Figure 8 TEMPERATURE - ENTOPY DIAGRAM [10]

The thermodynamics of the vapour compression cycle can be analyzed on a temperature versus entropy diagram as depicted in Figure 8. At point 1 in the diagram, the circulating refrigerant enters the compressor as a saturated vapour. From point 1 to point 2, the vapour is isentropically compressed (i.e., compressed at constant entropy) and exits the compressor as a superheated vapour.[10]

From point 2 to point 3, the superheated vapour travels through part of the condenser which removes the superheat by cooling the vapour. Between point 3 and point 4, the vapour travels through the remainder of the condenser and is condensed into a saturated liquid. The condensation process occurs at essentially constant pressure.

Between points 4 and 5, the saturated liquid refrigerant passes through the expansion valve and undergoes an abrupt decrease of pressure. This decrease in pressure causes adiabatic flash evaporation and auto-refrigeration of a portion of the liquid (typically, less than half of the liquid flashes). The adiabatic flash evaporation process is isenthalpic (i.e., occurs at constant enthalpy).

Between points 5 and 1, the cold and partially vaporized refrigerant travels through the coil or tubes in the evaporator where it is totally vaporized by the warm air (from the space being refrigerated) that a fan circulates across the coil or tubes in the evaporator. One can define the coefficient of performance (COP) of a cooling cycle as

$$COP = Qc/Ws$$

Where Qc is the amount of heat removed from the ‘system’ Ws is the compressor shaft work.[10]

2.3 EXERGY ANALYSIS OF LNG CYCLE

Exergy is a measure of the maximum amount of useful energy that can be extracted from a process stream when it is brought to equilibrium with its surroundings in a hypothetical reversible process [18]. This is a thermodynamic measure defined only in terms of stream enthalpy, *H*, and entropy, *S*, for the given stream conditions relative to the surroundings. For flow sheet unit operations at steady-state conditions, the kinetic and potential energy effects are ignored. The exergy, *Ex*, or useful available work, of a stream is therefore expressed as

$$\text{Exergy} = H - ToS \qquad 1$$

Exergy analysis is useful for evaluating and improving the efficiency of process cycles. It can identify the impact of the efficiency of individual equipment on the overall process and highlight areas in which improvements will produce the most benefits. For LNG processes, To is the temperature of the ambient air or cooling water since this is the ultimate heat sink for the process. The overall change in exergy of the streams flowing in and out of an equipment unit (e.g. heat exchanger, compressor) is the amount of lost work for that unit.

$$W_{los\ t}= W_{actual}- \Delta Ex. \qquad 2$$

In order to improve cycle efficiency, lost work must be reduced [19]. For a chosen feed condition and LNG product specification, the minimum possible amount of work required to produce the LNG product is determined by the difference in the exergy of the LNG and the feed. This can be expressed as:

$$W_{rev}=\Sigma\ (H-T_oS)_{LNG}-\Sigma\ (H-T_oS)_{feed}. \qquad 3$$

Irreversibilities exist in real systems. As a result the actual work required to be input to a process to change state is more than that which would be required in the ideal case. The actual amount of work required to produce LNG is greater than the minimum reversible work in all processes studied because of the irreversibilities within the processes. The major irreversibilities in the LNG processes are due to losses within the compression (and associated after-cooling) system, driving forces across the LNG heat exchanger and other exchangers, and losses due to refrigerant letdown. World-scale optimised LNG plants require more than 2.5 times the minimum theoretical power requirements.[18].

Efficiency in the main exchanger can be improved (lost work reduced) by reducing the temperature approach between the hot and cold streams. This leads to a reduction in specific power for the liquefaction process. This can be accomplished by increasing main exchanger surface area as illustrated in the figure in the appendix of this report. As the area is increased, the specific power is reduced. However, as the minimum achievable specific power is approached, the exchanger area increases substantially, indicating that there is an economic optimum. More information on exergy analysis can be found in [18,19].

3.0 SELECTED THEORIES ON CONTROL STRUCTURE DESIGN

The focus of this project is on self optimisation, optimal operation and simulation of LNG process plant. I think some important topics should be given proper description. Control structure design for chemical plants has been the major issue we have discussed this semester and I think in a project like, it is proper to emphasise on it.

According to Skogestad (2004)[1], Control structure design which is also known as plant wide control deals with the structural decisions that must be made before we start the controller design and it involves;

- Selection of controlled variables (cv).
- Selection of manipulated variables (mv).
- Selection of measurements v (for control purposes including stabilization).
- Selection of a control configuration (structure of the controller that interconnects measurements/set points and manipulated variables).
- Selection of controller type (control law specification, e.g. PID, decoupler, LQG, etc.).

There are several procedures involved in control structure design and they are;

1. **TOP DOWN:**

- Step 1: Degrees of freedom analysis (dynamic and steady state degrees of freedom)
- Step 2: Define the optimal operation (operational objectives)
  - Cost function J to be minimised
  - Operational constraints
- Step 3: What to control? Or Self optimisation control (Primary controlled variables c = y_1. (CVs))
- Step 4: Where to set the production rate? (inventory control)

2. **BOTTOM UP**

3. Regulatory control layer with respect to stabilization and local disturbance rejection.
4. Supervisory control layer. This involves selection of decentralized or multivariable control.
5. Real time optimization layer. This includes identification of active constraints and computation of optimal set points c_s for controlled variables.

In this project am going to elaborate more on the TOP DOWN steps because it is going to be implemented in this project.

3.1. DEGREES OF FREEDOM ANALYSIS FOR OPTIMISATION

In process systems, degree of freedom is the number of variables that can be manipulated. These include everything that can be manipulated in the process like; valves, compressor power and other adjustable objects. This is usually called degrees of freedom for operations (Nvalves). According to Skogestad, (2004)[1], the number of dynamic degrees of freedom is equal the number of manipulated variables.

The number of steady-state degrees of freedom can be found by counting the manipulated variables, subtracting the number of variables that need to be controlled but which have no steady-state effect on the remaining process (e.g. liquid level in a distillation column), and subtracting the number of manipulated variables with no steady-state effect. The number of degrees of freedom for steady-state optimization (here denoted u) is equal to the number of steady-state degrees of freedom. The number of unconstrained steady-state degrees of freedom is equal the number of steady-state degrees of freedom minus the number of active constraints at the optimum.

According to Skogestad (2004)[1], the number of degrees of freedom for control, Nm, is usually easily obtained from process insight as the number of independent variables that can be manipulated by external means (which in process control is the number of number of adjustable valves plus the number of other adjustable electrical and mechanical variables).

In this project report we are concerned with the number of degrees of freedom for optimization, Nopt =. Nc =. Nu, which is generally less than the number of control degrees of freedom, Nm. We have:

$$Nopt = Nm - No \qquad (1)$$

Where: $No = Nmo + Nyo$

- No;is the number of variables with no steady state effect Nmo; is number of manipulated input (u’) with no steady state effect
- Nyo; is the number of controlled output variable with no steady state effect

However, Nyo usually equals the number of liquid levels with no steady state effect including most buffer tank levels. We should not include in Nyo any liquid hold ups that are left uncontrolled, such as internal stage holdups in distillation columns. It should also be noted that some liquid levels do have a steady state effect such as the level in a non-equilibrium liquid phase reactor, and levels associated with adjustable heat transfer areas.

Optimization is generally subject to several constraints, and the Nopt degrees of freedom should be used to:

- satisfy the constraints, and
- Optimize the operation.

We consider the case where we have a feasible solution, that is, all the constraints can be satisfied. If the number of ``active" constraints (satisfied as equalities) is denoted Nactive, then the number of ``free" (unconstrained) degrees of freedom that can be used to optimize the operation is equal to

$$N_{opt,\ free} = Nopt - Nactive. \qquad (2)$$

More details on degrees of freedom analysis for optimisation can be found in Skogestad,(2004).

3.2OPTIMISATION

The whole essence of degree of freedom analysis is to obtain the variables for optimisation. Optimization can be defined as a process of selecting the best or optimum among the whole set of options using efficient quantitative method. Optimization techniques are applied in many cases of chemical engineering. For example, in sizing and pipeline lay out, plant and equipment design, operating equipment, planning and scheduling, etc. Applying optimization in plant operation can lead to improved plant performance, reduced energy consumption, longer time between shutdowns, etc.[11]

In every optimization problem, we can always find at least one objective function, which is typically named cost, energy, or profit function. It may or not be accompanied by constraints in the form of equality, inequality, or both.[12] In a more formal way, every optimization problem can therefore be stated as:

Given: $f(\mathbf{x})$
Subject to: $ai \leq gi\ (\mathbf{x}) \leq bi,\ i = 1 \ldots m$
And $lj \leq xj \leq uj,\ j = 1 \ldots n$

Where $f(\mathbf{x})$ is the objective function and $gi\ (\mathbf{x})$ is the constraint function.
Optimization problems can be categorized into:

- *Non linear* if one or more of $f(\mathbf{x})$, g1 ….gm are non linear
- *Unconstrained* if there is no constraint functions gi and no bounds on xi
- *Bound constrained* if only xi are bounded
- *Linearly constrained* if gi are linear, while $f(\mathbf{x})$ is non linear

3.2.1Optimization Method IN UNISIM

Optimising a process in unisim requires that we choose one among several options available on the drop down list. There are five scheme options available in Original mode; BOX, SQP, Mixed, Fletcher-Reeves and Quasi Newton.[17]

- BOX method solves problems with non-linear objective functions with non-linear inequality constraints. It requires no derivatives and does not handle equality constraints. It is not considered efficient with respect to number of feval (function evaluation). The method starts by creating complex of n+ 1 variable around feasible region then evaluates objective function on each point. New point is found through extrapolation.
- SQP is the most efficient method to solve problem with equality and inequality constraints, given a feasible initial point and small number of primary variables.
- Mixed method takes the advantage of SQP efficiency and global convergence characteristic of BOX. However, it handles only inequality constraint.
- Fletcher Reeves and Quasi Newton does not handle problem with constraints.

Table 2 OPTIMISATION METHODS IN UNISIM[17]

Method	Unconstrained Problems	Constrained Problems: Inequality	Constrained Problems: Equality	Calculates Derivatives
BOX	X	X		
Mixed	X	X		X
SQP	X	X	X	X
Fletcher-Reeves	X			X
Quasi-Newton	X			X

3.3. OPTIMAL OPERATION

When we have found the degrees of freedom for the process, is important for us to understand its further importance in the control structure design. Degrees of freedom analysis gives us the basis to define the optimal operations of the process.

According to Skogestad [3] optimal operation could be specified in terms of economic cost J [$/h] for any process which is to be minimised in this simple form:

Minimise J = (cost of raw materials) + (energy cost) - (value of product)

All the manipulated variables have associated constraints and there are constraints on the output including equality constraints on product quality and product rate.
A very important issue for optimal operation is to find the steady state degree of freedom available for optimisation.

According to (Skogestad,et al 2003)[3], the first step in optimal operation is to quantify the desired operation by defining a scalar cost function. The second step is to optimise the operation by minimising the cost function with respect to the available degree of freedom. The third step is the actual implementation of the optimal policy in the plant by the use of its control system.
Skogestad[3] went ahead to simplify optimal operation mathematically by assuming a pseudo- steady state condition. He also assumed that the optimal operation of a system can be quantified in terms of a scalar cost function 'J' which is to be minimised with respect to the available degrees of freedom $U \in R^n$ u

$$\mathrm{Min}_u\, J(x, u, d) \qquad 1$$

Subject to the constraints

$$\begin{aligned} f(x,u,d) &= 0 \\ g(x,u,d) &= 0 \end{aligned} \qquad 2$$

Where;

$f(x,u,d)$ - Process model
$g(x,u,d)$ - Other constraints in the process variables

$d \in R^n$ d - represents all the disturbances, including changes that affect the system (changes in feed), changes in the model (typically represented by changes in the function f), and changes in the parameters (prices), and changes in the specifications (constraints) that enter in the cost function and the constraints.

$X \in R^n$ - represents internal variables (states). The equality constraints (g1 =0) include the model equations, which give the relationship between the independent variables (u and d) and the dependent variable (x). The system must generally satisfy the several inequality constraints ($g_2 \leq 0$).

The scalar cost function J(u,x,d) is in many cases a simple linear function of the independent variables with prices as parameters. In many cases it is more natural to formulate the optimisation problem as a maximisation of the profit P, which may be formulated as a minimisation problem by selecting J = -P.

Inequality constraints at the optimal solution are in most cases active (ie g2 = 0). In such cases of active constraints, we separate the available degrees of freedom into the set U' necessary to satisfy the active constraints at the optimum. The value of U' and X is then a function of the remaining independent variables U and d ie X = X (U, d) and U' = U'(U,d). The cost function is then reformulated in terms of the unconstrained variables U and d. More on optimal operation can be seen in Skogestad 2000 and Skogestad et al, 2003)

3.4 WHAT TO CONTROL? (SELF OPTIMISATION CONTROL)

This is a very important step in the control structure design procedures and the answer to this question is that we need to control the variables that are directly related to optimal economic operation (primary control variables y1= C).

According to Skogestad 2004[1], we should control

- Active constraints. Active constraints consume degrees of freedom and they need to be controlled.
- Select unconstrained controlled variables so that with constant set points the process is kept close to its optimum in spite of disturbances and implementation errors (Skogestad, 2000). These are the less intuitive ones, for which they idea of self optimising control is very important.

3.4.1 SELF OPTIMISING CONTROL

.

The term self optimisation is similar to the term self regulating control, which is when acceptable dynamic performance can be achieved with no control (ie constant manipulated variables).

The concept was further explained that self optimisation control is when we achieve acceptable loss with constant set point values for the controlled variables without the need to reoptimise the process when disturbance occur. In other words, this is to choose controlled variables which characterise operation at optimum, and the value of this variable at the optimum should be less sensitive to the variations in disturbances than the optimal value of the remaining degrees of freedom. At the nominal optimum some process variables will lie at their constraints. [1]

This acceptable loss is defined in terms of economics (Economic loss) 'L' to mean the difference between the actual value of the cost function and the truly optimal value

Loss = J(u, d) – Jopt(d)

The truly optimal operation corresponds to **L = 0**, but in general should be **L>0.** This shows that the plant is operating at near optimum and a small loss of L is desired. Self optimisation control is achieved if acceptable loss results at constant set point without the need to reoptimise when disturbance occurs. The precise value of what is an acceptable loss must be selected based on engineering and economic considerations. The idea of self optimisation control is illustrated in the figure below. We can see that a loss occurs when we keep a constant set point rather than reoptimising when disturbance occurs. For the figure below it is better to keep the setpoint C_{1s} constant than to keep C_{2s} constant

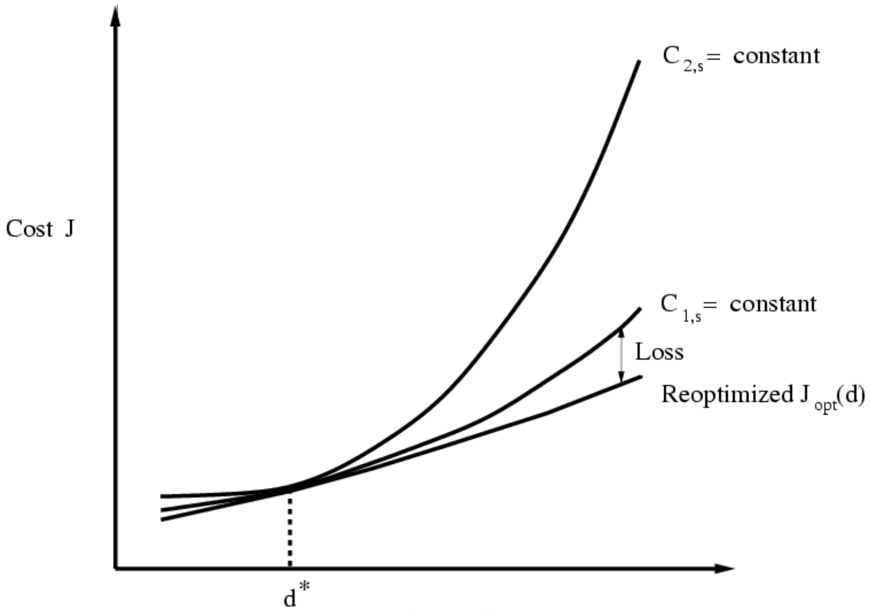
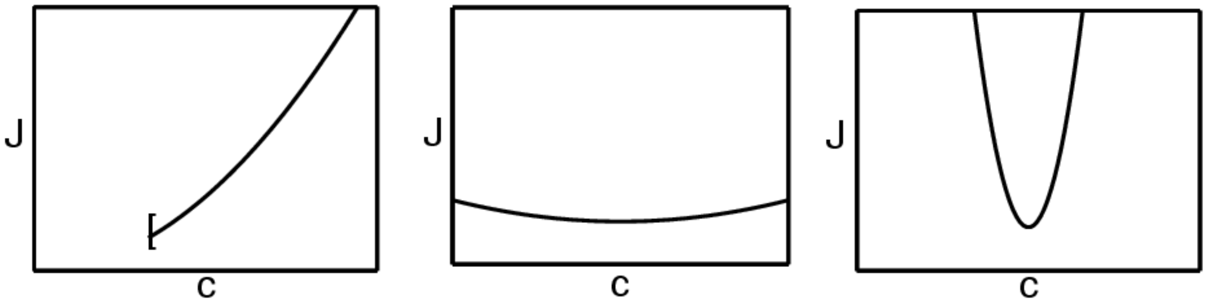


Figure 9 A TYPICAL ILLUSTRATION OF SELF OPTIMISATION CONTROL

The main point here is not to find the optimal set point but rather to find the right variables to keep constant. It is important to know that there is always a difference between constant set points **'Cs'** and actual **'C'**. This is due to implementation errors caused by measurement errors and imperfect control. To minimise the effect of the errors on the operating cost, the cost surface as a function of C should be as flat as possible. This is illustrated in figure 12 below. From this figure 11 , we were able to distinguish between three cases when it comes to the actual implementation



(a) Active constraint control: Implememtation easy

(b) Flat optimum: Implementation easy

(c) Sharp optimum: Sensitive to implementation erros

Figure 11 TYPICAL STRUCTURES OF OPTIMUM BEHAVIOURS

(A) Constrained optimum; In this case the optimal cost is achieved when one of the variables is at its maximum or minimum. There is no loss imposed by keeping the variable constant at its active constant. Implementation of an active constraint is usually easy, eg; is easy to keep a valve fully open or closed.

(B) Unconstrained flat optimum; in this case the cost is insensitive to the value of the controlled variable C.

(C) Unconstrained sharp optimum; In this case the cost operation is sensitive to the actual value of the controlled variable C and self optimising control is not possible. In this case, we would like to find another controlled variable C in which the optimum is flatter

According to Skogestad[1], self optimisation control is when near optimal operations can be achieved with **Cs** constant, in spite of disturbances ‘**d**’ and implementation error **‘n’ (n= C+ n).** In figure 4 below, you see the interaction between the local optimisation layer and the feedback control layer. The two layers interact through the controlled variables C, whereby the optimiser computes their optimal set points Cs, and the control layer attempts to implement them in practice that is to get C≈ Cs.

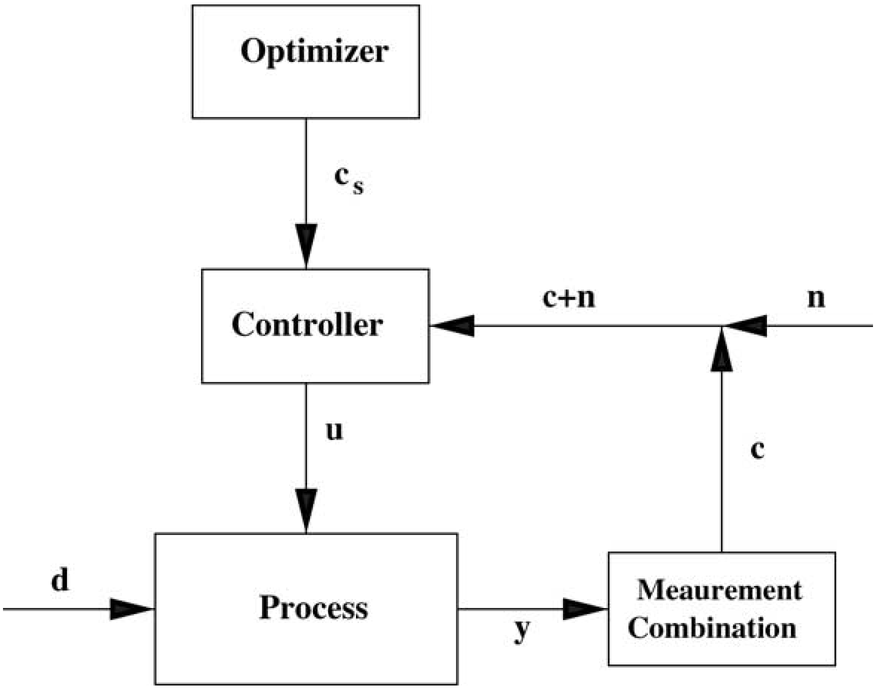


Figure 10 IMPLEMENTATION WITH SEPARATE OPTIMISATION AND CONTROL

3.5. SELECTION OF CONTROL VARIABLES:

According to Skogestad (2000), the following rules must be observed while selecting a controlled variable ‘CV’ suitable for constant set point control (Self optimisation)

- Rule 1: The optimal value for CV *c* should be insensitive to disturbances *d* (minimizes effect of setpoint error)
- Rule 2: *c* should be easy to measure and control (small implementation error *n*)
- Rule 3: *c* should be sensitive to changes in *u* (large gain |*G*| from *u* to *c*) or equivalently the optimum J_{opt} should be flat with respect to *c* (minimizes effect of implementation error *n*)
- Rule 4: For case of multiple CVs, the selected CVs should not be correlated.

The first rule minimises the effect of disturbances d. The second rule reduces the magnitude of n. The last two rules minimises the effect of the implementation error n. These four rules can be summarised by the following single rule;

Select controlled variables C for which the controllable range is large compared to the sum of optimal variation and control. (Skogestad and Postlethwaite, 1996) [13]. controllable range means the range that C may reach by varying the inputs (degrees of freedom) u, the ‘optimal variation’ is the expected variation in ‘Copt’ due to disturbance and the control error is the implementation error n.

There are some quantitave methods for control variable selection and they are explained below;

3.5.1. SELECTION OF CONTROL VARIABLE C BY MINIMUM SINGULAR VALUE RULE

Select controlled variables **c** such that we maximize the minimum singular value of the scaled gain matrix **G** (from **u** to **c** ; here **u** 's are the "original" degrees of freedom). This requires that the candidates **c's** have been scaled with respect to their span. This can be represented as;

$$Gs = S_1 GJuu^{-1/2} \qquad (1)$$

$$S_1 = \text{diag}\ [1/\text{span}\ (c_i)\ ] \qquad (2)$$

$$\text{Span ci} = [n_i^c\ ] + [\Delta ci,\ \text{opt}\ (d)] \qquad (3)$$

Span is the sum of optimal variation plus control error. Worst case loss is given below;

$$L_{\max} = \max_{\|e^1{}_c\| \le 1} \frac{1}{2} \| z \|_2^2 = \frac{1}{2(\sigma(S_1 G J_{uu}^{-1/2}))^2} \qquad (4)$$

The full derivation of this rule is given by Halvorsen et al. (2003). Although this rule is not exact, especially for plants with an ill-conditioned gain matrix like distillation columns,, it is very simple and it works well for most processes (Halvorsen et al., 2003).

Maximum gain rule has some limitations like;

- It may be impossible to calculate the value of Juu as a scalar times unitary matrix but this limitation can be corrected by defining $S_2 = Juu^{-1/2}$
- Maximum gain rule assumes that the worst case setpoint errors Δci,opt (d) for each CV can appear together. In general Δci,opt (d) are correlated.
- These limitations can lead to sub optimal set of CVs.

Hence, to overcome these limitations we use the **exact local method**.

3.5.2. SELECTION OF CONTROL VARIABLE C BY EXACT LOCAL METHOD

The exact local method was presented by Halvorsen et al. (2003). This method utilizes a Taylor series expansion of the loss function, and the exact value of the worst case loss is:

$$L_{\max} = \max_{\|e^1{}_c\| \le 1} L = \frac{(\sigma([M_d M_{n^y}]))^2}{2} \qquad (5)$$

Where: (6)

$$M_d = J^{1/2}{}_{uu}\ (J^{-1}uu\ Jud - G{-}1Gd)W_d$$

$$Mn^y = J^{1/2}{}_{uu}\ G^{-1}We \qquad (7)$$

Md represents the loss in the primary variables caused by disturbances and Mny represents the loss caused by implementation error. The magnitude of the disturbances and implementation error enter into the diagonal matrices Wd and We, respectively. The steady-state gains G and Gd and the second order derivatives Juu and Jud may be obtained numerically by applying small perturbations in the inputs u.

3.5.3. SELECTION OF CONTROL VARIABLE C BY MEASUREMENT COMBINATION METHOD

Another option is to use the linear combinations of measurements

$$c = Hy, \qquad (8)$$

where; *H* is a linear combination rather than a selection matrix. Y is the available measurements, including the inputs u used by the control system.
The goal of using several measurements is to further reduce the effect of disturbances and implementation errors. For a given set of measurements (*ys*), the combination matrix *H* can be evaluated in two different ways:

3.5.3.1 OPTIMAL METHOD:

In this method, we write the linear model in terms of the measurements y as;

$$y = G^y u + G^y d \qquad (9)$$

Locally, the optimal linear combination is obtained by minimizing σ([Md Me]) in (2) with We = HWny , where Wny contains the expected measurement errors associated with the individual measured variables; see Halvorsen et al.(2003). Note that H enters (2) indirectly, since G = HGy and Gd = HGy d depend on H. However, (2) is a nonlinear function of H and numerical search-based methods need to be used.

3.5.3.2 NULL SPACE METHOD

According to Alstad and Skogestad (2007)[4], null space method is the simpler method for finding H where we neglect the implementation error, i.e., Me = 0.Then, a constant set point policy (c = r) is optimal if Copt (d) is independent of d, that is, when Copt = 0 *d in terms of deviation variables. Note that the optimal values of the individual measurements yopt still depend on d and we may write

$$yopt = Fd \qquad (10)$$

where F is the optimal sensitivity of y with respect to d. We would like to find c = Hy such that Copt = Hy_{opt} = HFd = 0 * d for all d. To satisfy this, we must require HF = 0 or that H lies in the left null space of F. This is always possible, provided

$$n_y \geq n_u + n_d. \qquad (11)$$

This is because the null space of F has dimension $n_y = n_d$ and to make HF = 0, we must require that

$$nz = nu < ny - nd. \qquad (12)$$

This shows $Md = 0$ when $Hf = 0$. If there are too many disturbances, i.e. $ny < nu+nd$, then one should select only the important disturbances (in terms of economics) or combine disturbances with a similar effect on y (Alstad, 2007).

In the presence of implementation errors, even when $HF = 0$ such that $Md = 0$, the loss can be large due to non-zero Me. Therefore, the null space method does not guarantee that the loss L using a combination of measurements will be less than using the individual measurements. One practical approach is to select first the candidate measurements y, whose sensitivity to the implementation error is small (Alstad, 2007).

4.0 MODELLING AND SIMULATION OF TEALARC LNG PROCESS PLANT

The following chapter describes the modelling and simulation of Tealarc LNG process plant with major emphasis on:

- Process description
- Modelling of some important unit operations
- Specifications made for the entire process plant.

4.1 Process description

Tealarc LNG process plant is shown in figure below. It has two sections:

- Pre cooling cycle
- Liquefaction cycle

In this project, I tried to simulate and optimise each part of the Tealarc process plant separately.

4.1.1PRE-COOLING CYCLE

This pre-cooling cycle primarily consists of propane and ethane gas which is cooled from a temperature of about 86℃ to -64℃ at a five pressure level. It is also used to cool and partially liquefy the mixed refrigerant (MR) coming from the Liquefaction cycle to a low temperature of about -62℃. The cooling of this refrigerant is achieved by the help of the three heat exchangers (PLNG100; PLNG101; PLNG102). In the heat exchangers, the pre-cooling gas boils and evaporates in the shell side of the heat exchanger while the compressed and cooled mixed refrigerant flows in the immersed tube side. There is indirect heat exchange between the two and as a result pre cooling and cooling takes place. The centrifugal compressors recover the evaporated pre-cooling gas stream and compress the vapour to about 2260KPa at three pressure level. It is further condensed against water or air and recycled back to the heat exchangers to continue the circulation.

In the mixed refrigeration cycle (MR), the compressed and pre-cooled MR from the liquefaction cycle is partially liquefied and cooled to a temperature of about -62℃. This is made possible by the interaction in the heat exchangers between the pre-cooling gas stream and the MR stream. In the heat exchangers the MR is cooled, partially liquefied and reduced to a pressure of about 2400KPa. This cooled and partially liquefied MR is flashed at constant pressure resulting in cooled refrigerant liquid and cooled refrigerant vapour. This separated component is recycled back to the liquefaction cycle through the second heat exchanger where it is sub-cooled and it helps in sub-cooling the feed natural gas. This pre-cooling process flow sheet is shown in figure 11 below:

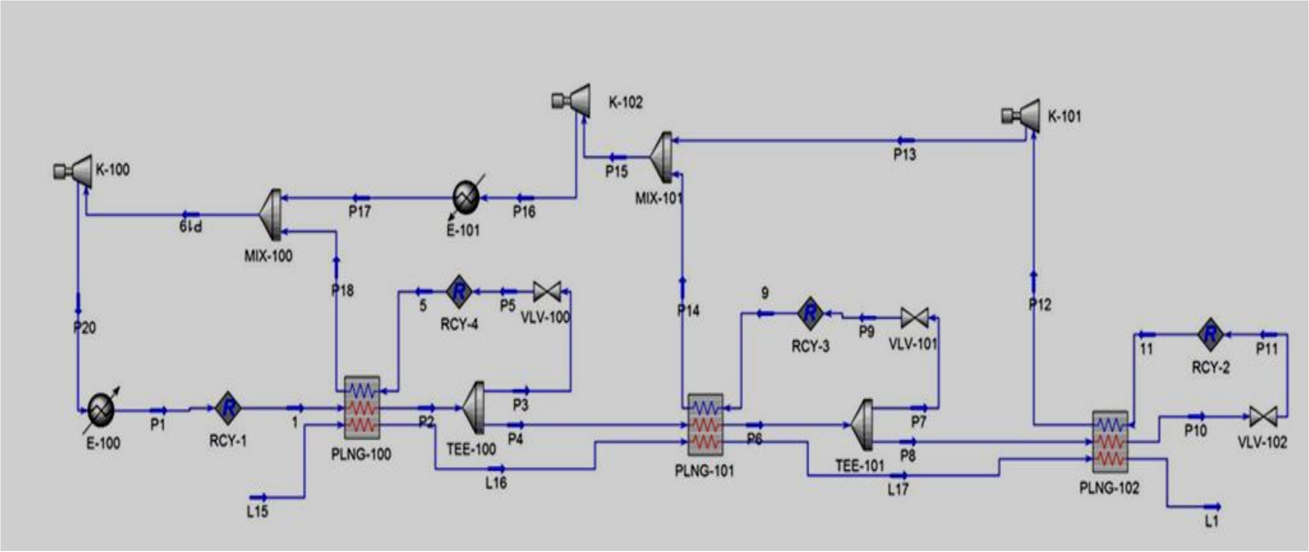


Figure 11 TEALARC PRE-COOLING FLOWSHEET

4.1.2 Liquefaction cycle

The liquefaction cooling cycle consists of three heat exchangers (LNG100; LNG101, LNG102), two compressors and a flash tank as you can see in the figure 12 below. In the flash tank as said above the mixed refrigerant from the pre-cooling cycle is split into a vapour and liquid fraction in other to utilise the different boiling points of the two components. These two fractions of the mixed refrigerant stream is further cooled in the second heat exchanger (LNG101) and then reduced in pressure and vaporized by heat exchange while cooling the feed gas and low level mixed refrigerant.

Treated natural gas is feed into the first heat exchanger (LNG 100) where, by indirect heat exchange between the mixed refrigerant and the natural gas stream, the temperature of the feed natural gas is cooled from 30℃ to about -52℃ while that of the mixed refrigerant is superheated from about -56℃ to about -4℃. This superheated mixed refrigerant is compressed by compressors (K102 and K101) and subsequently cooled before it is recycled back to the pre-cooling plant. The feed natural gas is further sub-cooled in the second heat exchanger by indirect heat exchange between the recycled, cooled and partially liquefied mixed refrigerant.

At the third heat exchanger, the sub-cooled natural gas is further cooled and liquefied at it's at a temperature of -157℃, which is about the boiling point temperature of methane gas. The pressure of the natural gas stream is further reduced to the atmospheric pressure with the help of the expander (K101) thereby reducing the temperature to about -163℃. The liquefied natural gas is flashed and stored in the LNG tank. The process is a continuous cycle.

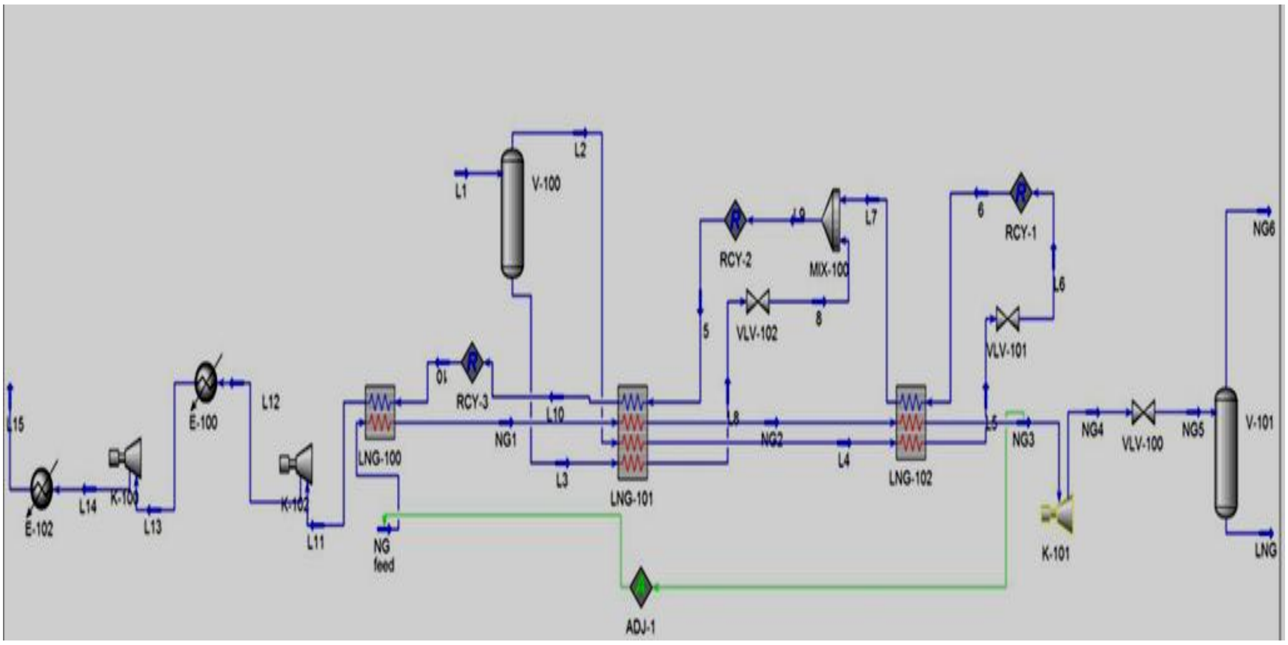
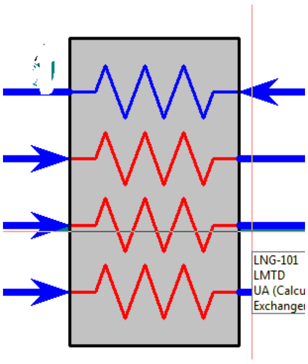


Figure 12 TEALARC LIQUEFACTION FLOWSHEET

4.2 MODELLING AND SIMULATION OF THE UNIT OPERATIONS

Generally, when simulating a process plant like this, one must model some important units of operations like; heat exchangers, compressors/expanders, coolers and valves. This steady state model of the Tealarc LNG process plant was built by selecting from UNISIM library the proper unit operation models, describing the heat and mass transfer taking place during the process, defining the streams, characterising heat exchangers, and other plant equipment and choosing the proper set of independent variable. SRK (Soave-Redlich-Kwong) fluid package was used for the thermodynamics calculations in this process model. Some of these modelled unit operations are:

4.2.1LNG Heat exchangers



LNG HEAT EXCHANGER

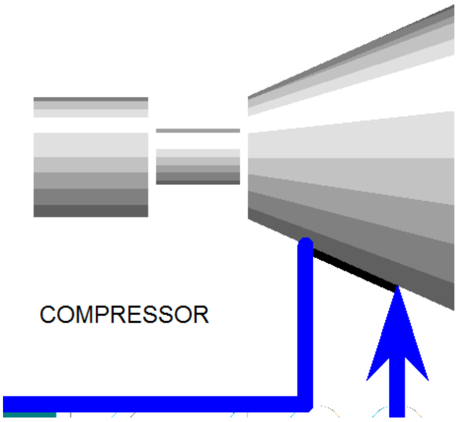
In modelling LNG Heat exchangers in UNISIM, you are expected to either specify the heat loss or allow the heat loss calculated from specified values (overall U value or ambient temperature). The heat transfer area (A) and minimum temperature approach are calculated by UNISIM. The heat loss is calculated using;

$$Q = UA(T-T_{amb})$$

I made sure that the pressure drop between inlet and out let streams are specified so that the necessary degrees of freedom in the heat exchanger will be achieved (Degree of freedom = 0). UNISIM sees the number of streams around the heat exchanger as variables and the number of parameters (temperature, pressure...) as constraints. Hence, for the heat exchanger to solve the degrees of freedom must be zero.

During the heat exchanger modelling I encountered some problems like Temperature cross in heat exchanger and insensitivity of heat exchanger. The problem of temperature cross means that the calculated heating curves for the different sides in the exchanger are touching or crossing each other at some point. I tried to solve this problem by making sure that the difference between the flowrate of the hot and cold stream is not much.

4.2.2 Compressor

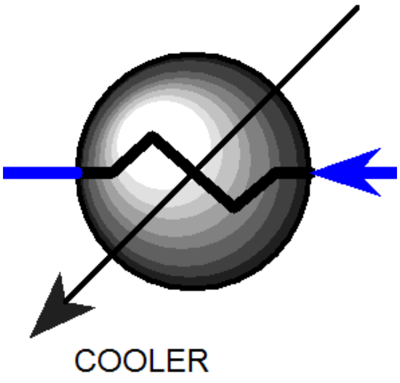


The centrifugal compressor was modelled with constant efficiency of 80%, meaning that shaft is the variable to be manipulated rather than compressor speed. I specified the outlet streams pressure and UNISIM calculated the compressor work Ws.

$$-\eta W_s = \int_{P_1}^{P_2} \frac{dp}{\rho}$$

One problem I encountered during the compressor modelling is the problem of water inlet into the compressor. In order to avoid water inlet into the compressor i made sure that the inlet stream is superheated. Superheated streams have temperature higher than there dew point temperature.

4.2.3 SW Coolers



Coolers are modelled by specifying the out let temperature and allowing UNISIM to calculate the duty Q. I also made sure that the maximum cooling temperature (30℃) of the cooler is specified for efficiency.

$$\text{SW cooler duty } Q = m\,(h_{out} - h_{in})$$

4.2.4 VALVES

For the valves, there are three different sizing methods; Cv , Cg and k methods. I selected universal gas sizing method for my model; it uses the equation below;

$$Q_{scfh} = \sqrt{520/GT}\; C_g P_1 \mathrm{SIN}\left[(59.64/C_1)\sqrt{\Delta P/P_1}\right]_{Rad.}$$

where:

$$C_1 = C_g/C_v$$

This universal gas sizing equation incorporates both the basic Cv and Cg equations into a single, dual-coefficient equation where the new factor Cf is introduced. Cf is the ratio the gas sizing coefficient and liquid sizing coefficient.

4.3 SPECIFICATIONS MADE FOR THE MODELLING AND SIMULATION OF THE ENTIRE PLANT

I made the following specifications while modelling the entire Teal arc LNG process plant and I will like to point out that I simulated the two sections of the plant sequentially(one after another). The specifications are shown in the table 4 and 5 below:

Table 3 GIVEN DATA FOR THE SIMULATION

Mixed Refrigerant composition (mol%)		Natural Gas composition(mol%)	
Methane	0.45	Methane	0.897
Ethane	0.45	Ethane	0.055
Propane	0.02	Propane	0.018
n-butane	0.00	n- butane	0.001
		Nitrogen	0.029
Nitrogen	0.08		
Pre-cooling temp -62℃		Natural gas feed pressure 4000KPa	
Pre-cooling pressure 2400KPa		Natural Gas feed temperature 30℃	

Table 4 SPECIFICATIONS FOR THE SIMULATION OF LIQUEFATION PLANT

Degrees of freedom for the design (11**DOF)**

LNG 100...........................	UA=9.6e6KJ/C-h
LNG 101...........................	UA1=1.92e8KJ/C-h
LNG 101............................	UA2=1.1e8KJ/C-h
LNG 101..............................	UA3=1.9e8KJ/C-h
LNG 102................................	UA1=2.0e7KJ/C-h
LNG 102................................	UA2= 1.2e7KJ/C-h
Choke valve (VLV-101) to set stream L6 pressure to	≥365KPa
Compressor (K-102) to set stream L12 pressure to	3000KPa
Compressor (K-100)to set stream L14 pressure to	3500KPa
Cooler (E-100) for stream L13 temperature	≥ 25℃
Cooler (E-102) for stream L15 temperature	=30.15℃

You will notice from this specification that the three heat exchangers (LNG100, LNG101.LNG102) had six specifications (UA), the choke valve had one, two coolers have two specifications and the two compressors have two specifications, making it a total of eleven specifications for the design.

Table 5 SPECIFICATIONS FOR THE SIMULATION OF THE PRE-COOLING PLANT

1.Choke valve(VLV-102) to set pressure of Stream P11.......≥170KPa
2. Choke valve(VLV-101) to set pressure of stream P9..........≥ 400KPa
3. Choke valve (VLV-100) to set pressure of stream P5........ ≥800KPa
4. SW cooler (E-101) to set temperature of stream P17. =30℃
5. Compressor (K-100) to set pressure of stream P20............. = 2260KPa
6.SW cooler (E-100) to set temperature of stream P1 =30℃
7. Molar flow rate of the split stream (P3)..... = 5.040e4kgmol/hr
8. Molar flow rate of the split stream (P7)..... = 2.262e4kgmol/hr
9.Molar flow rate of the pre cooling fluid.... = 8.733e4kgmol/hr
10.PLNG 100 UA1=3.1e7KJ/C-h
11.PLNG100 UA2=3.0e6KJ/C-h
12.PLNG101 UA1=6.3e7KJ/C-h
13.PLNG101 UA 2=2.1e7KJ/C-h
14.PLNG102 UA1=6.4e6KJ/C-h
15.PLNG102 UA2=5.1e7KJ/C-h

- PLNG is used here to represent the pre-cooling heat exchanger.
- It is 15 Degrees of freedom for the design

5.0 IMPLEMENTATION OF SELF OPTIMISING CONTROL

The theory of self optimisation as explained in chapter three of this report was implemented in optimising this process simulation. I optimised each part of the process separately and carefully followed the stepwise procedure for self optimisation as explained in Skogestad 2004[1]. Optimisation is usually carried out at optimal operation and I ensured that the simulated Tealarc LNG process plant was at its optimum before optimisation was carried out.

5.1Optimising the liquefaction section

The liquefaction process as described above is the part of the Tealarc process where the feed natural gas is cooled, sub-cooled and liquefied. In other to optimise the process I did the following;

5.1.1 Define Objective function

The objective function is to maximise the NG feed flow rate (36700kgmol/h). NG federate was taken as the objective function because we got the nominal optimal condition at low Efficiency (=Total Compressor work/NG feed rate). This implies that efficiency is inversely proportional to NG feed rate. Hence, this is represented in terms of cost function J;

$$\max_u J \quad [NOK/h]$$

$$J = q_{\ln g} v_{\ln g} - q_{energy} v_{energy}$$

Subject to

$$f(x,u,d) = 0$$
$$g(x,u,d) = 0$$

Where;

$x \in \mathbb{R}^{n_x}$ - State vector

$u \in \mathbb{R}^{n_u}$ - MVs

$d \in \mathbb{R}^{n_d}$ - DVs

$f(x,u,d)$ - Process model

$g(x,u,d)$ - Other constraints in the process variables

q_{lng} - Production rate of LNG [kg/h]

v_{lng} - Unit price of LNG

q_{energy} - Flow of consumed energy

v_{energy} - Unit price of energy

5.1.2Degrees of freedom for operation

These are the manipulating variables in the liquefaction process. There are four of them remaining after the process design because during operation heat exchangers (UA values) are fixed and there are no heat exchangers bypasses.

Table 6 DEGREES OF FREEDOM FOR OPERATIONS IN THE LIQUEFACTION PLANT

1) Choke valve (VLV-101) used to set stream L6 Pressure⩾365KPa
2) Compressor (K-102) used to set Stream L12 Pressure ≥3000KPa
3) Compressor (K-100) used to set Stream L14 Pressure ≥3500KPa
4) NG Feed rate (36700kgmol/h)

5.1.3Constraints

Constraints are those variables that consume degrees of freedom in the operation and are necessary that these constraints are identified and controlled in the process.

Table 7 CONSTRAINTS

1) Stream L6 pressure ≥ 365KPa
2) Stream L12 Pressure ≥3000KPa
3) Stream L14 Pressure ≥3500KPa
4) LNG Product Temperature = -157℃

5.1.4Disturbances

These are those variables that may have steady state effect on the objective function

Table 8 DISTURBANCES IN THE LIQUEFACTION PLANT

1) Natural gas feed composition
2) Natural gas feed pressure
3) Mixed refrigerant feed pressure (Stream L1)

5.2 Optimising the pre-cooling section

For the pre-cooling section, the same steps were taken to optimise the process and they are;

5.2.1Objective function

The objective function for this part of the process is to minimise the compressor work (Ws) and is defined in terms cost function (economics)

$$\text{Min } J \text{ [NOK/KWh]} = Wsp_{energy}$$

Subject to constraints $C \leq 0$

$$f(x,u,d) = 0$$
$$g(x,u,d) = 0$$

5.2.2Degrees of Freedom for Operation

Is assumed that there is no bypass in heat exchanger and heat exchanger UAs are fixed

Table 9 DEGREES OF FREEDOM FOR OPERATION IN THE PRE-COOLING PLANT

1) VLV-102 to set Stream P11....... 170KPa
2) VLV-101 to set stream P9..........= 400KPa
3) VLV-100 to set stream P5........ 800KPa
4) K-100 for stream P20..............= 2260KPa
5) Molar flow rate of the split stream(P3).....= 5.040e4kgmol/hr
6) Molar flow rate of the split stream(P7).....= 2.262e4kgmol/hr
7) Molar flow rate of the pre cooling fluid....= 8.733e4kgmol/hr

- The temperatures of the SW coolers were specified to their maximum cooling temperature (30°C), hence, they were not used in the optimisation process. So i am left with 7 degrees of freedom for the operation.

5.2.3 Constraints

Seven constraints were obtained during the design but we are applying four constraints in the operation since is assumed that heat exchanger value is fixed.

Table 10 CONSTRAINTS FOR THE PRECCOLING PLANT

1) Minimum temperature approach in the three heat exchangers(PLNG-100, PLNG-101, PLNG-102)= 3 constraints
2) Superheating in compressor feed streams.(P12,P14,P19) = 3 constraints
3) Temperature of stream L1 ≤ -62°C

5.2.4 Disturbances

1) Pre cooling composition

6.0 OPTIMISATION RESULTS AND DISCUSSIONS

6.1 Nominal optimum result for liquefaction plant

LNG flow rate is maximised with a fixed temperature of 30℃ after the SW cooler and 4 degrees of freedom for operation from the 11 degrees of freedom for design. The following results were obtained at optimum solution;

Table 11 NOMINAL OPTIMUM RESULT FOR LIQUEFATION PLANT

Obj. Function (Kgmol/h)	NG exit Temp (℃)	PL6 (KPa)	PL12 (KPa)	PL14(KPa)	Compressor work K-102(KJ/h)	Compressor work K-100(KJ/h)
36700	**-157**	**365**	**3000**	**3500**	**5.451e8**	**2.573e7**

6.2 ACTIVE CONSTRAINTS

Active constraints are those inequalities that turned to equalities at optimal solution. Most of the degrees of freedom are used to satisfy active constraints.[1]

After optimising the liquefaction process cycle, the following were found to be active constraints;

1. Pressure of stream L12 = 3000KPa
2. Pressure of stream L14 = 3500KPa
3. Pressure of stream L6 = 365KPa
4. Temperature of NG after cooling = -157℃

Hence, you can see from this result that there is no unconstraint degree of freedom, since all the available degrees of freedom are used to satisfy the active constraints. Natural gas feed rate was used to set the LNG product temperature at-157℃

6.3 Optimal Result with Disturbance

I considered the 3 disturbance variables in table 8 above, the disturbances were used in both maximum and minimum value. The result is shown below: bold face shows the result at the maximum disturbance value.

Table 12 OPTIMUM RESULT WITH DISTURBANCES

Optimisation Result Table at Nominal Optimal condition with disturbance d1(NG feed pressure at 3990KPa **and 4010KPa).**

Obj.Funct. (Kgmol/h)	NG exit Temp (°C	PL6 (KPa)	PL12 (KPa)	PL14(KPa)	Compressor work K-102(KJ/h)	Compressor work K-100(KPa)	Total compressor work(KJ/h
36636.7	-157	365	2999.83	3499.85	5.451e8	2.573e7	5.7083e8
36814.7	**-157.**	**365**	**2999.83**	**34999.85**	**5.442e8**	**2.573e7**	**5.6993e8**

Optimisation Result Table at Nominal Optimal condition with disturbance d2 (NG feed composition: CH_4=0.89, C2=0.05, C3= 0.028, n-C4=0.002, N_2=0.029, **CH_4=0.9, C2=0.052, C3= 0.018, n-C4=0.001, N_2=0.030)**

Obj. Funct (Kgmol/h)	NG exit Temp (°C	PL6 (KPa)	PL12 (KPa)	PL14 (KPa)	Compressor work K-102(KJ/h)	Compressor work K-100(KJ/h)	Total compressor work(KJ/h
36632.6	-157	365	2999.83	34999.85	5.432e8	2.573e7	5.6893e8
36894.2	**-157**	**365**	**2999.83**	**34999.85**	**5.465e8**	**2.573e7**	**5.7223e8**

- Bold face is the maximum value for the disturbance

Optimisation Result Table at Nominal Optimal condition with disturbance d3 (Mixed Refrigerant Pressure 2390KPa and **2410KPa)**

Obj. Funct (Kgmol/h)	NG exit Temp (°C	PL6 (KPa)	PL12 (KPa)	PL14 (KPa)	Compressor work K-102(KJ/h)	Compressor work K-100(KJ/h)	Total compressor work(KJ/h	
36700	-157	365	2999.83	34999.85	5.451e8	2.573e7	5.7083e8	
36843	-157	**365**	**2999.83**	**34999.85**	**5.460e8**	**2.573e7**	**5.7173e8**	

- Bold face is the maximum disturbance value

6.3 Optimisation result for the pre cooling part

I was able to do only one optimisation simulation for the pre-cooling part because of time constraints. The result is shown on the table below:

Table 13 OPTIMUM RESULT FOR THE PRE-COOLING PLANT

Obj.fuct(kJ/h)	Molar flow P3(kgmol/h)	Molar flow P7(kgmol/h)	Molar flow1(kgmol/h)	P5(KPa)	P9(KPa)	P11(KPa)	P20(KPa)
38.63e8	5.040e4	2.400e4	8.800e4	808	400	170	2260

6.3.1 Active constraint

Stream L1 (℃) = -61.91 was seen to be active at optimum solution.

6.4 DISCUSSION OF RESULTS

It can be seen from the optimisation result table with disturbance that the production rate was still maximised despite the influence of the disturbance variables during operation. Changing the operating parameters such as inlet gas pressure, and gas composition (particularly nitrogen content) has a big influence on plant performance. Decrease in the inlet natural gas pressure causes the decrease of Joule-Thompson effect.[8]

LNG production rate increased with increase in pressure of the NG feed and decreased with decrease in pressure of the NG feed. This is confirmed in the work done by Jørgen [16] Increase in the mixed refrigerant pressure caused a subsequent increase in the LNG production rate. This is caused by the heat exchange that takes place inside the heat exchanger between the pre cooled mixed refrigeration gas and the feed natural gas.[2]

The main energy loss of a natural gas liquefaction process exists in compressors and heat exchangers. In this paper, the matching of the heating and cooling curves between the feed gas and the mixed-refrigerant in heat exchangers was analyzed. The temperature difference and heat exchange load in heat transfer contribute to the energy loss, so large temperature difference and heat exchange load are the primary reasons of the energy loss in heat exchangers [18,19]. Sure the heat flow is not the only one affected by the changing temperature; the enthalpy and pressure also are the ones. The matching of the heating and cooling curves in heat exchangers in the liquefaction Processes was analyzed. It can be seen from the results shown in figure 14 and figure 15 below that heat loss is minimised in the heat exchangers at optimum solution. The minimum temperature approach for the three heat exchangers in both the liquefaction and pre-cooling plant (LNG100 <5℃, LNG101and LNG102<2℃), corresponds to the same value explained in [18] as the minimum temperature approach for high efficient cryogenic heat exchangers.

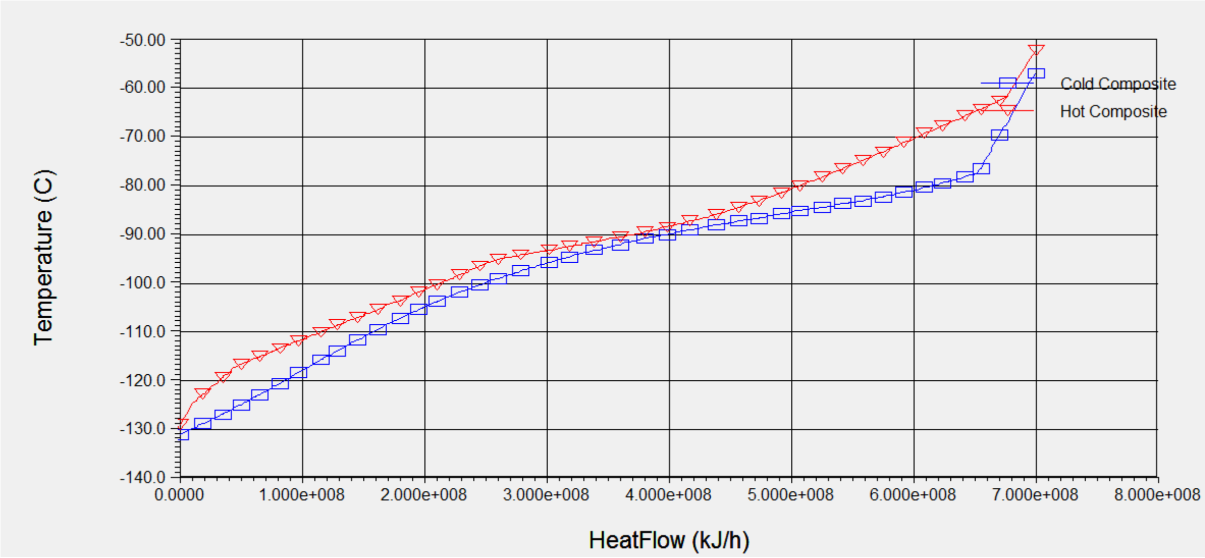


Figure 13 TEMPERATURE(C)- HEAT FLOW(KJ/h) FOR LIQUEFACTION PLANT.

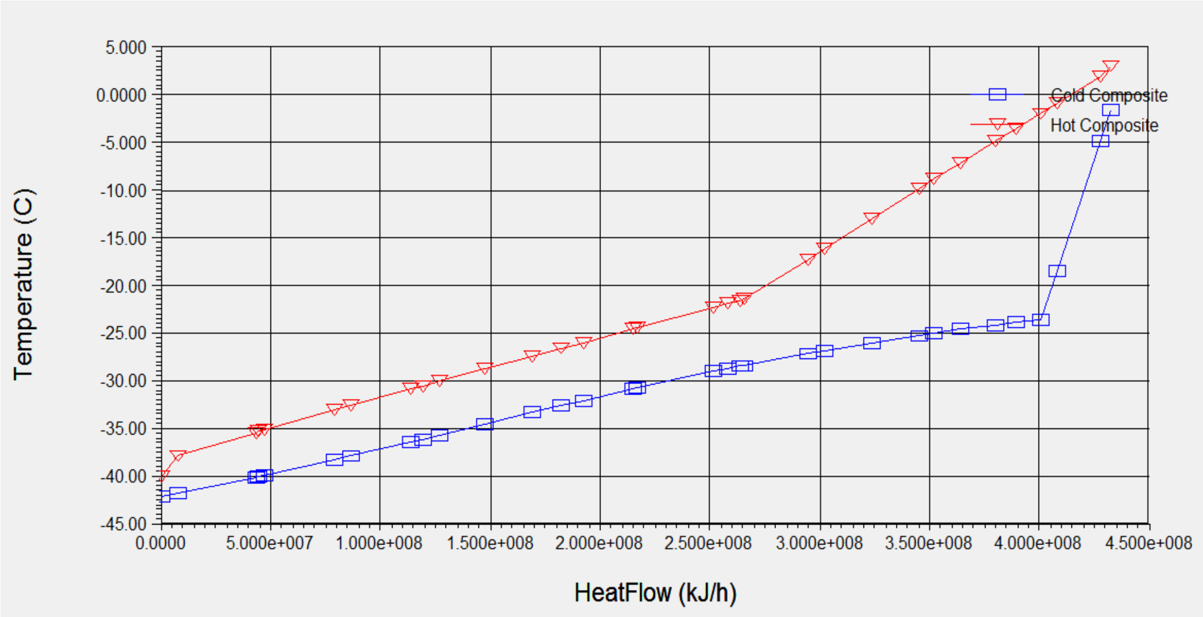


Figure 14 TEMPERATURE(C)- HEAT FLOW(KJ/H) FOR PRECOOLING PLANT

7.0 Conclusion

This project report on simulation, optimal operation and self optimisation of LNG process plant has shown the systematic implementation of Control structure/ Plant wide control to natural gas liquefaction industry. To obtain optimal solution, steady state simulation of the Tealarc LNG process plant was carried out by the help of UNISIM simulator. I simulated and optimised the pre-cooling and liquefaction part of the plant separately using the model specifications shown in table 4 and 5 in this report.

The simulated process was optimised using the plant wide control method by skogestad [1]. The optimal result in table 12 and 13 shows that the defined objectives function (cost function) for the liquefaction process was achieved subject to certain constraints and degrees of freedom. We also found out that all the degrees of freedom for operation were used to satisfy the active constraints during operation. Hence, self optimisation control could not be carried out because you need at least one unconstrained degree of freedom to perform self optimisation control.[1]

Heating curve of the heat exchangers for the pre-cooling and liquefaction plant (figure 15 and 16) shows reduced energy loss in the heat exchangers which is basically one major objective function that must be minimised in natural gas liquefaction process.[18]

8.0REFERENCES

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APPENDIX -A A TYPICAL TEMPERATURE-ENTHALPY CURVE

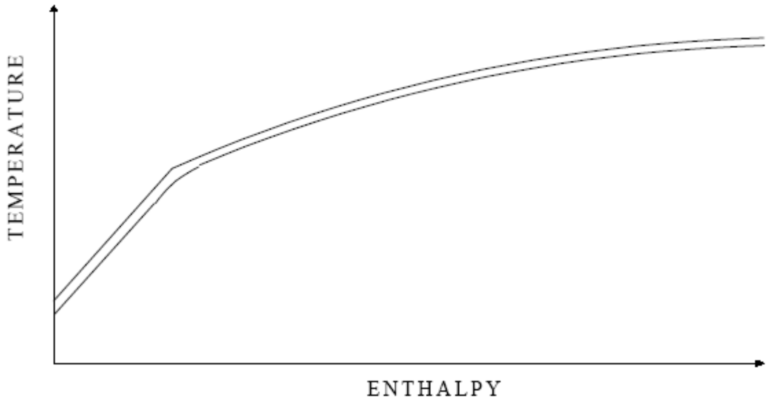
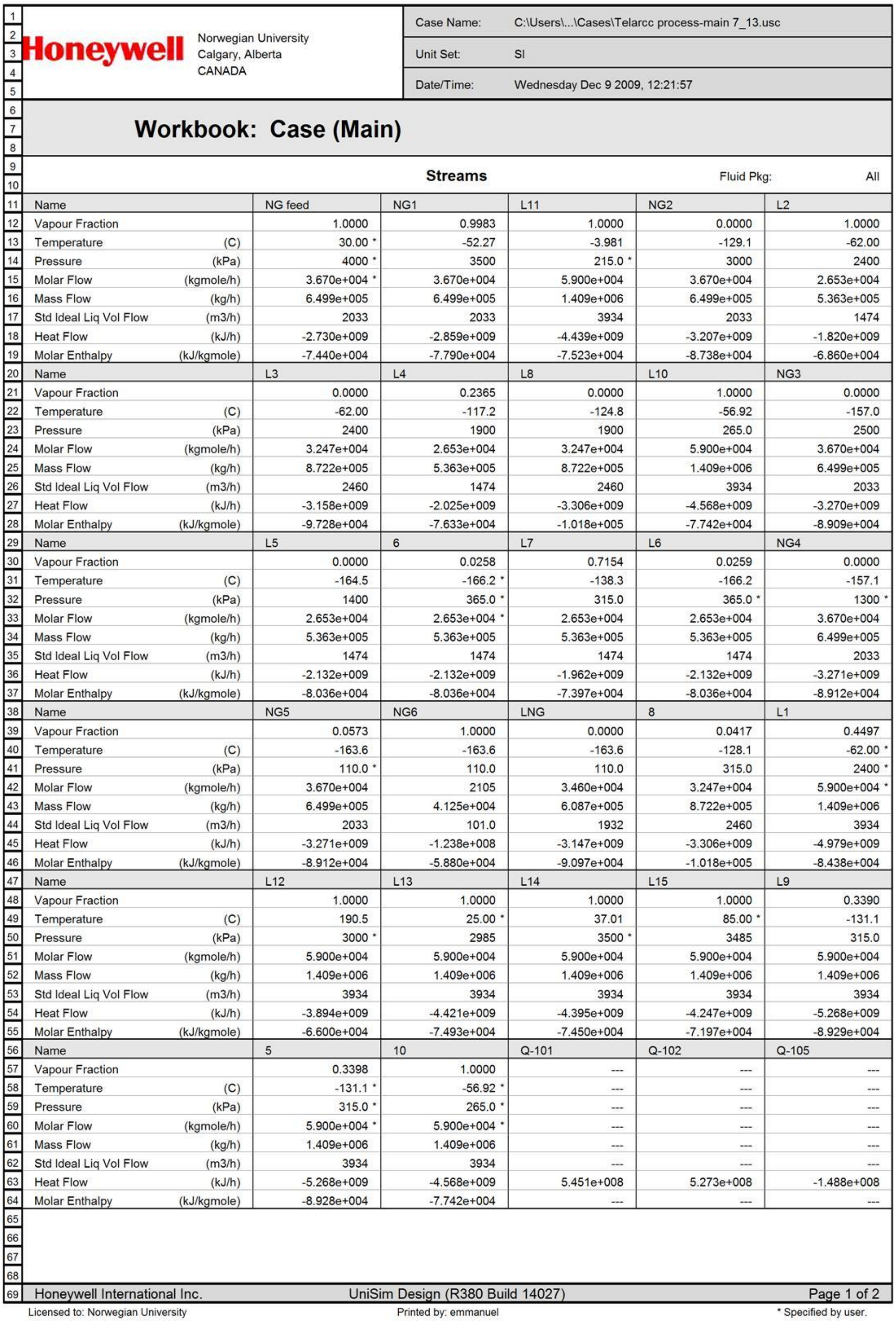


Figure 15 A TYPICAL TEMPERATURE ENTHALPY CURVE [SOURCE:15]

APPENDIX-B WORKBOOK OF ALL STREAMS AND PROPERTIES FOR BOTH LIQUEFACTION AND PRE-COOLING



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CANADA

Case Name:	C:\Users\...\Cases\Telarcc process-main 7_13.usc
Unit Set:	SI
Date/Time:	Wednesday Dec 9 2009, 12:21:57

Workbook: Case (Main)

Streams

Fluid Pkg: All

Name		NG feed	NG1	L11	NG2	L2
Vapour Fraction		1.0000	0.9983	1.0000	0.0000	1.0000
Temperature	(C)	30.00 *	-52.27	-3.981	-129.1	-62.00
Pressure	(kPa)	4000 *	3500	215.0 *	3000	2400
Molar Flow	(kgmole/h)	3.670e+004 *	3.670e+004	5.900e+004	3.670e+004	2.653e+004
Mass Flow	(kg/h)	6.499e+005	6.499e+005	1.409e+006	6.499e+005	5.363e+005
Std Ideal Liq Vol Flow	(m3/h)	2033	2033	3934	2033	1474
Heat Flow	(kJ/h)	-2.730e+009	-2.859e+009	-4.439e+009	-3.207e+009	-1.820e+009
Molar Enthalpy	(kJ/kgmole)	-7.440e+004	-7.790e+004	-7.523e+004	-8.738e+004	-6.860e+004

Name		L3	L4	L8	L10	NG3
Vapour Fraction		0.0000	0.2365	0.0000	1.0000	0.0000
Temperature	(C)	-62.00	-117.2	-124.8	-56.92	-157.0
Pressure	(kPa)	2400	1900	1900	265.0	2500
Molar Flow	(kgmole/h)	3.247e+004	2.653e+004	3.247e+004	5.900e+004	3.670e+004
Mass Flow	(kg/h)	8.722e+005	5.363e+005	8.722e+005	1.409e+006	6.499e+005
Std Ideal Liq Vol Flow	(m3/h)	2460	1474	2460	3934	2033
Heat Flow	(kJ/h)	-3.158e+009	-2.025e+009	-3.306e+009	-4.568e+009	-3.270e+009
Molar Enthalpy	(kJ/kgmole)	-9.728e+004	-7.633e+004	-1.018e+005	-7.742e+004	-8.909e+004

Name		L5	6	L7	L6	NG4
Vapour Fraction		0.0000	0.0258	0.7154	0.0259	0.0000
Temperature	(C)	-164.5	-166.2 *	-138.3	-166.2	-157.1
Pressure	(kPa)	1400	365.0 *	315.0	365.0 *	1300 *
Molar Flow	(kgmole/h)	2.653e+004	2.653e+004 *	2.653e+004	2.653e+004	3.670e+004
Mass Flow	(kg/h)	5.363e+005	5.363e+005	5.363e+005	5.363e+005	6.499e+005
Std Ideal Liq Vol Flow	(m3/h)	1474	1474	1474	1474	2033
Heat Flow	(kJ/h)	-2.132e+009	-2.132e+009	-1.962e+009	-2.132e+009	-3.271e+009
Molar Enthalpy	(kJ/kgmole)	-8.036e+004	-8.036e+004	-7.397e+004	-8.036e+004	-8.912e+004

Name		NG5	NG6	LNG	8	L1
Vapour Fraction		0.0573	1.0000	0.0000	0.0417	0.4497
Temperature	(C)	-163.6	-163.6	-163.6	-128.1	-62.00 *
Pressure	(kPa)	110.0 *	110.0	110.0	315.0	2400 *
Molar Flow	(kgmole/h)	3.670e+004	2105	3.460e+004	3.247e+004	5.900e+004 *
Mass Flow	(kg/h)	6.499e+005	4.125e+004	6.087e+005	8.722e+005	1.409e+006
Std Ideal Liq Vol Flow	(m3/h)	2033	101.0	1932	2460	3934
Heat Flow	(kJ/h)	-3.271e+009	-1.238e+008	-3.147e+009	-3.306e+009	-4.979e+009
Molar Enthalpy	(kJ/kgmole)	-8.912e+004	-5.880e+004	-9.097e+004	-1.018e+005	-8.438e+004

Name		L12	L13	L14	L15	L9
Vapour Fraction		1.0000	1.0000	1.0000	1.0000	0.3390
Temperature	(C)	190.5	25.00 *	37.01	85.00 *	-131.1
Pressure	(kPa)	3000 *	2985	3500 *	3485	315.0
Molar Flow	(kgmole/h)	5.900e+004	5.900e+004	5.900e+004	5.900e+004	5.900e+004
Mass Flow	(kg/h)	1.409e+006	1.409e+006	1.409e+006	1.409e+006	1.409e+006
Std Ideal Liq Vol Flow	(m3/h)	3934	3934	3934	3934	3934
Heat Flow	(kJ/h)	-3.894e+009	-4.421e+009	-4.395e+009	-4.247e+009	-5.268e+009
Molar Enthalpy	(kJ/kgmole)	-6.600e+004	-7.493e+004	-7.450e+004	-7.197e+004	-8.929e+004

Name		5	10	Q-101	Q-102	Q-105
Vapour Fraction		0.3398	1.0000	---	---	---
Temperature	(C)	-131.1 *	-56.92 *	---	---	---
Pressure	(kPa)	315.0 *	265.0 *	---	---	---
Molar Flow	(kgmole/h)	5.900e+004 *	5.900e+004 *	---	---	---
Mass Flow	(kg/h)	1.409e+006	1.409e+006	---	---	---
Std Ideal Liq Vol Flow	(m3/h)	3934	3934	---	---	---
Heat Flow	(kJ/h)	-5.268e+009	-4.568e+009	5.451e+008	5.273e+008	-1.488e+008
Molar Enthalpy	(kJ/kgmole)	-8.928e+004	-7.742e+004	---	---	---

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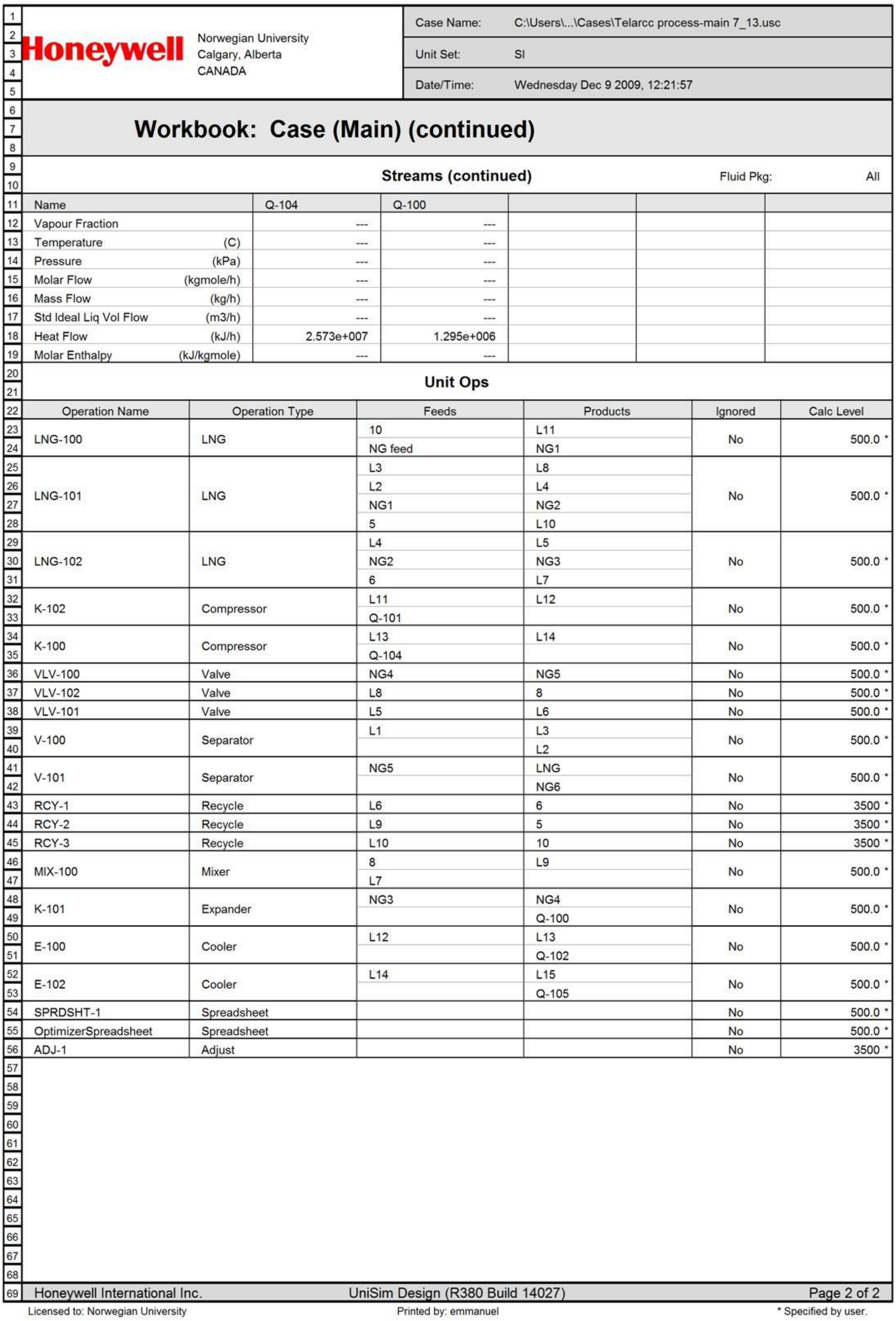

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Case Name:	C:\Users\...\Cases\Telarcc process-main 7_13.usc
Unit Set:	SI
Date/Time:	Wednesday Dec 9 2009, 12:21:57

Workbook: Case (Main) (continued)

Streams (continued)

Fluid Pkg: All

Name		Q-104	Q-100			
Vapour Fraction		---	---			
Temperature	(C)	---	---			
Pressure	(kPa)	---	---			
Molar Flow	(kgmole/h)	---	---			
Mass Flow	(kg/h)	---	---			
Std Ideal Liq Vol Flow	(m3/h)	---	---			
Heat Flow	(kJ/h)	2.573e+007	1.295e+006			
Molar Enthalpy	(kJ/kgmole)	---	---			

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
LNG-100	LNG	10 NG feed	L11 NG1	No	500.0 *
LNG-101	LNG	L3 L2 NG1 5	L8 L4 NG2 L10	No	500.0 *
LNG-102	LNG	L4 NG2 6	L5 NG3 L7	No	500.0 *
K-102	Compressor	L11 Q-101	L12	No	500.0 *
K-100	Compressor	L13 Q-104	L14	No	500.0 *
VLV-100	Valve	NG4	NG5	No	500.0 *
VLV-102	Valve	L8	8	No	500.0 *
VLV-101	Valve	L5	L6	No	500.0 *
V-100	Separator	L1	L3 L2	No	500.0 *
V-101	Separator	NG5	LNG NG6	No	500.0 *
RCY-1	Recycle	L6	6	No	3500 *
RCY-2	Recycle	L9	5	No	3500 *
RCY-3	Recycle	L10	10	No	3500 *
MIX-100	Mixer	8 L7	L9	No	500.0 *
K-101	Expander	NG3	NG4 Q-100	No	500.0 *
E-100	Cooler	L12	L13 Q-102	No	500.0 *
E-102	Cooler	L14	L15 Q-105	No	500.0 *
SPRDSHT-1	Spreadsheet			No	500.0 *
OptimizerSpreadsheet	Spreadsheet			No	500.0 *
ADJ-1	Adjust			No	3500 *

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Unit Set: SI

Date/Time: Wednesday Dec 9 2009, 12:32:16

Workbook: Case (Main)

Streams

Fluid Pkg: All

Name		L15	L16	P2	P1	P20
Vapour Fraction		1.0000	1.0000	0.0000	0.4620	1.0000
Temperature	(C)	30.15 *	1.959	3.115	30.00 *	86.26
Pressure	(kPa)	3485 *	3005	2225	2250	2260 *
Molar Flow	(kgmole/h)	5.900e+004 *	5.900e+004	8.800e+004	8.800e+004	8.800e+004
Mass Flow	(kg/h)	1.409e+006	1.409e+006	3.263e+006	3.263e+006	3.263e+006
Std Ideal Liq Vol Flow	(m3/h)	3934	3934	7549	7549	7549
Heat Flow	(kJ/h)	-4.416e+009	-4.491e+009	-9.682e+009	-8.994e+009	-8.065e+009
Molar Enthalpy	(kJ/kgmole)	-7.485e+004	-7.612e+004	-1.100e+005	-1.022e+005	-9.165e+004
Name		P4	P3	P5	P18	1
Vapour Fraction		0.0000	0.0000	0.1632	1.0000	0.4620
Temperature	(C)	3.115	3.115	-16.69	26.59	30.00 *
Pressure	(kPa)	2225	2225	808.0 *	783.0	2250 *
Molar Flow	(kgmole/h)	3.760e+004	5.040e+004 *	5.040e+004	5.040e+004	8.800e+004 *
Mass Flow	(kg/h)	1.394e+006	1.869e+006	1.869e+006	1.869e+006	3.263e+006
Std Ideal Liq Vol Flow	(m3/h)	3226	4324	4324	4324	7549
Heat Flow	(kJ/h)	-4.137e+009	-5.545e+009	-5.545e+009	-4.782e+009	-8.994e+009
Molar Enthalpy	(kJ/kgmole)	-1.100e+005	-1.100e+005	-1.100e+005	-9.489e+004	-1.022e+005
Name		P19	P17	L17	P6	P9
Vapour Fraction		1.0000	1.0000	0.7169	0.0000	0.0167
Temperature	(C)	28.05	30.00 *	-37.80	-39.99	-42.00
Pressure	(kPa)	783.0	783.0	2705	2200	400.0 *
Molar Flow	(kgmole/h)	8.800e+004	3.760e+004	5.900e+004	3.760e+004	2.400e+004
Mass Flow	(kg/h)	3.263e+006	1.394e+006	1.409e+006	1.394e+006	8.900e+005
Std Ideal Liq Vol Flow	(m3/h)	7549	3226	3934	3226	2059
Heat Flow	(kJ/h)	-8.341e+009	-3.559e+009	-4.761e+009	-4.299e+009	-2.744e+009
Molar Enthalpy	(kJ/kgmole)	-9.479e+004	-9.465e+004	-8.070e+004	-1.143e+005	-1.143e+005
Name		P14	P8	P7	L1	P10
Vapour Fraction		1.0000	0.0000	0.0000	0.4507	0.0000
Temperature	(C)	-1.533	-39.99	-39.99	-61.91	-62.99
Pressure	(kPa)	375.0	2200	2200	2400	2175
Molar Flow	(kgmole/h)	2.400e+004	1.360e+004	2.400e+004 *	5.900e+004	1.360e+004
Mass Flow	(kg/h)	8.900e+005	5.043e+005	8.900e+005	1.409e+006	5.043e+005
Std Ideal Liq Vol Flow	(m3/h)	2059	1167	2059	3934	1167
Heat Flow	(kJ/h)	-2.312e+009	-1.555e+009	-2.744e+009	-4.978e+009	-1.583e+009
Molar Enthalpy	(kJ/kgmole)	-9.632e+004	-1.143e+005	-1.143e+005	-8.437e+004	-1.164e+005
Name		P11	P12	P13	P15	P16
Vapour Fraction		0.0137	1.0000	1.0000	1.0000	1.0000
Temperature	(C)	-64.48	-42.29	2.026	-0.2424	37.41
Pressure	(kPa)	170.0 *	145.0	375.0	375.0	793.0
Molar Flow	(kgmole/h)	1.360e+004	1.360e+004	1.360e+004	3.760e+004	3.760e+004
Mass Flow	(kg/h)	5.043e+005	5.043e+005	5.043e+005	1.394e+006	1.394e+006
Std Ideal Liq Vol Flow	(m3/h)	1167	1167	1167	3226	3226
Heat Flow	(kJ/h)	-1.583e+009	-1.339e+009	-1.307e+009	-3.619e+009	-3.540e+009
Molar Enthalpy	(kJ/kgmole)	-1.164e+005	-9.844e+004	-9.610e+004	-9.624e+004	-9.415e+004
Name		11	9	5	Q-1	Q-2
Vapour Fraction		0.0135	0.0164	0.1632	---	---
Temperature	(C)	-64.48 *	-42.00 *	-16.69 *	---	---
Pressure	(kPa)	170.0 *	400.0 *	808.0 *	---	---
Molar Flow	(kgmole/h)	1.360e+004 *	2.400e+004 *	5.040e+004 *	---	---
Mass Flow	(kg/h)	5.043e+005	8.900e+005	1.869e+006	---	---
Std Ideal Liq Vol Flow	(m3/h)	1167	2059	4324	---	---
Heat Flow	(kJ/h)	-1.583e+009	-2.744e+009	-5.545e+009	2.759e+008	9.288e+008
Molar Enthalpy	(kJ/kgmole)	-1.164e+005	-1.144e+005	-1.100e+005	---	---

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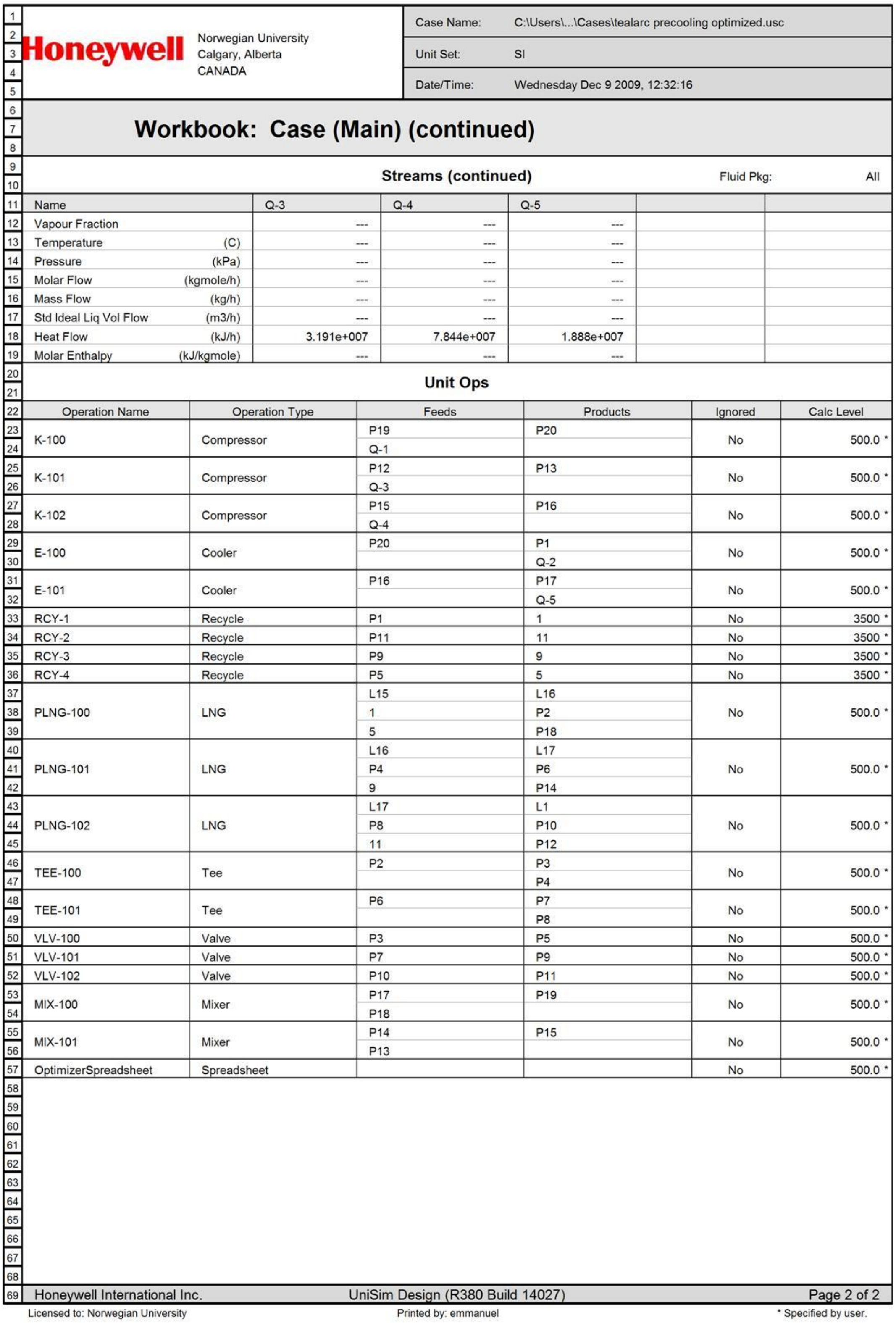

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Case Name:	C:\Users\...\Cases\tealarc precooling optimized.usc
Unit Set:	SI
Date/Time:	Wednesday Dec 9 2009, 12:32:16

Workbook: Case (Main) (continued)

Streams (continued)

Fluid Pkg: All

Name		Q-3	Q-4	Q-5		
Vapour Fraction		---	---	---		
Temperature	(C)	---	---	---		
Pressure	(kPa)	---	---	---		
Molar Flow	(kgmole/h)	---	---	---		
Mass Flow	(kg/h)	---	---	---		
Std Ideal Liq Vol Flow	(m3/h)	---	---	---		
Heat Flow	(kJ/h)	3.191e+007	7.844e+007	1.888e+007		
Molar Enthalpy	(kJ/kgmole)	---	---	---		

Unit Ops

Operation Name	Operation Type	Feeds	Products	Ignored	Calc Level
K-100	Compressor	P19 Q-1	P20	No	500.0 *
K-101	Compressor	P12 Q-3	P13	No	500.0 *
K-102	Compressor	P15 Q-4	P16	No	500.0 *
E-100	Cooler	P20	P1 Q-2	No	500.0 *
E-101	Cooler	P16	P17 Q-5	No	500.0 *
RCY-1	Recycle	P1	1	No	3500 *
RCY-2	Recycle	P11	11	No	3500 *
RCY-3	Recycle	P9	9	No	3500 *
RCY-4	Recycle	P5	5	No	3500 *
PLNG-100	LNG	L15 1 5	L16 P2 P18	No	500.0 *
PLNG-101	LNG	L16 P4 9	L17 P6 P14	No	500.0 *
PLNG-102	LNG	L17 P8 11	L1 P10 P12	No	500.0 *
TEE-100	Tee	P2	P3 P4	No	500.0 *
TEE-101	Tee	P6	P7 P8	No	500.0 *
VLV-100	Valve	P3	P5	No	500.0 *
VLV-101	Valve	P7	P9	No	500.0 *
VLV-102	Valve	P10	P11	No	500.0 *
MIX-100	Mixer	P17 P18	P19	No	500.0 *
MIX-101	Mixer	P14 P13	P15	No	500.0 *
OptimizerSpreadsheet	Spreadsheet			No	500.0 *

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