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Sensitivity Analysis of Proposed LNG liquefaction Processes for LNG FPSO

Sultan Seif Pwaga

Natural Gas Technology
Submission date: July 2011
Supervisor: Carlos Alberto Dorao, EPT

Norwegian University of Science and Technology
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Research advisor : Luis Castillo and Roy Scott Heiersted

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Problem Description

Background and objective

When selecting the right LNG FPSO process, some of the critical parameters are: different compositions over time, lean-rich gas composition (GCV), nitrogen content, thermodynamic efficiency etc. Other problems relates to: power and heat system under extreme conditions regarding high CO2 content and low feed gas pressure. For the case of LNG FPSO, there are several possible technologies available, but the best candidates for LNG FPSO based on MRC are Single Mixed Refrigerant (SMR) and Dual Mixed Refrigerant (DMR) while based on expander cycle are NICHE and dual nitrogen expander. These processes were identified in the previous project "Natural gas liquefaction processes onboard an LNG FPSO" submitted January 2011.

The main object of this project is to carry out a sensitivity analysis of the proposed liquefaction processes for LNG-FPSO and established a benchmarking of the processes considering thermodynamics, energy analysis and process design.

The following questions should be considered in the project work:

1. Developed a detail process model for each of the LNG technologies
2. Investigate the effect of quality of the feed gas composition, ambient temperature, pressures, train capacity, product specifications, etc.
3. Developed an strategy for selecting a particular technology according to the particular case scenario

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MASTER THESIS

for

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Spring 2011

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

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Department of Energy and Process Engineering, 17. January 2011

Olav Bolland
Department Head

Ivar Ertesvåg
for Carlos A. Dorao
Academic Supervisor

Research Advisors: Roy Scott Heiersted
Luis Castillo

Preface

This thesis has been written as final part of master degree program in Natural Gas Technology at Norwegian University of Science and Technology (NTNU), Norway. The thesis involve Höegh LNG and NTNU and was a continuation of semester project titled as "Natural Gas Liquefaction Process on Board an LNG-FPSO" submitted January 2011 which proposed the liquefaction processes that can be considered as good candidate for LNG FPSO. The main focus on thesis was thermodynamic analysis of the proposed liquefaction processes for LNG-FPSO and established a benchmarking of the processes considering energy analysis and process design.

I would like to thank almighty God for his immeasurable blessings and mercy for wonderful two years of stay in Norway. Second I would like to thank my supervisor associate professor Dorao, Carlos A. and my Phd supervisor Luis Castillo for their assistance from time I was doing semester project to the thesis. Also I would like to thank all individual who in one way or another contributed to the successively completion of this thesis.

Abstract

The four liquefaction processes proposed as a good candidate for LNG FPSO are simulated and evaluated. These processes include a single mixed refrigerant (SMR), dual mixed refrigerant (DMR), Niche LNG (CH_4 and N_2 process) and dual nitrogen expander. The steady state hysys simulation of the processes were undertaken to ensure that each simulated liquefaction process was compared on the identical parameters. An in-depth optimization has not been conducted but the simulation was aimed at obtaining an optimal efficient processes based on the simulated constraints.

This thesis presents the analysis of the effects of natural gas pressure, temperature and composition on the proposed liquefaction processes for LNG FPSO. During the simulations the effects were analyzed by examining specific power, power consumption and refrigerant flow rate of the proposed processes. To meet the demand of greater efficiency and large capacity for liquefaction processes, thermodynamic analysis on the liquefaction processes for LNG FPSO also has been evaluated.

The analysis of specific powers, power consumptions and refrigerant flow rates on the proposed processes shows that DMR specific power is lower than that of dual nitrogen expander by 50%, Niche LNG by 41.6% and SMR by 9.6%. The power consumption of DMR is lower that of dual nitrogen expander by 54%, Niche LNG by 47.8% and SMR by 9.6%. Also DMR has lowest refrigerant flow rate than that of dual nitrogen expander by 157.6%, Niche LNG DMR by 96.4% and SMR by 30.9%

The production capacity of simulated processes shows that DMR has higher production capacity per train of (0.91MTPA/Train) and dual nitrogen expander

has the lowest which is (0.61MTPA/Train) based on maximum duty of one LM6000 gas turbine. DMR production capacity exceeded that of dual nitrogen expander by 33%, Niche LNG by 29.7% and SMR by 8.8%.

The analysis of effect of natural gas supply temperature on the proposed processes shows that the change of natural gas supply temperature has major effect of SMR process compared to other process. The analysis shows that when natural gas supply temperature decreases from 15 to 5^0C SMR specific power and power consumption decrease by 14.99% and 15.10% respectively and when it is increases from 15 to 25^0C its specific power and power consumption increases by 39.27% and 39.19 respectively.

The analysis of the effect of natural gas supply pressure on the proposed processes shows that when natural gas supply pressure decrease dual nitrogen expander has the highest effect with specific power and power consumption increases by 22.41% and 23.25% respectively and when natural gas supply pressure increases DMR has highest effect on specific power and power consumption which are 13.06% and 13.67% respectively.

The effect of natural gas composition on the proposed natural gas liquefaction processes for LNG FPSO shows that for all proposed processes the change in natural gas composition may lead to increase or decrease of processes specific power, power consumptions or refrigerant flow rates.

Energy efficiency is important to LNG production as feed gas is consumed in order to carry out the liquefaction process .The exergy analysis of the proposed process shows that shows that DMR process has highest useful exergy about 31% compared to the other processes. Niche LNG and dual nitrogen expander has almost same useful exergy.

Contents

List of Figures

List of Tables

Nomenclature

DMR	Dual Mixer Refrigerant
FPSO	Floating Production Storage and Offloading
HHC	Heavy Hydro Carbon
HHV	High Heating Value
LMTD	Logarithimic Mean Temperature Difference
LNG	Liquefied Natural Gas
LPG	Liquefied Petroleum Gas
MFC	Mixed Fluid Cascade
MRC	Mixed Refrigerant Cycle
NGL	Natural Gas Liquid
SMR	Single Mixed Refrigerant
TCM	Trillion Cubic Meters
"UA"	Overall heat transfer multiply by Area

1. Introduction

About one-third of the world's natural gas reserves (60 TCM) are found in offshore fields [7] and 36% of the natural gas reserves are considered as stranded natural gas and most of them are found offshore [8].To realize its value it has to be brought onshore, to be processed to the required specification and sold to the transportation and distribution networks. Lack of means to bring stranded natural gas to the market leads to increase of remotely located natural gas reserves, flaring and re-injection of associated gas from offshore fields [3]. Flaring of associated gas has become an environmental issue with high degree of focus among approving authorities and oil companies. Handling of associated gas for oil developments has become a more critical issue than before. Therefore, this has led the oil and gas industry to seek solutions that can handle the associated gas in an acceptable manner both economically and environmentally. Due to the increasing demand for natural gas in the world and growing demand for cleaner energy, the pressure to bring the stranded natural gas to market is also increasing.

Companies and individuals are seeking full realization of offshore fields' potential for exploiting stranded natural gas. For many years' industry has proposed LNG FPSO as means of recovering and monetizing off-shore stranded natural gas [14]. Also many studies concluded that offshore stranded natural gas will require off-shore solution (LNG FPSO) to put the natural gas into exportable product. Industry has put attention on LNG-FPSO due to lower investment, shorter building period and being easer to move from one field to another [27]. Analyst estimated that the LNG-FPSO project might be 20-30% cheaper than a same size project at onshore and the construction time 25% faster compared to onshore plant which requires land acquisition [7, 27].

Four different natural gas liquefaction processes have been reported as suitable candidate for LNG FPSO. These processes were identified in the semester project with title "Natural gas liquefaction processes onboard an LNG FPSO" submitted January 2011[21]. The four processes considered as good candidates for LNG FPSO are single mixed refrigerant (SMR), dual mixed refrigerant (DMR), dual nitrogen expander and Niche LNG. These processes significantly differ in process designing and refrigerants. All the processes claim reasonable efficiency, minimum capital cost and offshore suitability

The main reasons for selecting SMR and DMR is its higher efficiency, being currently in operation in onshore LNG production plants and also they have higher production capacity per train. But the challenges are they are sensitive to changes in feed gas conditions as they rely on small temperature differences between the composite cooling and warming streams to give reasonable process efficiency. They also take longer time to start-up and stabilize than expander plants because of the need for precise blending of the refrigerant and also their refrigerants operate on liquid phase hence the FPSO motion will cause mal-distribution in equipments and pipelines, which will impact the performance of heat transfer.

The main reasons for selection of expander cycle are its refrigerants are in gaseous form hence it has negligible impact from vessel motion and there is no need for refrigerants storage thus increasing space availability. Expander cycle has high inherent to safety as no storage of hydrocarbon refrigerant is required and has low footprint. The major disadvantage is its relatively high power consumption.

Liquefaction process is one of the important part of LNG FPSO. There are many processes for liquefying natural gas, but few of them are in use (onshore plants) and many they do not have industrial reference. The liquefaction process cools natural gas to liquid form using various methods of cryogenic processes and also expands the liquefied natural gas to atmospheric conditions for easier and safer storage. The liquefaction processes (technologies) are divided into three main cycles; cascade cycle, mixed refrigerant cycle and expander cycle. Major factors which differentiate technologies are setup and designing but they use the same fundamentals principles [24].

Natural gas liquefaction process operates under low temperatures and the processes features one or more refrigeration cycles to remove heat from natural gas. The liquefaction process requires significant power demand for compression in refrigeration cycle so it is very important to achieve energy efficiency in designing and operation of refrigeration cycles [5, 9]. Energy efficiency is important to LNG production as feed gas is consumed in order to carry out the liquefaction process. However, the energy efficiency is not the only factor of importance, as any project must be cost effective, reliable, and tolerant of reasonably foreseeable feed changes. It must also be safely operated and maintained and for offshore, it should also be relatively compact [5].

In order to meet a number of challenges including the demand of greater efficiency and large capacity for liquefaction processes, thermodynamic analysis on the liquefaction process for LNG FPSO must be evaluated. To lower input power for natural gas liquefaction process it is crucial to reduce entropy generation due to temperature difference between feed natural gas and refrigerants flow in LNG heat exchangers. Natural gas is a mixture of different hydrocarbons and its specific heat capacity varies considerably during its liquefaction process, hence a variety of combined refrigerants are required to efficiently liquefy natural gas [5, 11, 18].

2. Theory

2.1 Liquefaction of Natural Gas

Natural gas liquefaction processes convert pre-treated natural gas into liquid suitable for transportation or storage. The liquefied natural gas (LNG) has a temperature of -162^{0}C at atmospheric pressure [22]. The liquefaction of natural gas reduces the volume of natural gas by a factor of 600 thus enabling transportation in tanks on board specialized ship [5, 22, 20].

Natural gas entering a liquefaction plant must be pre-treated to remove impurities such as water, acid gases (eg CO_2 and H_2S) and mercury to prevent freezing out in process equipment, corrosion, depositions on heat exchangers surface and controlling of heating values in the final product. Nitrogen is removed at end flash while heavy hydrocarbons may be removed at pre-cooling stage because they are valuable products as natural gas liquids (NGL), liquefied petroleum gas (LPG) and for refrigerant makeup. The compositions of Natural gas suitable for liquefaction process may contain a mixture of methane (about 85-95%), lighter hydrocarbons and small fraction of nitrogen [5].

Generally natural gas liquefaction plants consist of two main sections, pre-treatment and liquefaction. In the pre-treatment section, acidic gases (CO_2 and H_2S), water, mercury and any other impurities that may solidify when natural gas is refrigerated are removed and the liquefaction section removes sensible and latent heat from natural gas before it is expanded to atmospheric pressure [4].
A block diagram showing some of the key sections of an LNG Plant is presented in figure-1 below. The sections include gas reception (slug catcher), Pre-treatment (which consist of acid gas removal, molecular sieve dehydration and mercury removal) and liquefaction sections.

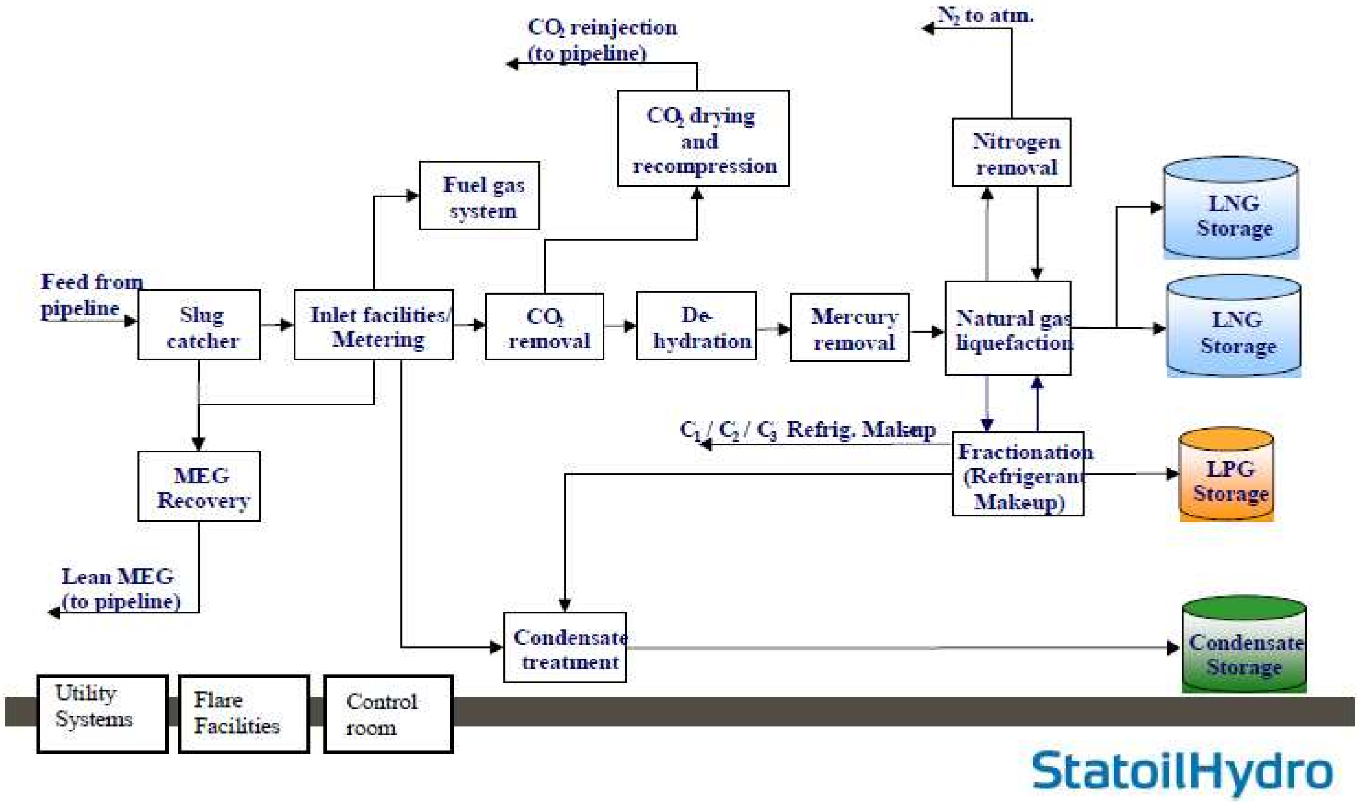

Figure 1: Principal block diagram for Liquefied natural gas plant (source [15])

2.1.1 Classification of natural gas liquefaction processes

The processes for liquefying natural gas can be classified in three groups, cascade process, mixed refrigerant liquefaction process and expander or turbine-based process. The classification of the mentioned natural gas liquefaction processes is shown in Figure-2 below.

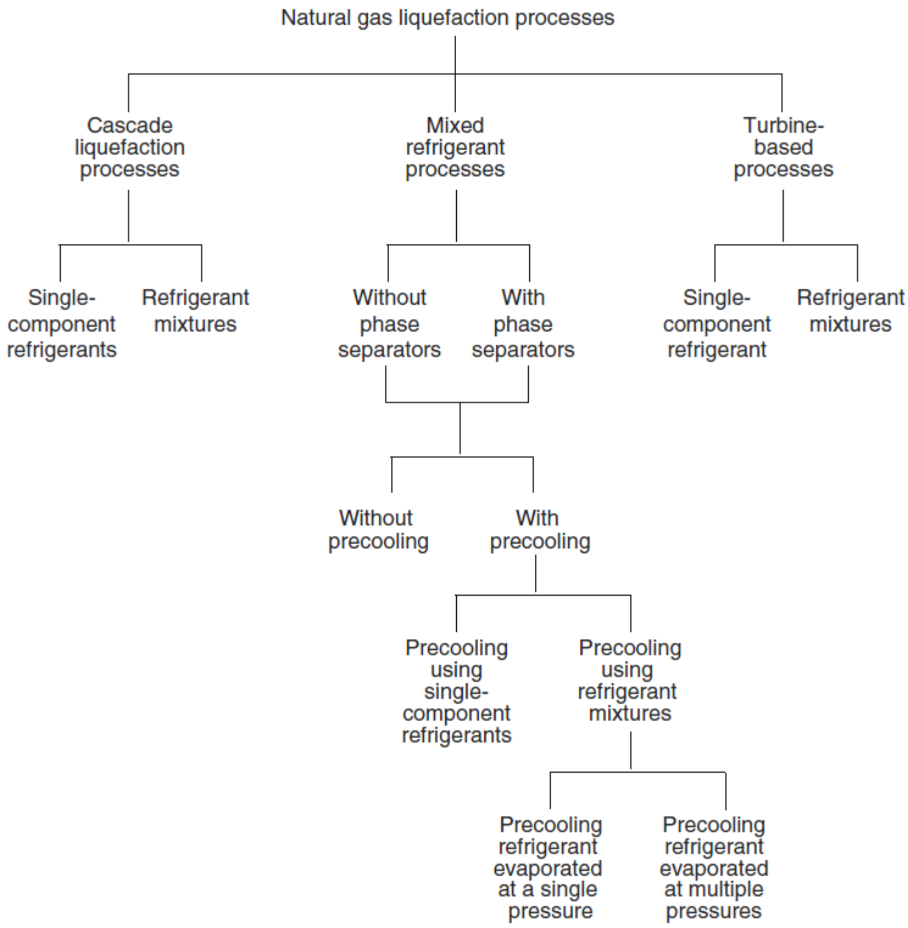

Figure 2: Classification of natural gas liquefaction processes (source [10])

The cascade process operates with pure component refrigerants which are methane, ethylene and propane. There are few LNG plants in the world operating on cascade process .The disadvantage of cascade cycle/group is its relatively high capital cost due to the number of refrigeration compression circuits which each requires its own compressor and refrigerant storage. Costs for maintenance and spares tend to be comparatively high. The main advantage of cascade cycle is its less power requirement compared to other liquefaction cycle, mainly because the flow of refrigerant is lower than in other cycles. It is also flexible in operation as each refrigerant circuit can be controlled separately. The mean temperature differences between the composite curves are wide relative to those of the mixed refrigerant cycle. Economies of scale dictate that the cascade cycle is most suited to very large train capacities where the low heat exchanger area and low power requirement offset the cost of having multiple machines [21].

The mixed refrigerant cycle (MRC) uses a single mixed refrigerant instead of multiple pure refrigerants as the cascade cycle. The mixed refrigerant normally consists of nitrogen, ethane, propane, butane and pentane. Such a mixture evaporates over a temperature trajectory instead of at a constant evaporating point and this has large benefits for the total process. The refrigeration effect will be distributed over a range of temperatures and accordingly the overall temperature difference between the natural gas and mixed refrigerant is small. Small driving temperature differences give operation nearer to reversibility; leading to a higher thermodynamic efficiency. Simultaneously, the power requirement will be lower and the entire machinery smaller [1,19] . Some of MRC technologies are; Single mixture process (SMR), Mixed refrigerant with propane pre-cooling (C3/MR), Dual mixed refrigerant process (DMR), Mixed fluid cascade process (MFC), AP-X and Small scale MRC . 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 used [10].

Expander cycle produces LNG by means of refrigeration generated by the isentropic expansion of gases used as refrigerant. There are various expander technologies, some of them use a single cycle, others use a dual expansion cycle and in other cases a pre-cooling cycle is added to improve the overall efficiency [23]. In the industry nitrogen expansion cycles have been used for low capacities LNG plant (0.02- 0.14 MTPA) especially in peak-shaving plants and also in re-liquefaction units located in very large LNG carriers [16, 23].

2.2.2 Type of LNG Plants

LNG plants can be group in three types Base-load, Peak-shaving and Small-scale plants.

Base-load plants – These are large plants which are directly linked to a specific gas field development, and serves to transport gas from the field. A base-load plant typically has a production capacity of above 3 MTPA (million tons per annum) of LNG. The main world-wide LNG production capacity comes from this type of plants [20].

Peak-shaving plants – These are smaller plants that are connected to a gas network. During periods of the year when gas demand is low, natural gas is liquefied and LNG is stored as a gas buffer. LNG is vaporized during short periods when gas demand is high. These plants have a relatively small liquefaction capacity (such as 200 tons/day) and large storage and vaporization capacity (such as 6000 tons/day). Especially in the USA many such plants exist [20].

Small-scale plants - Small-scale plants are connected to a gas network for continuous LNG production in a smaller scale. The LNG is distributed by LNG trucks or small LNG carriers to various customers with a small to moderate need of energy or fuel. This type of LNG plants typically has a production capacity below 500 000 TPA. In Norway and China several plants within this category is in operation [20].

2.2 Liquefaction and refrigeration

Liquefaction of gas is done by refrigerating the gas to the temperature below its critical conditions. Natural gas has a critical point at -80 to -90^{0}C and its liquefaction cannot be done by pressurization and expansion alone [22]. The process requires a refrigeration cycle that removes energy from natural gas in the form of sensible and latent heat. Selection of best refrigeration cycle for liquefaction of natural gas can be done after thorough study of local conditions [4].

The thermodynamic principles of liquefaction and refrigeration process are quite similar but the designing of the two systems are different. The refrigeration process/cycle for liquefaction of natural gas involves some equipment in which refrigerant is compressed, cooled to reject heat at ambient conditions and expanded to produce refrigerant capacity required. In refrigeration cycles which operate as close loop, refrigerant is constantly circulating as working fluid and there is no accumulation or withdraw of refrigerant from the cycle. The diagram showing a simple refrigeration circuit is given in figure-3 below. The system comprises of four components evaporator, compressor, condenser and throttling valve.

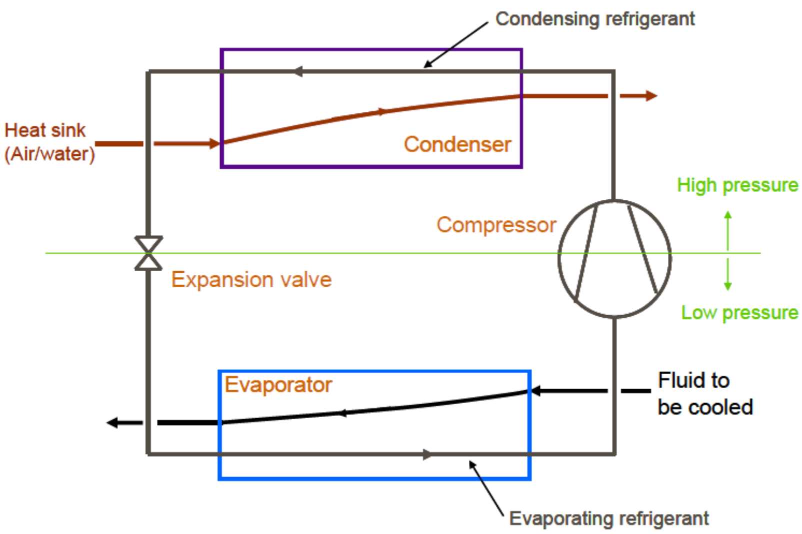

Figure 3: Flow diagram for single cycle refrigeration system (source [20])

The refrigerant is in closed circuit and circulated by compressor. By keeping the pressure of refrigerant low in evaporator, the refrigerant boil by absorbing heat from the fluid to be cooled and at the same time it continues to remove the vaporized refrigerant and compress it to the condensing pressure. The condensing pressure must be higher enough to make refrigerant condense at ambient conditions using water or air. Ambient temperature must be less than the critical temperature of the refrigerant to effect condensation using the environment as a coolant. The temperature of the evaporator is usually near the normal boiling point of the refrigerant, so the pressure of the evaporator may be approximately atmospheric. The throttling valve maintains a pressure difference between the higher and lower side of the refrigeration cycle [20]. Note that the work supplied to the refrigeration cycle increases with the temperature lift (difference from evaporating to condensing temperature) [20].

2.2.1 The effect of natural gas pressure on liquefaction processes

Temperature-entropy diagram of natural gas mixture with phase envelope in black colour and lines of constant pressure in blue is presented in figure-4 below.

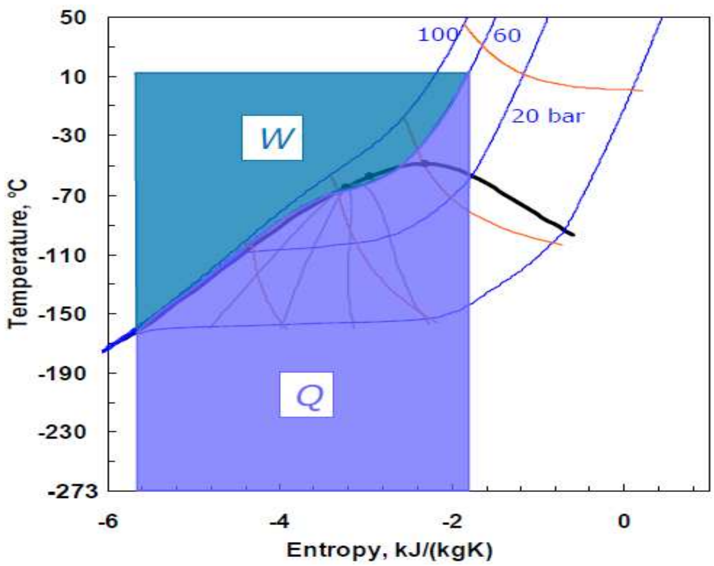

Figure 4: Temperature entropy diagram of natural gas with area showing heat removed (Q) and ideal work for reversible liquefaction process (W) (source [20])

From the diagram it can observed that if natural gas is liquefied at low pressure, work (area W) is increased and also there is some increase on amount of heat required to be removed from natural gas (area Q). Thermodynamically it is useful to liquefy natural gas at highest possible pressure so that work can be saved and reduce the heat load. Practically there are some constrains for example it is required that natural gas be below the critical pressure in order to obtain liquid/gas separation in the heavy hydrocarbon removal column. Another limitation is equipment design pressure eg. Heat exchangers hence natural gas pressure should be within heat exchanger designed pressure.

2.3 Exergy analysis of Natural gas Liquefaction Processes

Exergy analysis can be described as thermodynamic analysis technique which is based on the second law of thermodynamic and can provide alternative way of assessing and comparing thermodynamic processes/system realistically and significantly. Exergy analysis yields data/information's which shows how the actual performance of a thermodynamic system approaches the ideal. It also indicates thermodynamic losses and the effect of built environment on natural environment. Exergy analysis can describe how to design more energy efficient systems by reducing different thermodynamic losses in the system [13].

Exergy can be defined as useful work of a given amount of energy at a specific state or the work potential of energy contained in a system at specific state relative to a reference state [13], or measure of 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. It is a thermodynamic measure defined only in terms of stream enthalpy and entropy for the given stream conditions relative to the surroundings [5]

The exergy of a system at a given state depends on the condition of environment and properties of the system; hence exergy is a combination of system and environment. A system will be in dead state if it is in thermodynamic equilibrium with its environment, at this state, the system is at temperature and pressure of its environment and it has no kinetic or potential energy relative to its environment and does not react with its environment. The properties of dead state can be denoted by subscript zero (P_0, T_0, S_0, $h_{0,}$) unless specified otherwise. The idea is the system must go to the dead state at the end of the process so that the work out put can be maximized. This can be described by the assumption that if the system temperature at the end of the process is greater than that of its environment, additional work can be produced by running a heat engine between the two temperature levels or if the end pressure of the system is higher than its environmental pressure, additional work can be produced by letting the system expand [13].

It should be understood that exergy represents the maximum amount of work that a device can deliver without violating any thermodynamic laws and does not represent the amount of work that a work producing equipment will really deliver upon installation. Therefore there will be a difference large or small between exergy and actual work delivered by equipment. For example the internal energy and enthalpy of a thermodynamic system are not entirely available for work but only a portion of thermal energy of the system can be converted to work hence the exergy of the system will be larger than thermal energy available [13].

The exergy or useful available work, of a stream is therefore expressed as [5],

$Ex = (H - T_0S)_{T,P} - (H\text{-}T_0S)_{T0,P0}$ --(1)

When matter is taken from one state to another via a hypothetical reversible process, the reference terms cancel out and the change in exergy is given by

$\Delta Ex = (H - T_0S)_{state\ 2} - (H\text{-}T_0S)_{state1}$ --(2)

This change in exergy represents the minimum amount of work to be added or removed to change from state 1 to state 2 when there is an increase or decrease in internal energy or enthalpy resulting from the change [5]

Lost work can thus be defined as the difference between the work involved with the ideal reversible process and the real process chosen. Lost work, W-lost, can be expressed as

$W_{lost} = W_{actual} - \Delta Ex$---(3)

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 [5]

$W_{rev} = \sum(H-T_0S)_{LNG} - \sum(H-T_0S)_{feed}$ --(4)

Energy can enter or exit the system by heat, work and mass. In the actual process mass and energy are conserved while entropy is generated

$S_{in} - S_{out} + S_{generated} = \Delta S_{system}$ ---(5)

$Ex_{in} - Ex_{out} - Ex_{destroyed} = \Delta Ex_{system}$ --(6)

Exergy can be transferred from any system by heat, mass and work. Irreversibility's such as chemical reaction, heat transfer through finite temperature, friction, unrestrained expansion, mixing and anything that generates entropy destroys exergy. The exergy destroyed is proportional to entropy generated [13].

$Ex_{destroyed} = T_0S_{generated}$ ---(7)

2.3.1 Exergy losses in natural gas liquefaction processes

Exergy losses in natural gas liquefaction process are important parameters since such losses are to be compensated by more work or power input. The major losses are within the compression system (Compressors), heat transfer in heat exchangers (LNG heat exchanger and after coolers), losses due to refrigerant letdown and super heating of refrigerant (compressor discharge temperature). The losses can be categorized in three groups, heat transfer loss which includes losses in LNG heat exchangers and after collars, process losses which includes letdown losses and super heating of refrigerant and compressor losses [5]. Temperature entropy diagram showing the different losses is presented in figure-5 below.

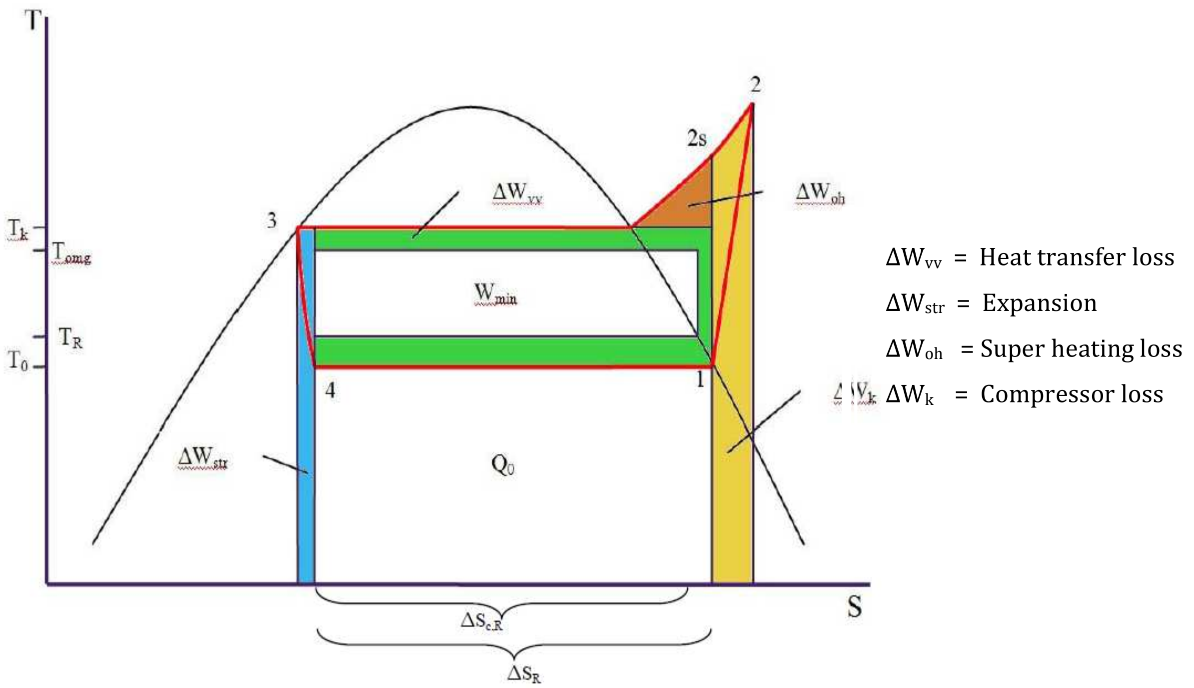

Figure 5: Temperature/entropy diagram showing different losses in refrigeration cycle

3. Proposed LNG liquefaction processes for FPSO

There are several technologies for liquefaction of natural gas, but for FPSO LNG the most proposed technology are based on mixed refrigerant cycle (MRC) which includes Single Mixed Refrigerant (SMR) and Dual Mixed Refrigerant (DMR) and expander based cycle which include Niche LNG (CH_4 and N_2) and dual nitrogen expander [21].

The proposed natural gas liquefaction cycles vary in both sophistication and power consumption. Choosing the optimum cycle for FPSO LNG is crucial and many factors are involved. Some of the major factors are [1, 2, 26];

a) LNG-FPSO liquefaction process should be light and compact due to the space and weight limitations
b) Should be adaptable/flexible to varying natural gas compositions,
c) Rapid start-up and shutdown in a safe and controlled manner, due to unstable condition at offshore. There is high probability of having many stops due to weather.
d) Ease operation and high uptime
e) Low requirement for handling potentially hazardous refrigerant
f) Reliable and insensitive to the motion of LNG-FPSO (minimize weather related downtime)
g) The process should also be low-cost and easy to maintain
h) Optimal power requirements to increase LNG production efficiency
i) Process option to recover LPG and NGL

3.1 Single Mixed Refrigerant Process (SMR-PRICO)

The process is considered as one of the simplest and most basic processes currently in operation in the industry. The mixed refrigerant used in this process contains methane, ethane, propane, pentane and nitrogen. Mixed refrigerant is compressed and passes through the main exchanger where it is condensed. It is then expanded across a Joule-Thomson valve and evaporated as it returns counter-currently through the exchanger back to the compressor. The simplest flow diagram of SMR is shown as figure-6 below. The process is simple and requires small equipment number but its capacity is limited to 1.2 MTPA per train, though the setup reduces capital costs significantly [25].

There is a considerable amount of refrigerant used in the process to facilitate the cooling of the natural gas which leads to a lot of compression work needed. Its low production capacity can be considered a disadvantage because more trains will be required to produce in high capacities. The production rate closely mirrors the capital cost not allowing for future improvement options without a total overhaul [25].

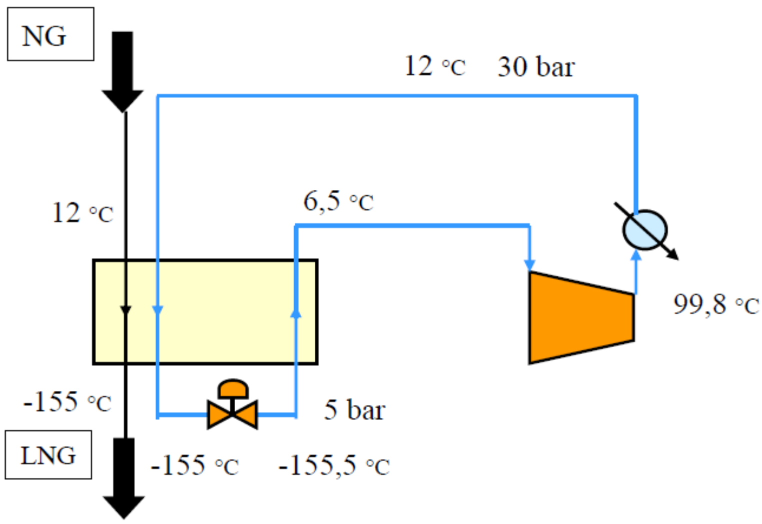

Figure 6: Principal flow diagram of SMR PRICO process (Source [20])

3.2 Dual Mixed Refrigerant process (DMR)

This process contains two refrigeration cycles. The refrigerant used in the first cycle is a mixture of ethane and propane while in the second cycle is a mixture of nitrogen, methane, ethane, propane and butane. The process is licensed by Shell Global Solutions and its capacity is reported at about 4.5 MTPA. The principles of the Shell-developed dual mixed refrigerant process is illustrates figure-7 below [25].

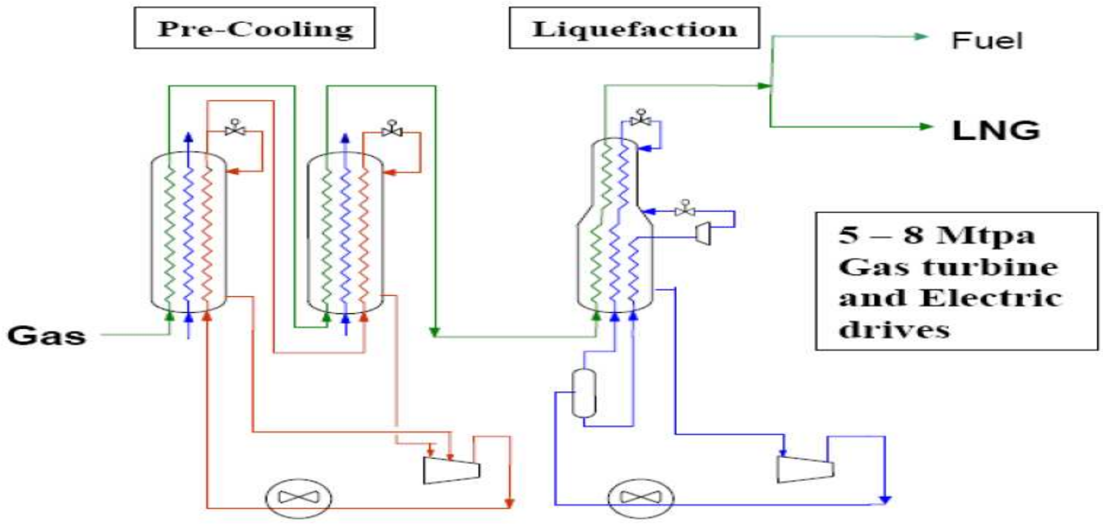

Figure 7: Diagram of dual mixed refrigerant process (Source [16])

A main argument for developing the DMR process is the need for a pre-cooling refrigerant that can cover a wider temperature range than propane, and thus give a better load distribution between the compressors. Especially in a cold or arctic climate, and for more optimal integration of HHC extraction, the minimum pre-cooling temperature needs to be extended. This has been solved in the DMR process by using a refrigerant mixture mainly based on ethane and propane. The process has been applied in the Sakhalin LNG Plant in Russia where cold climate, air cooling and large seasonal variation in temperature can be adapted to in a better way by using a mixed pre-cooling refrigerant where composition can be optimized to meet varying operating conditions [25].

3.3 Dual nitrogen expander

To eliminate the weakness of single nitrogen expander, a second stage expander was introduced on the process. A large part of refrigerant is expanded at warm temperature and only sufficient amount for sub-cooling is expanded at low temperature. There are many companies who have developed their technology based on dual nitrogen expander cycle some of them are Mustang Engineering, Hamworthy, Petroleum Pty Ltd, Dubar (PHB) , Kanfa Aragon and Statoil [6, 16]. The schematic flow diagram of dual nitrogen expander cycle is given in figure-8 below.

The industrial reference for dual nitrogen expander cycle is Kollsnes II, built by Hamworthy and the plant has an energy demand reported to be 510 kWh/ton LNG, which is a considerable reduction from the Snurrevarden plant [16].

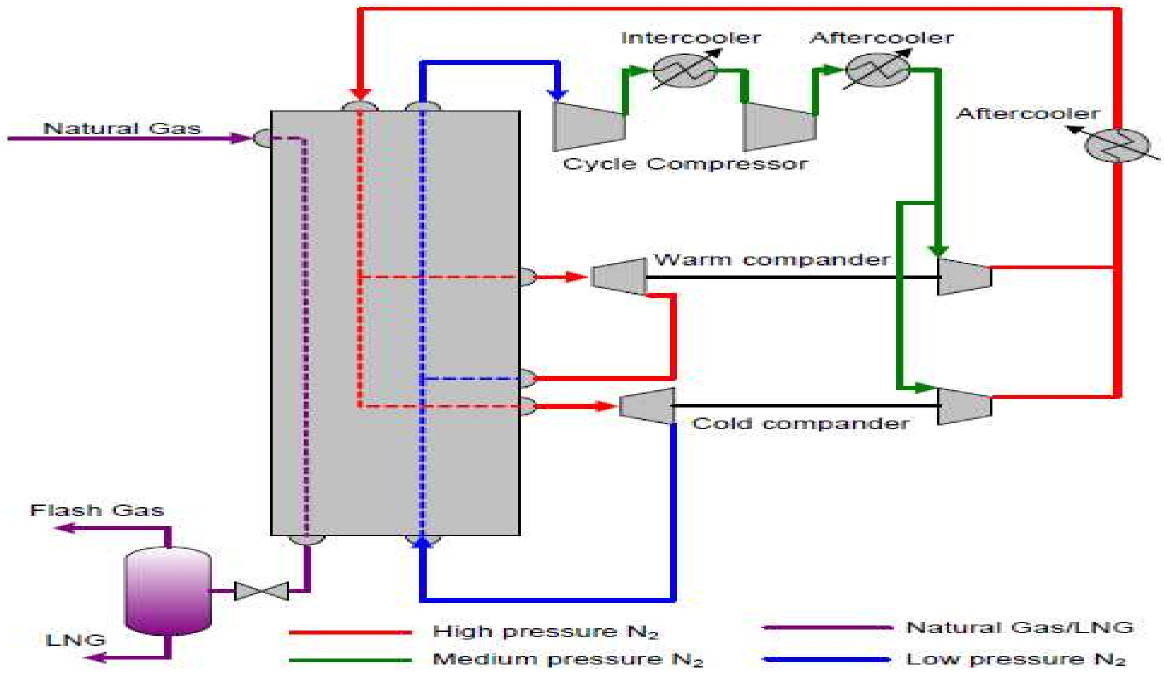

Figure 8: Diagram of dual nitrogen expander process (source [16])

3.4 Nitrogen and methane Expander (Niche LNG)

This process is offered by CB&I Lummus and consists of two cycles. The first cycle uses methane as refrigerant or feed natural gas after heavy hydrocarbon has been removed and provides cooling at warm and moderate level. The second cycle uses nitrogen and provides refrigeration at lower temperature levels. Both cycles use turbo expanders and their refrigerant is always in gaseous phase [6, 16, and 23].

4. Basis for simulation of proposed liquefaction processes for LNG FPSO

As described in previous sections, four LNG liquefaction process have been proposed as good candidates for LNG FPSO, the proposed processes are single mixed refrigerant (SMR-PRICO), dual mixed refrigerant (DMR), dual nitrogen expander and Niche LNG (CH_4 and N_2 liquefaction process). This chapter describes the assumptions for simulations, thermodynamic parameters simulated for each process and methodology used during simulation. An in-depth optimization has not been conducted but the simulation was aimed at obtaining an optimal efficient process based on the simulated constraints. The processes were simulated based on natural gas supplied at flow rate of 150mmscfd.

The effects of natural gas pressure, temperature and composition on the proposed liquefaction processes for LNG FPSO were investigated in the simulation. During the simulations the effects were analyzed by examining specific power, power consumption and refrigerant flow rate of the proposed processes. The results of each process are plotted in graphs and are presented in the chapter of results of each process. All four processes were simulated using commercial mass and energy balance software (hysys) with the Peng–Robinson equation of state.

4.1 Assumptions

The benchmarked simulation of each proposed liquefaction cycle was based on natural gas supplied at 60bar and ambient temperature of 15^{0}C. The ambient condition was assumed as natural gas inlet/supply temperature throughout the simulation. The refrigerant conditions and operation parameters of each process were found from patents and are presented in the detailed description chapters of each process. The simulated processes based on the above description were used as benchmark for all other simulated conditions and parameters.

Generally prior to liquefaction, the natural gas feed is pre-treated to remove acid gases (such as carbon dioxide and hydrogen sulfide), water, mercury and portion of heavier hydrocarbons. The pretreatment process is not covered in this thesis hence the natural gas feed used in simulation is assumed to be pretreated natural gas.

Molar fractions of natural gas composition used in the simulation of thermodynamic parameters of liquefaction processes mentioned above is presented in table-1 below and is termed as Composition-reference. The other composition used to analyze the effect of natural gas composition on the proposed liquefaction processes for LNG FPSO are presented in table (2, 3, 4, 5 & 6) and termed as Composition- A, B, C, D & E.

Table 1 : Composition Reference

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction Composition - reference	0.90043	0.07321	0.00346	0.000008	0.000001	0.0229	0.000005

Table 2: Natural gas Compostion-A

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction of Composition- A	0.9146	0.0570	0.0129	0.0000	0.0020	0.0133	0.00005

Table 3: Natural Gas composition-B

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction of Composition- B	0.8738	0.067	0.035	0.006	0.009	0.005	0.000005

Table 4: Natural gas composition-C

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction of Composition- C	0.97	0.008	0.013	0.003	0.004	0.002	0.000005

Table 5: Natural gas composition-D

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction of Composition- D	0.88	0.06	0.03	0.005	0.005	0.02	0.000005

Table 6: Natural gas composition-E

Component	C_1	C_2	C_3	iC_4	nC_4	N_2	CO_2
Mole fraction of Composition- E	0.82	0.112	0.04	0.012	0.009	0.007	0.000005

4.2 Methodology

The literature review on the patents of each proposed liquefaction process was conducted and one patent for each process was selected. The detailed description of each process based on the selected patent is presented in detailed description chapter of each process.

Based on the selected patent conditions and parameters the liquefaction processes were built on hysys using natural gas with composition presented in table-1 and supplied at pressure of 60bar, temperature of 15^0C and flow rate of 150 mmscfd. The processes were optimized by varying refrigerant flow rate to obtain the proposed minimum approach temperatures and LMTD. After optimization of the processes the key parameter recorded were “UA” values of LNG heat exchangers, power consumption, refrigerant flow rate and specific power. The simulated processes based on above conditions were taken as benchmark during variation of other parameters to analyze the effect of pressure, temperature and composition on the processes

The calculated “UA” value of all LNG heat exchangers was kept constant during variation of natural gas pressure, temperature and composition. The reason for keeping the calculated “UA” value constant is that, in real plant the designed “UA” of LNG heat exchanger cannot be changers unless you modify or replace heat exchanger. The optimization of processes was performed by varying refrigerant flow rate and refrigerant evaporating pressure to ensure “UA” values of LNG heat exchangers are kept constant as in the benchmarked process.

To analyze the effect of natural gas temperature on the benchmarked processes, the temperature of natural gas was varied between (5^0C) and 30^0C) and the effect of natural gas temperature on specific power, power consumption and refrigerant flow rate when natural gas is supplied at 60bar was determined and presented in graphs. The same procedure was done to analyze the same effects when natural gas is supplied at different pressures between (20 and 150bar).

To analyze the effect of natural gas pressure, simulation started by supplying natural gas at a temperature of 15^0C and varying natural gas pressure between 20 and 150bar, the graphs showing its effect on specific power, power consumption and refrigerant flow rate are drawn and presented in result section of each process. The same procedure was performed to analyze the same effects when natural gas is supplied at different temperatures (between 5 and 30^0C) and for each temperature natural gas pressure was varied between 10 and 150bar and its effect on specific power, power consumption and refrigerant flow rate was determined and presented in graphs in result section.

The effect of natural gas composition on processes has been analyzed using five different composition termed as composition-A,B,C,D and E as presented in table 2, 3, 4, 5&6 section 4.1 above. The compositions were analyzed by looking at its effect on process power consumption, refrigerant flow rate and specific power. All compositions were analyzed by supplying natural gas at 60bar, 15^0C and molar flow rate of 150mmscfd. Results are presented in graphs and tables in the results section of each process.

5. Simulation of Single Mixed Refrigerant Process (SMR- PRICO)

5.1 Detailed description of SMR process

The single mixed refrigerant (SMR) process with single compressor has been simulated and the effect of natural gas pressure, temperature and composition on the process is analyzed also useful exergy and exergy losses on the process are determined. The process flow diagram of the simulated SMR is presented in figure-9 below. The simulated process was based on patent with reference [19].

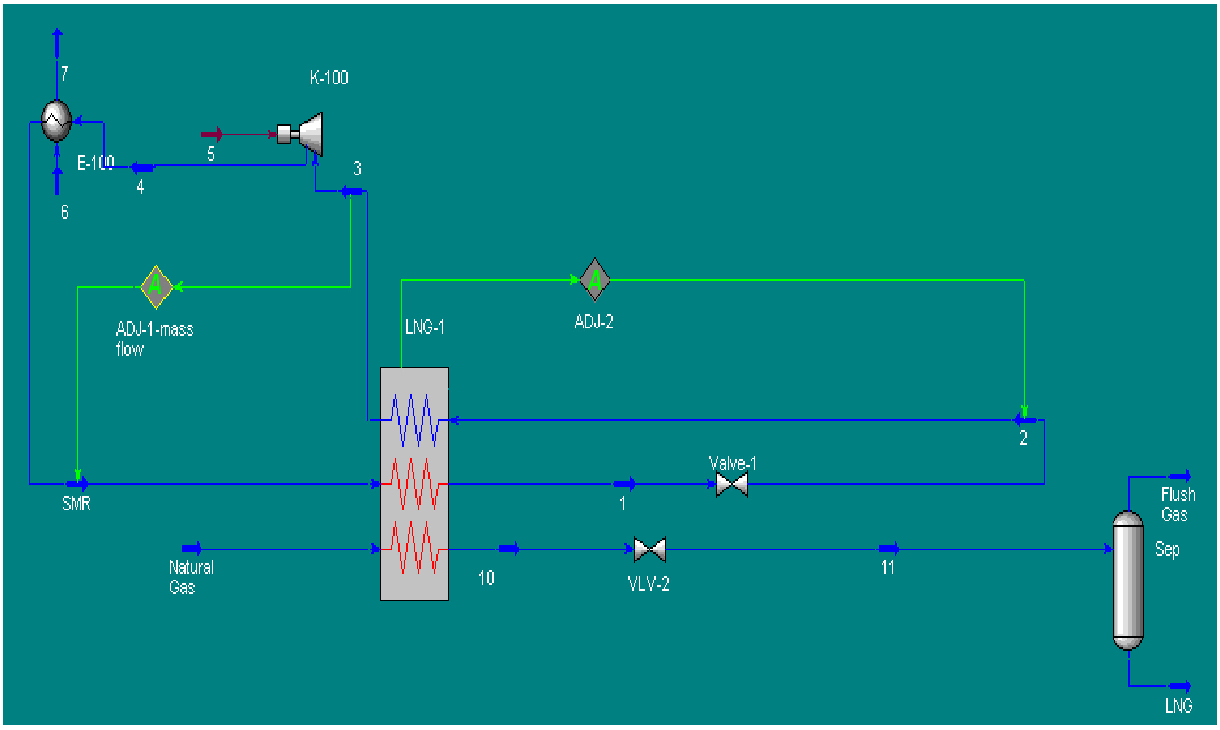

Figure 9: Process Flow Diagram of SMR (PRICO)

Referring to figure-9 above, Pre-treated natural gas (**stream natural gas**) at 60bar and 15^{0}C is cooled in LNG heat exchanger (**LNG-1)** to temperature of about -155^{0}C (**stream-10)** by vaporizing mixed refrigerant **(stream-2).** Cooled natural gas (**stream-10)** at -155^{0}C and 60bar is expanded in throttling valve **(VLV-2)** to -162.9^{0}C and 1.05bar (**stream-11).** The Produced LNG (**stream-11)** is sent to flash tank where LNG and Flash gas can be separated. The ratio of flash gas to inlet natural gas is 0.076, this means that 92.4% of natural gas is converted to LNG.

The mixed refrigerant **(stream-SMR)** enters LNG heat exchanger **(LNG-1)** at ambient temperature 15^{0}C and pressure 30bar and is cooled to -155^{0}C (**stream-1).** Cooled refrigerant **(Stream-1)** is reduced in pressure adiabatically across a throttling valve **(valve-1)** to pressure of about 5bar and its temperature is decreased to -158^{0}C (**stream-2)** and introduced to the cold end of LNG heat exchanger (**LNG-1)** to provide refrigeration to stream **(SMR and Natural Gas)**. Vaporized refrigerant **(stream-3)** is withdrawn from LNG heat exchanger (**LNG-1)** at 12^{0}C . The vaporized refrigerant **(stream-3)** at 5bar and 12^{0}C is returned to the compressor (**K-100)** and its pressure and temperature is increased to 30bar and 135.1^{0}C respectively (**steam-4)**. Hot refrigerant **(stream-4)** is cooled in cooler **(E-100)** to ambient temperature 15^{0}C (**stream-SMR**). Then stream-SMR enters LNG heat exchanger **(LNG-1)** and the cycle repeats as described above. The refrigerant molar flow rate is approximately 530mmsfcd.

The refrigerant composition is as given in table-7 which comprises mainly methane and ethane. The amount of propane is low compared to n-butane because of low assumed ambient temperature (15^{0}C). The compressor suction volume flow rate is 123,935.5 m^3/hr which gives a mass flow rate of 213.5kg/s .

Table 7: SMR process Refrigerant composition

Compound	Methane	Ethane	Propane	n-Butane	Nitrogen	Total
Mole Fraction	0.40	0.31	0.01	0.17	0.11	1.0

5.2 Simulation Results for Single Mixed Refrigerant (SMR- PRICO)

The process was simulated by hysys using the conditions described above. At minimum approach temperature of 2.91^{0}C and LMTD of 5.9^{0}C LNG heat exchanger "UA" value and process specific power was found to be 16,629.21kW/^{0}C and 0.387kWh/kg-LNG respectively. The "UA" value of the LNG heat exchanger was kept constant during variation of other parameters in simulation.

5.2.1 The effect of natural gas temperature on SMR PRICO process

a) The effect of natural gas temperature on specific power

The effect of natural gas temperature on specific power when natural gas is supplied at 60bar in SMR process is presented in figure-10 below. By Considering 15^0C as reference point, the graph shows that when natural gas temperature decreases from 15 to 5^0C, SMR process specific power decreases from 0.387 to 0.329 kWh/kg-LNG which amounts to 14.99% decrease and when the natural gas supply temperature increases from 15 to 25^0C, its specific power increases from 0.387 to 0.539 kWh/kg-LNG which amounts to 39.27% increase. In general it can be described that at given natural gas pressure and temperature, an increase in natural gas supply temperature increases specific power and a decrease in natural gas supply temperature decreases specific power.

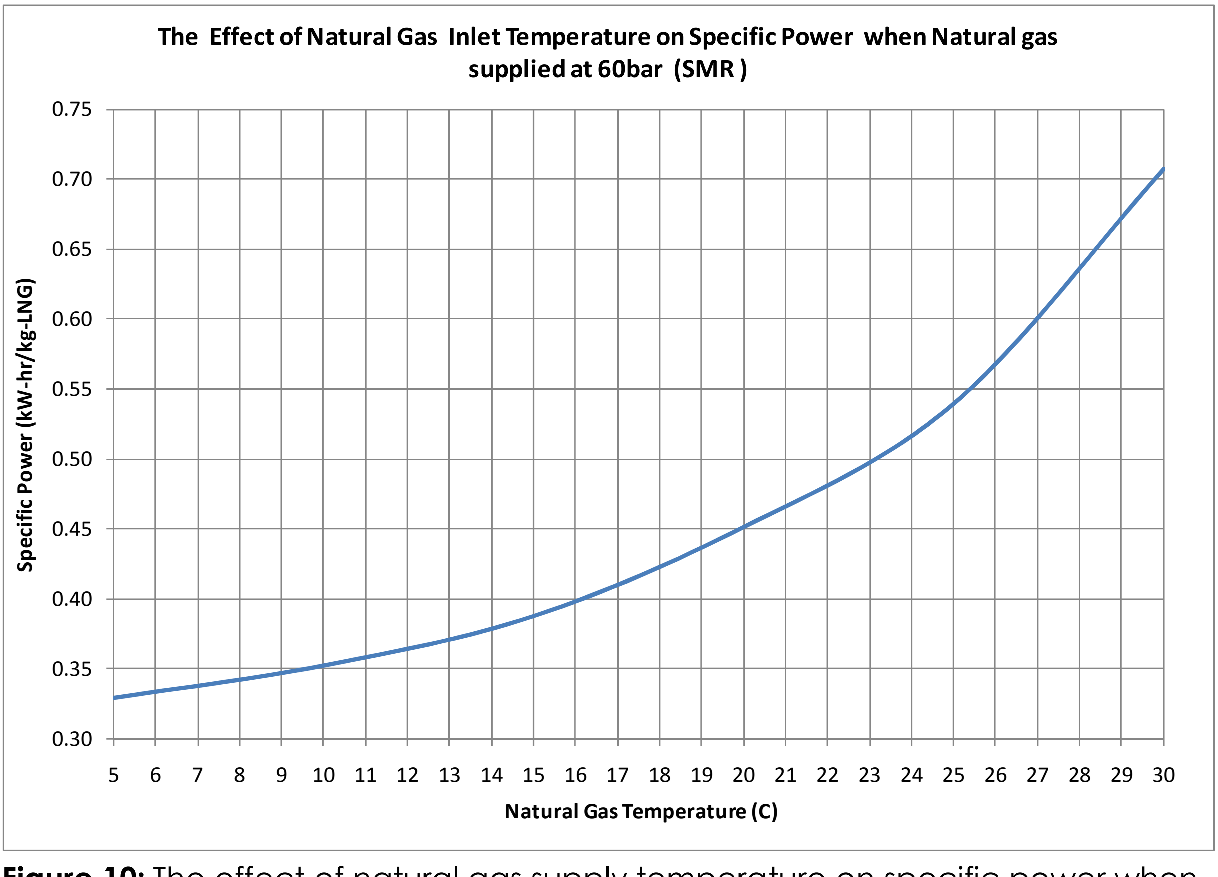

Figure 10: The effect of natural gas supply temperature on specific power when natural gas supplied at 60bar (SMR-PRICO)

The effect of natural gas temperature on specific power at different natural gas pressures (20, 40, 60, 80, 100, 120 and 140bar) supplied to SMR process is presented in figure-11 below. With reference to the simulated conditions of LNG heat exchanger (15^0C and 60bar), figure-11 below shows clearly that as temperature decreases from 15^0C to 5^0C at any natural gas supply pressure, specific power decreases and when supply temperature increases from 15^0C to 30^0C, specific power increases. Also the graphs depicted that natural gas supplied at higher pressure has low specific power (eg graph with pressure 140bar) and natural gas supplied at low pressure has higher specific power (eg graph with 20bar pressure) irrespective of natural gas supply temperatures. For example with reference to simulated conditions of this process, natural gas supplied at (15^0C and 90bar) and (15^0C and 60bar) have specific power of 0.348 and 0.387kWh/kg-LNG respectively which is a difference of 10.07% with respect to 60bar.

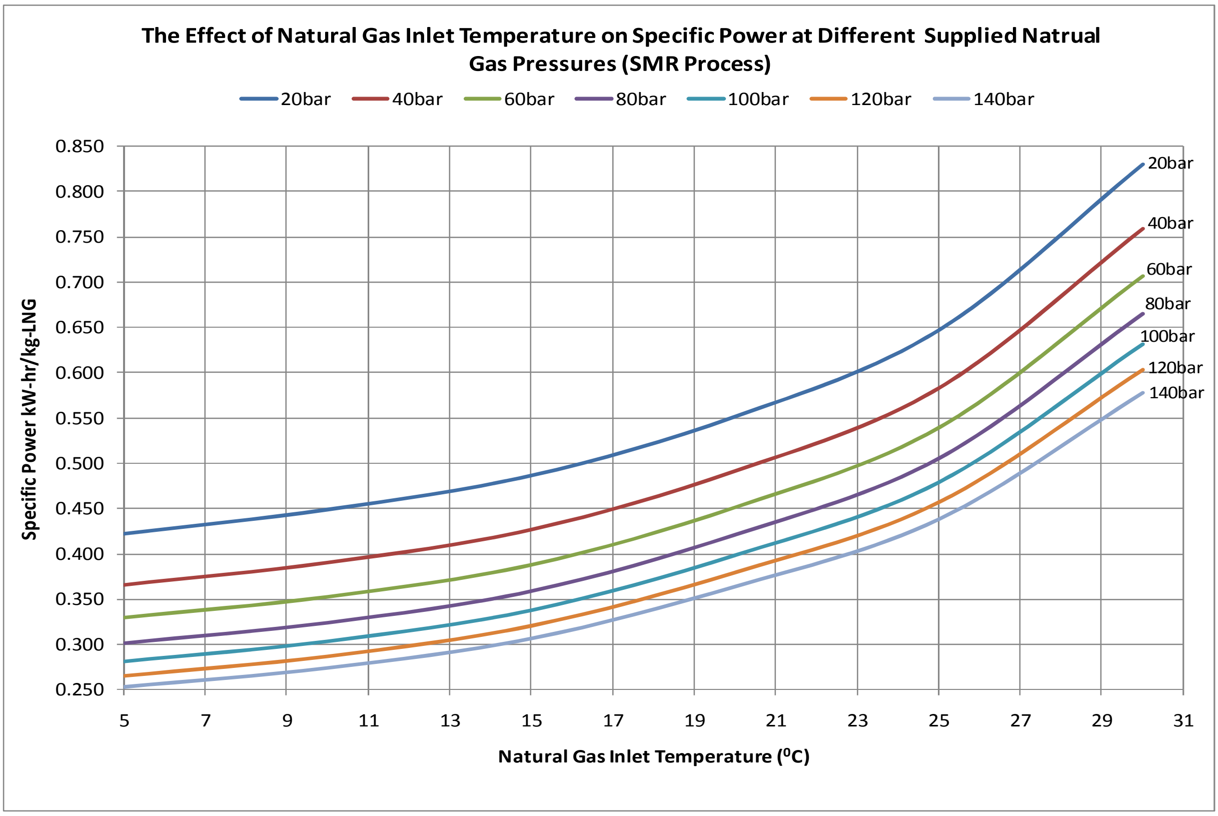

Figure 11: The effect of natural gas supply temperature on specific power when natural gas supplied at different pressures (SMR-PRICO)

b) The effect of natural gas temperature on power consumption

The effect of natural gas supply temperature on power consumption in SMR process is presented in figure-12 below. The graph shows that when natural gas supply temperature decreases from 15 to 5^{0}C, SMR process power consumption decreases from 46.42 to 39.41 MW which amounts to a 15.10% decrease and when supply temperature increases from 15 to 25^{0}C, its power consumption increases from 46.42 to 64.61 MW which amounts to a 39.19% increase. Thus, it can be described that when natural gas is supplied at a given pressure and temperature, an increase in natural gas supply temperature will increase power consumption and a decrease in natural gas supply temperature, will decrease power consumption.

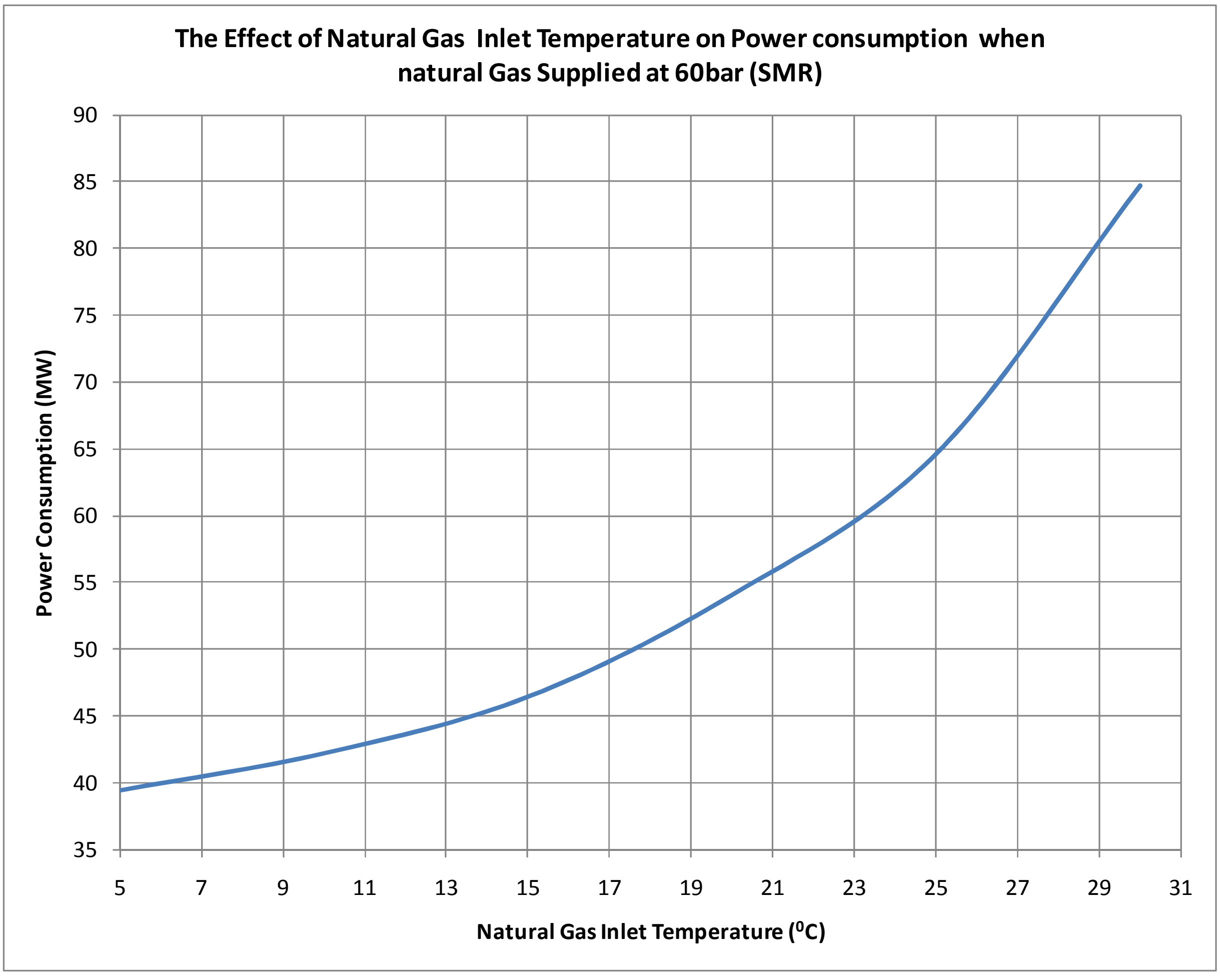

Figure 12: The effect of natural gas supply temperature on power consumption when natura gas supplied at 60ba (SMR PRICO)

The graphs showing the effect of natural gas inlet temperature on power consumption when natural gas is supplied at different pressures (20, 40, 60, 80, 100, 120 and 140bar) in SMR process is presented in figure-13 below. The figure shows that when natural gas temperatures at any supply pressure decreases from 15^{0}C to 5^{0}C, power consumption decreases and when it increases from 15^{0}C to 30^{0}C the power consumption increases. Also the graphs depicted that when natural gas is supplied at higher pressures, power consumption is low (eg graph with 140bar pressure) and when it is supplied at low pressures, power consumption is high (eg graph with 20bar) irrespective of supply temperature. For example natural gas supplied at (15^{0}C, 90bar) and (15^{0}C , 60bar) have power consumptions of 41.40 and 46.42MW respectively which amounts to a difference of 10.81%.

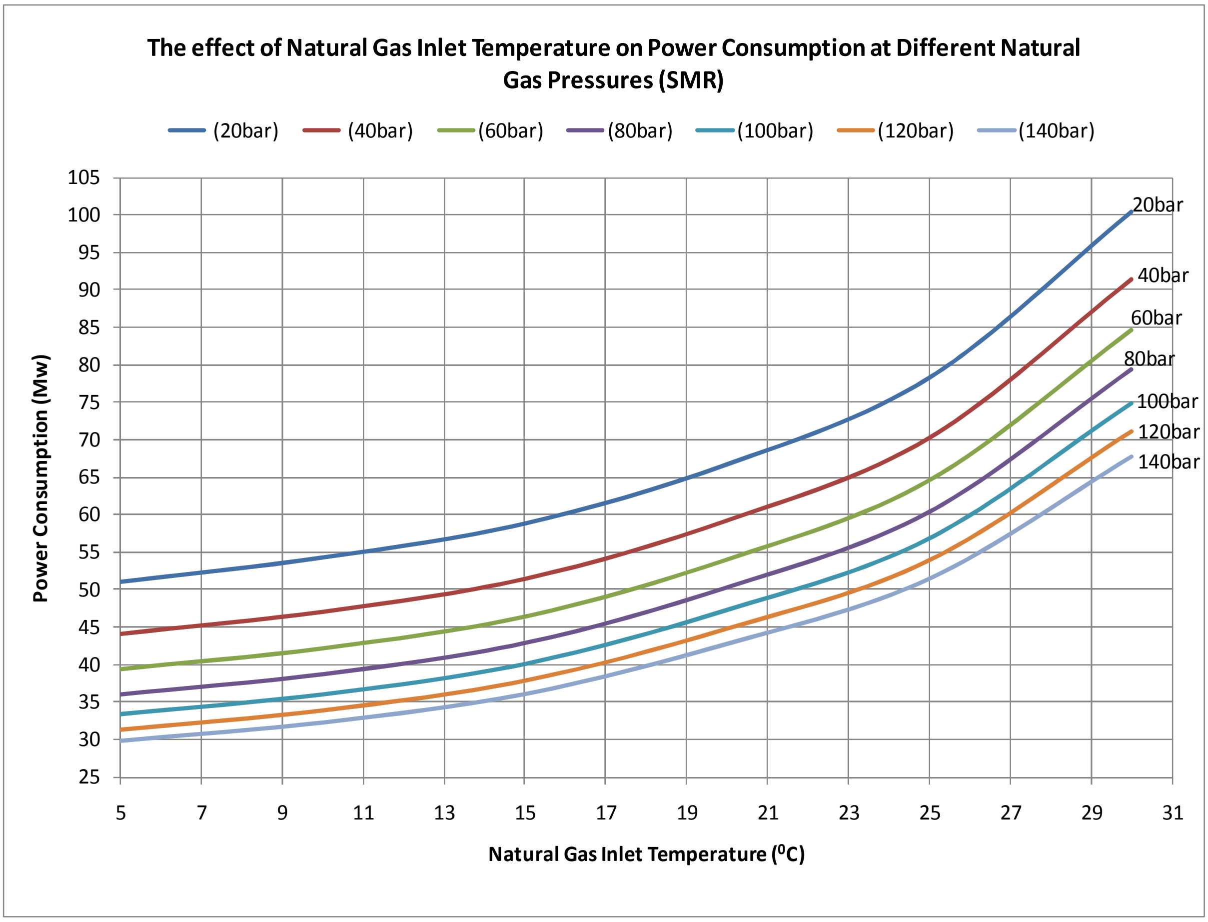

Figure 13: The effect of natural gas inlet temperature on power consumption at different Pressures (SMR PRICO)

C) The effect of natural gas inlet temperature on refrigerant flow rate

The effect of natural gas temperature on refrigerant flow rate when natural gas is supplied at 60bar in SMR process is given in figure-14 below. The figure shows that when natural gas temperature decreases from 15^{0}C to 5^{0}C, refrigerant flow rate decreases from 531.28 to 431.10 mmscfd which amounts to a 18.86% decrease and when supply temperature increases from 15^{0}C to 25^{0}C, the refrigerant flow rate increases from 531.28 to 713.04 mmscfd which amounts to 34.21% increase. Generally it can be described that when natural gas is supplied at a given pressure and temperature, an increase in supply temperature will lead to an increase in refrigerant flow rate and a decrease in supply temperature will result in a decrease refrigerant flow rate.

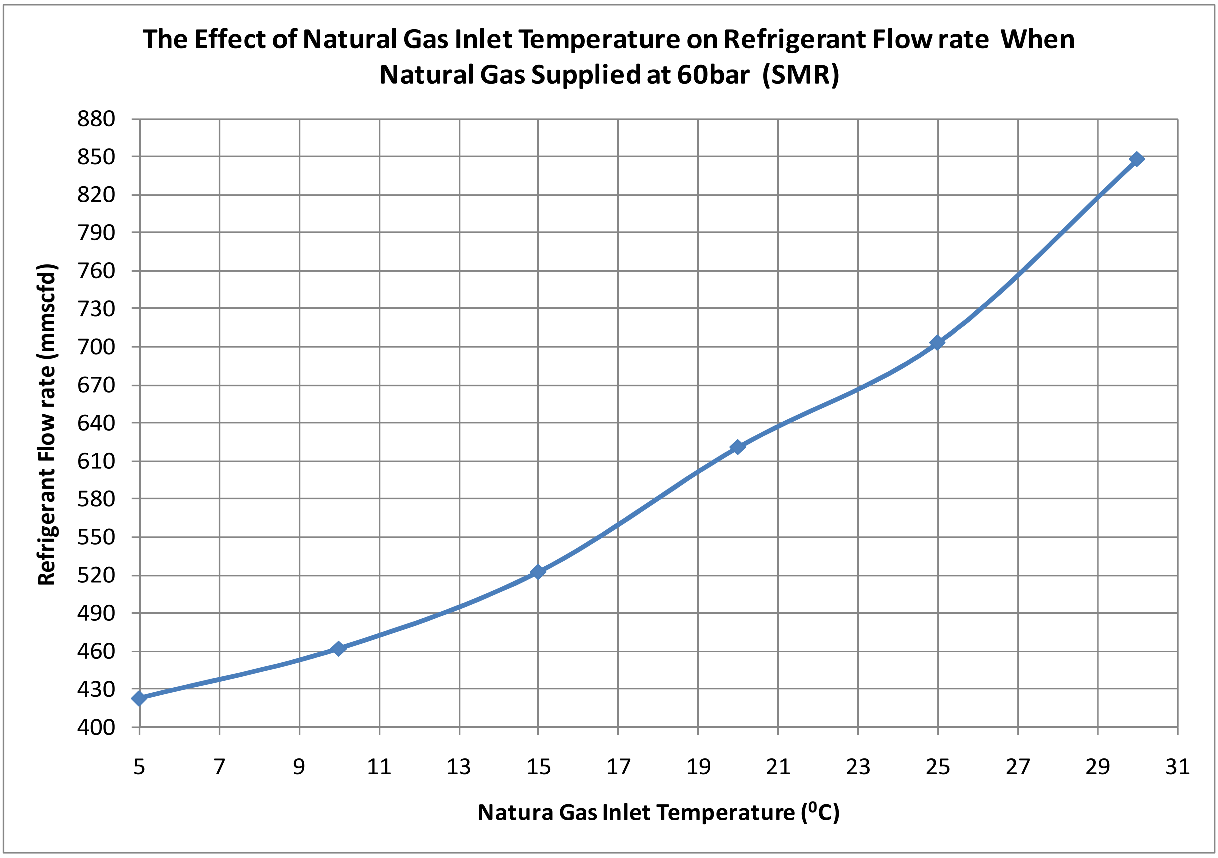

Figure 14 : The effect of natural gas inlet temperature on refrigerant flow rate when natural gas supplied at 60 bar (SMR PRICO)

The graphs which show the effect of natural gas temperature on refrigerant flow rate when natural gas is supplied at different pressures (20, 40, 60, 80, 100, 120 and 140 bars) in SMR process are presented in figure-15 below. The figure shows that when natural gas supply temperature at any supply pressure decreases from 15^0C to 5^0C refrigerant flow rate decreases and when supply temperature increases from 15^0C to 30^0 , refrigerant flow rate increases. The figure also depicts that when natural gas is supplied at higher pressure, refrigerant flow rate is low (eg graph with 140bar) and when natural gas is supplied at low pressures, the refrigerant flow rate is high (eg graph with 20bar) at any supply temperature. For example when natural gas is supplied at (15^0C, 90bar) and (15^0C, 60bar) refrigerant flow rate is 505.58 and 531.28 mmscfd respectively, which amounts to a difference of 4.84% with respect to 60bar.

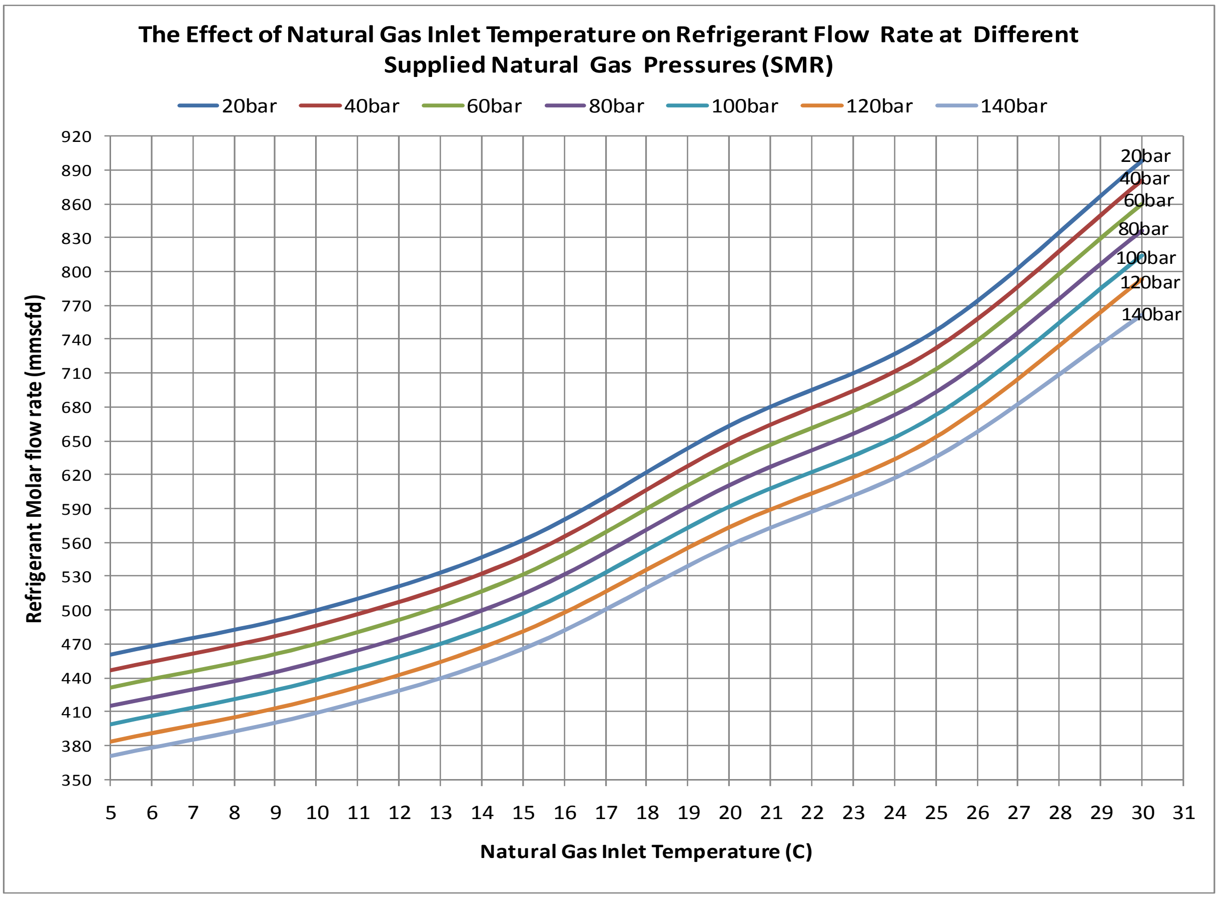

Figure 15: The effect of natural gas inlet temperature on refrigerant flow rate at different natural gas supplied pressures (SMR- Process)

5.2.2 The effect on Natural gas pressure on SMR-PRICO

a) The effect of natural gas pressure on specific power

The effect of natural gas pressure on specific power when natural gas is supplied at 15^{0}C in SMR process is presented in figure-16 below. The figure indicates that when natural gas pressure decreases from 60 to 30bar, specific power increases from 0.387 to 0.452 kWh/kg-LNG which amounts to a 16.80% increase and when natural gas pressure increases from 60 to 90bar specific power decreases from 0.387 to 0.348 kWh/kg-LNG which amounts to a 10.08 % decrease. In general when natural gas is supplied at a given pressure and temperature, specific power will increase when supply pressure decreases and specific power will decrease when supply pressure increases.

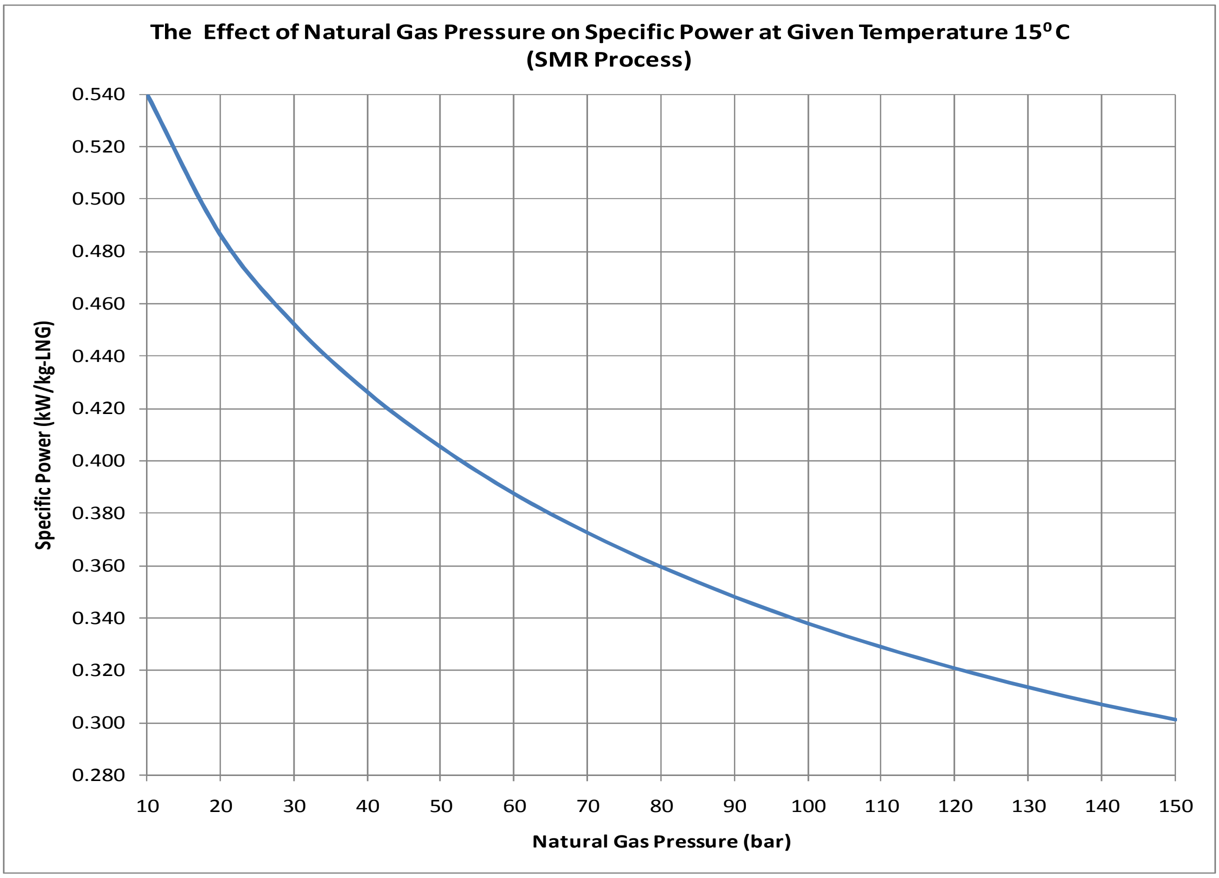

Figure 16: The effect of natural gas pressure on specific power when natural gas supplied at inlet temperature of 15^{0}C (SMR-PRICO)

The graphs showing the effect of natural gas pressure on specific power when natural gas is supplied at different temperatures in SMR process is presented in figure-17 below. The figure shows that at any supply temperature of natural gas, when natural gas pressure decreases from 60 to 20bar, specifc power increases and when supply pressure increases from 60 to 150bar specific power decreases. The graphs depicted that when natural gas is supplied at higher temperature, specific power is higher (eg graph with 30^{0}C) and when supplied at low temperature, specific power is low (eg graph with 5^{0}C) at any supply pressure. For example natural gas supplied at (15^{0}C, 60bar) and (30^{0}C, 60bar) have specific power of 0.387 and 0.707 kWh/kg-LNG which amounts to a difference of 8.27% with reference to 15^{0}C.

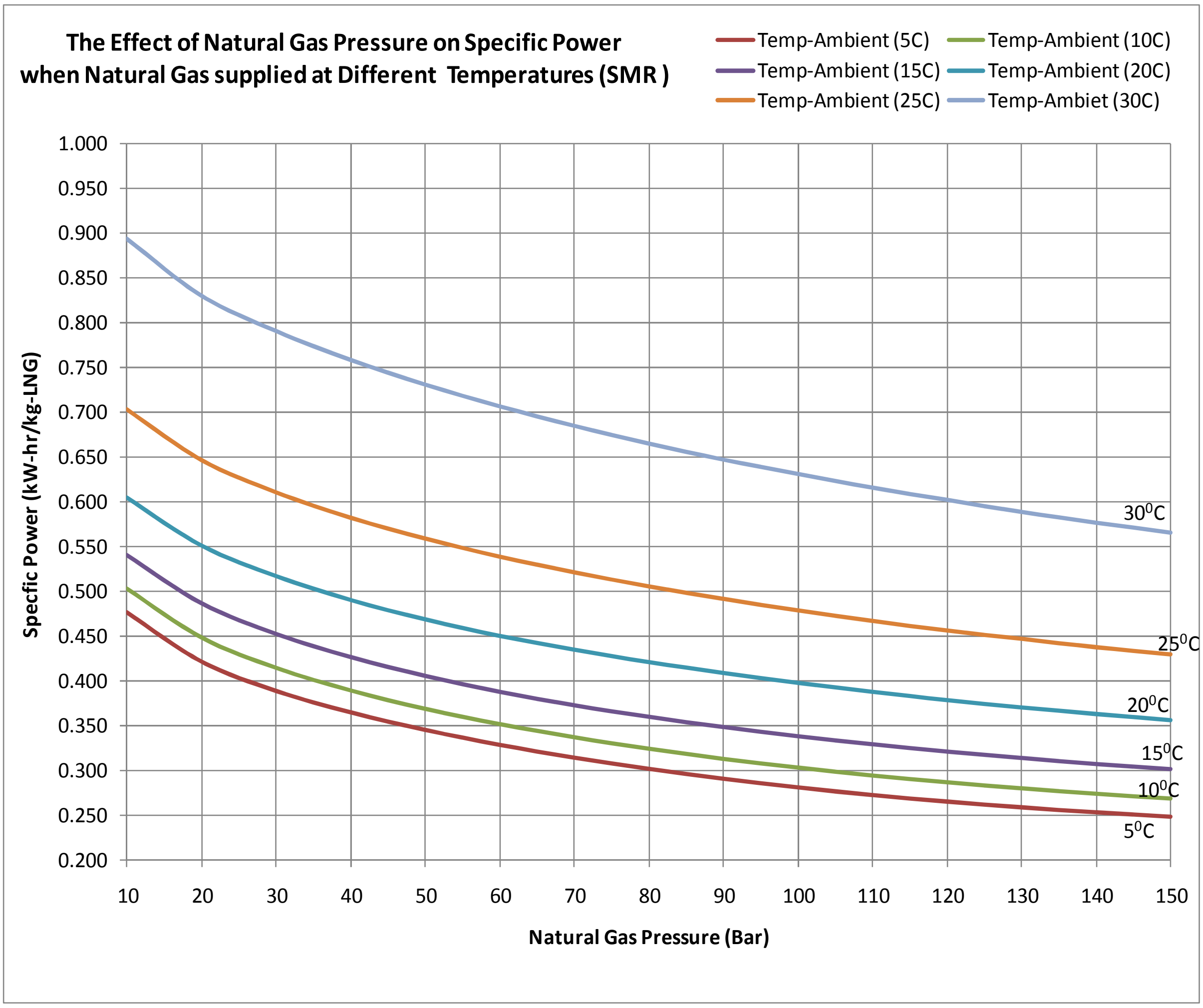

Figure 17 : The effect of natural gas pressure on specific power when natural gas supplied at different temperature (SMR-PRICO)

b) The effect of natural gas pressure on power consumption

The effect of natural gas pressure on power consumption when natural gas is supplied at 15^{0}C in SMR process is presented in figure-18 below. The figure indicates that when natural gas pressure decreases from 60 to 30bar, SMR process power consumption increases from 46.42 to 54.59MW which amounts to a 17.60% increase and when natural gas pressure increases from 60bar to 90bar, its power consumption decreases from 46.42 to 41.40 MW which amounts to a 10.81% decrease. Thus, when natural gas is supplied at a given pressure and temperature, a decrease in supply pressure will increase power consumption and an increase in supply pressure will decrease power consumption.

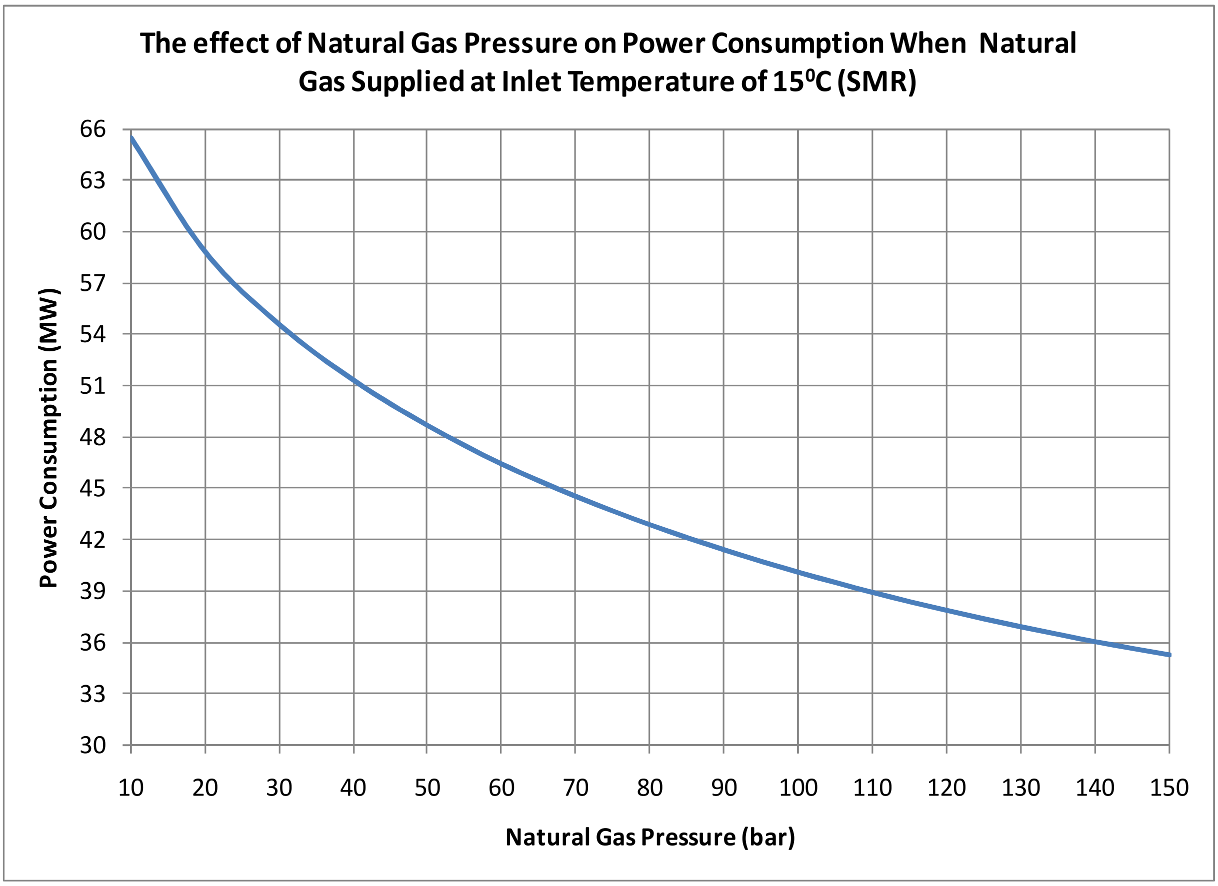

Figure 18: The effect of natural gas pressure on power consumption when natural gas supplied at inlet temperature of 15^{0}C (SMR-PRICO)

The graphs showing the effect of natural gas pressure on power consumption when natural gas is supplied at different temperatures (5, 10, 15, 20, 25, 30^{0}C) in SMR process are presented in figure-19 below. The figure shows that when natural gas supply pressure decreases from 60 to 20bar , power consumption increases and when natural gas supply pressure increases from 60 to 150bar, power consumption decreases at any natural gas supply temperature. The graphs depicted that natural gas supplied at low temperature has low power consumption (eg graph with 5^{0}C) and natural gas supplied at higher temperature has higher power consumption (eg graph with 30^{0}C) at any natural gas supply pressure. For example natural gas supplied at (15^{0}C, 60bar) and (25^{0}C, 60bar) have power consumption of 46.42 and 64.61 MW respectively which amounts to a difference of 39.19% with respect to 15^{0}C.

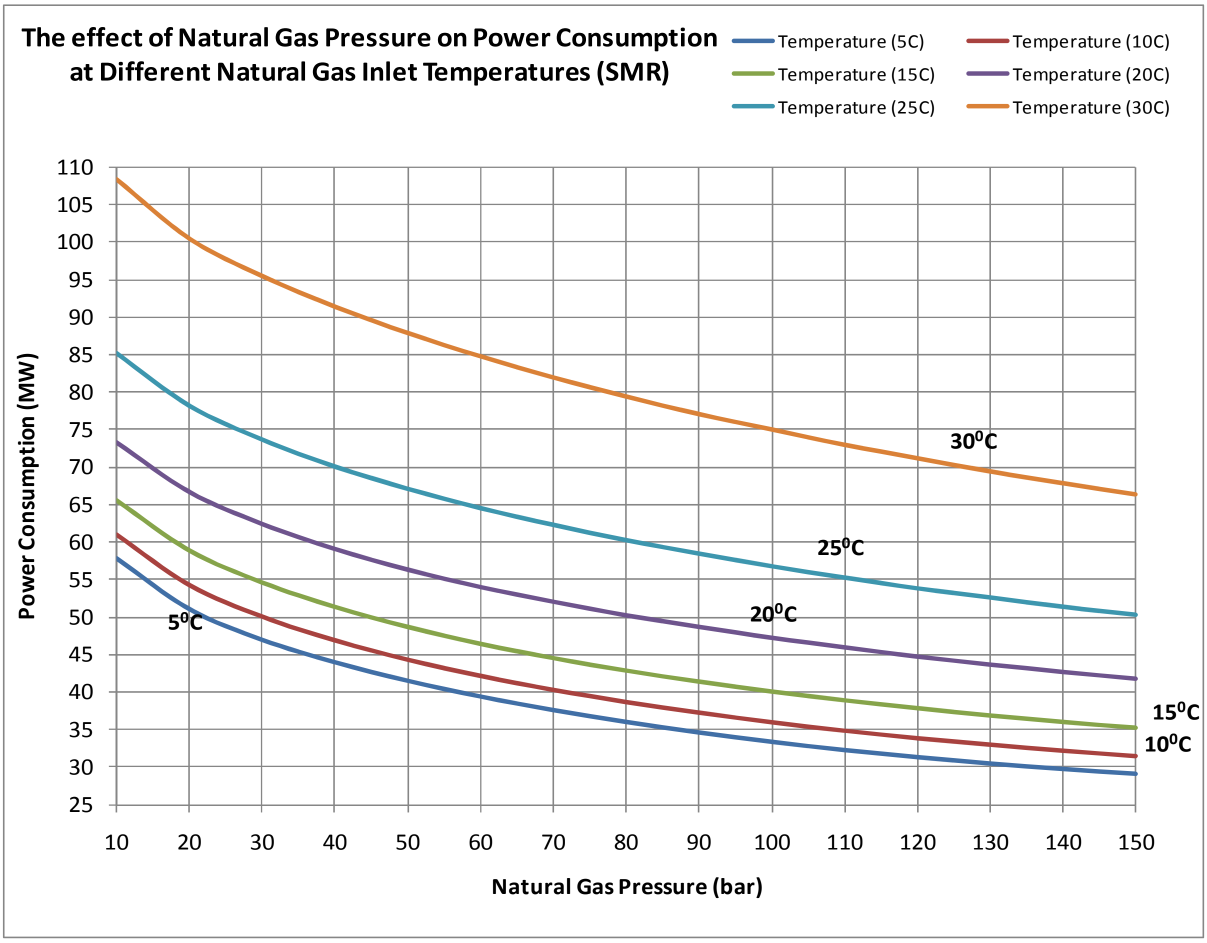

Figure 19: The effect of Natural gas pressure on power consumption when natural gas supplied at different temperatures (SMR-PRICO)

C) The effect of natural gas pressure on refrigerant flow rate.

The effect of natural gas pressure on refrigerant flow rate when natural gas is supplied at 15^{0}C in SMR process is presented in figure-20 below. The figure shows that when natural gas pressure decreases from 60 to 30bar, the refrigerant flow rate increases from 531.28 to 554.66 mmscfd which amounts to a 4.4% increase and when natural gas pressure increases from 60 to 90bar, refrigerant flow rate decreases from 531.28 to 505.58 mmscfd which amounts to a 4.84% decrease. It can be concluded that at a given natural gas pressure and temperature, a decrease in supply pressure will increase refrigerant flow rate and an increase in supply pressure will decreases refrigerant flow rate.

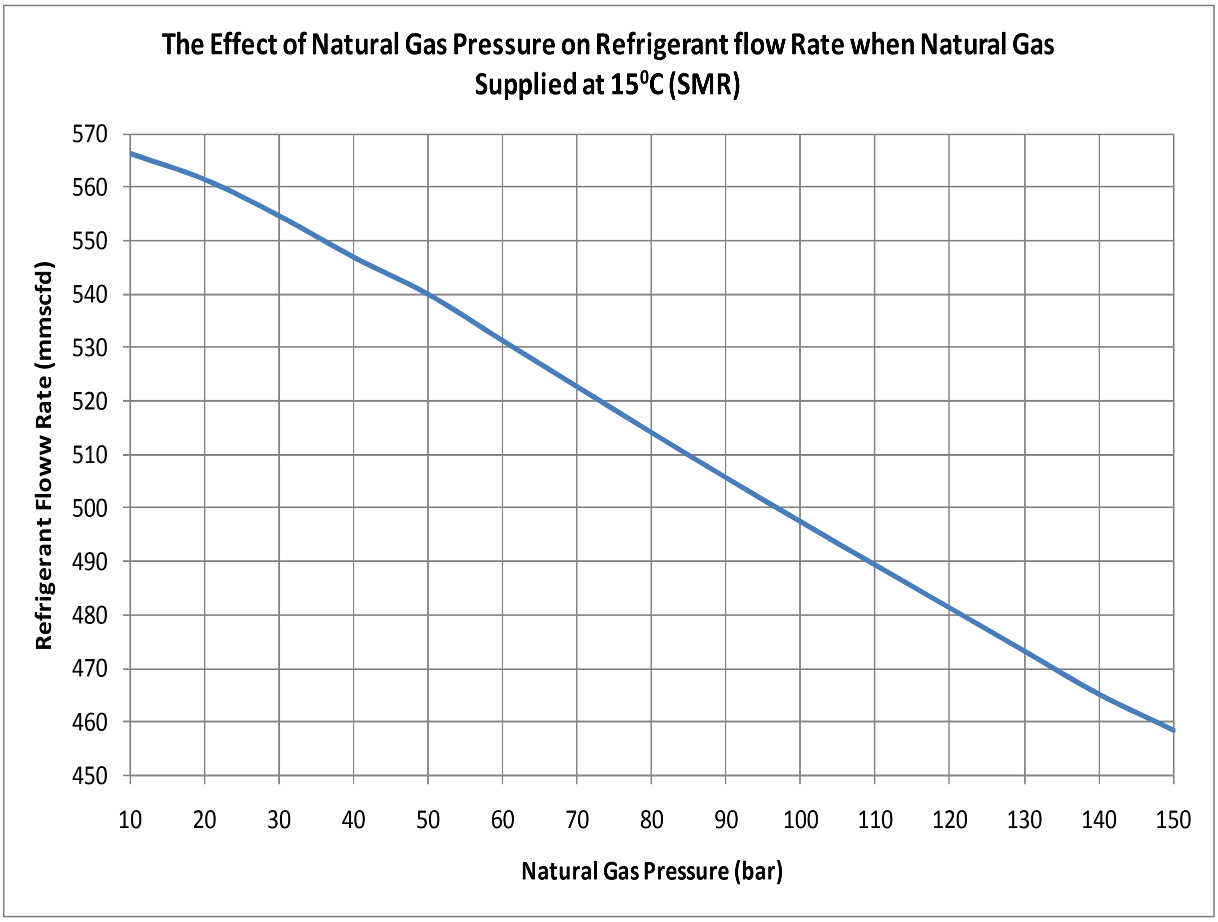

Figure 20: The effect of natural gas pressure on refrigerant flow rate when natural gas supplied at different temperatures (SMR-PRICO)

The graphs showing the effect of natural gas pressure on refrigerant flow rate when natural gas is supplied at different temperatures (5, 10, 15, 20, 25 and 30^{0}C) in SMR process is presented in figure-21 below. The figures shows that when natural gas supply pressure decreases from 60 to 20bar refrigerant flow rate increases and when natural gas supply pressure increases from 60 to 150bar, refrigerant flow rate decreases at any natural gas supply temperature. The graph indicates that when natural gas is supplied at low temperatures (eg graph with 5^{0}C), refrigerant flow rate is low and when natural gas is supplied at higher temperatures (eg graph with 30^{0}C) refrigerant flow rate is higher at any natural gas supply pressure.

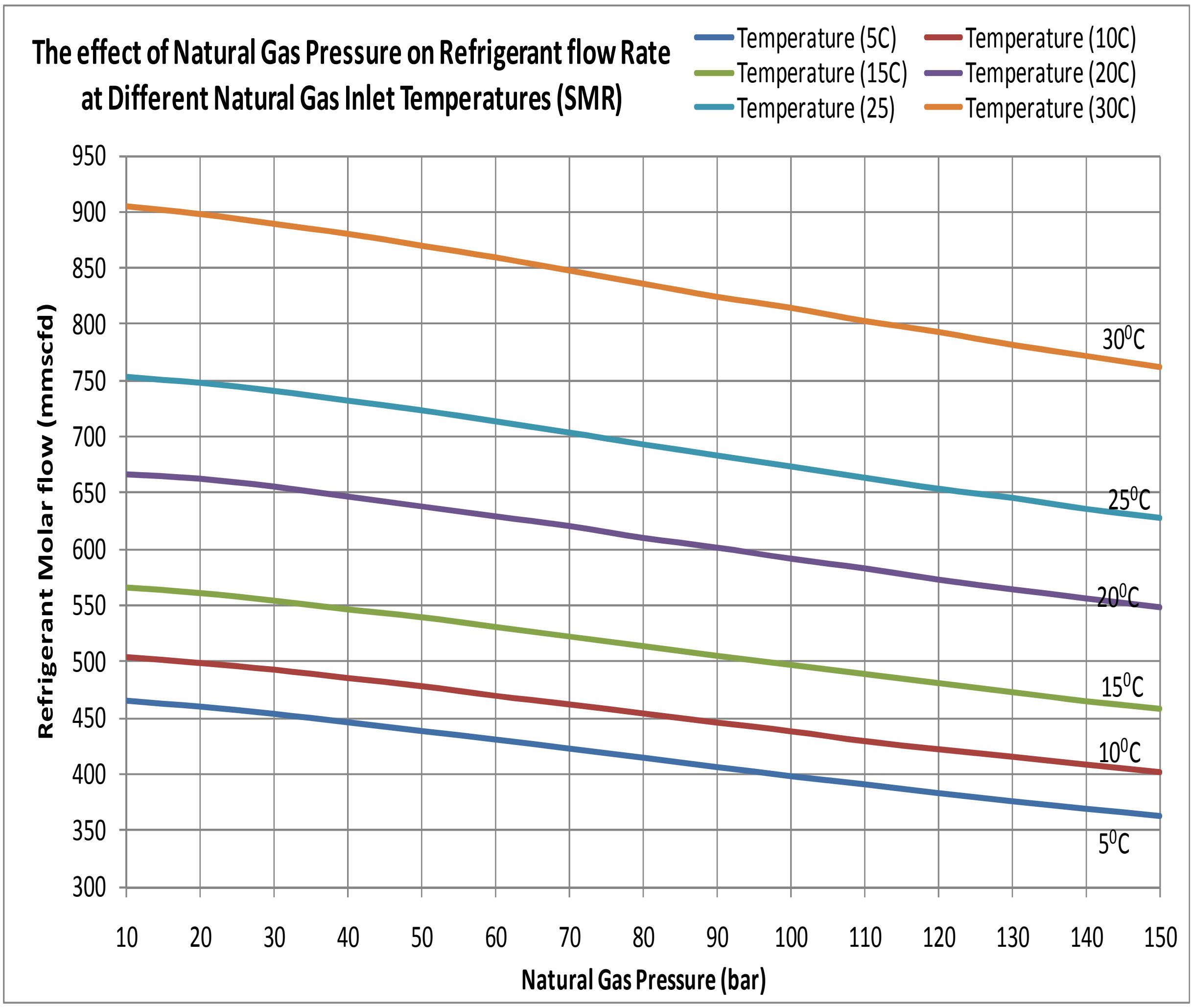

Figure 21: The effect of natural gas pressure on refrigerant flow when natural gas supplied at different temperatures (SMR-PRICO)

5.2.3 The effect of natural gas composition on SMR- PRICO

The effect of natural gas composition on specific power, power consumption and refrigerant flow rate for SMR-PRICO process is presented in figure 22, 23 &24 and table 8, 9&10 below. The natural gas composition used for simulating the benchmarked process is termed as composition-reference and its results were used as reference point to the other compositions. The figures and tables show that composition may have significant effect on specific power, power consumption or refrigerant flow. The effect may be an increase or decrease of the mentioned parameters.

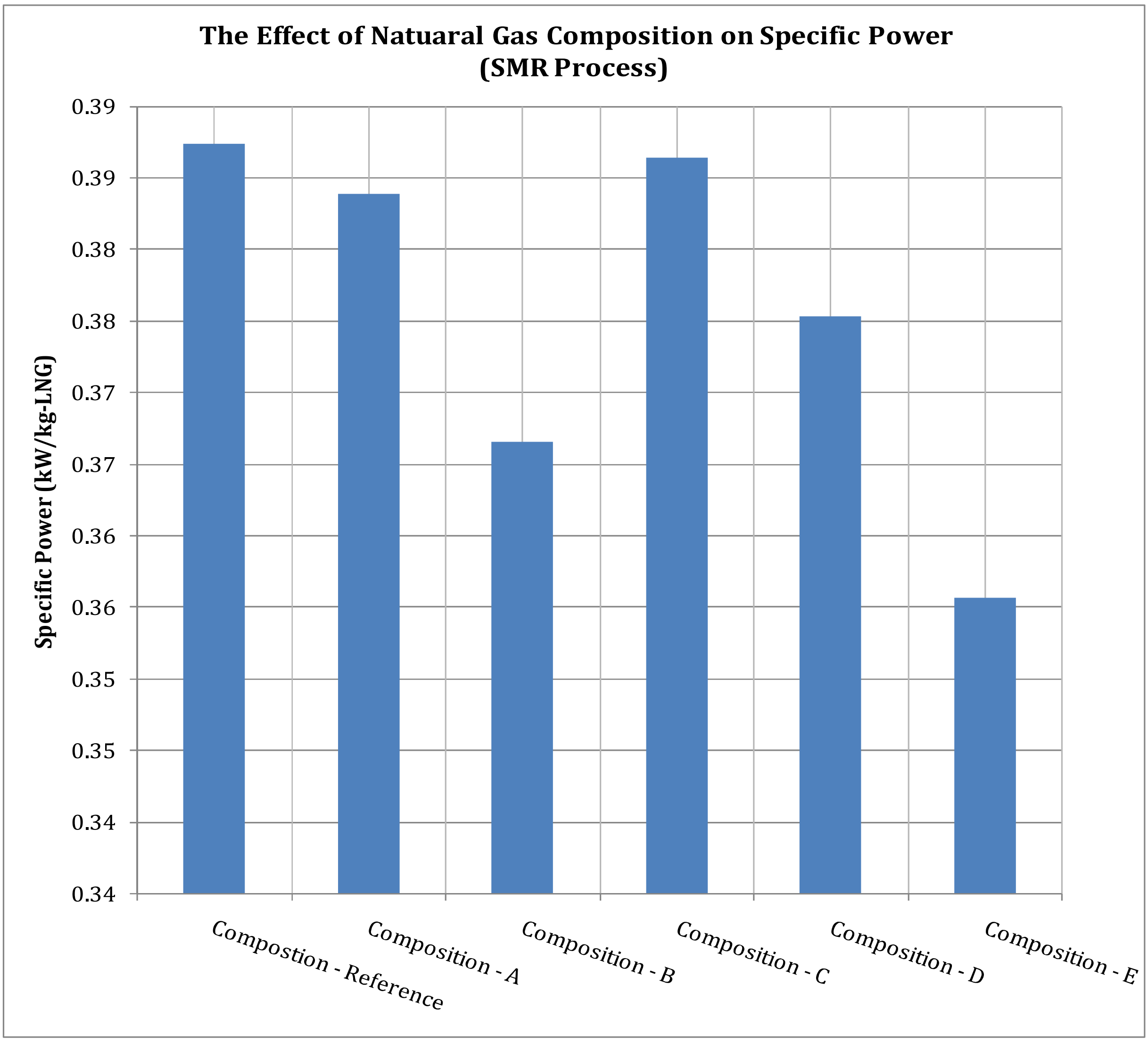

Figure 22: The effect of natural gas composition on specific power (SMR-PROCO)

Table 8: The effect of natural gas composition on specific power SMR process

Composition	Specific power (kWh/kg-LNG)	Increases or decreases
Composition-Reference	0.387	Used as reference for calculation
Composition - A	0.384	Decrease by 0.8%
Composition -B	0.367	Decreases by 5.2%
Composition –C	0.386	Decreases by 0.3%
Composition -D	0.375	Decreases by 3.1%
Composition -E	0.356	Decreases by 8.0%

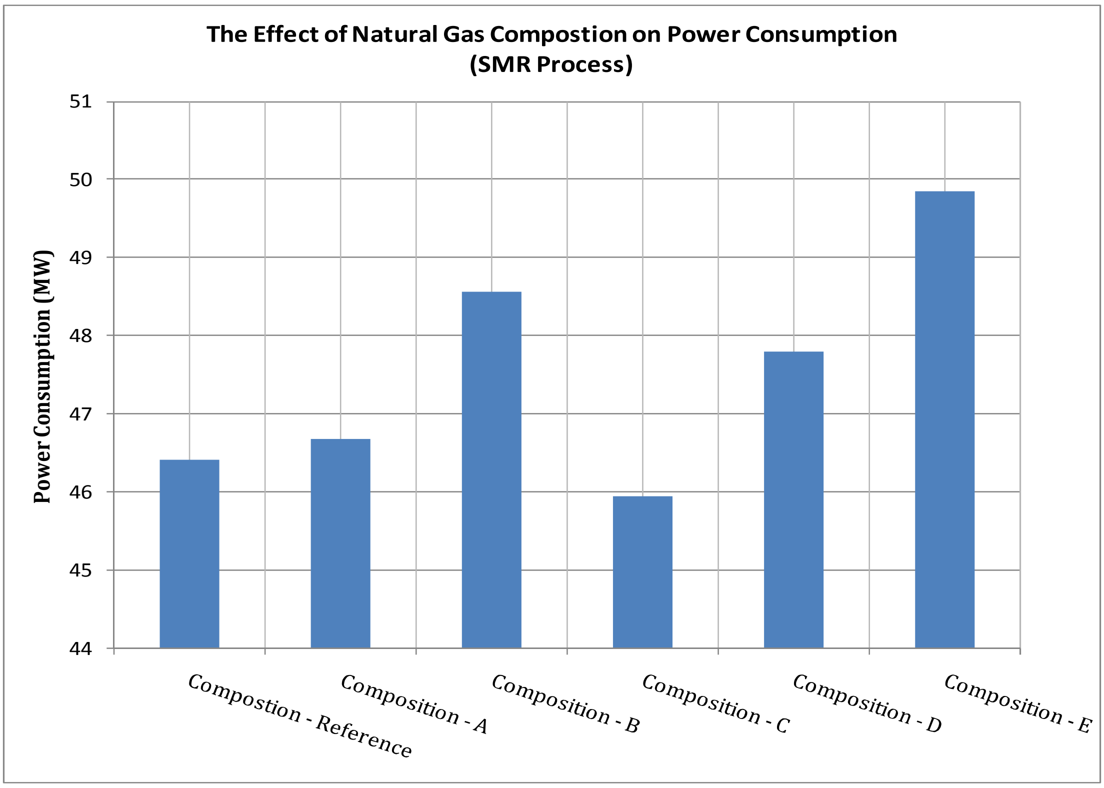

Figure 23: The Effect of natural gas composition on power consumption (SMR-PRICO)

Table 9: The effect of natural gas composition on power consumption SMR process

Composition	Power consumption (MW)	Increases or decreases
Composition-Reference	46.42	Used as reference for calculation
Composition - A	46.68	Increases by 0.6%
Composition -B	48.56	Increases by 4.6%

Composition -C	45.95	Decreases by 1.0%
Composition -D	47.80	Increases by 3.0%
Composition -E	49.85	Increases by 3.4%

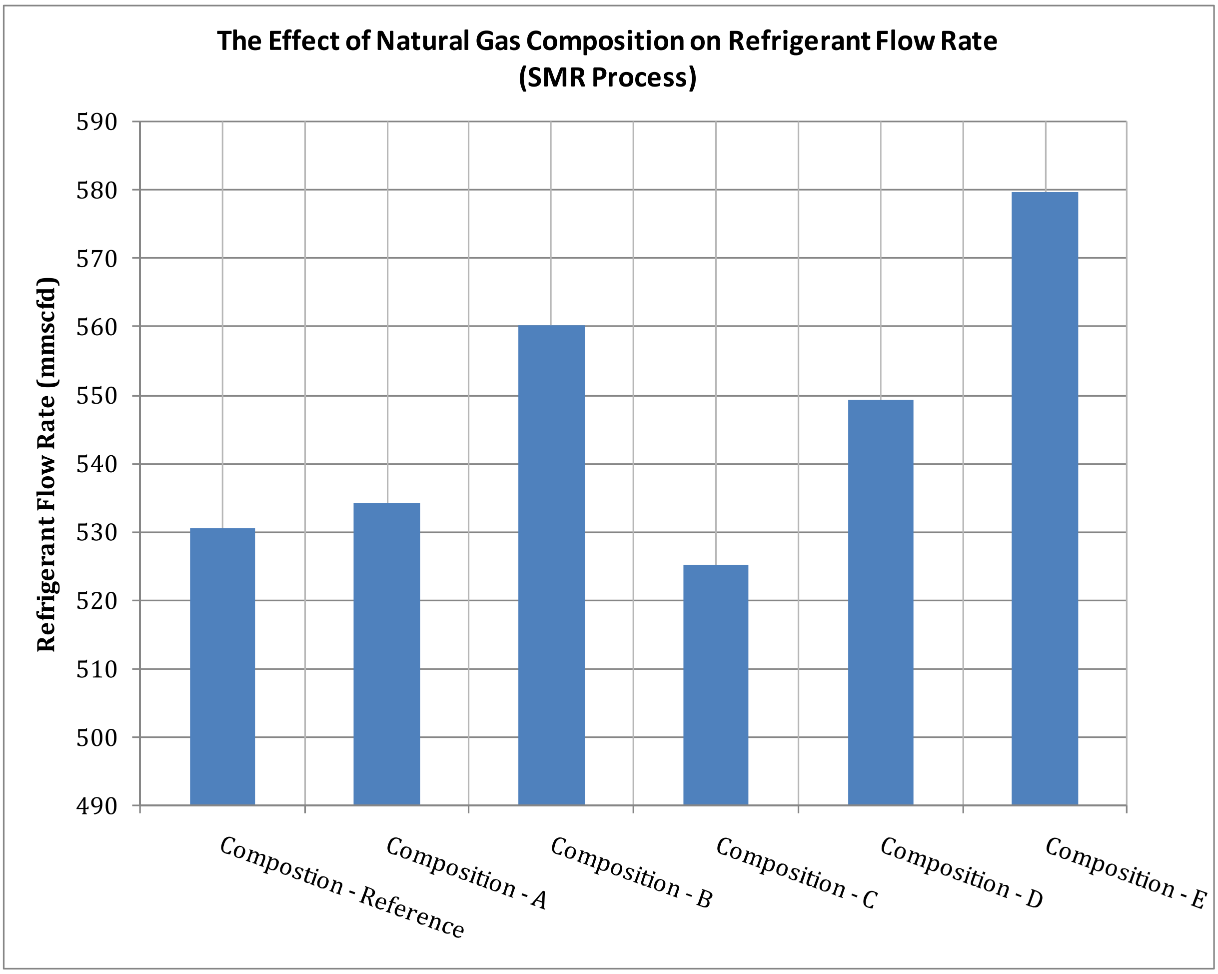

Figure 24: The effect of natural gas composition on refrigerant flow (SMR-PRICO)

Table 10: The effect of natural gas composition on refrigerant flow rate SMR process

Composition	Refrigerant flow (mmscfd)	Increases or decreases
Composition-Reference	530.51	Used as reference for calculation
Composition - A	534.30	Increased by 0.71%
Composition -B	560.17	Increase by 5.6%
Composition -C	525.09	Decreased by 1.0%
Composition -D	549.34	Increased by 3.5%
Composition -E	579.72	Increased by 9.3%

5.2.4 Exegy analysis for SMR-PRICO.

The useful exergy and distribution of exergy losses on different components in the process are as represented in figure-25 below. From figure-25 Useful exergy account for about 26% which indicates that about 74% of energy supplied by the compressor end up as losses in different components in the process liquefaction cycle. The largest loss occurs in LNG heat exchanger and after-cooler (cooler) which amounts to 25 and 24% respectively, the amount in mega watt (MW) of useful energy and losses are presented in table-11 below.

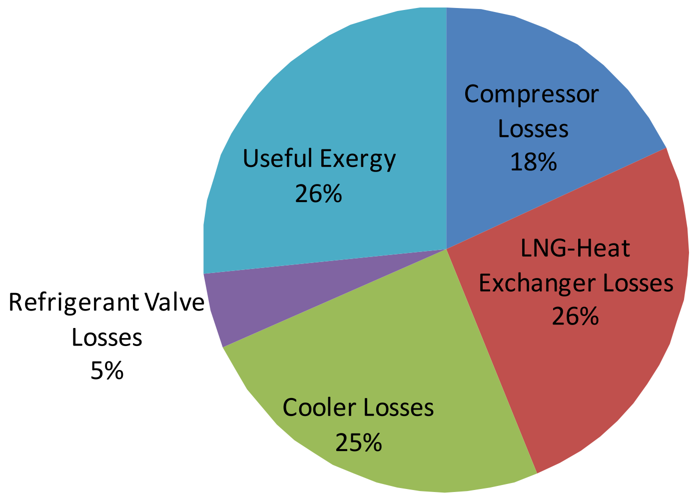

Figure 25: Distribution of Exergy losses on different components and useful exergy (SMR PRICO)

Table 11: Distribution of exergy Losses and useful exergy (SMR process)

Total Exergy Supplied	**46.42**	**MW**
Useful Effect	12.34	MW
Compressor Losses	8.44	MW
LNG-Heat Exchanger Losses	11.91	MW
Cooler Losses	11.41	MW
Refrigerant Valve Losses	2.32	MW

5.2.5 Product quality and production capacity SMR-PRICO

a) Production Capacity

The production capacity of the simulated single mixed refrigerant process was analyzed based on maximum duty of one LM6000 gas turbine which has a capacity of 40.7MW [31]. The assumption was made on natural gas supplied at temperature of 15^0C temperature and flow rate of 150mmscfd at different pressures.

The production capacity per train of single mixed refrigerant process when the process simulated based on conditions above is presented in figure-26 below. Considering natural gas supplied at 60bar the train production capacity is estimated to be 0.87MTPA/train. From simulations, when natural gas supplied at 150mmscf , 60bar and 15^0C LNG production is about 0.95MTPA.

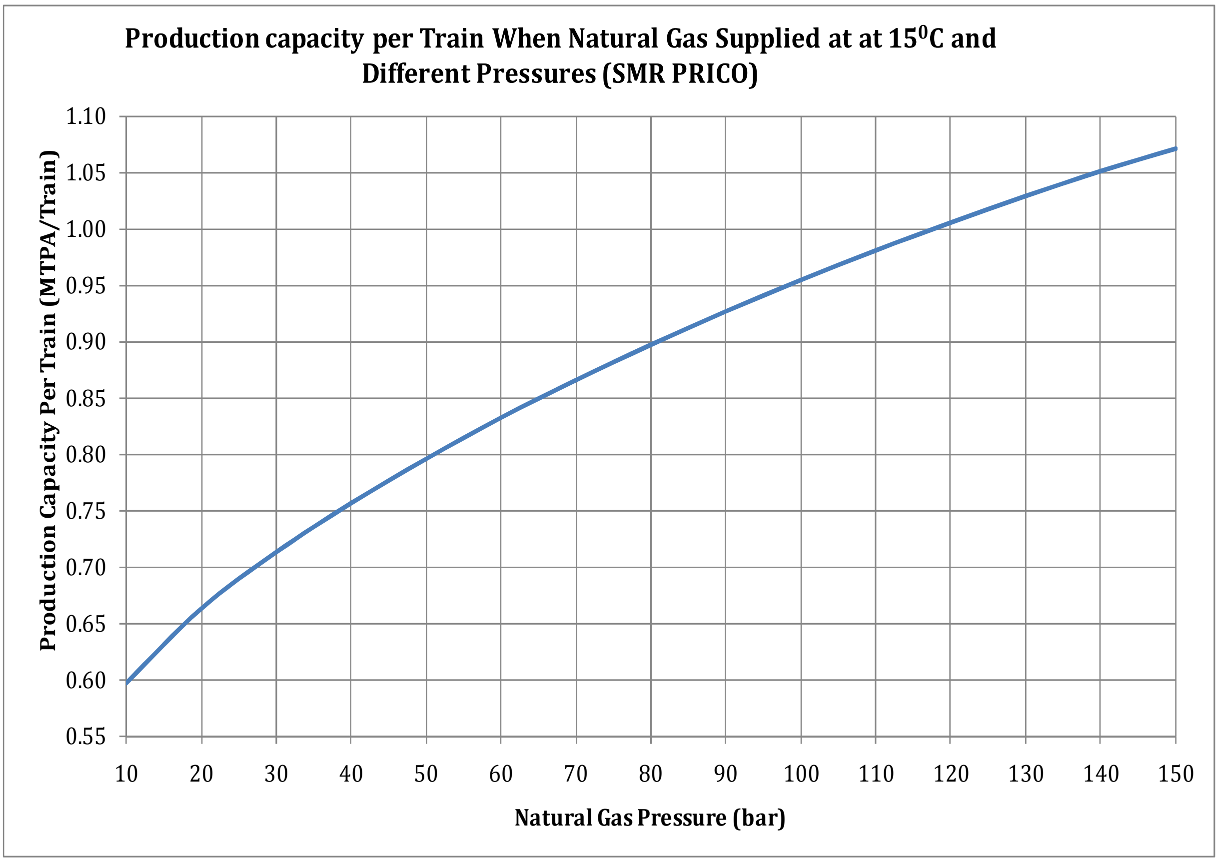

Figure 26: Production capacity of SMR process

b) Product Quality

The quality of LNG product produced by single mixed refrigerant is analyzed on two parameters; high heating value (HHV) and wobbe index. The LNG high heating value was calculated to be 39.59MJ/Sm^3 and wobbe index 51.14. These values are within range as described in detail in appendix-A.

5.3 Summary of simulation results of SMR-PRICO

Table 12: Summary of effect of natural gas supply temperature on SMR-PRICO Process

	Temperature decreased from 15^{0}C to 5^{0}C	**Temperature increased from 15^{0}C to 25^{0}C**
Specific power (kWh/kg-LNG)	Decreased from 0.387 to 0.329 4 which amounts to 14.99%	Increased from 0.387 to 0.539 which amounts to 39.27%
Power consumption (MW)	Decreases from 46.42 to 39.41 which amounts to 15.10%	Increases from 46.42 to 64.61 which amounts to 39.19%
Refrigerant Flow rate (mmscfd)	Decreases from 531.28 to 431.10 which amounts to 18.86%	Increases from 531.28 to 713.04 which amounts to 34.21%

Table 13: Summary of effect of natural gas supply pressure on SMPR-PRICO Process

	Pressure decreased from 60 to 30bar	**Pressure increased from 60 to 90bar**
Specific power (kWh/kg-LNG)	Increases from 0.387 to 0.452 which amounts to 16.80%	Decreases from 0.387 to 0.348 which amounts to 10.08 %
Power consumption (MW)	Increases from 46.42 to 54.59 which amounts to 17.60%	Decreases from 46.42 to 41.40 which amounts to 10.81%
Refrigerant Flow rate (mmscfd)	Increases from 531.28 to 554.66 which amounts to 4.4%	Decreases from 531.28 to 505.58 which amounts to 4.84%

6. Simulation of Dual Mixed Refrigerant (DMR) Process

6.1 Detailed description of DMR process

Dual mixed refrigerant (DMR) process consist of two stages refrigerant low and high. The low stage refrigerant represents pre-cooling stage while the high level refrigerant represents the liquefaction and sub-cooling stage. The process flow diagram is presented as figure-27 below. The basis for designing of system was based on the patent with reference [20].

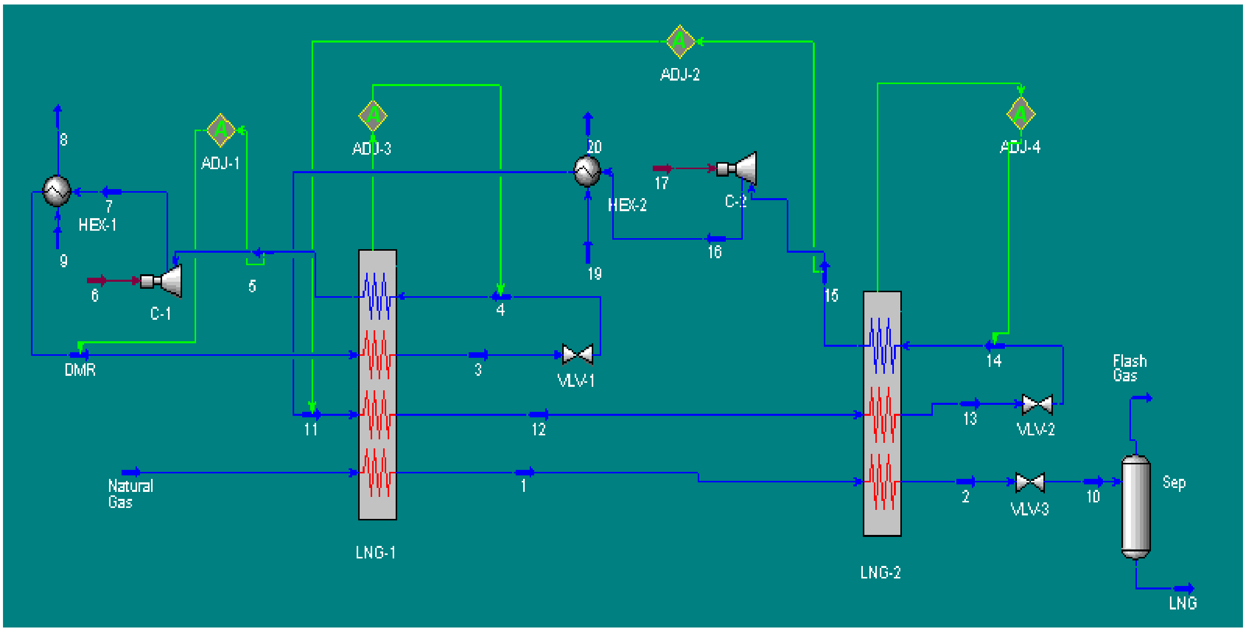

Figure 27: Process flow diagram for dual mixed refrigerant (DMR)

Feed stream **(natural gas)** at 15^{0}C and pressure of 60bar is subjected to initial cooling by LNG heat exchanger **(LNG-1)** against low temperature refrigerant **(stream-4).** The cooled natural gas exits the first LNG heat exchanger at -50.8^{0}C **(stream-1).** The cooled natural gas **(stream-1)** enters the second LNG heat exchanger **(LNG-2)** for liquefaction and sub-cooling. The liquefied natural gas exits the second heat exchanger at -155^{0}C **(stream-2)**. The liquefied natural gas **(stream-2)** is reduced in pressure from 60 to 1.05bar (**stream-10)** via LNG valve **(VLV-3)**. Then low pressure LNG (**stream-10)** is flashed in phase separator **(Sep).** The liquid phase of natural gas (LNG) is removed as bottom product (**steam-LNG**) and flash gas can be treated as fuel gas or otherwise.

The low level or pre-cooling refrigerant passes only in the first stage LNG heat exchanger (**LNG-1)**. The low level refrigerant **(stream-DMR)** enters the first stage LNG heat exchanger at 15^{0}C and 40bar and its temperature is reduced to -50.8^{0}C (**stream-3)** by low temperature refrigerant (**steam-4)** . The cooled refrigerant (**stream-3)** is reduced in pressure from 40 to 3.069bar and its temperature decreases from -50.8^{0}C to -53.71^{0}C respectively via throttling through the valve **(VLV-1)**. The low pressure and temperature refrigerant **(stream-4)** enters LNG heat exchanger **(LNG-1)** and is evaporated by absorbing heat from three streams, warm low level refrigerant **(stream-DMR),** warm high level refrigerant **(stream-11)** and natural gas **(stream-natural gas) .** The vaporized refrigerant **(stream-5)** at 10^{0}C returns to the compressor **(C-1)** , pressure and temperature are increased to 40bar and 151.1^{0}C respectively. The high pressure and temperature refrigerant is cooled to 15^{0}C in heat exchanger **(HEX-1)** and enters LNG heat exchanger **(LNG-1)** as warm refrigerant **(stream-DMR)**. The low level cycle is repeated as described above. The refrigerant flow rate for low level cycle is 181mmscfd.

The high level refrigerant at 30bar and 15^{0}C **(stream-11)** enters the first stage LNG heat exchange **(LNG-1)** and cooled to -50.8^{0}C **(stream-12)** by low temperature refrigerant (**stream-4).** The pre-cooled high level refrigerant **(stream-12)** enters the second stage LNG heat exchanger **(LNG-2)** and is further cooled to -155^{0}C **(stream-13).** The pressure of **(stream-13)** is reduced to 2.99bar and its temperature decreased to -157.9^{0}C **(stream-14)**. The low temperature high level refrigerant **(stream-14)** enters the second LNG heat **(LNG-2)** exchanger and is vaporized by absorbing heat from two streams, pre-cooled high level refrigerant **(stream-12)** and natural gas **(stream-1).** The vaporized high level refrigerant **(stream-15)** returns to the compressor **(C-2),** pressure and temperature are increased to 30bar and 108.7^{0}C respectively. The high pressure and temperature refrigerant **(stream-16)** is cooled in heat exchanger **(HEX-2)** to 15^{0}C **(stream-11)**. Stream-11 enters first stage LNG heat exchanger **(LNG-1)** and the cycle is repeated as described above. The refrigerant flow for high level cycle was 224.4mmscfd.

The suction flow rate of low level refrigerant compressor is 66,078.4m^3/hr which gives a mass flow rate of 94.26kg/s, while for the high level refrigerant suction flow rate is 65,128.9 m^3/hr which gives a mass flow rate of 74.56kg/s.

The compositions of low and high stage refrigerant are as presented below in table-14 and 15 respectively. The high level refrigerant cools natural gas from -50.8^0 C to -155^{0}C hence does not contain propane or other heavy hydrocarbon as they have high boiling point. The exact concentration of the various components is dependent upon the ambient conditions, the composition of feed natural gas and the temperature of external cooling fluids which are used in the liquefaction plant. Also it depends on the exact power shift or balance desired between the two cycles (cooling and liquefaction cycle).

Table 14: Low level refrigerant composition (DMR process)

Compound	Ethane	Propane	i-Butane	n-Butane	Total
Mole fraction	0.63	0.20	0.06	0.11	1.0

Table 15: High level refrigerant flow rate (DMR process)

Compound	Methane	Ethane	Nitrogen	Total
Mole Fraction	0.42	0.50	0.08	1.0

6.2 Simulation results for dual mixed refrigerant process (DMR)

The hysys simulation of DMR process was based on detailed descriptions as presented on section 6.1 above. In the simulation LNG heat exchangers "UA" values was calculated to be 5,213.25kW/^{0}C for LNG heat exchanger LNG-1 and 4,691.53kW/^{0}C for LNG heat exchanger LNG-2 at minimum approach temperature of 2.9^{0}C for both LNG heat exchangers and LMTD of 9.4 and 10.2^{0}C for LNG-1 and LNG-2 respectively. The specific power of the process based on the simulated conditions was calculated to 0.353kWh/kg-LNG. The calculated "UA" values of the LNG heat exchangers were kept constant during simulation by varying refrigerant flow rate and refrigerant evaporating pressure.

6.2.1 The effect of natural gas temperature on DMR process

a) The effect of natural gas temperature on specific power

The effect of natural gas temperature on specific power when natural gas is supplied at pressure of 60bar for DMR process is presented in figure-28 below. The figure shows that when natural gas temperature is decreased from 15^{0}C to 5^{0}C the specific power of DMR process decreases from 0.353 to 0.330kWh/kg-LNG which amounts to a 6.51% decrease and when the supply temperature increase from 15^{0}C to 25^{0}C, specific power increases from 0.353 to 0.382kWh/kg-LNG which amounts to an 8.23% increase. Generally it can be shown that when natural gas is supplied at a given pressure, the specific power decreases with decrease in natural gas inlet temperature and it increases with increase in natural gas inlet temperature.

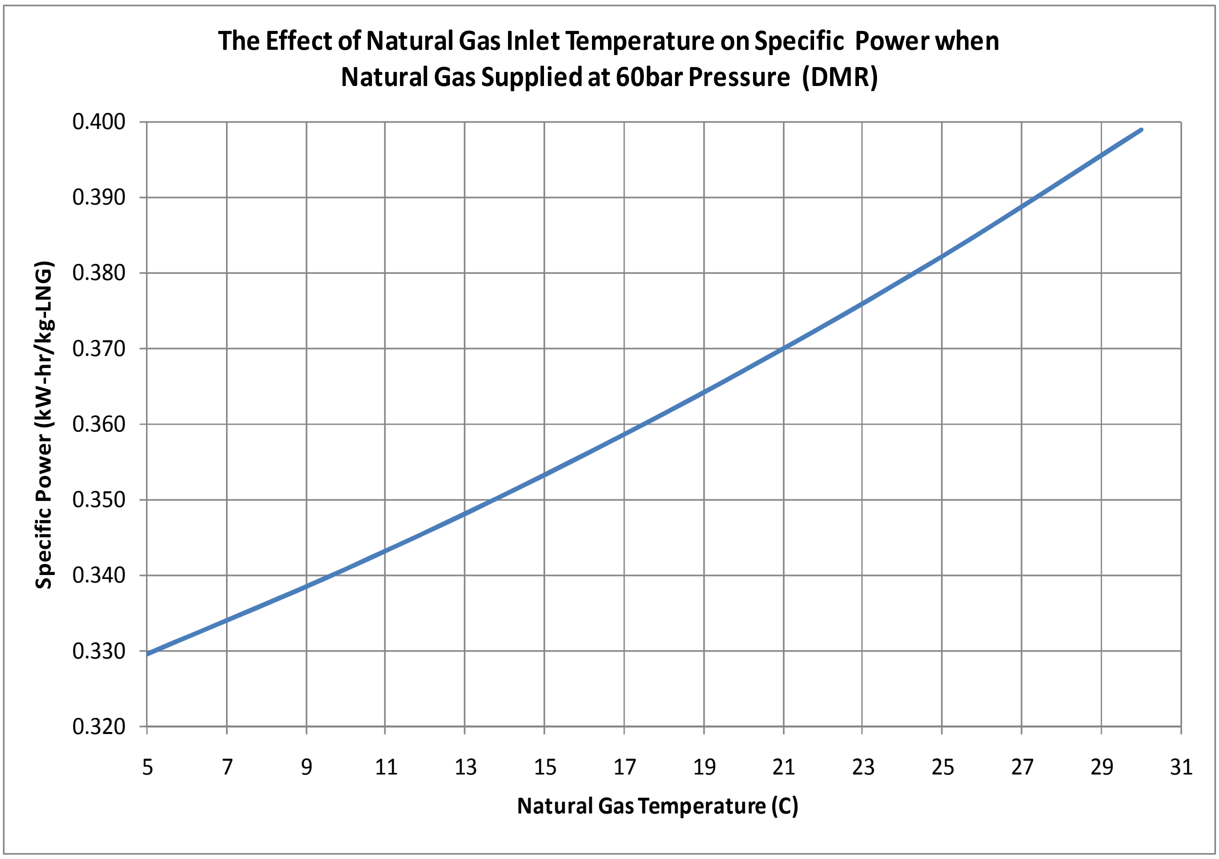

Figure 28: The effect of natural gas temperature on specific power when natural gas supplied at 60bar (DMR Process)

The effect of natural gas temperature on specific power when natural gas is supplied at different pressures (20, 40, 60, 80, 100, 120 and 140bar) in DMR process is presented in figure-29 below. The figure shows that DMR process specific power increases as natural gas supply temperature increases and decreases as natural gas supply temperature decreases irrespective of natural gas supply pressure. But also it can be depicted that the higher the natural gas supply pressure (eg graph with 140bar) the lower the specific power and the lower the natural gas supply pressure (eg graph with 20bar) the higher is the specific power at any natural gas supply temperature. For example natural gas supplied at (15^0C, 60bar) and (15^0C, 120bar) has specific power of 0.353 and 0.282kWh/kg-LNG which is a difference of 20.11% with reference to 60bar.

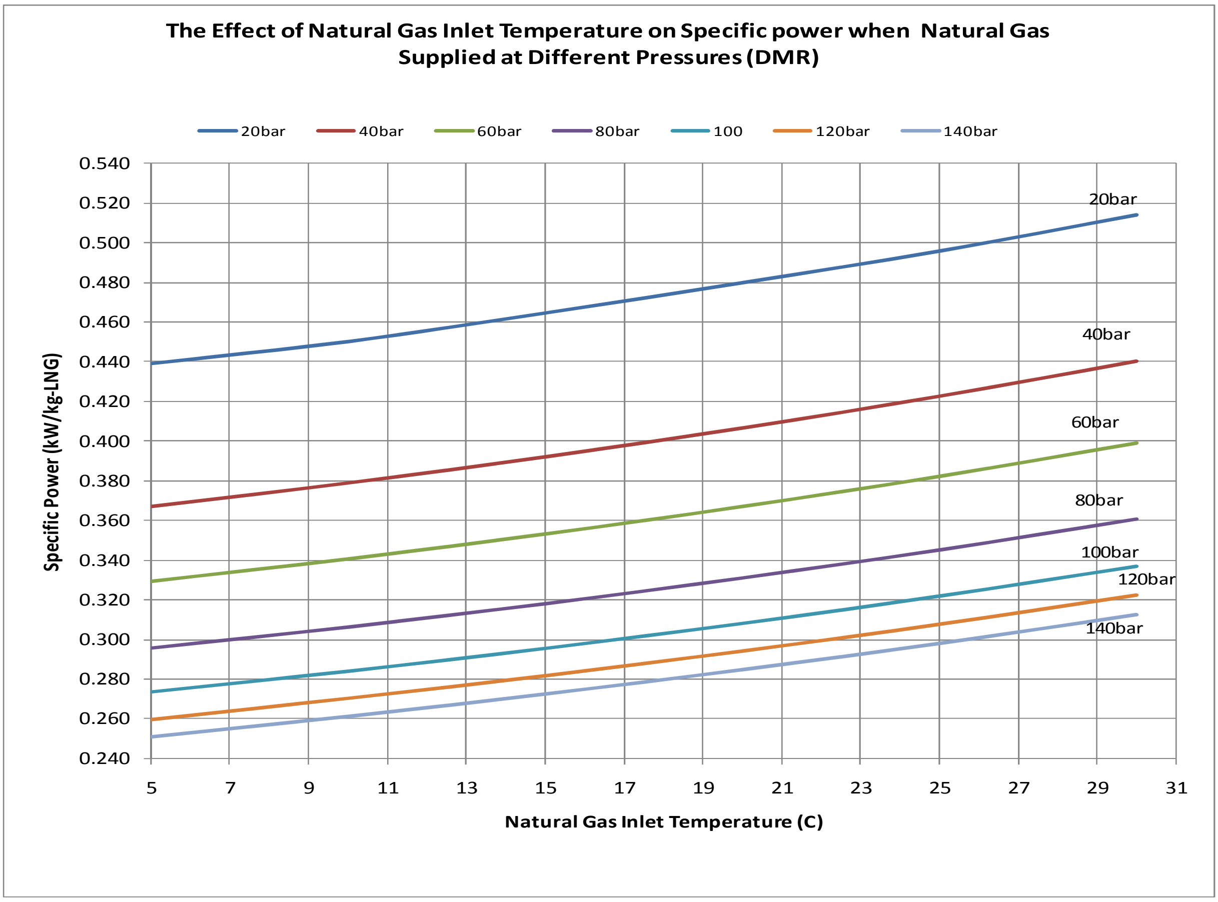

Figure 29: The effect of natural gas inlet temperature on specific power when natural gas supplied at different pressures (DMR Process)

b) The effect of natural gas temperature on power consumption

The effect of natural gas temperature on power consumption when natural gas is supplied at 60bar in DMR process is presented in figure-30 below. The graph shows that as temperature decreases from 15^0C to 5^0C, the power consumption of DMR process decreases from 42.34MW to 39.50MW this amounts to a 6.7% decrease and when the supply temperature increases from 15^0C to 25^0C, power consumption increases from 42.34 MW to 45.80 MW which amounts to an 8.2% increase. It can be shown that when natural gas is supplied at a given pressure and temperature, decreases in natural gas supply temperature will decrease power consumption and an increase in natural gas supply temperature will increase power consumption.

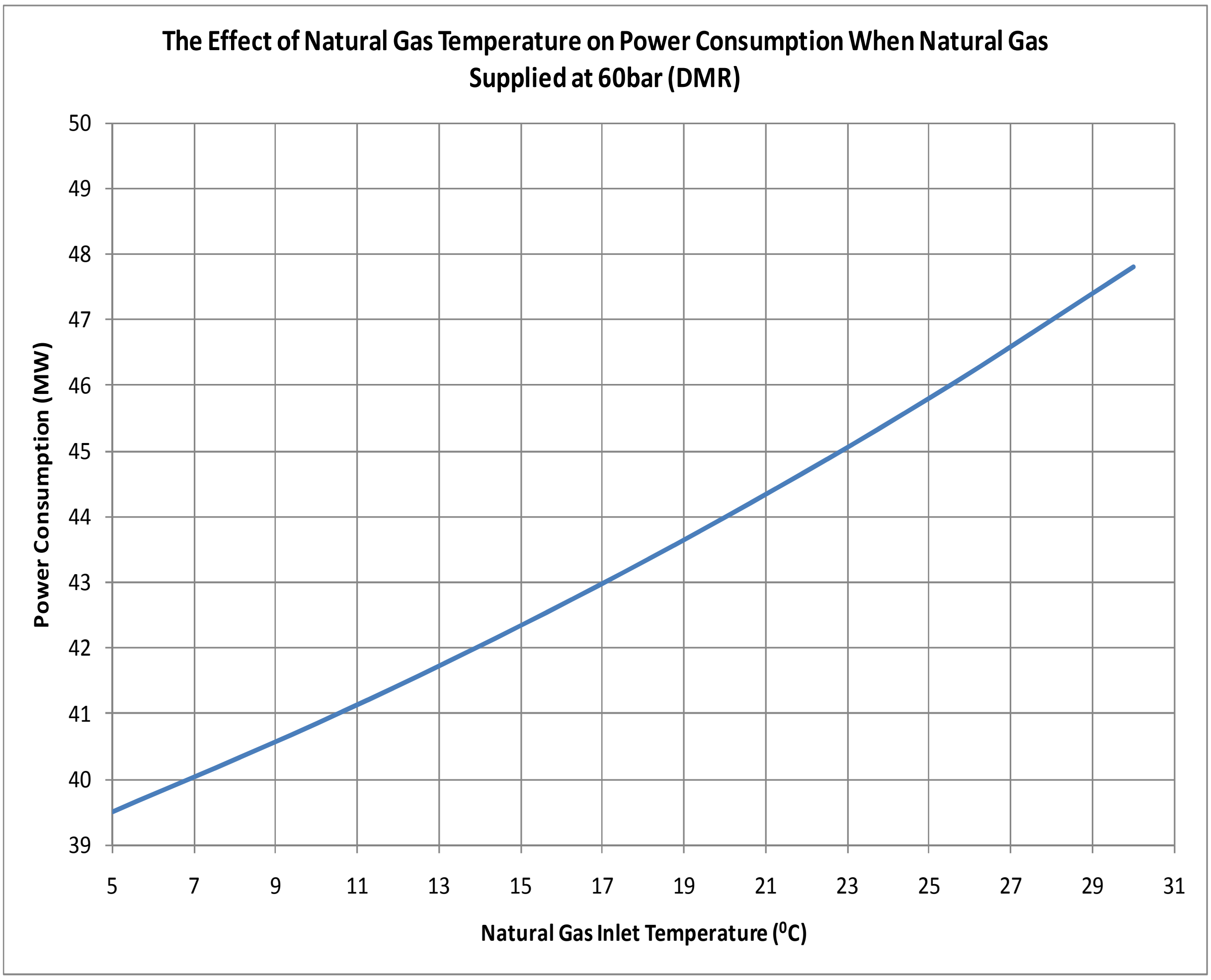

Figure 30 : The effect of natural gas temperature on power consumption when natural gas supplied at 60bar (DMR Process)

The effect of natural gas inlet temperature on power consumption when natural gas supplied at different pressures (20, 40, 60, 80, 100, 20 and 140bar) in DMR process is presented in figure-31 below. The figure shows that at any natural gas supply pressure, when natural gas supply temperature decreases from 15^0C to 5^0C, power consumption of DMR process decreases and when natural gas supply temperature increases from 15^0C to 30^0C the power consumption increases. But also the figure shows that natural gas supplied at higher pressure has low power consumption (eg. graph with 140bar pressure) and when supplied at low pressure, power consumption is higher (eg graph with 20bar pressure) at any natural gas supply temperature. For example consider natural gas being supplied at (15^0C,60bar) and (15^0C,120bar) its power consumption is 42.34 and 33.27MW which is a difference of 21.42% with reference to 60bar.

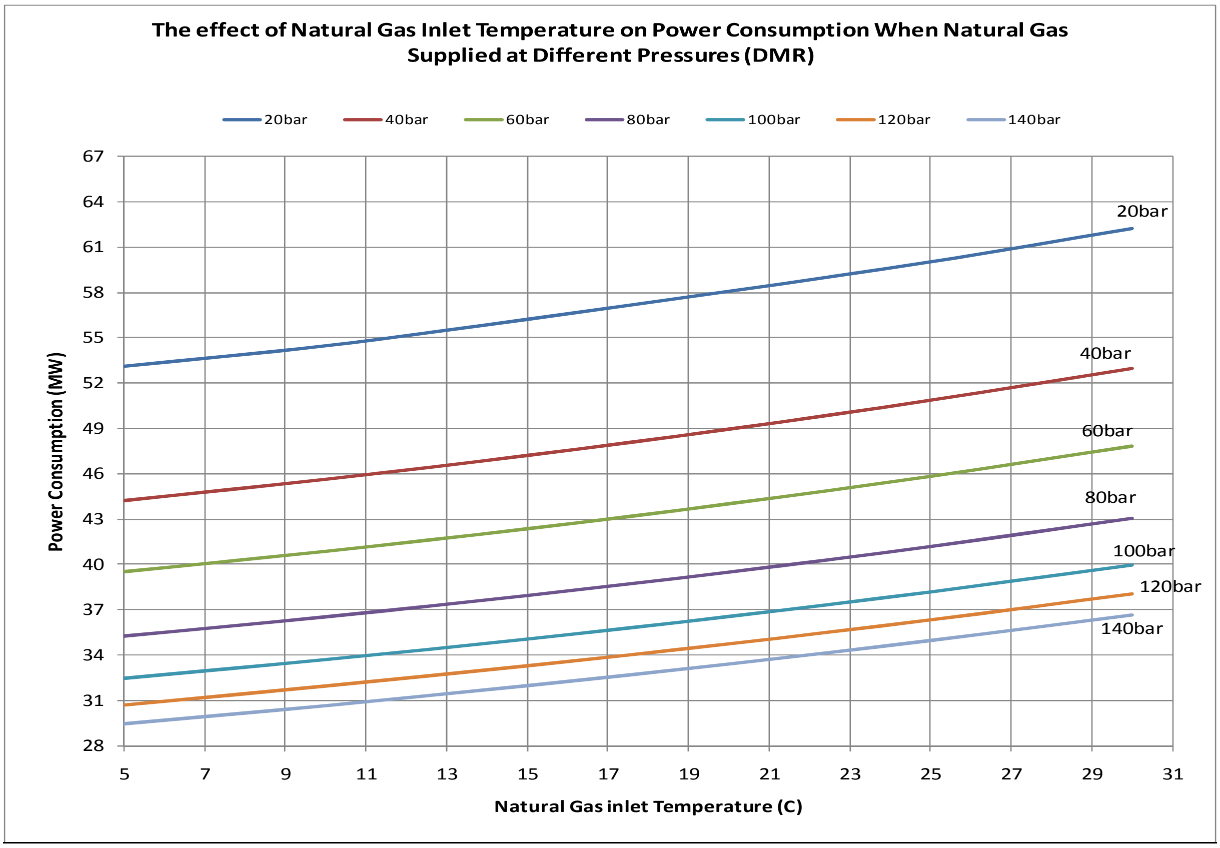

Figure 31: The effect of natural gas temperature on power consumption when natural gas supplied at different temperatures (DMR Process).

c) The effect of natural gas inlet temperature on refrigerant flow rate

The effect of natural gas inlet temperature on refrigerant flow rate in DMR process is presented on figure-32 below. The figure shows that when natural gas supply temperature decreases from 15^{0}C to 5^{0}C the DMR process refrigerant flow rate decreases from 405.67 to 385.78 mmscfd which amounts to a 4.9% decrease and when supply temperature increases from 15^{0}c to 25^{0}C, its refrigerant flow rate increases from 405.67 to 427.62 mmscfd which amounts to 5.41% increase. It can be shown that when natural gas is supplied at given pressure, flow rate will decrease with decrease in natural gas inlet temperature and increase with increase in natural gas inlet temperature.

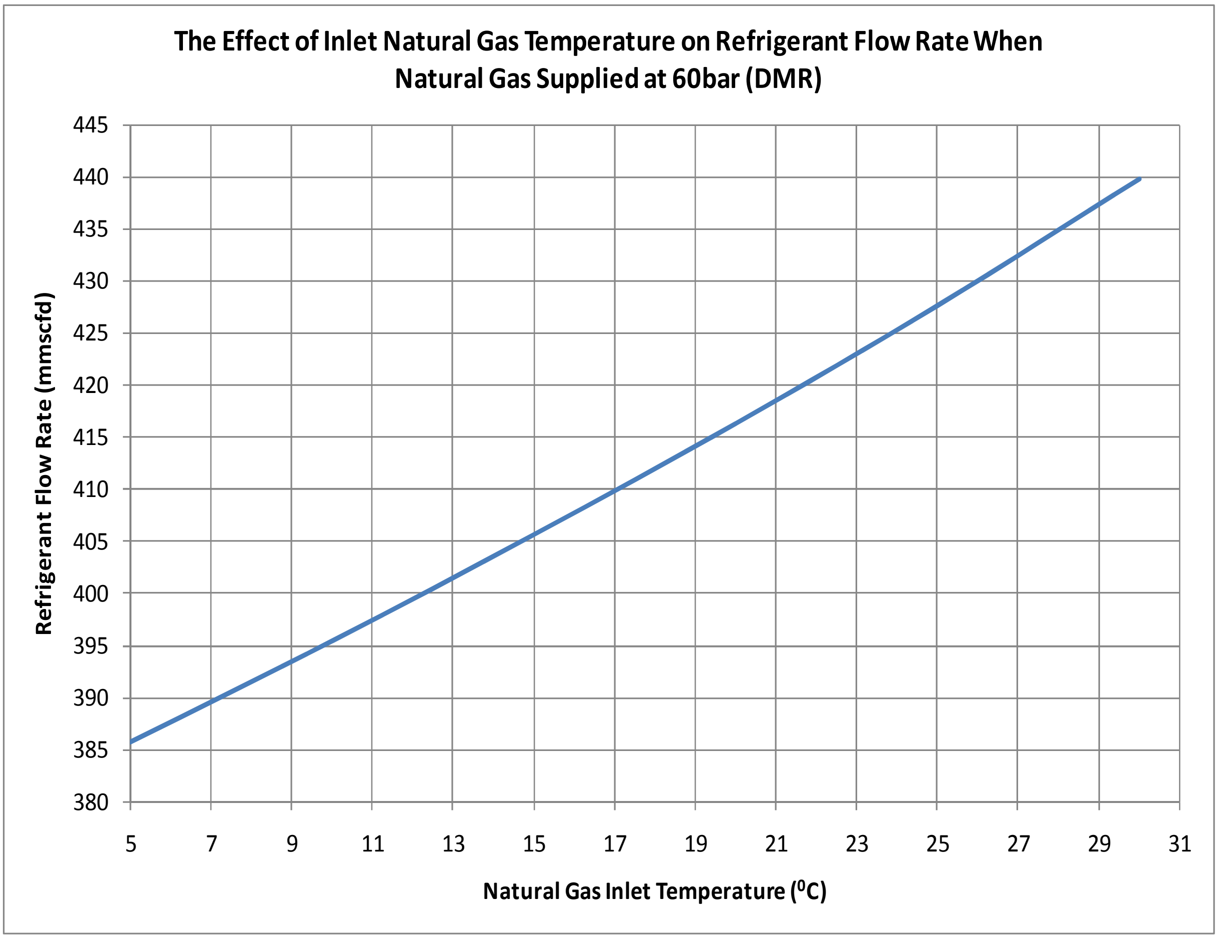

Figure 32 : The effect of natural gas inlet temperature on refrigerant flow rate when natural gas supplied at 60bar (DMR Process)

The effect of natural gas inlet temperature on refrigerant when natural gas is supplied at different pressures (20, 40, 60, 80,100, 120 and 140bar) in DMR process is given in figure-33 below. The graph shows that as natural gas inlet temperature decreases from 15⁰C to 5⁰C refrigerant flow rate decreases and as supply temperature increases from 15⁰C to 25⁰C refrigerant flow rate increases at any natural gas supply pressure. But also the figure shows that when natural gas is supplied at higher pressure refrigerant flow rate is low (eg graph with 140bar) and when natural gas is supplied at low pressure, refrigerant flow rate is higher (e.g graph with 20bar) at any natural gas supply temperature. For example when natural gas is supplied at (15⁰C, 60bar) and (15⁰, 120bar) refrigerant flow rate is 405.67 and 318.81 mmscfd which differs by 21.41% with respect to 60bar.

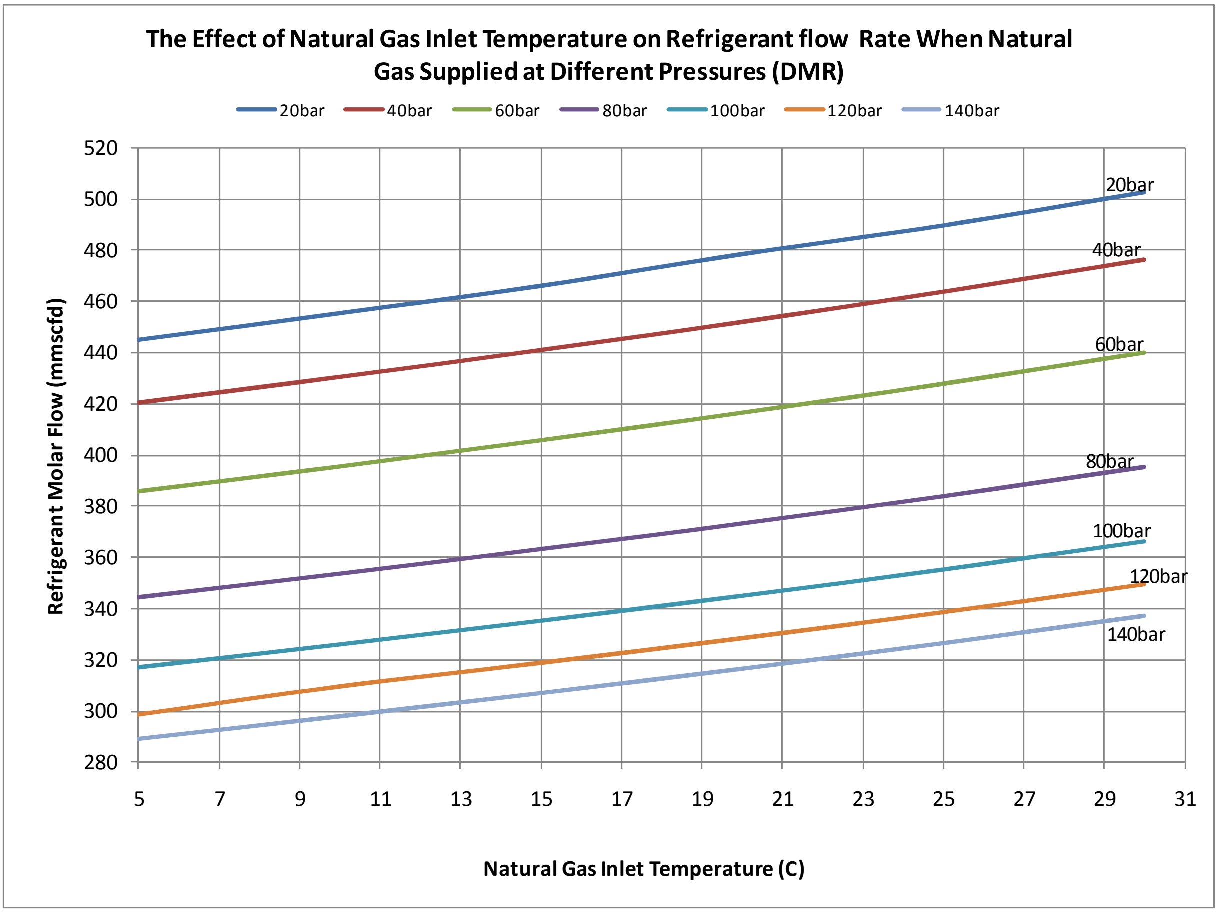

Figure 33: the effect of natural gas inlet temperature on power consumption when natural gas supplied at different temperatures (DMR Process)

6.2.2 The effect of natural gas pressure on DMR Process

a) The effect of natural gas pressure on specific power

The effect of natural gas pressure on specific power when natural gas is supplied at 15^{0}C in DMR process is presented in figure-34 below. When natural gas pressure decreases from 60 to 30bar, the DMR process specific power increases from 0.353 to 0.419kWh/kg-LNG which amounts to an 18.70% increase and when natural gas supply pressure increases from 60 to 90bar, its specific power decreases from 0.353 to 0.305 kWh/kg-LNG which amounts to a 13.60% decrease. In general it can be shown that as natural gas supply pressure increases, specific power decreases and as natural gas supply pressure decreases, specific power increases at given temperature.

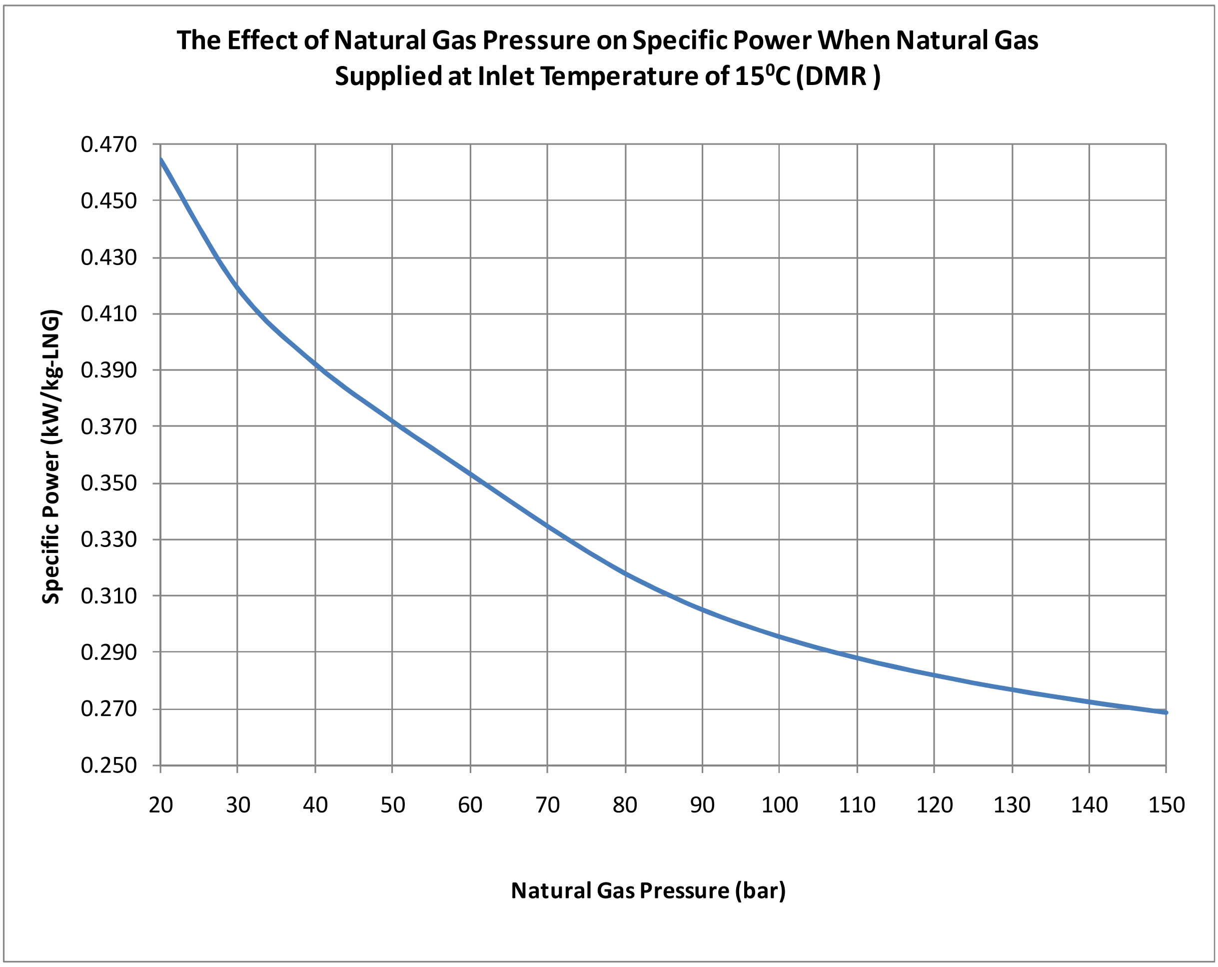

Figure 34: The effect of natural gas pressure on specific power when natural gas supplied at 60bar (DMR Process)

The effect of natural gas pressure on specific power when natural gas is supplied at different temperature (5, 10, 15, 20, 25 and 30^{0}C) in DMR process is given in figure-35. The figure shows that as natural gas pressure increases from 60 to 150bar DMR process specific power decreases and as pressure decreases from 60 to 20bar, specific power increases at any natural gas supply temperature. But also the figure shows that the higher the natural gas supply temperature in DMR process with reference to its simulated conditions the higher the specific power (eg graph with 30^{0}C) and the lower the temperature the lower the specific power at any natural gas supply pressure. For example natural gas supplied at (15^{0}C, 60bar) and (30^{0}C, 60bar) has specific power of 0.353 and 0.399 kWh/kg-LNG respectively which differs by 13.03% with reference to 15^{0}C.

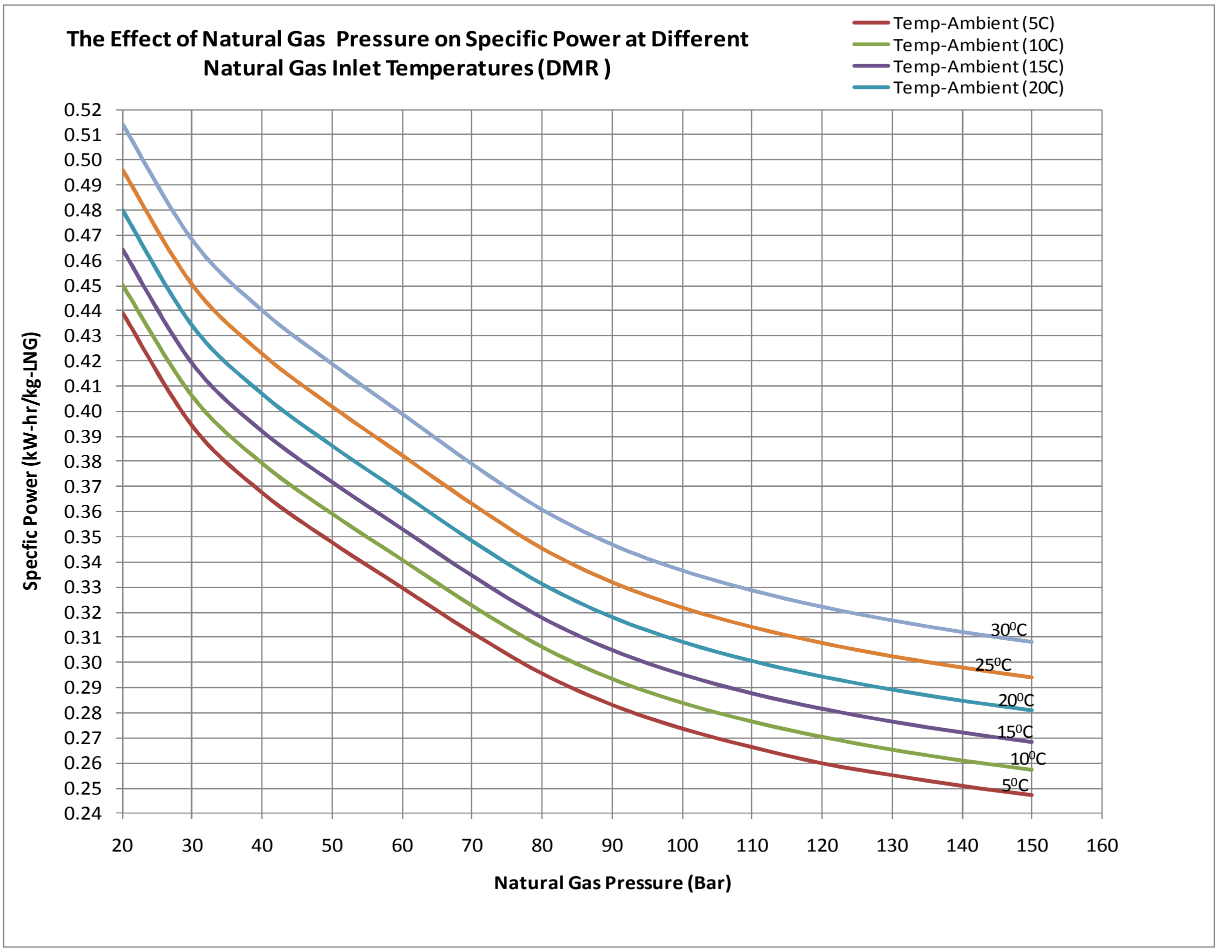

Figure 35 : The effect of natural gas pressure on specific power when natural gas supplied at different temperatures (DMR Process)

b) The effect of natural gas pressure on power consumption

The effect of natural gas pressure on power consumption when natural gas is supplied at 15^{0}C in DMR process is shown in figure-36 below. The figure indicate that as pressure increases from 60 to 150 bar, power consumption of DMR process decreases and as pressure decreases from 60 to 20bar power consumption increases. For example when natural gas supply pressure increases from 60 to 90bar, power consumption decreases from 42.34 to 36.30MW which amounts to a 14.27% decrease and when natural gas supply pressure decreases from 60 to 30bar power consumption increases from 42.34 to 50.57MW which amounts to a 19.44% increase.

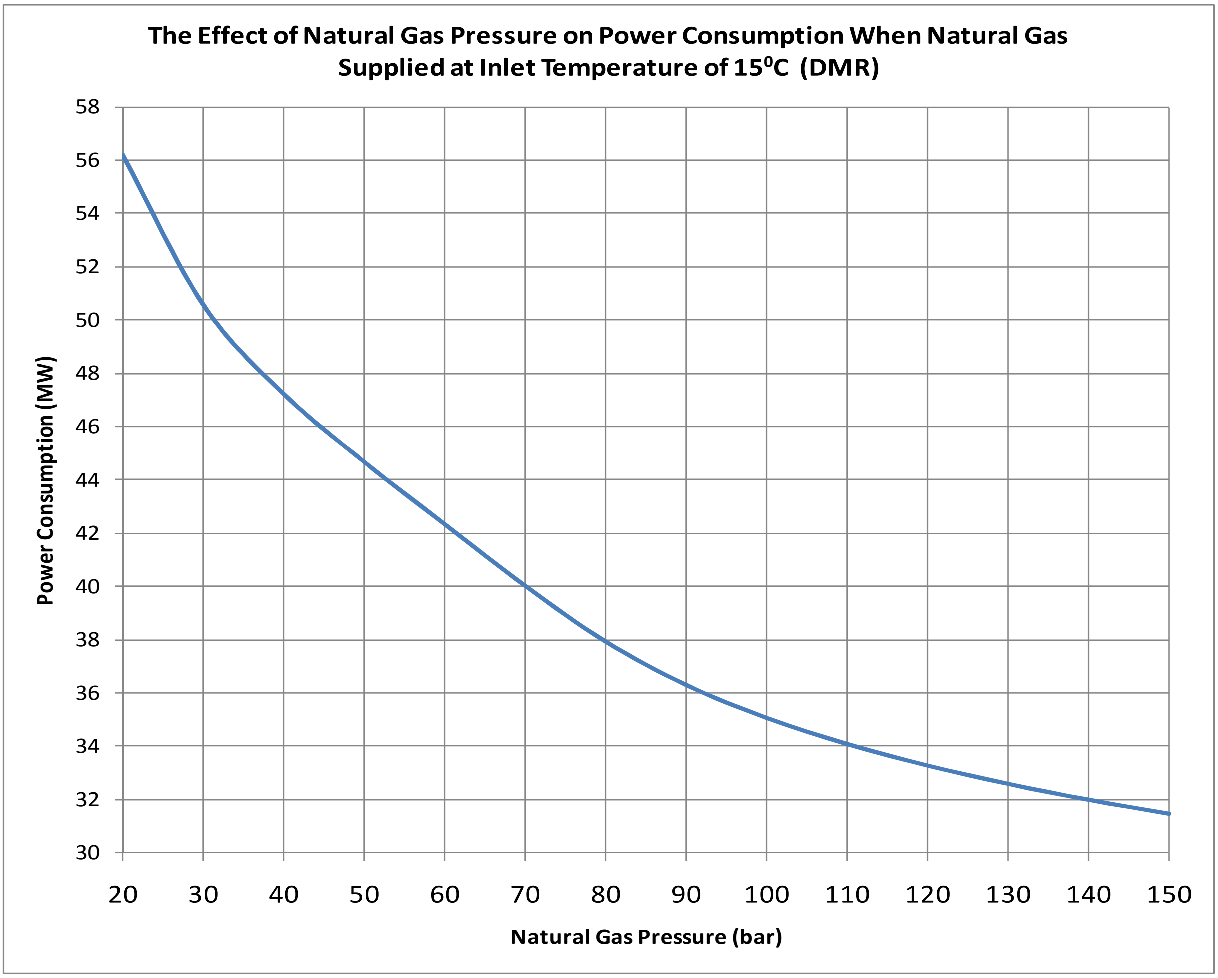

Figure 36: The effect of natural gas pressure on power consumption when natural gas supplied at 15^{0}C (DMR Process)

The effect of natural gas pressure on power consumption when natural gas is supplied at different temperatures (5, 10, 15, 20, 25 and 30^{0}C) in DMR process is shown in the figure-37 below. The figure indicates that as natural gas supply pressure increases from 60 to 150bar power consumption of DMR process decreases and as supply pressure decreases from 60bar to 20bar, power consumption increases at any natural gas supply temperature. With reference to the simulated process conditions, the graphs indicates that natural gas supplied at higher temperature has higher power consumption (eg graph with 30^{0}C) and natural gas supplied at low temperature has low power consumption (eg. graph with 5^{0}C) at any natural gas supply pressure. For example natural gas supplied at (15^{0}C, 60bar) and (30^{0}C, 60bar) has power consumption of 42.34 and 47.80 MW which is a difference of 12.90% with reference to 15^{0}C.

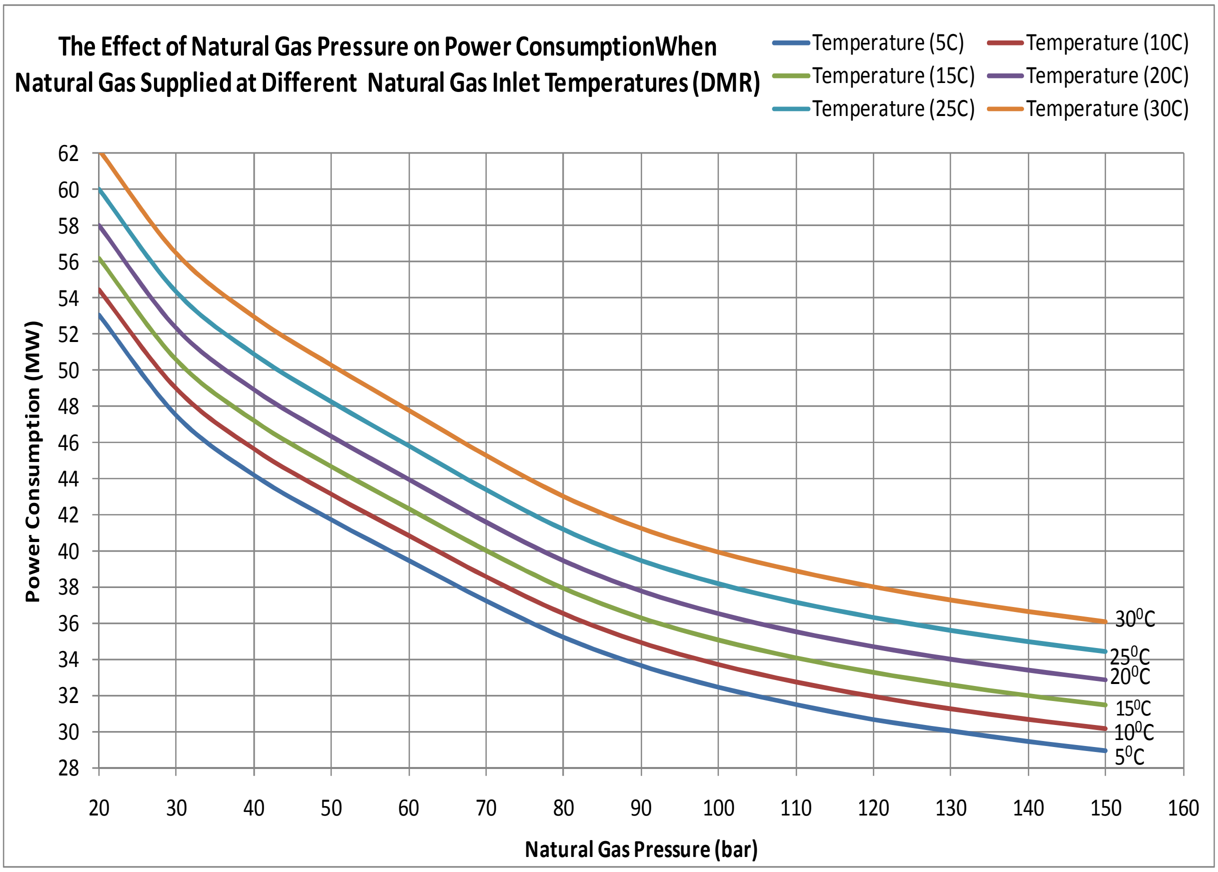

Figure 37: The effect of natural gas pressure on power consumption when natural gas supplied supplied at different temperature (DMR Process)

C) The effect of natural gas pressure on refrigerant flow rate

The effect of natural gas pressure on refrigerant flow rate when natural gas is supplied at 15^{0}C in DMR process is presented in the figure-38 below. With reference to natural gas supplied at 60bar to DMR process, the figure shows that as natural gas supply pressure increases from 60bar to 150bar, refrigerant flow rate decreases and as natural gas supply pressure decreases from 60 to 20bar refrigerant flow rate increases. For example as natural gas increases from 60 to 90bar refrigerant flow rate decreases from 405.67 to 347.18 mmscfd which amounts to a 14.42% decrease and as natural gas supply pressure decreases from 60 to 30 bar, refrigerant flow rate increases from 405.67 to 454.70 mmscfd amounts to a 12.09% increase.

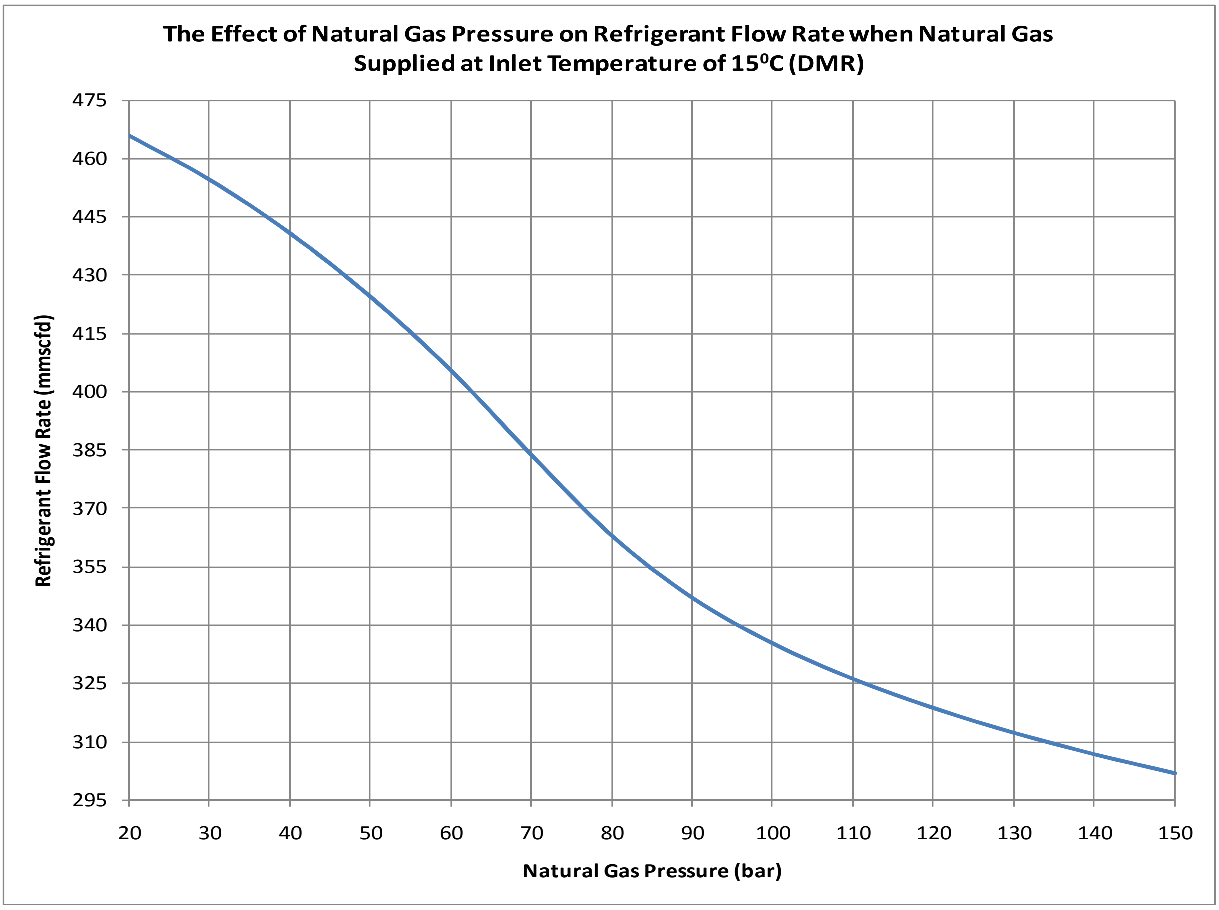

Figure 38: The effect of natural gas pressure on refrigerant flow rate when natural gas supplied at inlet pressure of 15^{0}C (DMR Process)

The effect of natural gas pressure on refrigerant flow rate when natural gas is supplied at different temperatures (5, 10, 15, 20, 25 and 30^{0}C) in DMR process is shown in figure-39 below. The figure indicates that as the supply pressure increases from 60 to 150bar refrigerant flow rate decreases and as supply pressure decreases from 60 to 20bar refrigerant flow rate increases at any natural gas supply temperature. Also the graph shows that when natural gas is supplied at low temperatures (e.g graph with 5^{0}C) it uses low refrigerant flow rate compared to when natural gas is supplied at high temperatures (e.g graph with 30^{0}C) for any natural gas supply pressure. Consider natural gas supplied at (15^{0}C, 60bar) and (30^{0}C, 60bar) uses refrigerant flow rate of 405.67 and 439.75 mmscfd which is a difference of 8.40% with reference to 15^{0}C.

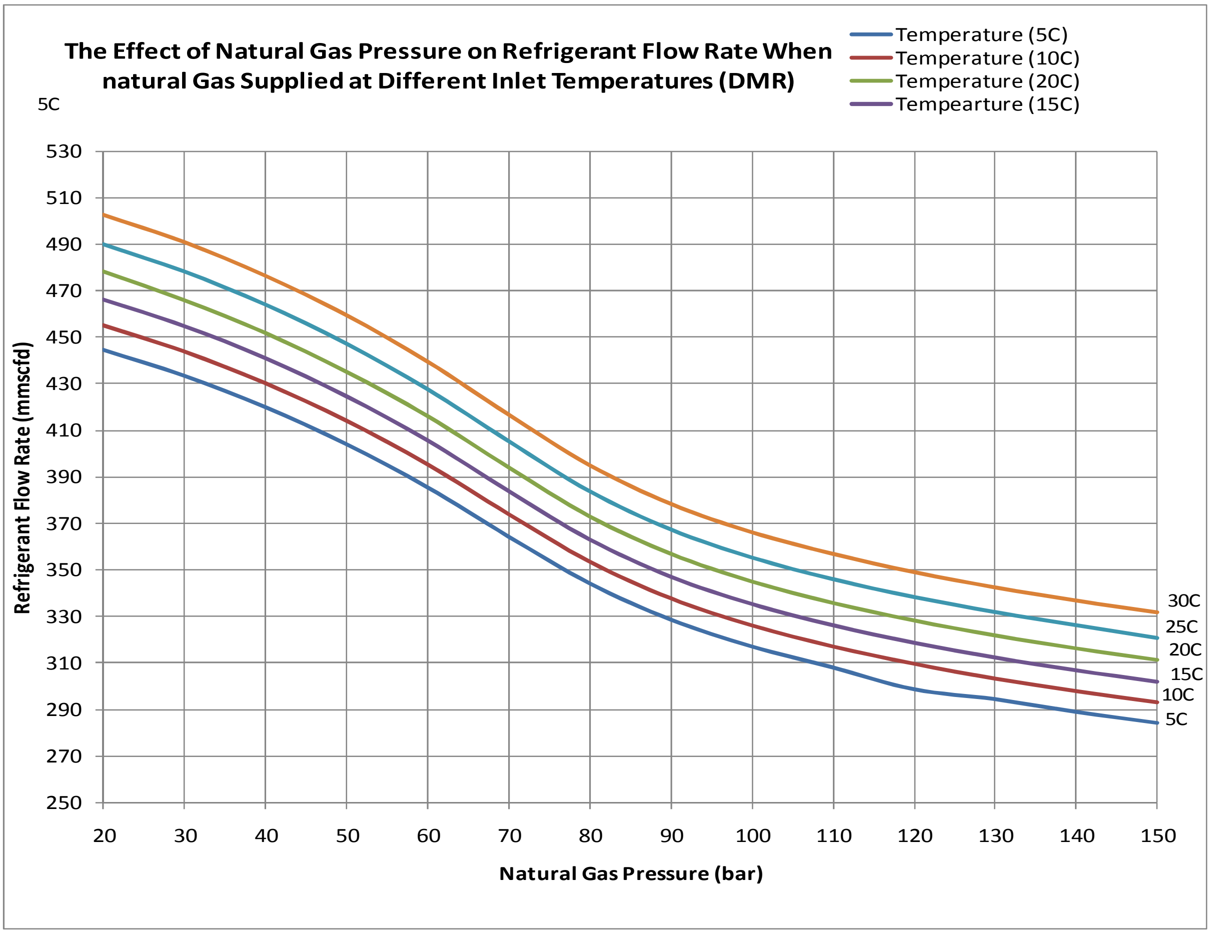

Figure 39: The effect of natural gas pressure on refrigerant flow when natural gas supplied at different temperatures (DMR).

6.2.3 The effect of natural gas composition DMR Process

The effect of natural gas composition on dual mixed refrigerant process has been analyzed using five different compositions termed as composition-A, B, C, D and E . The compositions were analyzed by looking at its effect on process power consumption, refrigerant flow rate and specific power. The effect of natural gas composition on specific power, power consumption and refrigerant flow rate is presented in figure (40, 41, & 42) and table (16, 17, &18). The figures and tables show changing natural gas composition may increase or decrease specific power, power consumption and refrigerant flow rate of the process.

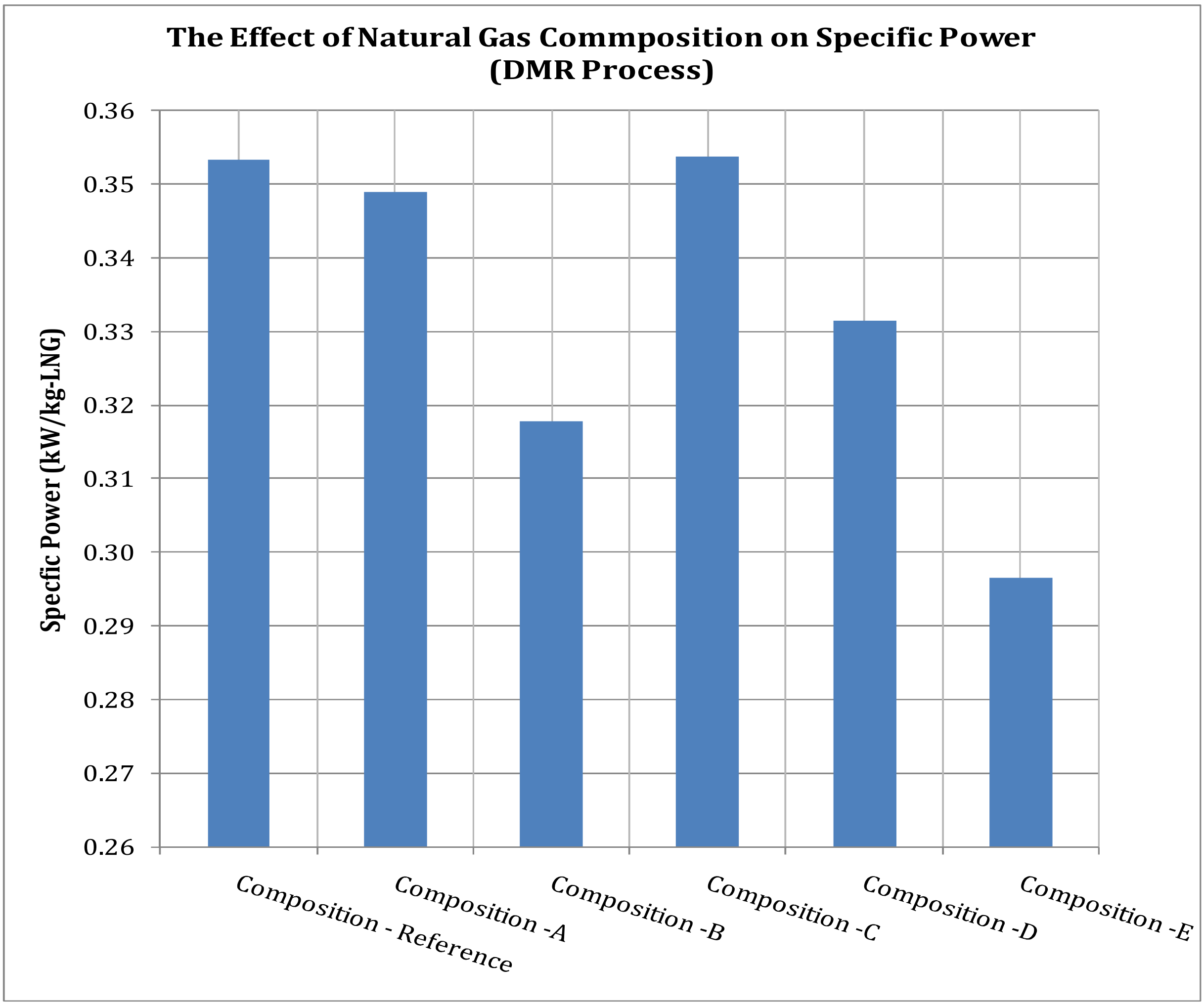

Figure 40: The effect of natural gas composition on specific power (DMR Process)

Table 16 : The effect of natural gas composition on specific power DMR process

Composition	Specific power (kWh/kg-LNG)	Increases or decreases
Composition-Reference	0.353	Used as reference for calculation
Composition - A	0.349	Decreases by 1.1%
Composition -B	0.318	Decreases by 9.9%
Composition –C	0.354	Increase by 0.3%
Composition -D	0.332	Decrease by 6.0%
Composition -E	0.296	Decrease by 16.2%

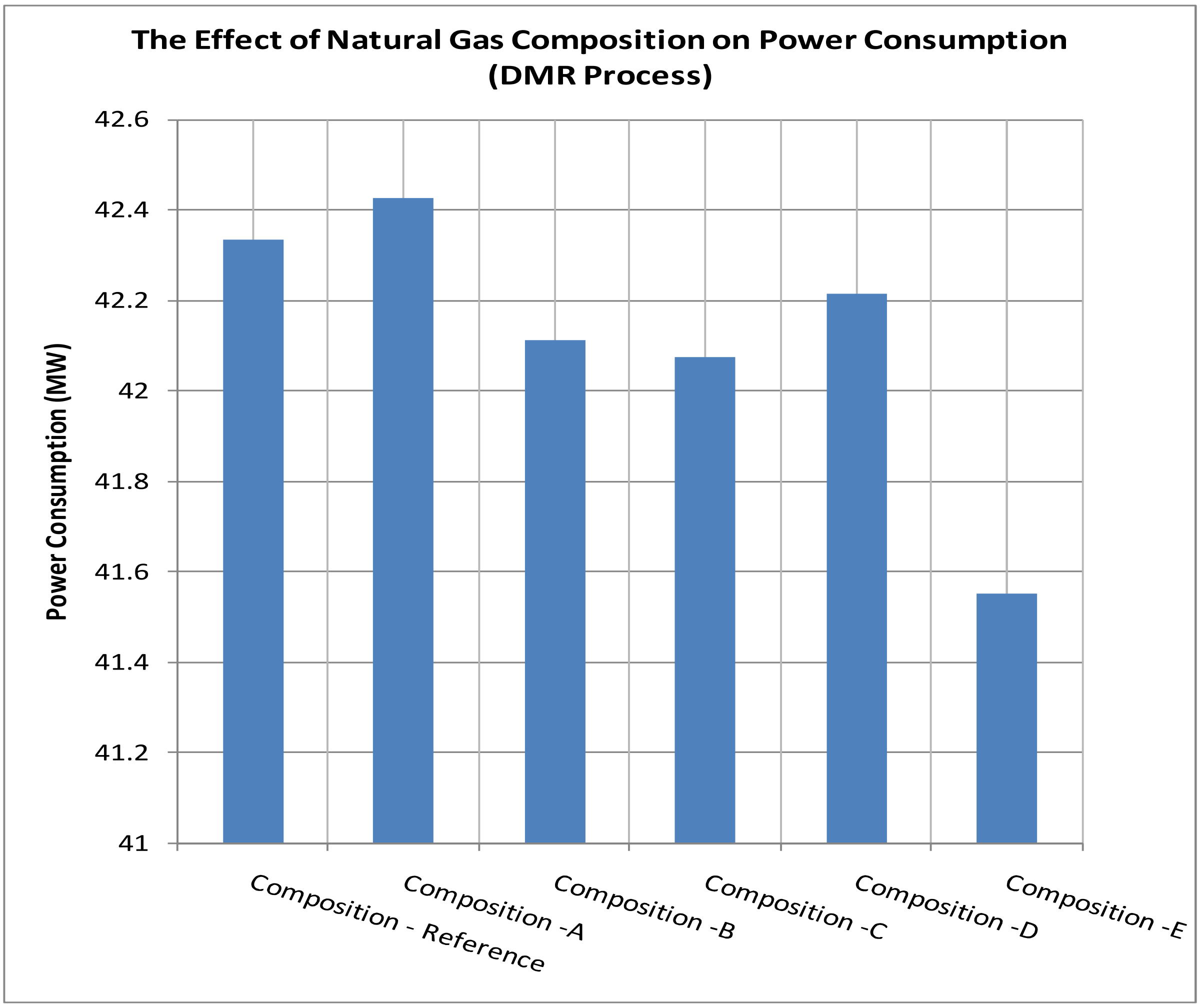

Figure 41: The effect of natural gas composition on power consumption (DMR Process)

Table 17: The effect of natural gas composition on power consumption DMR process

Composition	Power consumption (MW)	Increases or decreases
Composition-Reference	42.33	Used as reference for calculation
Composition - A	42.43	Increase by 0.2%
Composition -B	42.11	Decrease by 0.5%
Composition -C	42.07	Decrease by 0.6%
Composition -D	42.21	Decrease by 0.3%
Composition -E	41.55	Decrease by 1.8%

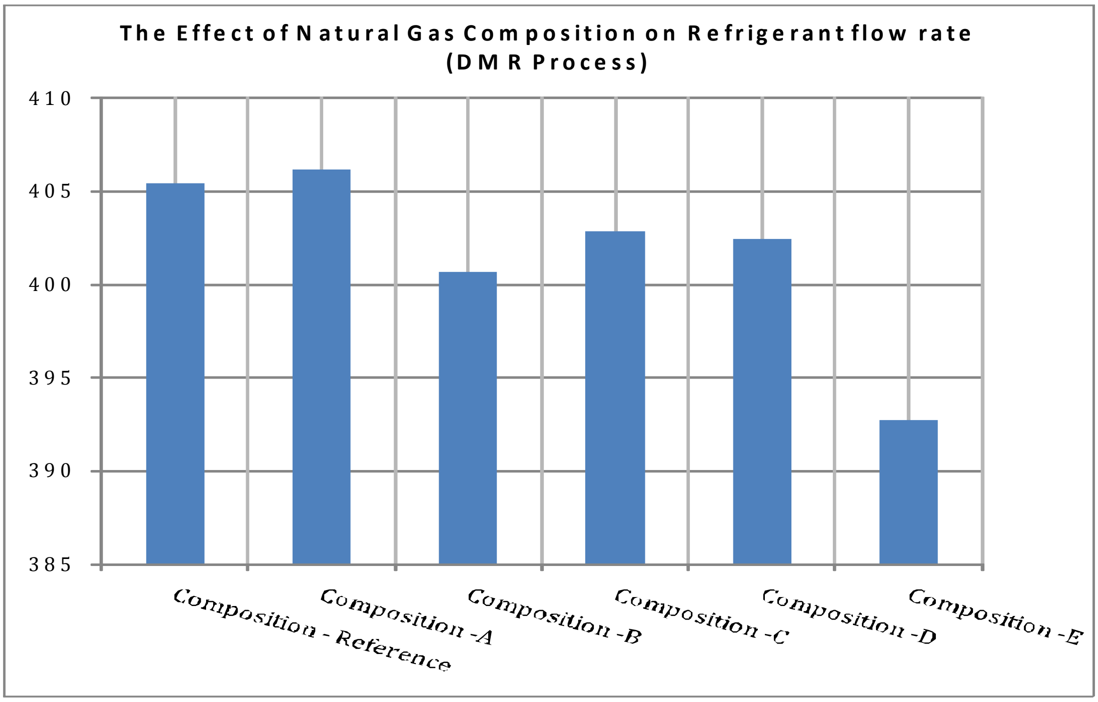

Figure 42 : The effect of natural gas composition on refrigerant flow rate (DMR Process)

Table 18: The effect of natural gas compostion on refrigerant flow rate DMR process

Composition	Refrigerant flow (mmscfd)	Increases or decreases
Composition-Reference	405.42	Used as reference for calculation
Composition - A	406.15	Increase 0.2%
Composition -B	400.68	Decrease by 1.2%
Composition -C	402.87	Decrease by 0.6%
Composition -D	402.47	Decrease by 0.7%
Composition -E	392.72	Decrease by 3.1%

6.2.4 Exergy analysis for DMR process

The exergy losses on different components and useful exergy for DMR process is presented in figure-43 below. The useful exergy accounted for 31% of total exergy and about 69% of exergy is lost on different process components where by cooler (after cooler) accounted for major losses. The amount of useful exergy and losses in (MW) are presented in table-19 below.

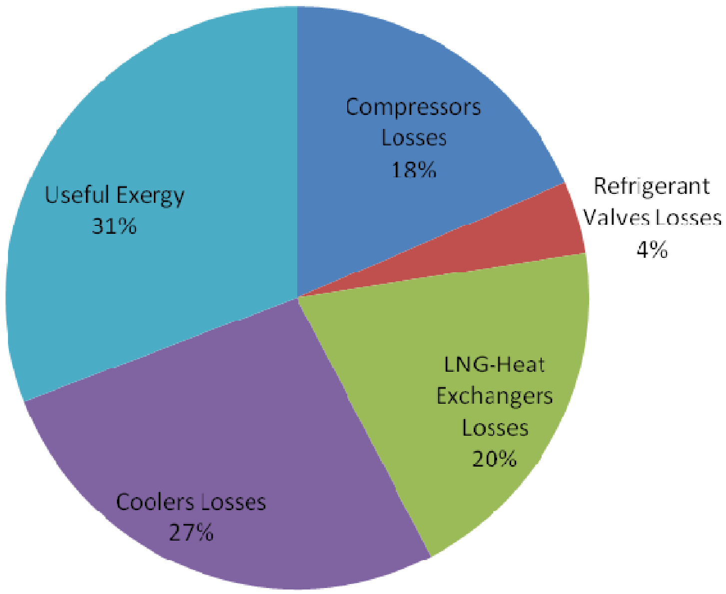

Figure 43: Distribution of exergy losses on different components and useful exergy (DMR)

Table 19: Distribution of exergy losses on different components and useful effect (DMR)

Total exergy supplied	42.33	MW
Useful effect	13.03	MW
Compressor losses	7.85	MW
LNG heat exchangers losses	8.41	MW
Cooler lossses	11.35	MW
Refrigerant valve losses	1.69	MW

6.2.5 Product quality and production capacity of DMR

a) Production Capacity

The production capacity of simulated dual mixed refrigerant process based on detailed description as presented in section 6.1 was analyzed based on maximum duty of one LM6000 gas turbine which has a capacity of 40.7MW [31]. The assumption was made on natural gas supplied at temperature of 15^{0}C and flow rate of 150mmscfd at different pressures.

The production capacity per train of dual mixed refrigerant based on the conditions described above is presented in figure-44 below. Considering natural gas supplied at 60bar the train production capacity is estimated to be 0.91MTPA/train. From simulation when natural gas supplied at 150mmscf , 60bar and 15^{0}C LNG production is about 0.95MTPA .

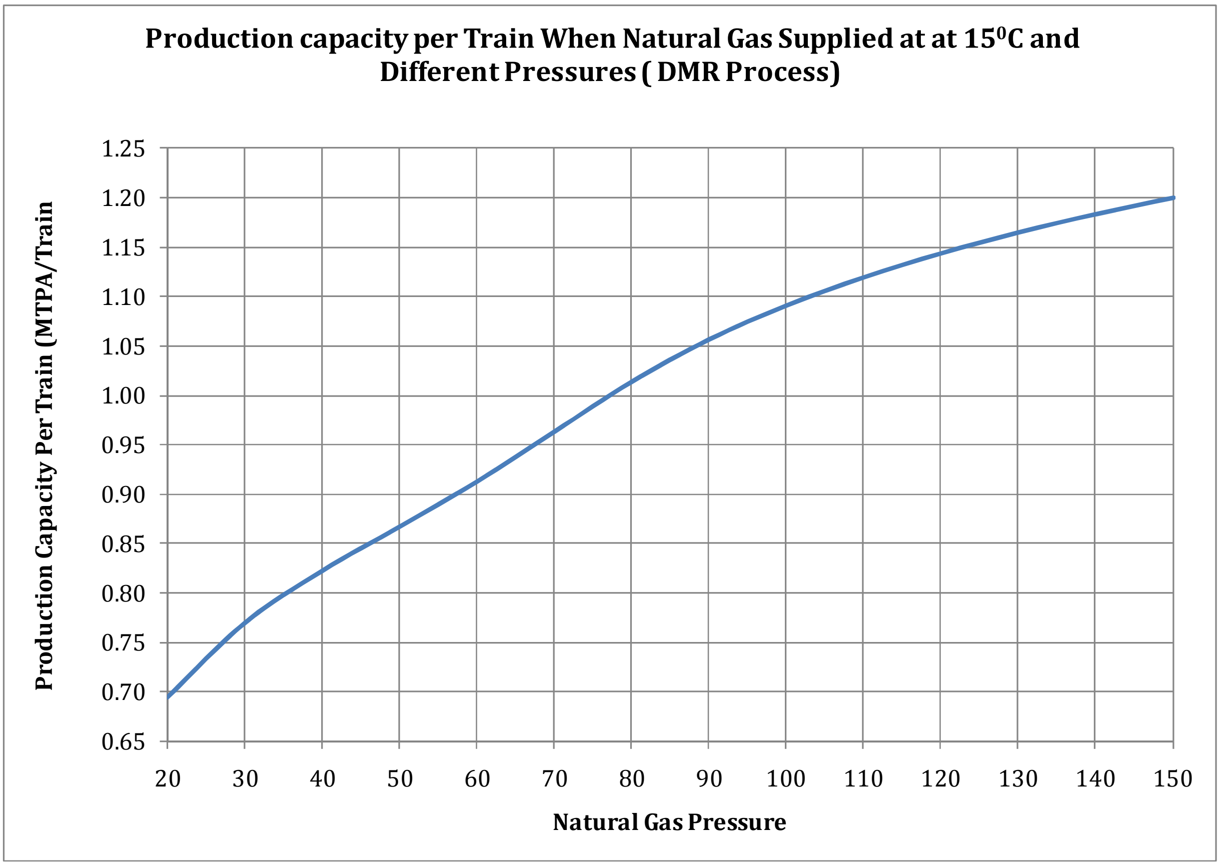

Figure 44: The production capacity of DMR

b) Product Quality

The quality of LNG product produced by dual mixed refrigerant is analyzed on two parameters high heating value (HHV) and wobbe index. The LNG high heating value was calculated to be 39.57MJ/Sm^3 and wobbe index 51.13 this values are within range as described in detail in appendix-A.

6.3 Summary of simulation results for DMR Process

Table 20: Summary of effect of natural gas temperature on DMR Process

	Temperature decreased from 15^{0}C to 5^{0}C	**Temperature increased from 15^{0}C to 25^{0}C**
Specific power (kWh/kg-LNG)	Decreases from 0.353 to 0.330 which amounts to 6.51%	Increases from 0.353 to 0.382 which amounts to 8.23%.
Power consumption (MW)	Decreases from 42.34 to 39.50 amounts to 6.7%	Increases from 42.34 to 45.80 which amount to 8.2%.
Refrigerant Flow rate (mmscfd)	Decreases from 405.67 to 385.78 which amounts to 4.9%	Increases 405.67 to 427.62 which amounts to 5.41%

Table 21: Summary of effect of natural gas pressure on DMR Process

	Pressure decreased from 60 to 30bar	**Pressure increased from 60 to 90bar**
Specific power (kWh/kg-LNG)	Increase from 0.353 to 0.419 which amounted to 18.70%	Decreases from 0.353 to 0.305 which amounts to 13.60%
Power consumption (MW)	Increases from 42.34 to 50.57 which amount to 19.44%.	Decreases from 42.34 to 36.30 which accounts about 14.27%
Refrigerant Flow rate (mmscfd)	Increases from 405.67 to 454.70 amounts to 12.09%.	Decreases from 405.67 to 347.18 which accounts about 14.42%

7. Simulation of Niche LNG process.

7.1 Process detailed description of Niche LNG Process

Niche LNG process consists of two refrigerant cycles. The first cycle uses methane as refrigerant and the second cycle uses nitrogen gas. The refrigerant cycles are designed to operate independently as presented in figure-45 below. The process was simulated based on the patented process [13].

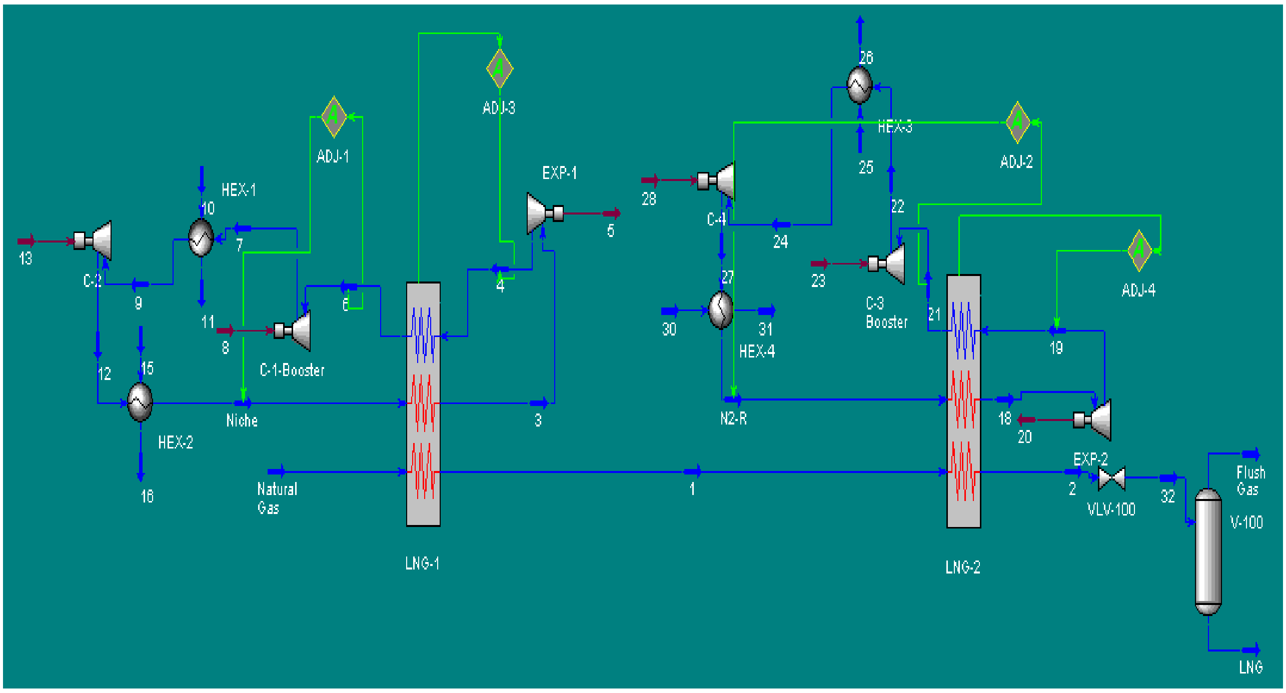

Figure 45: Process flow diagram for Niche LNG Process

The pre-treated natural gas, stream **(natural gas)** at 15^{0}C and 60 bar enters LNG heat exchanger **(LNG-1)** and is cooled to -118^{0}C **(stream-1).** The cooled natural gas (**stream-1)** enters LNG heat exchanger **(LNG-2)** and is cooled further to -159^{0}C **(stream-2).** The pressure of the cooled natural gas **(stream-2)** is decreased from 60 to 1.33 bar via an expansion valve **(VLV-100)** and its temperature decreases to -161.3^{0}C **(stream 32)**. The expanded liquefied natural gas **(stream-32)** enters a flash tank **(V-100)** where LNG and flash gas is separated.

In the first refrigerant cycle with methane as refrigerant, expanded methane **(stream-4)** enters LNG heat exchanger **(LNG-1)** at -124.1^{0}C and 10.08bar and exchanges heat with inlet natural gas stream and methane refrigerant inlet stream **(stream-**

Niche) and exits LNG heat exchanger **(LNG-1)** at 10⁰C **(stream-6)**. The warmed methane refrigerant (**stream-6)** is partially compressed in the first compressor **(C-1)** from 10 to 30bar (**stream-7)** and is cooled to ambient conditions in HEX-1 **(stream-9)**. The partially compressed and cooled methane **(Stream-9)** is then compressed in the second compressor **(C-2)** from 30 to 80bar (**stream-12)** and cooled to ambient temperature in HEX-2 (**stream-Niche**). Stream Niche is the starting point of the methane refrigerant cycle and it enters LNG heat exchanger **(LNG-1**) at 10^{0}C and 80bar and is cooled in LNG heat exchanger **(LNG-1)** to -25^{0}C **(steam-3)** . The cooled methane refrigerant **(stream-3)** is reduced in pressure by expansion in expander **(EXP-1)** from 80 to 10bar and its temperature decreases to -124.1^{0}C (**stream-4)** . Stream-4 is returned to the LNG heat exchanger **(LNG-1)** and the cycle is repeated as described above.

In the second refrigerant cycle with nitrogen refrigerant, expanded nitrogen **(stream-19)** at -168.7^{0}C and 10bar enters LNG heat exchanger **(LNG-2)** and exchanges heat with pre-cooled natural gas (**stream-1)** and inlet stream of nitrogen refrigerant **(stream N2-R)** and exits LNG heat exchanger **(LNG-2)** at 10^{0}C (**stream-21)**. The warmed nitrogen refrigerant **(stream-21)** is first compressed in the first compressor **(C-1)** from 10 to 30bar and cooled to 15^{0}C in HEX-3 **(stream-24)** then compressed in the second compressor from 30 to 80bar and cooled to 15^{0}C in HEX-4 **(stream-N2-R**). Stream N2-R enters LNG heat exchanger **(LNG-2),** is cooled to -90^{0}C **(stream-18)**, expanded in expander (**EXP-2)** from 80 to 10bar and its temperature decreases to -168.7^{0}C **(stream-19)** and the cycle is repeated as described above. Refrigerant flow rate in the first cycle is about 569.4 mmscfd and in the second cycle 225.1 mmscfd. Expander power generated EXP-1 is 13.59MW and EXP-2 3.88MW.

The suction flow rate of methane refrigerant cycle in the first stage compressor was 64,486.54 m3/hr which gives mass flow of 126.4kg/s and nitrogen refrigerant cycle first stage compressor has suction flow rate of 25897.09m3/hr which gives mass flow rate of 87.04kg/s.

7.2 Simulation results for Niche LNG Process

Based on detailed description of the process and conditions described above the processes were simulated and LNG heat exchangers "UA" values found to be 2256.84kW/C for LNG heat exchanger (LNG-1) and 1684.190.07kW/C for LNG heat exchanger (LNG-2) at minimum approach temperature of 5^0C for LNG heat exchanger (LNG-1) and 3.5^0C for LNG heat exchanger (LNG-2). The LMTD found to be 17.18^0C for LNG heat exchanger (LNG-1) and 10.22^0C for LNG-2 respectively . The specific power of the process at above conditions was found to be 0.500kWh/kg-LNG. The LNG heat exchangers "UA" calculated at above was kept constant in simulation by adjusting refrigerant flow rate and refrigerant evaporating pressure during variation of natural gas pressure, temperature and composition.

7.2.1 The effect of natural gas temperature on Niche LNG Process

a) The effect of natural gas temperature on specific power

The effect of natural gas inlet temperature on specific power when natural gas is supplied at 60bar in Niche LNG process is as presented in figure-46 below. From figure-46 below it can be shown that when natural gas inlet temperature decreases from 15^{0}C to 5^{0}C, the specific power of Niche LNG process decreases from 0.500 to 0.451kWh/kg-LNG which amounts to a 9.8% decrease and when the supply temperature increases from 15^{0}C to 25^{0}C its specific power increases from 0.500 to 0.557kW/kg-LNG which amounts to an 11.4% increase. Generally it can be concluded that the specific power of Niche LNG process decreases with decrease in natural gas inlet temperature and increases with increase in natural gas inlet temperature when natural gas is supplied at a given pressure.

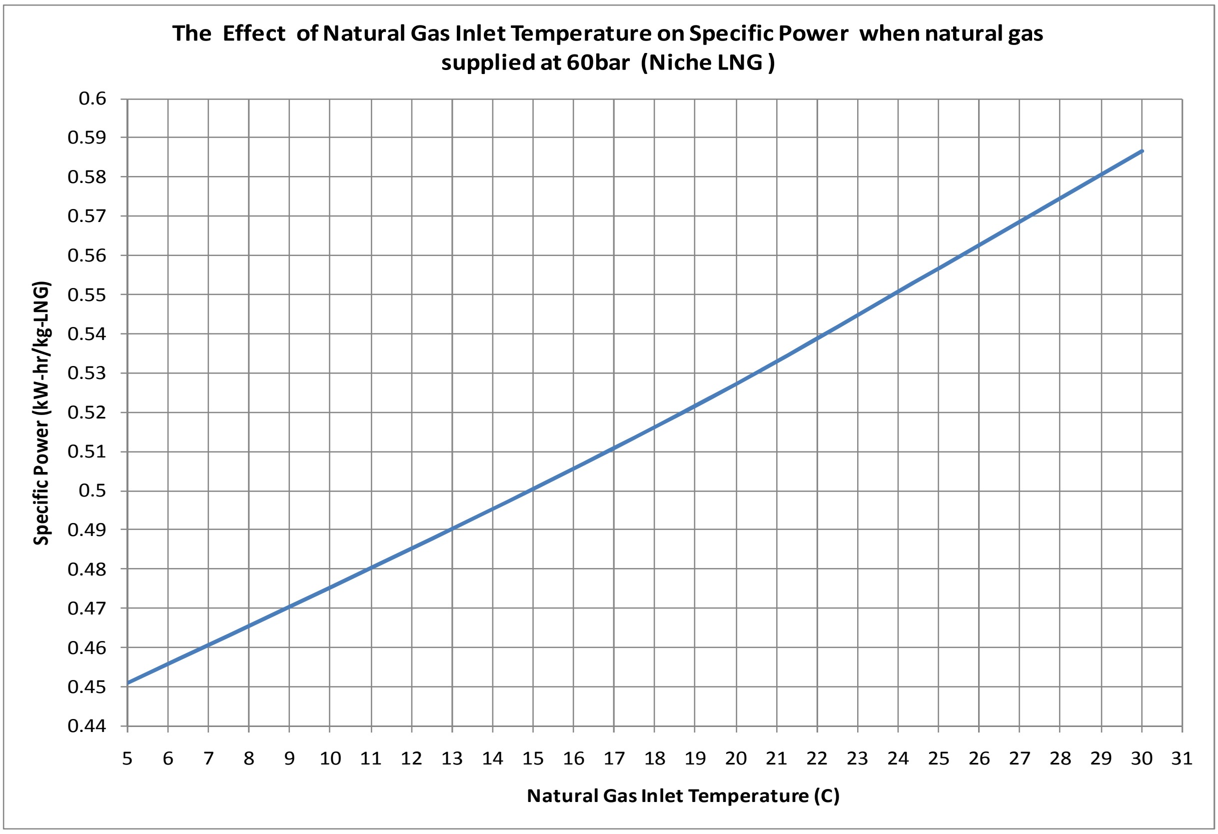

Figure 46: The effect of natural gas inlet temperature on specific power when natural gas supplied at 60bar (Niche LNG Process)

The graphs show the effect of inlet natural gas temperature on specific power when natural gas is supplied at different pressures (20, 40, 60, 80,100,120 and 140bar) in Niche LNG process are presented in figure-47 below. The figure shows that as natural gas temperature decreases from 15^{0}C to 5^{0}C Niche LNG process specific power decreases and when supply temperature increases from 15^{0}C to 30^{0}C, its specific power increases at any natural gas supply pressure. Also the figure shows natural gas supplied at higher pressure has lower specific power while natural gas supplied at low pressure has higher specific power. For example the graph with natural gas supplied at 60bar at 15^{0}C has a specific power of 0.500kWh/kg-LNG while the graph with 120bar at 15^{0}C has 0.444kWh/kg-LNG as its specific power, which amounts to a difference of about 11.2% with reference to 60bar.

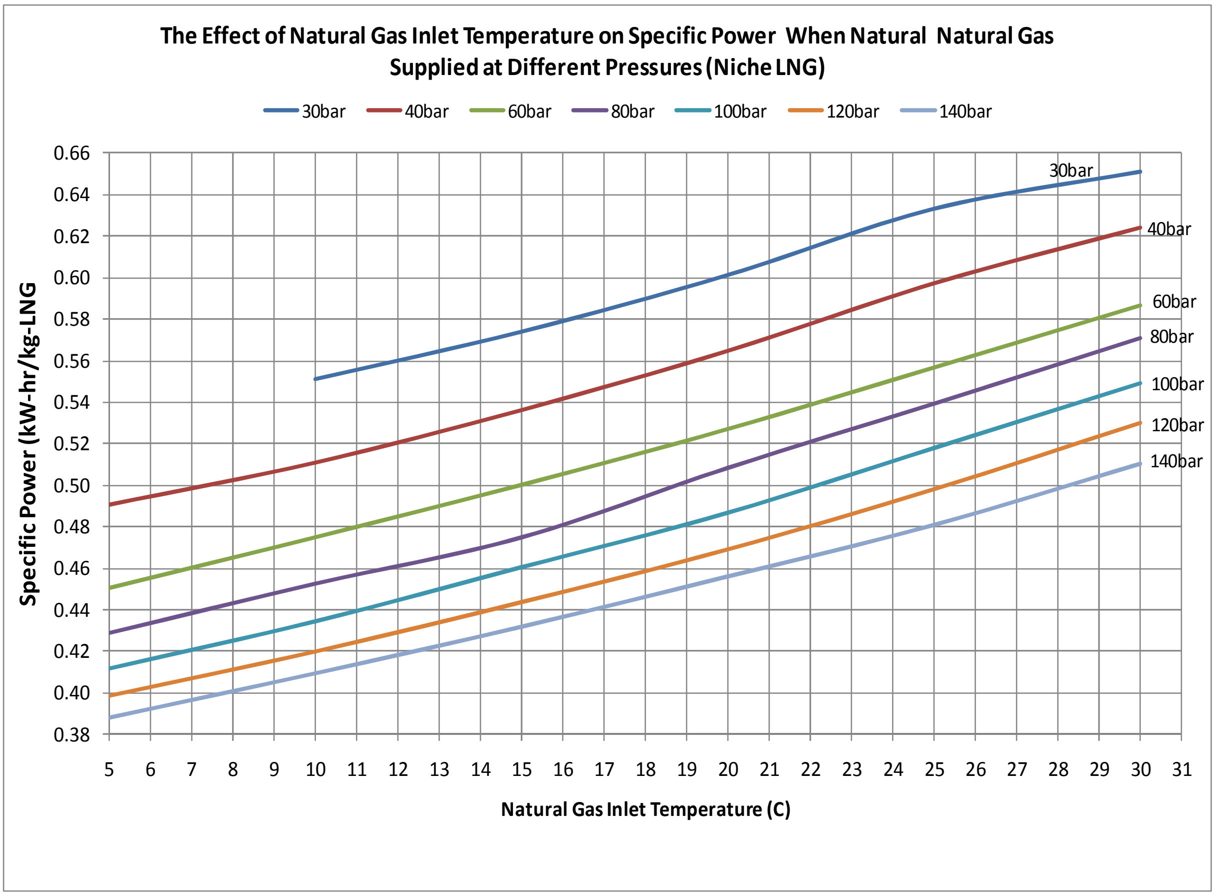

Figure 47: The effect of natural gas temperature on specific power when natural gas supplied at different pressures (Niche LNG Process)

b) **The effect of natural gas inlet temperature on power consumption**

The effect of natural gas inlet temperature on power consumption when natural gas is supplied at 60bar in Niche LNG process is presented in figure-48 below. The figure shows that when natural gas supply temperature decreases from 15^0C to 5^0C, power consumption of Niche LNG process decreases from 62.59 to 56.40MW which amounts to a 9.9% decrease and when natural gas supply temperature increases from 15^0C to 25^0C, the power consumption increases from 62.59 to 69.64MW which amounts to an 11.3% increase. Generally the figure shows that power consumption of Niche LNG process decreases with decrease in natural gas supply temperature and increases with increase in natural gas supply temperature.

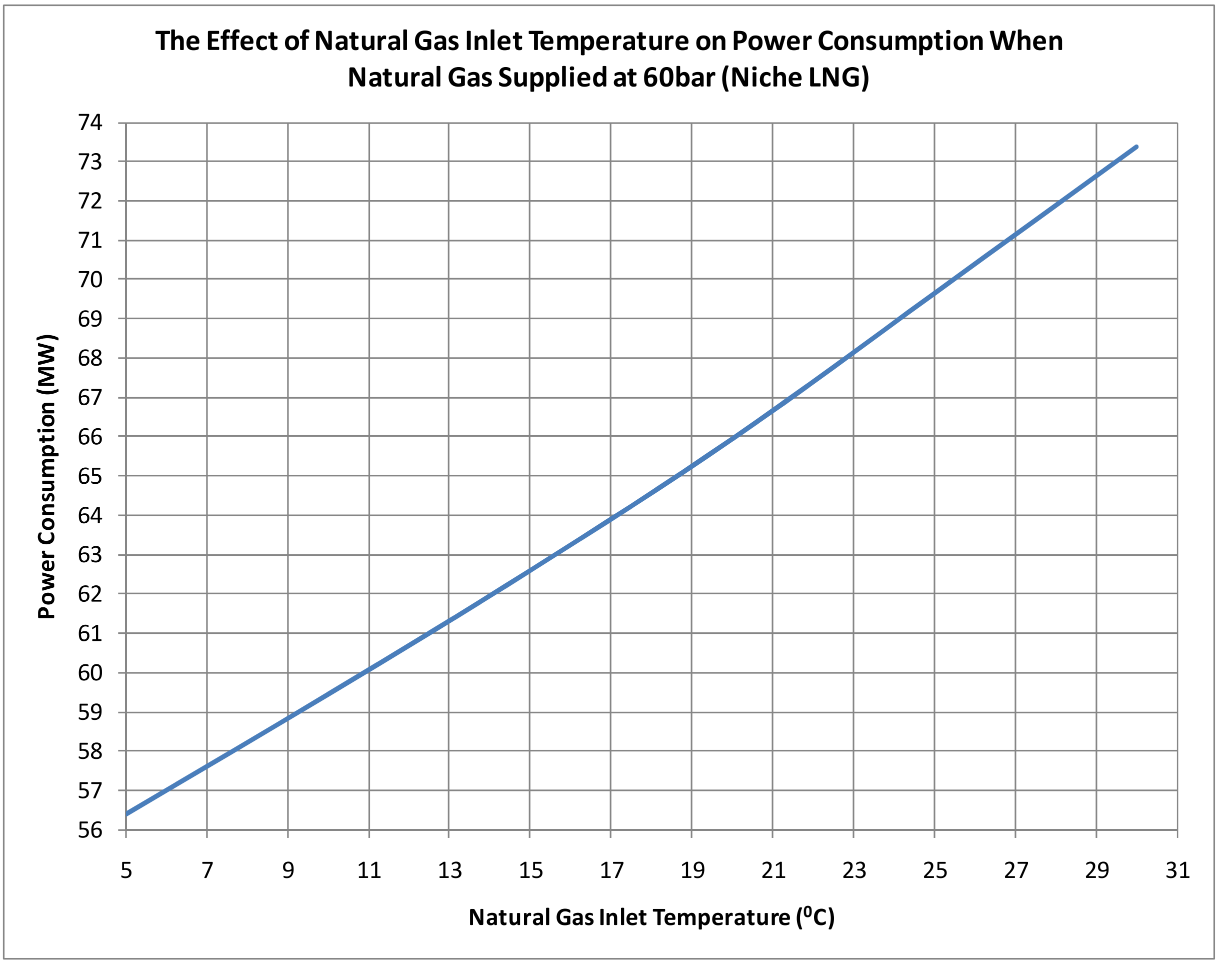

Figure 48: The effect of natural gas inlet temperature of power consumption when natural gas supplied at 60bar (Niche LNG)

The graphs showing the effect of natural gas temperature on Niche LNG process power consumption when natural gas is supplied at different pressures (40, 60, 80, 100, 120 and 140bar) are presented in figure-49 below. The graphs shows that as natural gas supply temperature decreases from 15^{0}C to 5^{0}C, Niche LNG process power consumption decreases and when the supply temperature increases from 15^{0}C to 30^{0}C, its power consumption increases for any supplied natural gas pressure. But also the graphs show that natural gas supplied at higher pressure (e.g graph with 140bar) has lower power consumption compared to natural gas supplied at low pressure (e.g graph with 40bar).

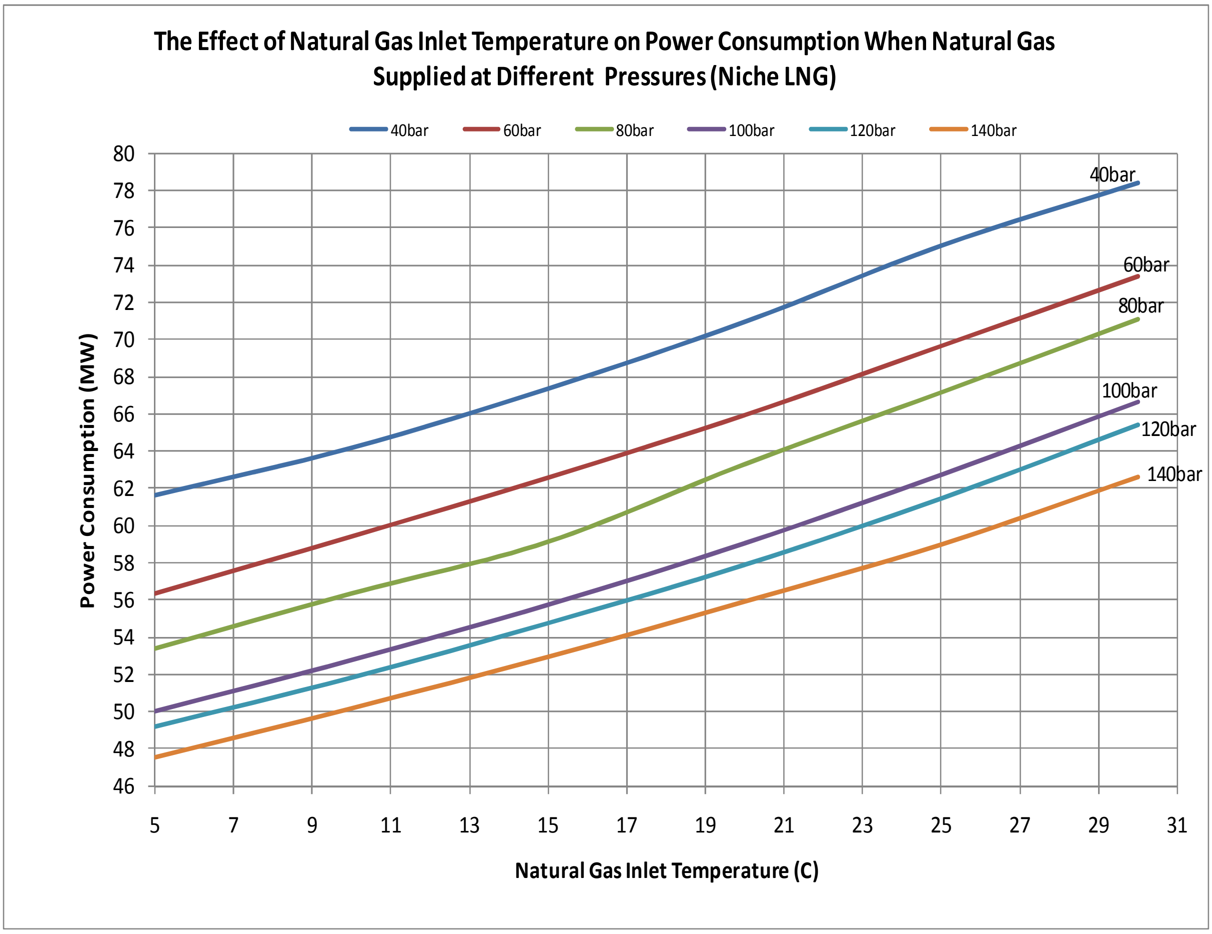

Figure 49: The effect of natural gas inlet temperature on power consumption when natural gas supplied at different temperatures (Niche LNG)

7.2.2 The effect of natural gas Pressure on Niche LNG Process

a) The effect of natural gas pressure on specific power

The effect of natural gas pressure on specific power when natural gas is supplied at 15^{0}C in Niche LNG process is presented in figure-50 below. From the figure it can be depicted that as the natural gas pressure decreases from 60bar to 30bar, the Niche LNG process specific power increases from 0.500 to 0.574kWh/kg-LNG which amounts to a 14.8% increase and when natural gas pressure increases from 60 to 90bar, the specific power decreases from 0.500 to 0.465kW/kg-LNG which amounts to a 7.0% decrease. General it can be described that when natural gas supply pressure decreases, specific power of Niche LNG process increases and when natural gas supply pressure increases, specific power decreases.

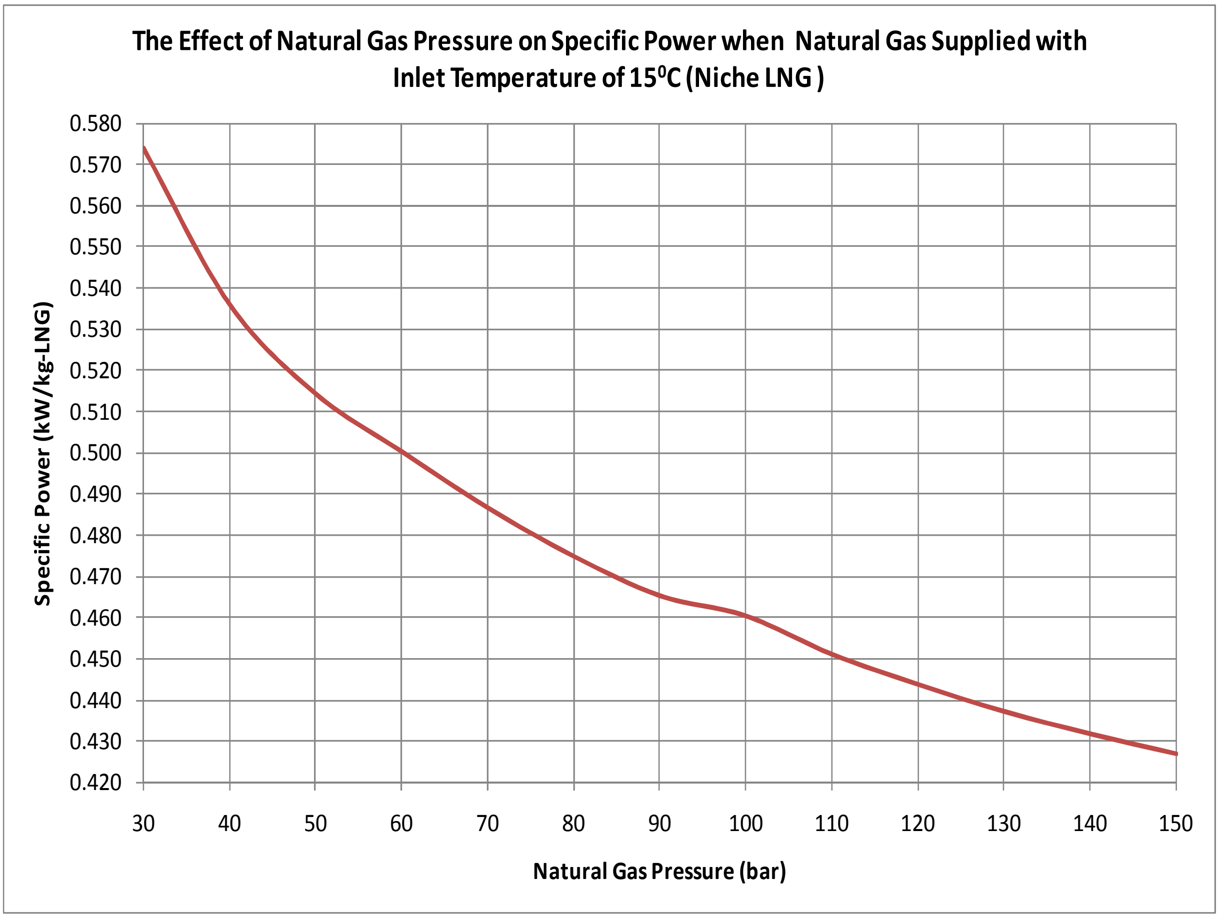

Figure 50: The effect of natural gas pressure on specific power when natural gas supplied with inlet temperature of 15^{0}C (Niche LNG Process)

The graphs which shows the effect of natural gas supply pressure on specific power when natural gas is supplied at different inlet temperatures (5, 10, 15,2 0, 25 and 30⁰C) in Niche LNG process are presented as figure-51 below. The figure shows that as natural gas pressure decreases from 60 to 30bar, the Niche LNG process specific power increases and when natural gas supply pressure increases from 60 to 90bar, its specific power decreases at any natural gas supply temperature. The graphs also show that when natural gas supply temperature is higher, the Niche LNG process specific power is higher (example graph with inlet temperature 30⁰C) and specific power is lower when natural gas supply temperature is low (e.g graph with inlet temperature of 5⁰C).

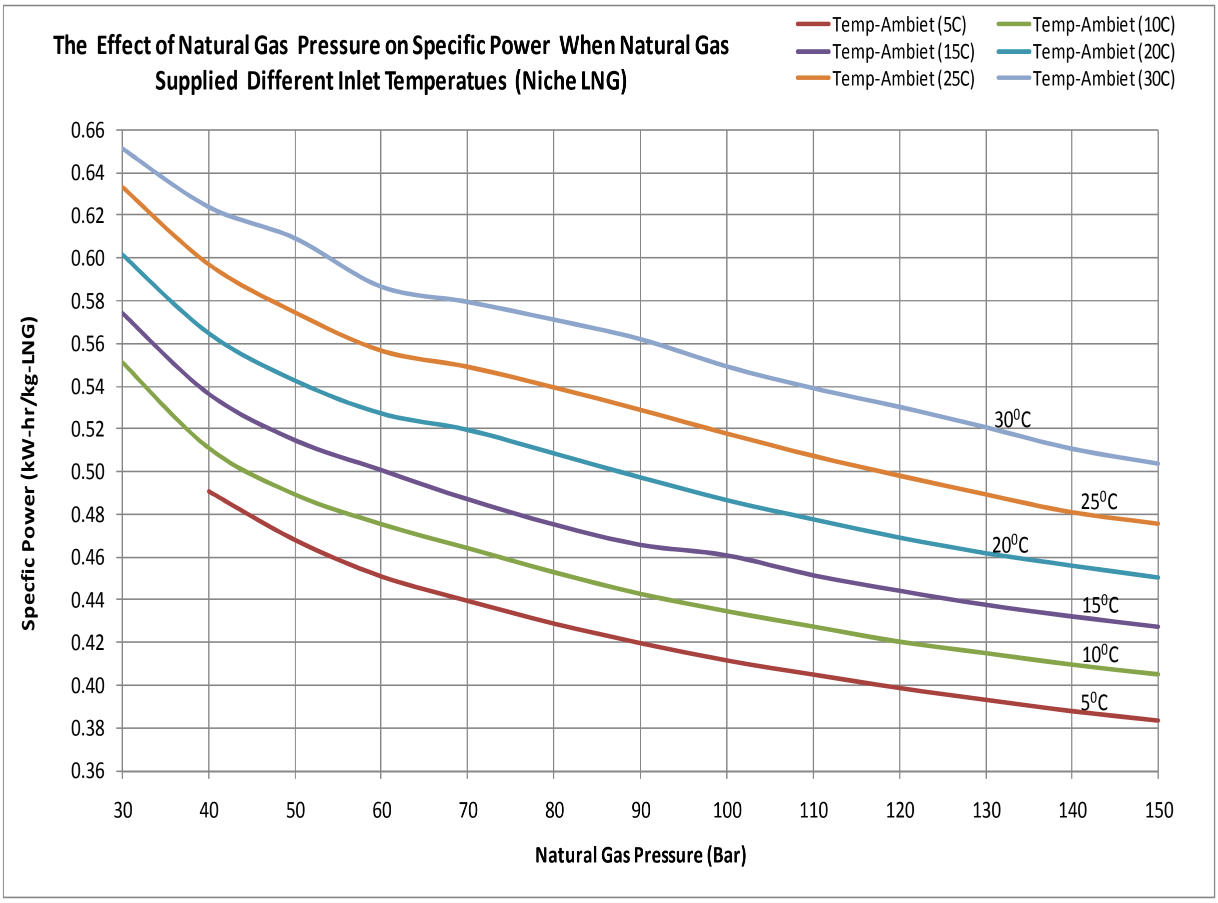

Figure 51: The effect of natural gas pressure on specific power when natural gas supplied at different temperature (Niche LNG Process)

b) The effect of natural gas pressure on power consumption

The effect of natural gas pressure on power consumption when natural gas is supplied with inlet temperature of 15⁰C in Niche LNG process is presented as figure-52. The graphs shows that when natural gas supply pressure decreases from 60 to 30bar, Niche LNG process power consumption increases from 62.59 to 72.29 MW which amounts to a 15.5% increase and when natural gas supply pressure increases from 60 to 90bar, Niche LNG process power consumption decreases from 62.59 to 57.8MW which amounts to a 7.7% decrease. Generally it can be shown that Niche LNG process power consumption increases with decrease in natural gas supply pressure and decreases with increase in natural gas supply pressure.

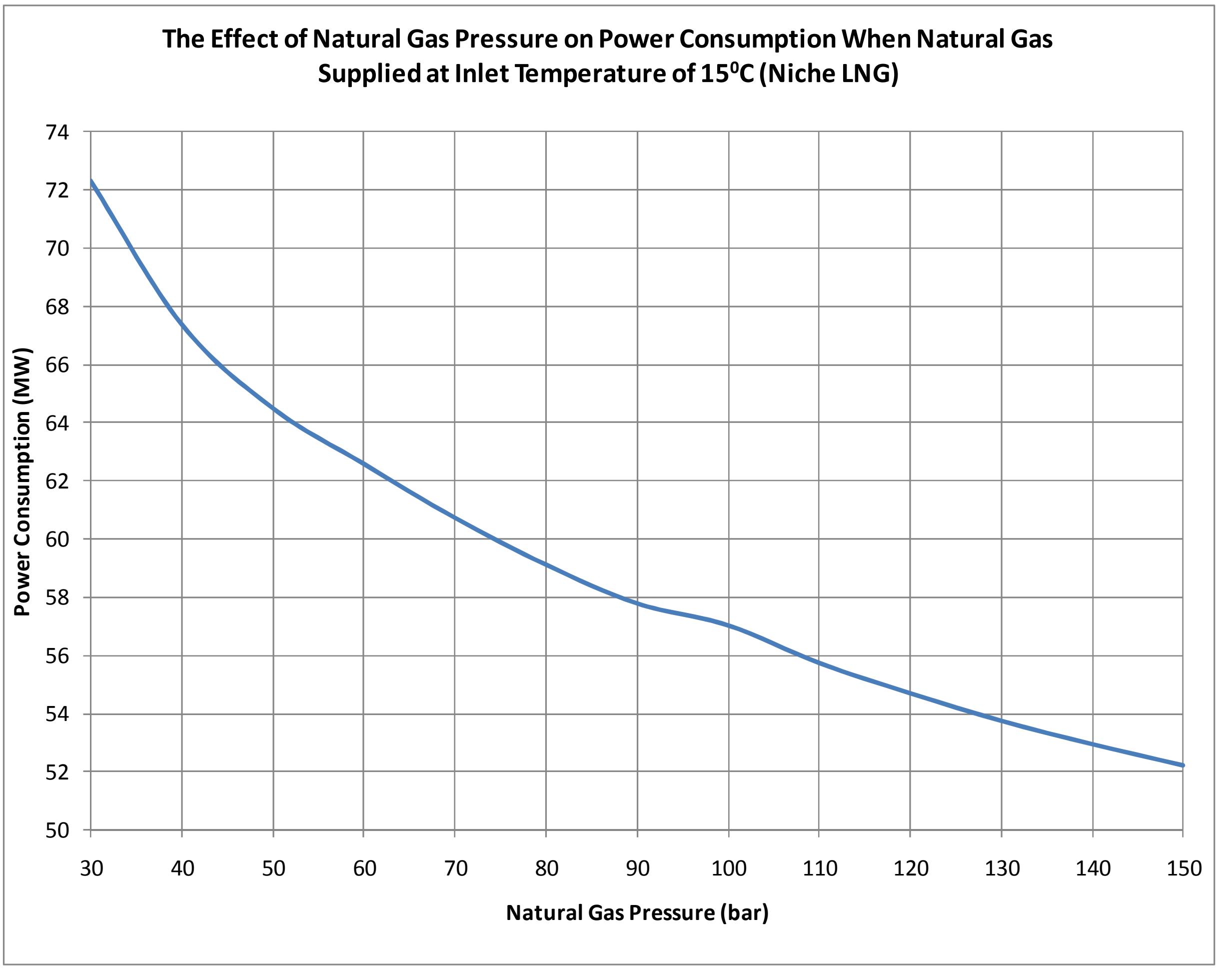

Figure 52 : The effect of natural gas pressure on power consumption when natural gas supplied at inlet temperature of 15^0C (Niche LNG Process).

The graphs showing the effect of natural gas pressure on power consumption when natural gas is supplied at different temperatures (5, 10, 15, 20, 25 and 30^{0}C) in Niche LNG process is presented in figure-53 below. The figure shows that as natural gas supply pressure decreases from 60 to 30bar, the Niche LNG process power consumption increases and when natural gas supply pressure increases from 60 to 150bar its power consumption decreases at any natural gas supply temperature. Also the graph indicates that Niche LNG process power consumption is low when natural gas supply temperature is low (e.g graph with natural gas inlet temperature of 5^{0}C) and is higher when natural gas supply temperature is high (e.g graph with natural gas inlet temperature of 30^{0}C) at any natural gas supply pressure.

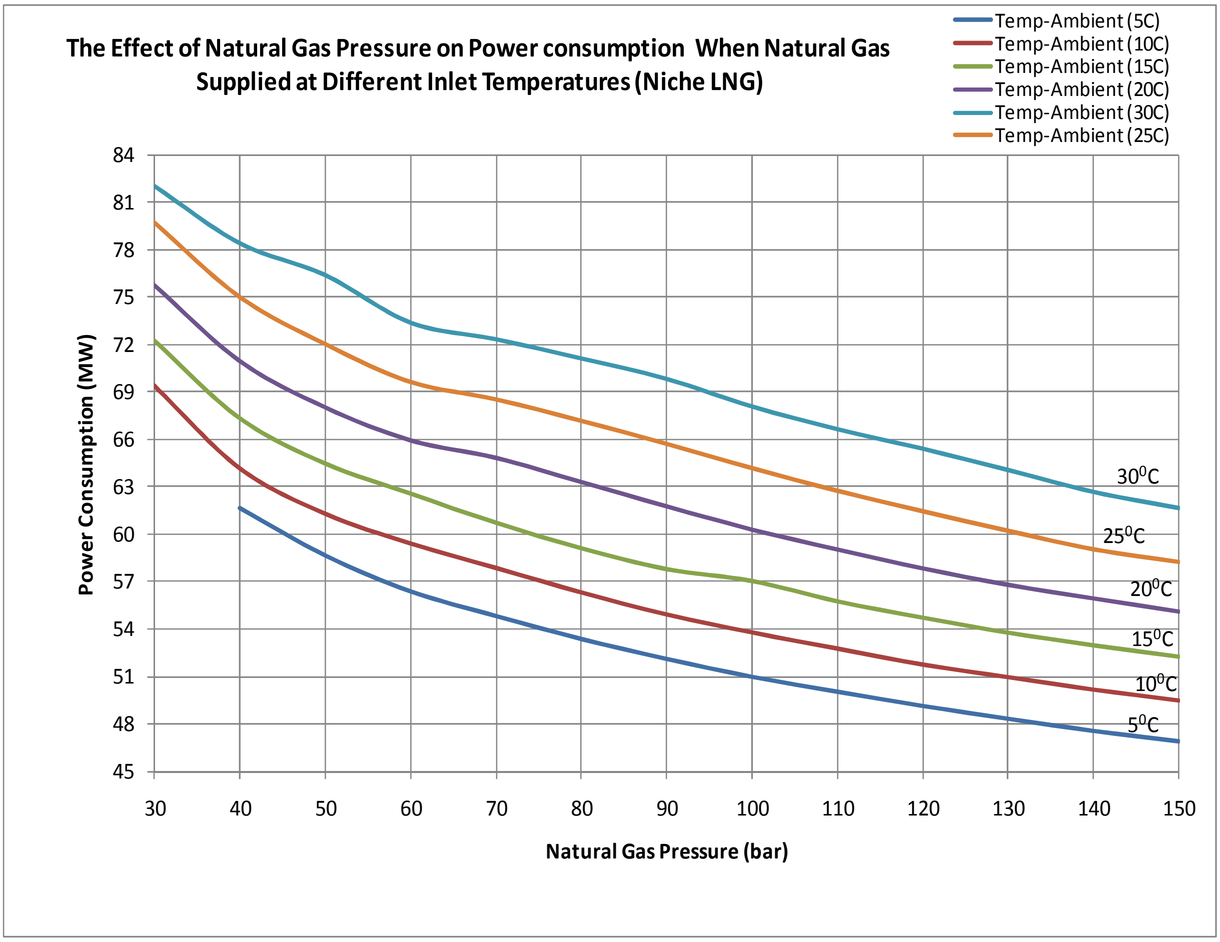

Figure 53: The effect of natural gas pressure on power consumption when natural gas supplied at different inlet temperatures (Niche LNG Process).

7.2.3 The effect of natural gas composition on Niche LNG Process

The effect of natural gas composition on specific power, power consumption and refrigerant flow rate in Niche LNG process is presented in figure 54, 55&56 below. The assumption was made on natural gas being supplied at 60bar and 15^{0}C for all compositions. The natural gas composition used for simulation of benchmarked process system is termed as composition-reference and its results were used as reference point to the other compositions. The other compositions were termed as composition -A, B, C, D and E and its effects are presented in table 22, 23 &24. The compositions shows that they may have significant effect on specific power, power consumption or refrigerant flow as shown on figures and tables below.

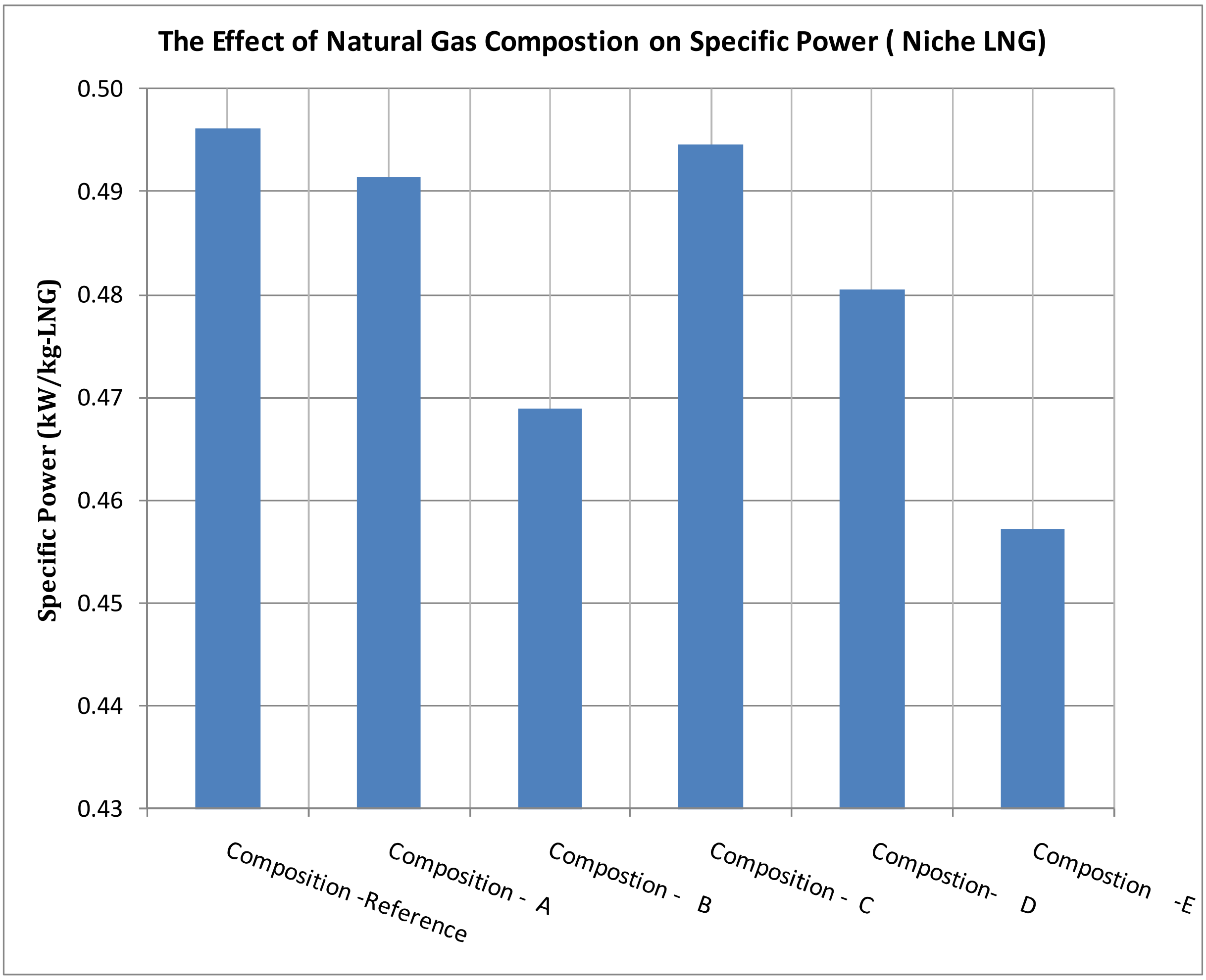

Figure 54: The effect of natural gas composition on specific power (Niche LNG)

Table 22: The effect of natural gas composition on specific power (Niche LNG)

Composition	Specific power (kWh/kg-LNG)	Increases or decreases
Composition-Reference	0.496	
Composition - A	0.491	Decrease by 1%
Composition -B	0.469	Decreased by 5.4%
Composition -C	0.495	Decrease by 0.2%
Composition -D	0.481	Decrease by 3.0%
Composition -E	0.457	Decrease by 7.9%

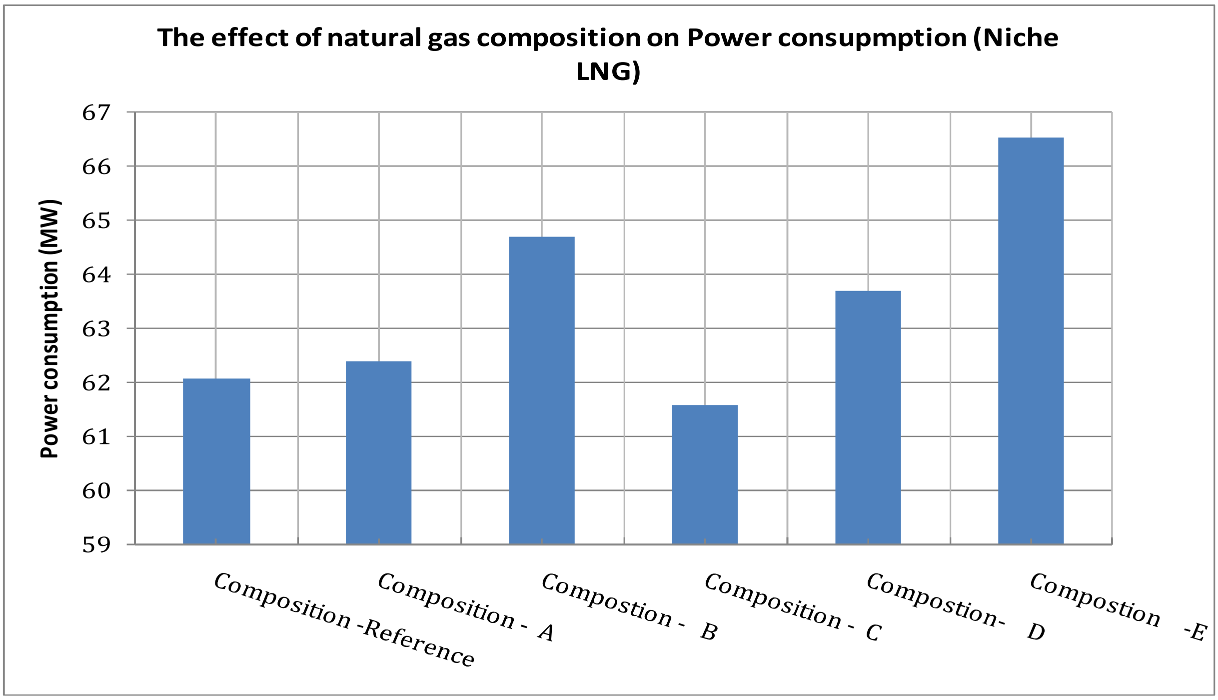

Figure 55: The effect of natural gas composition on power consumption (Niche LNG Process)

Table 23: The effect of natural gas composition on power consumption (Niche LNG)

Composition	Power consumption (MW)	Increases or decreases
Composition-Reference	62.07	
Composition - A	62.39	Increased by 0.5%
Composition -B	64,68	Increased by 4.2%
Composition -C	61.58	Decreased by 0.7%
Composition -D	63.70	Increased by 2.6%
Composition -E	66.53	Increased by 7.2%

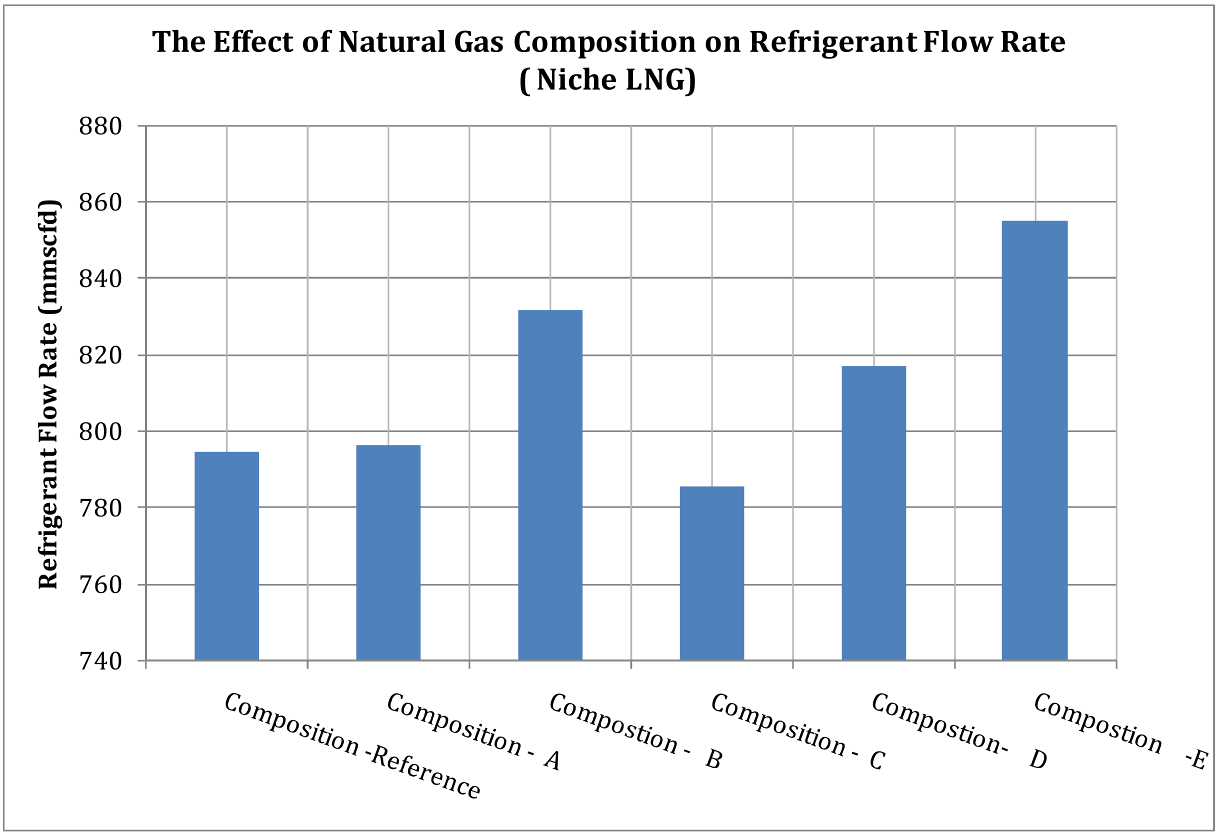

Figure 56: the effect of natural gas composition on refrigerant flow rate (Niche LNG)

Table 24: The effect of natural gas compositions on refrigerant flow rate (Niche LNG Process)

Composition	Refrigerant flow (mmscfd)	Increases or decreases
Composition-Reference	794.50	
Composition - A	796.41	Increase by 0.2%
Composition -B	831.65	Increases by 4.7%
Composition -C	785.67	Decreases by 1.1%
Composition -D	816.94	Increases by 2.8%
Composition -E	855.17	Increases by 7.6%

7.2.4 Exergy analysis for Niche LNG Process

The distribution of exergy losses and useful effect of Niche LNG process is presented in figure-57 below. It indicates that about 80% of exergy is lost in different process component. The amount of exergy losses of each component and useful effect is presented in table-25 below.

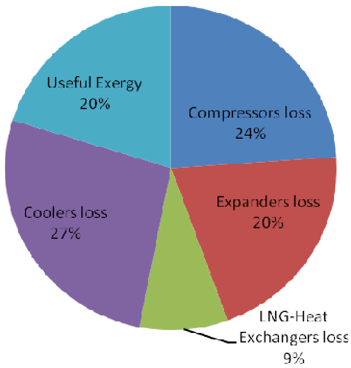

Figure 57: The distribution of exergy losses and useful effect in Niche LNG process

Table 25: The amount of exergy destroyed on each component and useful effect (Niche LNG)

Item	Exergy destroyed (MW)
Compressor (C-1)	6.74
Compressor (C-2)	3.44
Compressor (C-3)	2.04
Compressor (C-4)	1.96
Expander (EXP-1)	8.83
Expander (EXP-2)	3.32
LNG Heat exchanger (LNG-1)	3.45
LNG Heat exchanger (LNG-2)	1.67
Cooler (HEX-1)	8.49
Cooler (HEX-2)	2.83
Cooler (HEX-3)	2.23
Cooler (HEX-4)	2.27
Total Exergy destroyed	47.26
Total work supplied	59.25
Useful effect	11.99

7.2.5 Product Quality and Production Capacity of Niche LNG Process

a) Production Capacity

The production capacity of simulated Niche LNG process based on the detailed description as presented in section 7.1 was analyzed based on maximum duty of one LM6000 gas turbine which has a capacity of 40.7MW [31]. The assumption was made on natural gas supplied at temperature of 15^{0}C and flow rate of 150mmscfd at different pressures.

The production capacity per train of Niche LNG when Niche LNG was simulated based on the conditions mentioned above is presented in figure-58 below. Considering natural gas supplied at 60bar the train production capacity is estimated to 0.64MTPA/train. The simulation shows that when natural gas is supplied at 150mmscf , 60bar and 15^{0}C LNG production is about 0.99MTPA .

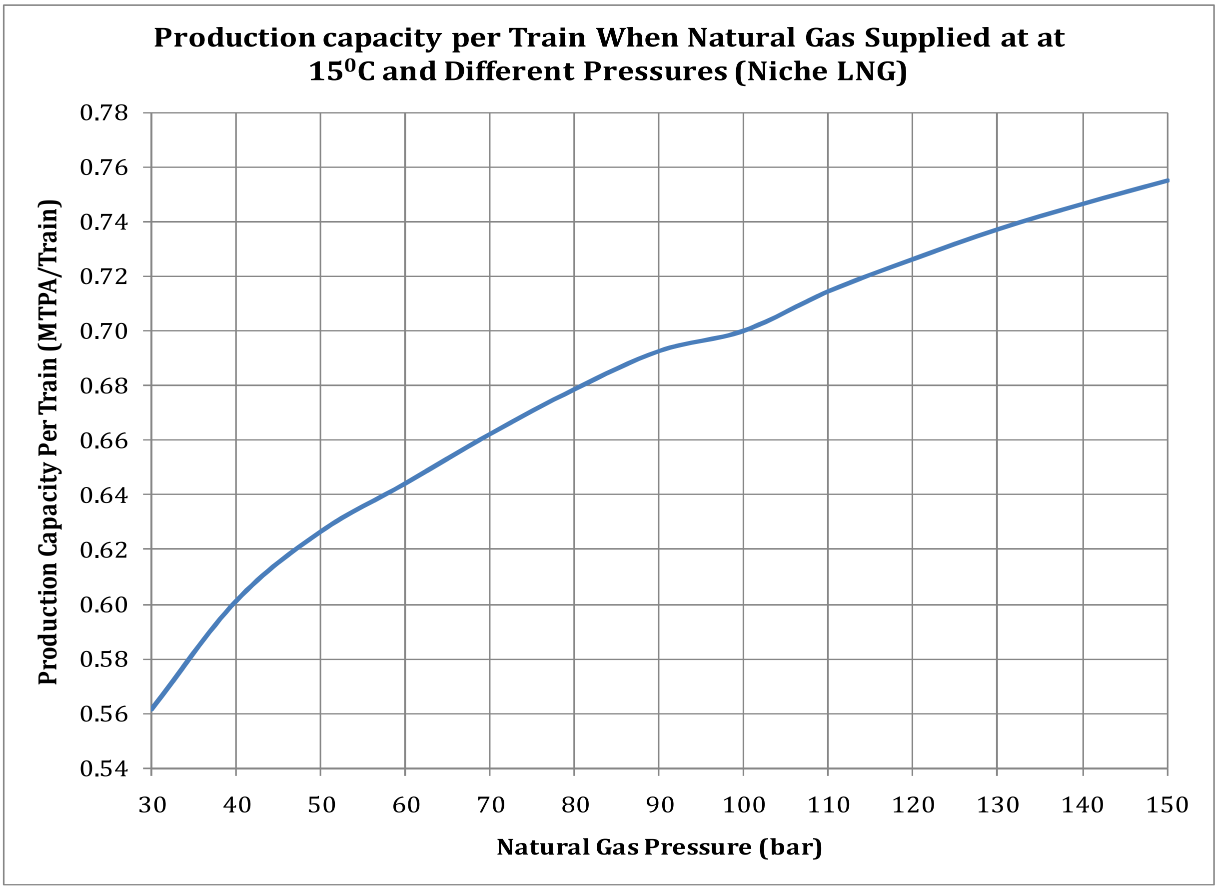

Figure 58: The production capacity of Niche LNG Process

b) Product Quality

The quality of LNG product produced by dual nitrogen expander is analyzed on two parameters high heating value (HHV) and wobbe index. The LNG high heating value was calculated to be 39.29MJ/Sm^3 and wobbe index 50.73 this value are within range as described in detail in appendix-A.

7.3 Summary of simulation results of Niche LNG process

Table 26: Summary of effect of natural gas temperature on Niche LNG process

	Temperature decreased from 15^{0}C to 5^{0}C	**Temperature increased from 15^{0}C to 25^{0}C**
Specific power (kWh/kg-LNG)	Decrease 0.500 to 0.451kWh/kg-LNG which amounts to 9.8%	Increases 0.500 to 0.557 which amounts to an 11.4% increase
Power consumption (MW)	Decreases from 62.59 to 56.40 which amounts to 9.9%	Increases from 62.59 to 69.64 which amounts to 11.3%.

Table 27: Summary of effect of natural gas pressure on Niche LNG process

	Pressure decreased from 60 to 30bar	**Pressure increased from 60 to 90bar**
Specific power (kWh/kg-LNG)	Increases from 0.500 to 0.574 which amounts to 14.8%	Decreases from 0.500 to 0.465kW/kg-LNG which amounted to about 7.0%
Power consumption (MW)	Increased from 62.59 to 72.29 which amounts to 15.5%	Decreases from 62.59 to 57.8 which amounts to 7.7%.

8. Simulation of Dual Nitrogen expander Process

8.1 Detailed Description of dual nitrogen expander process

Dual nitrogen expander process consists of two refrigerant cycles both use nitrogen gas as refrigerant. The cycles are dependent on each other but operate at different temperatures. The first cycle with large refrigerant flow rate is used for pre-cooling and liquefaction of natural gas while the second cycle with low refrigerant flow rate is used for sub cooling. The process flow diagram of dual nitrogen expander as simulated in hysys is presented in figure-59 below. The designed process was based on patent with reference [18].

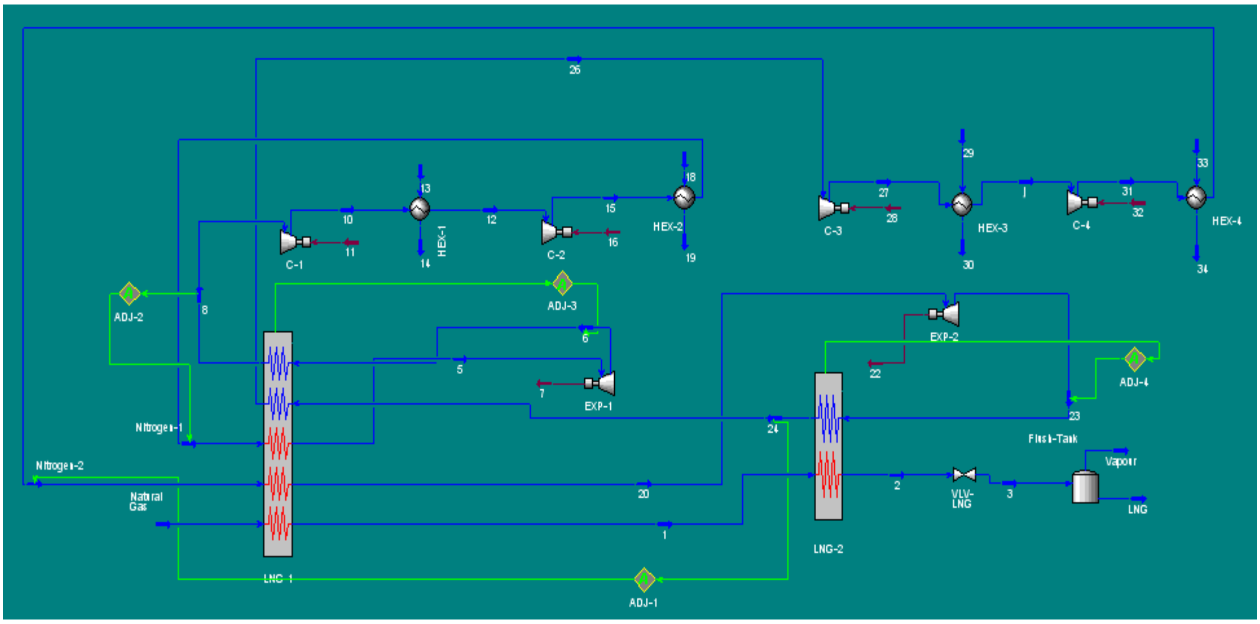

Figure 59 : Process flow diagram of dual nitrogen expander process

The pretreated inlet gas stream **(stream natural gas)** enters LNG heat exchanger **(LNG-1)** at 60bar and ambient temperature of 15^{0}C and is cooled to -90^{0}C **(stream-1)** by counter current flow of refrigerant **(stream-6 and stream-24).** The cooled natural gas (**stream-1)** enters LNG heat exchanger (**LNG-2)** and is cooled further to -159^{0}C (**stream-2).** Stream-2 is decreased in pressure via expansion valve **(VLV-LNG)** from 60bar to 1.05bar and thus the temperature is decreased to -161.3 **(stream-3)**. Stream-3 enters to flash tank where LNG and flash gas separated.

In the first refrigeration cycle expanded nitrogen stream (**stream-6)** enters LNG heat exchanger **(LNG -1)** at -105^{0}C and 15 bar and exchanges heat with pretreated natural gas stream, second refrigerant cycle (stream Nitrogen-2) and first refrigerant cycle stream (Nitrogen-1). Expanded nitrogen (**Stream-6)** exits LNG-1 at 10^{0}C (**stream-8**) and is compressed in first compressor (C-1) from 15 to 30bar, cooled to ambient conditions in **(HEX-1)** then compressed in second compressor from 30 to 80bar, cooled to ambient condition in **(HEX-2)** and returned to LNG heat exchanger LNG-1 (**stream Nitrogen-1)** at 10^{0}C and 80bar. Stream nitrogen-1 is cooled in LNG heat exchanger (**LNG-1**) to -15^{0}C (**stream-5**), its pressure decreased from 80 to 15bar by expansion in expander (**EXP-1**), its temperature decreased to -105^{0}C (**stream-6**) and the cycle loops as described above.

In the second refrigeration cycle, a cold expanded nitrogen **(stream-23)** enters LNG heat exchanger **(LNG-2)** at -169.1^{0}C and 10bar , exchanges heat with pre-cooled natural gas **(stream-1)** and cools natural gas to -159^{0}C **(stream-2).** The expanded nitrogen (**stream-23)** exit LNG heat exchanger **(LNG-2)** at -95^{0}C **(stream-24)** and enters LNG heat exchanger **(LNG-1),** exchanges heat and exits LNG heat exchanger **(LNG-1)** at 10^{0}C **(stream-26)**. Stream-26 is first compressed in compressor **(C-3)** from 10 to 30bar , cooled to ambient condition then compressed in second compressor **(C-4)** from 30 to 80bar and cooled to ambient conditions then returned to LNG heat exchanger **(LNG-1)** at 10^{0}C and 80bar as stream **(Nitrogen-2)**. Stream Nitrogen-2 is cooled in LNG heat exchanger **(LNG-1)** and exits LNG heat exchanger **(LNG-1)** **(stream-20**) at -90^{0}C and returned to the expander **(EXP-2)** and the cycle loops as described above.

The composition of refrigerant used contains nitrogen with mole fraction of 0.98 and oxygen 0.02. The first refrigerant cycle has flow rate of 775.73 mmscfd and second cycle 269.36 mmscfd. The calculated "UA" values of LNG heat exchangers was kept constant during variation of natural gas pressure, temperature and composition by adjusting refrigerant flow rate and varying refrigerant evaporating pressure of both cycles.

8.2 Simulation results for Dual Nitrogen expander

The hysys simulation of dual nitrogen expander process was based on process detailed description as presented in section 8.1 above. During the simulation some of the key parameters calculated were LNG heat exchangers "UA" values which was found to be 4,195.67kW/^{0}C for LNG heat exchanger LNG-1 and 852.094kW/C for LNG heat exchanger LNG-2 at minimum approach temperature of 5^{0}C for both heat exchanger and LMTD of 10.9^{0}C for both heat exchangers. The specific power of the process based on above conditions was found to be 0.531kWh/kg-LNG.

8.2.1 The effect of natural gas inlet temperature on dual nitrogen expander Process

a) The effect of natural gas temperature on plant specific power

The effect of natural gas temperature on plant specific power when natural gas is supplied at 60bar for dual nitrogen expander process is presented in figure-60 below. By taking into account 15^0C as benchmarked temperature the figure shows that as natural gas temperature decreases from 15^0C to 5^0C dual nitrogen expander process specific power decreases from 0.531 to 0.482kWh/kg-LNG which amounts to 9.2% and when temperature increases from 15^0C to 25^0C specific power increases from 0.531 to 0.590kWh/kg-LNG which amounted to 11.11%. Generally it can be described that for dual nitrogen expander process at given natural gas temperature and pressure when temperature of natural gas decreases specific power decreases and when temperature of natural gas increases also specific power increases.

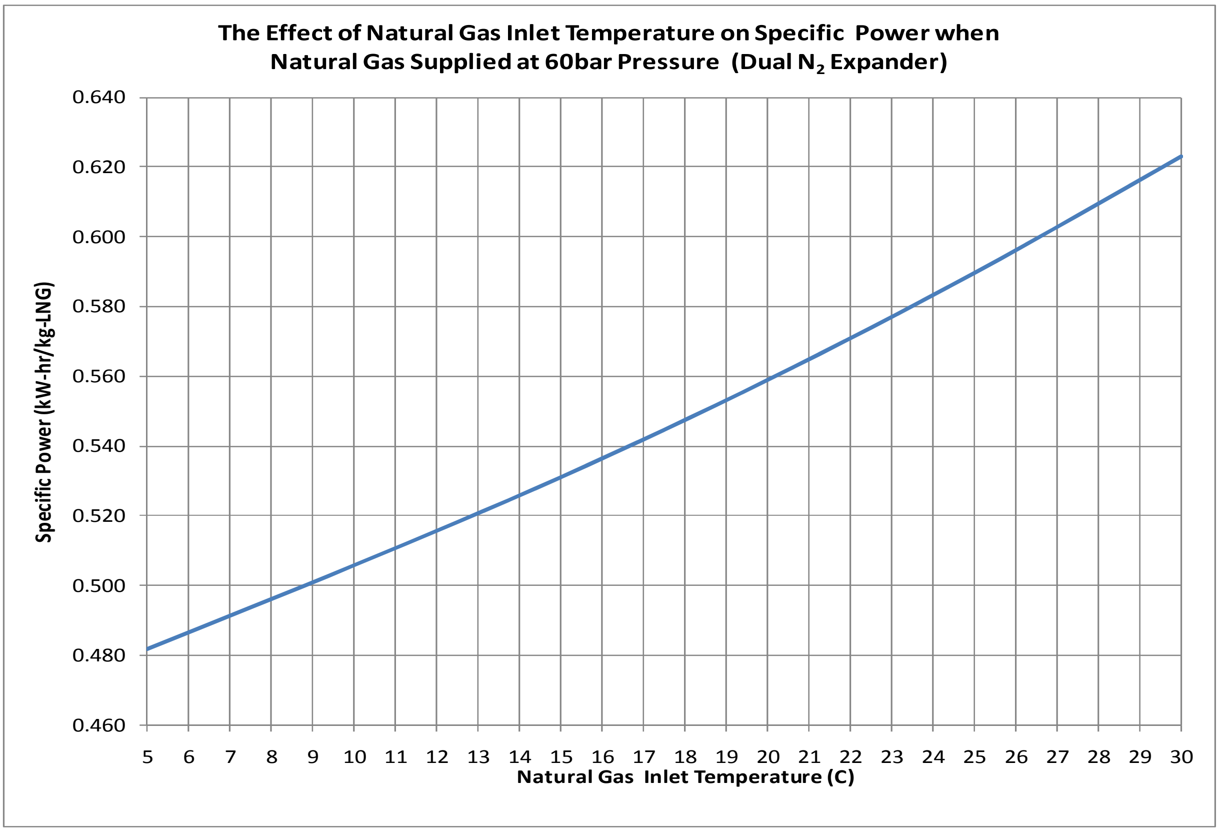

Figure 60: The effect of natural gas inlet temperature on specific power when natural gas supplied at 60bar (Dual N_2 Expander Process).

The graphs showing the effect of natural gas temperature on specific power when natural gas is supplied at different pressures (40, 60, 80, 100, 120 and 140bar) in dual nitrogen expander process are presented in figure-61 below. The graphs indicates that when natural gas temperature decreases from 15^{0}C to 5^{0}C dual nitrogen expander process specific power decreases and when natural gas temperature increases from 15^{0}C to 30^{0}C its specific power increases at any natural gas supplied pressure. Also the graphs depicted that when natural gas supplied is at higher pressure (eg. graph with 140bar), specific power is low compared to when natural gas is supplied at low pressure (e.g graph with 40bar) at any supplied natural gas temperature.

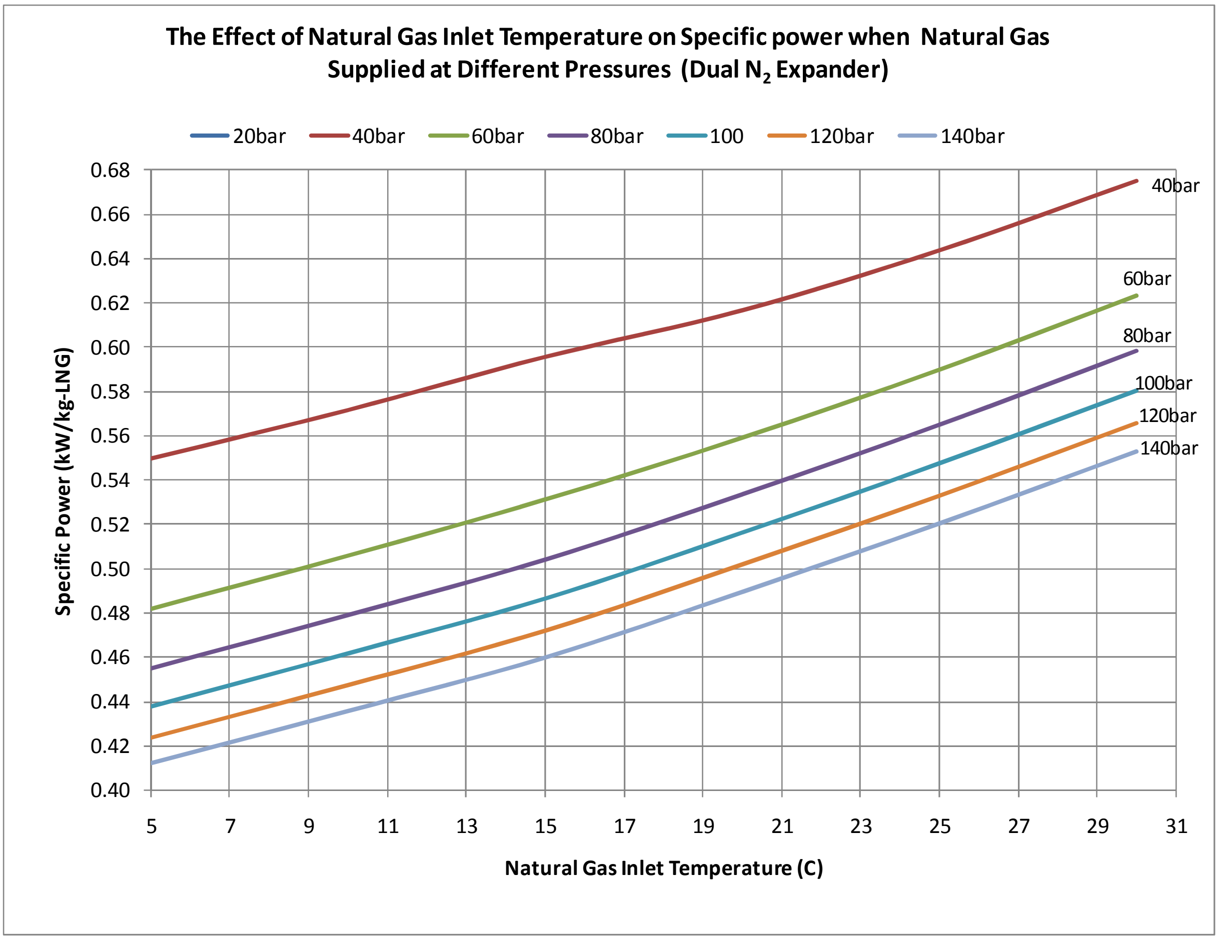

Figure 61: The effect of natural gas inlet temperature on specific power when natural gas supplied at different pressures (Dual Nitrogen Expander Process)

b) The effect of natural gas temperature on power consumption.

The effect of natural gas inlet temperature on power consumption when natural gas is supplied at 60bar in dual nitrogen expander process is presented in figure-62 below. The figure shows that as natural gas temperature decreases from 15^{0}C to 5^{0}C dual nitrogen expander process power consumption decreases from 65.21 to 59.17MW which amounts to a 9.27% reduction and when natural gas temperature increases from 15^{0}C to 25^{0}C the power consumption increases from 65.21 to 72.40MW which amounts to an 11.04% increase. It can be described that in dual nitrogen expander process when natural gas is supplied at given temperatures and pressures, decreases in natural gas temperature will lead to decreases in power consumption and increases in natural gas temperature will lead to increases in power consumption.

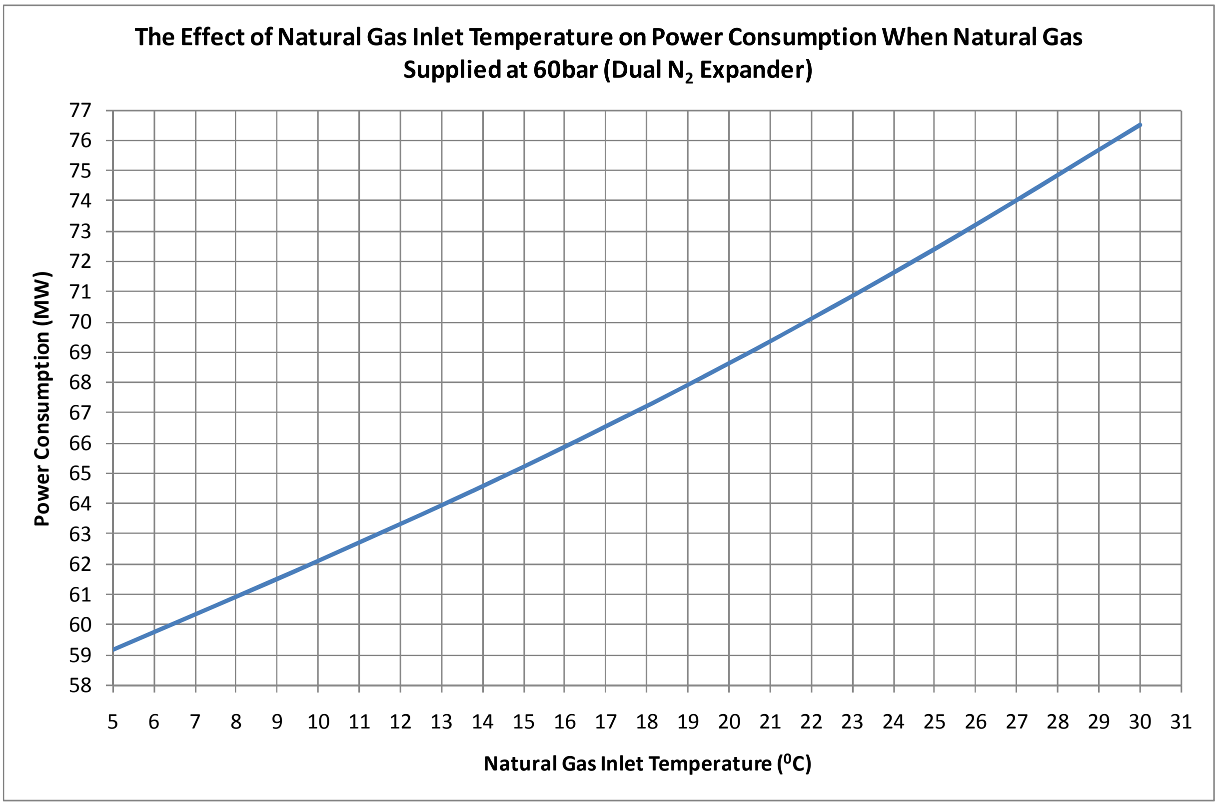

Figure 62: The effect of natural gas inlet temperature on power consumption when natural gas supplied 60bar (Dual Nitrogen Expander Process)

The graphs showing the effect of natural gas inlet temperature on power consumption when natural gas supplied at different pressures (40, 60, 80, 100, 120 and 140bar) in dual nitrogen expander process are presented in figure-63. The figure shows that when natural gas inlet temperature decreases from 15^{0}C to 5^{0}C the power consumption of dual nitrogen expander process decreases and when natural gas supply temperature increases from 15^{0}C to 30^{0}C then its power consumption increases at any natural gas supply pressure. The graphs also depicted that when natural gas is supplied at higher pressures the power consumption is lower (eg the graph with 140bar) compared to when natural gas is supplied at low pressures (eg. graph with 40bar) at any supply temperature. Also the graph indicates that when natural gas is supplied at pressures below 60bar, power consumption increases dramatically as shown in the space between graphs with 60bar and 40bar in the given figure below.

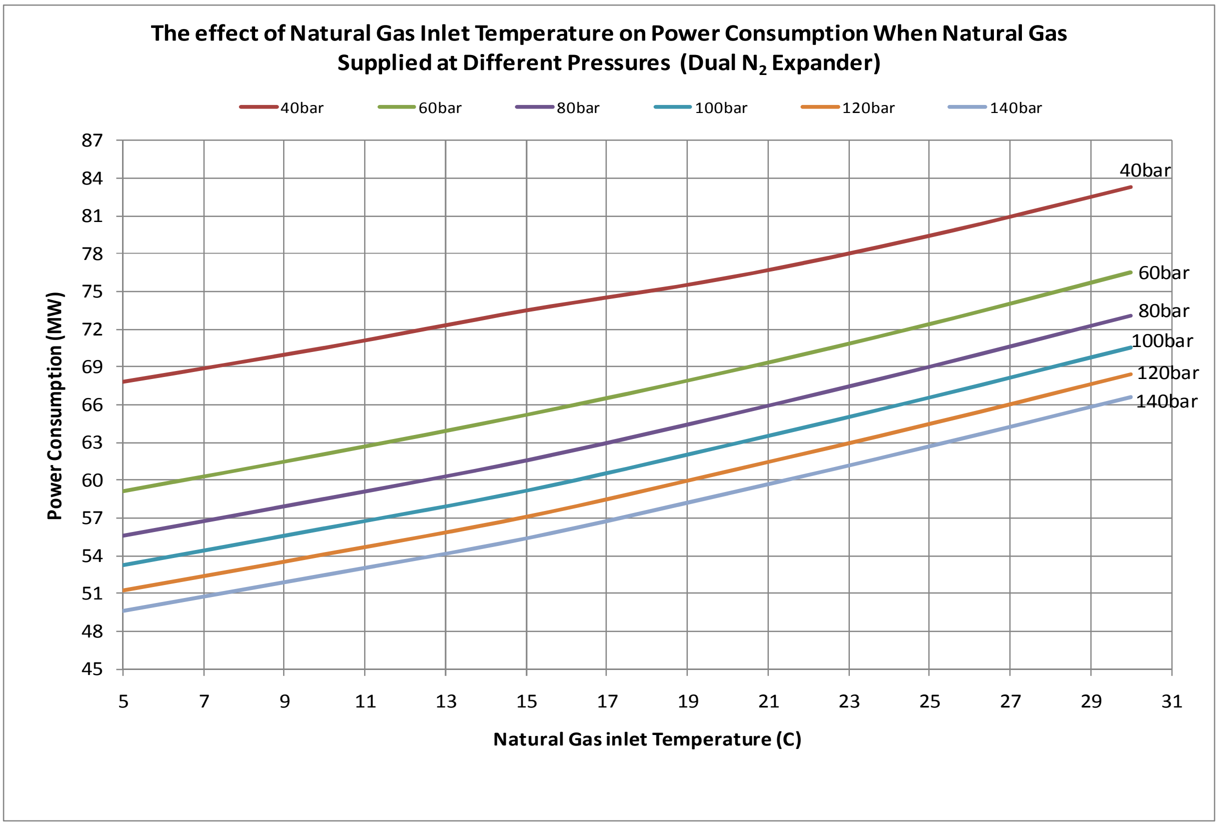

Figure 63: The effect of natural gas inlet temperature on power consumption when natural gas supplied at different pressures (Dual Nitrogen Expander Process).

8.2.2 The effect of natural gas pressure on dual nitrogen expander process

a) The effect of natural gas pressure on specific power

The effect of natural gas pressure on specific power when natural gas is supplied at 15^{0}C in dual nitrogen expander process is presented in figure-64 below. The figure indicates that when natural gas pressure decreases from 60bar to 30bar the dual nitrogen expander process specific power increases from 0.531 to 0.650 kW/kg-LNG which accounts to about 22.41 % increase and when natural gas pressure increases from 60bar to 90bar, specific power decreases from 0.531 to 0.495kW/kg-LNG which amounts to a 6.8% decrease. The graph also shows that when natural gas pressure decreases below 60bar there is a sharp increase of specific power.

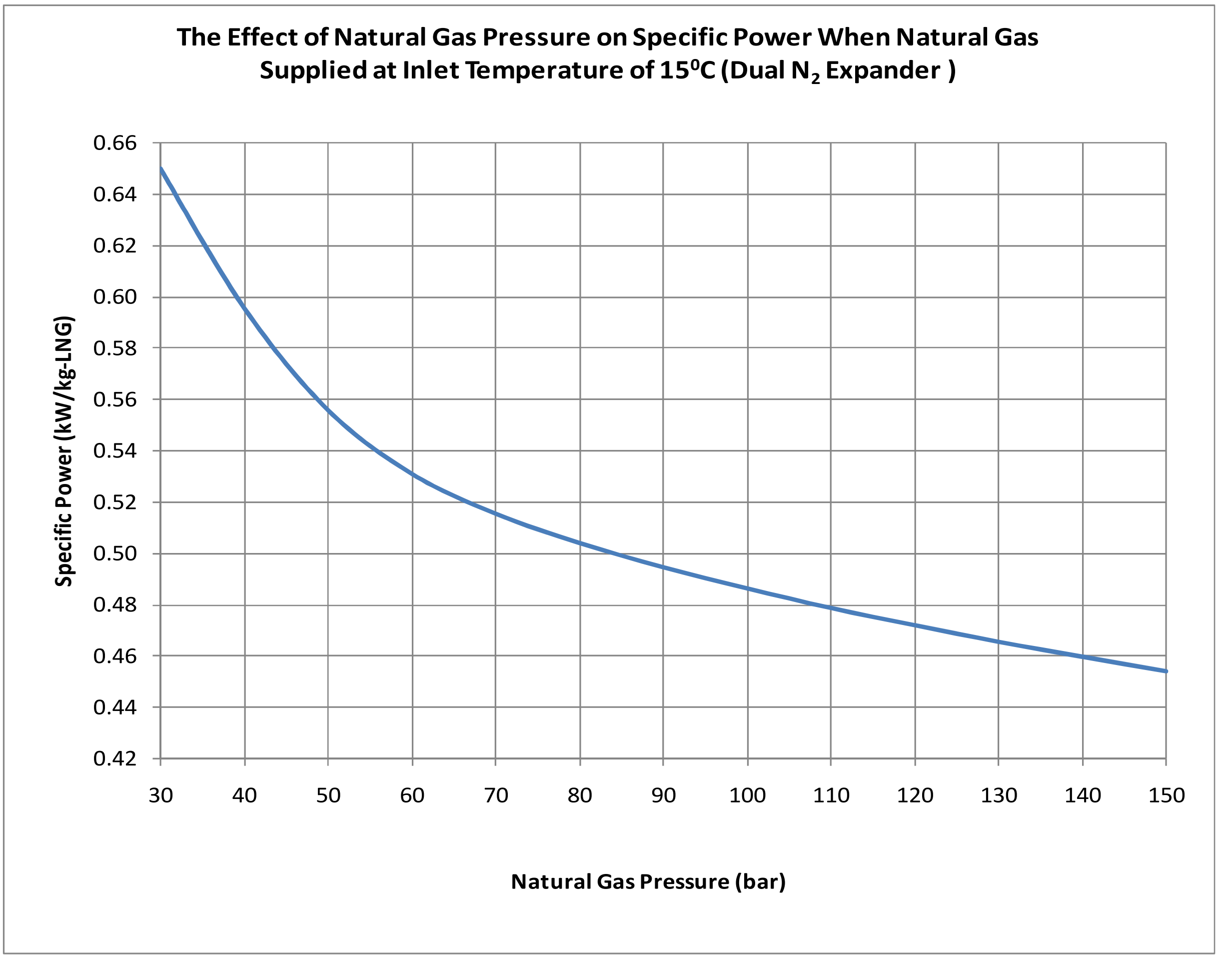

Figure 64: The effect of natural gas pressure on specific power when natural gas supplied at 15^{0}C (Dual nitrogen expander process)

The graphs showing the effect of natural gas pressure on specific power when natural gas is supplied at different temperatures (5, 10, 15, 20, 25 and 30^{0}C) in dual nitrogen expander process are presented in figure-65 below. The figure shows that when natural gas pressure decreases from 60bar to 30bar, the dual nitrogen expander process specific power increases and when natural gas pressure increases from 60bar to 150bar its specific power decreases at any natural gas supply temperature . Also the graphs indicates that when natural gas is supplied at low temperatures, specific power is low (eg graph with temperature 5^{0}C) and when natural gas is supplied at high temperatures specific power is high (eg graph with 30^{0}C) at any natural gas supply pressure. More over the figure shows that there is a sharp increase of specific power when natural gas is supplied with pressures less than 60bar.

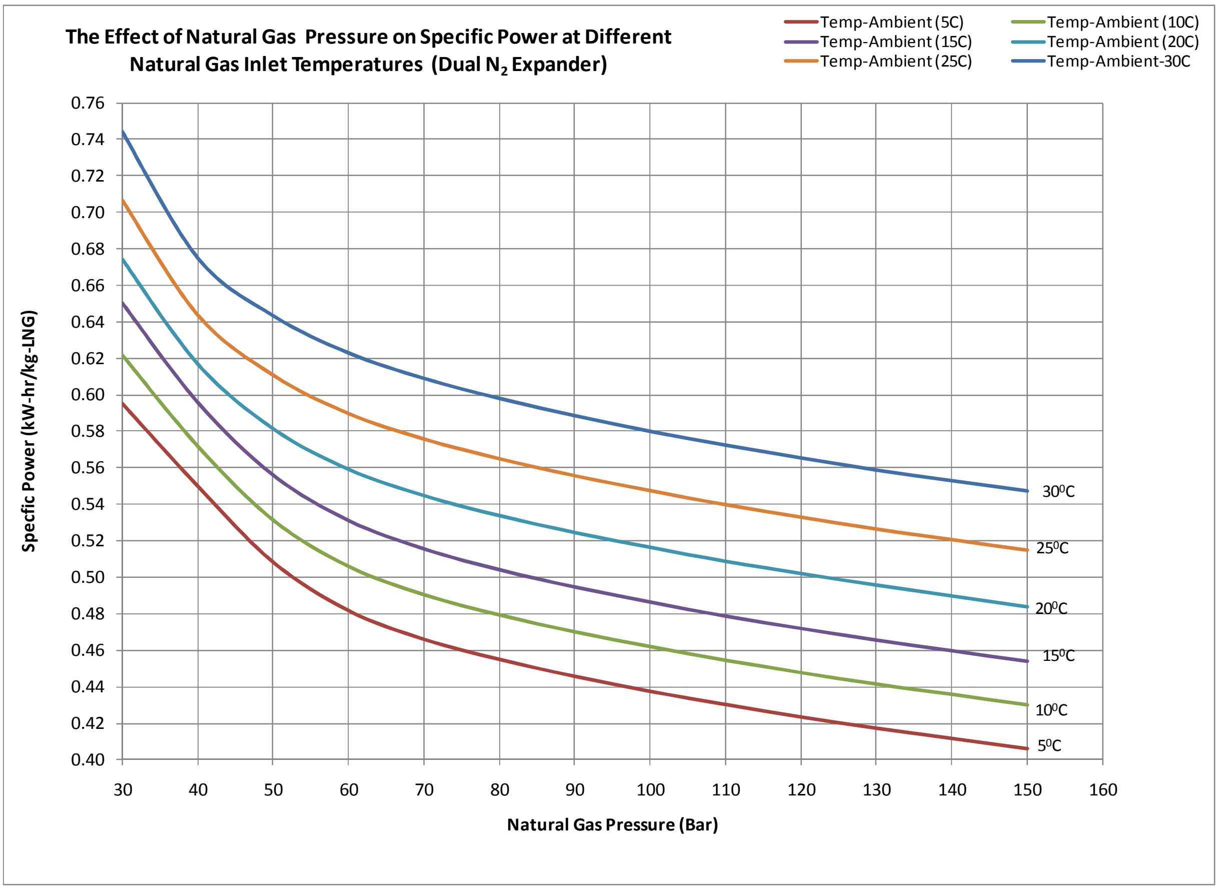

Figure 65: The effect of natural gas pressure on specific power when natural gas supplied at different inlet temperatures (Dual Nitrogen Expander Process).

b) The effect of natural gas pressure on power consumption

The effect of natural gas pressure on power consumption when natural gas is supplied at temperature of 15^{0}C in dual nitrogen expander process is given in figure-66 below. The figure shows that as pressure decreases from 60bar to 30bar, dual nitrogen expander process power consumption increases from 65.21 to 80.37 MW which amounts to a 23.25% increase and when natural gas pressure increases from 60bar to 90bar, its power consumption decreases from 65.21 to 60.31MW which amounts to a 7.51% decrease. Also the graph indicates that as pressure of natural gas decreases below 60bar the power consumption increases sharply.

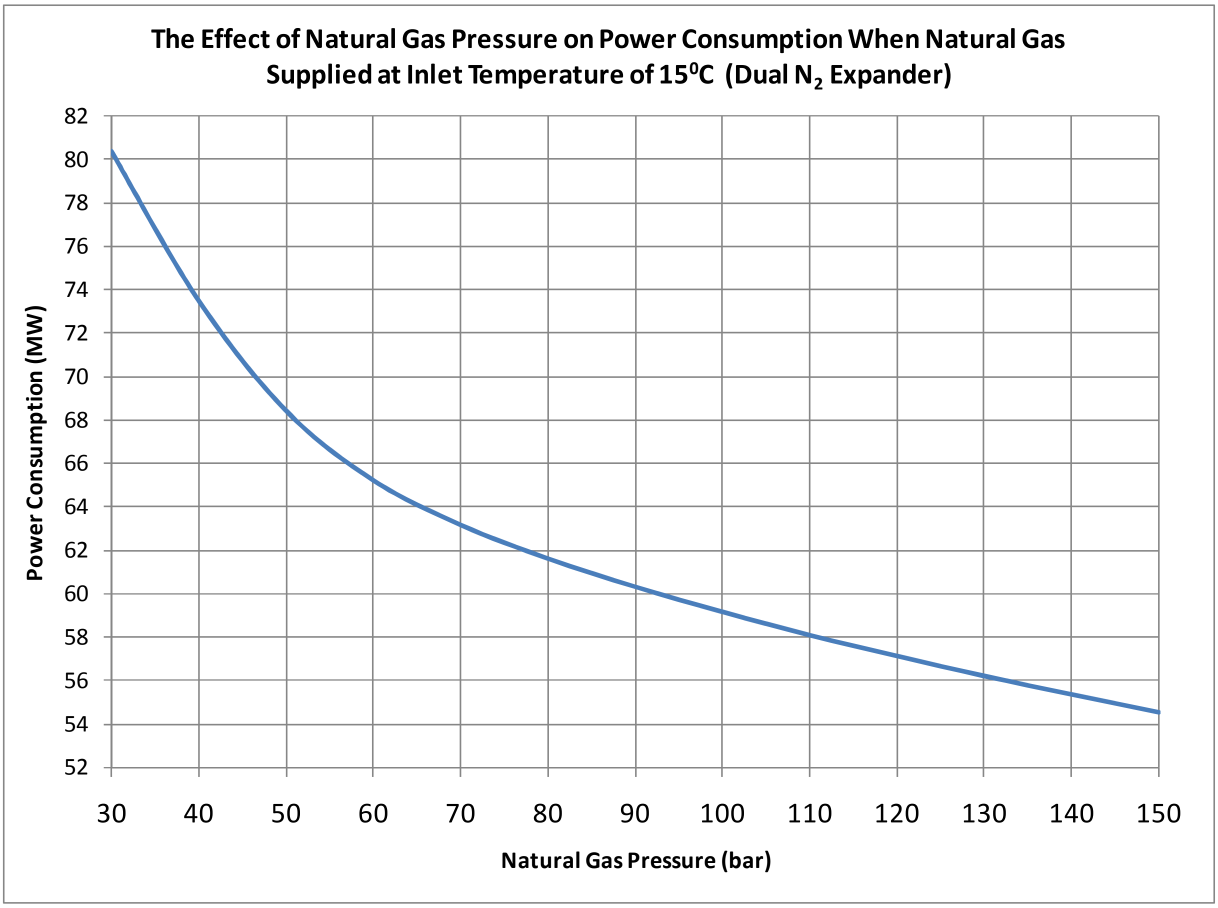

Figure 66 : The effect of natural gas pressure on power consumption when natural gas supplied at inlet temperature of 15^{0}C (Dual Nitrogen Expander Process).

The graphs showing the effect of natural gas pressure on power consumption when natural gas is supplied at different temperatures (5, 10,15,20,25 and 30^{0}C) in dual nitrogen expander process are presented in figure-67 below. The figure shows that as pressure decreases from 60 to 30bar, the dual nitrogen expander process power consumption increases and when natural gas pressure increases from 60 to 150bar its power consumption decreases at any natural gas supply temperature. The graph also indicates that when the supply temperature of natural gas is low, power consumption is low (eg graph with temperature of 5^{0}C) and when the supply temperature of natural gas is high (e.g graph with temperature 30^{0}C) the power consumption is high. Also from the graph it can be seen clearly that when natural gas pressure is below 60bar there is a sharp increase in power consumption at any natural gas supply temperature.

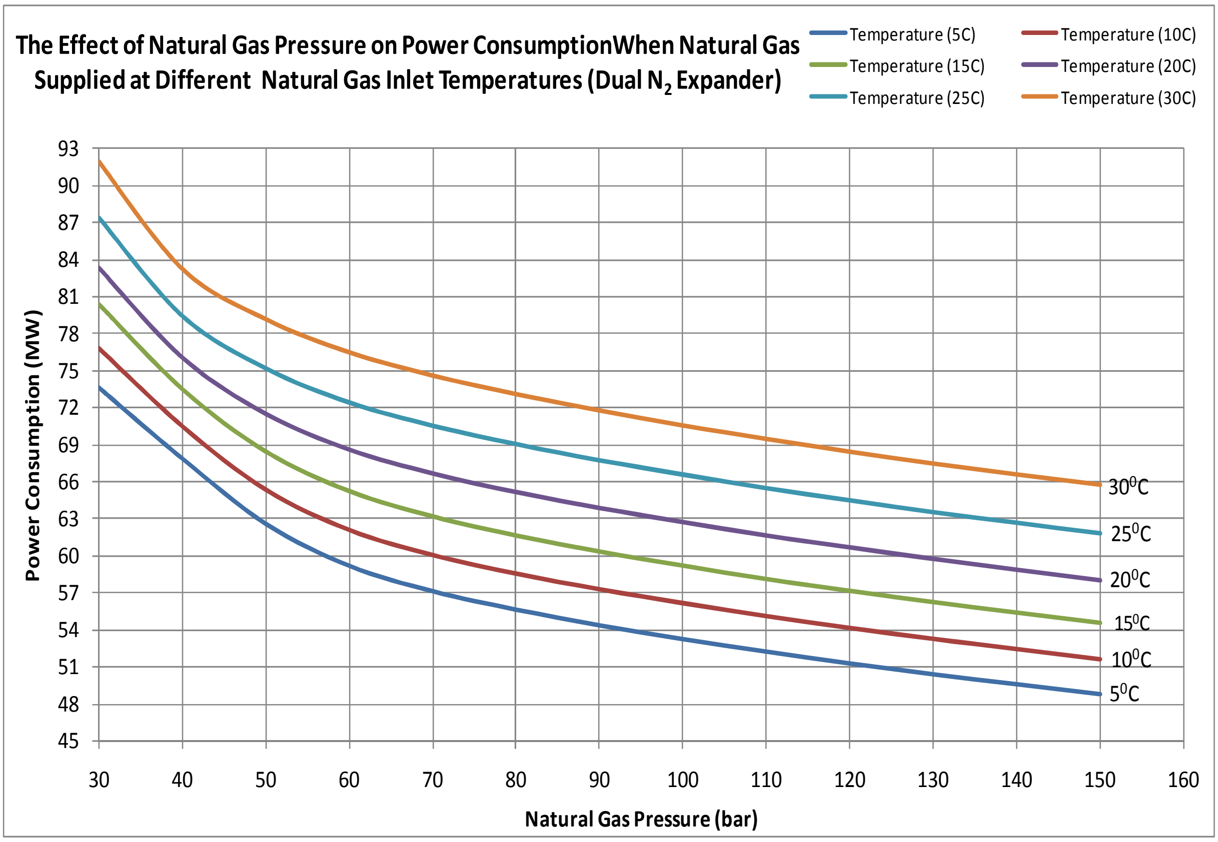

Figure 67: The effect of natural gas pressure on power consumption when natural gas supplied different temperatures (Dual nitrogen expander process).

8.2.3 The effect of natural gas composition on dual nitrogen expander Process

The effect of natural gas composition on dual nitrogen expander has been analyzed using five different natural gas composition termed as composition- A, B, C, D and E as presented in section 4.1 table (2 to 6). All the compositions were analyzed by comparison with a reference case (composition reference, table-1), with assumption that natural gas is supplied at 60bar and 15^{0}C for all compositions.

The effect of natural gas composition on specific power, power consumption and refrigerant flow rate was analyzed and presented in figure (68, 69&70) and table (28,29&30) below. The graph and tables indicates that changes in natural gas composition may lead to increase or decrease of process specific power, power consumption and refrigerant flow rate.

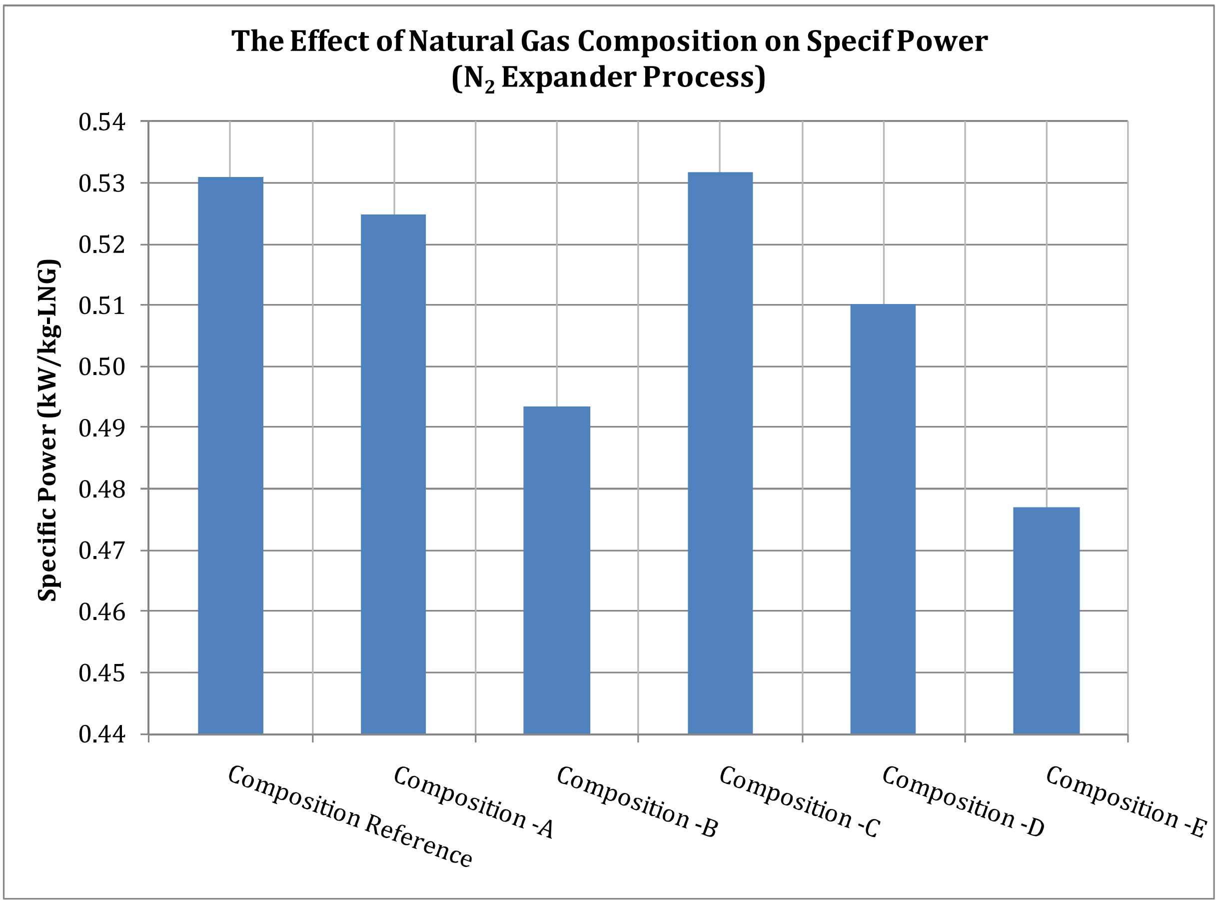

Figure 68: The effect of natural gas composition on specific power dual N_2 expander process

Table 28: The effect of natural gas composition on specific power dual N_2 expander process

Composition	Specific power (kWh/kg-LNG)	Increases or decreases
Composition-Reference	0.531	
Composition - A	0.525	Decreases by 1.1%
Composition -B	0.493	Decreases by 7.2%
Composition –C	0.532	Increase by 0.2%
Composition -D	0.510	Decrease by 4.0%
Composition -E	0.477	Decrease by 10.2%

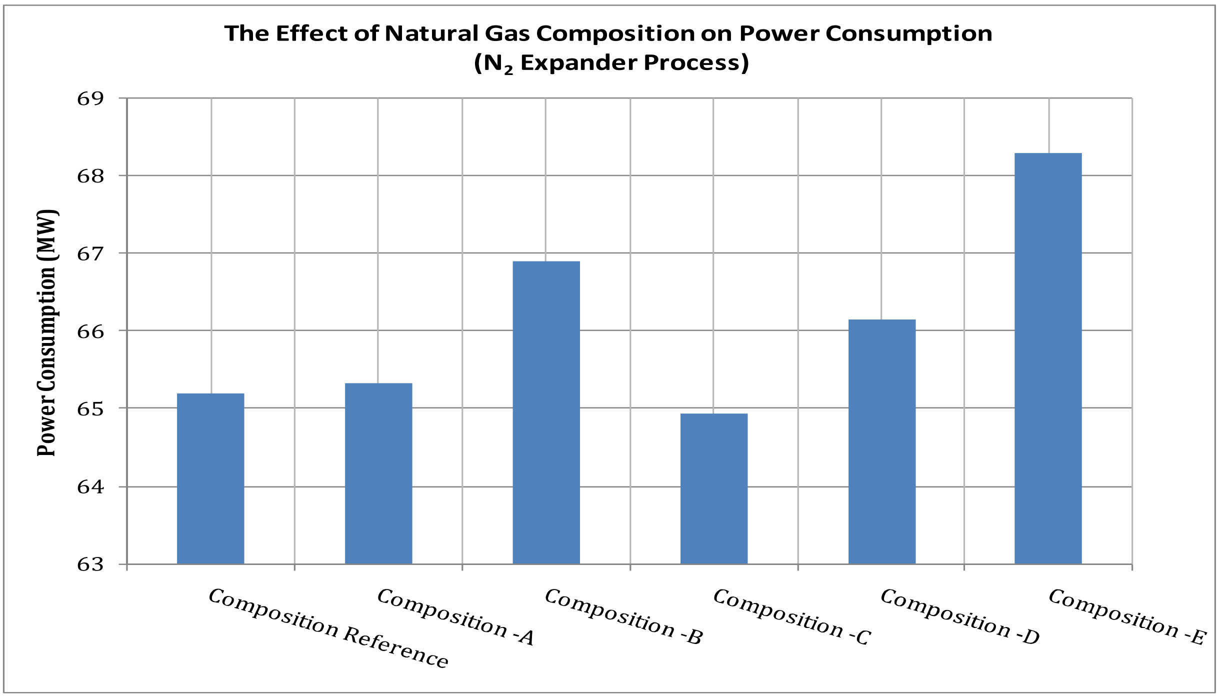

Figure 69: The effect of natural gas composition on power consumption dual N_2 expander Process

Table 29: The effect of natural gas composition on power consumption dual N_2 expander process

Composition	Power consumption (MW)	Increases or decreases
Composition-Reference	65.20	Used as reference for calculation
Composition - A	65.33	Increase by 0.2%
Composition -B	66.90	Increase by 1.7%

Composition -C	64.93	Decrease by 0.4%
Composition -D	66.15	Increase by 1.5%
Composition -E	68.30	Increase by 4.8%

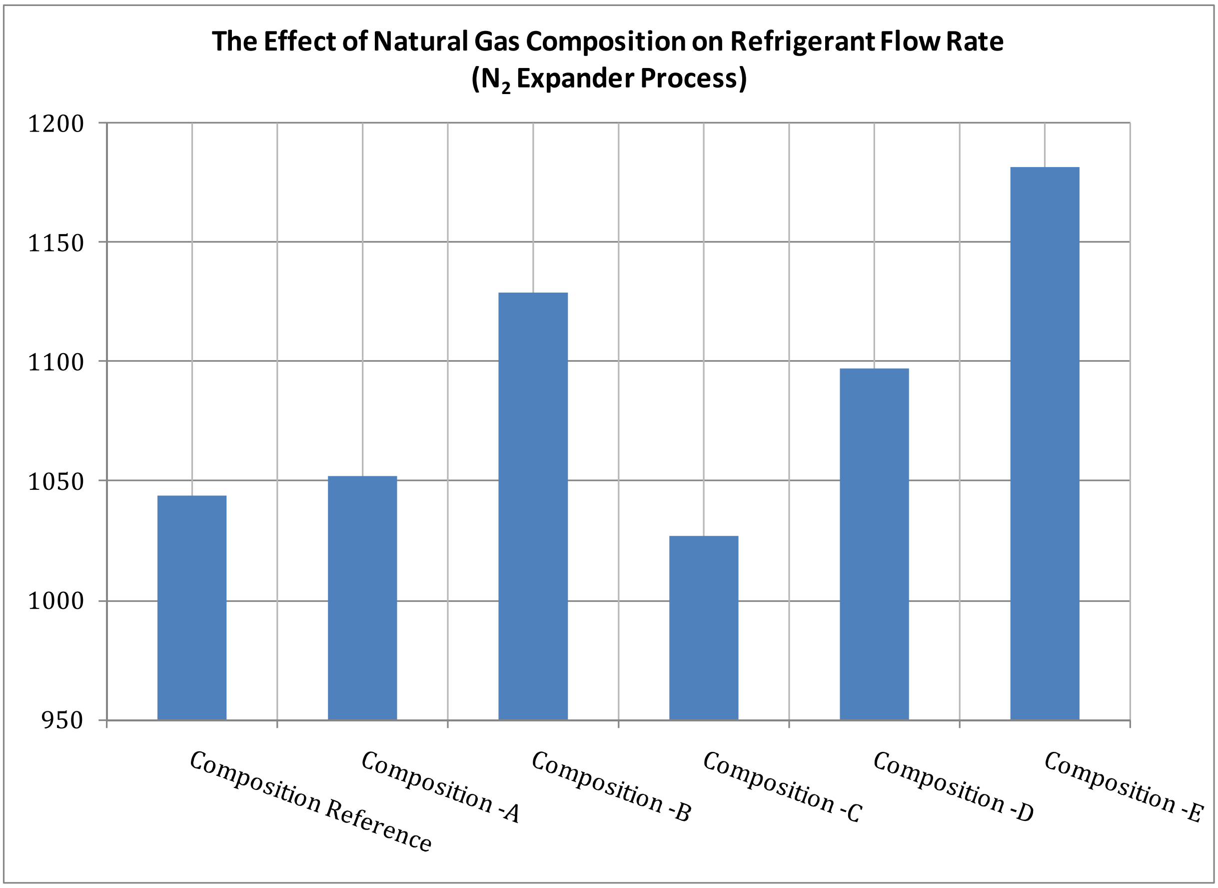

Figure 70: The effect of natural gas composition on refrigerant flow rate dual N_2 expander process

Table 30: The effect of natural gas composition on refrigerant flow rate dual N_2 expander process

Composition	Refrigerant flow (mmscfd)	Increases or decreases
Composition-Reference	1043.91	Used as reference for calculation
Composition - A	1052.29	Slightly increase 0.8%
Composition -B	1128.68	Increase by 8.1%
Composition -C	1027.31	Slightly decrease 1.6%
Composition -D	1097.22	Increase by 5.1%
Composition -E	1181.95	Increase by 13.2%

8.2.4 Exergy analysis

Distribution of exergy losses and useful exergy in different components is presented in figure-71 below. The useful energy/effect accounts about 22% and the amount of exergy losses on each component and useful exergy is present in table-31 below.

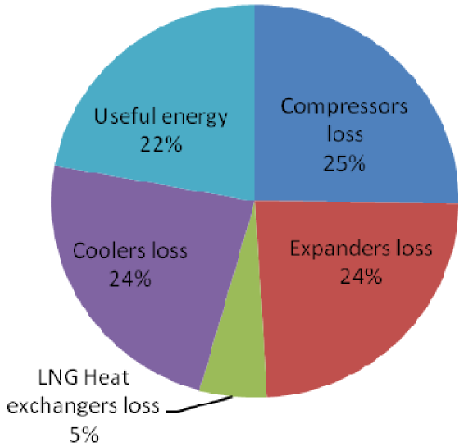

Figure 71: Exergy losses distribution at different component and useful effect (Dual N_2 expander process)

Table 31: The amount of exergy destroyed on each component and useful effect (Dual N_2 Expander Process)

Item	Exergy destroyed (MW)
Compressor (C-1)	4.30
Compressor (C-2)	5.99
Compressor (C-3)	2.83
Compressor (C-4)	2.54
Expander (EXP-1)	10.33
Expander (EXP-2)	4.52
LNG Heat exchanger (LNG-1)	1.65
LNG Heat exchanger (LNG-2)	1.67
Cooler (HEX-1)	2.63
Cooler (HEX-2)	6.30
Cooler (HEX-3)	3.02
Cooler (HEX-4)	2.67
Total Exergy destroyed	48.47
Total Energy supplied (W-real)	62.21
Useful Effect	13.74

8.2.5 Product Quality and Production Capacity of dual nitrogen expander Process

a) Production Capacity

The production capacity of simulated dual nitrogen expander process was analyzed based on maximum duty of one LM6000 gas turbine which has a capacity of 40.7MW [31]. The assumption was made on natural gas supplied at flow rate of 150mmscfd and plant operation for 330 days per annual.

The production capacity per train of dual nitrogen expander when natural gas is supplied at temperature of 15^{0}C and different pressures is presented in figure-72 below. The figure show that natural gas when supplied at higher pressure has higher production capacity per train. Considering natural gas supplied at 60bar, the train production capacity is estimated to be 0.61MTPA/train. Also the simulation indicated that natural gas supplied at 150mmscfd , 60bar and 15^{0}C can produce 0.97MTPA of LNG.

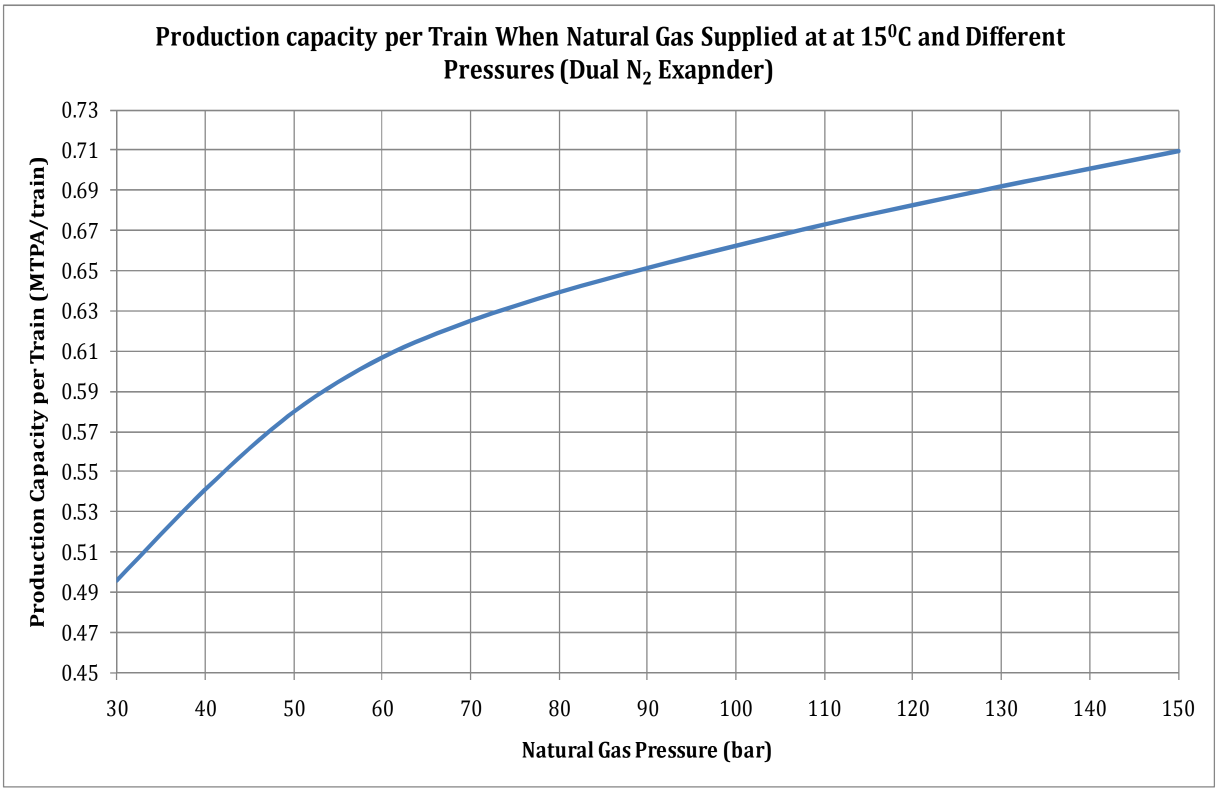

Figure 72: Production capacity per train when natural gas supplied at 15^{0}C and Different pressure (Dual Nitrogen Expander)

b) Product Quality

The quality of LNG product produced by dual nitrogen expander is analyzed on two parameters; high heating value (HHV) and wobbe index. The LNG high heating value was calculated to 39.42MJ/Sm^3 and wobbe index 50.94 this value are within range as described in detail in appendix-A.

8.3 Summary of simulation results of dual nitrogen expander Process

Table 32 : Summary of effect of natural gas temperature on dual nitrogen expander Process

	Temperature decreased from 15^{0}C to 5^{0}C	**Temperature increased from 15^{0}C to 25^{0}C**
Specific power (kWh/kg-LNG)	Decreases from 0.531 to 0.482 which amounts to 9.2%	Increases from 0.531 to 0.590 which amounts to 11.11%
Power consumption (MW)	Decreases from 65.21 to 59.17 which amounts to 9.27%	Increases from 65.21 to 72.40 which amounts to 11.04%

Table 33 : Summary of effect of natural gas pressure on dual nitrogen expander Process

	Pressure decreased from 60 to 30bar	Pressure increased from 60 to 90bar
Specific power (kWh/kg-LNG)	Increases from 0.531 to 0.650 which amounts to 22.41 %	Decreases from 0.531 to 0.495 which amounts to 6.8%.
Power consumption (MW)	Increases from 65.21 to 80.37 which amounts to 23.25%	Decrease from 65.21 to 60.31 which amounts to 7.51%

9. Discussion of results

The discussion of results is based on comparing the effect of natural gas pressure, temperature and composition of the four analyzed liquefaction processes proposed for LNG FPSO.

a) The Processes specific power, power consumption, refrigerant flow rate and train capacity

The comparison of specific powers of the proposed LNG liquefaction processes for LNG FPSO is presented in figure-73 and table-34 below. The presented comparison is based on natural gas supplied at pressure 60bar, temperature of 15^{0}C and molar flow rate of 150mmscfd .The figure shows that DMR has lowest specific power while dual nitrogen expander has the highest. The specific power of dual nitrogen expander process exceeded that of DMR by 50%, while Niche LNG exceeded by 41.6% and SMR by 9.6% as presented in table-34 below. It can be concluded that DMR has lowest specific power compared to other proposed process followed by SMR.

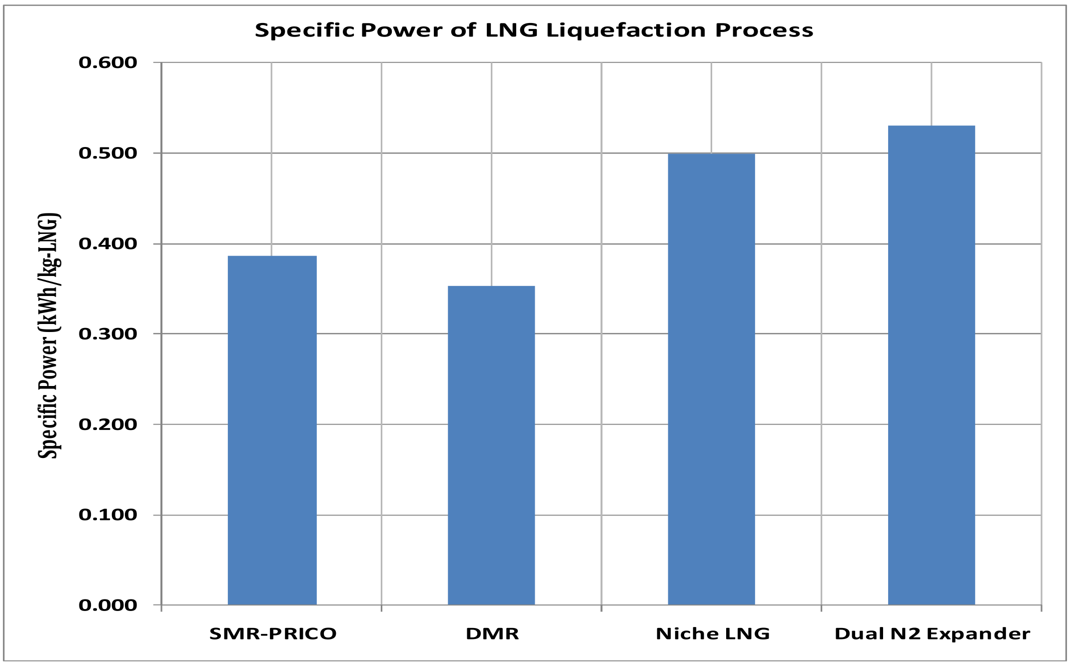

Figure 73 : The comparison of the specific power of proposed liquefaction processes for LNG FPSO

Table 34: The comparison of specific power of the proposed liquefaction processes for LNG FPSO

Processes	**Specific Power kWh/kg-LNG**	**Percentage Difference (%)**
SMR-PRICO	0.387	109.63
DMR	0.353	100.00
Niche LNG	0.500	141.64
Dual N_2 Expander	0.531	150.42

The comparison of power consumptions of the proposed LNG liquefaction processes for LNG FPSO is presented in figure-74 and table-35 below. The figure indicates that DMR has lowest power consumption compared to the other processes. Table-35 shows that DMR power consumption is lower than that of dual nitrogen expander by 54%, Niche LNG by 47.8% and SMR by 9.6% based on natural gas supplied at pressure 60bar, temperature of 15^0C and molar flow rate of 150mmscfd.

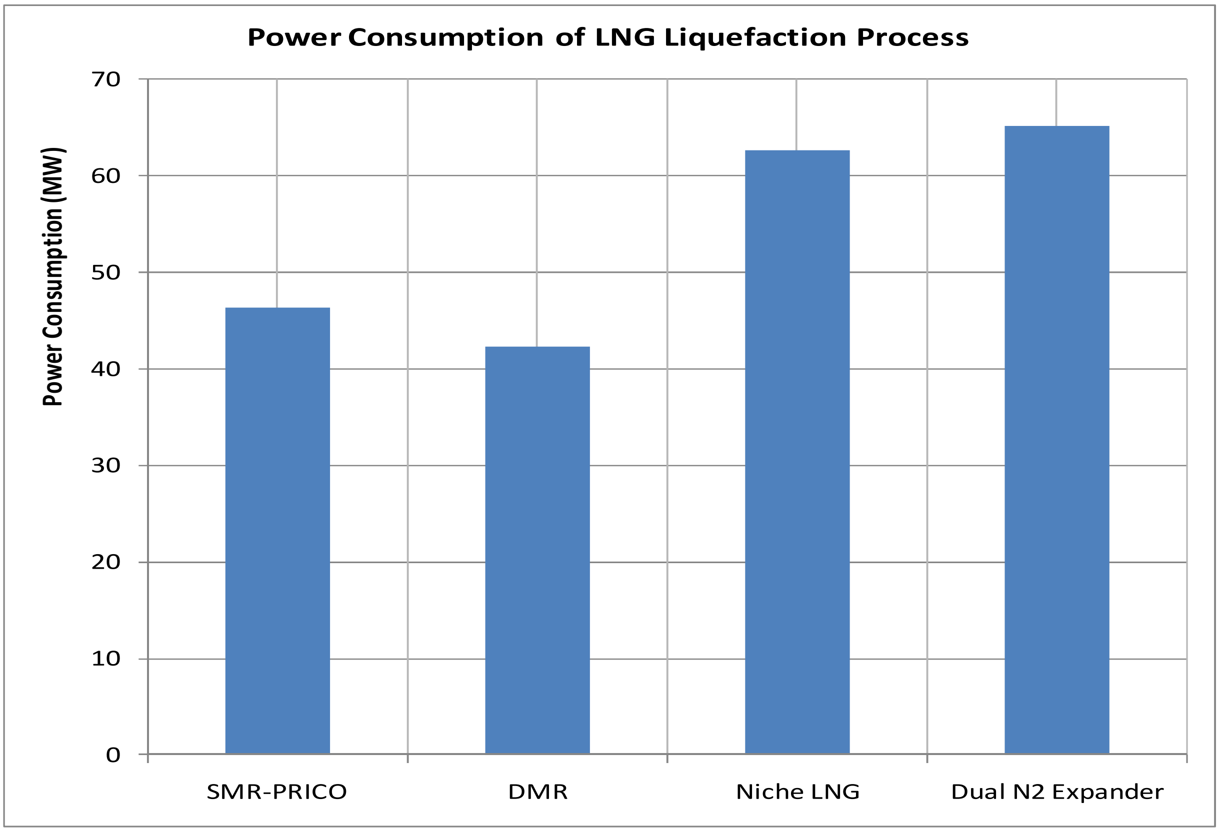

Figure 74: Comparison of power consumption of the proposed liquefaction processes for LNG FPSO

Table 35: The comparison of power consumption of the proposed LNG liquefaction process for LNG FPSO

Processes	Power Consumption (MW)	Percentage Difference (%)
SMR-PRICO	46.42	109.64
DMR	42.34	100.00
Niche LNG	62.59	147.83
Dual N_2 Expander	65.21	154.02

The comparison of refrigerant flow rate of proposed LNG liquefaction process for LNG-FPSO based on natural gas supplied at pressure 60bar, temperature of 15^{0}C and molar flow rate of 150mmscfd is presented in figure-75 and table-36 below. The figure shows that DMR has lowest refrigerant flow rate compared to the other process while dual nitrogen expander has highest flow rate. Table-36 shows that dual nitrogen expander refrigerant flow rate exceeded that of DMR by 157.6%, while Niche LNG exceeds DMR 96.4% and SMR 30.9%.

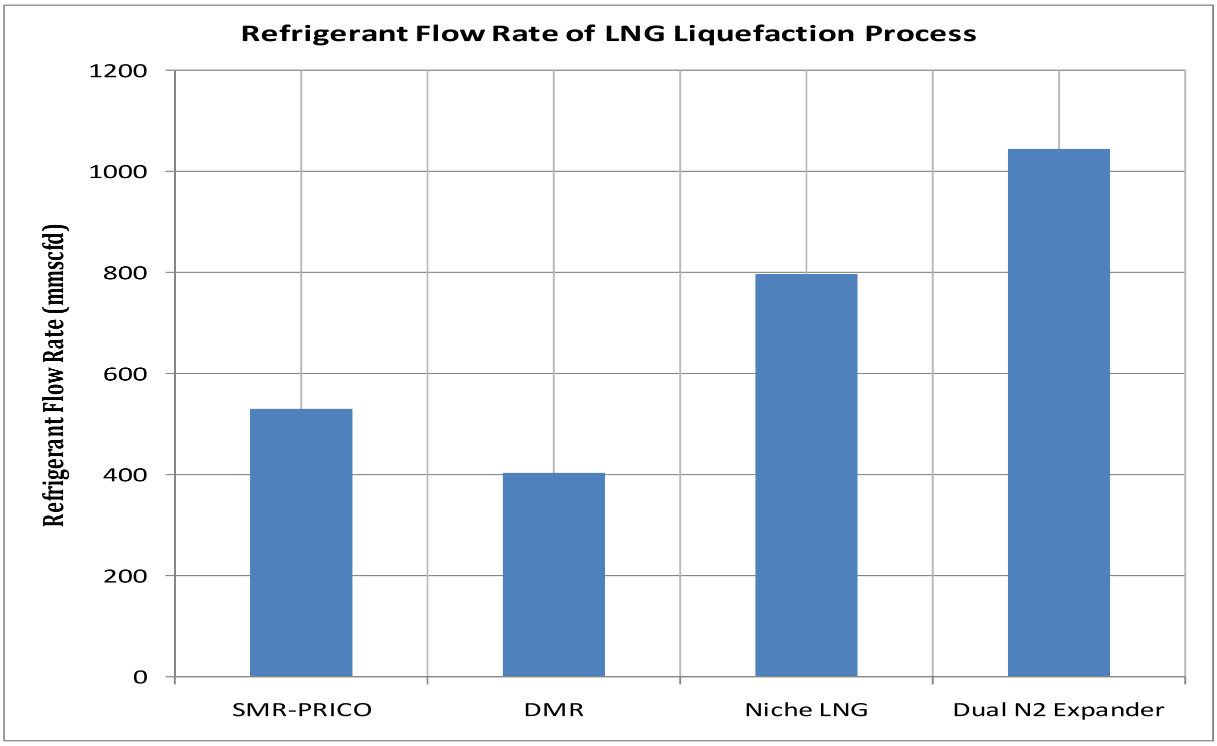

Figure 75: The comparison of refrigerant flow rate of the proposed LNG liquefaction process for LNG FPSO

Table 36: The comparison of refrigerant flow rate of the proposed LNG liquefaction Process for LNG FPSO

Processes	Refrigerant Flow rate (mmscfd)	Percentage Difference (%)
SMR-PRICO	531.28	130.96
DMR	405.67	100.00
Niche LNG	796.71	196.39
Dual N_2 Expander	1045.08	257.62

The comparison of the production capacity of the proposed LNG liquefaction processes for LNG FPSO based on natural gas supplied at pressure 60bar, temperature of 15^0C and molar flow rate of 150mmscfd with power supplied by one LM6000 turbine is presented in figure-76 and table-37 below. The figure shows that DMR has higher production capacity compared to the other process. The table-37 shows that DMR production capacity exceeded that of dual nitrogen expander by 33%, Niche LNG by 29.7% and SMR by 8.8%.

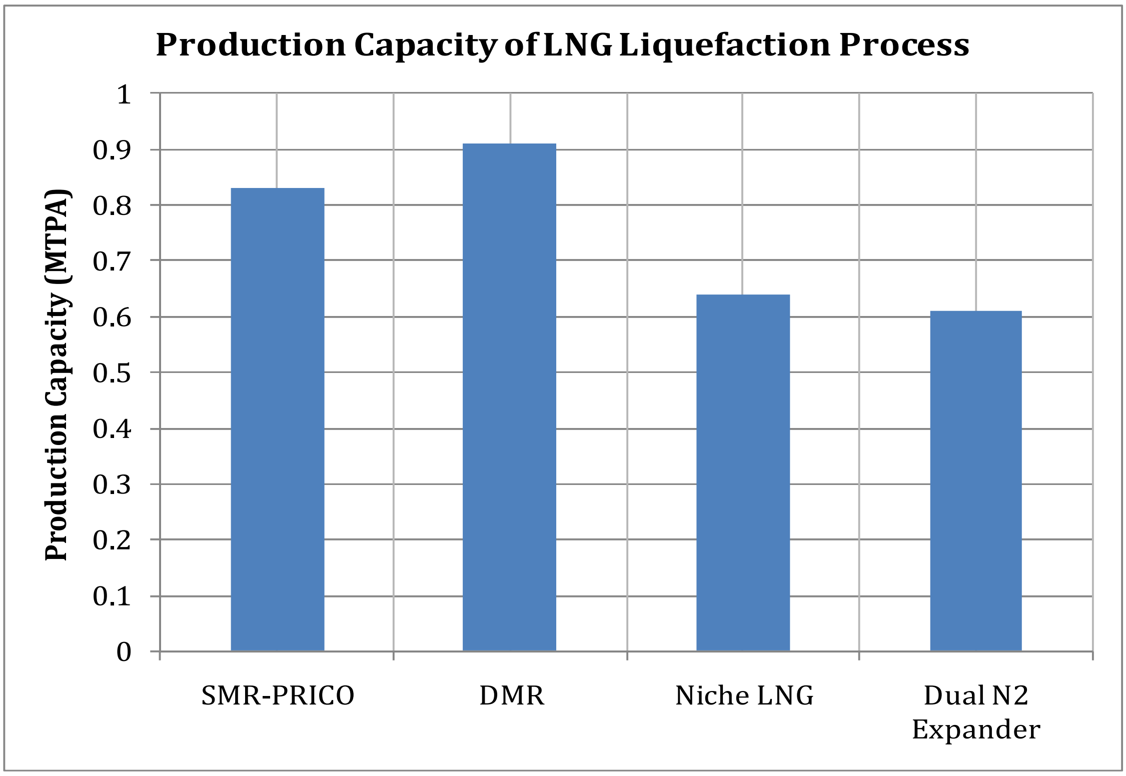

Figure 76: The comparison of production capacity of the proposed LNG processes for LNG FPSO

Table 37: The comparison of production capacity of the proposed LNG liquefaction processes for LNG FPSO

Processes	**Production Per train (MTPA/Train)**	**Percentage Difference (%)**
SMR-PRICO	0.83	91.21
DMR	0.91	100.00
Niche LNG	0.64	70.33
Dual N_2 Expander	0.61	67.03

b) The proposed Liquefaction process LNG heat exchangers "UA" values

Based on the benchmarked simulated processes the LNG heat exchangers "UA" values of each process was calculated and are presented in table-38 below. The table shows that dual nitrogen expander LNG heat exchangers have lowest "UA" value while MSR LNG heat exchanger has the highest values.

Table 38: The comparison of LNG Heat exchangers "UA" values (kW/^{0}C) for the proposed liquefaction process for LNG FPSO

Process	LNG-1 (UA Value)(kW/C)	LNG-2 (UA-Value)	Total (UA-Value)
SMR-PRICO	16,629. 21	-	16,629.21
DMR	5,213.25	4,691.53	9,904.78
Niche LNG	8,122.290	6,064.18	14,186.47
Dual N_2 Expander	4,414.55	841.10	5,255.65

c) The effect of natural gas temperature on the proposed liquefaction processes for LNG FPSO

The effect of natural gas temperature on specific power and power consumption for the proposed liquefaction processes for LNG FPSO with assumption that natural gas is at 60bar is presented in table 39 & 40 below. The tables shows that when natural gas supply temperature decreases from 15 to 5^{0}C the processes specific power and power consumption decreases at different magnitude and when it increases from 15 to 25^{0}C the processes specific power and power consumption increases. The tables show that change in natural gas supply temperature has high effect on SMR process compared to the other processes based on process specific power and power consumption.

Table 39: The comparison of the effect of natural gas temperature on specific power for LNG liquefaction Processes proposed for LNG FPSO

LNG Process	Decrease of temperature from 15^0 C to 5^0C, specific power decreases by	Increase of temperature from 15^0C to 25^0C , specific power increases by
SMR	14.99%.	39.27%
DMR	6.51%	8.23%.
Niche LNG	9.8%	11.4%.
Dual N_2 Expander	9.2%	11.11%.

Table 40: The comparison of the effect of natural gas supply temperature on power consumption for LNG liquefaction processes proposed for LNG FPSO

LNG Process	Temperature decrease from 15^0 C to 5^0C, power consumption decreases by	Temperature Increase from 15^0C to 25^0C, power consumption increase by
SMR	15.10%	39.19%
DMR	6.7%	8.2%.
Niche LNG	9.9%	11.3%.
Dual N_2 Expander	9.27%	11.04%.

d) The effect of natural gas pressure on the proposed liquefaction processes for LNG FPSO

The effect of natural gas pressure on specific power and power consumption on the proposed liquefaction processes for LNG FPSO is presented in table 41&42. The tables shows that as natural gas supply pressure decreases from 60 to 30bar processes specific power and power consumption increases at different magnitudes and when natural gas supply pressure increases from 60 to 90bar the specific power and power consumption decreases at different magnitude. Considering decrease in natural gas supply pressure which is negative effect the tables show that dual nitrogen expander has the highest effect while Niche LNG has lowest effect and considering an increase in natural gas supply pressure which is positive effect DMR has highest effect and Niche LNG has lowest effect.

Table 41: The comparison of the effect of natural gas pressure on specific power for the proposed liquefaction process for LNG FPSO

LNG Process	Pressure decreasing from 60 to 30bar, specific power increases by	Pressure increasing from 60 to 90bar, specific power deceases by
SMR	15.4%	10.26%
DMR	18.70%.	13.60%
Niche LNG	7.2%	7%
Dual N_2 Expander	22.41 %	6.8%.

Table 42: The comparison of the effect of natural gas pressure on power consumption for the proposed liquefaction process for LNG FPSO

LNG Process	Pressure decreasing from 60 to 30bar, power consumption increases by	Pressure increasing from 60 to 90bar, power consumption decrease by,
SMR	17.60%	10.81%.
DMR	19.44%.	13.67%
Niche LNG	15.5%	7.7%.
Dual N_2 Expander	23.25%	7.51%.

e) The effect of natural gas composition on the proposed liquefaction processes for LNG FPSO

The effect of natural gas composition on the proposed natural gas liquefaction processes for LNG FPSO is well elaborated in results chapters of each process. The analysis shows that for all proposed processes the change in natural gas composition may lead to increase or decrease of processes specific power, power consumptions or refrigerant flow rates. The comparison of the effect of natural gas composition on the proposed processes using one of the analyzed composition (composition-E) is presented in table (43, 44 & 45) below.

Table 43 : The comparison of the effect of composition-E on specific power for proposed liquefaction processes for LNG FPSO

Processes	Effect of composition on specific power
SMR	Decreases by 8.0%
DMR	Decrease by 16.2%
Niche LNG	Decrease by 7.9%
Dual Nitrogen Expander	Decrease by 10.2%

Table 44 : The comparison of the effect of composition-E on power consumption for proposed liquefaction process for LNG FPSO

Processes	Effect of composition on power consumption
SMR	Increases by 3.4%
DMR	Decrease by 1.8%
Niche LNG	Increased by 7.2%
Dual Nitrogen Expander	Increase by 4.8%

Table 45 : The comparison of the effect of composition-E on refrigerant flow rate for proposed liquefaction process for LNG FPSO

Processes	Effect of composition on refrigerant flow rate
SMR	Increased by 9.3%
DMR	Decrease by 3.1%
Niche LNG	Increases by 7.6%
Dual Nitrogen Expander	Increase by 13.2%

f) The Exergy analysis of the proposed liquefaction process for LNG FRSO

The usefully exergy of the proposed natural gas liquefaction processes is presented in table-46 below. The table shows that DMR have highest useful exergy of about 31% compared to the other processes. Niche LNG and dual Nitrogen expander have almost same useful exergy.

Table 46: The comparison of useful exergy of proposed liquefaction processes for LNG FPSO

Processes	**Usefeul exergy (%)**
SMR	26
DMR	31
Niche LNG	20
Dual Nitrogen Expander	22

10. Conclusions

The comparison of specific powers, power consumption, refrigerant flow rate and production capacity concluded that DMR is the best compared to the other process. The specific power of DMR is lower than that of dual nitrogen expander process by 50%, Niche LNG by 41.6% and SMR by 9.6%. Its power consumption is lower than that of dual nitrogen expander 54%, Niche LNG 47.8% and SMR 9.6%. Also its refrigerant flow rate is lower than that of dual nitrogen expander by 157.6%, Niche LNG 96.4% and SMR 30.9%. More over its

Production capacity based on natural gas supplied at pressure power supplied by one LM6000 turbine is lower than that of dual nitrogen expander by 33%, Niche LNG by 29.7% and SMR by 8.8%. Based on analyzed parameters above DMR may consider as good candidate for LNG FPSO compared to the other processes if any only if the arguments on the impact of vessel motion and inherent to safety on hydrocarbon refrigerant proved to have negligible effect.

All process proved that natural gas supply temperature and pressure has significant impact. The impacts may be in positive or negative way depends on increase on decrease of supply temperature or pressure. One of the analyzed condition show that when natural gas supply temperature decreases from 15 to 5^0C SMR specific power and power consumption decrease by 14.99% and 15.10% respectively and when it is increases from 15 to 25^0C its specific power and power consumption increases by 39.27% and 39.19 respectively. Other analyzed condition on pressure shows that when natural gas supply decrease dual nitrogen expander has the highest effect with specific power and power consumption increases by 22.41% and 23.25% respectively and when natural gas supply increases DMR has highest effect on specific power and power consumption which are 13.06% and 13.67% respectively.

The effect of natural gas composition shows that for all proposed processes the change in natural gas composition may lead to increase or decrease of processes specific power, power consumptions or refrigerant flow rates.

Energy efficiency is important to LNG production as feed gas is consumed in order to carry out the liquefaction process.The exergy analysis of the proposed process shows that shows that DMR process has highest useful exergy about 31% compared to the other processes. Niche LNG and dual nitrogen expander has almost same useful exergy.

11. References

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12. Appendices

Appendix A: Calculation of Product Quality.

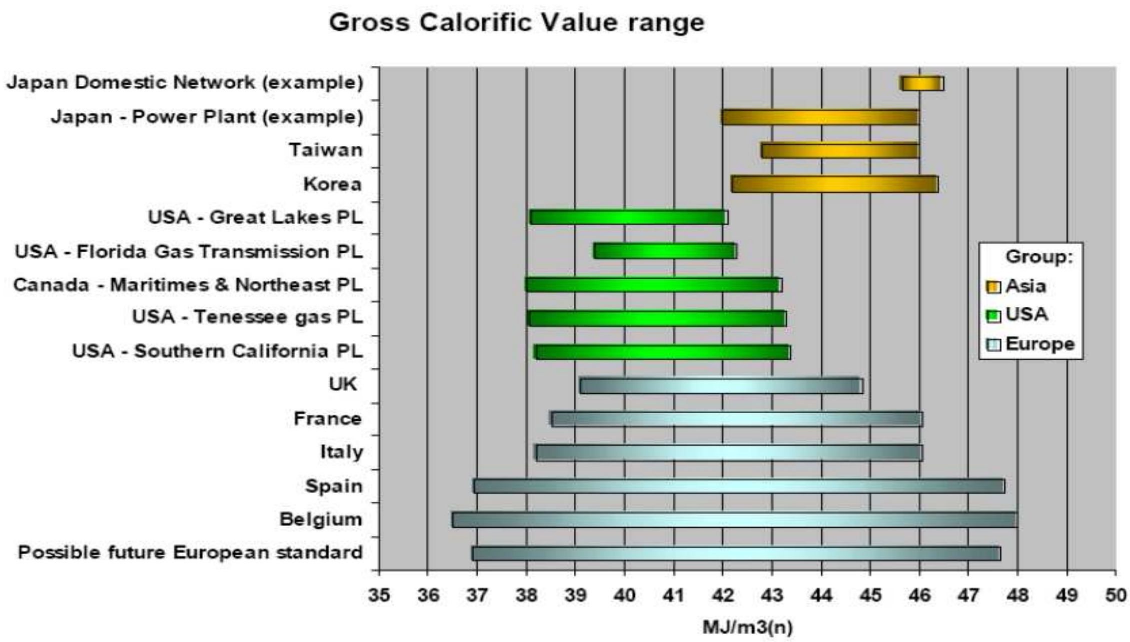
Wobbe Index (WI) is a density-corrected heating value, and this parameter reflects the behaviour of the fuel gas in a burner. It is an important factor in gas "interchangeability", i.e. evaluation if a specific gas composition can be fed into an existing gas distribution system. The Wobbe Index is defined as

$$\boxed{\mathrm{WobbeIndex} = \frac{\mathrm{GHV}}{\sqrt{\mathrm{spgr}}} = \frac{\mathrm{GHV}}{\sqrt{\frac{\mathrm{MW}}{28.964}}}}$$

Where

GHV: Gross Heating Value (MJ/Sm3) (same as Upper Heating Value)
spgr: specific gravity (-) – i.e. density in relation to air
MW: Molecular weight (kg/kmol)

An example of tolerable GCV (HHV) range in various gas transport/distribution systems is shown in Figure below.



Appendix B: The effect of natural gas supply pressure on SMR process when it is supplied at a temperature of 15⁰C

Natural Gas Pressure (bar)	Compressor Power (MW)	Refrigerant flow rate (mmscfd)	LNG Mass Flow (kg/s)	Specific Power (kW-hr/kgLNG)	LNG (MTPA)	Production per train
10	65.51	566.34	33.67	0.540	0.96	0.60
20	58.85	561.48	33.60	0.487	0.96	0.66
30	54.59	554.66	33.52	0.452	0.96	0.71
40	51.33	546.93	33.44	0.426	0.95	0.76
50	48.68	539.97	33.36	0.405	0.95	0.80
60	46.42	531.28	33.28	0.387	0.95	0.83
70	44.53	522.65	33.20	0.373	0.95	0.87
80	42.87	514.05	33.12	0.359	0.94	0.90
90	41.40	505.58	33.04	0.348	0.94	0.93
100	40.08	497.34	32.96	0.338	0.94	0.95
110	38.91	489.27	32.87	0.329	0.94	0.98
120	37.86	481.21	32.79	0.321	0.93	1.01
130	36.90	473.15	32.70	0.313	0.93	1.03
140	36.03	465.07	32.62	0.307	0.93	1.05
150	35.26	458.38	32.53	0.301	0.93	1.07

Appendix C: Comparison of Natural gas pressure, temperature and specific power on (SMR Process)

Temperature (C)	5	10	15	20	25	30
Natural gas Pressure (bar)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)
10	0.477	0.503	0.540	0.605	0.704	0.893
20	0.422	0.449	0.487	0.552	0.647	0.830
30	0.389	0.415	0.452	0.518	0.611	0.791
40	0.365	0.390	0.426	0.491	0.583	0.759
50	0.346	0.369	0.405	0.469	0.559	0.731
60	0.329	0.352	0.387	0.451	0.539	0.707
70	0.315	0.337	0.373	0.436	0.522	0.685
80	0.302	0.325	0.359	0.422	0.506	0.665
90	0.291	0.313	0.348	0.410	0.492	0.647
100	0.281	0.304	0.338	0.399	0.479	0.632
110	0.273	0.295	0.329	0.389	0.468	0.616
120	0.265	0.287	0.321	0.379	0.457	0.603
130	0.259	0.280	0.313	0.371	0.447	0.589
140	0.253	0.274	0.307	0.364	0.438	0.577
150	0.249	0.269	0.301	0.357	0.430	0.566

Appendix D: Comparison of Natural gas pressure, ambient temp and power Consumption (SMR Process)

Temperature (C)	5	10	15	20	25	30
Natural gas pressure (bar)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)
10	57.83	61.03	65.51	73.33	85.28	108.29
20	51.07	54.31	58.85	66.72	78.27	100.39
30	46.98	50.09	54.59	62.47	73.75	95.40
40	43.98	46.94	51.33	59.11	70.15	91.33
50	41.52	44.35	48.68	56.37	67.20	87.80
60	39.41	42.19	46.42	54.05	64.61	84.69
70	37.61	40.34	44.53	52.09	62.37	81.89
80	36.01	38.72	42.87	50.29	60.34	79.35
90	34.61	37.28	41.40	48.75	58.53	77.00
100	33.37	36.02	40.08	47.30	56.86	74.95
110	32.28	34.89	38.91	46.02	55.33	72.92
120	31.33	33.89	37.86	44.79	53.92	71.14
130	30.49	33.01	36.90	43.72	52.68	69.36
140	29.75	32.21	36.03	42.74	51.45	67.79
150	29.10	31.51	35.26	41.83	50.38	66.31

Appendix E: Comparison of natural gas pressure, ambient temp and refrigerant flow (SMR Process)

Temperature (C)	5	10	15	20	25	30
Natural gas pressure (bar)	Refg(mmscfd)	Refg(mmscfd)	Refg(mmscfd)	Refg(mmscfd)	Refg(mmscfd)	Refg(mmscfd)
10	465.22	504.60	566.34	666.86	752.66	905.16
20	459.93	499.23	561.48	662.89	747.34	898.06
30	453.63	493.26	554.66	655.87	740.35	889.23
40	446.21	485.84	546.93	647.13	731.63	880.41
50	438.52	478.58	539.97	638.30	722.93	869.82
60	431.10	469.90	531.28	629.50	713.04	859.25
70	423.06	462.24	522.65	620.87	703.14	847.54
80	414.93	454.02	514.05	610.33	692.60	835.83
90	406.81	446.01	505.58	601.76	682.71	824.11
100	398.81	438.18	497.34	591.85	672.81	814.22
110	391.40	429.27	489.27	583.22	662.92	802.51
120	383.87	422.06	481.21	573.31	653.03	792.61
130	376.75	415.23	473.15	564.74	644.96	781.09
140	370.07	408.32	465.07	556.67	635.08	771.19
150	363.68	401.44	458.38	548.60	627.01	761.30

Appendix F: The effect of natural gas supply pressure on DMR process when it is supplied at temperature of 15^0C

Natural Gas Pressure (bar)	K-100 (MW)	K-101 (MW)	Cycle-1 Refg flow (mmscfd)	Cycle-2 Refg flow (mmscfd)	LNG Mass Flow (kg/s)	Total Power (MW)	Total refg flow (mmscfd)	Specific Power (kW-hr/kgLNG)	LNG (MTPA)	Production per train
20	32.52	23.67	196.91	269.10	33.60	56.19	466.01	0.465	0.96	0.69
30	27.38	23.19	193.82	260.88	33.52	50.57	454.70	0.419	0.96	0.77
40	24.58	22.64	190.06	250.86	33.44	47.21	440.92	0.392	0.95	0.82
50	22.64	22.03	185.73	238.85	33.36	44.67	424.57	0.372	0.95	0.87
60	20.94	21.39	181.14	224.53	33.28	42.34	405.67	0.353	0.95	0.91
70	19.30	20.72	176.02	207.90	33.20	40.02	383.91	0.335	0.95	0.96
80	17.88	20.05	171.03	192.14	33.12	37.92	363.16	0.318	0.94	1.01
90	16.87	19.43	166.67	180.52	33.04	36.30	347.18	0.305	0.94	1.06
100	16.19	18.87	162.98	172.48	32.96	35.06	335.46	0.296	0.94	1.09
110	15.70	18.39	159.66	166.65	32.87	34.08	326.30	0.288	0.94	1.12
120	15.31	17.96	156.64	162.17	32.79	33.27	318.81	0.282	0.93	1.14
130	15.01	17.58	153.87	158.57	32.70	32.58	312.45	0.277	0.93	1.16
140	14.75	17.23	151.33	155.59	32.62	31.99	306.92	0.272	0.93	1.18
150	14.54	16.93	148.98	153.07	32.53	31.46	302.05	0.269	0.93	1.20

Appendix G: Comparison of Natural gas pressure, temperature and specific power on (DMR Process)

Temperature (C)	5	10	15	20	25	30
Natural gas Pressure (bar)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)
20	0.439	0.450	0.465	0.480	0.496	0.514
30	0.394	0.406	0.419	0.434	0.450	0.468
40	0.367	0.379	0.392	0.407	0.423	0.440
50	0.348	0.359	0.372	0.386	0.402	0.419
60	0.330	0.341	0.353	0.367	0.382	0.399
70	0.312	0.323	0.335	0.348	0.363	0.379
80	0.296	0.306	0.318	0.331	0.345	0.361
90	0.283	0.294	0.305	0.318	0.332	0.347
100	0.274	0.284	0.296	0.308	0.322	0.337
110	0.266	0.277	0.288	0.300	0.314	0.329
120	0.260	0.271	0.282	0.294	0.308	0.322
130	0.255	0.265	0.277	0.289	0.302	0.317
140	0.251	0.261	0.272	0.285	0.298	0.312
150	0.247	0.257	0.269	0.281	0.294	0.308

Appendix H: Comparison of Natural gas pressure, ambient temp and power Consumption (DMR Process)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure(bar)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)
20	53.099	54.444	56.193	58.042	59.982	62.194
30	47.523	48.963	50.566	52.339	54.295	56.466
40	44.229	45.642	47.214	48.953	50.866	52.974
50	41.754	43.132	44.671	46.364	48.233	50.298
60	39.496	40.843	42.336	43.986	45.799	47.801
70	37.259	38.566	40.018	41.614	43.374	45.297
80	35.241	36.514	37.924	39.476	41.165	43.030
90	33.671	34.920	36.298	37.809	39.460	41.274
100	32.471	33.704	35.063	36.550	38.171	39.948
110	31.511	32.734	34.082	35.551	37.151	38.909
120	30.677	31.934	33.272	34.728	36.310	38.038
130	30.042	31.251	32.584	34.028	35.595	37.305
140	29.461	30.666	31.987	33.420	34.974	36.666
150	28.952	30.151	31.463	32.882	34.420	36.103

Appendix I: Comparison of natural gas pressure, ambient temp and refrigerant flow (DMR Process)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure(bar)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)
20	444.86	455.23	466.01	478.47	489.74	502.76
30	433.72	443.97	454.70	466.02	478.05	490.99
40	420.24	430.35	440.92	451.99	463.89	476.57
50	404.30	414.29	424.57	435.34	447.16	459.61
60	385.78	395.47	405.67	416.31	427.62	439.75
70	364.60	374.09	383.91	394.24	405.36	416.95
80	344.44	353.61	363.16	373.15	383.82	395.22
90	328.83	337.83	347.18	357.07	367.45	378.59
100	317.27	326.20	335.46	345.16	355.42	366.40
110	308.22	317.11	326.30	335.92	346.08	357.10
120	298.82	309.67	318.81	328.37	338.46	349.23
130	294.56	303.26	312.45	321.95	331.98	342.69
140	289.11	297.86	306.92	316.38	326.35	337.00
150	284.35	293.04	302.05	311.43	320.84	331.98

Appendix J: The effect of natural gas supply pressure on Niche LNG process when it is supplied at temperature of 15⁰C

Natural Gas Pressure (bar)	C-1 (MW)	C-2 (MW)	C-3 (MW)	C-4 (MW)	EXP-1 (MW)	Exp-2 (MW)	Cycle-1 Refg flow (mmscfd)	Cycle-2 Refg flow (mmscfd)	LNG Mass Flow (kg/s)	Total Refg flow (mmscfd)	Total Power (MW)	Specific Power (kW-hr/kgLNG)	LNG (MTPA)	Production per train
30	50	16	13	11	15	4	541	229	34.99	770.05	72.29	0.574	1.00	0.56
40	44	17	13	11	14	4	564	227	34.91	790.92	67.38	0.536	1.00	0.60
50	41	17	13	11	14	4	572	225	34.83	796.92	64.50	0.514	0.99	0.63
60	39	17	12	11	14	4	570	226	34.75	796.71	62.59	0.500	0.99	0.64
70	37	17	12	11	13	4	564	222	34.67	786.23	60.74	0.487	0.99	0.66
80	35	17	12	11	13	4	560	221	34.58	781.49	59.13	0.475	0.99	0.68
90	34	17	12	11	13	4	553	220	34.50	772.77	57.80	0.465	0.98	0.69
100	34	16	12	11	12	4	542	218	34.42	759.87	57.05	0.460	0.98	0.70
110	32	16	12	11	12	4	535	217	34.33	752.07	55.76	0.451	0.98	0.71
120	31	16	12	11	12	4	527	217	34.25	744.00	54.73	0.444	0.98	0.73
130	31	16	12	11	12	4	518	216	34.16	733.92	53.78	0.437	0.97	0.74
140	30	15	12	11	11	4	509	216	34.08	724.70	52.98	0.432	0.97	0.75
150	29	15	12	11	11	4	501	215	33.99	715.87	52.25	0.427	0.97	0.75

Appendix K: Comparison of Natural gas pressure, temperature and specific power on (Niche LNG Process)

Temperature (C)	5	10	15	20	25	30
Natural gas Pressure (bar)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)
30		0.551	0.574	0.601	0.633	0.651
40	0.491	0.511	0.536	0.565	0.597	0.624
50	0.468	0.489	0.514	0.543	0.575	0.609
60	0.451	0.475	0.500	0.527	0.557	0.587
70	0.439	0.464	0.487	0.520	0.549	0.579
80	0.429	0.453	0.475	0.509	0.540	0.571
90	0.420	0.442	0.465	0.497	0.529	0.562
100	0.412	0.434	0.460	0.487	0.518	0.549
110	0.405	0.427	0.451	0.478	0.508	0.539
120	0.399	0.420	0.444	0.469	0.498	0.530
130	0.393	0.415	0.437	0.462	0.489	0.521
140	0.388	0.409	0.432	0.456	0.481	0.511
150	0.383	0.405	0.427	0.451	0.476	0.504

Appendix L: Comparison of Natural gas pressure, ambient temp and power Consumption (Niche LNG Process)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure (bar)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)
30		69.41	72.29	75.74	79.77	82.03
40	61.65	64.20	67.38	70.95	75.04	78.41
50	58.66	61.32	64.50	68.02	72.05	76.39
60	56.40	59.44	62.59	65.94	69.64	73.37
70	54.84	57.90	60.74	64.84	68.54	72.32
80	53.39	56.35	59.13	63.31	67.17	71.11
90	52.13	54.96	57.80	61.77	65.71	69.81
100	51.00	53.82	57.05	60.31	64.18	68.06
110	50.05	52.80	55.76	59.05	62.73	66.63
120	49.16	51.79	54.73	57.86	61.43	65.39
130	48.35	51.00	53.78	56.82	60.19	64.05
140	47.58	50.21	52.98	55.96	59.02	62.65
150	46.92	49.53	52.25	55.13	58.21	61.63

Appendix M: Comparison of natural gas pressure, ambient temp and refrigerant flow (Niche LNG Process)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure (bar)	Reg-Flow (mmscfd)	Reg-Flow (mmscfd)	Reg-Flow (mmscfd)	Reg-Flow (mmscfd)	Reg-Flow (mmscfd)	Reg-Flow (mmscfd)
30		747.48	770.05	994.17	803.00	1021.40
40	744.63	770.01	790.92	788.10	820.01	840.11
50	751.13	777.22	796.92	807.36	826.58	942.09
60	749.91	775.96	796.71	814.21	828.16	904.03
70	743.14	766.49	786.23	810.64	815.81	832.22
80	735.57	758.95	781.49	799.93	807.55	820.77
90	725.72	750.01	772.77	793.13	800.83	813.07
100	713.91	740.52	759.87	786.45	794.00	810.69
110	701.75	729.43	752.07	781.47	789.56	803.76
120	689.66	716.82	744.00	774.97	785.06	799.19
130	678.93	707.46	733.92	766.82	778.78	793.76
140	668.22	697.10	724.70	758.69	773.07	791.23
150	659.32	688.19	715.87	750.47	765.71	784.97

Appendix N: The effect of natural gas supply pressure on dual nitrogen expander process when it is supplied at temperature of 15^{0}C

Natural Gas Pressure (bar)	C-1 (MW)	C-2 (MW)	C-3 (MW)	C-4 (MW)	EXP-1 (MW)	EXP-2 (MW)	Cycle-1 Refg flow (mmscfd)	Cycle-2 Refg flow (mmscfd)	LNG Mass Flow (kg/s)	Total Refg (mmscfd)	Total Power (MW)	Specific Power (kW-hr/kgLNG)	LNG (MTPA)	Production per train
30	28.64	29.18	24.50	22.45	16.70	7.69	582.25	447.85	34.35	1030.10	80.37	0.650	0.98	0.50
40	34.07	33.40	16.13	14.15	19.35	4.92	666.46	282.37	34.27	948.83	73.49	0.596	0.98	0.54
50	26.12	36.45	15.86	13.78	18.98	4.81	727.16	274.97	34.19	1002.14	68.42	0.556	0.97	0.58
60	20.57	38.88	15.62	13.50	18.64	4.72	775.72	269.36	34.11	1045.08	65.21	0.531	0.97	0.61
70	17.27	40.14	15.42	13.26	18.29	4.64	800.96	264.66	34.03	1065.62	63.16	0.516	0.97	0.63
80	15.27	40.54	15.23	13.07	17.92	4.58	808.94	260.82	33.94	1069.75	61.61	0.504	0.97	0.64
90	13.90	40.54	15.06	12.91	17.57	4.53	808.85	257.52	33.86	1066.37	60.31	0.495	0.97	0.65
100	12.88	40.33	14.89	12.76	17.23	4.47	804.60	254.54	33.78	1059.13	59.16	0.486	0.96	0.66
110	12.10	39.90	14.74	12.64	16.87	4.43	796.14	252.10	33.69	1048.24	58.08	0.479	0.96	0.67
120	11.47	39.46	14.62	12.51	16.55	4.39	787.30	249.65	33.61	1036.95	57.12	0.472	0.96	0.68
130	10.98	38.89	14.51	12.39	16.22	4.35	775.96	247.30	33.52	1023.26	56.20	0.466	0.96	0.69
140	10.52	38.41	14.35	12.33	15.94	4.32	766.37	245.93	33.44	1012.30	55.35	0.460	0.95	0.70
150	10.18	37.79	14.24	12.24	15.63	4.29	753.89	244.25	33.35	998.13	54.53	0.454	0.95	0.71

Appendix O: Comparison of Natural gas pressure, temperature and specific power on (dual nitrogen expander Process)

Temperature (C)	5	10	15	20	25	30
Natural gas Pressure (bar)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)	Specific power (kW-hr/kg-LNG)
20						
30	0.595	0.622	0.650	0.674	0.706	0.744
40	0.550	0.572	0.596	0.617	0.644	0.675
50	0.508	0.531	0.556	0.581	0.611	0.643
60	0.482	0.506	0.531	0.559	0.590	0.623
70	0.466	0.490	0.516	0.545	0.576	0.609
80	0.455	0.479	0.504	0.534	0.565	0.598
90	0.446	0.470	0.495	0.524	0.556	0.589
100	0.438	0.462	0.486	0.516	0.547	0.580
110	0.430	0.454	0.479	0.509	0.540	0.573
120	0.424	0.448	0.472	0.502	0.533	0.566
130	0.418	0.441	0.466	0.495	0.526	0.559
140	0.412	0.436	0.460	0.490	0.521	0.553
150	0.406	0.430	0.454	0.484	0.515	0.548

Appendix P: Comparison of Natural gas pressure, ambient temp and power Consumption (dual nitrogen expander)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure(bar)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)	Power (MW)
30	73.58	76.88	80.37	83.33	87.35	91.98
40	67.84	70.53	73.49	76.07	79.40	83.26
50	62.57	65.38	68.42	71.53	75.16	79.19
60	59.17	62.11	65.21	68.63	72.40	76.50
70	57.11	60.07	63.16	66.71	70.53	74.62
80	55.63	58.56	61.61	65.20	69.03	73.10
90	54.36	57.29	60.31	63.92	67.73	71.78
100	53.23	56.16	59.16	62.77	66.57	70.56
110	52.22	55.11	58.08	61.69	65.47	69.47
120	51.28	54.15	57.12	60.73	64.48	68.43
130	50.39	53.25	56.20	59.80	63.53	67.47
140	49.59	52.44	55.35	58.93	62.67	66.59
150	48.79	51.60	54.53	58.07	61.82	65.74

Appendix R: Comparison of natural gas pressure, ambient temp and refrigerant flow (dual nitrogen expander)

Temperature (C)	5	10	15	20	25	30
Natural Gas Pressure(bar)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)	Refg-Flow (mmscfd)
30	991.41	1011.22	1030.10	1113.66	1185.28	1251.33
40	897.37	923.63	948.83	1014.38	1071.52	1123.88
50	950.91	978.51	1002.14	1068.43	1124.37	1174.54
60	999.41	1023.66	1045.08	1107.50	1161.56	1207.73
70	1021.72	1045.42	1065.62	1124.50	1176.72	1221.57
80	1025.10	1049.57	1069.75	1127.28	1179.29	1224.68
90	1017.30	1044.12	1066.37	1123.67	1175.42	1221.81
100	1005.37	1034.38	1059.13	1116.57	1168.81	1215.01
110	992.02	1022.11	1048.24	1105.68	1158.46	1207.36
120	977.25	1008.50	1036.95	1095.13	1147.85	1197.25
130	962.08	994.46	1023.26	1082.71	1136.27	1186.82
140	947.71	981.32	1012.30	1070.17	1124.95	1176.59
150	931.74	965.32	998.13	1055.45	1112.06	1165.39

Appendix S: Effect of natural gas composition on proposed liquefaction Processes for LNG FPSO

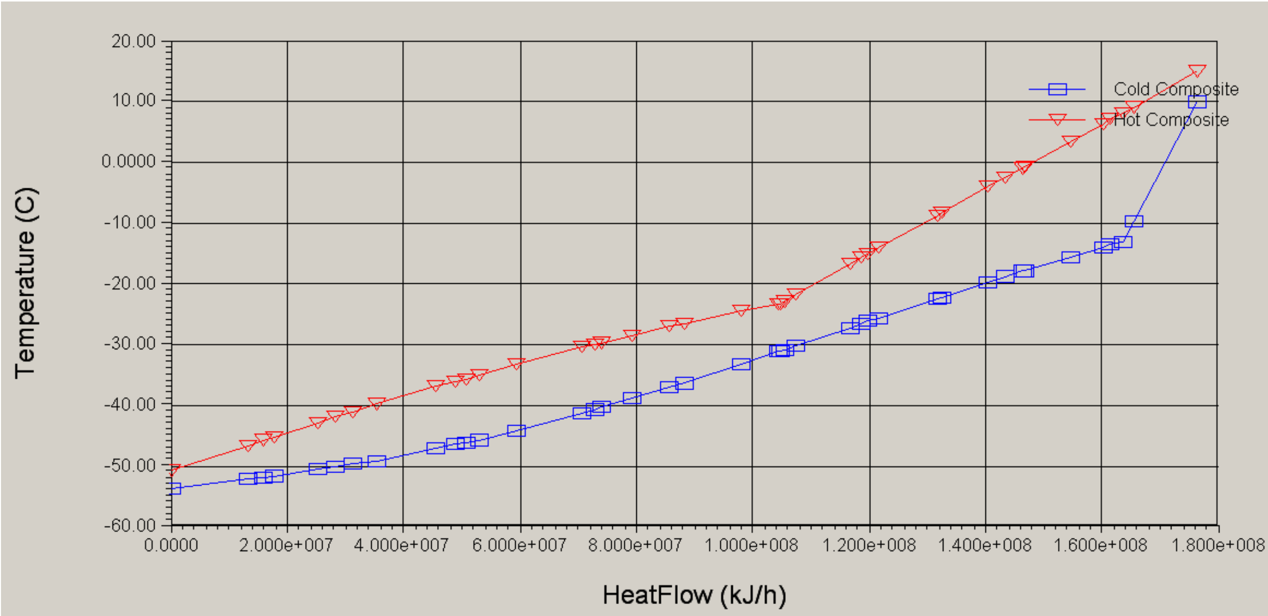
DMR			
	Power Consumption (MW)	**Refrigerant Flow Rate (MW)**	**Specfic Power (MW)**
Composition - Reference	42.33	405.42	0.35
Composition -A	42.43	406.15	0.35
Composition -B	42.11	400.68	0.32
Composition -C	42.07	402.87	0.35
Composition -D	42.21	402.47	0.33
Composition -E	41.55	392.72	0.30

N2 expander Process			
	Power consumption (MW)	Refrigerant Flow Rate (mmscfd	Specific Power (kW/kg-LNG)
Composition Reference	65.20	1043.91	0.53
Composition -A	65.33	1052.29	0.52
Composition -B	66.90	1128.68	0.49
Composition -C	64.93	1027.31	0.53
Composition -D	66.15	1097.22	0.51
Composition -E	68.30	1181.85	0.48

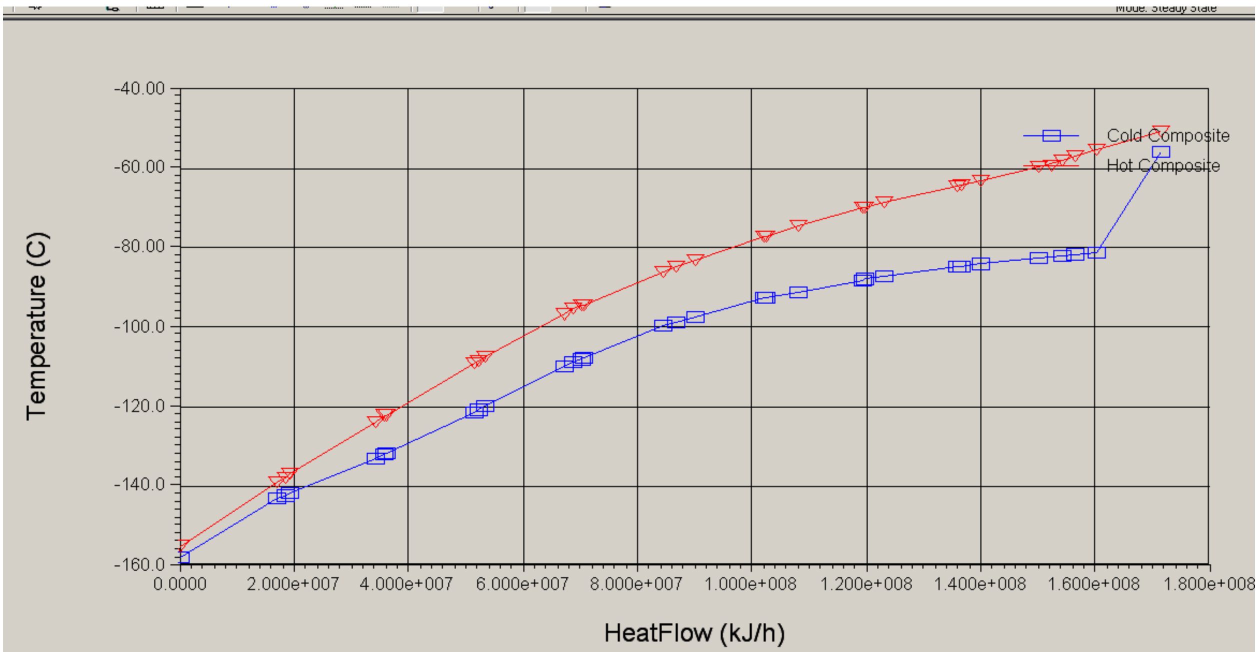
Niche Process			
	Power Consumption (MW)	Refrigerant Flow (mmscfd)	Specific Power (kW/kg-LNG)
Composition -Reference	62.07	794.50	0.50
Composition - A	62.39	796.41	0.49
Compostion - B	64.68	831.65	0.47
Composition - C	61.58	785.67	0.49
Compostion- D	63.70	816.94	0.48
Compostion -E	66.52	855.17	0.46

SMR Process			
	Power Consumption (MW)	Refrigerant (mmscfd)	Specific Power (kW/kg-LNG)
Compostion - Reference	46.42	530.51	0.39
Composition - A	46.68	534.30	0.38
Composition - B	48.56	560.17	0.37
Composition - C	45.95	525.09	0.39
Composition - D	47.80	549.34	0.38
Composition - E	49.85	579.72	0.36

Appendix T: Composite curve for DMR process

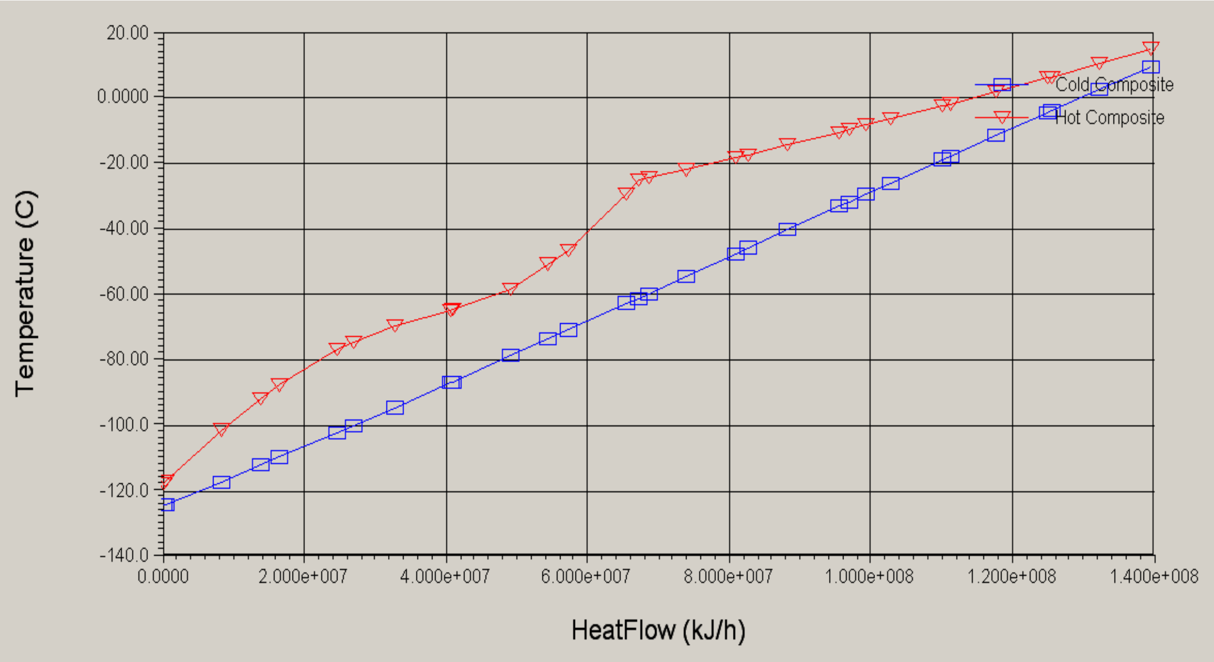


Composite curve for LNG heat exchanger LNG-1 (DMR process)

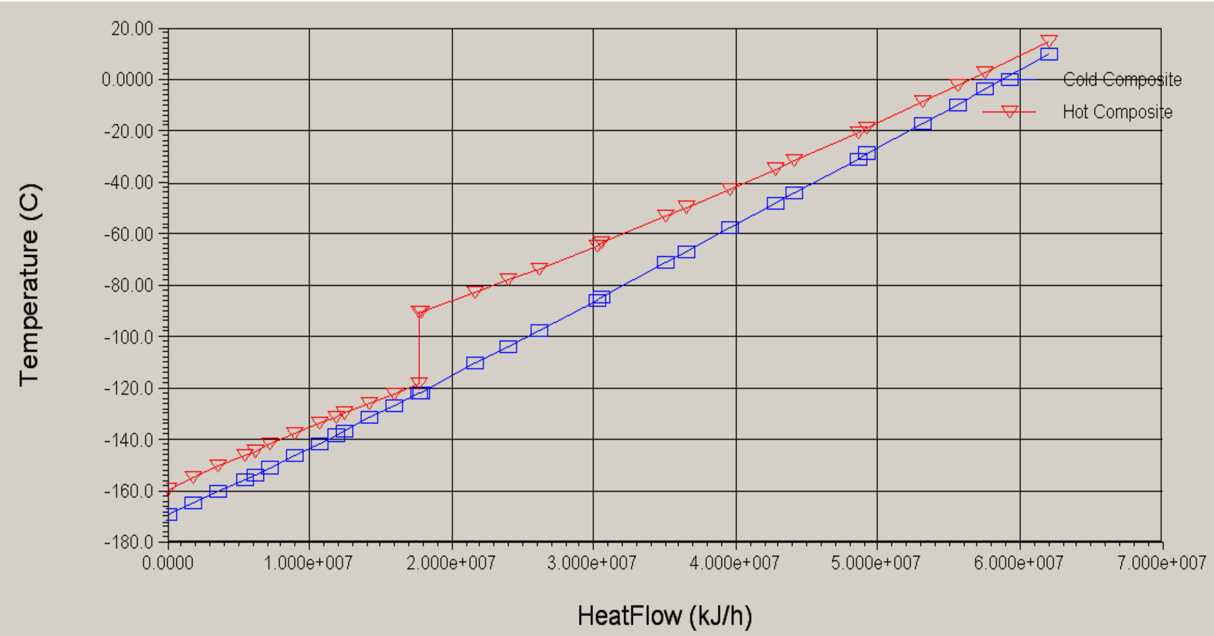


Composite curve for LNG heat exchanger LNG-2 (DMR process)

Appendix U: Composite curve for Niche LNG process

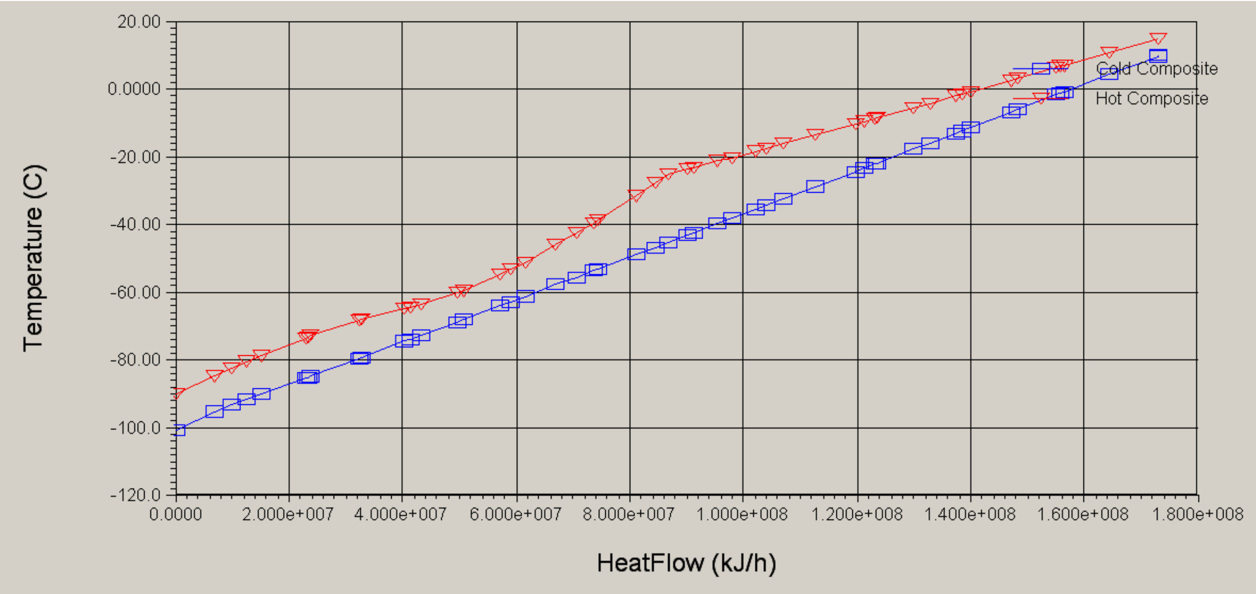


Composite curve for LNG heat exchanger LNG-1 (Niche LNG process)

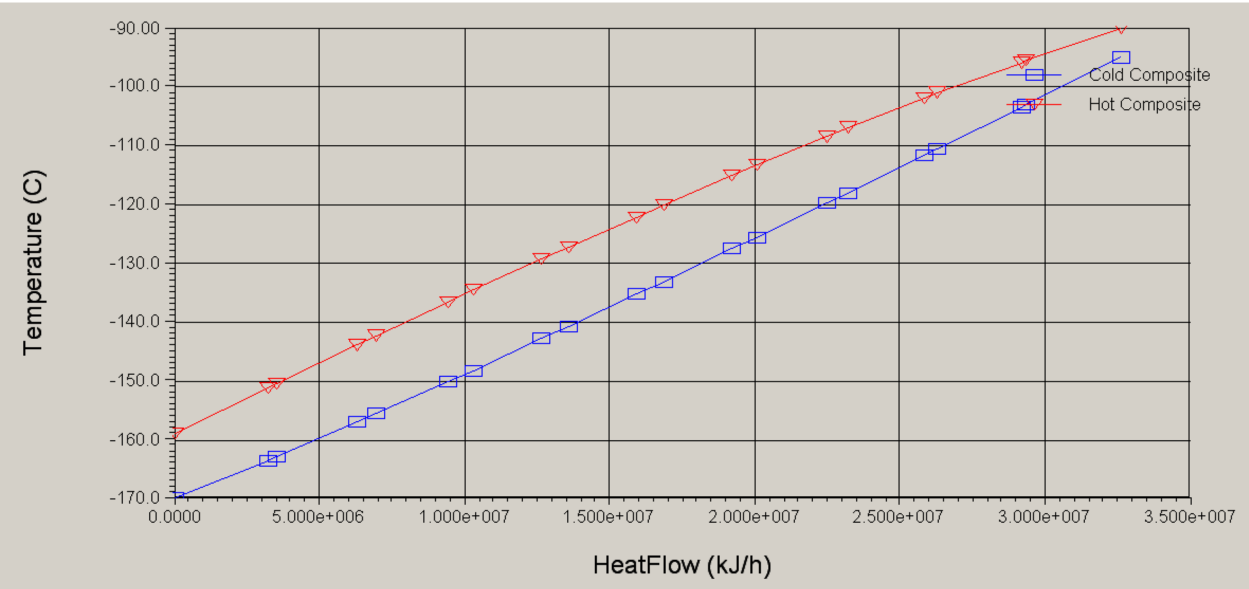


Composite curve for LNG heat exchanger LNG-1 (Niche LNG process)

Appendix V: Composite curve for dual nitrogen expander process



Composite curve for LNG heat exchanger LNG-1 (dual nitrogen expander process)



Composite curve for LNG heat exchanger LNG-2 (dual nitrogen expander process)

Appendix W: Composite curve for SMR process