Evaluation of LNG Technologies

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Abstract

Due to the increasing demand for natural gas in the world today, transportation of natural gas from different parts of the world has become a necessity. Liquefying the natural gas provides a safer and cheaper alternative for transportation and also increases its storage capabilities. The liquefaction process requires the natural gas to be cooled using various methods of cryogenic processes and also be depressurized to atmospheric conditions for easier and safer storage. This paper analyzes eleven total methods for the liquefaction of natural gas based on their capacity, fixed costs and their operating efficiency. Seven out of the eleven processes are currently established in various parts of the world. The remaining four processes are in developmental stages and each patent description is used to simulate each process and determine its feasibility in industry. At the reported maximum capacities of each process, the equipment and utility costs are determined. These prices per capacity are then related along a possible range of annual capacities to determine the most economical process based on specific capacities.

Contents

Abstract - 1 -

Introduction - 4 -

Natural Gas - 4 -

Liquefied Natural Gas - 5 -

Refrigeration Cycles - 6 -

TQ Diagrams - 7 -

Plate Fin Heat Exchangers - 8 -

Spiral Wound Heat Exchangers - 10 -

Simulation Method - 12 -

- LNG Heat Exchanger Area - 13 -
  - Plate fin Heat exchanger - 14 -
  - Spiral Wound Heat exchanger - 14 -

LNG Liquefaction Techniques - 15 -

- Processes Removed From Study - 16 -
- Description of Processes and Specific Simulation Methods - 17 -
  - Black and Veatch’s Prico Process - 17 -
  - Dual Mixed Refrigerants - 20 -
  - APCI C3-MR - 25 -
  - APCI AP-X - 30 -
  - Axens Liquefin Process - 34 -
  - ExxonMobil Dual Multi-component process (US patent no.: 6,250,105) - 39 -
  - Technip-TEALARC Process - 44 -
  - BP Self Refrigerated Process (US pat number: 6,564,578) - 51 -
  - Mustang Engineering Smart Liquefaction Process (Patent Number 7225636) - 55 -
  - ConocoPhilips Simple Cascade - 59 -
  - Mixed Fluid Cascade Process - 64 -
  - CO_2 Pre-cooled Linde Process - 71 -

Economic Analysis - 76 -

Recommendations .... - 78 -

References .... - 80 -

Appendices .... - 81 -

  Patent Search .... - 81 -
  Results of Patent Searches .... - 82 -
  Mustang Smart Liquefaction Stream Specifications and PFD from Patent **Error! Bookmark not defined.**

Introduction

The design of liquefaction plants is an important part of the natural gas industry. There are several methods available depending on the capacity of natural gas being produced and the amount of equipment desired. Several new processes for gas liquefaction have been developed but are not in use, even though they might be more efficient, require less equipment, and cost less to run. The purpose of this project is to simulate several processes and evaluate them based on cost of equipment, operating costs, and capacity.

Natural Gas

Natural gas is what is used to describe a gas composed of a mixture of different hydrocarbons, compounds made of carbon and hydrogen, being pulled from the ground. It contains different proportions of methane (CH_4), ethane (C_2H_6), propane (C_3H_8), and butane (C_4H_{10}). When the gas is initially extracted from the ground at the well sites, it also contains impurities in the form of hydrogen sulfide, carbon dioxide, and water, along with trace elements such as halogen gases. The normal compositions of these compounds in natural gas are listed below:

Table 1: Natural Gas Composition

Component	Composition
Methane	70-90%
Ethane	0-20%
Propane	
Butane	
Carbon Dioxide	0-8%
Oxygen	0-0.2%
Nitrogen	0-5%
Hydrogen Sulfide	0-5%
Rare gases	trace

Source: http://www.naturalgas.org/overview/background.asp

There are two forms of natural gas: wet and dry. Wet natural gas is where the methane composition is on the low side of the estimation in the table above, between 70% and 80%. The other compounds are present in their higher amounts. When methane is present in very high percentages, above 80%, with less percentages of the higher hydrocarbons present, it is considered dry.

After the gas has been removed from the ground it is transported to a processing facility. In the processing plant, there are four treatments that the gas goes through before it is ready to be transported for use: oil and solids removal, dehydration, separation of natural gas liquids (NGL), and scrubbing to remove hydrogen sulfide and carbon dioxide.

When underground, the pressure causes the natural gas to dissolve in oil. When it is released from the ground, the natural gas can separate from the oil as the pressure is decreased depending on the composition and pressure of the formation from which the gas is removed. Under the most basic removal conditions a simple tank with streams exiting from both the top and bottom is required, to allow gravity to naturally separate the liquid from the gas. Under other conditions a separator may be needed.

Methods for dehydrating the natural gas include glycol dehydration, in which a liquid desiccant with an affinity for water is used to absorb water vapor from the gas[1], and solid desiccant dehydration, the primary form, which uses two adsorption towers filled with solids[1].

Natural gas liquids are removed either by absorption, by a cryogenic expansion process, or by fractionation which separates each liquid component separately. Sweetening the gas is a term used to describe the process of removing hydrogen sulfide. To sweeten the gas, amine solutions are used to absorb and remove hydrogen sulfide. NGL and sulfur removed from the process are each sold after removal, as each is profitable on its own. Once the gas has gone through each of these processes, it is scrubbed to remove other large impurities such as sand. All of these processes take place at a plant near the well site.

Liquefied Natural Gas

Once natural gas has been cleaned and scrubbed, it is transported to its final destination. Depending on where the gas is being transported to, it needs to be liquefied, or cooled to a liquid. The liquefaction of natural gas involves cooling the natural gas to -260 F, at which point the gas has become a liquid. After liquefaction, the volume of the gas is reduced by a factor of 600.

Reducing the volume allows for more of the gas to be transported with less equipment, especially when being transported overseas.

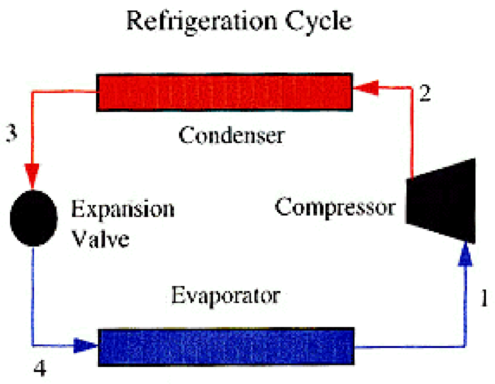
There are hundreds of processes for liquefying natural gas, but very few are actually in use. Most of the processes in use are not patented, with mostly new processes being patented. The most used processes for liquefaction are processes developed by Air Products and Chemical Inc., ConocoPhillips, and Linde, each of which has a capacity between two and eight MTPA. Depending on the process, different refrigerants are used to cool the natural gas but with the use of similar equipment. The factor that differs between processes is the setup and design. The primary equipment used are plate fin heat exchangers, spiral wound heat exchangers, shell and tube heat exchangers, compressors, expanders, and valves. Most of the work required to cool the natural gas takes place in the plate fin or spiral wound heat exchangers.

Newer processes tend to reduce the amount of equipment used in the cycle and have an increased capacity of five to nine MTPA. The lower costs of implementation and increased capacity make the new processes very viable in industry. Unfortunately many of the older processes have been in use for such a long time, that the cost of upkeep and initial cost have created a lull in the need for new liquefaction techniques. Companies do not want to have to tear down old equipment and pay for the installation of new equipment when they have a process that is based on proven technology.

Refrigeration Cycles

A basic refrigeration cycle consists of two heat exchangers, a valve, and a compressor. The refrigerant flows through the evaporator where it is heated. The evaporator represents the cooling that a gas or liquid would receive from the refrigerant. From the evaporator, the refrigerant flows through a compressor to get the stream back to the design pressure. It also converts the stream from two phases to one phase. After the evaporator the refrigerant might be at or past its boiling point. After the compressor, the refrigerant flows through a condenser to get it to its bubble point. The refrigerant then flows through an expansion valve, after which it is cool enough to absorb the heat that is transferred in the evaporator. A diagram of a simple refrigeration cycle (Figure 1[2]) is shown below.

Figure 1: Refrigeration Cycle[2]



Using plate fin or spiral wound heat exchangers, liquefaction plants use variations of this method to cool any refrigerants needed and then to cool the natural gas to the required temperature.

TQ Diagrams

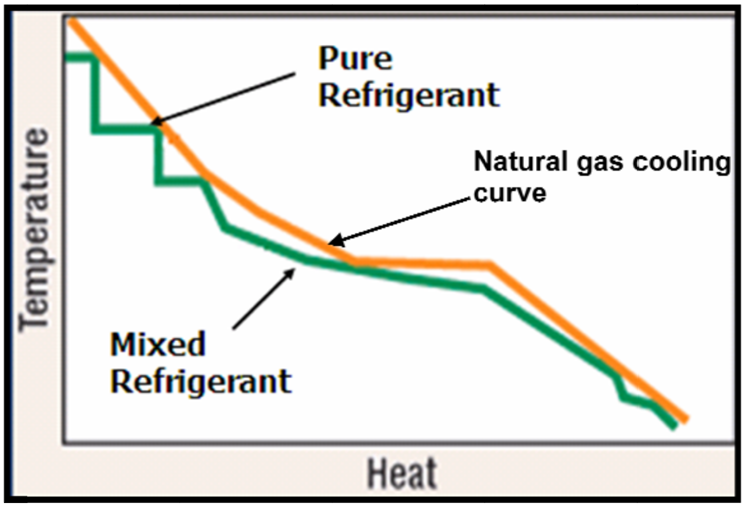


Figure 2: Natural gas refrigerant cooling curve[3]

Above is an example of what a typical temperature-heat diagram or cooling curve, for the cooling of natural gas using both pure and mixed refrigerants would look like. During cooling, it is desired to have an efficient process. One method of determining the efficiency of a cycle is to review the cooling curve. The closer the line depicting the refrigerants is to the curve of the natural gas, the more efficient is the cycle. Increasing the efficiency of the process reduces the amount of work done by the heat exchangers. The amount of work done by the heat exchangers is indicated by the spaces between the curves.

When using pure refrigerants, such as propane or nitrogen, the curve of the refrigerant is typically a stair step curve against the natural gas as indicated by the beginning of the diagram. A mixed refrigerant curve is typically smoother, allowing it to come closer to the curve of the natural gas, as shown at the end of the diagram. This implies that the use of a mixed refrigerant, typically including methane, ethane, propane, and butane, is a better choice for refrigerants. The composition of the mixed refrigerant is also a factor in how close the refrigerant curve is to the natural gas curve.

Plate Fin Heat Exchangers

These heat exchangers are widely used in cryogenic application because of their low cost, small size, low weight, high thermal capacity and effectiveness relative to other types of heat exchangers. The result of the improved effectiveness is the achievement of true counter current flow where there is an increase in the temperature spread and a closer approach to ideality. This means that the refrigerant cooling curve is closer to the natural gas cooling curve.

The exchanger is made up of manifolds or headers which consist of elements. A manifold and an element are shown below, Figures 4 and 5 respectively. An element is made up of a corrugated die-formed fin plate placed between flat metal separator plates. There are side bars along the outside of the fin sections. A stack of the elements is welded to form a rigid matrix and can be designed to meet any configuration and size. The stacks are welded onto the manifolds. Depending on the application, a number of manifolds can be assembled to form the heat exchanger.

Figure 3: Manifold or Header (xchanger.com)

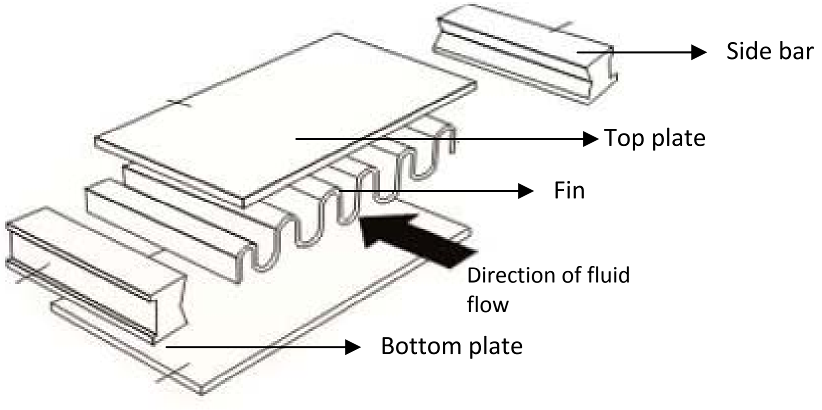


Figure 4: An element

The wavy configuration of the fin promotes turbulence and therefore improves heat transfer. This increase in heat transfer is accompanied by an increase in pressure drop. This is a problem with low density fluids like gases because of the extra work required to surmount the pressure drop.

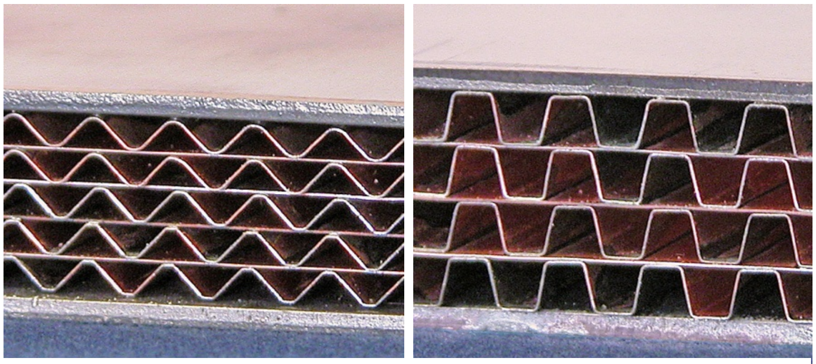


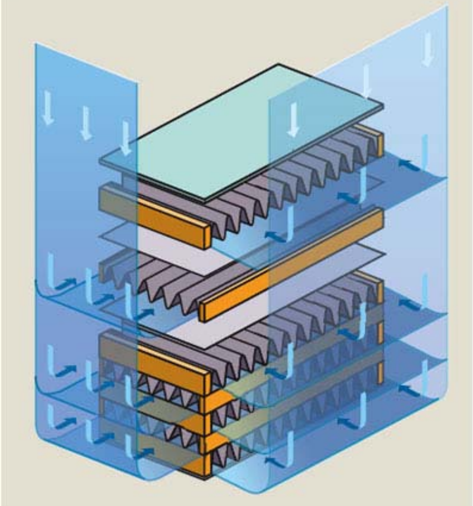
Figure 5: Fin Configurations

This work is often much higher than the increase in heat transfer acquired from the fins. For applications where it is proposed any fin configuration other than the simplest, a thorough analysis of the effect on the system should be conducted.

In designing the plate fin heat exchanger, it is possible to have different heights of the alternating fin plates. There is no requirement to have the same height or spacing of separator plates. This is a useful freedom to have in situations where the difference in density of the hot and cold fluids is large. In cryogenic systems, the refrigerant stream entering the expander has a higher density than the stream coming out of the expander. In such a case as this, it is necessary to use a larger height for the lower density stream so that a common Reynolds number and therefore heat transfer coefficient, U, can be attained.

Aluminum is the usual material of construction of plate fin heat exchangers for cryogenic applications. (Walker, 104-110)

Figure 6: Element Geometry (Cross-fl)



Spiral Wound Heat Exchangers

These heat exchangers have a broad temperature and pressure range and are used in single phase and two phase applications. They are able to handle many streams depending on the customers need. Pressures of up to 3625 psi can be handled sufficiently. The materials of construction of this exchanger include austenitic steels, aluminum alloys, carbon steel, nickel and chrome/moly alloys. This wide range of materials and the special features associated with the geometry of the heat exchanger allow for a wide range of applications.

The surface area of heat transfer can be as high as 20,000 m^2 but the maximum diameter is limited to 7,500 mm and a weight of approximately 260 tons per unit. It is possible to have larger diameters or higher weights but their feasibility has to be checked depending on the customer’s needs.

The material of construction chosen depends on the customer’s needs. Simulations are used for strength calculations of pressure containing vessels. Features such as tube spacing, tube support, bundle support and flexible mandrel are selected based on proven design technology developed over many years.

The exchanger consists of a center hollow tube with tube bundles that spiral round it. The center tube and the tube bundle are housed in a metal column. The outer column does not have a uniform diameter. The bottom part has a larger diameter than the top part. Figures A and B show the outer shell and the tube bundles respectively.

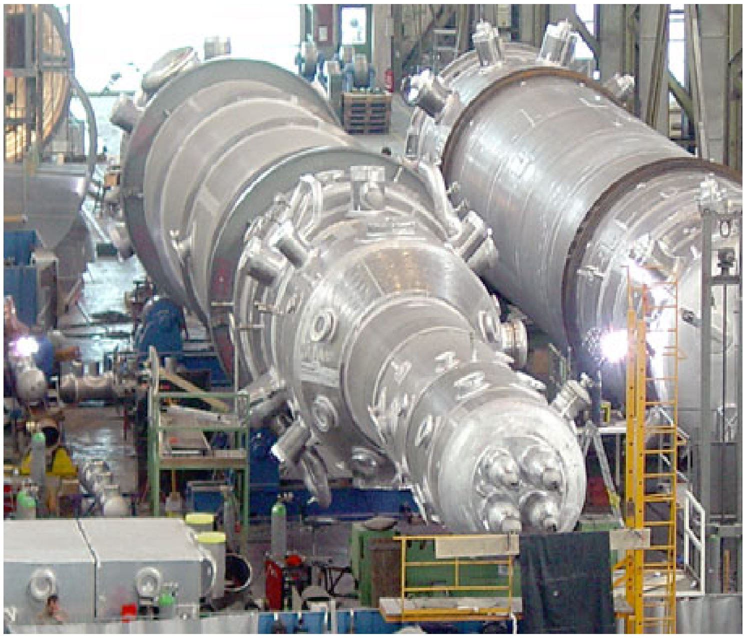


Figure 7

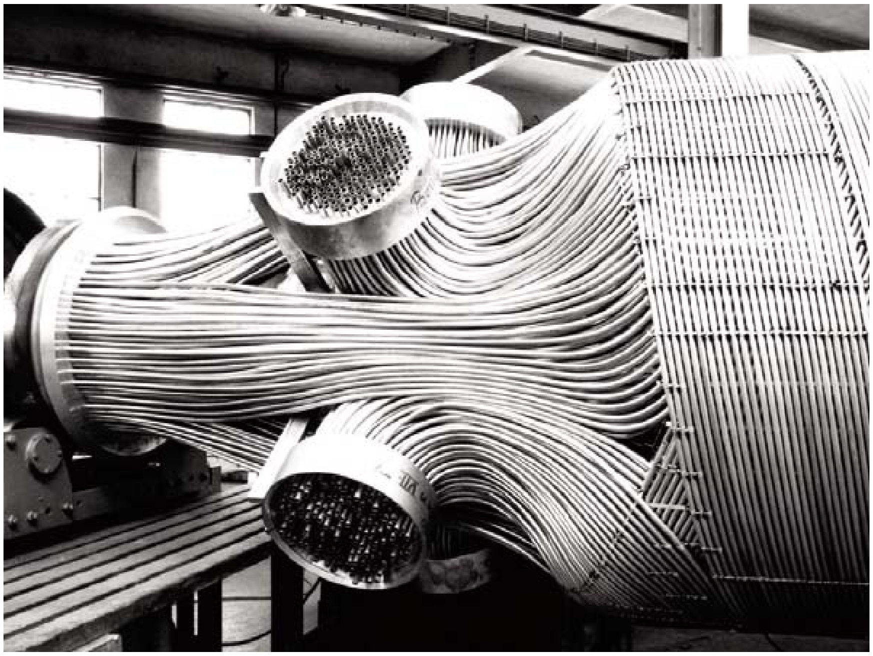


Figure 8

Simulation Method

SIMSCI PRO II is used for simulation. All processes were simulated in a similar fashion with some deviations where necessary in some of the processes. The processes are first translated into simpler simulations using simple heat exchangers. Using the known process specifications, namely temperature and pressure of LNG after each refrigeration stage, the simple simulations are made to mimic the real more complicated processes. The next step is to transfer all calculated process specifications to the real process simulation. The specifications transferred include refrigerant temperature and pressure before and after refrigeration stages, compressor outlet pressures, simple heat exchanger duties, utility flow rates, flash tank product fractions, etc. Some of the specifications are available for the patented processes. They are used as first approximations, after which they are further optimized. It is assumed that the data provided in the patents are products of optimization.

The refrigerant composition for each stage of refrigeration and the compressor work are then optimized. This can be done in the simple translation of the real process or it can be done in the real process after specifications found in the simple translation have been transferred. The refrigerant composition is optimized by finding the composition that produces the lowest temperature. The compressor work is optimized by changing the refrigerant flow rate. There are two restrictions to consider. These include the second law of thermodynamics and the surface area of heat transfer of the LNG heat exchangers. PRO II does not take the second law into consideration therefore it allows heat to be transferred from cold streams to hot streams. The temperatures of the streams are checked and adjusted to make sure this is not happening. The surface area of heat transfer in each cell of the LNG heat exchangers all have to be equal otherwise the unit cannot be built. PRO II does not take this into consideration so adjustments have to be made to the simulation. This is done by setting the difference between the areas of the cells of the LNG heat exchangers to zero by changing the outlet temperature of the cells. At this point, adjusting specifications of the simulation is done. The final step is to determine the utilities required by the simple heat exchangers. At this point, the data from the simulation is ready to be collected for economic calculations.

At the end of the simulation procedure, the point attained is not necessarily the global minimum; it could be a local minimum. Further optimization could be done to reach a global minimum but, due to time constraints, the point reached is accepted as the best possible situation.

LNG Heat Exchanger Area

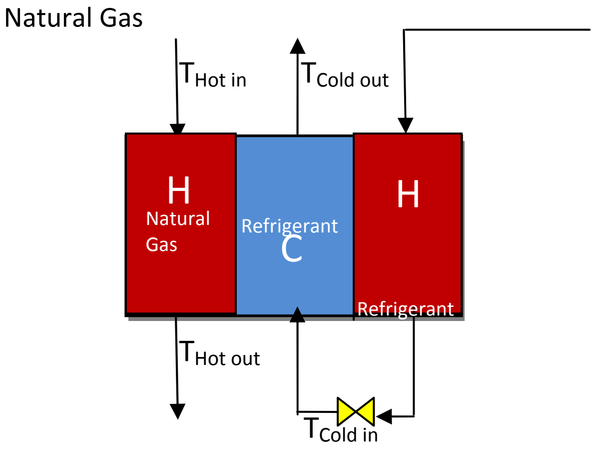


Figure 9

To use the LNG heat exchanger in PRO II, the outlet temperature of all the cells except one is set. PROII calculates the outlet temperature of that one cell based on the inlet temperatures, inlet pressures, flow rates and phases of all the cells. In the application of this project, the outlet temperatures of the natural gas cell and the hot refrigerant cells are set while the outlet temperature of the cold refrigerant cell is left to be calculated by PROII.

$$Heat\ Transfer\ Area\ (ft^2) = \frac{Q\ (Btu/hr)}{U\ (Btu/°Fhr) * \Delta T_{lm}(°F)} \qquad (1)$$

For Countercurrent flow:

$$\Delta\ T_{lm}(°F) = \frac{(T_{Hot\ in} - T_{Cold\ out}) - (T_{Hot\ out} - T_{Cold\ in})}{ln\frac{(T_{Hot\ in} - T_{Cold\ out})}{(T_{Hot\ out} - T_{Cold\ in})}} \qquad (2)$$

U is assumed to be 700 for Plate fin heat exchanger (PFHE)

U is assumed to be 900 for Spiral wound heat exchanger (SWHE)

These values for U were chosen based on literature. Literature reports that plate fin heat exchangers can achieve high heat transfer coefficients. 700 was a reasonable number to pick based on the heat transfer area values calculated in the simulations. Literature reports that the SWHE has an unlimited operating range in terms of temperature and is more effective overall in heat transfer in comparison to the plate fin heat exchanger. The SWHE U value, 900, was chosen based on the previous point.

Plate fin Heat exchanger

The area of each cell in the PFHE has to be the same based on the geometry. PROII does not do this automatically. It is done by calculating the area of each cell and using a controller to set the difference between the areas to zero by changing the outlet temperature of the cells, except the natural gas cell and the cold refrigerant cell (as mentioned earlier this is calculated by PROII). The heat transfer area is calculated using Equation 1 and the PFHE is priced directly using these areas. It is priced as a welded plate heat exchanger using McGraw Hill pricing website.

Spiral Wound Heat exchanger

The cells in the SWHE do not need to have the same area, based on geometry. As such, there is no limit on the temperatures the SWHE can handle. The heat transfer area is calculated with the same equation used for the PFHE but the SWHE is sized differently because of its complicated geometry. In the figure below, the cylinder with diameter D represents the central hollow tube and the wrapped cylinder with diameter d represents the tube bundle that spirals round the hollow tube. Although the tube bundles comprise many small tubes, for the sake of sizing, we assume that the tube bundle is one large tube with diameter d.

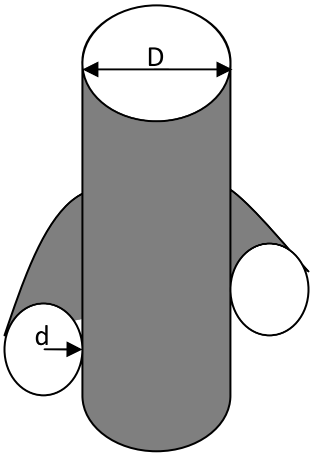


Figure 10: SWHE geometry

$$Heat\ Transfer\ Area\ (ft^2) = \frac{Q\ (Btu/hr)}{U\ (Btu/\ ^{o}F ft^2 hr) * \Delta T_{lm}(\ ^{o}F)} = N * \pi D(ft) * \pi d(ft) \tag{3}$$

$$Height\ of\ column(ft) = N * d \tag{4}$$

$$Diameter\ of\ column(ft) = d + \frac{D}{2} \tag{5}$$

D is assumed to be 3ft based on inspection of images of the SWHE. The diameter of the tube bundles, d, is calculated using equation (3). N is the number of times the tube bundles coil round the central tube full circle. In equation (3), the heat transfer area is equated to the surface area of the tube bundle with diameter, d. The term πD is the circumference of the central tube and is taken to be the length of the tube bundle. The length of the tube bundle is not uniform and so there is error in the calculation of the diameter, d. As there was no documented method for sizing the SWHE found, the method used in this project can be considered a rough estimation. The height and diameter of the column are calculated using equation (4) and equation (5) respectively. There was no direct way found to price the SWHE from diameter and height so it was priced as a storage tank and multiplied by 100 because the price as a storage tank was much lower than what was expected. It is known that the price of the SWHE is somewhere in the millions but the storage tank price ranges in the ten thousands.

LNG Liquefaction Techniques

In determining the processes to consider in this study, searches were conducted based on established processes currently in use by industry. This was followed by a patent search that revealed several more choices, including a few already in our repertoire, and several of which were modifications and improvements of the already established processes.

The list of liquefaction processes were then categorized by the type of refrigerant the processes used (if any): all mixed refrigerants, all pure refrigerants, both mixed and pure refrigerants, or other. The last category was compiled based on processes using self-refrigeration, turbo-expanders, and other methods, as shown in Table 1 below.

Table 2: Processes Under Consideration

Mixed Refrigerants	**Pure Refrigerants**	**Both**	**Other**
Linde- MFCP	CoP Simple Cascade	APCI C3 MR	*BP Self refrigerated process*
Axens Liquefin Process	*CoP Enhanced Cascade*	APCI AP-X	***ABB Randall Turbo-Expander***
Dual Mixed Refrigerant		*CO_2 MFCP*	***Williams Field Services co.***
Technip-TEALARC			*Mustang Group Smart Liquefaction*
ExxonMobil Dual Multi-component			
Black and Veatch Prico Process			
Technip- Snamprogetti			

* *Italicized processes signify Patent searched processes.*
* **Bolded processes signify processes not included in scope of project.**

A total of sixteen processes were chosen to be reviewed for the possibility of simulation. After reviewing all the processes, twelve processes were retained because of their currently established use in industry or there was sufficient information about the process to help in its simulations and determination of its optimum operating conditions.

Processes Removed From Study

The Technip-Snamprogetti process was removed due to its similarity to the TEALARC process. It is nearly identical and thus not essential to the project. There was also a lack of information about this process specifically

The ConocoPhillips Enhanced Cascade process was removed because of the use of the ConocoPhillips Simple Cascade method. The Enhanced Cascade is a modification of the Simple Cascade process that is established in industry. The modifications are designed to make the process more efficient. Its basic principles are the same and it is considered a more beneficial approach to optimize the "tried and true" original process rather than to optimize the enhanced cascade to compare it to other tried and true processes.

The Williams Field Service Company process was not chosen because, when looking over the patent, the description was a process for liquefying methane rich, or dry, natural gas with the purpose of using the end product as a fuel in cars. The process also claimed to use natural gas residue from a cryogenic plant as the feed stock for the process. With no knowledge of the composition of the residue gas to be fed into the process, there is no acceptable way of simulating the process.

The ABB Randall Corporation process was also disqualified from the project on the basis that information on the composition of its refrigerants was not found. With the refrigerants not known for the mechanical heat exchange process the simulation could not be completed.

Description of Processes and Specific Simulation Methods

The processes studied in this project are described below.

Black and Veatch's Prico Process

Description

This process is considered as one of the simplest and most basic processes currently in operation in the industry. It is considered a very basic setup with one large heat exchanger network and a single mixed refrigerant refrigeration cycle. Though the simple setup limits the capacity 1.2 MTPA per train, it reduces capital costs significantly.

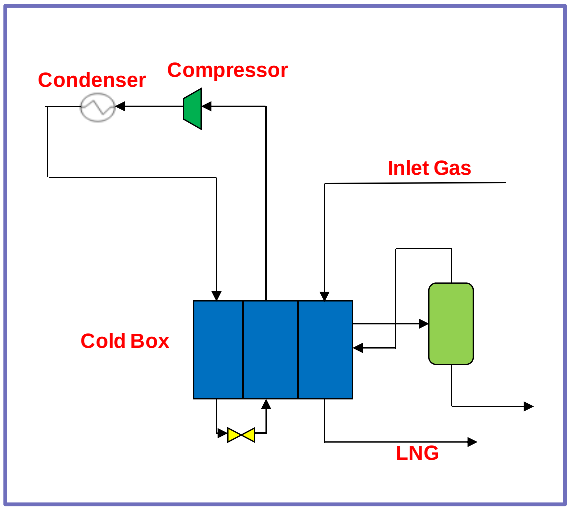


Figure 11: Prico Process

The mixed refrigerant used in this process is methane, ethane, propane, pentane and nitrogen. The composition of this mixed refrigerant can be chosen to match its boiling curve with the cooling curve of the natural gas. Due to the small train capacity that can be handled at time by this process, the curves are relatively easy to match thereby increasing its efficiency.

The 'cold box' in the setup is a collection of highly efficient plate fin heat exchangers that help in the heat exchange process between the compressed refrigerant and the raw natural gas. This heat exchange enables the cooling of natural gas to about a 260 F through forced convection due to the turbulent nature of its flow. 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.

In the cost analysis of the process (in the appendix), the largest cost comes from the cooling compression in the refrigeration cycle. The T-Q diagram below shows a supposedly efficient procedure, however the cost of compression greatly outweighs all other costs. This makes this process very fragile and the refrigerant composition very essential to its successful implementation. There is a significant reduction in capital and maintenance cost due to its simple setup.

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.

Simulation Method

Using a simple model of a single compression and a single refrigeration cycle, the Prico process has a very basic configuration. The input stream parameters of the natural gas are set to the standard temperature, pressure and composition.

The input refrigerant stream, REFRIGERANT, is all liquid at its bubble point temperature of about 270F and a pressure of about 800psia. This stream is mixed refrigerant composed of mostly hydrocarbons. The composition used in the simulations shown in Table 3 below.

Table 3: Refrigerant Composition

Component	Composition (%)
Methane	13.07
Ethane	11.81
Propane	10.09
I-Pentane	21.43
Nitrogen	43.60

The refrigerant stream enters the plate-fin Heat exchanger at its coldest temperature and exits at about -135 F at stream S2. The exit stream, S2, is now both in liquid and vapor phase. The refrigerant stream is then sent into an expansion valve where its pressure is reduced by 100psia. This expansion is meant to further re-cool the refrigerant before being sent back into the heat exchanger to cool the natural gas stream to our desired temperature. That pressure drop in the valve is appropriate because it prevents temperature crossing and reduces the area under the curve in T-Q diagram below.

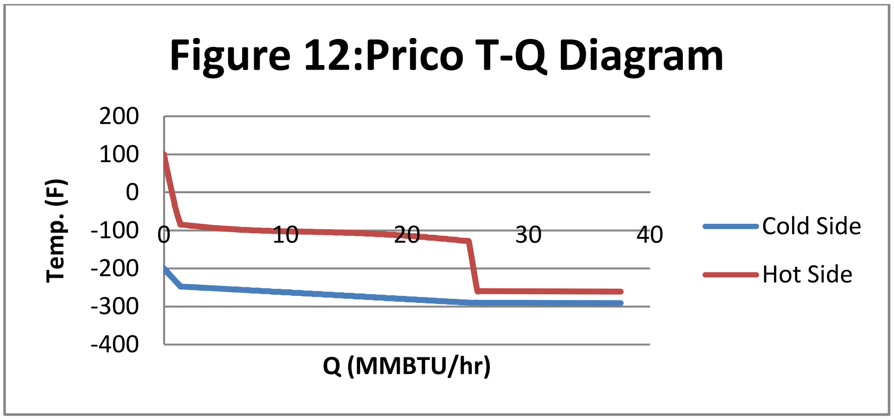


Figure 12:Prico T-Q Diagram

After the final cooling of the natural gas by the mixed refrigerant to about -150F, the refrigerants phase split is almost equal between the vapor and the liquid phases. This mixture is now recompressed back to the original pressure of the input refrigerant stream of 800 psia which was lost during the isenthalpic expansion by the valve, V1.

A significantly hotter mixed refrigerant stream is obtained after recompression and it now has to be re-cooled by a separate refrigeration (propane) cycle before being sent back into the heat exchanger. The separate refrigerant cycle is simulated as well to obtain the amount of propane and nitrogen needed to accomplish the re-cooling.

To obtain the final specification needed for the LNG, the expander reduces the pressure from a high 750 psia to atmospheric conditions. It provides the final needed cooling to obtain LNG at about -257F.

In calculating the area of heat transfer required, a controller and calculator were used to set the appropriate temperatures to equate the areas of the cells in the plate fin heat exchanger. An optimizer was also used to obtain the minimal work needed to produce the LNG by the compressor and the expander. The optimizer reduces the flow rate of the refrigerant while still satisfying other constraints.

Dual Mixed Refrigerants

Description

This process is similar to the APCI C3-MR process. It consists of 2 refrigeration cycles. The refrigerant used in the first cycle is a mixture of ethane and propane. The refrigerant used in the second cycle is a mixture of nitrogen, methane, ethane, propane and butane. The capacity of this process is reported at about 4.5 MTPA.

Figure 13: PFD of DMR

Simulation Method

For this process, the initial refrigerant composition used was obtained from literature. The compositions used are shown in the table below along with the first guess of total flow rate.

Table 4:Refrigerant Composition and Flow Rate		
	Stage 1 (%)	Stage 2 (%)
Methane	1	35
Ethane	47	41
Propane	6	1
n-Butane	30	■
i-Butane	16	■
Nitrogen	■	14
Total flow rate (lbmol/hr)	24900	20000

These parameters were entered into the simulation and temperature crossing was observed in both the first and second stage. In both stages, the outlet temperatures of the hot cells were lower than the inlet temperature of the cold cell. This means that heat was being transferred from the cold cell to the hot cells. For the first stage, the flow rate was first adjusted in an attempt to remove the temperature crossing but this method proved to be ineffective. The flow rate was returned to the initial value and the same procedure was carried out in the second stage. This was also ineffective. The next adjustment made was to the refrigerant composition. To reduce the inlet temperature of the refrigerant, the amount of heavier components (both butanes), was reduced and the amount of methane was increased by the balance. This method proved to be effective in both stages. The final compositions are shown in table 5.

Table 5:Refrigerant Composition and Flow Rate

	Stage 1 (%)	Stage 2 (%)
Methane	21	49
Ethane	47	41
Propane	6	5
n-Butane	25	[redacted]
i-Butane	1	5

At this point the simulation was working in that the natural gas target temperatures were being achieved without any temperature crossing. The compressor works before optimization are shown in table 6. The T-Q diagrams at this point are shown in Figures 14 and 15.

Table 6:Compressor Specifications

	Compressor Work (HP)	Outlet Pressure (Psia)
C1 (Stage 1)	23149	500
C2 (Stage 2)	28294	495
C3 (Stage 2)	5102	1000

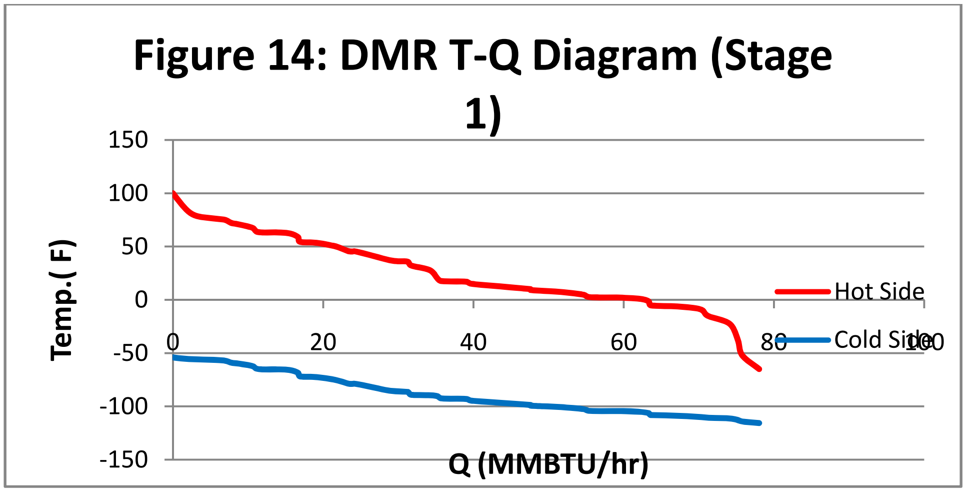


Figure 14: DMR T-Q Diagram (Stage 1)

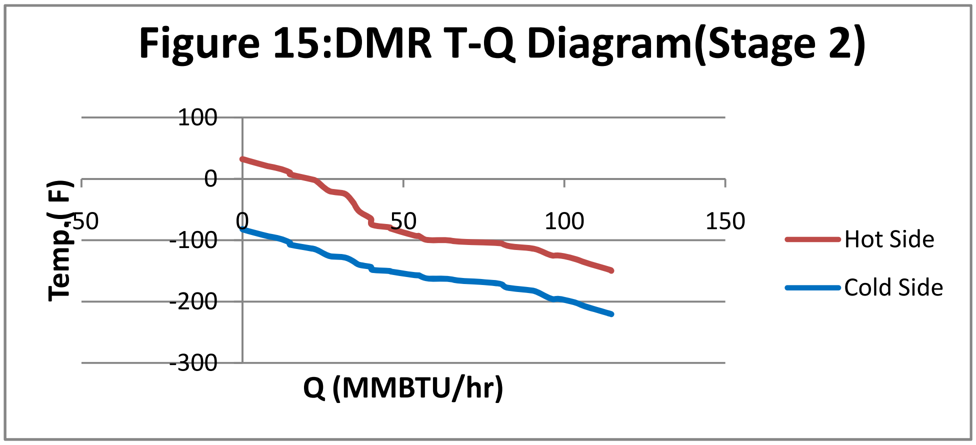


Figure 15:DMR T-Q Diagram(Stage 2)

The next step is to optimize the compressor work. The refrigerant flow rate was first lowered in first stage and the compressor work went down by a significant amount. Reducing the flow rate of the second stage refrigerant caused temperature crossing, the outlet temperature of the cold cell was higher than the inlet temperature of the natural gas hot cell. At the present conditions, 20000 lb/mol was the optimal flow rate for the second stage refrigerant. The refrigerant compositions and flow rates at this stage of optimization are shown in table 7.

Table 7:Refrigerant Composition and Flow Rate

	Stage 1 (%)	Stage 2 (%)
Methane	21	49
Ethane	47	41
Propane	6	5
n-Butane	25	[redacted]
i-Butane	1	5
Total flow rate (lbmol/hr)	18000	20000

The compressor works were further optimized by decreasing the temperature approach of the streams in both stages. This is done by using an optimizer to minimize the total compressor work by changing the valve outlet pressure and adjusting the outlet temperatures of the hot streams to decrease the temperature approach. The outlet temperature of the natural gas is not changed as there is a set target. In the first refrigeration stage the outlet pressure of the valve increased from 15 psia to 66 psia. In the second stage, it increased from 15 psia to 67 psia. At this point optimization is at a stopping point. The compressor works at the end of optimization are shown in table 8. Figures 16 and 17 show the T-Q diagrams of the first and second stage, respectively, after optimization is complete.

Table 8: Compressor Specifications

	Compressor Work (HP)	Outlet Pressure (Psia)
C1 (Stage 1)	13960	500
C2 (Stage 2)	10759	495
C3 (Stage 2)	3675	1000

Figure 16:DMR T-Q Diagram (Stage 1)

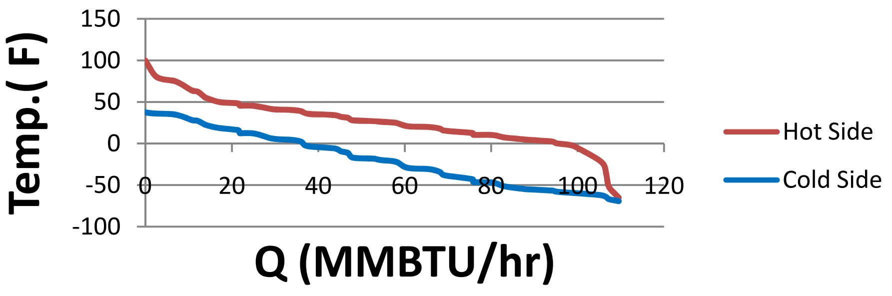
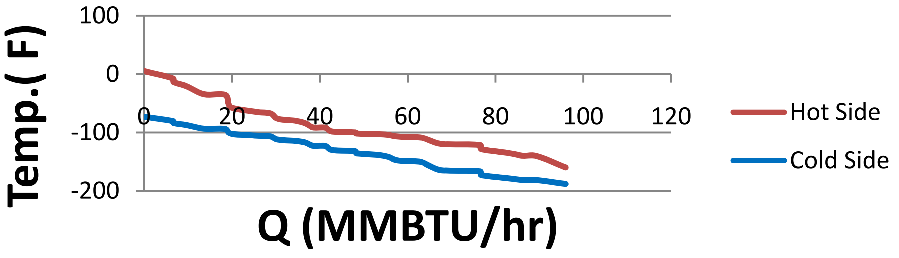


Figure 17: DMR T-Q Diagram (Stage 2)



APCI C3-MR

Description

The APCI C3-MR process has two cooling cycles. One is a series of heat exchangers using propane to pre-cool the natural gas. The three heat exchangers in the series each have propane at a different pressure: high pressure, medium pressure, and low pressure. The different pressures used allow the propane to be cooled to different temperatures to allow the natural gas to be cooled to an initial temperature before it enters a spiral wound heat exchanger that accomplishes most of the cooling.

A propane pre-cooling series is also used on the second cooling loop used in the spiral wound heat exchanger. Again the series runs high pressure, medium pressure, and low pressure propane to pre-cool the refrigerant after it is compressed. It then enters a flash tank and separated into two streams and from here the spiral wound heat exchanger to be further cooled, and then used for cooling. Another method that allows the refrigerant to be cooled is expansion through a valve after each stage of cooling in the spiral wound heat exchanger.

The mixed refrigerant used is composed of methane, ethane, propane, butane, and nitrogen. The suggested composition of the mixed refrigerant is shown in the first table, table 9. The work done by the compressor was optimized based on the composition of the refrigerant. The determined composition of the refrigerant after compressor optimization is shown in the second table, table 10.

Table 9

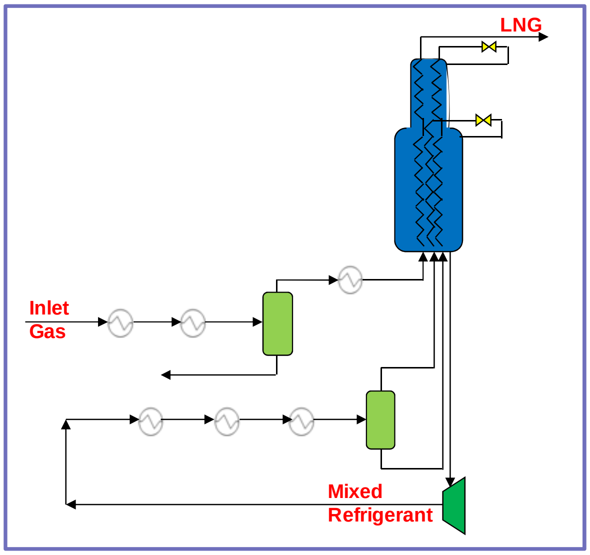
Component	Composition
Nitrogen	1%
Methane	27-30%
Ethane	50%
Propane	18-20%
Butane	1-2%

Table 10

Component	Composition
Nitrogen	.01
Methane	.27
Ethane	.50
Propane	.2
Butane	.01

The APCI C3-MR process is one of the most used processes for liquefaction, being used in almost 90% of the liquefaction industry. As one of the "original" processes and longest used, it has one of the largest capacities.

Figure 18: Process Flow Diagram of C3-MR



The maximum capacity of the C3-MR process is roughly 5 MTPA which is one of its advantages. One of the main disadvantages of the C3-MR process is its high equipment cost. With the use of the propane exchangers, there is a high utility cost associated, as well as large cost of the spiral wound heat exchanger. The amount of work required by the compressor is very large, which also increases the cost of the processes. A large compressor or multiple numbers of compressors are required to compress the amount of refrigerant in the process.

Simulation Method

When beginning the simulation of the C3-MR process, I tried to find information on specifications for the refrigerant, temperatures in and out of the heat exchangers (both LNG and regular heat exchangers), and information on the pressures of the pre-cooling heat exchangers. Most of this information was unavailable, except for the composition of the mixed refrigerant.

This process uses a spiral wound heat exchanger, which is not available as equipment in ProII. In order to simulate this, two LNG heat exchangers were used. I placed these in the process, inputted the composition of the natural gas streams and of the refrigerant, set up any other equipment associated with the LNG heat exchangers as it was used to simulate the spiral wound heat exchanger, including a mixer and two valves, and ran the simulation. The hot and cold streams needed to be labeled in the process and all streams flowing in the same direction as the natural gas stream were considered to be hot streams and all the streams flowing in the opposite direction of the natural gas stream were considered cold streams. Once this worked when the simulation was run, I began to place in the other equipment associated with the process based on the PFD I found for the process.

Figure 19: C3- MR T-Q Diagram (Stage 1)

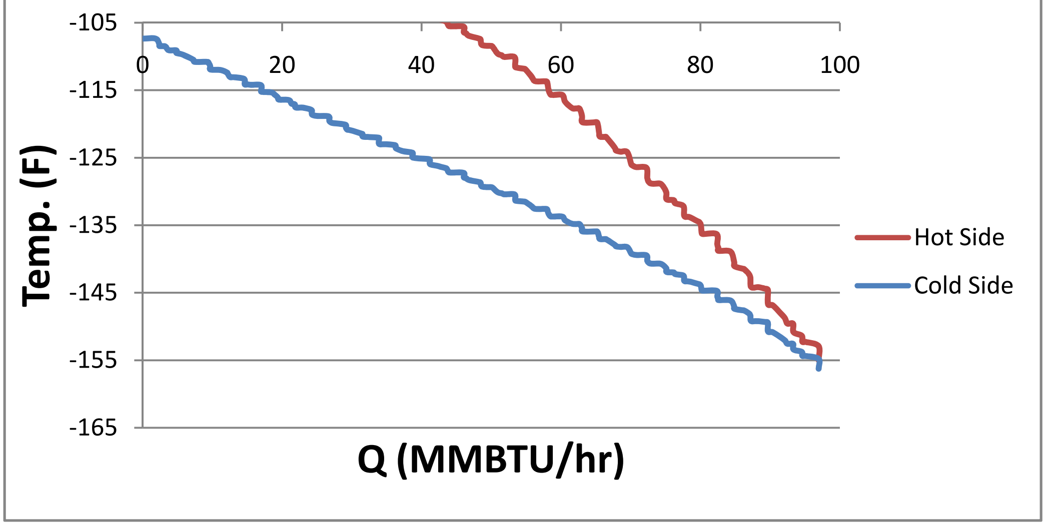
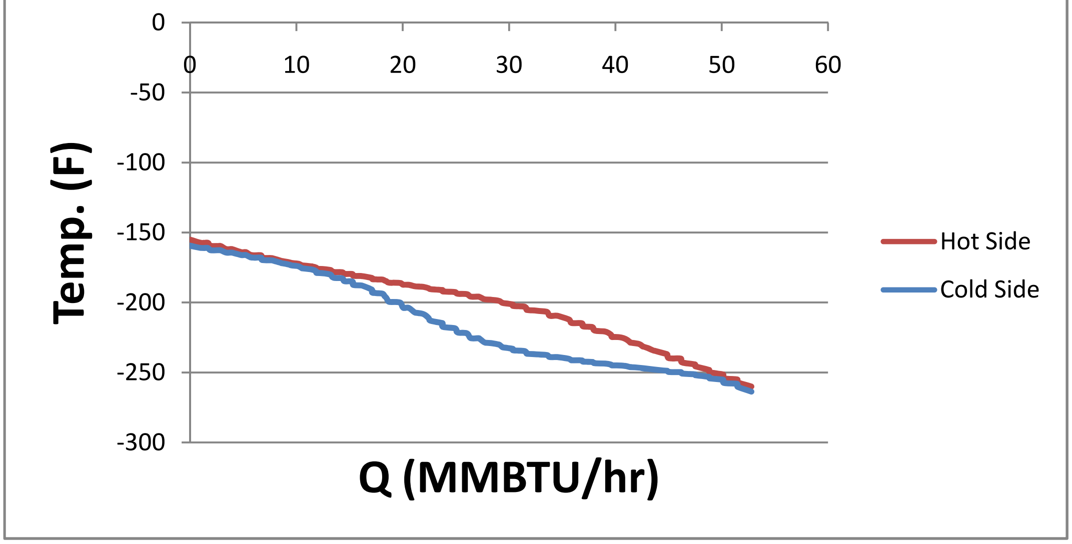


Figure 20: C3- MR T-Q Diagram (Stage 2)



The other equipment included, in pre-cooling the natural gas, three propane cooled heat exchangers and a flash tank. The propane cooled heat exchangers each ran with a different pressure of propane based on the temperature the natural gas needed to be cooled to. Altering the pressure of the refrigerant changes the properties of the propane and allows different temperatures to be reached. I selected the temperature I wanted heat exchangers to cool the natural gas to before it reached the LNG heat exchangers. I then set the duty of the flash tank to be zero, with a pressure drop of zero. I ran this to make sure the process still worked.

After I finished the pre-cooling loop, I began on the refrigeration loop. This loop also contained propane pre-cooling heat exchangers and a flash tank, but to complete the refrigeration loop, a compressor was also included. The valve at the end of the spiral wound heat exchanger is the last piece in the refrigeration loop. Again, I set temperatures that I wanted the heat exchangers to cool the refrigerant to before it entered the spiral wound heat exchanger to be cooled further, then used for refrigeration. I also selected a pressure to which I wanted the compressor to compress the refrigerant to after it left the spiral wound heat exchanger. This pressure was the original pressure that I had decided on for the refrigerant. The two valves in the process reduce the pressure to further cool the refrigerant, so after it leaves the spiral wound heat exchanger, it needs to be compressed.

Before the refrigerant enters the spiral wound heat exchanger, but after it has been pre-cooled, it enters a flash tank. This flash tank separates the liquid refrigerant from the gas refrigerant. It also allows for the refrigerant to be cooled further if needed. In this case, it is not needed and the specifications of the flash tank were the same as for the natural gas flash tank: zero duty and no pressure drop.

From here, I began to set temperatures in my LNG heat exchangers. All streams labeled as hot streams needed to be cooled to temperatures less than they entered, and all cold streams needed to leave hotter than they entered. I began to adjust the temperatures of each of the heat exchangers to achieve this. As the temperatures were adjusted, the specifications in most of the other equipment were adjusted as well. The valve outlet pressures were adjusted to help cool the refrigerant where needed. The flow rate of the refrigerant was also adjusted. Temperatures in the pre-cooling heat exchangers were also adjusted to try and help with cooling as well. I varied the specifications in each piece of equipment based on the outcome after running the simulation until I obtained the same output conditions for the refrigerant as I had for input conditions.

Once good operating conditions were achieved, the compressor work was optimized based on the composition of the refrigerant. To do this, separate streams with the flow rates of each specific component were inputted into individual streams entering a mixer. An optimizer was placed in the simulation once this was done, with the condition of minimizing compressor work by varying the flow rate of the refrigerants. Once this was done, the flow rate in the stream following the mixer was obtained, and the individual streams were removed in place of a single one. The specifications obtained from the simulation on the composition of the flow rates and percentages were included in a single stream and used to run the simulation. If any adjustments of other equipment were needed, they were made and checked as before until the right conditions were met.

The final step in this process was using calculators to calculate the log mean temperature difference between the cold and hot streams in the process. Once these were obtained, the areas between each cell in the heat exchanger were calculated based on the log mean temperature difference between the cells.

Components that could be changed in the process were valve outlet pressures, LNG heat exchanger cell temperatures, temperatures to which the natural gas and refrigerant could be pre-cooled to, and the refrigerant composition.

In changing the temperatures and pressures, the values chosen can either help or hurt the simulation. I found that in order for my simulation to work, that I could not choose a temperature below -170 F or above -120 F in my first heat exchanger. The input temperature of the streams also had to be between 10 F and -100 F or else the heat exchanger gave crossing temperatures. It was better to have temperatures between -50 F and -100 F. In the second heat exchanger, the values inputted had to be between -150 F and -250 F for the temperatures to not cross or to have heat flowing in the right direction. This was also paired with a valve. The valve had to have a pressure that outputted a temperature at least -20 below the desired output temperatures, depending on the flow rate of the refrigerant, for the temperatures to not cross and the heat exchanger to work. The selected value was an outlet pressure of 150 psi.

The second valve, had to output a temperature that was less than the output temperature of the first heat exchanger and to achieve this, the refrigerant always had to be reduced to atmospheric pressure in order for it to work. The output temperature of the valve very rarely changed with the flow rate of the refrigerant. The final valve in the process expanded the natural gas stream to atmospheric pressure and a temperature of -257 F.

APCI AP-X

Description

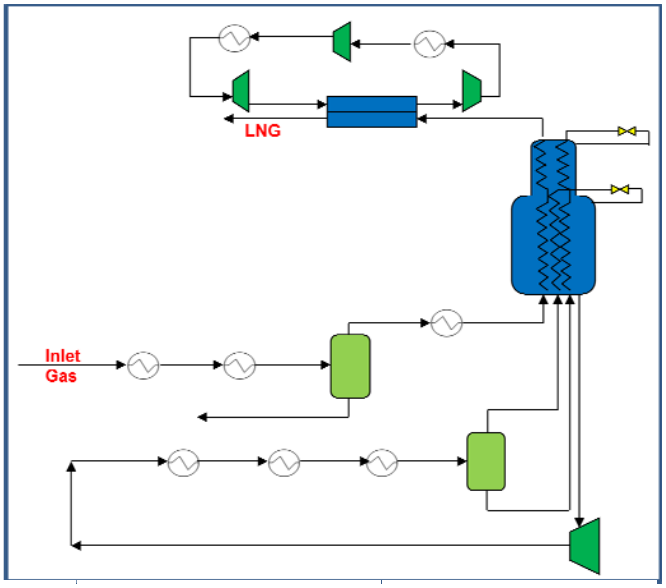


Figure 21: Process Flow Diagram of AP-X

The APCI AP-X process is a hybrid modification of the APCI C3-MR. The modification is the addition of a nitrogen sub-cooling loop after the spiral wound heat exchanger. The addition of this loop increases the efficiency and capacity of the process. It does this by reducing the amount of work done in the spiral wound heat exchanger, just as the pre-cooling loop does. The first two loops are the same as the C3-MR process: propane pre-cooling followed by the mixed refrigerant loop, with the mixed refrigerant loop using the propane pre-cooling as well.

The mixed refrigerant loop also has the same components of methane, ethane, propane, butane, and nitrogen as the C3-MR process. The composition was again changed to optimize the compressor flow rate. The initial and final compositions of the refrigerant in the AP-X process were the same as the refrigerant composition for the C3-MR process.

Table 11	
Component	Composition
Nitrogen	1%
Methane	27-30%
Ethane	50%
Propane	18-20%
Butane	1-2%

Table 12	
Component	Composition
Nitrogen	.01
Methane	.27
Ethane	.5
Propane	.2
Butane	.01

The final loop added is pure nitrogen compressed by two compressors and cooled by two water-cooled heat exchangers and then expanded through an expander. The nitrogen is able to be cooled to a much lower temperature than the mixed refrigerant, which allows more of the cooling to be done in the final loop and lowers the amount of mixed refrigerant needed through the spiral wound heat exchangers.

The addition of the nitrogen sub-cooling loop is what creates the hybridization. It pairs a spiral wound heat exchanger that utilizes mixed refrigerants with a plate fin heat exchanger using a pure refrigerant that is not a hydrocarbon, which is usually what is used in a pure component refrigerant. This also increases the capacity from 5 MTPA to almost 9 MTPA, which is an indication of how much more efficient the process is supposed to be.

The AP-X process is increasing in popularity due to the simple modifications to the C3-MR process. The main advantage of this process is the increased capacity. Very few processes have such a high capacity that it makes it an appealing process to implement. The disadvantage of the AP-X process is the same as the C3-MR. It has very costly equipment with the amount of work the compressors are required to perform, which are the main costs aside from the spiral wound heat exchanger.

Simulation Method

Since the AP-X process is a modification of the C3-MR, I took the simulation from the C3-MR and added the changes needed for the AP-X. This included a nitrogen sub-cooling section. This section had four water cooled heat exchangers, two compressors, an expander, and an LNG heat exchanger.

Components that could be varied during the course of the simulation were the valve outlet pressures, the LNG heat exchanger outlet temperatures, temperatures to which the natural gas and refrigerant could be cooled, and refrigerant composition.

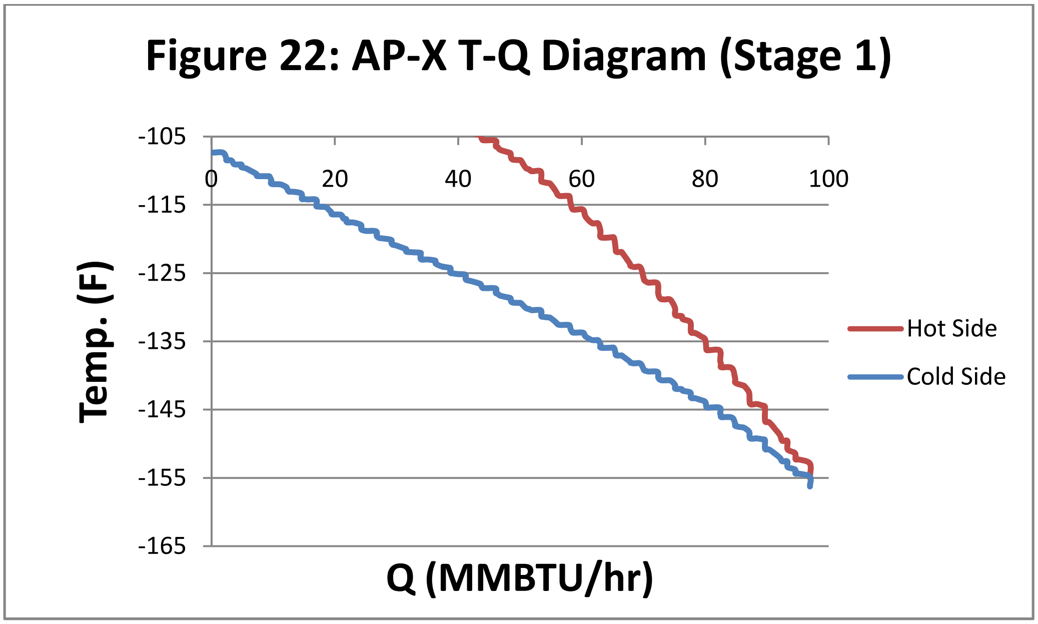


Figure 22: AP-X T-Q Diagram (Stage 1)

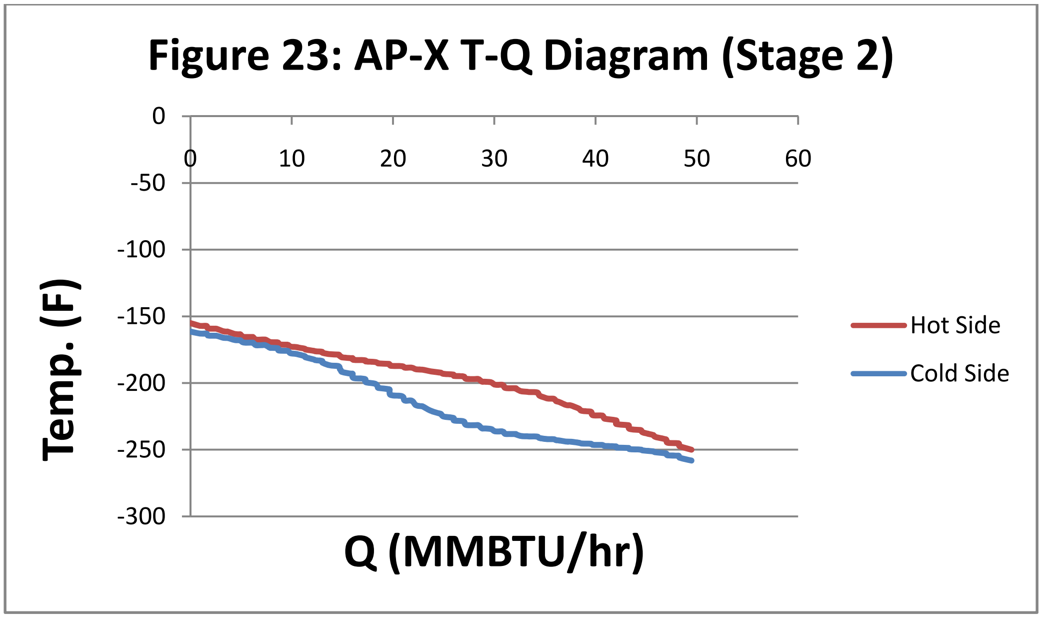


Figure 23: AP-X T-Q Diagram (Stage 2)

Once these were set up, I made my specifications for the LNG heat exchanger, whether a stream was hot or cold and what temperature I wanted the natural gas stream to be cooled to. I used the heat exchangers and compressors to increase the pressure of the nitrogen stream and to cool it, before it was expanded before it entered the LNG heat exchanger. I varied the specifications in each piece of equipment based on the outcome after running the simulation until I obtained the same output conditions as I had for input condition.

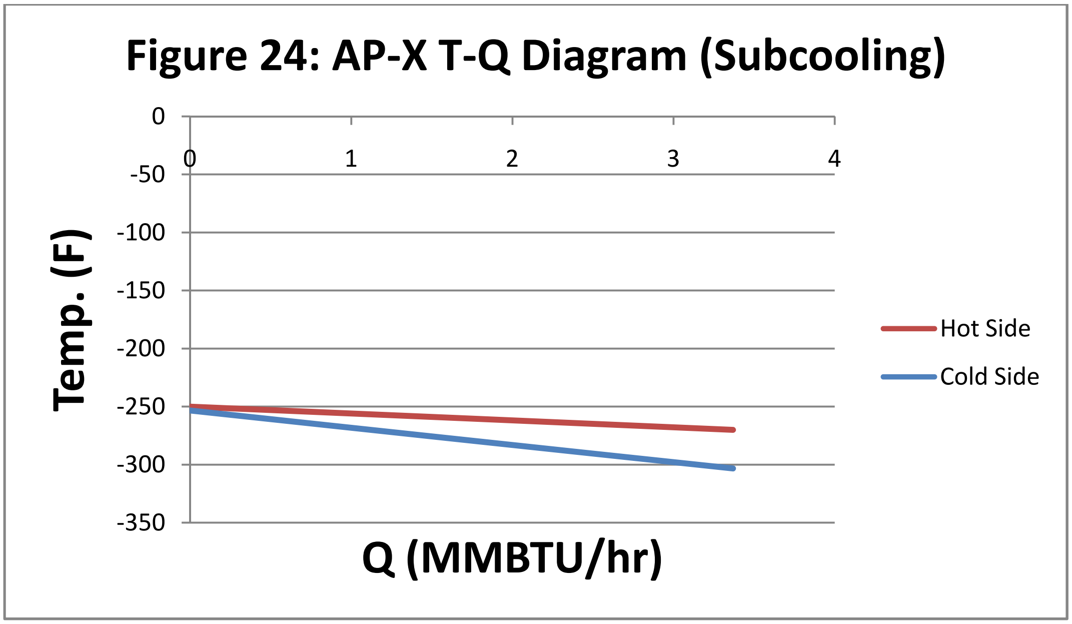


Figure 24: AP-X T-Q Diagram (Subcooling)

Another calculator was set up to calculate the log mean temperature difference between the cells and to calculate the area between the cells.

The values for the heat exchangers for the AP-X process were different from the C3-MR process because of the addition of the nitrogen sub-cooling loop. The desired temperature for the initial heat exchanger to cool to was -160 F. As mentioned above for the C3-MR process, the range of temperatures over which the heat exchanger did not cross temperatures or have heat flowing in the wrong direction were input stream temperatures between 10 F and -100 F and output temperatures between -120 F and -170 F.

The second heat exchanger had input stream temperatures of what came from the initial heat exchanger, and an output of -205 F. A range of temperatures between -202 F and -250 F allowed the heat exchanger to work properly with no crossed temperatures and no reversal of heat transfer. The valve paired with this process had an outlet pressure of 150 psi and required a temperature difference of 20 F, depending on refrigerant flow rate, in order to operate properly. The valve between the heat exchangers required an outlet pressure of atmospheric pressure in order to have a lower temperature in the stream cooling the initial refrigerant.

The sub-cooling stage cooled the natural gas to a temperature of -260 F. The pressure at which this cooling took place was atmospheric pressure based on refrigerant flow rates.

Axens Liquefin Process

Description

The major feature of the Liquefin process is that it consists of its two refrigeration cycles both using mixed refrigerants where one cycle is involved in pre-cooling the raw natural gas and the second cycle completes the liquefaction procedure. This division of tasks between the refrigeration cycles helps to increase production capacity and the amount of refrigeration to be used in the process.

The maximum capacity of the Liquefin process is reported at about 6 MTPA. In the pre-cooling stage, the natural gas is cooled effectively to about -70 F to -80 F. Due to the composition of the refrigerant, a smaller amount of the mixed refrigerant will be used thereby reducing the amount of work being put in by compressors at that stage.

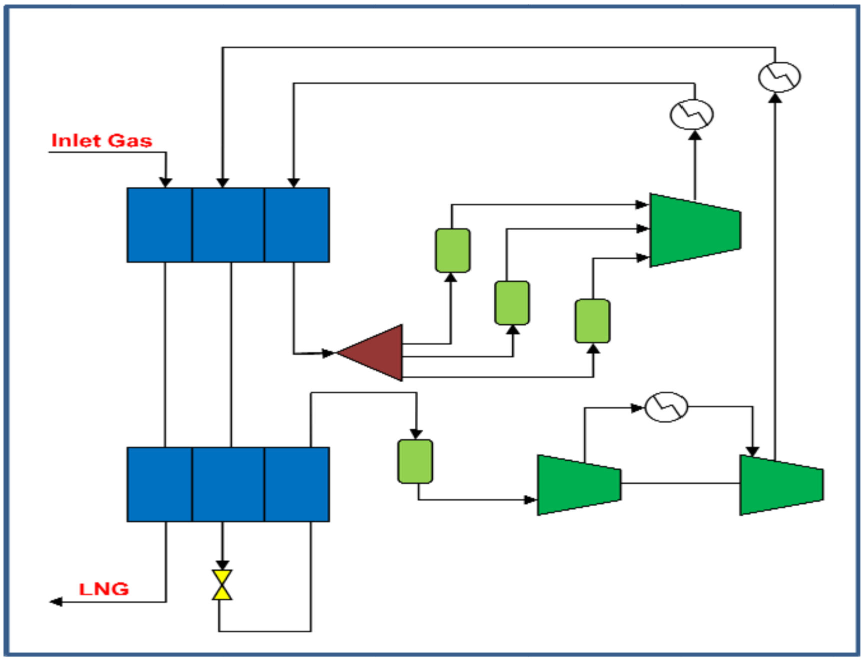


Figure 25: Liquefin Process

The final liquefaction stage would require the most amount of compression and cooling. This is facilitated by the use to two separate compressors and a larger amount of the refrigerant mixture. Not only can it operate at the maximum capacity stated earlier, it can operate in intermediate ranges due to the flexibility in the amount of refrigerant that can be used. The mixed refrigerant is composed of methane, ethane, propane, butane and nitrogen all of which are readily available from the natural gas itself. The more natural gas available, the more of the refrigerant can be available to produce the natural gas stream.

The heat exchanger network in the process is a combination of multiple plate fin heat exchangers. These PFHE are of similar dimensions and assembled side by side in a cold box. The design and efficiency of the PFHE makes the transfer from the exchangers in a cold reliable and easily maintained. Figure 26 below shows the schematic representation of the cold box and how the PFHE’s are arranged.

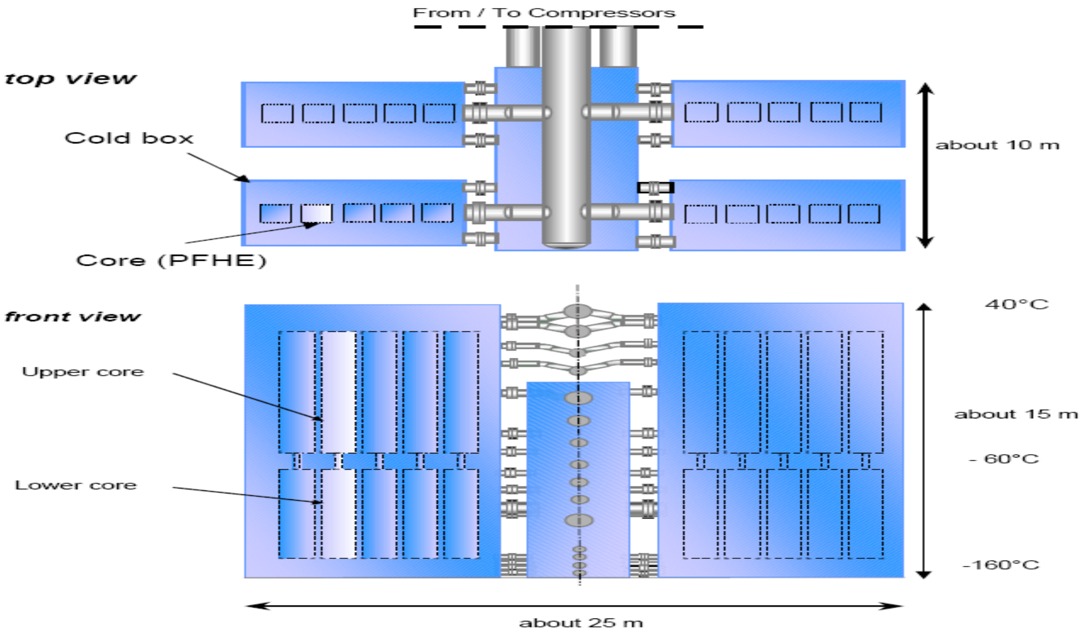


Figure 26: Typical Cold Box Arrangement

The flexibility in the ability to increase its production capacity by increasing compressor sizes and refrigerant flow rates is considered a great advantage. These capacity increases can be upwards of about 7-8 MTPA. With a smaller amount of rotating equipment, maintenance costs are significantly lowered when compared to larger processes. Energy usage can be alternated between the two refrigeration cycles which can reduce downtime and costs. The pre-cooling stage of the process is rather rigid due to the setup of the cold box involved.

Simulation Method

The cold box configuration in this process is essential in this simulation process. Due to its extensive design, in order to obtain constant area through each plate, the natural gas is first split into 4 equal natural gas streams before being fed into the heat exchanger. The split is due to the process flow diagram and the complex nature of the cold box setup described in earlier sections. A mixed refrigerant of mainly of propane and nitrogen is employed in this process. The specific composition of the hydrocarbons and nitrogen used in both cycles is shown in the table 13 below.

Table 13

Component	Composition (%)
Methane	2.3
Ethane	2.3
Propane	45.6

Butane	15.4
Nitrogen	34.4

The first refrigeration stream is separated into 3 equal portions as well with the help of a splitter before being fed into the LNG heat exchanger. The splitting of the streams is to help in calculating the area of heat transfer. This portion is meant to pre-cool the natural gas to temperatures around -70 F. The stream passes through the heat exchanger, gains heat from the natural gas and a vapor phase is formed during the process. This cold stream is then recycled back after an isenthalpic expansion that helps in cooling the natural gas further till the right temperature is obtained.

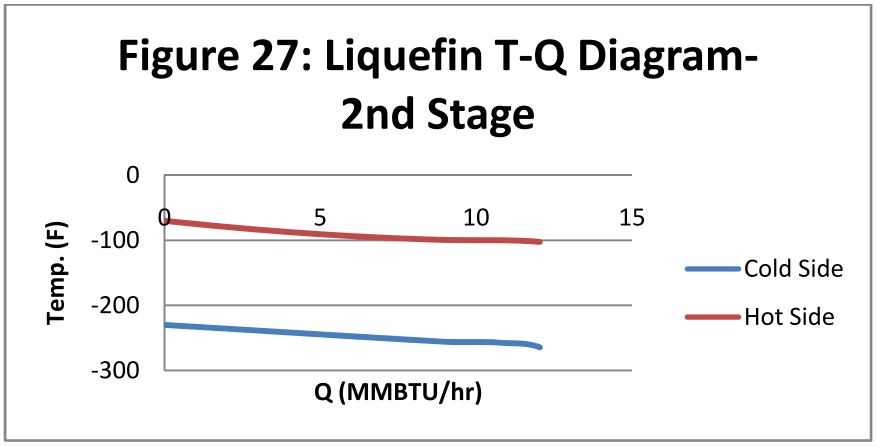
In this stage, the area required to cool the natural gas is small due to splitting up the natural gas. This shows how the process really capitalizes on the plate –fin heat exchanger features. It also helps in reducing the amount of compression that has to be done in the first refrigeration cycle.

After cooling the natural gas, there is the need to recompress the cold stream back into an entirely liquid stream. This is obtained by the use of three separate flash tanks that operate at low temperatures. The flash tanks are specified to cool the mixture to at least yield a 90% liquid fraction through the cooling. It is then totally liquefied by the compressor after the pressure is raised back to the initial 450psia. After the liquid is re-compressed to the adequate temperature, it is then cooled by propane back to its bubble point temperature of about -289 F and then the cycle is repeated.

The 2nd cooling stream is introduced in the first pre cooling stage as well but substantial heat gain does not occur there. The flow rate of the refrigerant in this stage is significantly greater than that of the first stage which accounts for the minimal heat loss experienced. Also only one-fourth of the LNG stream is being cooled by the second refrigerant stream. The stream therefore gains only a minimal amount of heat before leaving at about -269 F.

Pre-cooled natural gas at about -70 F is all sent into a mixer before further cooling is performed. At this point some liquid has already begun forming before being passed into the second stage LNG heat exchanger. The final cooling occurs in this stage and the final LNG stream exits at almost -130 F. The refrigerant stream enters the heat exchanger then is recycled back in similar to the first stage to further help the cooling process. At this point, the cold stream gains the most heat from the natural gas thereby heating itself up to about -110 F. At this stage, additional power is required to re-cool and recompress the stream back to it required temperature.

Refrigeration of the cooling stream is obtained with the help of two propane-cooled heat exchangers and two compressors in order to obtain a refrigerant close to its bubble point. The T-Q diagram in Figure 27 below represents the efficiency of this cooling process and indicates how much needs to be done by the stream to cool the natural gas stream to required temperatures.



In order to determine the areas of the plate fins in the LNG heat exchanger, the controller is used to determine the expected outlet temperatures through each cell in order to obtain the uniform area. In the calculator, the formula for the area is calculated based on the current outlet temperatures and heat gained values. A controller is then used to control the temperatures until all the area calculations become equal. This equality then shows how much cooling or heating a section of the heat exchanger is responsible for.

An optimizer is used to control the flow rates of the refrigerants that will reduce the compressor work to its minimum while satisfying the simulation specifications. It is further simplified by summing up the work involved in each stage and then minimizing the sum by changing the flow rates and maintaining all other conditions as well.

ExxonMobil Dual Multi-component process (US patent no.: 6,250,105)

Description

This process was patented by ExxonMobil in 1999. It was designed similarly to APCI's DMR process with the use of two mixed refrigerant cycles cooling each other and the natural gas as well. This process however switches that trend and has the two refrigerant cycles operating at high and low pressures.

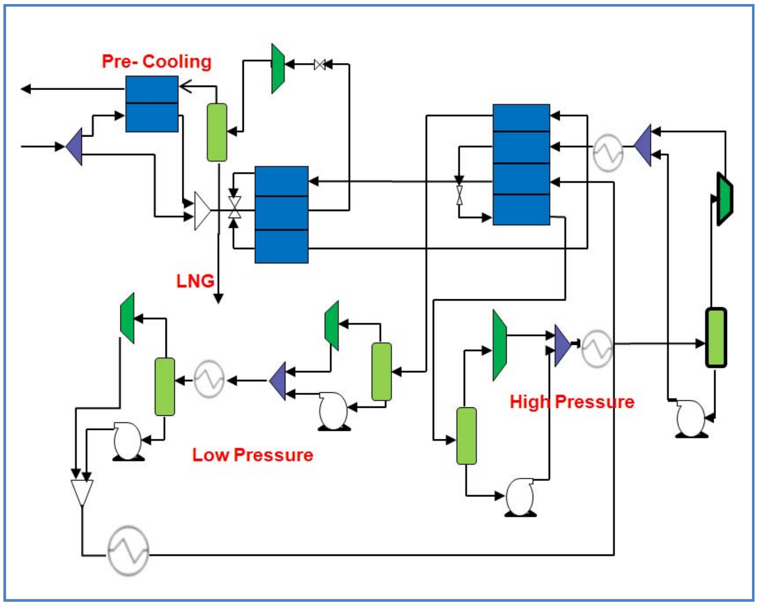


Figure 28: ExxonMobil Process

The refrigeration cycle operating at high pressure has the function of primarily cooling the low pressure refrigerant. This thereby reducing the amount of work required in the cooling of the both cycles with one cycle accounting for most of the energy required.

Another alteration to the DMR process is the re-introduction of the raw natural gas in the pre-cooling stage. During the separation of the cooled LNG from its cooled vapor component, after both refrigeration cycles have been employed, the cooled vapor is reintroduced into the feed stream to help pre-cool to a temperature between 20 F and 40 F. The heated gas is collected as either fuel gas or it is reintroduced into the system to increase the fraction of LNG to be produced.

The mixed refrigerant is composed of mostly hydrocarbons namely methane, ethane, propane, butane and pentane. The low pressure cycle is run at pressures ranging from 100 to 150 psia. The two compressors in that setup have the capability of handling the load very efficiently. However, in case of any failures, one compressor is capable of running the cycle to avoid any downtime that could be experienced. The high pressure cycle is run at pressures ranging from 300 to 350psia. The same basic principles apply in both cycles but higher rated compressors are needed in the high pressure area.

This process is capable of handling large capacities, up to 9 MTPA. This is due to the division of selected tasks of different regions in the process. It can also help reduce the downtime if any equipment fails by being converted into a single refrigeration stream if necessary. However, the numerous types of equipment lead to increased capital, utility and maintenance costs. Also, the partial liquefaction of the natural gas leads a large amount of unused gas that is can be resold as fuel gas or reinserted into the process.

Simulation Method

A combination of two mixed-refrigerant refrigeration cycles to produce a LNG stream proves to be very complex with the number of equipments involved.

The inlet gas stream, NG, is first split into 2 streams with one stream to be pre-cooled. Most of the stream is passed on into a mixer where the pre-cooled stream is input as well. This is to help cool the total natural gas stream before it is sent to the heat exchanger. In the pre-cooling section, the final stage of the entire process plays a role in cooling this stage. It uses the residue vapor available in cooling the specified portion of to significantly lower temperature. The amount of cooling obtained here, dictates how low a temperature can be obtained from the heat exchanger.

The first heat exchanger is comprised of three cells with the natural gas as the only hot stream. The coldest stream in this heat exchanger is obtained from the dual refrigeration cycle. This refrigerant stream is at a temperature of about -209F and a pressure of about 150psia. This stream is labeled as a cold stream in order to cool some of the natural gas before it is passed to an expansion valve. It is obtained as the coldest stream from the separate refrigerant cycle. The expansion valve helps in sub-cooling the natural gas further. There is pressure drop in the expansion valve by about 100psi which leads to a slight temperature drop. The pressure drop value was selected because it accommodates both refrigeration cycles and allows the simulation to converge.

The cooled natural gas stream is then passed on into another valve for further expansion and cooling. However, the patent flow diagram denotes all pressure drops occurring in valves without expanders. In order to obtain the LNG at atmospheric conditions, expanders have to be implemented eventually and that can be taken advantage of early in the process. Hence an expander is added to the process to obtain additional cooling and pressure requirements.

The T-Q diagram in Figure 29 shows the amount of cooling done strictly by the refrigerant in heat exchanger in addition to the pre-cooling of the natural gas by the refrigerated gas. The pre-cooling stage occurs efficiently to about -126F which could be attributed to the lower flow rate of the natural gas that is sent to be pre-cooled. The splitter is set to send about 25% of the natural gas feed be sent to pre-cooling heat exchanger. The small amount of the gas enables the cooling to be done more efficiently than it is done in the 2nd stage.

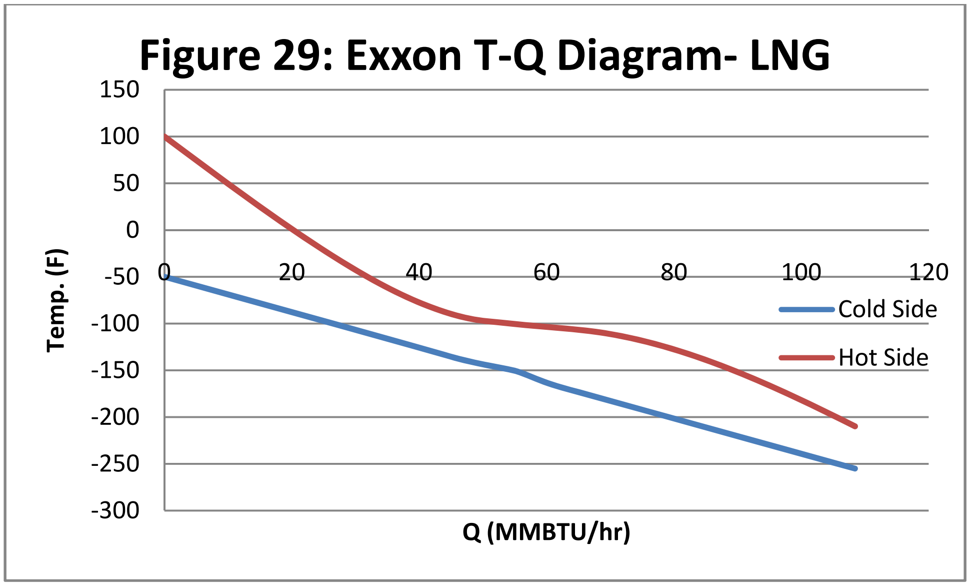


Figure 29 shows the overall cooling from the inlet natural gas. It proceeds with a rather stable efficiency but the recovery percentage of actual LNG occurs at merely 15%. This would mean that the bulk of natural gas injected into the system would merely become residue gas to be recycled into the process. A controller is in place to ensure the final LNG stream obtained matches the requirement of the project (220000 lb/hr).

The controller allows the flash tank to determine the amount of cooling it needs to input to obtain the project goals. Increasing the tank duty would increase the amount of LNG produced but will increase the compressor work involved in the process.

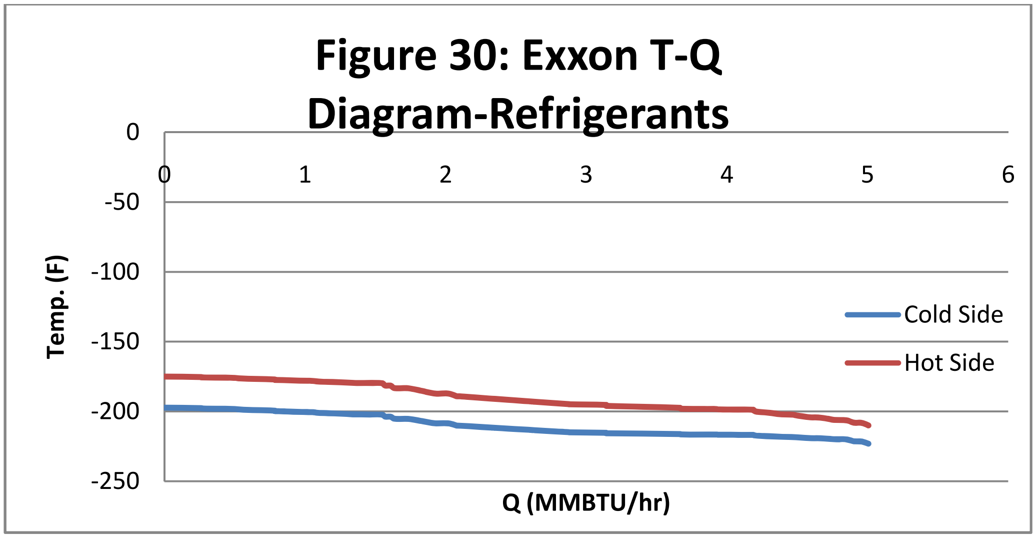
The bulk of the work in this process comes from the high and low pressure refrigerant cycles. There are different refrigerant compositions for the separate refrigeration cycles. The composition of the refrigerant is shown in Table 14 below.

Table 14:Refrigerant Composition

	Composition (%)	
Component	Low pressure	High pressure
Methane	71.4	19
Ethane	7.1	27.5
Propane	7.4	[redacted]
Butane	6.9	25.5
Pentane	7.2	28

In order to determine the appropriate composition to use, an optimizer is used to vary the flow rates of each component into a mixer. The input streams of each component would have the same flow rate to begin with but the optimizer, OP1, varies each streams flow rate until the coldest temperature out of the heat exchanger.

The setup on cooling the refrigerants involves the use of compressors, pumps and propane cooled heat exchangers. The simulation techniques at both pressure levels are similar with the compressor and refrigerant composition differing. Using multiple equipment configurations to cool the refrigerant allows for increased efficiency but might affect the cost involved. Its efficiency can be judged in the T-Q chart below.



The low and high pressure value is suggested by the patent information. The low pressure adopted by the simlation is 150psia and that of the high pressure stream at about 350psia. The high pressure refrigerant is solely responsible for cooling the low pressure refrigerant. The LP refrigerant is then responsible for cooling the natural gas feed as highlighted in Figure 2 above. The amount of refigerants used is comparatively low compared to LNG produced. The heat exchangers involved are propane cooled (as stipulated by the utility streanm in Pro II)in order to get the recycled refigerant cool enough for LNG cooling.

Another heat exchanger is where the high pressure refiigerant sub-cools the low pressure refrigerant before it is sent up for LNG cooling. The areas of the three heat exchangers are controlled by another controller. It controls the individual temperatures of the cells as well as the flowrates of the refrigerants. Selecting the individual temperatures is key to obtaining a successful simulation.

An optimizer was introduced in a separate simulation document to reduce the compressor work by adjusting the flow rates of the natural gas and the refrigerants. It however leads to very unfeasible areas and compressor work values as well.

Technip-TEALARC Process

Description Method

This process consists of two refrigeration cycles. The first cycle cools a mixed refrigerant stream. The refrigerant in this first cycle consists of a mixture of ethane and propane. The second cycle cools the natural gas with the mixed refrigerant from the first cycle.

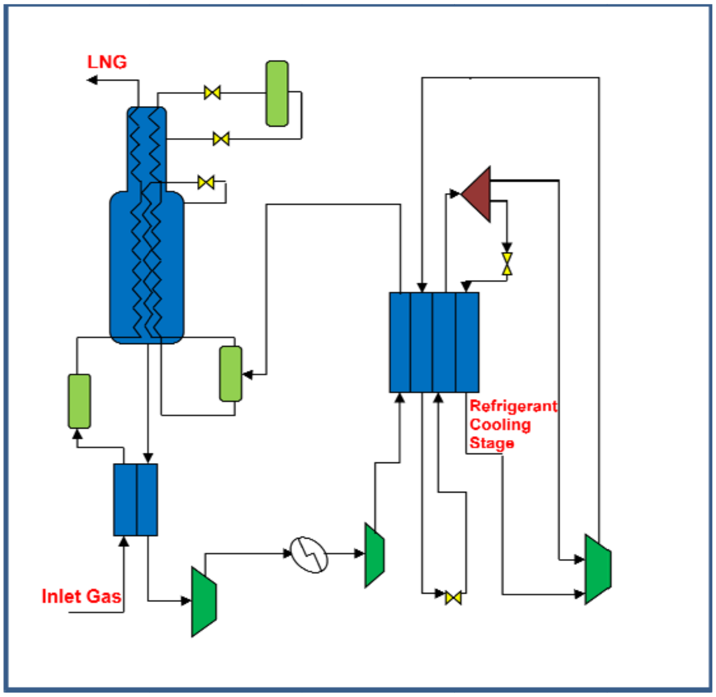


Figure 31: Technip TEALARC

Simulation method

There was no information found on the refrigerant composition of the natural gas cooling stage so a guess was made. The refrigerant composition of the refrigerant cooling stage was reported to be a mixture of ethane and propane. The compositions used are shown in the table below along with the first guesses of total flow rate.

Table 15

	Natural Gas Cooling Stage (%)	**Refrigerant Cooling Stage (%)**
Methane	80	
Ethane	5	5
Propane	5	95
n-Butane	5	
Nitrogen	10	

Total flow rate (lbmol/hr)	35000	24000

These parameters were entered into the simulation and temperature crossing was observed in the refrigerant cooling stage. The temperature crossing was present on both the inlet and outlet side of the heat exchanger. To alleviate this problem, the flow rate was first adjusted but this method proved to be ineffective. This decreased the approach temperature by a little bit but the temperature crossing was still present. The flow rate was returned to the initial value. The next adjustment made was to the refrigerant composition. To reduce the inlet temperature of the refrigerant, the amount of the heavier component (propane), was reduced and the amount of ethane was increased by the balance. This was also ineffective in fixing the temperature crossing problem. To further decrease the inlet temperature of the refrigerant, the outlet pressure of the valves were reduced but this proved ineffective as well. At this point, other components were added to the refrigerant namely, methane and butane. This approach was successfully in the removal of the temperature crossing. The final compositions are shown in table 16.

Table 16

Refrigerant Composition and Flow Rate

	Natural Gas Stage (%)	**Refrigerant Stage (%)**
Methane	80	5
Ethane	5	65
Propane	5	20
n-Butane	5	10
i-Butane	10	[redacted]

At this point the simulation was working in that the natural gas target temperatures were being achieved without any temperature crossing. The compressor works before optimization are shown in table 17. The T-Q diagrams at this point are shown in Figures 32 to 35

Table 17

	Compressor Work (HP)	**Outlet Pressure (Psia)**
C1 (Natural Gas Stage)	60701	520
C2 (Natural Gas)	9255	800
C3 (Stage 2)	23636	13000

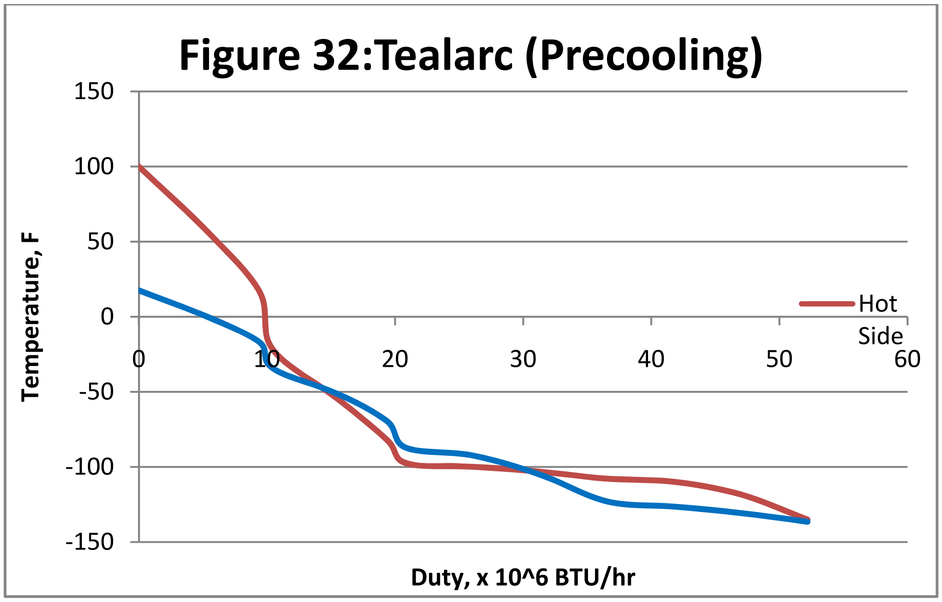


Figure 32:Tealarc (Precooling)
150
100
50
0
-50
-100
-150
Temperature, F
0
10
20
30
40
50
60
Hot Side
Duty, x 10^6 BTU/hr

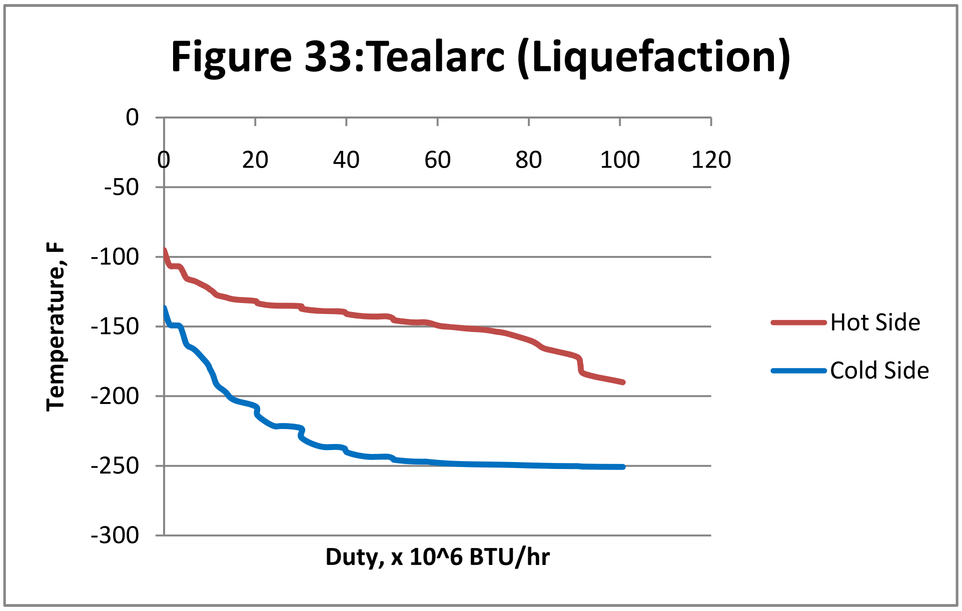
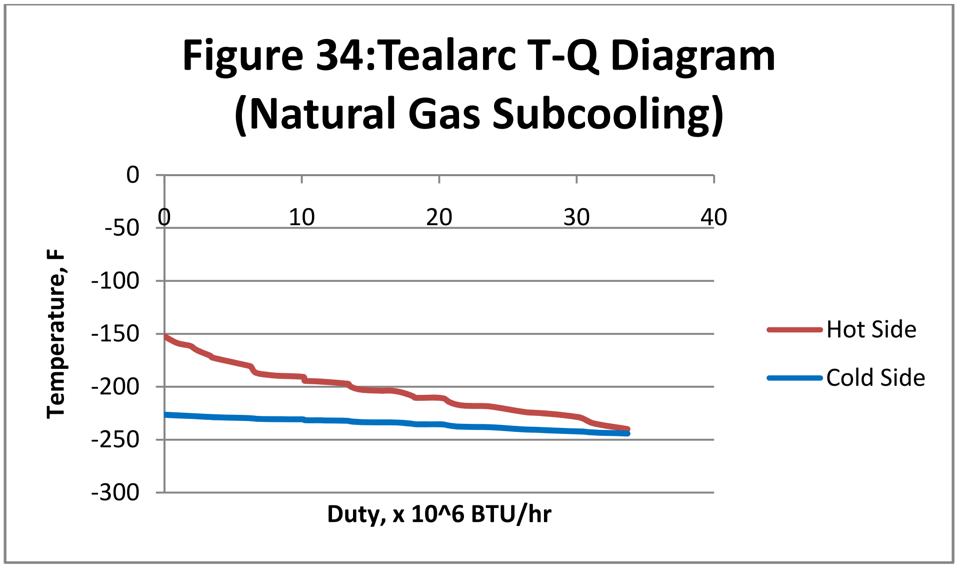
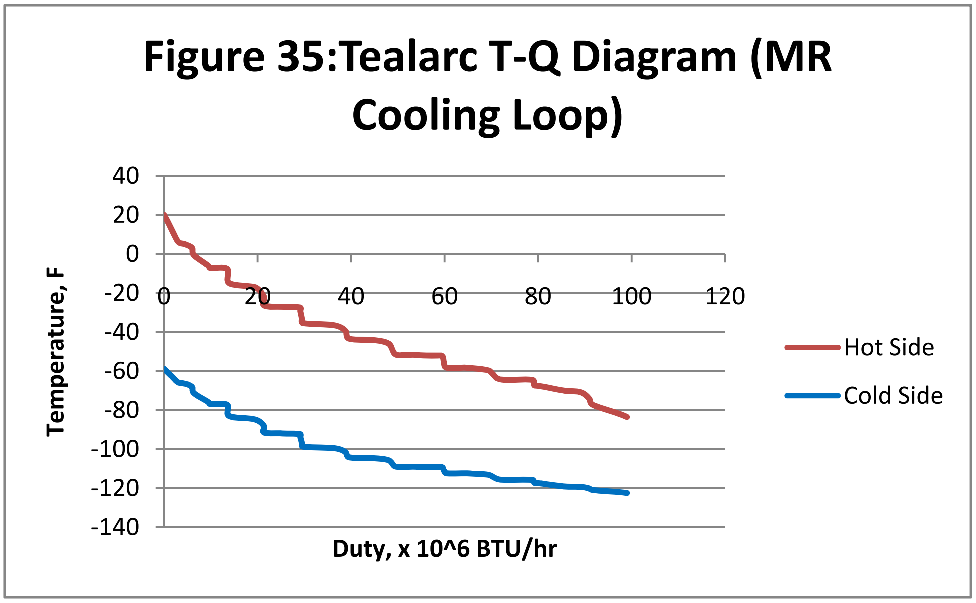


Figure 33:Tealarc (Liquefaction)
0
-50
-100
-150
-200
-250
-300
Temperature, F
0
20
40
60
80
100
120
Hot Side
Cold Side
Duty, x 10^6 BTU/hr





The next step is to optimize the compressor work. The refrigerant flow rate was first lowered in both stages. The total compressor work decreased in both cases. To further decrease the total compressor work, the temperature approach of the streams in both stages was optimized. This was first attempted by using an optimizer to minimize the compressor work by changing the outlet pressures of the valves. Using this method, the simulation would not converge. A solution was still being attained so the results were noted and the optimizer was taken out of the simulation. The results, outlet pressures of the valves, were entered into simulation without the optimizer and the simulation converged. The liquefaction stage valve pressure increased from 20

Psia to 40 Psia, the sub-cooling stage valve pressure increased from 50 Psia to 65 Psia and the refrigerant cooling stage valve pressure increased from 15 Psia to 20 Psia. The outlet temperatures of the hot streams are also adjusted to decrease the temperature approach. At this point, the compressor work was further decreased by further reducing the refrigerant flow rate. The refrigerant compositions and flow rates at the end of optimization are shown in table 18. The compressor works at the end of optimization are shown in table 19. Figures 36 to 39 show the T-Q diagrams after optimization is complete.

Table 18

Refrigerant Composition and Flow Rate

	Natural Gas Stage (%)	**Refrigerant Stage (%)**
Methane	80	5
Ethane	5	65
Propane	5	20
n-Butane	5	10
i-Butane	10	
Total flow rate (lbmol/hr)	31000	19000

Table 19

	Compressor Work (HP)	**Outlet Pressure (Psia)**
C1 (Stage 1)	41834	540
C2 (Stage 2)	7449	800
C3 (Stage 2)	16480	1300

Figure 36:Tealarc T-Q Diagram (Natural Gas Precooling)

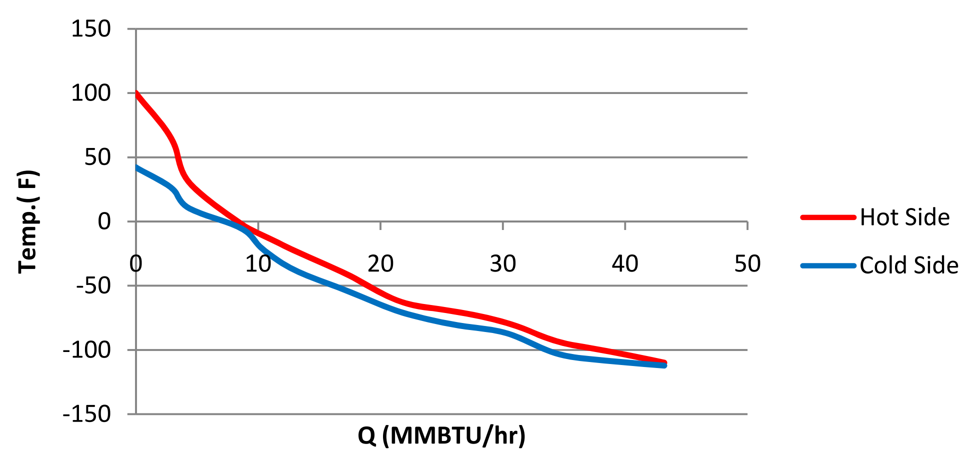
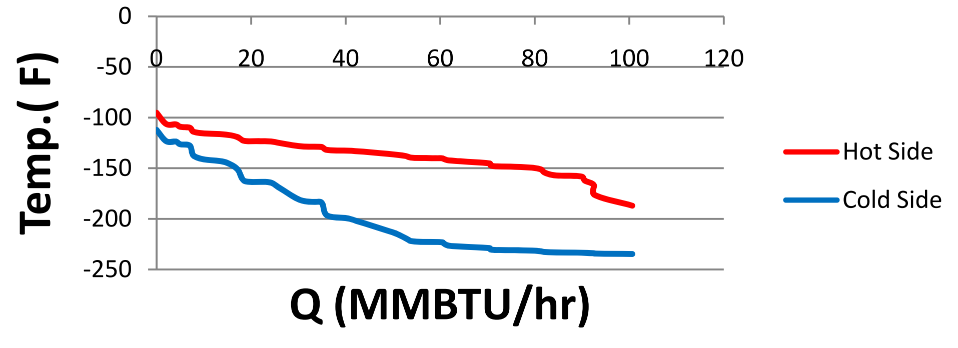


Figure 37:Tealarc T-Q Diagram (Natural Gas Liquefaction)



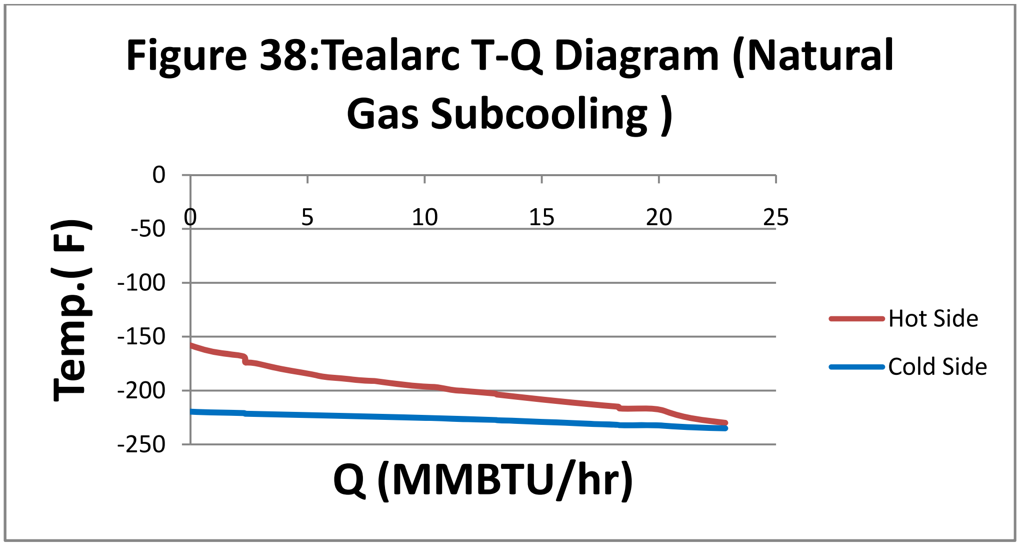


Figure 38:Tealarc T-Q Diagram (Natural Gas Subcooling)

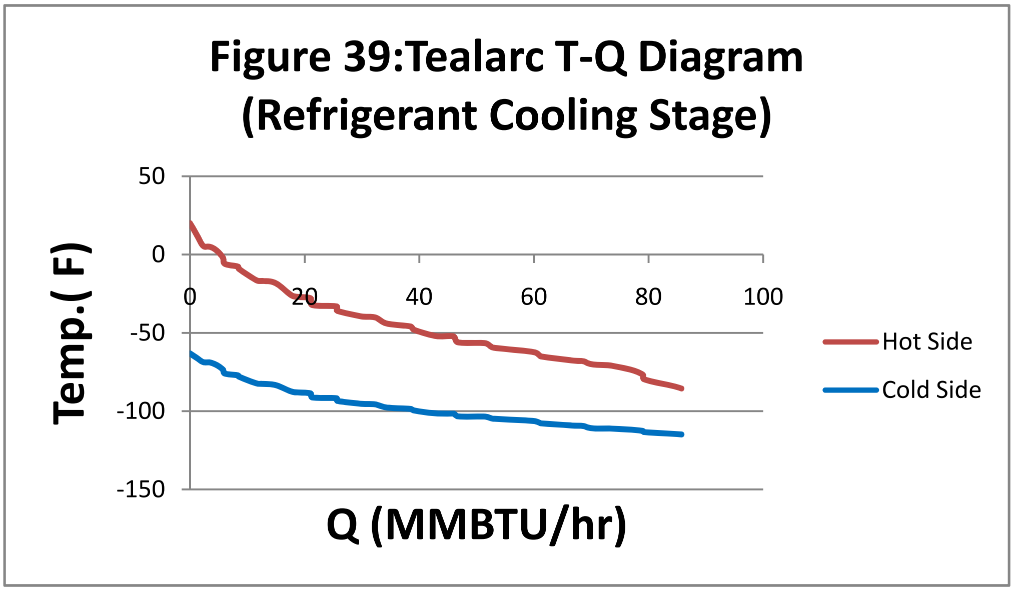


Figure 39:Tealarc T-Q Diagram (Refrigerant Cooling Stage)

The natural gas cooling stage refrigerant flow rate cannot be lower than 2990 lb-mol/hr in order to maintain a reasonable minimum temperature approach (MTA) in the pre-cooling heat exchanger. The liquefaction stage valve outlet pressure cannot go above 40 psia for the same reason. Also, to maintain a reasonable MTA in the sub-cooling stage heat exchanger the outlet pressure of the valve in this stage cannot go above 65 psia. Finally, to avoid temperature crossing in the refrigerant cooling heat exchanger, the outlet pressure of the valve in this stage cannot go above 20 psia. Making any further changes to the above mentioned parameters decreases the total compressor work but the resulting design is not feasible.

BP Self Refrigerated Process (US pat number: 6,564,578)

Description

This process was patented by the BP Corporation in 2002. It was considered a revolutionary method in LNG liquefaction because of its non-use of external refrigerants. It natural gas stream is flashed at cold temperatures to achieve a vapor and a liquid phase. The cold vapor stream is then used to cool the natural gas stream in the previous stage. This process is repeated about four to five times until the required LNG temperature is attained.

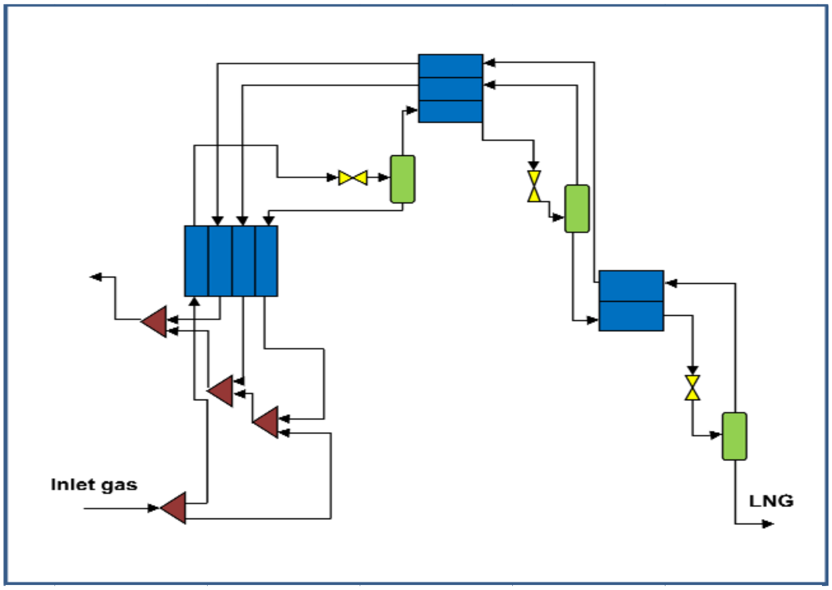


Figure 40: BP process

This process brings up the interesting subject that the natural gas, which is typically rich in methane can be the primary refrigerant in the process. This eliminates the need for the selection of adequate refrigerants and their compositions. It also eliminates the need for storage and recompression in a typical refrigeration cycle.

With no compressors and turbines, there is a reduced amount in the daily utility cost of the process. However, there is expected to be a great increase in the capital cost due to the number of heat exchangers required but the strict use of the PFHE significantly reduces that cost effect. The bulk of the cost comes from the cooling tanks used to separate the liquid and vapor phases.

The feed to LNG produced ratio is very low. Since the natural gas is required in the cooling, the amount of liquid preserved is only a fraction of the first initial separation. The rigid configuration of the PFHE's makes the process less flexible. Since each stage is dependent on the previous stage, production rates and temperatures cannot easily be altered without a total operations overhaul. Low capacity when compared to the sizes and magnitude of the equipment involved.

Simulation Method

The natural gas stream is initially split into two portions at a splitter. One portion is sent into the series of mixers where all the gas used in cooling is mixed together and reintroduced into the system if necessary or stored for future use. The process stream, NG, is what is supposed to be converted into natural gas. Hence, the splitter is configured to send about 70% of the natural gas for process and the rest for its cooling.

At the first heat exchanger, the stream NG is passed through each cell and it is cooled from 100F down to a lower temperature which is determined by its dimensions. There is not a substantial amount of cooling occurring in the heat exchangers until it the later expansion stages.

This expansion occurs in a valve with a pressure drop of about 250psia. This helps reduce the temperature down to about 51F. The expanded stream is then introduced into a flash tank which operates at colder temperature. The stream is then separated into a vapor and a liquid stream at this point with the vapor stream being reintroduced into the heat exchanger to help in the cooling process. At this point the natural gas stream is cooled to about -140F with about 90% of it in the liquid phase.

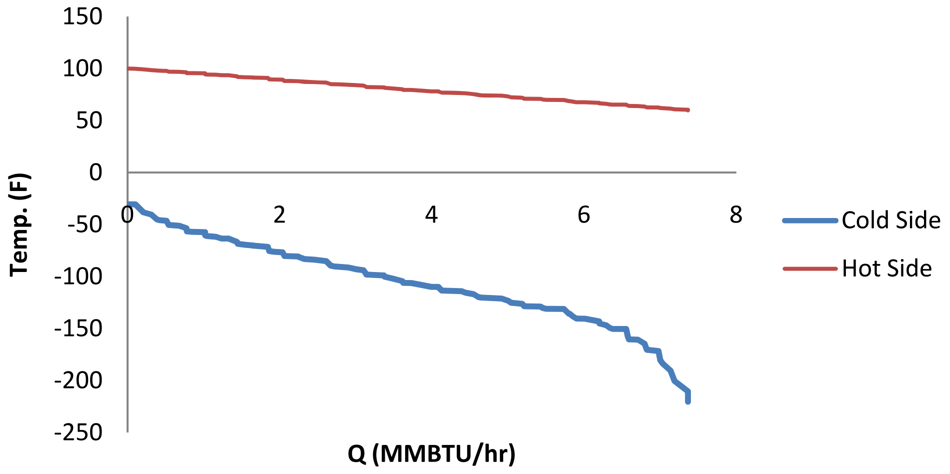
The cooled liquid phase is then sent into the second heat exchanger where it is supposed to be cooled by the cold vapor stream. However not a lot of cooling occurs at this point because there is not a great difference in the temperature of the cooling gas and the liquid natural gas. Also the amount of vapor available for cooling is also greatly reduced during the cooling in the flash tank as well. The same rule applies for the other heat exchangers in the as well. These sections can be called the 'conditioning' phase where any additional cooling obtained is just an additional advantage.

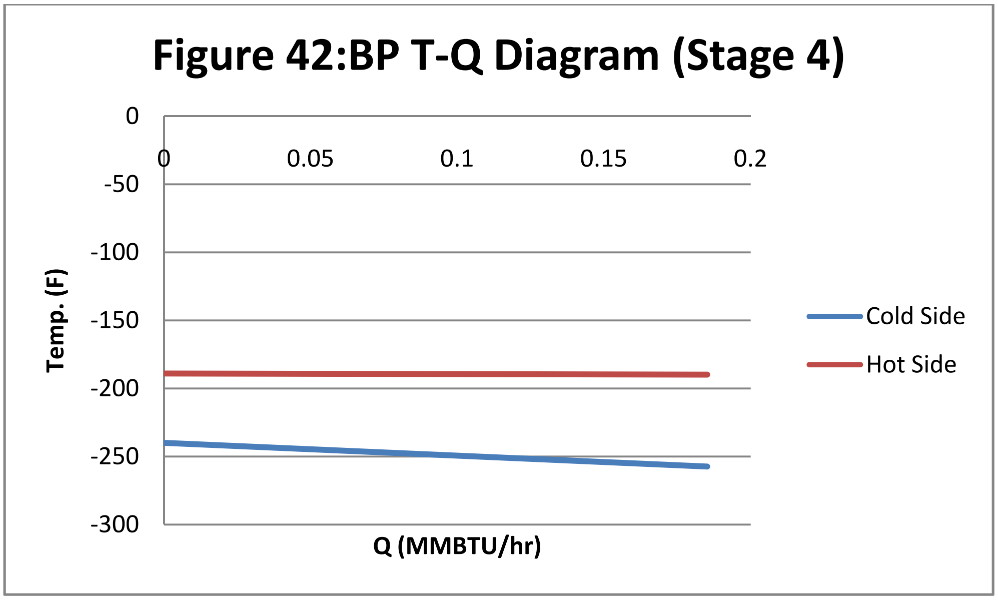
The second expansion valve reduces the pressure by another 150psia. This causes another temperature reduction before it is introduced into another flash tank. The separation in the flash tank leads to a cold vapor stream being input into LNG heat exchanger E2. The liquid portion is sent to the next heat exchanger before the process is repeated again until the desired temperature and pressure of -260F and 14.7psia is obtained in the last stage. It is clear that a bulk of the cooling occurs in the first heat exchanger, the valves and the flash tanks. This is due to limited amount of cooling vapor available as the number of stages is increases. If the amount of vapor is to be increased, there is a limited amount of LNG produced in the last stage.

Due to constant separation of the liquid and vapor phase, the amount of Liquefied Natural Gas obtained is always decreasing stage to stage. The optimizer in the simulation helps is set to maximize the amount of LNG obtained by changing the parameters of the splitter, on how much should be sent in cooling and the processing the LNG.

The efficiency of this process depends largely on the composition of the natural gas. The impurities like nitrogen and pentane could help in cooling the gas more significantly than is indicated in the T-Q diagrams below.

Figure 41:BP T-Q Diagram (Stage 1)





Stage 1 and 4 are the stages where most of the cooling occurs in heat exchangers. the curves are not close together which would indicate a low process efficiency as expected. Most of the cooling occurs in the expanders, valves and flash tanks.

An optimizer is inplace to maximize the amount of LNG that can be produced in the final stage by varyng the temperatures of the critical flash tanks and the heat exchanger cells. The fact that the feed has almost no impurities in its composition, most of the cooling is obtained only in the first and last stage.

Mustang Engineering Smart Liquefaction Process (Patent Number 7225636)

Description

Mustang Engineering claims that its process can be easily shutdown and moved to other areas very easily because of its simple design. Other established process cannot make this claim, which is one of the reasons most processes are not changed out or moved from place to place when needed. Exploring its simple design and less bulky equipment setup, makes this process important to be researched.

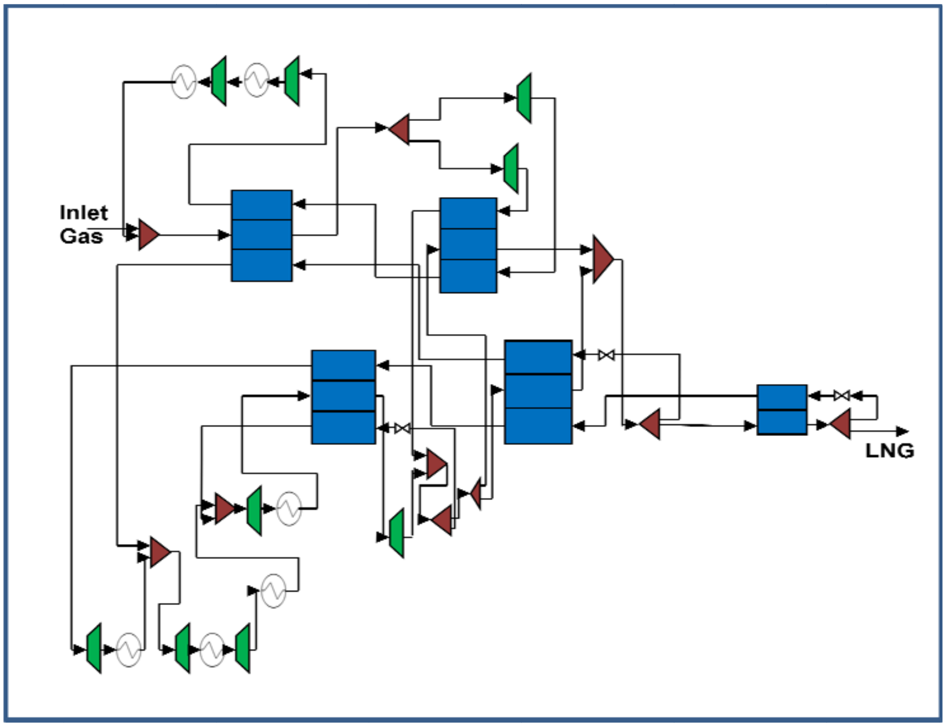


Figure 43: Mustang Engineering Process

The simple design of the process is what created the claim that it could be easily shut down and moved from place to place. Unfortunately, this process has lots of equipment that would utilize lots of resources to shut down and move. It would also have a high capital and utility cost associated with the equipment, yet all the equipment are all items that seem to be able to shut down and moved relatively easily.

It seems that the intention of this process is to be used on a floating barge for natural gas wells in oceans or off-shore. Mustang Engineering has a natural gas barge that it also advertises with this process, which would back up the idea of using it for wells in oceans.

This process has no external refrigerants. Instead, the process splits the natural gas stream and uses it to cool itself, which is an advantage because there are no external storage tanks needed for refrigerants. Unfortunately, the final liquefied natural gas is only a portion of the input gas, which is a disadvantage. This also requires more equipment to compress and expand the natural gas separated to be used as the refrigerant to allow it to reach temperatures where it can cool itself.

Simulation Method

The Mustang process I obtained from a patent. In the patent there were a few specifications for equipment, so I set up the simulation in its entirety and set in the specifications that I knew from the patent. Other specifications for equipment, I determined based on what I already knew about the process.

Once all the specifications were entered, I ran the simulation to make sure that it worked. Any adjustments that needed to be made to allow the simulation to run properly were made and the corrections were obtained from output filed for the equipment not running properly. Once all the equipment was running properly, I began to make adjustments to the temperature in the LNG heat exchangers. I ran the simulations until the condition that the hot streams exiting colder caused the cold streams to exit hotter was met.

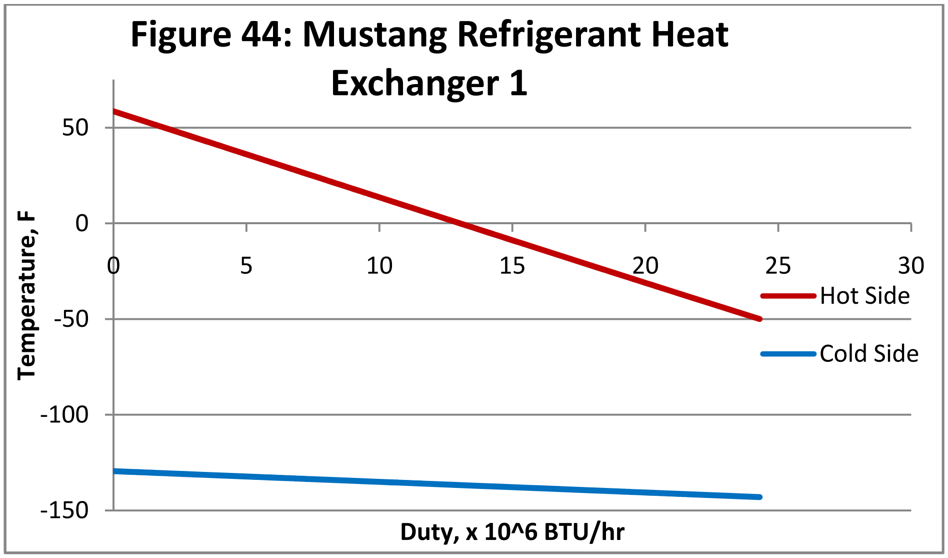


Figure 44: Mustang Refrigerant Heat Exchanger 1

I found that it was much easier to accomplish this by splitting the initial LNG heat exchanger into two. The top heat exchanger cooled one hot stream with two cold streams, and the lower heat exchanger cooled on hot stream with two cold streams.

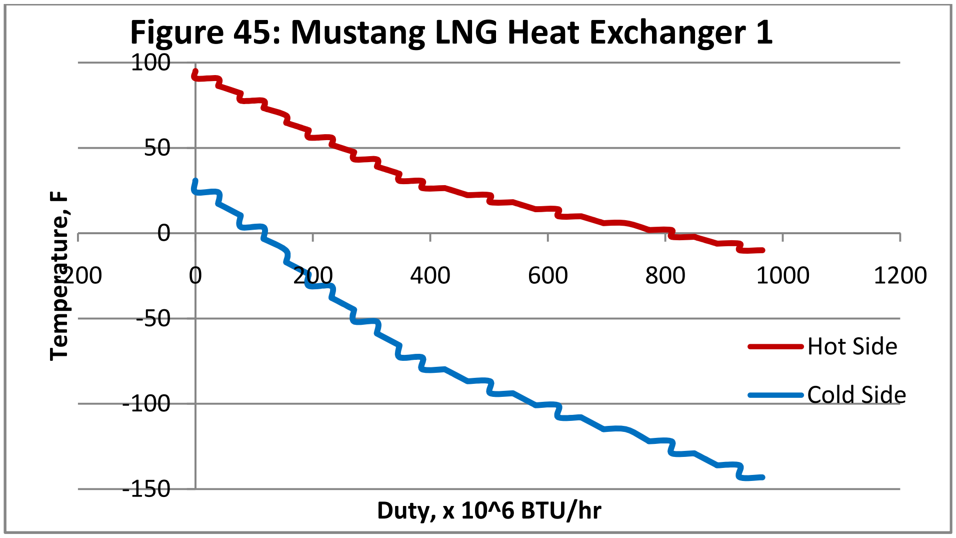


Figure 45: Mustang LNG Heat Exchanger 1

Once this was accomplished it was then time to calculate the log mean temperature difference between the hot and cold streams. A calculator was added for each LNG heat exchanger to accomplish this. A second calculator was then added to calculate the areas between the cells of the heat exchanger using the log mean temperature difference.

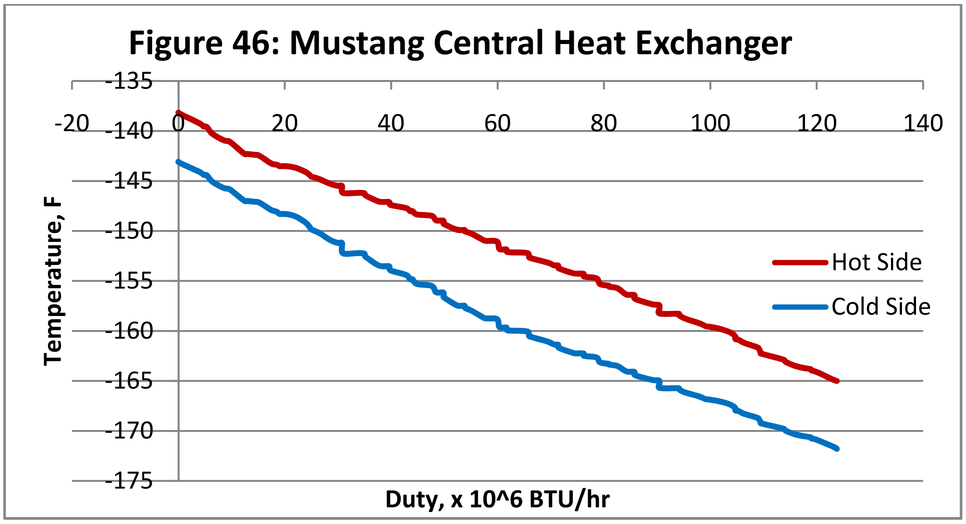
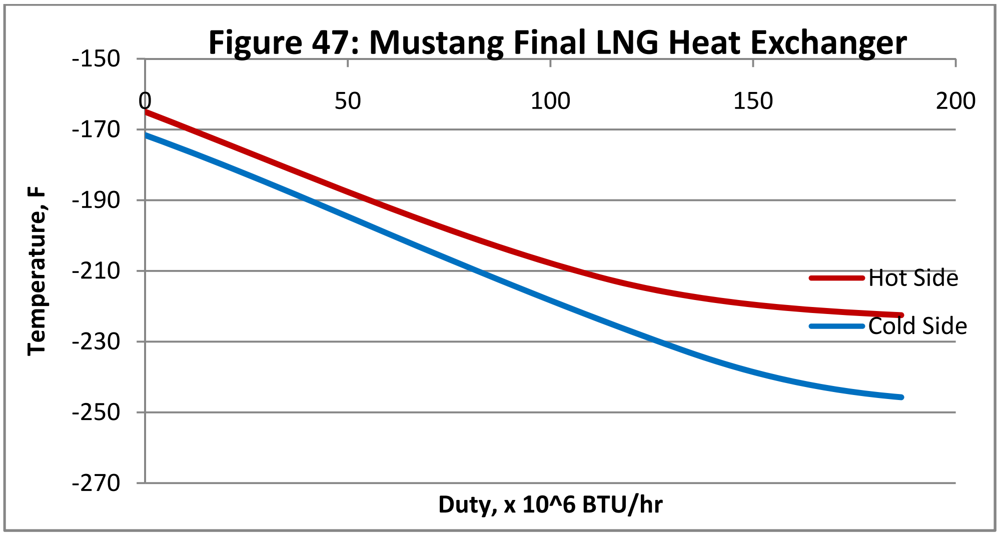


Figure 46: Mustang Central Heat Exchanger

Once the log mean temperature differences and the areas were calculated, controllers were then added to the simulation. These controllers were added to make the areas the same between the cells of the heat exchanger. The specifications included in the controllers were changing the specifications of splitters to change the flow rate of a stream in a cell to help or not help so much with cooling, changing pressured in valves to change the temperature of a stream entering a cell, and the temperatures of the cells themselves. All of these specifications helped to create a uniform area between each of the cells. Adjustments needed to other equipment based on the changes caused by the controller were then made and the simulation run again, to make sure everything worked.



In the Mustang process, most of the specifications for processes came from the patent from which the process was found. The patent had a range of temperature and pressure values for roughly half of the streams. These temperature and pressure ranges were used to determine the streams for which the values were available. From here, the various other temperature and pressure ranges for other streams were relatively easily determined. To be listed below is a table, paired with the PFD from the patent, which gave relative temperature and pressure ranges. The original PFD will be attached as an appendix.

The values that were chosen for the final values were also chosen because they were the ones that allowed the simulation to run properly, even after a few runs, without continually messing up.

Variables that were able to be adjusted to help the simulation run were splitter specifications, small changes to compressor and expander pressures, small changes to LNG heat exchanger outlet temperatures, and valve outlet pressures. These changes could mostly be varied slightly due to the specifications given in the patent and in order to keep temperatures from crossing and heat from flowing in the wrong direction.

In order to control the areas, the controller used varied splitter specifications and valve pressures. The refrigerant used was a split stream from the natural gas. When it came time to optimize the compressor work, since the pressures were specified in the patent, I did not change those. The refrigerant composition could not be changed because it was a split stream from the natural gas. The other variables such as splitters, that would reduce the flow rate into the compressor, and the pressures of valves, could not be changed, either because they were already controlled by other controllers or because any change in pressure would cause crossing streams or heat flowing in the wrong direction.

ConocoPhillips Simple Cascade

Description

This process consists of 3 refrigeration cycles. The first cycle consists of 3 refrigeration stages and the refrigerant used is propane. The second cycle consists of 5 refrigeration stages and the refrigerant used is ethylene. The natural gas feed stream is combined with one or more recycle streams (compressed open methane cycle gas streams) at a variety of locations in the second cycle thereby producing a liquefaction stream. The liquefaction stream is condensed in the last refrigeration stage of the second cycle to produce LNG. The third refrigeration cycle is an open-end methane cycle which uses a portion of the feed gas as a source of methane to be used in the refrigeration cycle and includes a multi-stage expansion cycle to further cool and reduce the pressure of the LNG from the second refrigeration cycle. (Ransbarger 2008, p. 3)

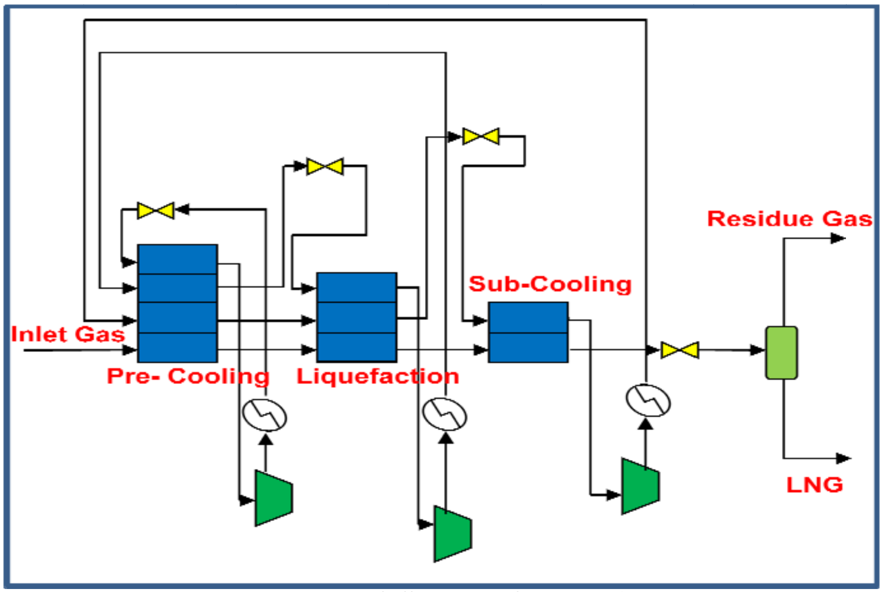


Figure 48: ConocoPhillips Simple Cascade Process

Simulation Method

All the stages in this process use pure refrigerant. Propane is used in the first stage (Pre-cooling stage), ethylene is used in the second stage (liquefaction stage) and methane is used in the last stage (liquefaction stage). The initial flow rates used in each stage are shown in the table below.

Table 20	
	Total flow rate (lbmol/hr)
Pre-cooling Stage	60000
Liquefaction Stage	20000
Sub-cooling stage	40000

These parameters were entered into the simulation, natural gas target temperatures were set and the outlet temperatures of all other hot streams were set. There was no temperature crossing observed. The compressor works before optimization are shown in table 21. The T-Q diagrams at this point are shown in Figures 49 to 51.

Table 21		
	Compressor Work (HP)	**Outlet Pressure (Psia)**
C2 (Propane Stage)	40311	188
C3 (Ethylene Stage)	33664	304
C4 (Methane Stage)	47415	478

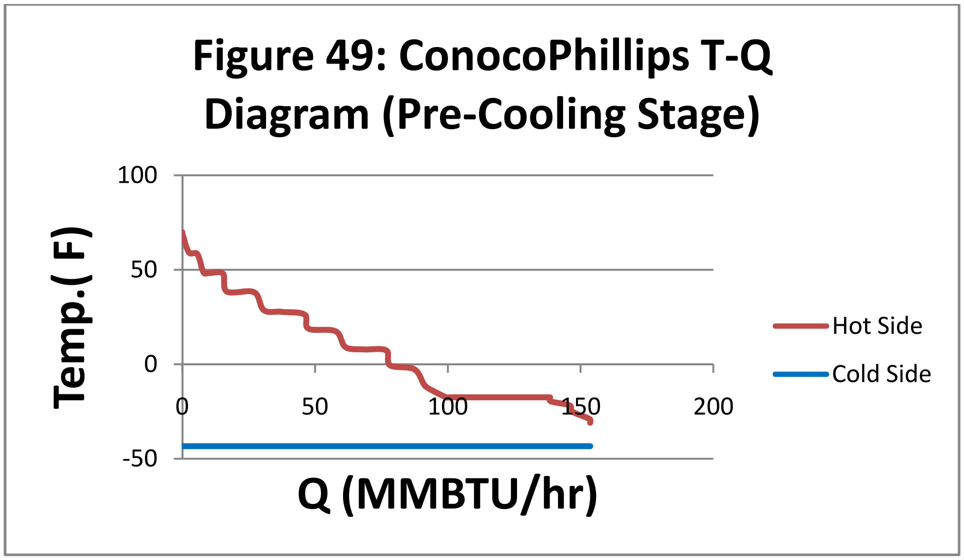


Figure 50: ConocoPhillips T-Q Diagram (Liquefaction Stage)

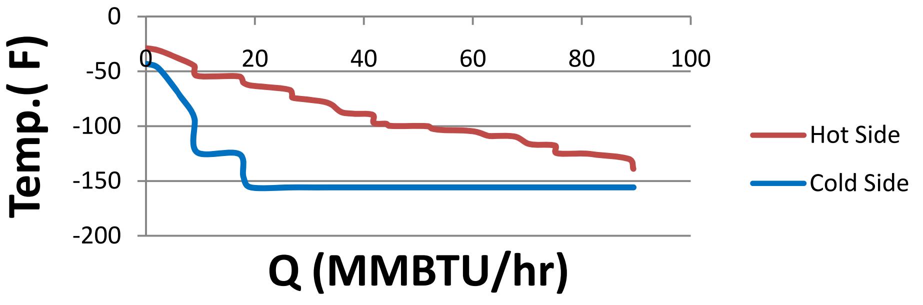
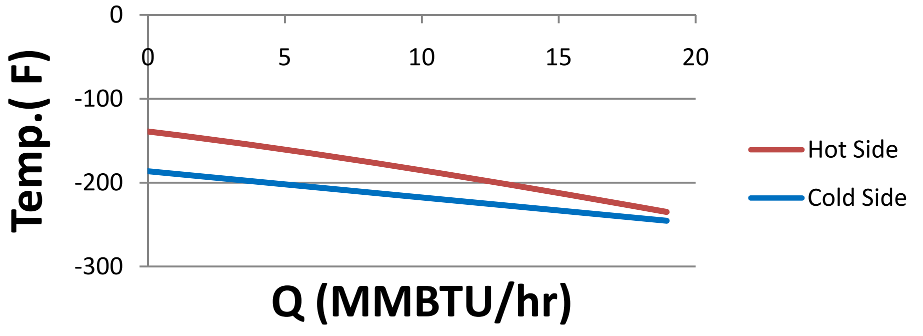


Figure 51: ConocoPhillips T-Q Diagram (Sub-Cooling Stage)

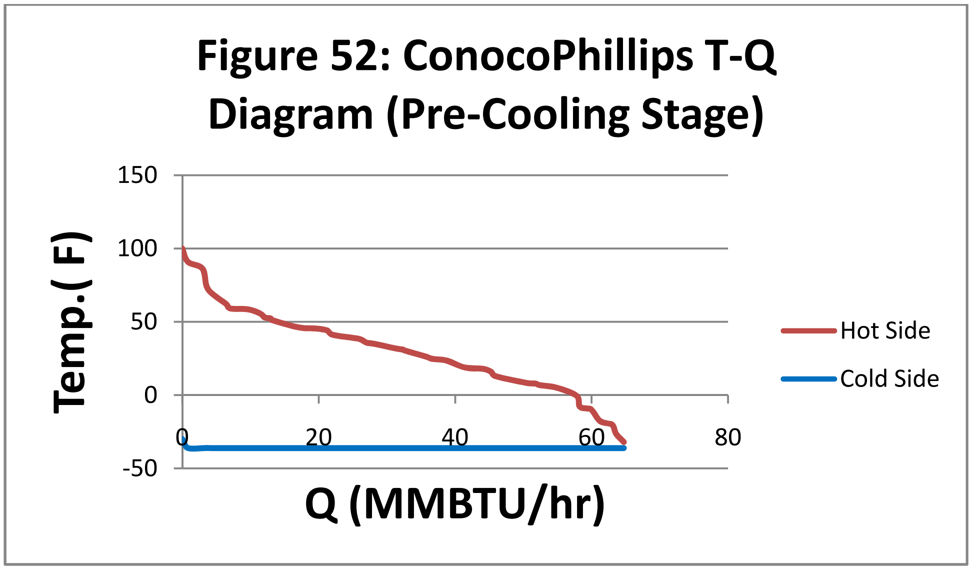


The cold side curve of the pre-cooling stage has a slope of zero because the refrigerant stream is completely liquid. The liquefaction and sub-cooling stage T-Q diagrams deviate from this because there is some gas present in the stream. This is addressed in the optimization stage.

The next step is to optimize the compressor work. The refrigerant flow rate was lowered in all stages. This was done using an optimizer. The simulation did not converge but the calculated flow rates were noted. The flow rates were entered into the simulation after taking out the optimizer and the simulation converged. The total compressor work was decreased by this approach. To further decrease the total compressor work, the temperature approach of the streams in all the stages were optimized. This was first attempted by using an optimizer to minimize the compressor work by changing the outlet pressures of the valves. Using this method, the simulation would not converge. After several trial and error exercises it was found that only the ethylene and methane stage valves could be handled by the optimizer. The propane stage valve's optimum outlet pressure was found to be 17.5 psia. If anything above that was selected the minimum temperature approach (MTA) would not be reasonable. The T-Q diagrams and refrigerant phase of all the stages were reviewed after every trial and error exercise to make sure the slope of the cold side curve was as close to zero as possible. The optimal flow rates and outlet valve pressures are shown in table 22. Figures 52 to 54 show the T-Q diagrams after optimization is complete. As can be seen from the T-Q diagrams, the issue of the cold side slope was addressed and fixed. The sub-cooling stage cold side slope is not completely zero but it is a significant improvement from that shown in Figure 51.

Table 22

	Total flow rate (lbmol/hr)	Valve pressure (psia) before optimization	Valve pressure (psia) after optimization
Pre-cooling Stage	60000	14	17.5
Liquefaction Stage	20000	14	27
Sub-cooling stage	40000	14	259



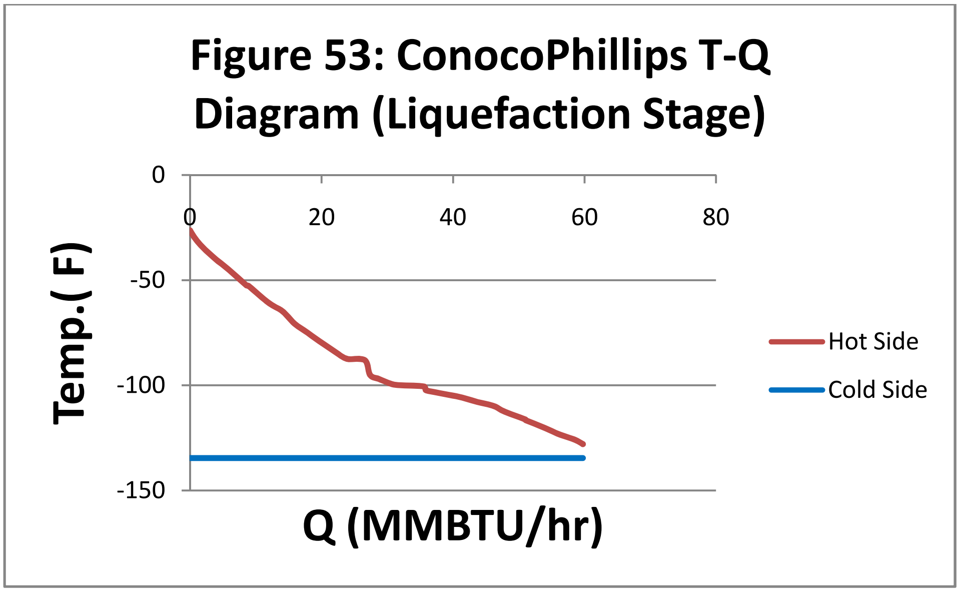


Figure 53: ConocoPhillips T-Q Diagram (Liquefaction Stage)

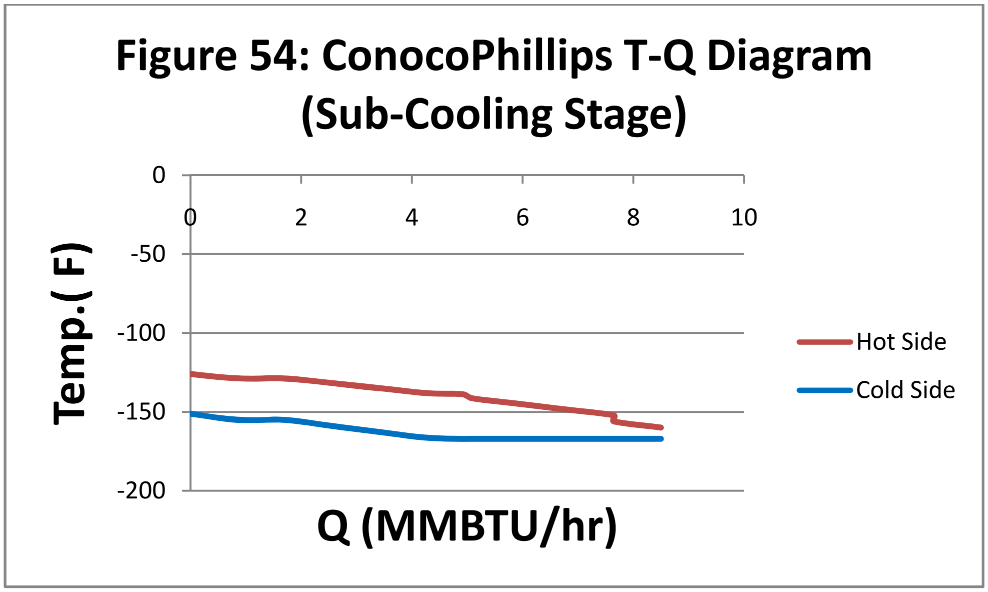


Figure 54: ConocoPhillips T-Q Diagram (Sub-Cooling Stage)

The flow rate of the sub-cooling stage refrigerant cannot be lower than 20000 lb-mol/hr because the fluid after the expander would contain more gas. This makes the cold side curve of the T-Q diagram deviate from the zero slope trend it should show. It also causes the heat transfer area of the heat exchanger to increase thereby increase the capital cost of the process. The same is also true of the pre-cooling stage refrigerant. For this refrigerant the flow rate cannot be lower than 11950 lb-mol/hr.

Mixed Fluid Cascade Process

Description

This process involves three distinct stages: Pre-cooling, Liquefaction and Sub-cooling. Each stage is controlled by three separate mixed refrigerant cycles. The mixed refrigerants are composed of methane, ethane, propane and nitrogen at different compositions. This process is set exactly like a classical simple cascade except with one major difference, the mixed refrigerant. The mixed refrigerant improves flexibility and thermodynamic efficiency. This process is fairly new and without any industrial references but the idea is built on well known and established elements[3]. Its capacity is stated at about 8 MTPA.

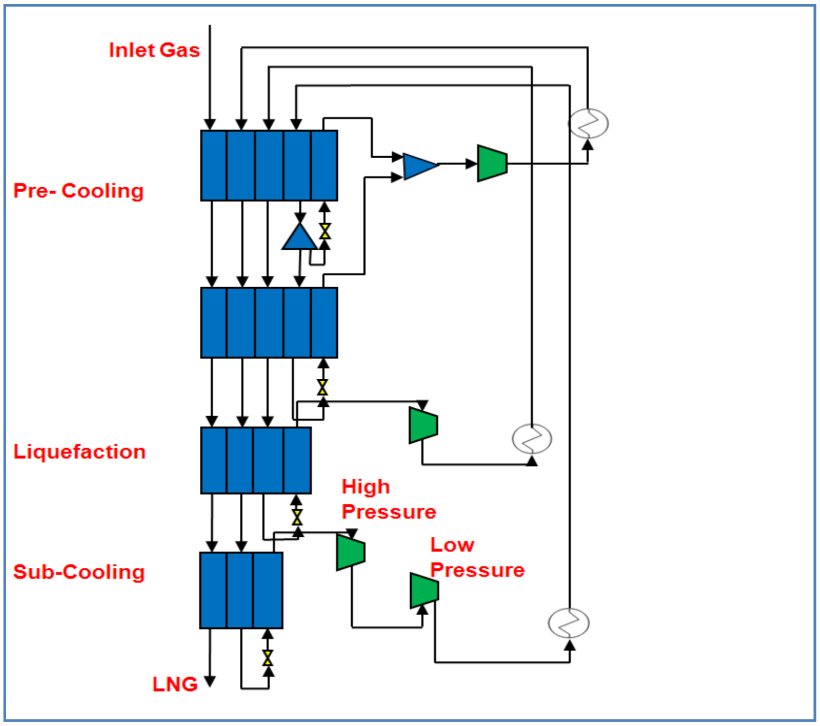


Figure 55: Mixed Fluid Cascade Process

Simulation Method

There was no information found on the refrigerant composition of any of the stages. The initial refrigerant compositions used in each stage were guessed. The guiding principle for this guess was that the later stages would need refrigerants that boiled at very low temperatures close to the target temperature of natural gas in the stage of interest. The refrigerants in the earlier stages would not need to boil at such low temperatures. The compositions used are shown in the table below along with the first guess of total flow rate.

Table 23: Refrigerant Composition and Flow Rate

	Pre-cooling		Liquefaction	Sub-cooling
	Stage 1	**Stage 2**		
Methane	10	10	80	80
Ethane	28	28	12	10
Propane	60	60	3	7
Nitrogen	2	2	5	3
Total flow rate (lbmol/hr)	16500	13000	23000	20000

These parameters were entered into the simulation and temperature crossing was observed in all the stages. The flow rates of all the stages were adjusted and the temperature crossing was alleviated. The flow rates chosen are shown in table 24.

Table 24

		Total flow rate (lbmol/hr)
Pre-cooling Stage	**1**	15300
	2	13000
Liquefaction Stage		22000
Sub-cooling stage		19000

At this point the simulation was working in that the natural gas target temperatures were being achieved without any temperature crossing. The compressor works before optimization are shown in table 25. The T-Q diagrams at this point are shown in Figures 56 to 59.

Table 25

	Compressor Work (HP)	Outlet Pressure (Psia)
C1 (Pre-cooling 1)	15165	300
C2 (Pre-cooling 2)	11925	650
C3 (Liquefaction)	75489	3500
C4 (Sub-cooling)	25085	4000

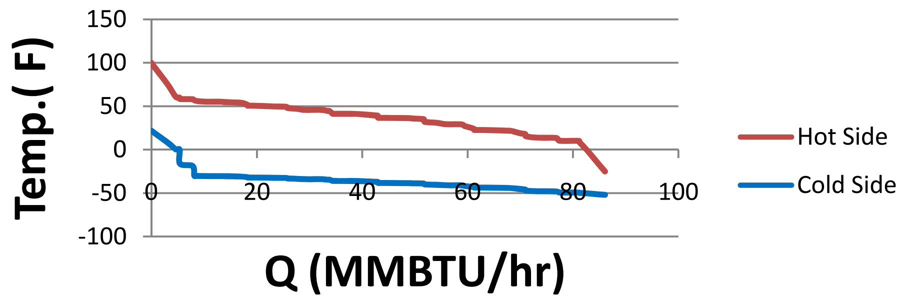


Figure 56: MFCP T-Q Diagram (Precooling Stage 1)

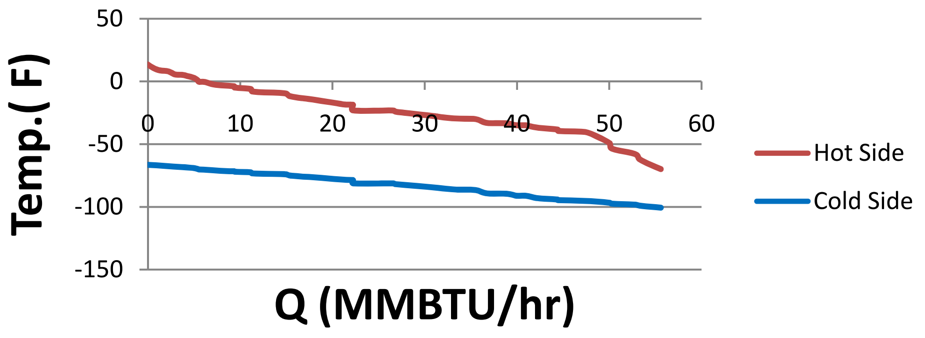


Figure 57:MFCP T-Q Diagram (Precooling Stage 2)

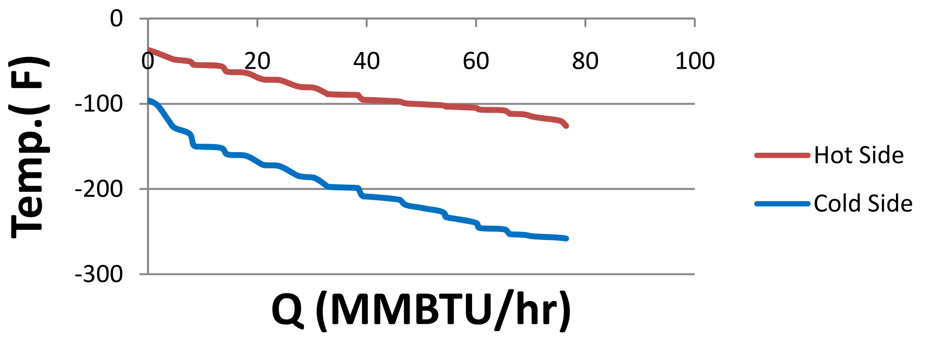
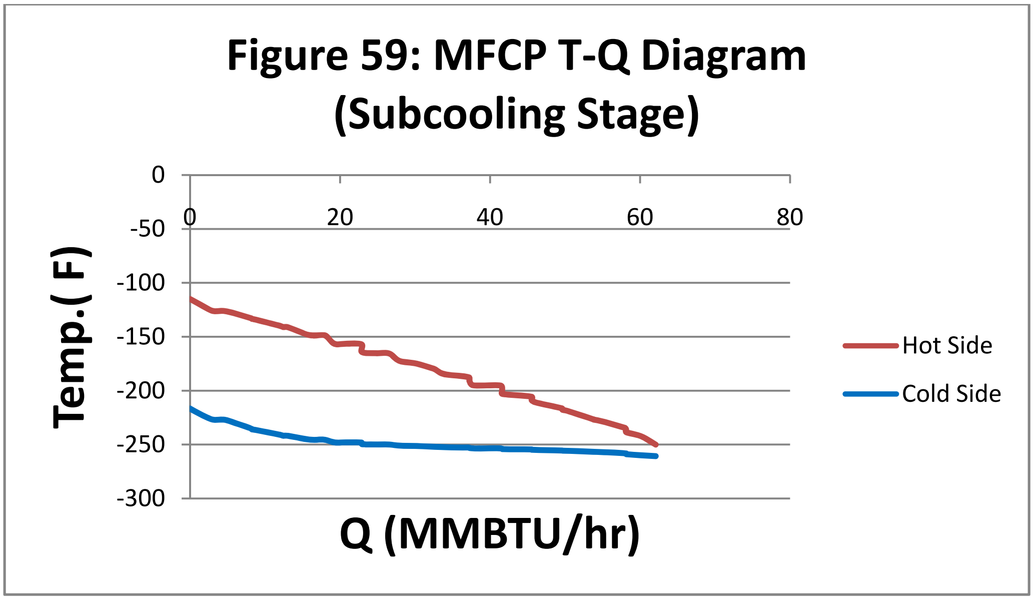


Figure 58: MFCP T-Q Diagram (Liquefaction Stage)



The next step is to optimize the compressor work. The refrigerant flow rates of all the stages were reduced and the compressor work went down by a significant amount. There was one stage that differed. It was found that the optimal flow rate of the refrigerant in the second pre-cooling stage was the same as the initial guess. The compressor works were further optimized by decreasing the temperature approach of the streams in all the stages. This was done by adjusting the outlet temperatures of the hot streams and adjusting the outlet pressure of the valves in each stage. The outlet temperature of the natural gas is not changed as there is a set target. The last step taken to optimize the simulation is to use an optimizer to minimize the total compressor work by changing the outlet pressure of all the compressors. At this point optimization is at a stopping point. The compressor works at the end of optimization are shown in table 26. The flow rates and valve pressures are shown in table 27. Figures 60 and 63 show the T-Q diagrams after optimization is complete.

Table 26

	Compressor Work (HP)	**Outlet Pressure (Psia)**
C1 (Pre-cooling 1)	11242	300
C2 (Pre-cooling 2)	10002	650
C3 (Liquefaction)	17812	3500
C4 (Sub-cooling)	22635	4000

Table 27

		Total flow rate (lbmol/hr)	Valve pressure (psia) before optimization	Valve pressure (psia) after optimization
Pre-cooling Stage	**1**	15300	32	50
	2	13000	15	30
Liquefaction Stage		22000	10	320
Sub-cooling stage		19000	15	20

Figure 60: MFCP T-Q Diagram (Precooling Stage 1)

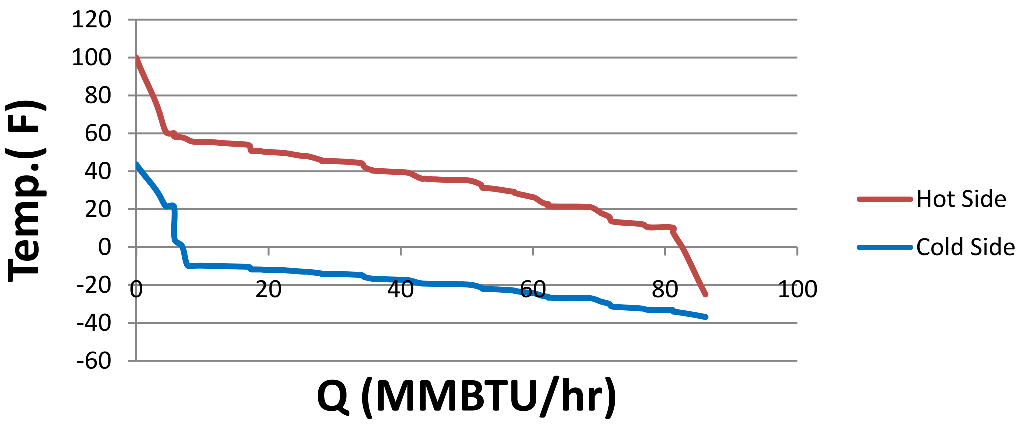
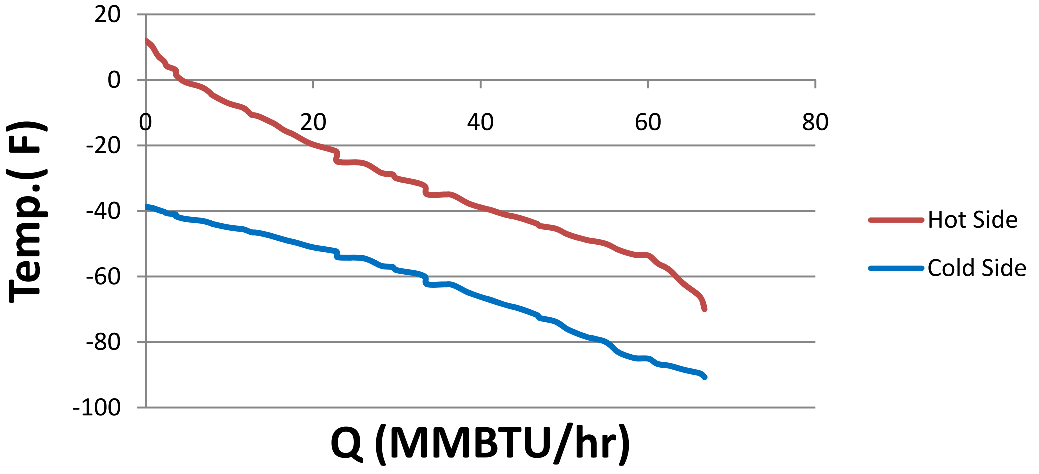
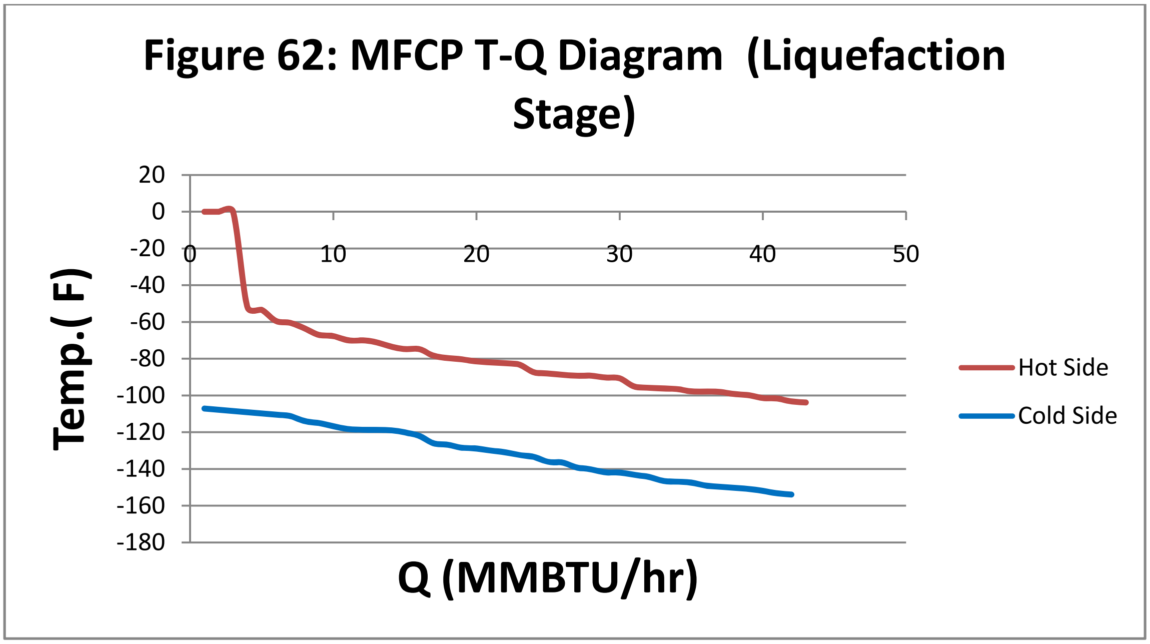
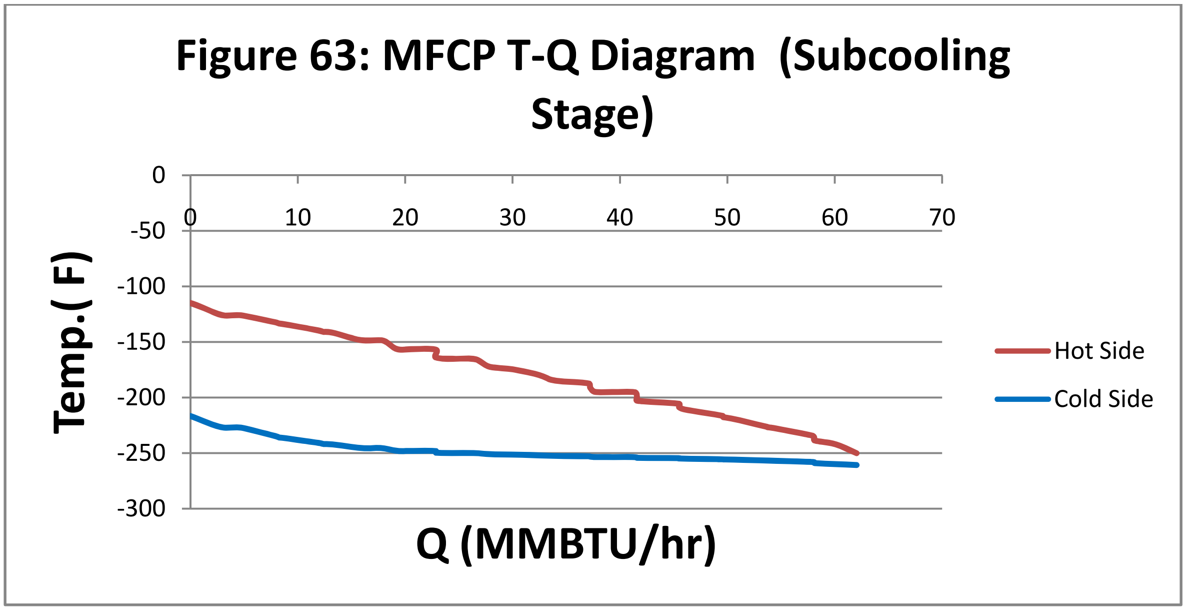


Figure 61: MFCP T-Q Diagram (Precooling Stage 2)







The optimal flow rate of the pre-cooling stage 1 refrigerant is 15300 lb-mol/hr. If it is lowered, the work of the pre-cooling stage 1 compressor reduces but that of the pre-cooling stage 2 and the liquefaction stage increases. The optimal flow rate of the pre-cooling stage 2 refrigerant is 13000 lb-mol/hr. If it is lowered, the work of the liquefaction stage compressor decreases but that of the pre-cooling stage 2 increases. If the flow rates of liquefaction stage and sub-cooling stage refrigerant are reduced, temperature crossing occurs in their respective heat exchangers.

CO_2 Pre-cooled Linde Process

Description

This process is similar to the MFCP except Carbon dioxide is used as the sole refrigerant in the pre-cooling stage. The reason for this is safety. Storing hydrocarbons is an important safety issue, especially in an offshore environment where the area is confined. The pre-cooling stage uses the most hydrocarbons so it is the first choice to be replaced by carbon dioxide. This is the best and only choice to make as attempting to operate the liquefaction or sub-cooling stage using carbon dioxide would produce dry ice as the freezing point of carbon dioxide is -69.88°F. (Bauer 2008, 2) The capacity of this process is estimated at about 8 MTPA.

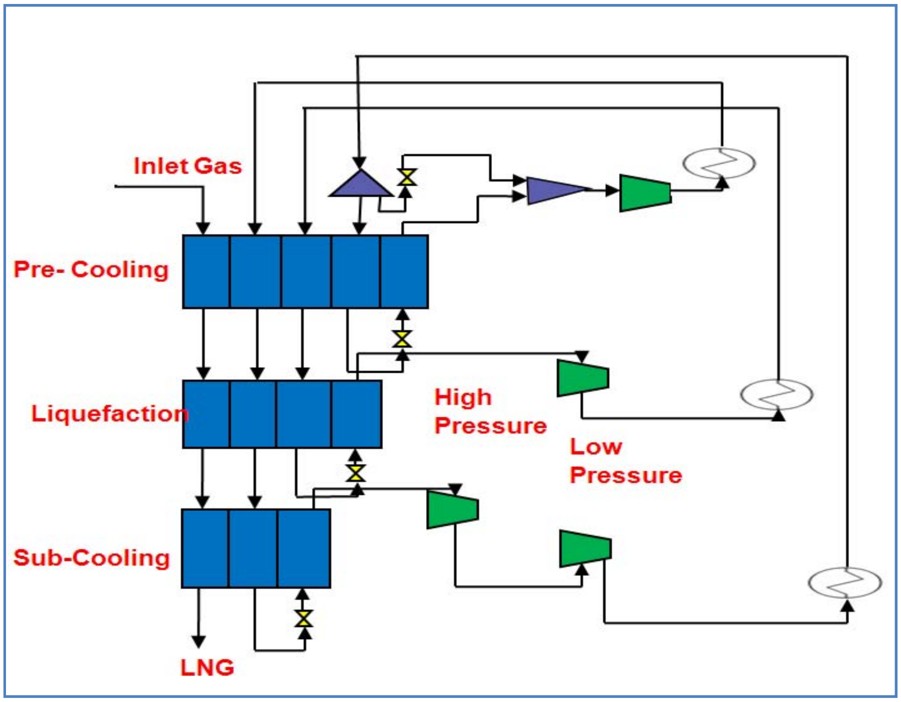


Figure 64: CO_2 Pre-cooled Mixed Fluid Cascade

Simulation Method

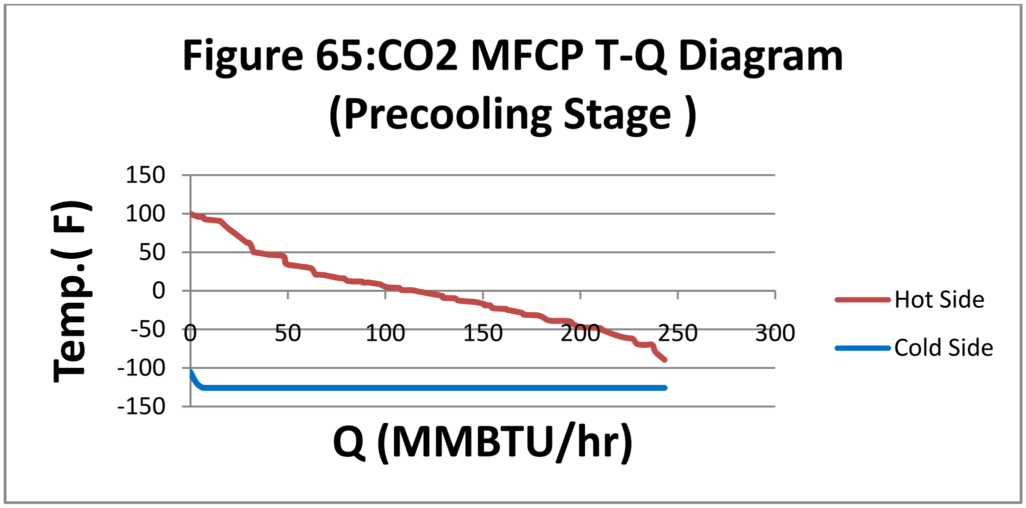
The pre-cooling stage refrigerant is pure carbon dioxide. There was no information found on the refrigerant composition of any of the other two stages. The initial refrigerant compositions used in the last two stages were guessed. The guiding principle for this guess is the same as that for the MFCP. The compositions used are shown in the table below along with the first guess of total flow rate.

Table 28			
	Pre-cooling	Liquefaction	Sub-cooling
Methane		51	80
Ethane		9.17	8
Propane		9.17	5
Carbon dioxide	1		
Nitrogen		30.58	7
Total flow rate (lbmol/hr)	40000	26000	24000

This process was simulated using the MFCP simulation as a starting point therefore there was no temperature crossing or any other problems except in the pre-cooling stage. This problem arose because of the change from a mixed refrigerant to a pure refrigerant. The temperature crossing in the first stage was fixed by adjusting the carbon dioxide flow rate and inlet pressure.

At this point the simulation was working in that the natural gas target temperatures were being achieved without any temperature crossing. The compressor works before optimization are shown in table 29. The T-Q diagrams at this point are shown in Figures 65 to 67.

Table 29		
	Compressor Work (HP)	Outlet Pressure (Psia)
C1 (Pre-cooling 1)	69500	450
C2 (Liquefaction)	39761	900
C3 (Sub-cooling 1)	13323	1500
C4 (Sub-cooling 2)	8118	5000



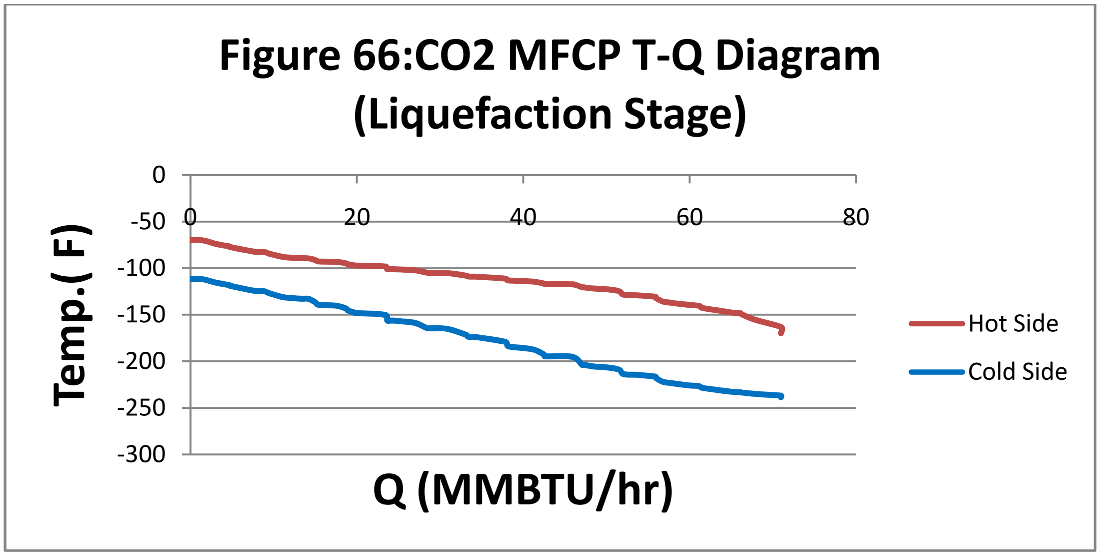


Figure 66:CO2 MFCP T-Q Diagram (Liquefaction Stage)

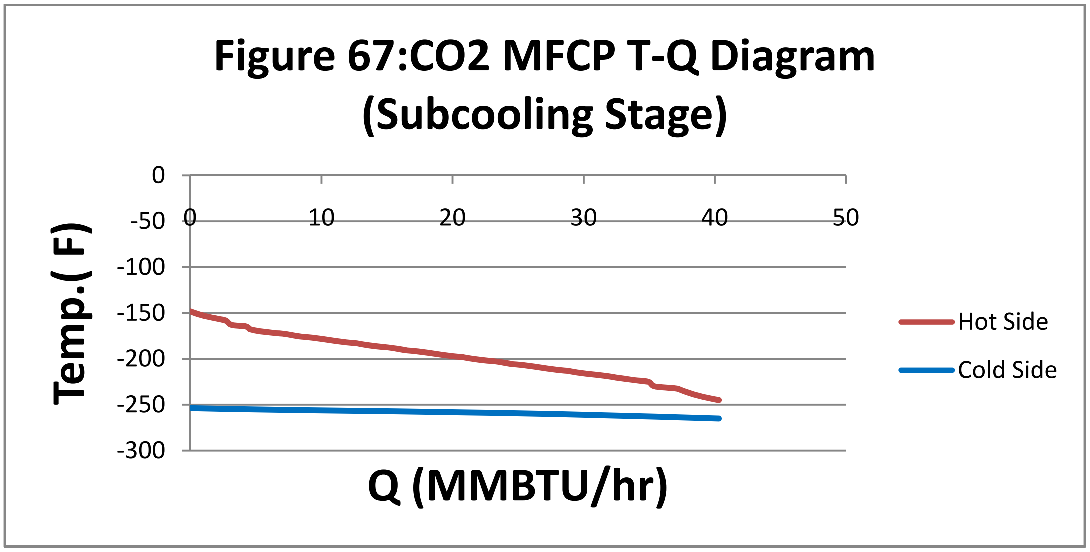


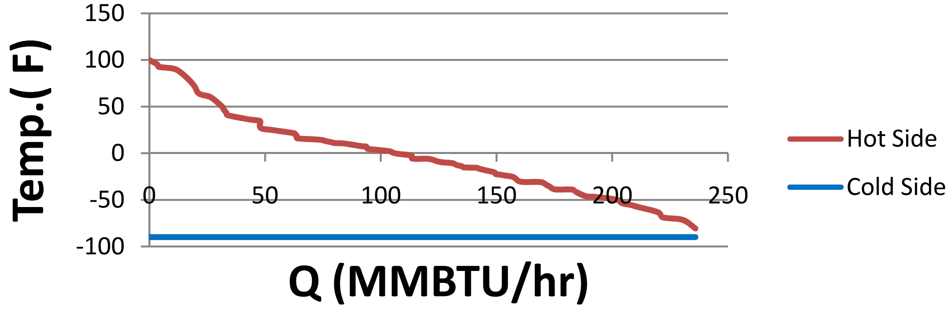
Figure 67:CO2 MFCP T-Q Diagram (Subcooling Stage)

The next step is to optimize the compressor work. The minimum approach temperatures (MAT) of the streams in all the stages were adjusted in order to minimize the total compressor work. This was done by adjusting the outlet temperatures of the hot streams and adjusting the outlet pressure of the valves in each stage. The outlet temperature of the natural gas is not changed as there is a set target. The refrigerants were also adjusted but only adjustment of the pre-cooling stage refrigerant was successful in reducing the compressor work. The optimal flow rate of the liquefaction stage refrigerant was found to be the initial guess, 26000 lb-mol/hr. Any decrease in this value caused temperature crossing in the liquefaction stage heat exchanger. Temperature crossing was also observed in the sub-cooling stage heat exchanger when the sub-cooling refrigerant flow rate was decreased from its optimal value of 24000 lb-mol/hr. The compressor works at the end of optimization are shown in table 30. The flow rates and valve pressures are shown in table 31. Figures 68 to 70 show the T-Q diagrams after optimization is complete.

Table 30		
	Compressor Work (HP)	**Outlet Pressure (Psia)**
C1 (Pre-cooling 1)	31838	450
C2 (Liquefaction)	21305	900
C3 (Sub-cooling 1)	12527	1500
C4 (Sub-cooling 2)	8747	5000

Table 31			
	Total flow rate (lbmol/hr)	**Valve pressure (psia) before optimization**	**Valve pressure (psia) after optimization**
Pre-cooling stage	38000	15	45
Liquefaction stage	26000	30	105
Sub-cooling Stage	24000	15	35

Figure 68:CO2 MFCP T-Q Diagram (Pre-Cooling Stage)



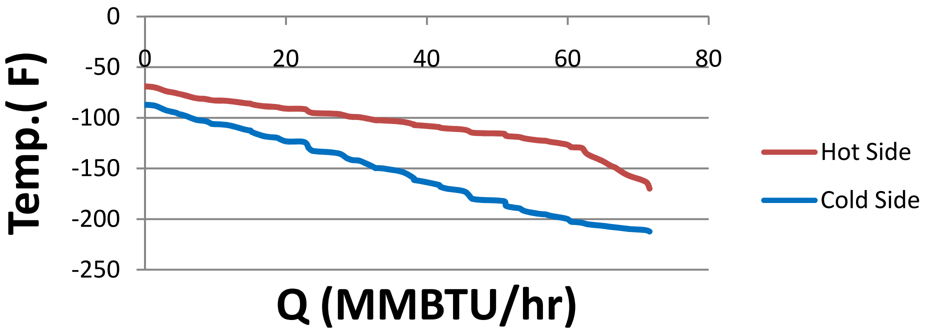


Figure 69:CO2 MFCP T-Q Diagram (Liquefaction Stage)

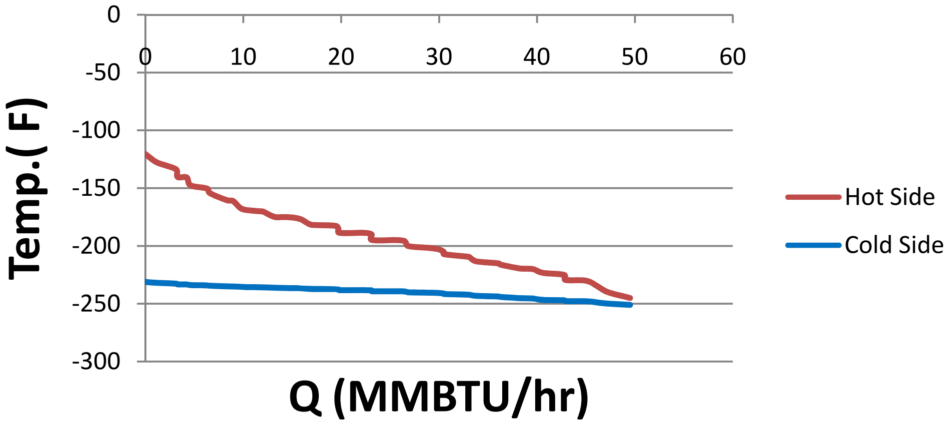


Figure 70:CO2 MFCP T-Q Diagram (Sub-Cooling Stage)

Economic Analysis

Selecting the appropriate process to cater to specific needs could be very challenging. However, the economic effect the process is expected to produce usually gets higher priority in the final selection stages. Not only is the cost per capacity produced relevant but also the utility costs associated with running the process on a daily basis.

Due to the unavailability of specific process information, a basic model was developed to compare all processes evenly. The total economic life expectancy of each project was set at twenty years. This enables any fixed equipment cost to be amortized on a yearly basis. Pricing each piece of equipment used in a liquefaction process was based on the documented maximum operating capacity of one train. It is assumed that whoever selects a process is ready to operate at its maximum capacity. This would mean purchasing large enough compressors, tanks and heat exchangers to meet the maximum design. The two websites primarily used in pricing the equipment were Matche.com and Plant Design and Economics for Chemical Engineers equipment website (both listed in the references section). Equation (6) below shows the capital cost per year.

$$Capital\ Cost\ \left(\frac{\$}{year}\right) = \left(\frac{Total\ Equipment\ Cost(\$)}{Economic\ Life}\right) \tag{6}$$

Another important factor in each process is the amount of water and electricity needed to process a selected minimum amount on LNG. This amount of ‘energy’ obtained to produce this minimum amount is then used as basis for which larger energy requirements are obtained. The average cost of water and electricity throughout the whole United States in 2007 was used in this analysis. The results produced by the simulation were calculated on an hourly basis before being converted to a yearly basis. Equation (7) is used in this determination.

$$Energy\ Cost\ \left(\frac{\$}{year}\right) = \frac{Energy\ Cost\ at\ Min.\ Capacity\left(\frac{\$}{yr}\right)}{Min.Capacity(MTPA)} * Capacity(MTPA) \tag{7}$$

Since both equations (6) and (7) are on a cost per year basis, their direct summation would yield an expected energy and capital cost per year for each process. Since each process has different maximum operating capacities, it becomes necessary to compare the cost per ton of the product based on only a maximum train capacity. This relationship becomes advantageous because it limits each train to its maximum operating capacity. The equation used in its determination is as follows:

Each train was setup at a maximum capacity thereby any additional capacity beyond the reported maximum capacity would mean the addition of another train. This would double the total equipment cost for the new train as well as increase the energy cost involved in the process. Fig. 71 below shows the relationship between the capital cost and an operating capacity. The spikes in the chart signify an increase in the capital cost upon the addition of another train.

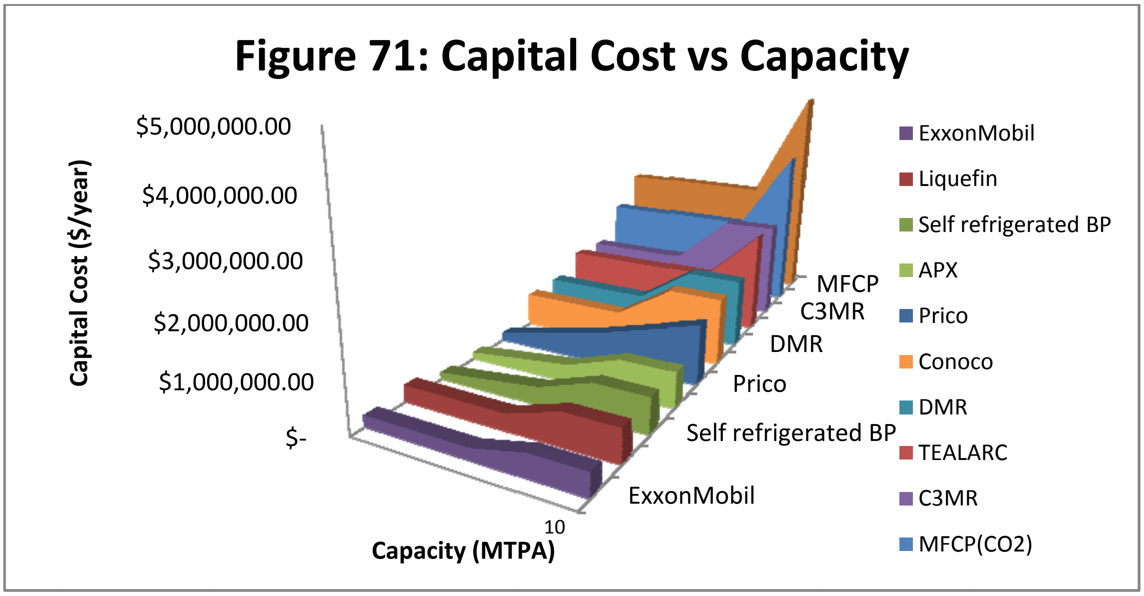


Figure 71: Capital Cost vs Capacity

The straight line signifies the range at which the number of trains remains constant as well as the equipment cost. This trend is not realized in the energy cost chart illustrated in Figure 72 below. There is always a proportional increase in the energy cost, in relation to the capacity being measured.

Figure 72: Energy Cost vs. Capacity

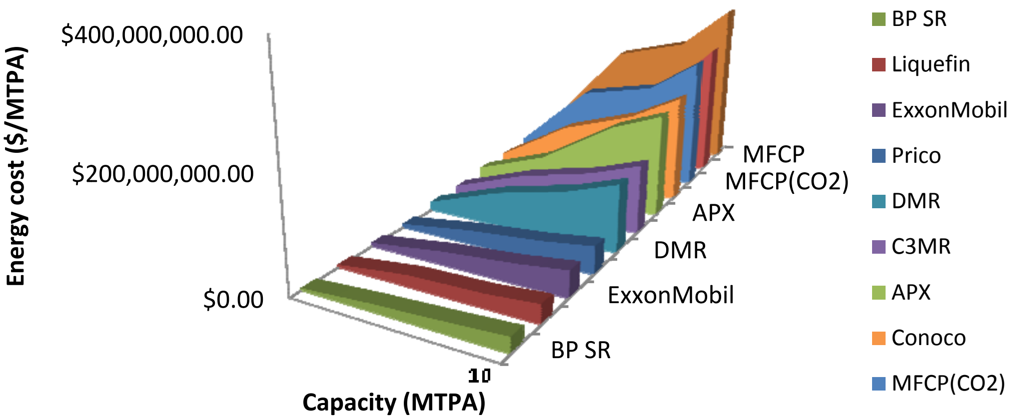


Table 32 below shows the estimated cost per ton and the maximum operation capacity of each process. This is aimed at making a direct comparison between the processes in order to help make decisions on what processes should be selected.

Table 32: Process Cost per Ton

Process	**Cost per ton ($)**	**Max capacity (MTPA)**
Prico	5.12	1.20
Liquefin	3.41	6.00
ExxonMobil	4.83	4.80
DMR	12.58	4.80
APX	19.20	7.80
MFCP	31.73	7.20
MFCP(CO2)	24.77	7.20
TEALARC	25.35	6.00
C3MR	12.93	4.80
Conoco	20.15	5.00

Recommendations

Newer processes on the list, which were considered relatively more efficient show lower cost per ton and fare better in direct comparisons. Processes like Liquefin and ExxonMobil show a lower cost per ton while producing relatively high capacities. Larger processes like the MFCP, TEALARC tend to spend more on capital cost and energy cost in order to obtain high train capacities.

More factors would be involved in making a final decision on selecting a final process. Some companies have proprietary rights over these techniques and installation costs might include licensing fees as well. Some processes are sold outright to customers and some others are contractual agreements usually involving revenue from selling natural gas. Trade sanctions and conflict situations in international locations could also drive up the installation costs.

With each company in charge of negotiations involving process installations a cost, a rigid blueprint on the economic impact cannot be easily determined externally. The use of simulations based on available information and research show an initial overview and can serve as a starting point before a final decision is to be made.

References

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[14] Plant Design and Economics for Chemical Engineers, 2003. <http://www.mhhe.com/engcs/chemical/peters/data/ce.html>

Appendices

Patent Search

There were several searches conducted to find the processes chosen to be simulated. Initially there were several documents available to search through for processes. These documents included the most popular choices of liquefaction today, most of which were included in those to be simulated. Second, it was felt that the easiest method to find other process and information on the processes already chosen would be to use a search engine on the internet.

The second search was conducted through Google.com. In the entry box, the phrase LNG liquefaction processes was entered. This gave the expected results of a few new processes not already included and several processes that were. Of the processes already included, this provided several new sources of information. For the processes that were not fully understood, it provided alternative choices. Once this was completed, the names of individual processes were placed in the search box. This provided more specific information on each process. The information needed, was process flow diagrams, information about equipment, and information about the refrigerant streams used. All this information was tracked and used in the simulation process.

A search of patents in the United States Patent Office was also conducted. In searching the patents, a quick search was completed. In the first search box, the phrase LNG liquefaction was typed in. This allowed for both words to be kept together and for the search to use them separately. In this manner the search was made more thorough. In later searches where the words were included separately or other words were added, more patents were brought up, most of which had nothing to do with liquefaction. A search was also conducted in the European

Patent Office website. The key phrases that yielded relevant patents are LNG system and natural gas liquefaction.

In the relevant search, roughly sixty patents were found with the words LNG liquefaction. A brief skim of the abstracts usually removed the patents or included them for further research. In this first elimination processes, it was found that most of the patents were not for liquefaction processes, but for ships and or re-gasification processes and such that are part of the natural gas process as a whole. This eliminated roughly forty processes. Of the remaining twenty or so patents left, further inquiries eliminated most, once it was determined that these were for processes already included, processes that were similar, or processes in which the end result was not for natural gas used commercially.

Of the processes found in the patents search only four were kept on the bases discussed earlier. The processes included in the search are described in more detail in the rest of the report. Included in a later appendix are the results of our searches.

Results of Patent Searches

European Patent Office lists

Natural gas liquefaction

RESULT LIST
*Approximately **426** results found in the Worldwide database for:*
***natural gas liquefaction** in the title*
(Results are sorted by date of upload in database)

1	**Natural gas liquefaction**

Inventor: WILKINSON JOHN D (US); HUDSON HANK M (US); (+1)
Applicant: ORTLOFF ENGINEERS LTD
EC:
IPC: ***F25J1/00***; **F25J1/00**
Publication info: **US2008028790** - 2008-02-07

2 Natural Gas Liquefaction Process

Inventor: BAUER HEINZ (DE); FRANKE HUBERT (DE); (+6)
Applicant: LINDE AG (DE); STATOIL ASA (NO)
EC: F25J1/02D
IPC: ***F25J1/02***; *F25B1/10; F25B5/02* (+5)
Publication info: **US2008006053** - 2008-01-10

3 Apparatus for the liquefaction of natural gas and methods relating to same

Inventor: WILDING BRUCE M (US); BINGHAM DENNIS N (US); (+6)
Applicant: BECHTEL BWXT IDAHO LLC (US)
EC: F25J3/08; F25J1/02; (+2)
IPC: ***B01D21/26; F25J1/02; B01D36/04*** (+10)
Publication info: **EP1867940** - 2007-12-19

4 Natural gas liquefaction system

Inventor: HOWARD HENRY E (US)
Applicant: PRAXAIR TECHNOLOGY INC (US)
EC: F25J1/02
IPC: ***F25J1/00***; **F25J1/00**
Publication info: **CN101048636** - 2007-10-03

5 Natural gas liquefaction.

Inventor: WILKINSON JOHN D HUDSON HANK M (US)
Applicant: ORTLOFF ENGINEERS LTD (US)
EC: F25J1/02; F25J3/02A2; (+4)
IPC: ***F25J1/00; F25J1/02; F25J3/00*** (+4)
Publication info: **CN101006313** - 2007-07-25

6 NATURAL GAS LIQUEFACTION METHOD

Inventor: BELJAEV ALEKSANDR ALEKSEEVICH (RU); GLAZUNOV VIKTOR DMITRIEVICH (RU); (+3)
Applicant: ZAO KRIOGAZ (RU)
EC:
IPC: ***F25J1/00***; **F25J1/00**
Publication info: **RU2306500** - 2007-09-20

7 Novel second-order mixed refrigeration process used for natural gas liquefaction

Inventor: LIANG SHIXI ZHANG (CN)
Applicant: HENAN ZHONGYUAN GREEN ENERGY H (CN)
EC:
IPC: ***F25J1/02***; **F25J1/00**
Publication info: **CN101008545** - 2007-08-01

8 APPARARTUS FOR THE LIQUEFACTION OF NATURAL GAS AND METHODS RELATING TO SAME

Inventor: TURNER TERRY D (US); WILDING BRUCE M (US); (+1)
Applicant: BATTELLE ENERGY ALLIANCE LLC (US)
EC: F25J1/02; F25J3/08
IPC: ***F25J1/00***; **F25J1/00**

Publication info: **WO2007133233** - 2007-11-22

9 **APPARATUS FOR THE LIQUEFACTION OF NATURAL GAS AND METHODS RELATING TO SAME**

Inventor: WILDING BRUCE M (US); MCKELLAR MICHAEL G (US); (+2)
Applicant: BATTELLE ENERGY ALLIANCE LLC (US)
EC: F25J1/02; F25J3/00B; (+4)
IPC: ***F25J1/00***; **F25J1/00**
Publication info: **WO2007130108** - 2007-11-15

10 **Method for liquefying hydrocarbon-rich flow, particularly natural gas flow, involves subjecting hydrocarbon-rich flow to absorptive water separation, before its liquefaction, where cooling of liquefied hydrocarbon-rich flow is up streamed**

Inventor: BOELT MANFRED (DE); FOERG WOLFGANG (DE); (+3)
Applicant: LINDE AG (DE)
EC: F25J1/02D
IPC: ***F25J1/00; F25B43/00***; **F25J1/00** (+1)
Publication info: **DE102006021620** - 2007-11-15

Lng system list

RESULT LIST
Approximately ***180*** *results found in the Worldwide database for:*
lng system *in the title*
(Results are sorted by date of upload in database)

1 *METHOD FOR TREATING OFFSHORE LNG REGASIFICATION SYSTEM FOR LNG REGASIFICATION SHIP*

Inventor: YOON JEONG SIK (KR); BAE JAE RYU (KR)
Applicant: DAEWOO SHIPBUILDING & MARINE (KR)
EC:
IPC: B63B27/30; B63B27/34; F17C9/02 (+2)
Publication info: KR100779779B - 2007-11-27

2 *LNG-BASED POWERAND RBGASIFICATION SYSTEM*

Inventor: KLOCHKO MARAT (IL); KAPLAN URI (IL); (+4)
Applicant: ORMAT TECHNOLOGIES INC (US)
EC:
IPC: F01K27/00; F01K25/00; F01K25/08 (+2)
Publication info: EP1888883 - 2008-02-20

3 *SYSTEM TO INCREASE CAPACITY OF LNG-BASED LIQUEFIER IN AIR SEPARATION PROCESS*

Inventor: DEE DOUGLAS PAUL; CHOE JUNG SOO; (+1)
Applicant: AIR PROD & CHEM
EC: F25J3/04F
IPC:

Publication info: SG138574 - 2008-01-28

4 LNG SYSTEM EMPLOYING STACKED VERTICAL HEAT EXCHANGERS TO PROVIDE LIQUID REFLUX STREAM
Inventor: EATON ANTHONY P (US); MESSERSMITH DAVID (US); (+1)
Applicant: CONOCOPHILLIPS CO (US)
EC: F25J3/02C4; F25J1/02; (+3)
IPC: F25J1/00; F25J1/00
Publication info: US2008022716 - 2008-01-31

5 LNG SYSTEM WITH ENHANCED REFRIGERATION EFFICIENCY
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EC:
IPC: F25J1/00; F25J1/00
Publication info: WO2008014091 - 2008-01-31

6 Lng Containment System And Method Of Assembling Lng Containment System
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EC: F17C1/00
IPC: E04H1/00; F17C1/00; E04H1/00 (+1)
Publication info: US2008016788 - 2008-01-24

7 SEAM BUTT TYPE INSULATION SYSTEM HAVING WELDABLE SECONDARY BARRIER FOR LNG TANKS
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EC:
IPC: B65D90/06; B65D90/02
Publication info: WO2008007837 - 2008-01-17

8 LNG pressure regulating system and pressure regulating method thereof
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Applicant: CHINA INTERNAT MARINE CONTAINE (CN)
EC:
IPC: B67D5/60; B67D5/60
Publication info: CN101058401 - 2007-10-24

9 LNG SYSTEM WITH OPTIMIZED HEAT EXCHANGER CONFIGURATION
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EC: F25J1/02
IPC: F25J1/00; F25J1/00
Publication info: US2007283718 - 2007-12-13

10 AUTO WELDING MACHINE FOR SETTING INSULATION SYSTEM INSIDE CARGO TANK OF LNG SHIP
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Applicant: KOREA GAS CORP (KR)
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IPC: B23K37/02; B23K9/00; B23K37/02 (+1)
Publication info: KR20070096633 - 2007-10-02

United States Patent Office Search

LNG Liquefaction

PAT. NO.		Title	
1	7,325,415	T	Process and device for production of LNG by removal of freezable solids
2	7,318,319	T	Apparatus for cryogenic fluids having floating liquefaction unit and floating regasification unit connected by shuttle vessel, and cryogenic fluid methods
3	7,311,055	T	Vessel with deep water transfer system
4	7,299,643	T	Method for recovering LPG boil off gas using LNG as a heat transfer medium
5	7,234,323	T	Recovering natural gas liquids from LNG using vacuum distillation
6	7,225,636	T	Apparatus and methods for processing hydrocarbons to produce liquified natural gas
7	7,219,512	T	Apparatus for the liquefaction of natural gas and methods relating to same
8	7,210,311	T	Natural gas liquefaction
9	7,204,100	T	Natural gas liquefaction
10	7,137,260	T	Method and substance for refrigerated natural gas transport
11	7,101,118	T	Multi hull barge
12	7,073,457	T	Duplex yoke mooring system
13	7,010,937	T	Natural gas liquefaction
14	7,007,623	T	Retrieval and connection system for a disconnectable mooring yoke
15	6,978,638	T	Nitrogen rejection from condensed natural gas
16	6,962,061	T	Apparatus for the liquefaction of natural gas and methods relating to same
17	6,945,075	T	Natural gas liquefaction
18	6,889,523	T	LNG production in cryogenic natural gas processing plants
19	6,889,522	T	LNG floating production, storage, and offloading scheme
20	6,886,362	T	Apparatus for the liquefaction of natural gas and methods relating to same
21	6,851,994	T	Disconnectable mooring system and LNG transfer system and method
22	6,742,358	T	Natural gas liquefaction
23	6,581,409	T	Apparatus for the liquefaction of natural gas and methods related to same
24	6,564,578	T	Self-refrigerated LNG process
25	6,530,240	T	Control method for mixed refrigerant based natural gas liquefier
26	6,526,777	T	LNG production in cryogenic natural gas processing plants
27	6,463,740	T	Compressor starting torque converter
28	6,434,948	T	LNG load transfer system
29	6,367,258	T	Method and apparatus for vaporizing liquid natural gas in a combined cycle power plant
30	6,298,688	T	Process for nitrogen liquefaction
31	6,250,105	T	Dual multi-component refrigeration cycles for liquefaction of natural gas
32	6,070,429	T	Nitrogen rejection system for liquified natural gas

33	5,842,357	T	Landfill gas recovery
34	5,803,005	T	Ship based system for compressed natural gas transport
35	5,791,160	T	Method and apparatus for regulatory control of production and temperature in a mixed refrigerant liquefied natural gas facility
36	5,755,114	T	Use of a turboexpander cycle in liquefied natural gas process
37	5,724,833	T	Control scheme for cryogenic condensation
38	5,681,360	T	Landfill gas recovery
39	5,669,238	T	Heat exchanger controls for low temperature fluids
40	5,615,561	T	LNG production in cryogenic natural gas processing plants
41	5,421,165	T	Process for denitrogenation of a feedstock of a liquefied mixture of hydrocarbons consisting chiefly of methane and containing at least 2 mol % of nitrogen
42	5,325,673	T	Natural gas liquefaction pretreatment process
43	4,970,867	T	Liquefaction of natural gas using process-loaded expanders
44	4,901,533	T	Process and apparatus for the liquefaction of a natural gas stream utilizing a single mixed refrigerant
45	4,541,852	T	Deep flash LNG cycle
46	4,392,346	T	Cogeneration process using augmented Brayton cycle
47	4,305,256	T	Cryogenic gel having a methane component and process for making same
48	4,168,673	T	Floating island for extracting or processing gas
49	4,130,077	T	Single-point mooring system
50	4,057,972	T	Fractional condensation of an NG feed with two independent refrigeration cycles