NATURAL GAS LIQUEFACTION PROCESS AND OPTIMIZATION

(FINAL PROJECT REPORT)

SUBMITTED BY:

ANURADHA MEHRA
MALVIKA BAHUGUNA

Under the able guidance of
Dr.R.P.Badoni

In partial fulfillment of the requirements for

BACHELOR OF TECHNOLOGY
IN
APPLIED PETROLEUM ENGINEERING

COLLEGE OF ENGINEERING
UNIVERSITY OF PETROLEUM AND ENERGY STUDIES
DEHRADUN
2007

UNIVERSITY OF PETROLEUM & ENERGY STUDIES

CERTIFICATE

This is to certify that the project work on ***"NATURAL GAS LIQUEFACTION PROCESS AND OPTIMIZATION"*** submitted to University of Petroleum & Energy Studies, Dehradun, by Ms.Anuradha Mehra and Malvika Bahuguna, in partial fulfillment of the requirement for the award of Degree of Bachelor of Technology in Applied Petroleum Engineering (Academic Session 2003 – 2007) is a bonafied work carried out by them under my supervision and guidance. This work has not been submitted anywhere else for any other degree or diploma.

R.P. Badoni
09.08.2007

Dated:

Dr. R.P.BADONI

Distinguished Professor

Corporate Office :
Hydrocarbons Education & Research Society
3rd Floor, PHD House 4/2, Siri Institutional Area
August Kranti Marg, New Delhi-11001 India
Ph + 91-11-41730151-53 Fax +91-11 1730154

Main Campus :
Energy Acres, PO Bidholi, Via Prem Nagar,
Dehradun-248 007 (Uttaranchal) India
Ph. : +91-135-2261090-91, 2694201/203/208
Fax : +91-135-2694204

Regional Centre (NCR) :
SCO 9-12, Sector-14, Gurgaon 122 007
(Haryana), India
Ph : + 91-124-4540300
Fax : +91 124 4540 330

E-mail . info@upes.ac.in URL:www.upes.ac.in

ACKNOWLEDGEMENT

We would like to express our most sincere gratitude to many people, notably among them are numerous scholars at UPES who have given guidance at various levels and gave us an opportunity to work with them. They provided us necessary impetus with all their zeal and motivation and without their guidance, support and extremely helpful attitude this project would not have seen the light of the day.

We express our deep sense of gratitude and indebtedness to our guide **Dr. R.P.Badoni** for his invaluable and efficient guidance.

Anuradha Mehra

Malvika Bahuguna

EXECUTIVE SUMMARY

Natural gas is fast gaining pivotal prominence in meeting the energy demand worldwide, especially in Asia Pacific region. Several factors support an increased role of gas. Apart from economic consideration, the environmental superiority of natural gas over other fossil fuels in response to growing environmental awareness is one of the main drivers pushing up the share of natural gas in the energy mix.

Furthermore the market for LNG is growing at an enormous rate and is expected to double in the next decade. Gas trading usually confined to regional markets by pipelines can now be extended world wide on account of very large capacity LNG tankers.

The rise of LNG promises to change the notion "that gas traditionally needed elaborate system of pipelines to get it from well-head to customer". Put simply, gas can be frozen into liquid form near its source, shipped to the market in refrigerated tankers, warmed back into gaseous form on foreign shores and injected into the local pipeline system. This technological advancement has made gas a fungible global commodity like oil. Our nation is also gradually adopting LNG as the complementary source to make up for the deficit in supply.

In recent years, process selection and comparison studies for base load LNG production have grown in scope and have increasingly considered multiple cycle option. The reason include interest in larger trail capacities, efforts to minimize carbon di oxide emissions, different compressor drive arrangements plus specialized applications such as floating production storage and off-loading(FPSO) plants, LNG peak-shaving and associated gas liquefaction.

The liquefaction system described in this report are: the conventional process using a propane refrigeration, pre-cooling step followed by mixed refrigeration condensation cycle (C3/MR)

and the newer technologies that include C3MR followed by a nitrogen refrigeration cycle, 3-refrigerant cascade, dual-mixed refrigerant(DMR) with spiral wound exchangers, and DMR with plate fin-exchangers.

Selecting the best fit technology is a complex affair as it is sensitive to number of design parameters, therefore merit careful attention.

Accordingly, effect of several parameters is analyzed namely: type refrigeration cycle on condenser size, compressor size on Mach number, end-flash quantity, compressor efficiency, and LPG recovery on LNG production. Careful consideration of these parameter enable scientifically and technologically appropriate decision leading to correct selection and comparison of liquefaction technology and capital cost comparison between APCI and AP-X^{TM} process. ,

Main objective for optimizing Liquefaction process: Maximize LNG production and minimize energy consumption. This objective can be accomplished by integrating LNG plant with NGL and GTL, nitrogen removal unit, vapour recovery, efficient equipments, tuebo-expander, suitable driver configuration, etc.

TABLE OF CONTENTS Page No.

LIST OF TABLES

LIST OF FIGURES

I. NATURAL GAS LIQUEFACTION PROCESS AND OPTIMIZATION

Liquefaction technology is based on the principle of a refrigeration cycle, where a refrigerant by means of successive expansion and compression transports heat from a lower to a higher temperature. LNG plants consist of a number of parallel units, called trains, which treat and liquefy natural gas and then send the LNG to several storage tanks. The capacity of a liquefaction train is primarily determined by the liquefaction process, the refrigerant used, and largest available size of the compressor / driver combination that drives the cycle and the heat exchangers that cool the natural gas.

The first LNG plants were built in Algeria, the US and Libya. They used either a Cascade process, with three-pure refrigerants, or a single mixed refrigerant. In Brunei, a two-cycle process was implemented for the first time, using propane and a mixed refrigerant, capturing the benefit of the two previous processes. This Propane Mixed Refrigerant (MR) process - developed by Air Products & Chemicals Int. (APCI) - started to dominate the industry from the late seventies on.

Economy of scale drove the size of propane-MR trains from 1.1 million tons per annum (Mtpa) for the first train built in Brunei to 3.9 Mtpa for the last LNG train that was recently started up. This scale increase was realized by close cooperation of operating companies, process licensors and equipment vendors. As compressors are key equipment of any refrigeration cycle, compressor vendors focus on increasing it's capacity.

The earliest compressors were driven by steam turbines, but Shell and Nuove Pignone introduced gas turbines as mechanical drivers. The earliest applied gas turbines delivered around 25 MW of power (GE Frame 5), while some plants apply gas turbines that deliver 75 MW of power (GE Frame 7). Nowadays, frame-9 gas turbines (deliver 110MW) and variable speed electric motor drivers are used. Last but not least, the size of spool wound cryogenic heat exchangers, and plate fin heat exchanger were steadily increased to match the developments of compressors and drivers.

However, costs are not the only driver. The restriction on CO2 production under the Kyoto protocol has made efficiency a second driver for considering new process options. The Oman LNG plant was built with a specific capital expenditure of 200 $/ton, but the plant was also designed for a high energy efficiency of 92%.

All these developments have led to the current state of the art in LNG technology, which is the two-cycle propane-MR process. The LNG trains in Shell-advised Oman LNG are thought to be the leading example of cost and energy efficiency. Innovation in liquefaction process and technology will be judged against these benchmarks of size, cost and efficiency, but also operability and reliability will be important parameters.

Two areas can be identified where major improvements are possible: better integration of the various processes like NGL extraction, LPG recovery, GTL with LNG plant, and use of efficient equipments like large capacity compressors, plate-fin heat exchanger, drivers, and proper refrigerants in LNG process. This helps in optimizing natural gas liquefaction process in terms of cost, equipments and efficiency.

II. SCOPE OF PROJECT

I. Study of Basics of Liquefaction Process

II. Comprehensive Study of Various conventional and Emerging Liquefaction process: The conventional process which is used for the liquefaction of natural gas has been modified with better and innovative & creative techniques for maximum LNG production rate to meet the energy demand and minimize energy consumption, create flexibility and ease in the operations, minimizing space requirements which results in reduced operating cost of the plant and the overall plant is optimized.

III. Techno-economic assessment for the conventional and emerging liquefaction technology : C_3MR and AP-X™ Process

IV. Scope for optimizing the Liquefaction process which can be accomplished by integrating LNG plant with NGL and GTL, additional nitrogen removal unit, vapor recovery, efficient equipments selection, suitable driver configuration, etc.

III. BACKGROUND

1. LNG-MARKET-GLOBAL SCENARIO

The world LNG market has seen a lot of growth, especially in the regions undergoing economic development. A market research report by RNCOS, "Global LNG market Worldwide (2005-2015)" has brought out a detailed picture of the world LNG scenario. It gives an in-depth analysis of the trends and behavior of the world LNG market. According to the research report, the LNG market has shown a tremendous growth worldwide. About one to 40 LNG projects exist or are in the pipeline in every country. This goes well with the demand and supply statistics, the sooner the projects will get over, the better the supply/capacity of LNG will become.
As per the market research report, the delays that are taking place in the liquefaction projects are affecting the market share for LNG. A better pace will lead to reduction in the LNG prices. If the gestation gap compresses, the LNG market economics can be balanced very well. A boom in the production and consumption of LNG was seen in the year 2005. The global liquefaction capacity for LNG was 150MTPA in 18 plants spanning13countries.

A calculation reveals that there were 14 importers of LNG in 2005. According to experts, the LNG market is expected to increase with a number of 35 importers joining the race by 2015. The world LNG market will double its size with countries like Taiwan, Japan and Korea, and other Asian countries becoming prominent in the demand and supply for the fuel. The reason attributed to the growth of LNG worldwide is the need for electricity produced by environmentally friendly fuel. Domestic natural gas resources have added to the demand. Most of the gas consuming countries stress on commercialization of their resources.

Natural gas is transported either by pipeline (73% of internationally traded gas in 2005), mainly across landmasses and by LNG transportation across the oceans (27% of internationally traded gas in 2005). The rapid expansion of LNG infrastructure worldwide in the past decade is enabling natural gas to penetrate many more isolated gas markets and the development of remote gas reserves once considered to be stranded and uneconomic to develop.

The liquefied natural gas (LNG) industry is set for a large and sustained expansion as improved technology has reduced costs and improved efficiency along the entire supply chain over the course of the past decade. This shift in the dynamics of the natural gas market will further commoditize and diversify the natural gas globally.

Table2: Worldwide Distribution & Capacity of Gas Liquefaction Plants (2006)

Country	Number of Liquefaction Trains	Capacity- million tones per year (mtpa)
AFRICA		
Algeria	21	23.1
Egypt	3	12.0
Libya	3	2.3
Nigeria	5	16.9
ASIA		
Australia	4	11.7
Brunei	5	7.2
Indonesia	14	30.6
Malaysia	8	22.7
MIDDLE EAST		
Oman	3	9.3
Qatar	7	27.3
UAE	5	5.6
NORTH AMERICA		
USA	1	1.5
SOUTH AMERICA		
Trinidad and Tobago	4	15.1
TOTAL	**81**	**185.3**
Total LNG Exported in 2005	**~141**	

DEMAND FOR LNG

Consuming countries in Asia, Europe and North America, are continuing to develop LNG importing infrastructure and are looking at LNG not just in terms of diversification and cleaner energy production, but because some of the largest consumers (e.g. China and USA) have growing energy supply deficits i.e. production of domestic energy resources is falling to satisfy growth in energy demand. Therefore, LNG is attractive to a variety of consuming nations:

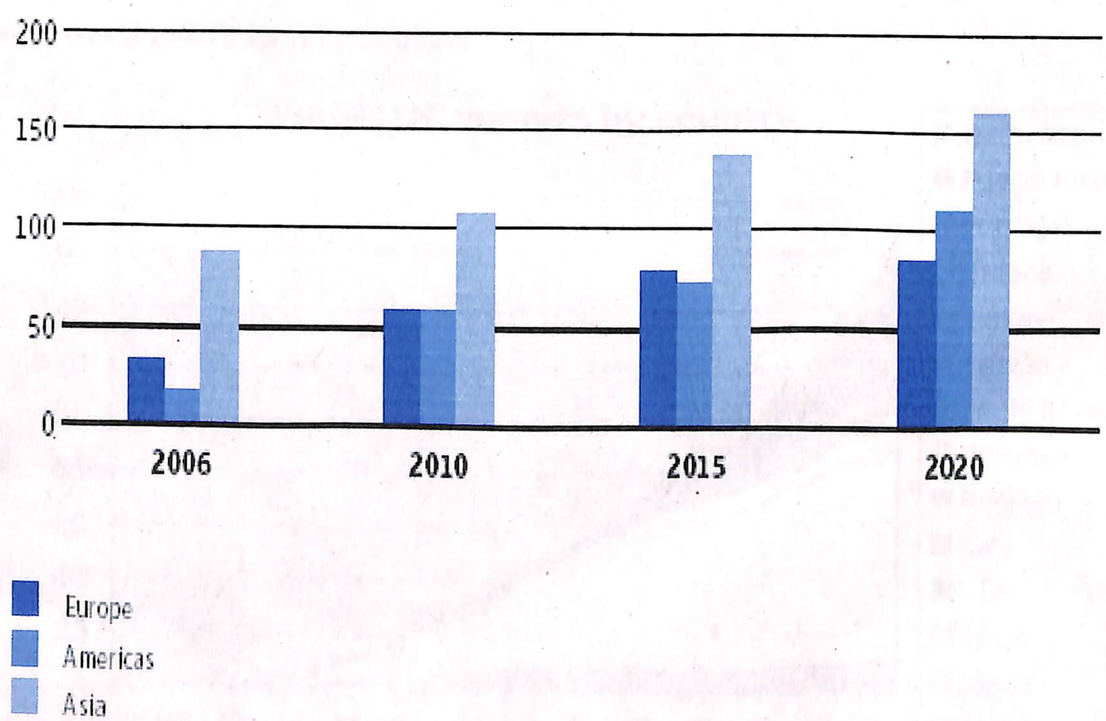
- Those without significant reserves, seeking reduced oil dependence, such as- Japan and Korea, but also major EU economics (e.g. Spain, Italy, France, Netherlands and other);
- Those whose indigenous industries have hit or passed peak output, such as the United States and United Kingdom, which have highly developed gas delivery infrastructure but are increasingly net importers; and
- Developing economies which are energy hungry for their fast economic growth- obviously china and India, both of which have under developed gas handling and pipeline infrastructures but potential for huge growth in the number of end-users.

There are approximately 60 LNG receiving terminals located worldwide.

Asia holds more than three-quarters of current worldwide LNG import storage tank capacity.

Global LNG demand outlook

(million tones per year)



Source: Cambridge Energy Research Associates (CERA), 2006.

Figure1: Global LNG Demand Outlook

LNG EXPORT AND IMPORT

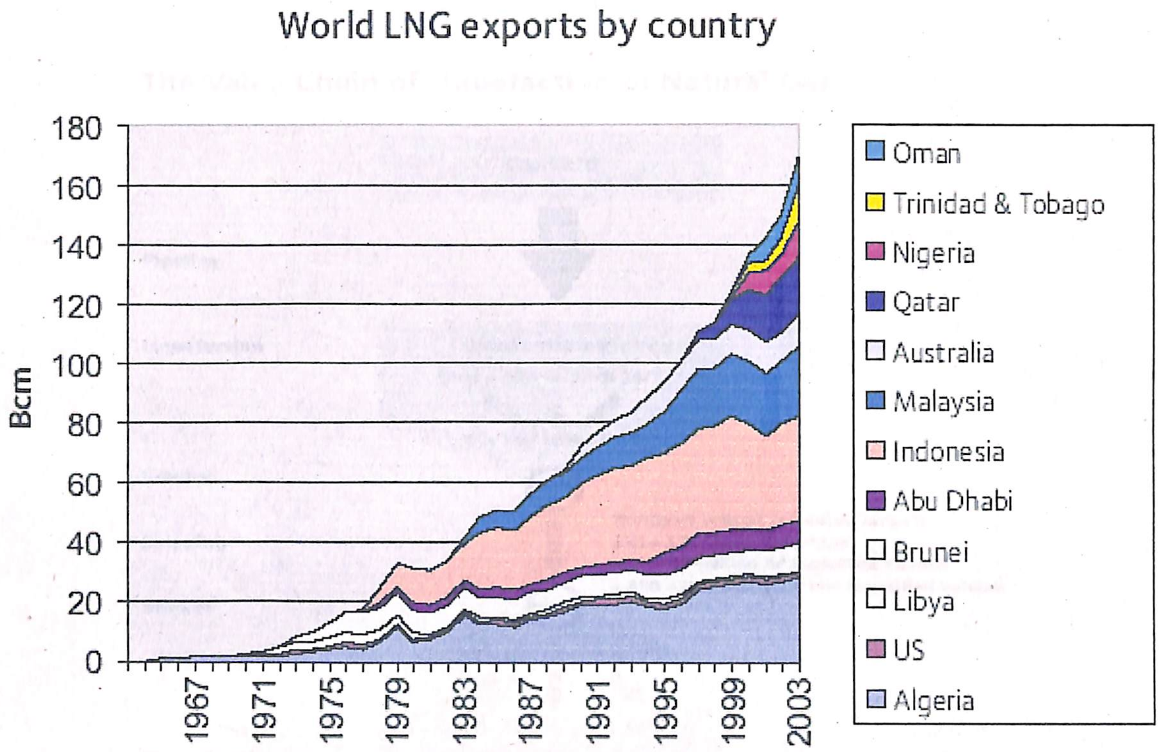


Figure2: World LNG Export by Countries

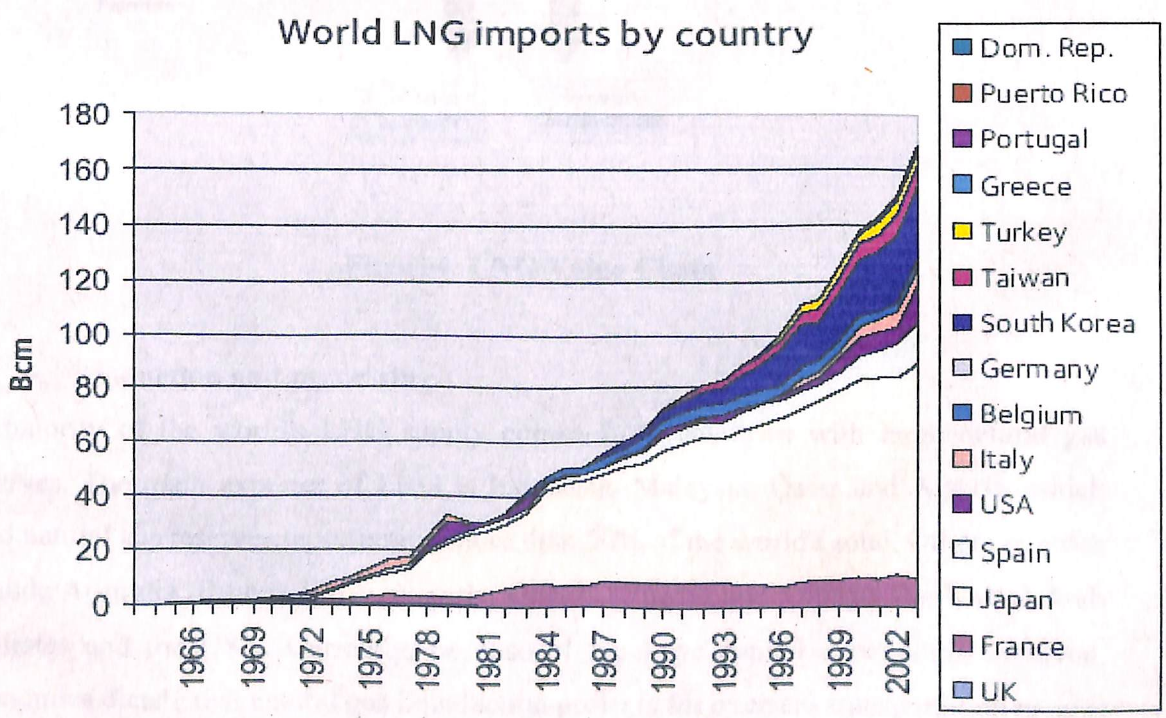


Figure3: World LNG import by Countries

2. LNG VALUE CHAIN

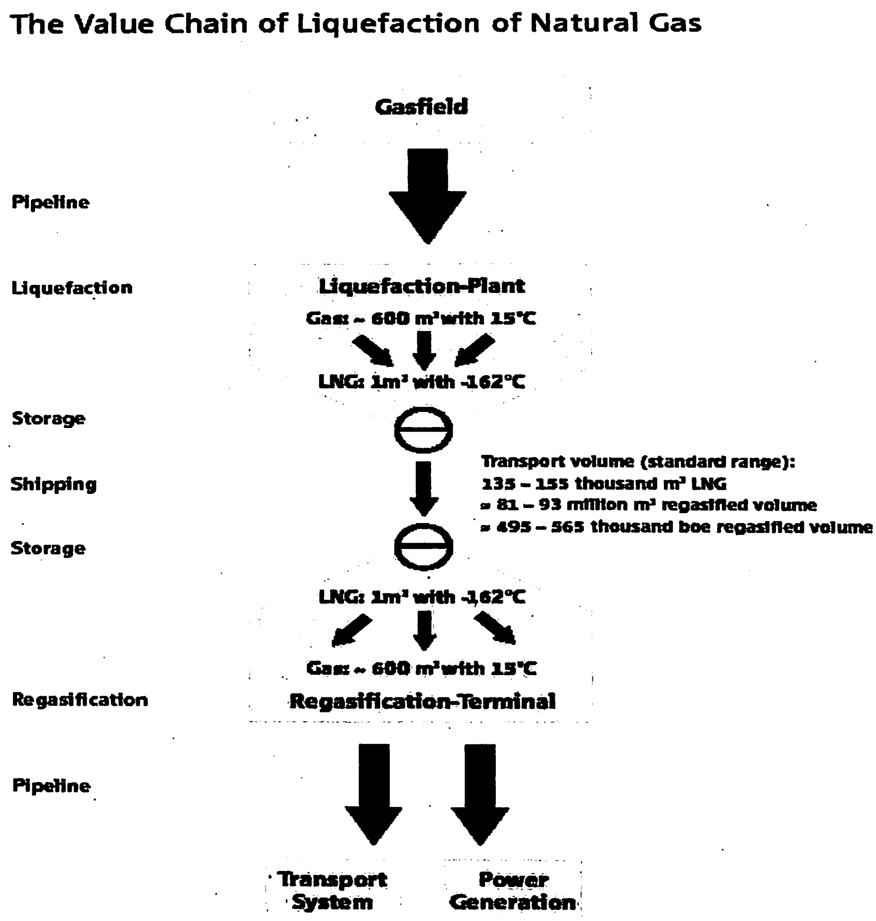


Figure4: LNG Value Chain

- **Production and processing**

A majority of the world's LNG supply comes from countries with large natural gas reserves. The main exporter of LNG is Indonesia, Malaysia, Qatar and Algeria, which hold natural gas reserves representing more than 50% of the world's total. Other countries include Australia, Brunei, Libya, Nigeria, Oman, Trinidad and Tobago, the United Arab Emirates and the USA. Currently, because of the large capital expenditure involved, economics dictate that natural gas liquefaction projects for overseas transportation require large gas reserves able to produce large quantities of gas for at least 20 years. Worldwide, there were 28 LNG export terminals in 2005.

- **Liquefaction**

At the production site, natural gas is converted into liquid form through a cooling process called liquefaction, achieved through refrigeration cycles. Before liquefaction, the natural gas is purified from humidity, carbon dioxide, sulphur components, heavy hydrocarbons and other impurities. The purified natural gas is cooled to its atmospheric boiling point (about -161°C). The set of units where natural gas is purified and liquefied is called a train. In 2005, there were 75 trains, which means a global LNG liquefaction capacity of roughly 160 million tonnes. This production capacity is expected to increase to 197 million tones in 2007 based on the facilities under construction. There are 3 main liquefaction processes:

I. **The classical cascade,** where refrigeration and liquefaction of the gas is achieved in a cascade process using three pure refrigerants: propane, ethylene and methane;

II. **The single flow mixed refrigerant process**, where the mixed refrigerant made up of nitrogen, methane, ethane, propane and isopentane, is compressed and circulates using a single compression train;

III. **The propane precooled mixed refrigerant process** where precooling is achieved by a multi-stage propane cycle and liquefaction and subcooling are accomplished by a two-stage mixed refrigerant cycle, which is so far the most common process, used since 1972 in 8 different countries.

- **Transport**

It is easy to transport LNG from remote production areas to consumer areas than conventional gas pipelines. In international trade, LNG is transported in double-hulled ships, specifically designed to handle the low temperature of LNG, to a receiving terminal where it is stored and regasified. These carriers are insulated to limit the amount of LNG that boils off or evaporates. This boil-off gas is sometimes used to supplement fuel for the carriers or it is re-liquefied by an on-board liquefier

In May 2005, 181 LNG tankers were operating, with another 74 under construction for delivery in the 2005-2007.

- **Storage**

In production plants and in reception terminals, LNG is stored in specifically built insulated flat bottom storage tanks, operating at atmospheric pressure. These tanks can be above or below ground and keep the liquid at very low temperatures (cryogenic condition) to minimize evaporation. The largest tanks currently in operation are 160 000 m3 above ground and 200 000 m^3 under ground. The tanks are insulated by special powder insulation, filled between the inner and outer shells. In 2005, worldwide, there were 50 LNG import terminals in Asia (India, Japan, South Korea, Taiwan), Europe (Belgium, France, Greece, Italy, Portugal, Spain, Turkey, United Kingdom), North America (the United States) and South America. In 2005, the main LNG importers were Asia with 64.8%, in particular Japan with 40.4%, and Europe with 25.2% of total imports. However, the LNG market is undergoing global expansion.

- **Regasification and distribution**

At the import terminal, LNG is warmed to a point where it converts back to its gaseous state. This is accomplished using special exchangers fed with high-pressure pumps for achieving the final gas pressure. Two types of exchangers are used mainly: open-rack vaporizers (ORV) using sea water and submerged combustion vaporizers (SCV) that are water baths heated by combustion of fuel gas. While 2% of the imported gas is used as fuel gas for LNG vaporization, nearly all the gas is distributed to end-users via a conventional gas pipeline network with recompressing stations at distances 150 to 250 km. Organizing direct distribution of LNG from import terminals to end-users by road trailers or rail cars will be a challenge in the years to come.

- **Local liquefiers**

LNG may also be produced by liquefying gas taken from a pipeline or from a small local well. The typical capacity of such installations is 40 to 100 tonnes/day. Some of them are used as peak-shaving plants, for liquefaction of gas in periods of low demand, and for vaporization back into pipeline, when the demand is high. The others are used for further distribution of LNG to satellite plants. It can be based again on a mixed refrigerant cycle, or it can be a straddle plant, in parallel to a pressure-reduction station, where the gas is throttled from the transit pipeline to a local net, thus using the available pressure energy for liquefaction.

- **Small scale distribution**

LNG from marine terminals or from local liquefiers can be further distributed for one of the following purposes:

– **satellite plants**: LNG may be transported in special tanker trucks to small facilities, called satellite plants, where it is stored and regasified into a local pipeline for heating or process technology. This system may be very beneficial for countries with no existing or insufficient pipeline infrastructures. There are still a lot of opportunities enabling developing countries to speed up their industrial growth. LNG is also used in industrialized countries where opportunities abound for gas delivery to scattered settlements and areas where natural protection is in force. Some satellite plants can be operated for peak shaving or for back-up purposes in parallel to delivery by pipeline;

TYPES OF LNG PLANT

• Base-load plants

– Large plants which are directly based on a specific gas field development and are the main plants for handling the gas. A base-load plant has typically a production capacity of above 2 mtpa (million tons per annum) of LNG. The main world-wide LNG production capacity comes from this type of plants

• Peak-shaving plants

– Smaller plants that are connected to a gas network. During the period of the year when gas demand is low, natural gas is liquefied and LNG is stored. LNG is vaporized during short periods when gas demand is high. These plants have a relatively small liquefaction capacity (as 200 tons/day) and large storage and vaporization capacity (as 6000 tons/day). Especially in the US many (57) such plants exist

• Small-scale plants

– Small-scale plants are plants that are connected to a gas network for continuous LNG production in a smaller scale. The LNG is distributed locally by LNG trucks, in a range of about 300 km from the production facility, to various customers with a small to moderate need of energy or fuel. This type of LNG plants typically has a production capacity below 100 000 TPA. In Norway three plants within this category is in operation

IV. POSSIBILITIES, PERSPECTIVES AND CHALLENGES

- **Reducing costs and improving efficiency**

LNG production and transport still require important investments and the large liquefaction process represents 50% of the investment cost in the entire LNG value chain. For instance, the cost of a large liquefaction train with a capacity of 4.7 million tons per annum was referenced as 1.5×10^9 USD (average current costs are 250×10^6 USD/MTPA). The cost of an LNG tanker with 138 000 m^3 cargo volume is 0.2×10^9 USD. The cost of a regasification plant is 0.3×10^9 USD, or 0.5×10^9 (for a 6 MTPA).

Hence, in order for LNG to remain competitive, production costs must be reduced through technological developments as well as improvements in the design and construction of the entire plant.10 Complete LNG complexes currently cost between 2-5 109 USD each, and costs have already been reduced significantly since the mid 1990s. For example, liquefaction costs were around 560 USD/tonne in 1995 and were reduced to 222 USD/tonne in 2004. Developments increasing efficiency include the use of liquid expanders, enabling power reduction for the liquefier process, or alternately an increase in output for the same power.

- **Changes & Challenges with regard to construction and capacity**

It seems likely that future LNG export terminals will be larger, based in remote locations with no infrastructure and subjected to extreme weather conditions. Therefore, conventional construction approaches will no longer be cost and time-effective. Innovative efficient solutions are being developed. For example, recently, the major part of an LNG process plant was developed in a very compact form on a purpose-built barge that can be transported after assembly to the production site where it will serve as a permanent base and foundation for a new LNG process plant. Other projects are developed with modularization of the facility. Plant capacities presently reach 5 million tonnes LNG/year and designs are aiming for 7.8 million tonnes LNG per year. Storage tanks are also increasing in size: presently around 120 000-160 000 m^3.

- **Contract trends**

Currently, the LNG market is driven by long-term contracts, on a 20-25-year scale. However, they have become increasingly flexible in recent years. Some newer long-term contracts are designed to provide only a base supply of LNG, which can be supplemented by short-term contracts during periods of high demand. Medium-term and short-term contracts (or spot LNG trading) are emerging: growth has risen from 1% of the LNG market in 1992 to 8% in 2002 and could reach over 15% in the next decade. For the short-term market to expand, uncommitted ships and flexible contracts are required. These trends should promote LNG as a fuel of choice, either as the main fuel generator, or as a back-up system to compensate shortages at peak times.

- **Optimizing the safety**

Sustainable design implies LNG processes with highest safety, possibly by using non-hydrocarbon refrigerants in the natural gas liquefaction processes.

V. GAS MONETIZATION

Advanced Technologies for Gas Monetization

The world's oil resources are limited. Natural gas, by contrast, offers a substantially larger time perspective and is by far the cleaner energy source. Therefore, the trend in the foreseeable future is expected to be focused towards the processing and upgrading of natural gas. Already, the route from gas to petrochemicals and fuels is proving to be more economic than the classic petroleum route. Natural gas and associated gases from oilfields, as well as from coal and biomass, can be economically and efficiently converted into petrochemical products and synthetic fuels which are needed worldwide.

Following are the ways for gas-monetization:

1. **Natural gas to synthetic gas**

Increased concern about world's dependence upon petroleum oil in the light of its fast depleting reserves has provoked increased interest in the efficient utilization of natural gas. Methane (a major constituent of natural gas) may be converted to synthesis gas. It is a mixture of carbon monoxide and hydrogen, a highly versatile feedstock used in ethanol, Fischer-Tropsch synthesis processes, carbonylation, hydrogenation, hydroformylation and other industrially important reactions. The present section focuses on the industrial importance of synthesis gas and its diverse applications in synthesis of chemicals and liquid fuels and the literature on the conversion of methane to syngas. Syngas, the technical jargon for mixtures of hydrogen and carbon oxides, is crucially important building block of the chemical industry and is widely used in a variety of processes.

Currently, the production of H2 and CO, synthesis gas, (syngas), is carried out by steam reforming of methane, partial oxidation of fuel oil , coal gasification and naphtha reforming. Synthesis gas is widely used and can be converted into petrochemicals, higher alcohols and synthetic fuels. Hydrogen is used in ammonia synthesis and petroleum refining industries, while carbon monoxide is widely used in the production of plastics, paints, foams, pesticides and insecticides.

A powerful and highly economic production of synthesis gas is the key to modern plants with large capacities.

2. Methanol

Basis for more valuable products process, the "chemical" liquefaction of natural gas, represents a novel technology for converting natural gas to methanol at low cost and in large quantities. It permits the construction of highly efficient single-train plants. The currently operating facility using this technology is today producing more than 5,000 tons per day. This is more than twice the output of conventional plants and results in significant reduction in capital and operating costs.

This technology contributes to saving resources and decisively reduces the dependence on scarce crude oil reserves. The hitherto flared-off crude-oil associated gases can be utilized economically and ecologically.

3. Gas to Petrochemicals

Demand for methanol is substantial and the utilization of methanol generated from natural gas and associated gas is set to rise appreciably over the next few years. Methanol is one of the main feed stocks for syntheses in the petrochemical industry. It is thought that the percentage of plastics produced from natural gas will increase fivefold over the next 20 years. Lurgi developed the Methanol to Propylene (MTP®) process as a competitive alternative to produce propylene. It constitutes a simple, cost-effective and highly selective technology, yielding an excellent value-added product for the utilization of natural gas reserves

4. Gas to Synthetic Fuels

The generation of synthetic fuels from natural gas via the methanol, or Fischer Tropsch, routes is gaining momentum. The improved combustion of these fuels in engines and the fact that they are free from sulphur and aromatics results in more environmentally-friendly and more efficient automobiles. Not surprisingly the automotive industry is interested in adding such fuels to conventional petroleum based fuels. Substitution of 20–

30% of conventional fuel with synthetic fuels reduces emissions by up to 80%. Fuel production represents one of the largest markets in the world. For Europe, it is estimated that the share of petroleum-based fuels will drop from currently over 90% to below 40% in about 30 years time, while the share of synthetic fuels made from natural gas is expected to rise by an almost similar percentage.

VI. NATURAL GAS LIQUEFACTION PROCESS

1. BASICS OF LIQUEFACTION PROCESS

REFRIGERATION,

Refrigeration is the process of removing heat from an enclosed space, or from a substance, and rejecting it elsewhere for the primary purpose of lowering the temperature of the enclosed space or substance and then maintaining that lower temperature. To satisfy the Second Law of Thermodynamics, mechanical work must be performed to accomplish this.

REFRIGERATION PROCESS

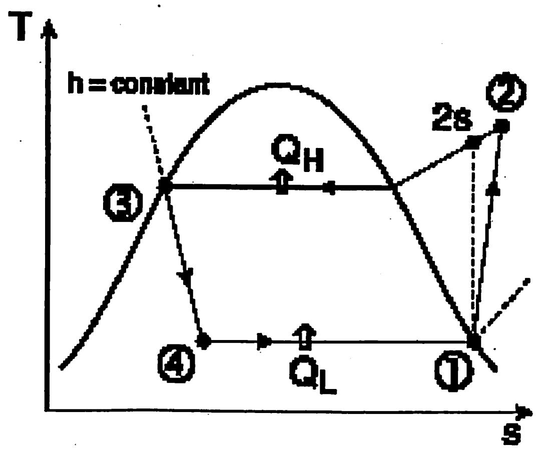


Figure5: T-S Diagram

1-2s: A reversible, adiabatic (isentropic) compression of the refrigerant. The saturated vapour at state 1 is superheated to state 2.

$$w_c = h_{2s} - h_1$$

2s-3: An internally, reversible, constant pressure heat rejection in which the working substance is desuperheated and then condensed to a saturated liquid at

3. During this process, the working substance rejects most of its energy to the condenser cooling water.

$$Qh=h2s-h3$$

3-4: An irreversible throttling process in which the temperature and pressure decrease at constant enthalpy.

$$h3=h4$$

4-1: An internally, reversible, constant pressure heat interaction in which the working fluid is evaporated to a saturated vapour at state point 1. The latent enthalpy necessary for evaporation is supplied by the refrigerated space surrounding the evaporator.

The amount of heat transferred to the working fluid in the evaporator is called the refrigeration load.

$$q_L=h_1-h_4$$

The thermal efficiency of the cycle can be calculated as:

$$\text{Efficiency}=q_{evap}/w_{comp}=h_1-h_4/h_{2s}-h_1$$

Refrigeration Cycle: PH diagram

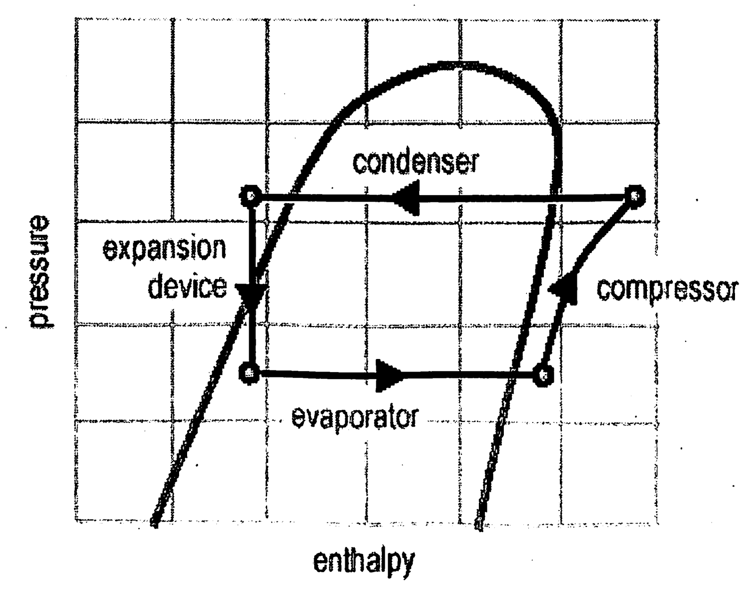


Figure6: P-H Diagram

- A diagram of a typical vapor-compression refrigeration cycle is superimposed on a pressure-enthalpy (P-*h*) chart to demonstrate the function of each component in the system.
- The **pressure-enthalpy chart** plots the properties of a refrigerant—refrigerant pressure (vertical axis) versus enthalpy (horizontal axis).

Major Components in Vapour Compression Cycle

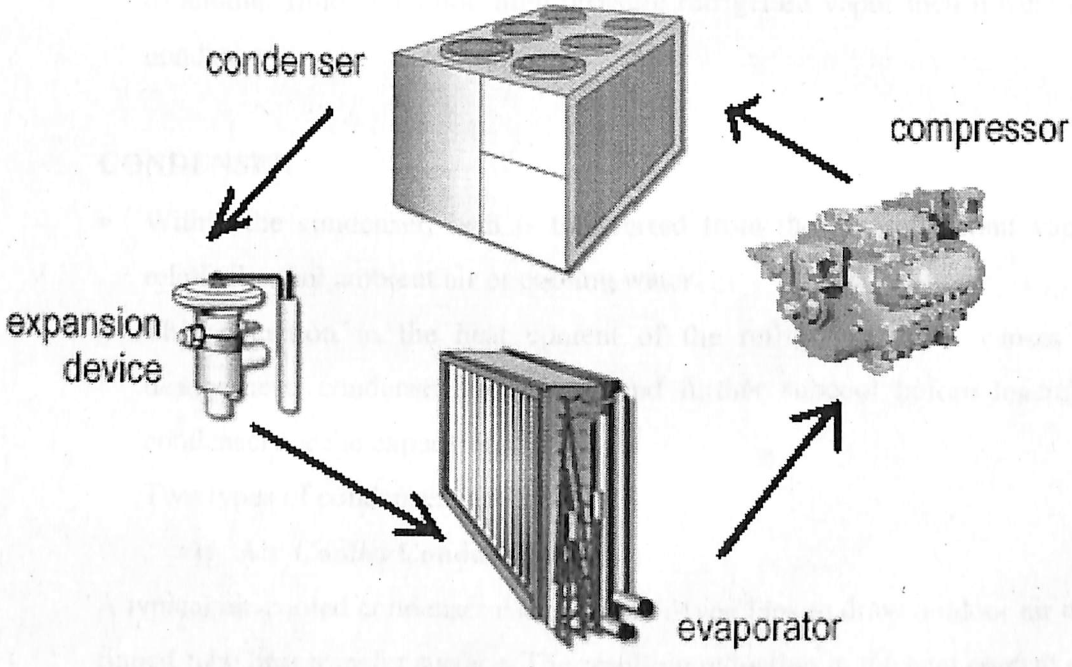


Figure7: Major component in Vapor-Compression Cycle

EVAPORATOR:-

- When the low-pressure mixture of liquid and vapor refrigerant entering the evaporator where it absorbs heat from the relatively warm air, water, or other fluid that is being cooled. This transfer of heat boils the liquid refrigerant in the evaporator, and this superheated refrigerant vapor is drawn to the compressor.

COMPRESSOR

- The compressor draws in the saturated refrigerant vapor and compresses it to a pressure and temperature high enough & superheated it that it can reject heat to another fluid. This hot, high-pressure refrigerant vapor then travels to the condenser.

CONDENSER

- Within the condenser, heat is transferred from the hot refrigerant vapor to relatively cool ambient air or cooling water.
- This reduction in the heat content of the refrigerant vapor causes it to desuperheat, condense into liquid, and further subcool before leaving the condenser for the expansion device.

  Two types of condensers are there:

  1. **Air-Cooled Condensers**

A typical air-cooled condenser uses propeller-type fans to draw outdoor air over a finned-tube heat transfer surface. The resulting reduction in the heat content of the refrigerant vapor causes it to condense into liquid.

  2. **Water-Cooled Condensers**

- The shell-and-tube is the most common type.
- Water is pumped through the tubes while the refrigerant vapor fills the shell space surrounding the tubes.
- As heat is transferred from the refrigerant to the water, the refrigerant vapor condenses on the tube surfaces.
- Hot refrigerant vapour enters the water cooled condenser at the top

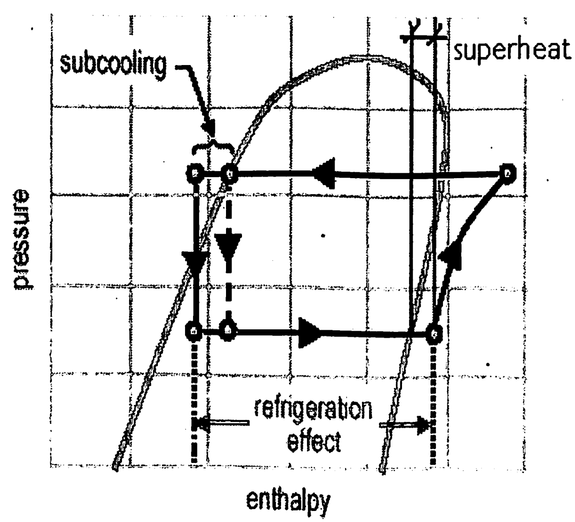
- The condensed liquid refrigerant then falls to the bottom of the shell at which is subcooled by the subcooler

EXPANSION DEVICE

- The high-pressure liquid refrigerant flows through the expansion device, causing a large pressure drop that reduces the pressure of the refrigerant to that of the evaporator. This pressure reduction causes a small portion of the liquid to boil off, or flash, cooling the remaining refrigerant to the desired evaporator temperature. The cooled mixture of liquid and vapor refrigerant then enters the evaporator to repeat the cycle.

REFRIGERATION EFFECT

- The change in enthalpy that occurs in the evaporator is called the **refrigeration effect.**
- It is the amount of heat that each kg of liquid refrigerant will absorb when it evaporates.
- In comparison, the same system without subcooling produces less refrigeration effect.
- The system without subcooling must evaporate substantially more refrigerant within a larger coil to produce the same capacity as the system with subcooling.



REFRIGERANTS

Definition of Refrigerant:

A refrigerant is a fluid used for heat transfer in a refrigeration system.

- Refrigerants absorb heat during evaporation at low temperature and low pressure and reject heat during condensation at a higher temperature and higher pressure.
- Some refrigerants produce a refrigeration effect when they throttle and expand in the refrigeration cycle.

Classification of Refrigerants

Refrigerants most commonly used refrigeration systems can be classified into four groups:

- Hydrocarbons
- Halocarbons
- Azeotropes
- Inorganic compounds

Hydrocarbons

Refrigerants belonging to the hydrocarbon group are ethane, propane, butane and isobutane.

- They are produced from petroleum in an oil refinery.
- This group of refrigerants is used in the refrigeration systems in oil refineries and the petrochemical industry due to their low cost and ready availability.
- Hydrocarbons are flammable and so safety precautions are of utmost importance in the petrochemical industry.

Selection of a suitable refrigerant

It depends on many factors, such as the evaporating temperature required during operation, the coefficient of performance COP, safety requirements, and the size and location of the refrigeration plant.

In order to select a suitable refrigerant for a refrigeration system of known size and evaporating temperature, the following factors must be considered:

- The volume flow rate required per kW of refrigeration capacity
- The coefficient of performance COP
- Safety requirements
- Physical properties
- Operating properties
- Cost.

Ideal refrigerants

- Having the desired thermodynamic properties,
- Nontoxic
- Non-flammable
- Completely stable inside a system
- Environmentally benign-even with respect to decomposition products- and abundantly available or easy to manufacture
- Self-lubricating
- Compatible with other materials used to fabricate and service refrigeration systems
- Easy to handle and detect
- System not operating at extreme pressures, either high or low

EQUIPMENTS USED IN LIQUEFACTION PROCESS

There are two types of equipment that are the key to the liquefaction process: the compression equipment and the heat exchange equipment.

Compressors & Drivers

Two types of compressors are employed:

- Centrifugal.
- Axial.

The axial type is just starting to be used in existing plants, but is being considered for the mixed refrigerant type plants. Single line units of up to 135,000 hp have been built. The centrifugal type compressors have been used for the existing plants, and are generally being considered for both the Cascade and MR plants now under study. The centrifugals used are of fairly standard, proven design, of large size but for relatively low pressure. MR type plants require a discharge pressure on the order of 450 psig, and for the Cascade type plant, the discharge pressure range about 200 psig for propane, 230 psig for ethylene, and 620 psig for methane.For the large Cascade type plants, the centrifugal compressor lines being considered range up to 55,000 hp requiring one body for propane, two bodies for methane, and two bodies for ethylene. For MR type plants, the numbers of compressor bodies vary widely depending upon the manufacturer and the line size. The compressors are provided with intermediate nozzles for entry of economizer streams.

Drivers used have been of three basic types:

- Steam turbines.
- Gas turbines.
- Electric motors

The gas turbines have the advantage of a simpler installation, little water requirement, and lower initial cost. Gas turbines are available of proven design up to 64,000 hp. Steam turbines are well proven as variable speed compressor drivers up to 30,000 hp. Steam turbine systems may be slightly more efficient than industrial gas turbines without waste heat recovery, but the efficiency comparison varies for each plant situation. It depends on steam conditions, design of steam system, and the particular gas turbine design. Each type of driver has its advantages, and should be evaluated for a specific plant location and situation. Steam turbines have an efficiency that is almost independent of load while normal overall efficiency is about 31% (on LHV). Also, steam turbines easily fit with compressor power and speed requirements.Steam, however, has serious disadvantages because in most sites fresh water has to be recovered from sea water; and because it requires large quantities of cooling water.

Two types of gas turbines are available:

- Industrial gas turbines, heavy duty type, where efficiencies reach 27%.
- Jet type turbines derived from aircraft engines in which efficiencies may exceed 30%.

For cold climates, gas turbines are very attractive because they are efficient and do not require cooling water.

Electric motor drives are being planned for the Western LNG project. The selection of electric motor drives is based on the availability of electrical power and its relative cheapness as compared with gas. Also, there has been considerable experience in using large electric motors in ethane extraction plants and gas compression facilities in Alberta (in the 30,000 hp or bigger size). A thorough review of the appropriate driver and compressor type is most important so as to produce LNG at the minimum cost.

Heat Exchangers

Four basic types of heat exchangers have been employed. These are:

- Air fin coolers.
- Conventional shell and tube.
- Brazed plate-fin type.
- Wound type exchangers

The air fin coolers are only suitable for cooling to within 20°F to 40°F of the ambient air and are used as compressor inter and after coolers. The shell and tube exchangers are normally used at moderate temperatures, down to approximately -50°F, as water coolers and precoolers. These have been discussed in Chapter 8. The cryogenic (cold box) exchangers are normally either of the brazed type or wound type. These will be discussed further.

Brazed Plate-Fin Type - These are made from a stack of layers, each one consisting of corrugated aluminum sheets (fins) between flat aluminum separating plates to form individual fluid passages. Each layer is closed at the edge with solid aluminum bars, and the stack is bonded together by a brazing process to yield an integral rigid structure with a series of passages. These passages normally have integral welded headers. Several sections may be connected together to form one larger exchanger. They are Compact about nine times as much surface per cubic foot as shell and tube exchangers. Weight is kept to a minimum and the design pressures go up to 1440 psig and from cryogenic temperatures to +150°F. This type of exchanger is used extensively in ethane extraction plants.

Wound Type –

Wound type exchangers are made by spirally winding layers of small aluminum tubes, with spacers. The tubes tie into headers and the shell side (low pressure) fluid flows through the bundle, around the tubes. The tube side may be provided with headers to allow multiple separate flows and functions. They can be made huge with multiple bundles in one shell, eliminating some headers, piping etc .They can be built for higher tube side pressures, up to about 2,000 psig. Spiraling of tubing eliminates contraction problems. The maximum present size of this equipment is about 15 feet (4.6m) in diameter, 250 feet (76m) long, and weighs about 400 tons (367 tonnes). This size is restricted primarily by transportation restrictions.

2. NATURAL GAS LIQUEFACTION PROCESS-CONVENTIONAL

There are three main types of liquefaction processes. These are the expansion system, the cascade system, and the mixed refrigerant system. These are discussed further on.

1. Cascade Liquefaction process

The classical Cascade process reduces the irreversible heat exchange losses by utilizing several refrigeration cycles whose refrigerants vaporize at different but constant temperatures. The process flow sheet may be found on Figure 9. Propane is liquefied by heat exchange with cooling water and then expanded in stages to 5, 2.5, and 1 atmosphere with the result that cooling zones at temperatures of about 0°C, -20°C, -40°C respectively are present in the propane cycle vaporizers. In this way, natural gas and cycle gas are cooled in stages. Propane vaporized at the various expansion levels will be cycled to corresponding inlets in the propane recycle compressor. Ethylene is compressed to about 310 psia and is liquefied in the coldest of the propane evaporators, and is then used as a refrigerant - again in three stages of cooling. In the coldest ethylene evaporator, methane and natural gas are liquefied. The liquefied methane is used at three different temperature levels as a refrigerant, in the same way as propane and ethylene are used in the early process. Following condensation, the methane for the three cooling stages will be returned to the methane cycle compressors. Liquefied natural gas taken off in the last and coldest of the methane cycle vaporizers is expanded to tank storage pressure. The specific energy requirements of the classical Cascade liquefaction cycle approach the upper limit of maximum attainable thermodynamic efficiency; however, the requirement for machinery, heat exchangers, separators and condensers, and control systems is exceptionally high. The first two base load liquefaction plants at Kenai in Alaska and Arzew in Algeria used this liquefaction cycle.

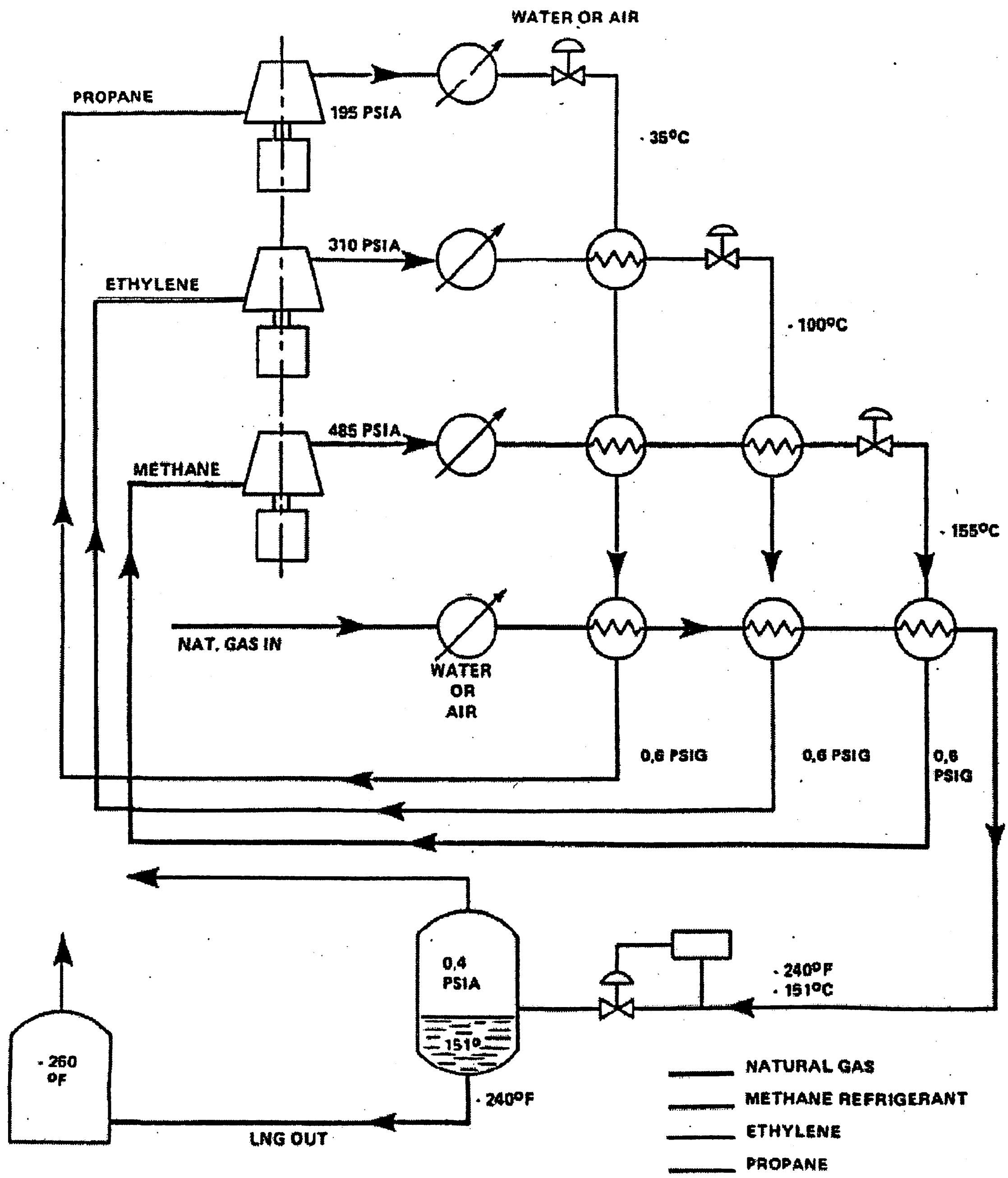


Figure 8: Liquefaction of Natural Gas Classical Cascade

Feature of Cascade Process

- Many plant still being designed and built using the cascade process – simple and reliable
- Three pure components used for refrigeration: Propane pre-cooling
Ethylene
Methane
- Propane pre-cooling
  - -Centrifugal compressors
  - - Typically 2 x ~30 MW Gas Turbines (e.g. Frame 5)
- Ethylene and Methane cycles
  - - Centrifugal compressors
  - - Typically 2 x ~30 MW Gas Turbines (e.g. Frame 5) for each cycle

3. Air Products Propane Precooled Mixed Refrigerant Process

The flow sheet for this process may be found on Figure10. The gas stream is cooled directly as shown. The refrigeration stream flow is more complicated and is as follows: The low pressure refrigerant stream leaving the cryogenic heat exchanger is compressed to a high pressure in a compressor and cooled to about -35°C by propane refrigeration in the evaporators. The partially condensed stream is separated into liquid and vapour fractions in a phase separator. The liquid is cooled in the warm bundle of the cryogenic heat exchanger, flashed across a pressure letdown valve, and distributed over the shell side of the exchanger. The vapour stream is condensed and subcooled in both the warm bundle and the cold bundle, flashed across a letdown valve, and distributed over the shell side of the cold bundle. The stream from the bottom of the cold bundle mixes with the low pressure stream before redistribution over the warm bundle, The propane streams from the high, medium, and low level evaporators of the plant are compressed in a single multi-stage compressor system. The high pressure propane is cooled and condensed against cooling water or air and stored in an accumulator that supplies propane to the evaporators. The three primary operating parameters for the entire mixed refrigerant system are: composition, suction pressure, and discharge pressure. In the propane system, the low level evaporator pressure is set to obtain the desired temperature of the feed and mixed refrigerant streams to the cryogenic heat exchanger; the high level evaporator pressure is set sufficiently high to avoid hydrate formation in the feed gas; and the medium level evaporator pressure is allowed to float as dictated by compressor performance. The cooling water temperature determines the discharge pressure of the propane compressor. Thus, the only primary control parameter in the propane cycle is compressor speed. Air Products have reported that the propane precooled mixed refrigerant process can be operated to accommodate large turndown in LNG production. At reduced capacity the efficiency of the mixed refrigerant system can be maintained by changing composition and lowering suction pressure. Similarly, by adjusting the composition of the mixed refrigerant and changing the suction and discharge pressures of the mixed refrigerant compressor, two different feed streams can be liquefied at the same plant without any significant reduction of efficiency. Their conclusion was, that by adjusting the primary operating variables of composition, suction pressure, and discharge pressure, close to optimum efficiency can be maintained even with significant changes in LNG production rates, LNG composition, and off-design equipment performance.

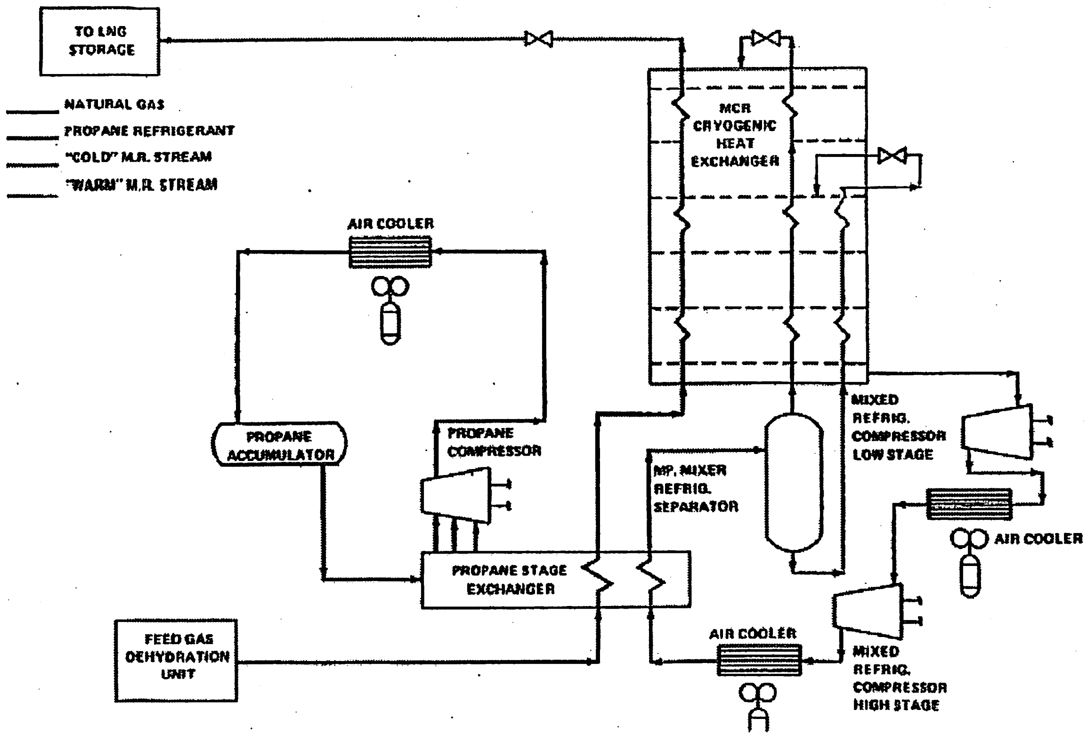


Figure 9: Liquefaction of Natural Gas Air Products Propane Precooled Mixed Refrigerant Process

Features of APCI Process

- Most of existing plant are using the APCI process with 3 – 3.3 MTPA Fr 6 / Fr 7 combination
- Train capacities up to 4.7 MTPA built or under construction using Fr 7 / Fr 7 combination
- Higher Capacities to 7.9 MTPA being announced with Frame 9 GT
- Two main refrigeration cycles: Propane pre-cooling

  Mixed refrigerant liquefaction and sub-cooling
- Propane pre-cooling

  -Centrifugal compressor (to 15 – 25 bar)

  -Side-streams at 3 pressure levels

  -Typically requires a ~40 MW Gas Turbine (e.g. Frame 6) plus Helper Motor or Steam Turbine

-Compressor sizes reaching maximum capacity limits Added aerodynamic constraint; high blade Mach numbers due to high mole weight of propane (44) Prevents utilisation of full power from larger gas turbines (Frame 7)

- Mixed refrigerant liquefaction and sub-cooling
  -Axial LP for Shell Advised Plant
  -Centrifugal HP compressor (45 – 48 bar)
  -Typically requires ~70 MW Gas Turbine (e.g. Frame 7) plus Helper Motor or Steam Turbine
- Mixed refrigerant liquefaction and sub-cooling
  -Large volumetric flows
  -Two casing arrangements (LP and an HP)
  -Axial LP / centrifugal HP compressor (45 – 48 bar)
  -Typically requires ~70 MW Gas Turbine (e.g. Frame 7) plus Helper Motor or Steam Turbine
  -LP and HP compressor speeds compromised
  -LP axial compressor (higher efficiency)
  -HP centrifugal compressor

3. EMERGING TECHNOLOGIES

- **Phillips optimized cascade process**

This process, a modified version of a process used in an earlier plant in Alaska during the 1960s, was used for the Atlantic LNG plant in Trinidad and for a baseload plant under construction in Egypt. Train capacities of up to 3.3 million tpy have been constructed with larger trains in development. This process is illustrated in Figure 4.Refrigeration and liquefaction of the process gas is

achieved in a cascade process using three pure component refrigerants; propane, ethylene and methane, each at two or three pressure levels. This is carried out in a series of brazed aluminum PFHEs arranged in vertical cold boxes. Precooling could be carried out in a core-in-kettle type exchanger. The refrigerants are circulated using centrifugal compressors. Each refrigerant has parallel compression trains. Frame 5 gas turbine drivers were used.

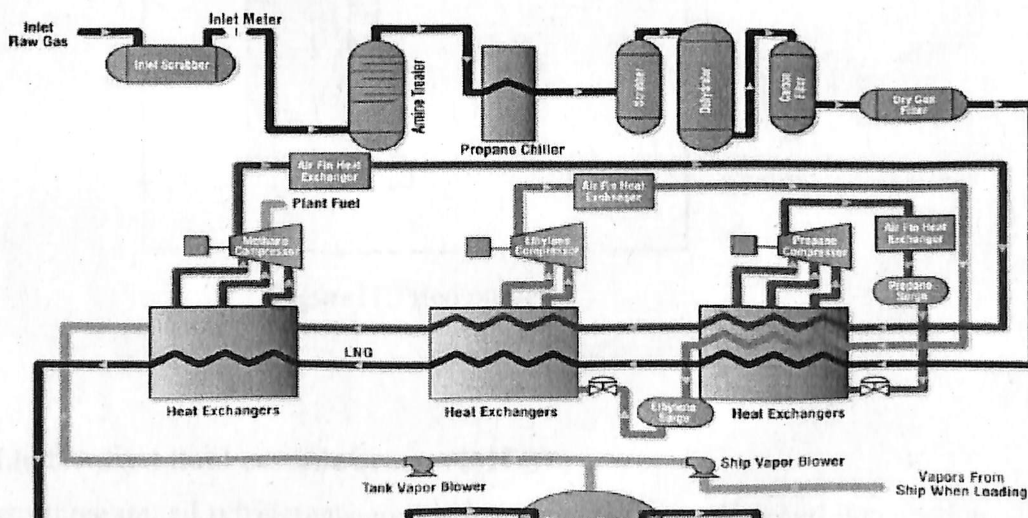


Figure10: Phillips Cascade Process

- **Black & Veatch PRICO process**

This is a single mixed refrigerant process used on an earlier baseload plant in Algeria. Train capacity has been uprated to 1.3 tpy per train. It is illustrated in Figure 5. The mixed refrigerant is made up of nitrogen, methane, ethane, propane and iso-pentane. The cooling and liquefaction is carried out at several pressure levels, in PFHEs in cold boxes. The refrigerant is compressed and circulated using a single compression train. In the Algerian plant axial compressors driven by steam turbines were used.

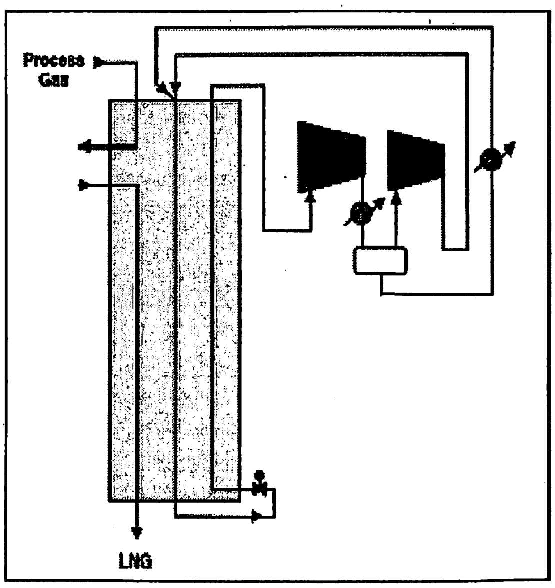


Figure11:Prico process

- **Statoil/Linde mixed fluid cascade process (MFCP)**

In this process three mixed refrigerants are used to provide the cooling and liquefaction duty. It has been selected for the Snøhvit LNG project (Ekofisk, Norway) which is under design/construction. This is a single train 4 million tpy LNG plant. The process is illustrated in Figure 6. Pre-cooling is carried out in PFHE by the first mixed refrigerant, and the liquefaction and sub cooling are carried out in a spiral wound heat exchanger

(SWHE) by the other two refrigerants. The SWHE is a proprietary exchanger made by Linde. It may also be used for the pre-cooling stage. The refrigerants are made up of components selected from methane, ethane, propane and nitrogen. The three refrigerant compression systems can have separate drivers or integrated to have two strings of compression. Frame 6 and Frame 7 gas turbine drivers have been proposed for large LNG trains (> 4 million tpy). A novel feature of the Snøhvit project is that all motor drivers will be used for the main refrigerant compressors, with sizes up to 60 MW. The SWHE itself is being installed with other liquefaction processes, in new and expansion projects or as a replacement for old cryogenic exchangers.

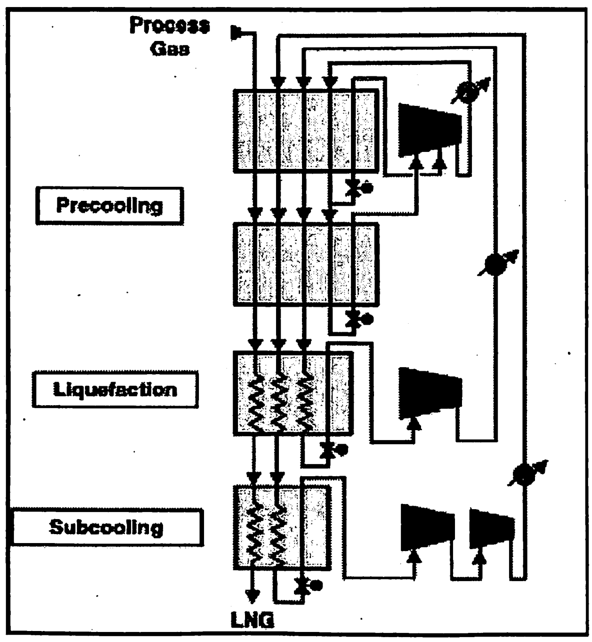


Figure12: Linde Mixed Fluid Cascade Process

Linde Process

- Mixed refrigerants for pre-cooling, liquefaction and sub-cooling duties
- Minimum of Three Gas Turbine or electric motors needed for compressor driver
- 4.3 MTPA plant under construction with VSD motor drivers and onsite power generation with aero-derivative gas turbines.

- **Axens Liquefin™ process**

This is a two-mixed refrigerant process, which is being proposed for some new LNG base load projects of train sizes up to 6 million tpy. It is illustrated in Figure 7. Detailed studies have been made including input from main equipment vendors. All cooling and liquefaction is conducted in PFHE arranged in cold boxes. The refrigerants are made up of components from methane, ethane, propane, butane and nitrogen. The first mixed refrigerant is used at three different pressure levels to precool the process gas and precool and liquefy the second mixed refrigerant. The second mixed refrigerant is used to liquefy and subcool the process gas. Using a mixed refrigerant for the precooling stage allows a lower temperature to be achieved (for example, -60 °C) depending on refrigerant composition. The PFHEs are non-proprietary and can be supplied by independent vendors. Two large drivers can drive the refrigerant compression systems. Frame 7 gas turbines are being proposed for the large LNG trains.

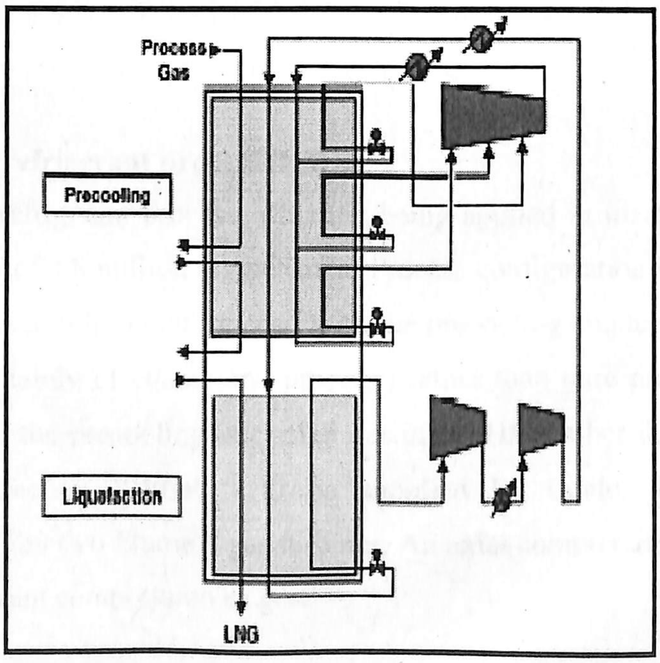


Figure13: Liquefin Process

Features of Axens Liquefin Process

- Mixed refrigerants for pre-cooling, liquefaction and sub-cooling duties

- Liquefin development studies presently oriented towards increasing capacity to 6 MTPA with: 2 x Frame 7 Gas Turbines for main compression

  2 x Frame 5 Gas Turbines for power generation
- Higher capacities possible using: Frame 9 GT's ,

  Electric motors,

  Steam-turbines etc.
- Similar to APCI with Propane compressor replaced with Mixed Refrigerant for pre-cooling.
- Allows more balanced flows, refrigeration loads and power between the two compressors.
- Avoids the process design limits associated with Propane compressors.

• Shell double mixed refrigerant process (DMR)

This is a dual mixed refrigerant process, which is being applied in the Sakhalin Island project with a capacity of 4.8 million tpy per train. Process configuration is similar to the propane pre-cooled mixed refrigerant process, with the precooling conducted by a mixed refrigerant (made up mainly of ethane and propane) rather than pure propane. Another main difference is that the precooling is carried out in SWHEs rather than kettles. The precooling and liquefaction SWHEs will be supplied by Linde. The refrigerant compressors are driven by two Frame 7 gas turbines. An axial compressor is also used as part of the cold refrigerant compression stages.

Features of Shell DMR Process

- Similar to Axens but with twin parallel compressor trains for each process stream
- Use of aero-derivative or VSD motors
- Shell claim 4.5 - 5.5 MTPA and lower cost

CYCLE EFFICIENCY COMPARISON

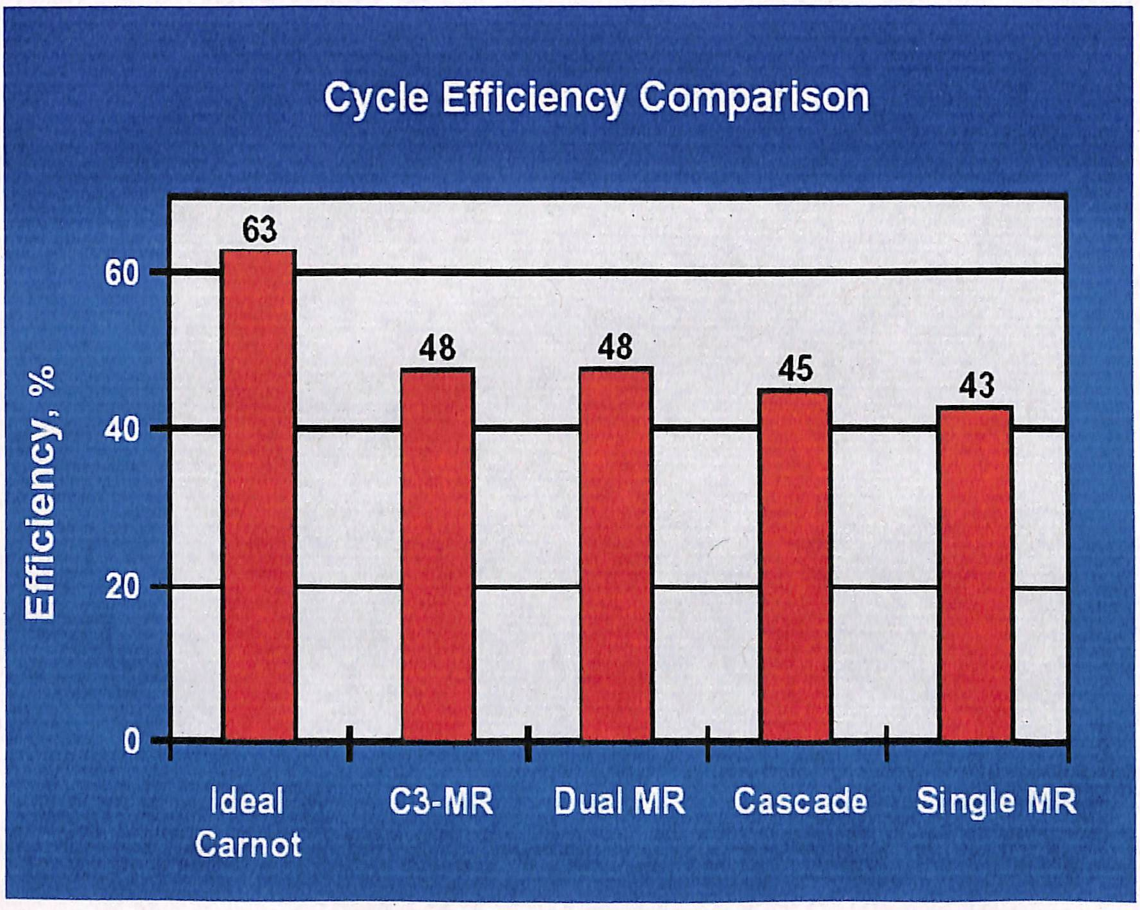


Figure14: Liquefaction Cycle Efficiency Comparison

Above figure shows that both the mixed refrigerant process i.e. C3MR and DMR, have higher cycle efficiency as compared to pure refrigerant process like-Cascade

PROCESS SELECTION PARAMETER

To compare one process with another, it is necessary to be very careful about several parameters, which can change hugely the result: end flash quantity, compressor efficiencies, condenser temperature approach, LPG recovery

- **END-FLASH QUANTITY**:-is used to reduce nitrogen content in the LNG and provide fuel gas. The quantity is strongly related to plant fuel gas consumption. If it is possible to increase this quantity through fuel gas export to other plants, recycle, etc. the cold end temperature of the main exchanger line will increase, and the efficiency of the plant , thus the quantity of LNG produced will increase. If for any reason no end-flash is desired, there will be a large reduction in LNG production. In any cases however, the quantity of fuel gas cannot be decreased below a certain quantity because of the nitrogen content of the feed gas. The figure shows the production variation with the quantity of end-flash. If the end-flash can be increased by 40%, the LNG production will be increased by 5% with the same power on the refrigeration compressors. This also must be checked carefully for process comparison.

- **COMPRESSOR EFFICIENCY**:- has a significant effect on LNG production. LNG production is increased by nearly 10% when the polytrophic efficiency is changed from 79 to 85%.Realistic polytrophic efficiencies must be chosen when making comparison.

For any process, a large condenser in the first refrigerant cycle removes the heat produced by the refrigeration compressors. At the outlet of this condenser ia at the bubble point, modifying the condenser outlet temperature will change the discharge pressure of the corresponding compressor, its power consumption and finally the overall process efficiency. The temperature approaches in the other cooler will also have an impact although less, on power consumption.

Closer the temperature approaches, larger the LNG production.

COMPRESSOR PERFORMANCE CURVE

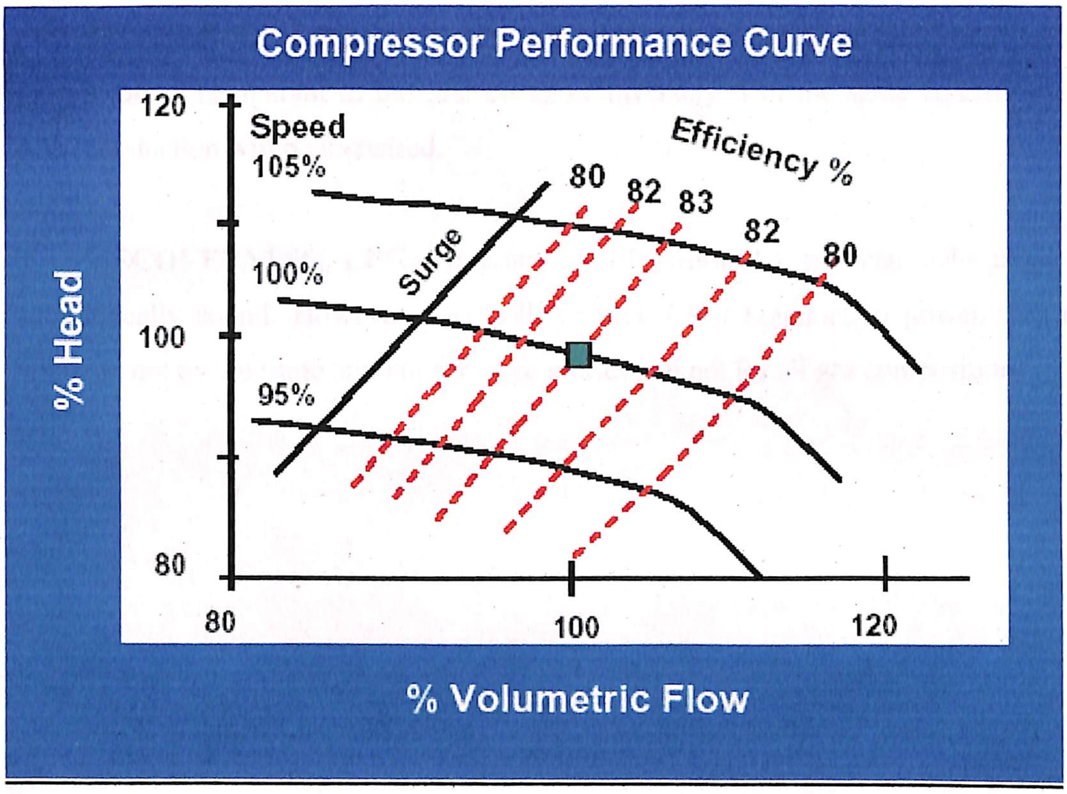


Figure15: Compressor Performance Curve

- **Temperature Approach on The Main Condenser**

Whatever the process, a large condenser on the first refrigerant cycle must evacuate the heat produced by the refrigeration compressors. As the outlet of this condenser is at bubble point, to modify the outlet temperature of this condenser will change the discharge pressure of the corresponding compressor, so the power of this compressor and the overall efficiency. The temperature approach of the other coolers will also have an impact on the power, but less important than this one.

We have plotted for a Liquefin case the capacity versus the temperature approach of the condenser. The closer the temperature approach, larger the LNG production. However, in air cooling case, the size of this condenser can be a problem, as it governs more or less the size of the plant area, so a part of the cost.

The size of the condenser will depend upon whether the refrigerant of the first cycle is a pure component, or a mixed refrigerant. With a pure component, the condensation is done at a fixed temperature (the dew point temperature is the same as the bubble point temperature), whereas with mixed refrigerant the temperature varies linearly between the dew-point temperature and the bubble point temperature. Either the condenser will be much smaller with a mixed refrigerant in the first cycle, or inversely with the same condenser size, the LNG production will be increased.

- **RECOVERYING LPG:-**To recover LPG from the gas can help make project economically sound. However, this will increase LNG liquefaction power consumption, although not by the same amount for all processes and not for all gas composition

V. TECHNO-ECONOMIC ASSESSMENT

1. LIQUEFACTION PLANT COST

The capital cost of a liquefaction plant is a critical component of the overall cost of an LNG delivery chain. In fact, total costs of a facility can run as high as $2 billion. While this is certainly a huge expenditure, costs on a per unit basis have dropped significantly in the last 26 years. The initial liquefaction plants were small in size compared to those in the planning stages today, with no trains over 2 mtpa built until the 1990s. As the LNG trade became more than a small niche market, plant owners began looking for ways to lower costs. One key was to take advantage of economies of scale by building larger facilities. As you can see from the chart on figure, the cost of liquefaction has fallen significantly – from over $500 a tonne in 1988 to below $200 a tonne today. This drastic reduction in costs is due to a number of influences. Certainly, technology was pushed to gain economies of scale. But in addition, organizational learning, research and development, project management, and technology supplier competition had a hand in reducing the cost of liquefaction. And we may still see future cost reductions. Gains are not likely to come from component prices since steel and concrete have experienced 100% to 400% price increases during the last two years. Construction time, however, is actually more expensive than the materials required building the asset.

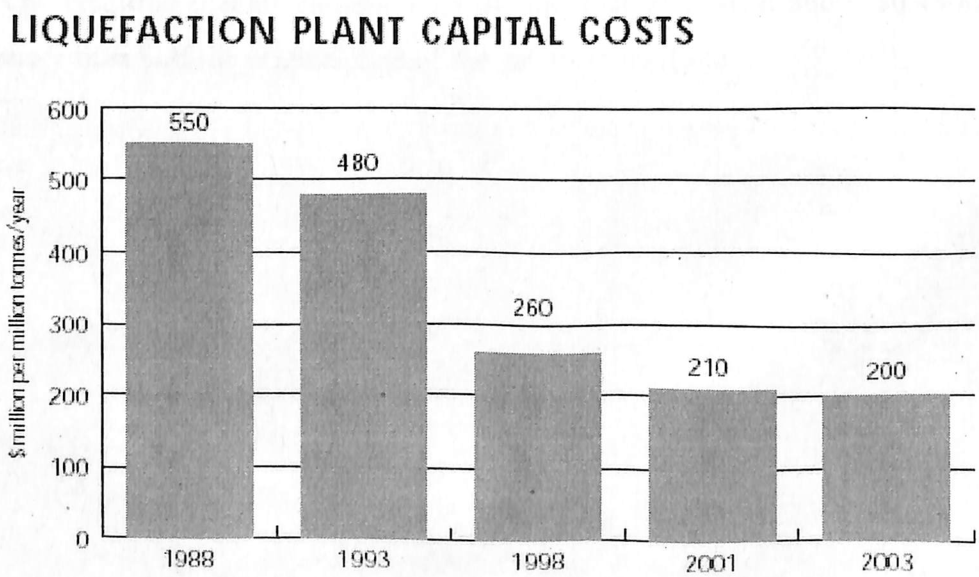


Figure15: Liquefaction Plant Capital Cost

Economies of scale are also a significant factor in obtaining reductions in liquefaction unit costs. Many of the new plants being considered will include trains in the 4 mtpa, 5 mtpa, and 8 mtpa range. While larger trains can accomplish a substantial reduction in costs, there are concerns. One of the major hurdles is the significant reduction in LNG production that occurs when one of these super-size compressor trains goes out of service unexpectedly or for planned maintenance. Thus operational flexibility can be an important factor in design decisions. The future costs and reliability of mega trains remain to be seen, and at this point the jury is still out on their feasibility, especially given cost and reliability concerns.

Variable Costs

Ongoing costs to operate a liquefaction unit are also an important factor in the overall cost of liquefaction. Important factors include use of natural gas as fuel in the liquefaction plant, taxes paid to the local government and general operating and maintenance (O&M) costs. A typical liquefaction unit might use 11% of the plant's input gas as fuel. If we assume a fuel cost of $0.75/Mcf (current supply costs range from $0.55 to about $1/Mcf) then the operating cost associated with use of fuel is approximately $0.08/Mcf. Taxes will vary depending on where the facility is located but night be on the order of $0.15/Mcf and O&M costs are typically about $0.20/Mcf. The resulting overall variable cost of liquefaction is then about $0.43/Mcf, which is more than half the original cost of the gas from the field.

2. TECHNO-ECONOMIC COMPARISON: APCI & AP-X™

Economies of scale continue to favor increasing train size for baseload LNG plants in order to drive down the unit cost of LNG. The AP-X™ LNG Process achieves not only high capacity in a single train, but also can incorporate high LPG recovery, lower LNG heating value for new markets, and maximum efficiency. This range of applications demonstrates that the AP-X™ LNG Process brings significant economies of scale to the industry, reducing the capital cost of LNG while maintaining the efficiency, flexibility, and reliability of the C3MR process.

AP-X™ LNG PROCESS

While the C3MR process remains the preferred option in many cases, there exists substantial developing demand for larger train sizes. For example, trains using multiple GE Frame 7 or Frame 9 gas turbine drivers or large electric motors can be configured. While it remains feasible to further increase train capacity with a C3MR process, new designs must be developed for several major equipment items at capacities exceeding 5.0 Mta. For example, the propane and centrifugal MR compressors are approaching single casing flow limits at current world scale LNG plant production levels. In response to continuing customer demand for increased LNG train capacity and lower unit cost, Air Products has developed and patented the AP-XTM LNG Process.

The AP-XTM process cycle is an improvement to the C3MR process in that the LNG is subcooled using a simple, efficient nitrogen expander loop instead of mixed refrigerant. Other embodiments include a dual MR version where another MR refrigeration loop is used for pre-cooling and nitrogen is likewise used for subcooling.

In addition to improving the efficiency, the use of the nitrogen expander loop makes greatly increased capacity feasible. It does this by reducing the flow of both propane and mixed refrigerant. Volumetric flow of mixed refrigerant at the low-pressure compressor suction is about 60% of that required by the C3MR process for the same production. Mass flow of propane is about 80% of that required by the C3MR process.

With the new AP-XTM process, train capacities in excess of 8 Mta are feasible in tropical climates, in existing compressor frame sizes, without duplicate/parallel compression equipment, and using a single spool-wound MCHE of a size currently being manufactured.

The nitrogen expander loop is a simplified version of the cycle employed by Air Products in hundreds of air separation plants and nitrogen liquefiers worldwide. Experience has shown these plants to be simple to operate and very reliable. Many of these plants are remotely operated, including shutdowns and restarts. The nitrogen cycle has also been employed by Air Products with similar success in small, stand-alone LNG peak-shaving plants.The AP-XTM process cycle is depicted below in figure 2. As is the case with C3-MR process, propane is used to provide cooling to a temperature of about -30 °C. The feed is then cooled and liquefied by mixed refrigerant, exiting the MCHE at a temperature of about -120 °C. Final subcooling of the LNG is done using cold gaseous nitrogen from the nitrogen expander. Figure 3 shows the equipment layout for the liquefaction and subcooling sections of an AP-XTM train. Coil-wound heat exchangers are used to liquefy and subcool the LNG, while the nitrogen economizer uses brazed aluminum plate-fin heat exchangers.

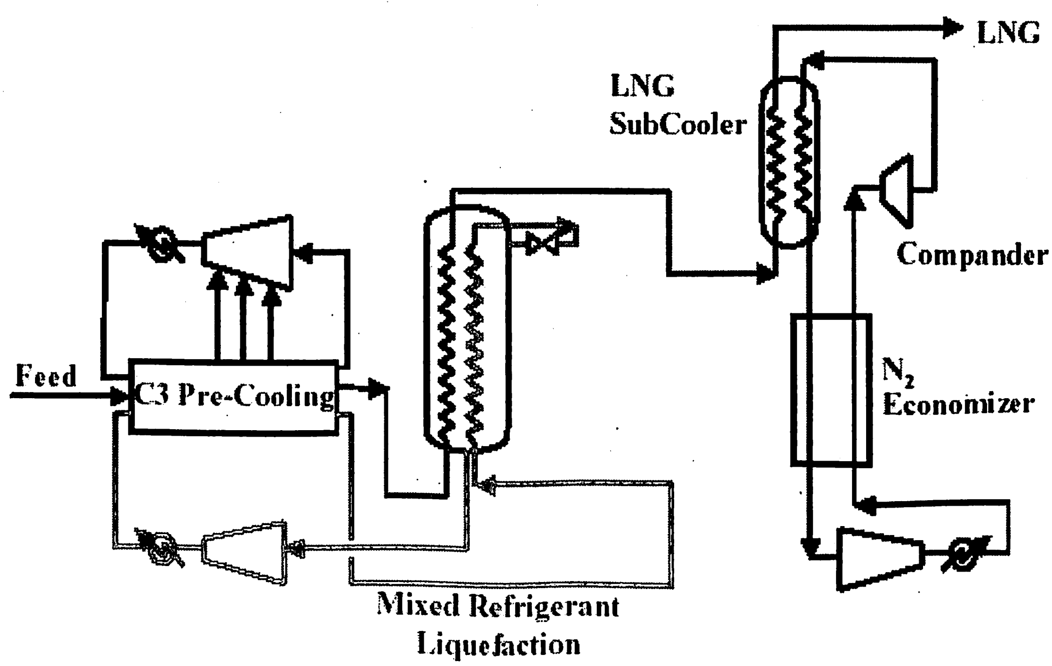


Figure17: AP-X^{TM} Process

It is possible to operate an AP-XTM train at a reduced production rate of about 65% without the nitrogen expander loop by adjusting the composition of the mixed refrigerant inventory. The ability to operate an AP-XTM process plant in C3MR mode can also be exploited as an expandable plant. For example, a producer may choose to invest in a nominal 5 Mta C3MR train with plans to expand production later to a level of up to 8 Mta by adding the nitrogen expander cycle as the market develops.

Retrofitting an existing plant with a nitrogen expander cycle to subcool the LNG is also possible, although the production increase will be more modest due to bottlenecks with existing equipment.

COMPARISON: APCI'S C3-MR AND AP-X™ TECHNOLOGY

	APCI'S C3-MR	AP-X™
TECHNOLOGY	Well-Known	New and Complex
No. Of Refrigeration Cycle	• 2 • Propane and Mixed refrigerant cycle	• 3 • Propane, Mixed refrigerant and nitrogen cycle
COMPRESSOR	Centrifugal	Centrifugal
COMPRESSOR DRIVER	• Propane compressor-40 MW Gas-Turbine (Frame6)+helper motor • MR-70MW-frame7 GT + helper motor	• Propane compressor-102MW Frame9 Gas Turbine • MR-Compressor-102MW Frame9 Gas Turbine • Nitrogen Compressor-70MW Frame7 Gas turbine
LPG-EXTRACTION	No	Yes
HEAT-EXCVHANGERS	Spiral wound heat exchangers	• Propane-Kettle type • MR-Spiral wound • Nitrogen-Plate-Fin Heat Exchanger
MAXIMUM CAPACITY	4Mtpa	7 to 8 Mtpa

Table: Technical Comparison between APCI's C3-MR and AP-X™

ECONOMIC ASSESSMENT:-

The AP-XTM meets the industry demand for the economies of scale that can result from larger single train capacities. The unit cost of LNG production is lowered significantly with the AP-XTM process. Merlin Associates have recently completed an independent study comparing the cost of a single 8 Mta AP-XTM train to that of two 4 Mta C3MR trains. The results are summarized in Table 1. Overall plant facility costs are reduced by about 11% with a total cost saving of about 140 MM$. For the C3MR option, some process units such as acid gas removal and the endflash system could potentially be combined depending on owner preference. The resulting savings are only about 10 MM$3 however.

Table6: capital Cost Comparison

Estimate Basis		
Process	**C3/MR**	**AP-X**
Plant Capacity, mtpa	8	8
Number of trains	2	1
Train capacity, mtpa	4	8
Capital Cost-millions US$		
Plant Facilities	1,272	1,131
Marine Facilities	24	24
Temporary Infrastructure	52	52
Total	1,348	1,206
US$/tpa	168	151

VIII. OPTIMISATION

The classical Cascade process reduces the irreversible heat exchange losses by utilizing several refrigeration cycles whose refrigerants vaporize at different but constant temperatures. The process flow sheet may be found on Figure 23.8. Propane is liquefied by heat exchange with cooling water and then expanded in stages to 5, 2.5, and 1 atmosphere with the result that cooling zones at temperatures of about 0°C, -20°C, -40°C respectively are present in the propane cycle vaporizers. In this way, natural gas and cycle gas are cooled in stages. Propane vaporized at the various expansion levels will be cycled to corresponding inlets in the propane recycle compressor. Ethylene is compressed to about 310 psia and is liquefied in the coldest of the propane evaporators, and is then used as a refrigerant - again in three stages of cooling. In the coldest ethylene evaporator, methane and natural gas are liquefied. The liquefied methane is used at three different temperature levels as a refrigerant, in the same way as propane and ethylene are used in the early process. Following condensation, the methane for the three cooling stages will be returned to the methane cycle compressors. Liquefied natural gas taken off in the last and coldest of the methane cycle vaporizers is expanded to tank storage pressure. The specific energy requirements of the classical Cascade liquefaction cycle approach the upper limit of maximum attainable thermodynamic efficiency; however, the requirement for machinery, heat exchangers, separators and condensers, and control systems is exceptionally high. The first two base load liquefaction plants at Kenai in Alaska and Arzew in Algeria used this liquefaction cycle.

Table7: Natural Gas Composition (Feed gas)

COMPONENT	MOLE FRACTION (%)	CRITICAL PRESSURE	CRITICAL TEMPERATURE	MOLECULAR WEIGHT
N2	4	493	227.3	28.013
CH4	87.5	667.8	343.1	16.043
C2H6	5.5	707.8	549.8	30.070
C3H8	2.1	613.3	665.7	44.097
i-C4H10	0.3	529.1	734.7	58.124
n-C4H10	0.5	550.7	765.4	58.124
i-C5H12	0.1	490.4	828.8	72.151

Pseudo Critical Pressure (P_{pc}) = $Y_i\, P_{ci}$ = 650 psia (From KAY'S mixing rule)

Pseudo Critical Temperature (T_{pc})= $Y_i\, T_{ci}$ = 350 ^{0}R

Natural Gas Temperature (T) = 34 ^{0}C = 553.2 ^{0}R

Natural Gas Pressure (P) = 60 bar = 882 psia

Reduced Temperature T_{pr} = T / Tpc = 1.58

Reduced Pressure Ppr = P/ Ppc = 1.30

Specific Heat of Natural Gas Cp= 4.52 cal/gram mole ^{o}K

Density of Natural Gas = 2.37 lbm/ft^3

Mass Flow Rate of Natural Gas (m) = 100 MMSCFD = 9.88 X10^6 lbm /hr =4.40 X10^9

Efficiency of Centrifugal Compressor (E) = 82%

FOR PROPANE CYCLE——

Chiller Duty Calculation ----Q = m Cp T

T = 34 + 31 = 65 0C = 338 ^{0}K

Qc= (4.40 X 10 ^{9}) X (4.52)X (338.0) = 7.4 X10^6 KW

Enthalpies from P-H diagram---:

Enthalpy at point C (saturated Vapor) = h_c =250.0KJ/kg

Enthalpy at point A (saturated liquid) = h_a = 15.0 KJ/kg

Enthalpy at point D (For constant entropy) =340 KJ/kg

Refrigerant mass Flow Rate (Mp) = Qc/hc-ha = 31489.36 kg/sec

Power requirement for Propane Compressor (Wp) = $(h_d^{isen} - h_c)$ Mp/E = 3456149.46 KW = $3.456X10^6$ KW

FOR ETHYLENE CYCLE-----

Chiller Duty Calculation ----Q = m Cp T

T = -31 –(-90)= 59 ^{0}C = 332 ^{0}K

Qc= ($4.40 X 10^9$) X (4.52)X (332.0) = $7.278X10^6$ KW

Enthalpies from P-H diagram---:

Enthalpy at point C (saturated Vapor) = hc =125.0KJ/kg

Enthalpy at point A (saturated liquid) = ha = 36.0 KJ/kg

Enthalpy at point D (For constant entropy) =145 KJ/kg

Refrigerant mass Flow Rate (Me) = Qc/hc-ha = 81777.94 kg/sec

Power requirement for Ethylene Compressor (We) = $(h_d^{isen} - hc)$ Mp/E = 1994583.9 KW = $1.99X10^6$ KW

FOR METHANE CYCLE-----

Chiller Duty Calculation ----Q = m Cp T

T = -90 –(-160)= 70 ^{0}C = 343 ^{0}K

Qc= ($4.40 X 10^9$) X (4.52)X (343.0) = $7.519X10^6$ KW

Enthalpies from P-H diagram---:

Enthalpy at point C (saturated Vapor) = hc =250.0KJ/kg

Enthalpy at point A (saturated liquid) = ha =22.0 KJ/kg

Enthalpy at point D (For constant entropy) =570 KJ/kg

Refrigerant mass Flow Rate (Mm) = Qc/hc-ha = 32978.07 kg/sec

Power requirement for Methane Compressor (Wm) = (h_d^{isen} – hc) Mp/E = 1286940.84 KW = 1.2X 10^6 KW

Total Compressor Power W = Wp + W e +Wm = 6.65 X 10^6 KW

In the Classical Cascade LNG process the refrigerant mass flow rate, the compressor power requirement are higher which make the process less efficient and expensive .This process can be optimize by using suitable driver/compressor configuration, efficient main cryogenic heat exchanger, integrating with NGL extraction and GTL unit, nitrogen removal unit.

OPTIMIZATION OF CLASSICAL CASCADE PROCESS

Classical Cascade process can be optimized by:

- Adding nitrogen removal unit
- Vapor Recovery
- Integrating LNG production with NGL recovery
- LNG and GTL integration
- Suitable driver configuration
- Equipment selection

- **NITROGEN REMOVAL PROCESS-**

Nitrogen removal process allows the direct liquefaction of approximately 6.2% of the natural gas and at the same time allows the temperature of the LNG at the MCHE outlet to be increased by 4 0C, which is the equivalent of an additional 4.8% capacity. In total this process offers an increase in LNG production of 11% with no modification of the liquefaction process. Nitrogen removal process is used to increase the liquefaction plant capacity with no impact on the other cycles of cascade LNG process. Addition of nitrogen removal unit in the cascade LNG process increases LNG production by 11% and the process can be optimize.

- **INTEGRATION OF LNG PRODUCTION WITH NGL RECOVERY**

The combination of changing global markets for natural gas liquids (NGL) with the simultaneous increase in global demand for liquefied natural gas (LNG) has stimulated an interest in the integration of NGL recovery technology with LNG liquefaction technologies.

As such, the integration of the two processes was not a priority. Integrating state-of-the art NGL recovery technology within the CoP LNG% 'Process, formerly the Phillips Optimized Cascade LNG Process, results in a significant reduction in the specific power required to produce LNG, while maximizing NGL recovery.

This corresponds to a production increase in both LNG and NGL for comparable compression schemes as compared to stand-alone LNG liquefaction and NGL extraction facilities. In addition, there are potential enhancements to the overall facility availability and project economics using the integrated concept. In these cases, LNG production has increased by approximately 7%,while using the same process horsepower. --

- **Process Description of an Integrated NGL and LNG plant**

A block diagram for an integrated LNG and NGL process is presented in Figure below. the CoP LNG Process was used for the LNG liquefaction technology. Treated natural gas is first cooled by utilizing refrigeration from within the liquefaction process in one or more stages and then introduced into a distillation column, or Heavies Removal Column. Figure 18 represents the simplest embodiment of NGL integration, where the Heavies Removal Column is not refluxed other than with condensed liquids contained within file column feed. Once the feed has entered the column, it is separated or in this case stripped. A bottoms stream, primarily comprised of NGL components, and a methane rich overhead stream are fonned. The methane rich overhead stream is chilled, condensed, and in most cases subcooled within the liquefaction process. Once liquefied and subcooled, the stream is subsequently flashed to near atmospheric pressure in one or more steps in preparation for LNG storage. Flashed vapor is used as methane.

Proper integration of NGL recovery technology within LNG liquefaction technology results in significant advantages by lowering overall capital cost requirements and improving both LNG and NGL productions. Through careful process selection and heat integration, the integrated LNG& NGL facility results in lower specific power consumption, an approximate 7% LNG production gain is realized through careful integration. Utilization of deethanizer overhead as reflux to the Heavies Removal Column improves separation efficiency while maintaining higher column pressures for efficient

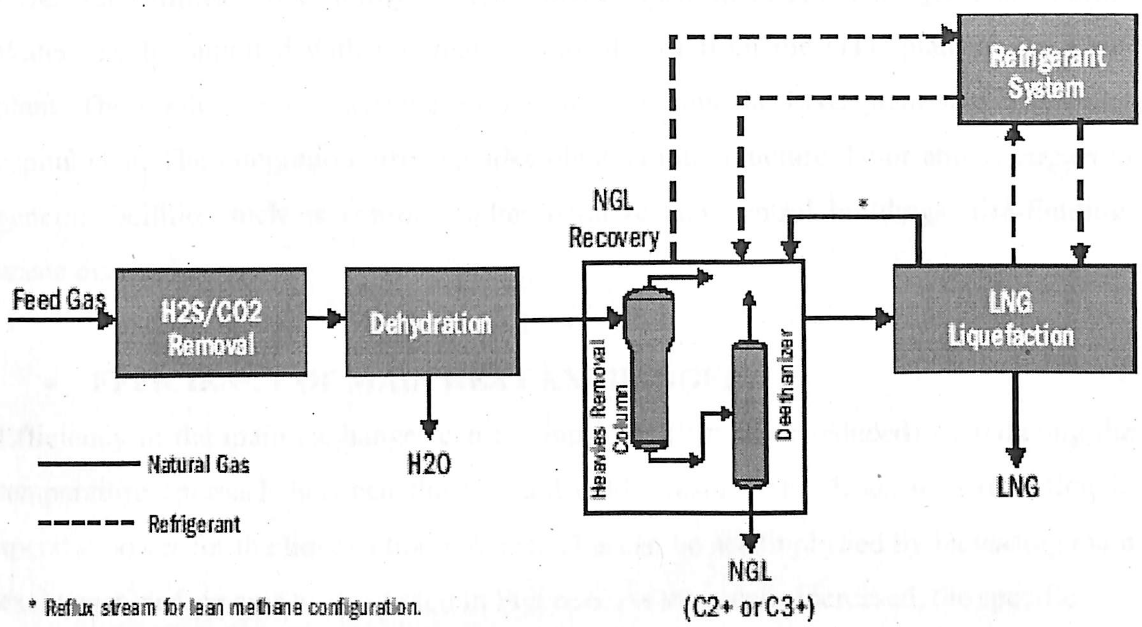


Figure18: Block Diagram showing integrated NGL and LNG process

- **LNG and GTL Integration**

Compared to standalone facilities, a combined GTL-LNG integrated facility would benefit from reductions in both capex and opex. However, there is very little scope for integration within the process units, and most of the integration options reside in the offsites and utilities area. The most significant integration would be in the selection of machinery drives, and in the optimization of the power and steam systems between the two plants.

Some of the key integration advantages are outlined below:

• LNG Product Increased: The LPG product from the GTL facility can be routed to final LNG product thereby eliminating separate LPG handling facilities for GTL LPGs. This also enhances the LNG product quality by increasing the heating value of the LNG.

• Steam/Power Utility Synergies: Integration of steam/power systems (where applicable) can be expected to create significant capital savings. Depending upon the overall project definitions Foster Wheeler studies have shown a reduction of $30-50 million.

• General Utilities: Some utility systems such as Instrument Air, Nitrogen and Cooling Water can be supplied with minimal additional cost from the GTL plant to the LNG plant. These allow certain systems to be removed from the LNG plant scope reducing capital cost. The integration also provides obvious infrastructure, labor and synergies in general facilities such as common administrative and control buildings, fire-fighting, waste disposal etc.

- **EFFICIENCY OF MAIN HEAT EXCHANGER**

Efficiency in the main exchanger can be improved (lost work reduced) by reducing the temperature approach between the hot and cold streams. This leads to a reduction in specific power for the liquefaction process. This can be accomplished by increasing main exchanger surface area as illustrated in Figure 6. As the area is increased, the specific power is reduced. However, as the minimum achievable specific power is approached, the exchanger area increases substantially, indicating that there is an economic optimum.

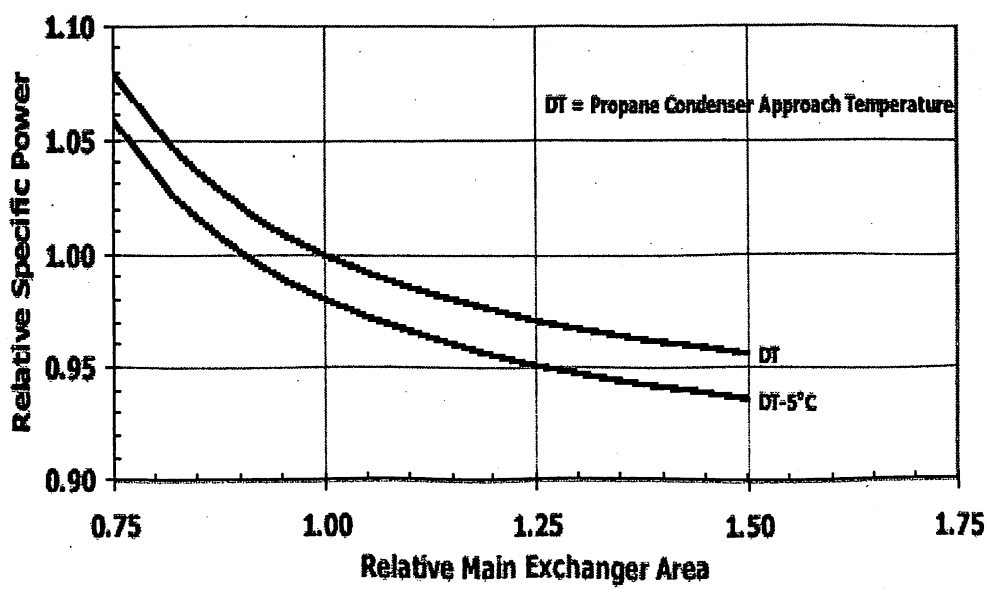
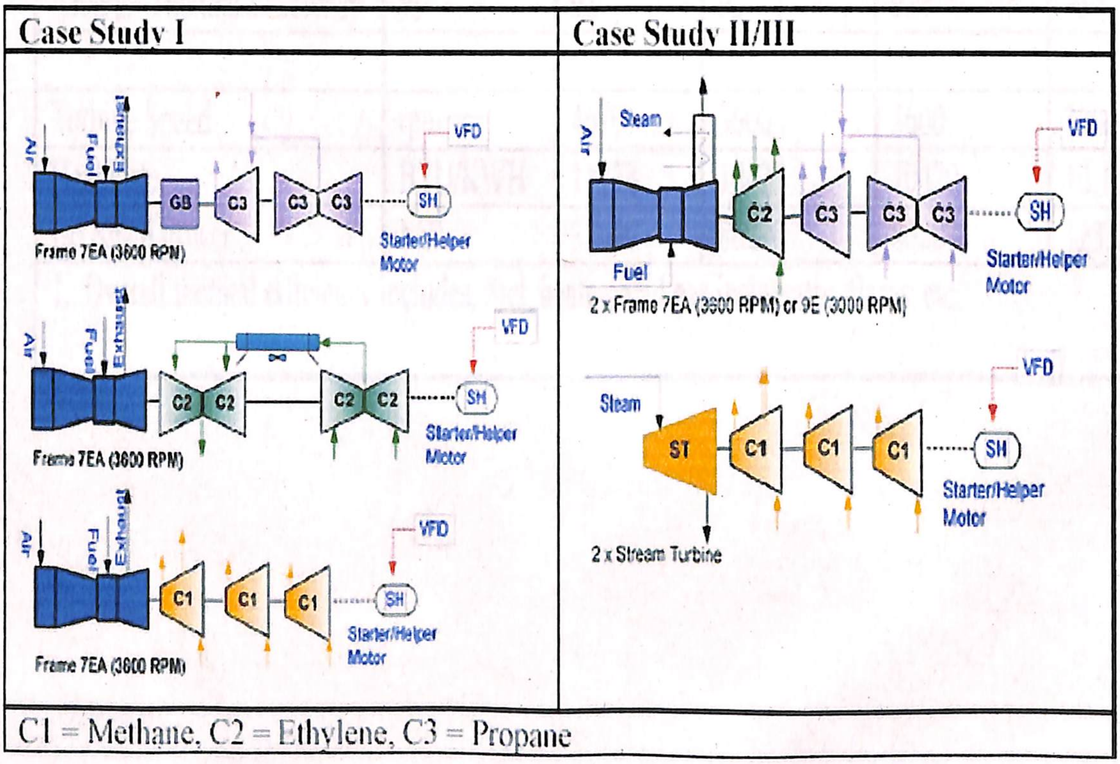


Figure19: Effect of main exchanger area and propane condenser approach Temperature on specific power for a typical LNG liquefaction process

- **Driver/Compressor Case Study Configurations**

Three of the case studies evaluated by the ConocoPhillips-Bechtel LNG PDC are discussed in this report. In Case-I the study used three Frame 7EA's for the refrigerant cycles, while in Case II we used two Frame 7EA gas turbines equipped with waste heat recovery and two parallel steam turbines. Case III was similar to Case II, but Frame 9 gas turbines were used instead of Frame 7's. These configurations are shown in Figure 3. The driver/compressor arrangement in Case I required a gearbox between the gas turbine and the propane compressor to reduce the nominal speed from 3600 rpm to about 2400. Further optimization of these configurations continues as part of the PDC ongoing efforts.



Unlike the split-shaft gas turbines, single-shaft turbines such as the Frame 7EA series turbines require starting assistance beyond what is typically provided by the vendor. For the driver/compressor configurations presented it was necessary to equip the Frame 7EA with a starting motor that would in turn be utilized as helper motor to achieve the desired LNG production rate. The motor size rating is dependent upon the refrigerant pressure within the piping and compressor system and their respective isolation valves, and the incremental power required to achieve target LNG production rate. Depending on the design premise of the respective facility, the starter/helper motors can range from 5 to 30 MW.

Table8: Driver Comparison

	Units	Frame 5D Base	Frame 7EA Case I	Frame 7EA Case II	Frame 9E Case III
LNG Production	mta	3.3	5.9	5.7	8
[1]Overall Thermal Efficiency	%	91	92	93+	93+
Turbine Speed	rpm	4670	3600	3600	3000
Heat Rate	BTU/KWH	11278	10420	10420	10,100
Iso Rated Power	KW	32600	86200	86200	123,400
1. Overall thermal efficiency includes, fuel, heater, acid gas incinerator, flares, etc.					

- **EQUIPMENT SELECTION**

Table9: Equipment Selection

Equipment Selection Items	Pros.	Cons.
Spiral wound Heat Exchanger	Flexible Operation	Proprietary/ more expensive
PFHE	Competitive vendors available. Lower pressure drop & temperature differences	Require careful design to ensure good 2-phase flow distribution in multiple exchanger configuration
Axial Compressors	High efficiency	Suitable only at high flow-rates
Large Gas Turbines	Proven, efficient and cost effective	Less reliable/strict maintenance cycle/more complicated control/fixed speed
Large Motor Drivers	Efficient, flexible and more available	Untried in LNG at speeds needed/require large power plant
Air Cooling(compared to sea-water cooling)	Lower cooling system CAPEX	Less efficient process / higher operating costs
Fluid Medium Heating(compared to steam)	Eliminates the need for steam generation and water treatment	Higher reboiler cost
Larger Train Capacity	Lower specific costs (CAPEX per tonne LNG)	Some equipment/ process may require further development

TWO-PHASE LNG EXPANDERS

LNG liquefaction plants have a complex structure with numerous systems interacting to produce the desired output. Capital investment, operation costs, environmental burden of these plants are relatively high. These high economic demands have initiated new and additional efforts to reduce the costs and the environmental impact of natural gas liquefaction plants.

The conventional liquefaction process for natural gas is to operate at high pressure through the condensation phase, after which the high pressure of the liquefied natural gas is reduced by expansion across a JT-valve. All LNG plants commissioned before 1996 are operating with an inefficient expansion valve. By replacing the existing JT-valve with a cryogenic LNG liquid turbine to expand the condensed liquefied natural gas from high pressure to low pressure the thermodynamic efficiency of the existing refrigeration process is substantially improved resulting in an increase of the total LNG output between 3 to 5%. Two-phase expanders compared to conventional valve expansion increase the total LNG output between 5 to 8%.

SPECIAL FEATURES OF THE OPTIMIZED CASCADE LNG PROCESS

Among the special features of the Optimized Cascade LNG Process are:

1. PRODUCTION FLEXIBILITY. As the composition of feed gas changes, withdrawal points for optimum production of LNG and NGL change. Multiple-stage processing lends itself to making withdrawals as necessary to minimize the cost of producing NGL products.

2. NITROGEN REMOVAL. Removal of nitrogen from the feed gas minimizes the power requirements per million Btu of product and lowers marine transportation costs. Nitrogen is removed in a unique rejection scheme. Also, fuel is provided in a manner that eliminates the need for a separate fuel gas compressor.

3. VAPOR RECOVERY. Storage tank vapor is returned to the methane refrigeration system to recover both the vapor and its refrigeration. No special equipment other than a vapor blower is required to recover the vapor and its refrigeration since the vapor is processed through existing liquefaction equipment.

4. RATE FLEXIBILITY. Liquefaction unit operation is possible over a range of near zero to one hundred percent of capacity. Recycle capability is designed into the compressors to permit operation completely independent of actual feed gas load. Heat exchangers operate smoothly at all load levels. This allows smooth operation of the total plant during start-up, shutdown and periods of reduced throughput.

5. MINIMAL SPACE REQUIREMENTS. Compact plate-fin type heat exchangers are used extensively in this process. The use of this type of equipment allows construction of a compact, easily modularized plant.

6. EASE OF OPERATION. The process utilizes pure component refrigerants of essentially constant molecular weight. This fact greatly simplifies the operation of the compression systems).

CONCLUSION

- The Liquefin process has two key elements provide advantages over existing alternatives, namely, the use of dual-mixed refrigerants and high efficiency plate-fin heat exchanger. Combining these advantages results in the liquefaction process being around 15% more efficient then current commercial processes, this can be translated into producing 15% more LNG for the same installed power generation capacity.
- A modification of conventional APCI process, known as AP-X™ process has reduced LNG capital cost. The AP-X™ LNG Process achieves not only high capacity in a single train, but also can incorporate high LPG recovery, lower LNG heating value for new markets, and maximum efficiency. This range of applications demonstrates that the AP-X™ LNG Process brings significant economies of scale to the industry, reducing the capital cost of LNG while maintaining the efficiency, flexibility, and reliability of the C3MR process.
- Optimization of Cascade LNG process reduces cost, maximize LNG production, minimize energy consumption leads to: Production Flexibility, vapor Recovery, ease of operation, rate of Flexibility, nitrogen removal, minimizes space requirement.

REFERENCES

1. Roberts, et. al. "Hybrid Cycle for the Production of Liquefied Natural Gas" US Patent 6,308,521
2. Roberts, M.J., Bronfenbrenner, J.C., Solomon, J., "Technology selection", Hydrocarbon Engineering, May, 2004.
3. Phillips Petroleum Company, Phillips Optimized Cascade Cycle for Liquefaction of Natural Gas, 1993.
4. Shell-Global Solutions Int., The Netherlands Liquefaction Plants: Development of technology & innovation, 22nd World Gas Conference, Tokyo 2003
5. ROBERTS, M.J., and AGRAWAL, R., 'Dual Mixed Refrigerant Cycle for
6. Gas Liquefaction' US Patent 6,119,479.
7. Chiu, C.-H., LPG-Recovery Processes for Baseload LNG Plants Examined. Oil & Gas Journal, 1997:p. 59-63.
8. Hudson, H.M., Wilkinson, J.D., Cuellar, K.T. and Pierce, M.C. Integrated Liquids Recovery Technology Improves LNG Production Efficiency. 82nd Annual Convention of the Gas Processors Association. 2003.
9. Swenson, L.K., Single Mixed Refrigerant Closed Loop Process for Liquefying Natural Gas, U.S. Patent 4,033,735.
10. Efficiency improvement of open-cycle cascaded refrigeration process for LNG production, US Patent No.6289692B1
11. http:// www.mpptech.com/techpp/tech_home.htm
12. http://www.lngfacts.org/
13. http://www.atlanticlng.com
14. http://www.ipt.ntnu.no/~jsg/studenter/projekt/salako2005.pdf/
15. David Wood, Natural Gas Takes Centerstage, Petromin Exploration & Production Journal, 2007: p. 18-23