LNG- NEXT GENERATION FUEL

By
Pooja V Gupta
M.Tech (Process Design Engineering)
R670209018
SAP ID 500006971

College of Engineering
University of Petroleum & Energy Studies
Dehradun
May, 2011

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A thesis submitted in partial fulfilment of the requirements for the Degree of
Master of Technology
(Process Design Engineering)

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Under the guidance of

Mr. Vasdev Singh

HOD (Chemical Engg Department)

Approved

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CERTIFICATE

This is to certify that the work contained in this thesis titled "LNG- NEXT GENERATION FUEL" has been carried out by Pooja V Gupta under my supervision and has not been submitted elsewhere for a degree.

Vasdev Singh 27.04.2011.

Mr. Vasdev Singh
HOD (Chemical engg Department)
COES, UPES.

Date 25/04/11

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 41730154

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

CERTIFICATE

L&T-CHIYODA LIMITED

Registered Office

L & T House, Ballard Estate, Mumbai 400 001

ISO 9001
ISO 14001
OHSAS 18001
REGISTERED FIRM

BS7799
IT Services

REPLY TO : Winchester B Wing, First Floor, Hiranandani Business Park, Hiranandani Garden, Powai, Mumbai 400 076

TEL : (91) (022) - 67722772
Fax : (91) (022) - 67722727

14 April 2011

To whomsoever it may concern

This is to certify that Ms. Pooja Gupta a student of M. Tech Process Design from University of Petroleum and Energy Studies- Dehradun has completed summer internship training from 01 January 2011 to 15 March 2011 in our Organization.

She has completed the project which comprises study of

- LNG - The Next Generation Fuel,
- PSV Calculations,
- Thermal rating of Heat Exchangers
- Preparation of Equipment/ Instrument Data sheets

Ms. Pooja has taken a keen interest in knowing details of the subject and worked with zeal and sincerity. She had maintained the punctuality in the completion of tasks given to him and in attendance as well.

We wish her success in her future endeavors.

for **L&T-CHIYODA LIMITED**

K. Sudhakar

K. Sudhakar,
Head - Mumbai Growth Centre

ABSTRACT

LNG - A Future Generation Fuel is being produced from natural Gas. The process includes the midstream and downstream process. The resource i.e. Natural Gas contains undesired Acid gas and mercury content, removing these and liquefying Natural gas is basic need to get the alternative fuel.

L & T Chiyoda Ltd handles such projects in its process department. The Australia based project-PNG LNG sustains on the soft ware as used for the Hydraulics, Equipments sizing, and process Simulation.

The process as described is followed for the equipment designing and further optimization using the basic software under the norms followed by L & T Chiyoda Ltd.

The data sheets developed are based on the real life data, produced after going through all procedures followed for it. Debottlenecking of a heat exchanger as described in brief gives an option over opting for investing on new equipment. Separator sizing calculations as performed results in the data sheets presented.

Design and rating of the heat exchanger is done using HTRI. Optimization and integration of condenser and rating it to the economic level. Fractionation process is simulated using HYSIS.

ACKNOWLEDGEMENT

Sometimes words fall short to show gratitude, the same happened with me during this project. The immense help and support received from L & T Chiyoda Limited overwhelmed me during the project.

First and foremost I would like to acknowledge my institute – University of Petroleum and Energy Studies and **Mr. Vasdev Singh (HOD- Chemical Department, COE, UPES)** for providing me an opportunity to work on this project and his guidance.

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I give my vote of thanks to **Mr. Deepak Shinde & Mr. Shafiulla** and to all the executives of the company without where support this report contact would never have been completed.

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Pooja Gupta

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NOMENCLATURE

V_L = required separator liquid volume , m^3

q_L = liquid throughput, m^3/d

t = Retention time , min

V_g = allowable gas velocity at the operating conditions, m/sec

ρ_l = liquid density at the operating conditions, Kg/m^3

ρ_g = gas density at the operating conditions, Kg/m^3

K = separation coefficient

D = Pipe inside diameter [m]

f = Friction factor [-]

g = Gravity (=9.80665) [m/s2]

Hg= Void fraction[-]

k = Ratio of specific heat (Cp/Cv) [-]

L = Pipe length [m]

m = Total mass flux [kg/m2s]

ml, mg = Mass flux of each phase (l: Liquid, g: Vapor) [kg/m2s]

m = ml + mg { ml = pl . U l

mg = pg. U g

Ma1, Ma2 = Mach number (1: Inlet, 2: Outlet) [-]

MW= Molecular weight [kg/kmol]

P1, P2= Pressure (1: Inlet, 2: Outlet) [Pa]

Ql, , Qg = Volume flow rate of each phase (l: Liquid, g: Vapor) [m3/h]

R = Gas Constant (=8314.51) [J/kmol.K]

Re= Reynold number [-]

Rs= Slip ratio [-]

T1, T2= Temperature (1: Inlet, 2: Outlet) [.C]

U l , U g = Velocity of each phase (l: Liquid, g: Vapor) [m/s]

z = Compressibility factor (z=1 for Ideal gas) [-]

Z1, Z2 = Elevation (1: Inlet, 2: Outlet) [m]

ΔP = Total difference of pressure in piping [Pa]

ΔPf = Frictional loss in piping [Pa]

ΔPh = Head loss in piping [Pa]

ΔPm = Pressure loss in piping due to increase of momentum [Pa]

ε = Surface roughness of pipe inside [m]

μ l , μ g = Absolute viscosity of each phase [kg/m.s]

θ = Angle of piping from horizontal [0]

pl , pg = Density of each phase [kg/m3]

Greek Alphabet

Alpha α	Iota ι	Rho ρ	Beta β
Kappa κ	Sigma σ	Gamma γ	Lambda λ
Tau τ	Delta Δ	Mu μ	Upsilon υ
Epsilon ε	Nu ν	Phi Φ, φ	Zeta ζ
Xi ξ	Kai χ	Eta η	Omicron O
Psi ψ	Theta θ	Pi π	Omega Ω, ω

CHAPTER 1 - INTRODUCTION

1.1 Company Profile

L & T Chiyoda Limited (LTC) is an engineering consultancy organization formed by larsen & Toubro Limited, India's premier engineering , manufacturing and construction company (holding 50 % equity) and Chiyoda corporation, Japan , world renowned Engineering company five decades of experience in Hydrocarbon and related fields (holding 50 % equity).

Incorporated on 19th November 1994, LTC commenced operations in February 1995 and is catering to national and international clients, both directly and through its parent companies, LTC offers international grade engineering and project management services eith integrated engineering concepts, supported by state of the art computer hardware and software facilities operating in a networking environment.

LTC, the youngest organization of its kind to get the ISO 9001 accreditation certification, has established an independent identity amongst major clients and process know how suppliers globally, through its indigenous and export engineering credentials. It has already upgraded its ISO certification to ISO 9001:2000 and also achieved certifications for ISO 14001:2004, ISO 27001:2005, OHSAS 18001:2007 and CMMI maturity level 5.

Working towards positive engineering through plant modeling in electronic media, LTC offers a creative response to clients' needs. The actual plant itself is 'visualized' to a very close reality in the engineering office during the detailed engineering stage, resulting in high efficiency and accuracy in engineering and ease of construction.

LTC has specialized in the engineering for fast track EPC jobs of multiple complexities; repeatedly proving it's adaptability from time to time.

The major industries in which LTC adds significant dimension includes Petroleum refining Petrochemicals, chemicals, Fertilizers, Oil and Gas and LNG & LPG.

LTC executed projects for the numerous clients either through its parent companies or directly. Some of the major clients are:

Indian Clients:

1. Oil India Limited(OIL),Assam
2. Gas Authority of India Limited, Baroda
3. Indian Petrochemical Corporation Limited, Baroda
4. Cairn Energy India Pty. Ltd., Chennai
5. Chennai Petroleum Corporation Limited, Chennai
6. Indian Oil Corporation Ltd. Delhi for petroleum refining projects in their various refineries like, Panipat Refinery, Digboi Refinery, Barauni Refinery, Haldia Refinery, Gujarat Refinery, Guwahati Refinery etc.
7. Kochi Refineries Limited, Kochi
8. Shriram Fertilizer & Chemicals Ltd., Kota
9. Bharat petroleum Coproration Limited, Mumbai
10. Hindustan Petroleum Corporation Limited(HPCL),Mumbai
11. Oil & Natural Gas Corporation Ltd., Mumbai
12. ONGC-Mangalore Petrochemicals Limited (OMPL), Mangalore
13. Indian Oil Corporation Limited including Paradip Refinery, Mathura Refinery, Haldia Refinery
14. HPCL-Mittal Energy Investments Limited (HMEL), Bhatinda

Overseas Clients:

1. Abu Dhabi Gas Industries Co.(GASCO), Abu Dhabi
2. Abu Dhabi Gas Liquefaction Co.(ADGAS), Abu Dhabi
3. Abu Dhabi Oil Refining Company(Takreer), Abu Dhabi
4. Ruwais Fertilizer Industries, Abu Dhabi

LNG accounts for about 4% of natural gas consumption worldwide, and is produced in dozens of large-scale liquefaction plants. It is produced by cooling natural gas to a temperature of minus 260 degrees F (minus 160 Celsius). At this temperature, natural gas becomes liquid and its volume reduces 615 times. LNG occupies 1/600th the volume of natural gas at atmospheric temperature and pressure. The gas have high energy density, which makes it useful for energy storage in double-walled vacuum-insulated tanks.

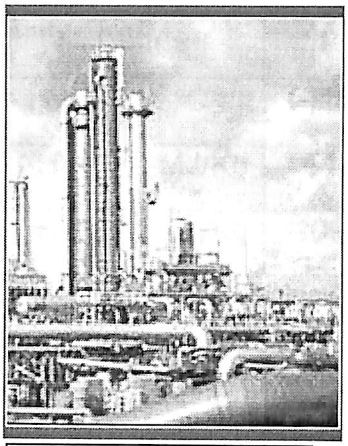


Fig 1.1 plant view

The production process of LNG starts with, Natural Gas, being transported to the LNG Plant site as feedstock. After filtration and metering in the feedstock reception facility, the feedstock gas enters the LNG plant and is distributed among the identical liquefaction systems. Each LNG process plant consists of reception, acid gas removal, dehydration/mercaptan removal, mercury removal, gas chilling and liquefaction, refrigeration, fractionation, nitrogen rejection and sulfur recovery units.

Composition mol%	Minimum	Maximum	Average	Specification
Nitrogen	0.14	0.62	0.28	Max. 1.0
Methane	89.95	92.42	91.32	Min. 85.0
Ethane	3.76	5.01	4.2	
Propane	2.45	3.47	2.86	
i-Butane	0.56	0.86	0.69	Max. 2.0
n-Butane	0.53	0.8	0.63	
i-Pentane	0	0.04	0.02	Max. 0.1

n-Pentane	0	0.01	0	
CO2 , ppmv	-	100	-	
H2S , mg/NM3	-	5.0	-	
Total S , mg/NM3	-	30.0	-	
Molecular weight	17.76	18.41	18.03	
Density , Kg/M3	452.9	463.5	457.9	
Temperature , °C	(-)161.3	(-)159.6	(-)160.3	
CV (HHV) , MJ/Kg	54.17	54.9	54.44	
CV (HHV) , MJ/Kg	1098.0	1134.0	1112.0	1050 – 1150

Table 1.1 LNG compositions and properties

Whenever the source of natural gas production is a long distance from the location of potential usage and a pipeline is not a viable solution, liquefaction of the natural gas may be an economical choice. The liquefaction of natural gas reduces its volume about 600 fold and allows the gas to be exported to distant ports as a liquid in LNG tankers.

New LNG (Liquefied Natural Gas) production plants are constantly being built to satisfy the growing global demand for natural gas. Likewise, in order to reduce the unit production cost, liquefaction line capacity has been increasing year by year and is currently topped by Qatar's mega LNG lines, each producing about 8 Mtons/y of LNG.

Technology innovation and economies of scale have been the two key contributors to the industry's progress. GE Oil & Gas has a long history of leadership in the evolution of LNG technology. Our sustained commitment to innovative design and world-class engineering, and our production and testing capabilities have allowed us to push the envelope of highly reliable, advanced LNG solutions.

1.3 Brief about technology

The natural gas from the field is first treated in a gas processing unit to remove higher molecular weight hydrocarbons, sulfur compounds and water. It is then fed to the liquefaction process where it is, depending on the process used, cooled in two or three cascade cooling cycles down to the liquefaction temperature of -160°C (- 256 °F). The cold liquid LNG is then transferred to heavily insulated storage tanks at atmospheric pressure, and from there it is loaded into LNG tankers for shipment.

Each of the cooling cycles requires a very large compression train which is typically driven by a gas turbine. The 8 Mtons/y plants in Qatar are based on three trains, each driven by a 125MW Fr9E gas turbine.

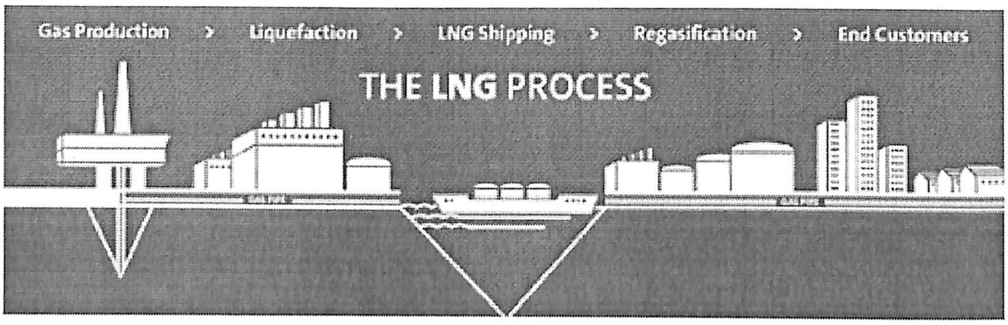


Fig 1.2 The value chain for LNG process

There is a four-step 'process' to get natural gas into the UK Natural Gas transmission system and on to homes and businesses in the UK from gas fields in remote locations.

There are several steps involved in LNG life cycle including:

- Exploration to find natural gas in the earth's crust and production of the gas for delivery to gas users. Most of the time natural gas is discovered during the search for crude oil.

- Liquefaction to convert natural gas into a liquid state so that it can be transported in ships.
- Shipping the LNG in special purpose vessels.
- Storage and Regasification, to convert the LNG stored in specially made storage tanks, from the liquefied phase to the gaseous phase, ready to be moved to the final destination through the natural gas pipeline system.

1.4 Features of LNG Plant

There exist a vast number of natural gas liquefaction plants designs, but, all are based on the combination of heat exchange and refrigeration. The gas being liquefied, however, takes the same liquefaction path. The dry, clean gas enters a heat exchanger and exits as LNG. The capital invested in a plant and the operating cost of any liquefaction plant is based on the refrigeration.

Though, Liquefied Natural Gas can also be extracted from cryogenic hydrocarbon extraction and petro-chemical processes, but it requires careful consideration at these facilities to assure the process gas is liquefiable.

1.5 LNG - Scenario

LNG trade first developed in the 1960s to move gas from producer to consumer, where pipelines were not economic because of distance or the need to cross deep water. It has grown rapidly over the last 30 years, reaching nearly 64 m in 1994. Yet, it still accounts for 4% of global gas consumption and less than 1% of total primary energy use.

The prospects for LNG in the future are encouraging. The efficiency, convenience and environmental advantages of natural gas make it the fuel of choice. As lower cost gas reserves close to market become fully developed, consumers have to look to more distant reserves to satisfy demand growth. This means that LNG will potentially play an increasing role in future natural gas supply. However, LNG projects involve large capital investment which must be economically attractive for investors: a major challenge.

1.6 History of LNG

The technology for the liquefaction of natural gas was used first in the United state over 50 years ago, for the storage of gas to meet daily and seasonal fluctuations in demand. The first international trade in LNG took place in 1959 when a cargo of LNG was transported from the United state to the UK via "Methane Pioneer" a specially constructed LNG vessel.

This proved that transportation of LNG over long distances was feasible. Algeria was the first commercial LNG exporting country. The first shipments were made to British Gas in UK, commencing in 1964. Japan received its first LNG from the Kenai LNG plant, in the south of Alaska, in 1969. Since then, Japan has been the dominant LNG market, and today imports some 65% of the total world LNG production. The United States received first LNG in the late 1970s from Algeria. The first Middle East LNG supply was from Abu Dhabi to the pacific region in 1977 and also more recently from Qatar.

1.7 LNG Scenario in India

India has launched a major initiative centering on the use of liquefied natural gas as fuel for future power generation projects. According to the official estimates of future electric power demand, India needs to install nearly 57000 MW of power capacity. Presently, there are no power plants in India that use LNG as a source of fuel, although natural gas is widely used in the Petrochemical , Fertilizer and Power sectors.

Indian officials have approved the formation of a joint venture (JV) company, "Petronet LNG " to import and create infrastructure for LNG projects throughout in India. The joint venture (JV) company consists of Indian Oil Corporation Ltd. (IOCL), Oil & Natural Gas Corporation (ONGC), Bharat Petroleum Corporation Ltd. (BPCL) , and Gas Authority of India Ltd. (GAIL) With authorized capital of about $480 million , the JV will seek to advance LNG projects and pursue a partner with upstream project interests. The JV will have an equity position of 50% (12.5% each by IOCL, ONGC, BPCL & GAIL) and the balance of the equity will be offered to private parties, either Indian or Foreign.

Gas Authority of India Ltd. (GAIL), acting on behalf of Petronet LNG, a consortium of four public sector enterprises including GAIL, expects to construct a number of LNG receiving terminals along the western coast of India, starting with one at Dahej in Gujarat state and one at Cochin in the state of Kerala. The Dahej terminal will have a capacity of 5.0 million metric tons per year, while the size of the Cochin terminal will be 2.5 million metric tons per year. The company is now in the process of short listing LNG suppliers, with whom it will sign the long term purchase contract. Gaz de France, one of the world's largest importers of LNG will assist Petronet LNG in this endeavor. The LNG suppliers being considered are, Total of France , Ras Laffan LNG Co. of Qatar, Woodside Petroleum Ltd. of Australia, Pertamina of Indoanesia, Petronas of Malaysia, Chevron Asiatic Ltd. of Australia and Shell International Petroleum Co. Ltd. of U.K.

The IOCL/ONGC/BPCL/GAIL group also plans to build two LNG terminals with a capacity of 2.5 million metric tons per year each. One will be built at Ennore in Tamilnadu state and the other at Manglore in Karnataka state. The Petronet LNG is also planning with Amoco Corporation, a 5.0 million metric tons per year LNG terminal at Hazira in Gujarat state. The complex will handle importing, unloading, regasification and supply.

British Gas, BG plc has a joint venture with India's Gujarat Pipavav Port Limited to build an LNG import terminal at the port of Pipavav in the state of Gujarat at an estimated cost of $400 million. BG plc has signed a memorandum of understanding (MOU) with Yemen LNG Co. , that sets a framework for further talks on delivery of LNG from Yemen to India. The dedicated LNG terminal will have a capacity of 2.5 million metric tons per year and is expected to be expanded later to 5.0 million metric ton per year. The import facilities will consists of berthing and offloading facilities for the tankers, storage tanks and regasification facilities. First delivery of LNG is envisaged between mid 2002 and mid 2003, continuing for approximately 25 years. BG chose Pipavav after carrying out a detailed, 18 month feasibility study of the port. The study also focused on potential power projects in the state of Gujarat and northern India.

Hindustan Petroleum Corporation Ltd. (HPCL) has a joint venture with the Total to construct LNG terminal with an initial capacity of 2.5 million metric tons per year in the eastern part of Andhra Pradesh for a capital cost of $500 million. The target is a capacity of 6 - 7 million metric tons per year.

As on today, the total numbers of proposed LNG terminals in India can be summarized as under.

PROPOSED LNG TERMINALS IN INDIA		
Terminal Location	Capacity	Owner
Dahej , Gujarat	5.0 mmt / Year	Petronet LNG
Cochin , Kerala	2.5 mmt / year	Petronet LNG
Ennore , Tamilnadu	2.5 mmt / year	Petronet LNG
Manglore , Karnataka	2.5 mmt / year	Petronet LNG
Hazira , Gujarat	5.0 mmt / year	Petronet LNG & Amoco
Hazira , Gujarat	5.0 mmt / year	Reliance Industries Ltd.
Jamnagar , Gujarat	5.0 mmt / year	Reliance Industries Ltd.
Hazira , Gujarat	2.77 mmt / year	Royal Dutch / Shell & Essar Group of India
Pipavav , Gujarat	2.5 - 5.0 mmt / year	BG plc & GPPL
Andhrapradesh	2.5 mmt / year	HPCL & Total

Table 1.2 Proposed LNG terminals in India

1.8 LNG – Import and Export

Over a period of past 30 years, a dramatic turnaround has been witnessed in the growth of liquefied natural gas (LNG). This has also tracked equally strong natural gas demand growth. Much of this growth has been for power generation in Asia.

Today, there are nine LNG exporting countries and nine importing countries. The details of the LNG importing and exporting countries, up to 1996 are tabulated here below.

LNG IMPORTING COUNTRIES			
Country	Million Metric Tons	Billion cu m of Gas	% of total
Japan	44.237	56.034	60.93
Korea	9.470	11.995	13.05
Spain	5.568	7.053	7.67
France	5.104	6.466	7.03
France / Belgium *	0.165	0.209	0.23
Belgium	2.862	3.625	3.94
Taiwan	2.575	3.262	3.55
Turkey	1.767	2.238	2.43
U.S.	0.850	1.077	1.17
Total	**72.598**	**91.959**	**100**

* Joint purchase

Table 1.3 LNG importing Countries

LNG EXPORTING COUNTRIES			
Country	Million Metric Tons	Billion cu m of Gas	% of total
Indonesia	25.372	32.139	34.95
Algeria	14.412	18.256	19.85
Malaysia	12.112	15.342	16.68
Australia	7.151	9.058	9.85
Brunei	6.038	7.649	8.32
Abu Dhabi	5.315	6.732	7.32
U.S.	1.312	1.662	1.81
Libya	0.885	1.121	1.22
Total			
Qatar also recently entered in the world of LNG exporting countries.			

Table 1.4 LNG exporting countries

1.9 LNG need

Natural gas as fuel for power generation grows at a faster pace than other fuels for electricity generation. The growth rate is 5.7% per year compared with 3.5% per year for all fuels as represented here below.

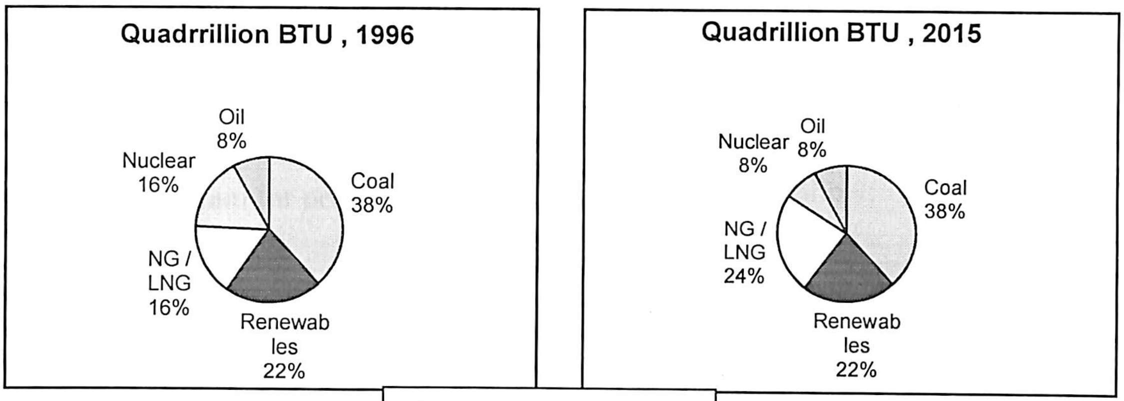


Fig 1.3 Need of LNG

1.10 PNG – LNG in LTC

Client- Esso Highlands Ltd

Site – Papua New Guinea, Port Moresby (25 km north west from port Moresby)

Project name – EPC3 for PNG LNG project

Guarantee Items – LNG production, Auto consumption, Emissions and LNG/ Condensate loading rate.

Name of Joint Venture – Chiyoda – JGC joint venture (CJJV)

Capacity – 6.9 MTA (1000 Mscfd Feed gas Guaranteed LNG production Base)

Work done under this project by the L & T Employees includes the hydraulics, which covers the single phase, two phase calculation and pump detailing.

This project report concludes with a comparative result for the hydraulics done by soft ware and hand calculation. It also includes the rating of a condenser, separator sizing and simulation of part of a single train.

CHAPTER 2 – PROCESS AND TECHNOLOGY

The Block diagram for processing Natural gas to LNG is ass follow:

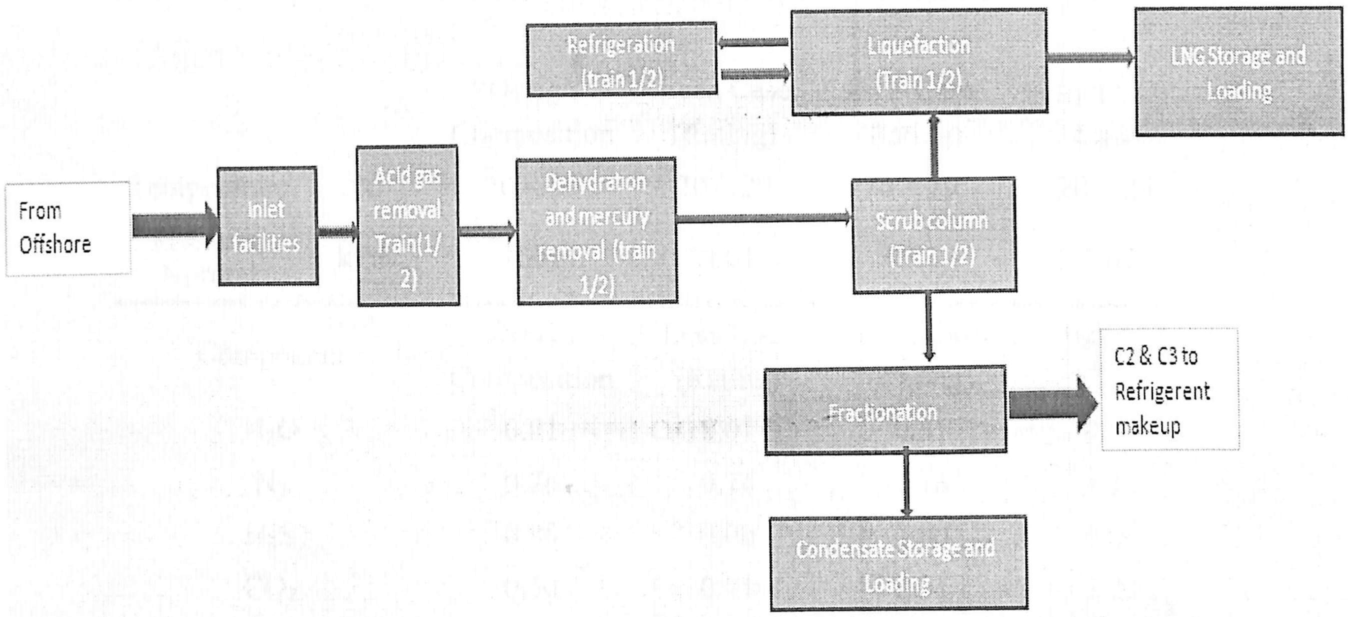


Fig 2.1 Block Diagram – LNG Processing

The process for processing obtained natural gas to LNG consist of three sections

- Inlet system
- Hot section
- Cold section

2.1 Inlet system

Feed gas from the offshore gas pipeline is introduced to the inlet receiving and treatment system of the LNG plant via this Gas Pipeline Inlet System. The purpose of this system is to connect offshore gas line to the inlet receiving and treatment system. The Pig receiver is designed to carry out pigging operation in the upstream offshore feed gas pipeline.

Feed gas from the upstream pipeline is introduced to the LNG plant via this Inlet Receiving and Treatment System. The Inlet Receiving and Treatment System is intended to stabilize feed gas supply pressure to the LNG plant, to separate entrained liquid from the feed gas stream and provide allocation flow measurement. The capacity of the Inlet Receiving and treatment System is based on the peak feed rate of 1226 kSm3/hr (1041 Mscfd).

		Design Composition	Lean Case (Rating)	Rich Case (Rating)	High CO_2 Case
Temperature	°C	20 ~ 29	20 ~ 29	20 ~ 29	20 ~ 29
Pressure, Normal	kPaa	7601	7601	7601	7601
Component		Design Composition	Lean Case (Rating)	Rich Case (Rating)	High CO_2 Case
H_2O		0.01	0.01	0.01	0.00
N_2		0.76	0.74	1.16	1.10
H_2S		0.00	0.00	0.00	0.00
CO_2		0.50	0.71	2.04	3.09
C_1		87.94	89.98	83.37	83.81
C_2		6.83	5.71	8.01	7.21
C_3		2.44	1.72	3.70	3.30
iC_4		0.48	0.36	0.59	0.51
nC_4		0.58	0.40	0.77	0.69
iC_5		0.18	0.16	0.16	0.12
nC_5		0.12	0.09	0.10	0.08
C_6		0.10	0.07	0.05	0.05
C_7		0.04	0.04	0.03	0.03
C_8		0.01	0.01	0.01	0.01
C_9		0.00	0.00	0.00	0.00
MOL%		100.00	100.00	100.00	100.00
Molecular weight		18.64	18.16	19.70	19.70

Fig

Table 2.1 Feed compositions to the inlet receiving system

The LNG project gas pipeline delivers the high- pressure gas at 7601 kPaa, between 20°C and 29°C to the inlet facilities in the LNG plant. The pipeline will deliver single phase conditioned at the upstream facilities to a water content of < 7 lbs/Mscf and a maximum hydrocarbon dew point of 5°C.

2.2 Hot section

The hot section includes following processes:

2.21 Acid gas removal

Feed gas from the upstream Inlet Receiving and treatment System is sent to the AGR to remove Acid Gas (CO_2) from the feed gas to avoid freezing out and Blockage in the downstream liquefaction unit. The plant shall consist of two separate AGR trains after Inlet Receiving and Treatment System. The capacity of each AGR train is based on 50% of the peak feed gas flow rate and high CO_2 (3.09 mol % CO_2) gas composition.

Feed gas from the inlet area is sent to AGR for removal of the CO_2 in the gas stream. The AGR design is done by UOP which uses Ucarsol AP – 814 as solvent for absorption. The AGR equipment is sized based on a Ucarsol AP – 814 circulation rate corresponding to high CO_2 case inlet gas composition at 50% of peak flow rate of feed gas 1227.1 kSm^3/hr for each train. The specification of the treated gas from the AGR is 30 ppm(v) CO_2 and 4 mg/m3 H_2S. The feed gas is sent through AGR filter separator to remove the entrained liquids and solids from natural gas feed stream. The normal pipeline feed gas temperature is too low for effective CO_2 absorption kinetics, so the feed gas is preheated in the feed gas preheater with hot oil to a temperature of between 35°C and 40°C by heat exchange.

The warmed feed gas, at approximately 37 °C, is contacted with Ucarsol AP – 814 solvent in the amine absorber. The solvent absorbs CO_2 (and H_2S,if any) in the feed gas and reduces

CO_2 concentration to less than 30 ppm(v). this sweet gas is further washed with water at the top section of amine absorber to minimize solvent carry over with the gas. The wash water is in circulation and makeup water is added to the wash water stream to compensate the loss of water from the amine system.

The CO_2 rich Ucarsol AP – 814 solvent is sent to amine regenerator to recover lean amine solvent. To minimize the dissolved gas hydrocarbon carryover to the amine regenerator, rich amine solvent is flashed in rich amine flash drum operating at about 750 kPaa. Flash gas from rich amine flash drum is sent to the LP fuel gas header. The rich amine solvent is preheated by regenerator bottom lean amine solvent in lean rich amine exchanger before it is sent to amine regenerator.

Amine regenerator strips off CO_2 and (H_2S if any) from the amine solvent, lean amine solvent at amine regenerator bottom is sent back to amine absorber after heat recovery in lean rich amine exchanger by lean amine boosted pump followed by the high pressure multistage lean amine pump.

2.22 Dehydration and Mercury removal system

Treated gas from AGR system is introduced to Dehydration and Mercury removal system. Dehydration and mercury removal system dries the water – saturated treated gas down to less than 0.1 ppm(v) of water and removes any mercury present to less than 10 ng/Nm3.

The dehydration system is designed to remove water from the feed gas prior to enter the liquefaction unit. The dehydration system uses molecular sieves, crystalline structure which traps the water, as the adsorbent to accomplish this task. The dehydration system is located downstream of the acid gas removal system. The dehydration system will dry the water –

saturated treated gas down to less than 0.1 ppm(v) o water to avoid freezing and hydrate issues in the liquefaction unit.

The treated gas from the AGR unit is delivered to dehydration pre-cooler and cooled with propane refrigerant to 25 °C, approximately 5 °C above the hydrate formation temperature. The cooled gas flows to dehydration feed separator where the condensed free water is recovered along with any solvent carryover. This recovered water stream is returned to the AGR system to minimize the make-up water requirement.

The remaining water in the feed gas is adsorbed from the gas in the molecular sieve driers, three- bed (plus 1 spare bed) molecular sieve system. The regenerator gas for the molecular sieve beds are heated to approximately 288 °C in a fired heater.

The beds are regenerated using BOG from BOG compressor. If this flow rate is npt enough for regeneration, the dehydrated feed gas from the molecular sieve drier outlet is used as supplement. Spent regeneration gas downstream of the mole sieves is used as a source of HP fuel gas for the plant.

The dry feed gas enters the mercury adsorber to remove any mercury present to less than 10 ng/Nm3. The catalyst is sulphur- impregnated type activated carbon. The treated gas will then flow through mercury adsorber after filters, which prevents catalyst dust entering liquefaction system.

2.23 Amine storage system

The purpose of amine storage system is to hold inventory of the amine solvent for AGR system. Freshly prepared amine solvent in amine sump drum is transferred to the amine storage tank equal to the volume required to fill up one AGR trains. Similarly, in case of draining of total amine solvent holdup from AGR system during shutdown of the unit, all the amine solvent is transferred to amine storage tank via amine sump drum and pump.

The amine solvent from amine sump drum is sent to amine storage tank through amine make-up filter. Amine storage tank pump transfers the amine solvent back to the bottom of amine regenerator of AGR system through the amine make-up filter. Amine storage tank serves as a source as well as sink of amine solvent inventory for acid gas removal system.

Operating pressure of the amine storage tank is kept at pressure, slightly above the atmospheric pressure with the help of the pressure regulator on nitrogen blanket gas supply line.

2.3 Cold section

The cold section consist of the following

2.31 Fractionation

The purpose of Fractionation System is to produce Ethane, Propane, Butane, Pentane and Condensate products from the natural gas liquids (NGLs). The Ethane and Propane produced by this system are utilized as a make-up for the refrigerant of PNG LNG Plant. The systems which are supplied these refrigerant are Propane and Mixed Refrigerant Systems (System 666). Produced Ethane, Propane, and Butane are combined and injected to the Natural Gas stream to MCHE for heating value control. The remained C2, C3 & C4 are utilized as a feedstock to HP Fuel Gas System (System 966). The produced Pentane is utilized only for fuel gas. The bottom product: Condensate is sent to Condensate Storage System (System 634) for shipping out.

De-ethanizer system

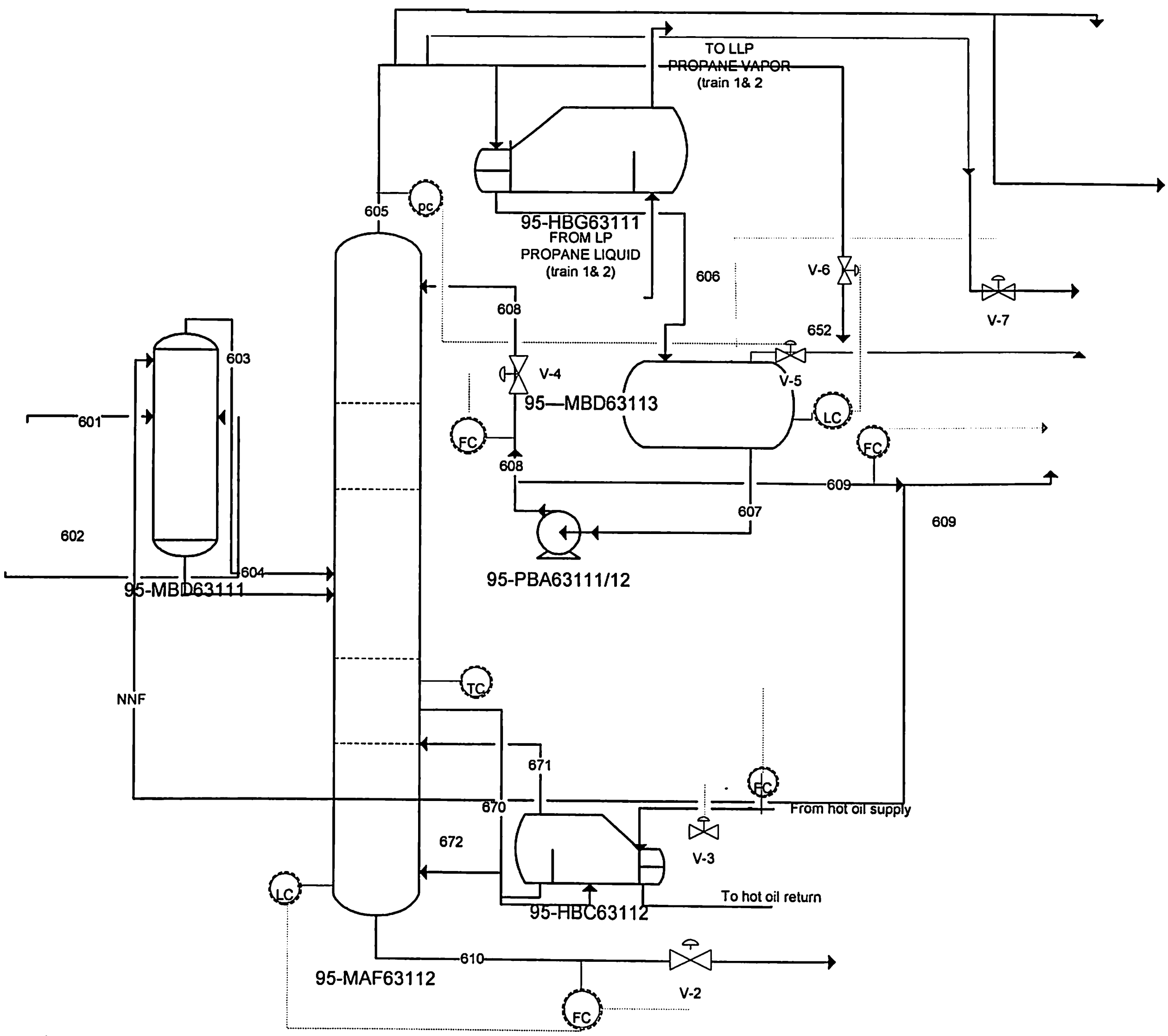


Fig 2.2 Process flow diagram for De-ethanizer system

The hydrocarbon liquids from the bottom of scrub columns (91/92-MAF 69511) are fed to de-ethanizer (95-MAF63112). Ethane rich vapor is recovered at the top of the column.

Propane and heavier components are drawn of from the bottom of the column and sent to de-propanizer (95-MAF63121).

The vapor from the de-ethanizer overhead is partially condensed through de-ethanizer condenser (95-HBG63111) chilled by LP propane. It is then fed to de-ethanizer reflux drum (95-MBD63113). The vapor from the reflux drum is sent to the LPG reinjection facility. The condensed liquid from the reflux drum is pumped back to the top of de-ethanizer as a reflux by de-ethanizer reflux pump (95-PBA63111/12).

When required, a portion of the reflux stream can be diverted to ethane refrigerant storage drum (95-MBJ69810) for future servicing as ethane refrigerant make-up. Lines are also provided to send a part of the de-ethanizer overhead stream as make-up for MR to LP/MR suction drums in the both the trains (91/92-MBD66610/20/30).

De-propanizer system

The hydrocarbon liquid from the bottom of de-ethanizer (95-MAF63112) is fed to the de-propanizer (95-MAF63121). The vapor from the de-propanizer overhead is totally condensed through air cooled de-propanizer condenser (95-HFF63121). Then the liquid is sent to de-propanizer reflux drum (95-MBD63122). A portion of the condensed liquids from the reflux drum is pumped back to the top of the de-propanizer as a reflux by de- propanizer reflux pump (95-PBA63121.22/23). The remaining portion is sent to the LPG reinjection facility by propane reinjection pump (95-PBA63124/25). When required, a portion of the liquid propane stream can be sent to propane refrigerant storage sphere (95-MBJ69820) for future serving as make-up for the propane refrigerant and/or MR.

For vapor make-up to propane refrigerant and MR circuits, lines are provided to route part of thr de- propanizer overhead to the suction line propane compressor LLP and LP. MR suction drums (91/92-MBD66610/20/30) of the both trains. The liquid from the bottom of the de-prpanizer is sent to de- botanizer (95-MAF62131).

De-butanizer system

The hydrocarbon liquids from the bottom of the de-propanizer (95-MAF63121) are fed to the de-butanizer (95-MAF63131).

The de-butanizer overhead vapor is totally condensed through air cooled de-butanizer condenser (95-HFF63131), and then sent to de-butanizer reflux drum (95-MBD63132). A portion of the condensed liquid from the reflux drum is pumped back as reflux to the top of the column by debutanizer reflux pump (95-PBA63131/32).

The remaining portion is sent to the LPG reinjection facility by the butane reinjection pump (95-PBA63133/34). When required, a portion of the butane product stream is directly sent to HP fuel gas system to adjust the HHV of LNG products. The liquid from the bottom of the de- botanizer is sent to the de- pentanizer (95-MAF63141).

De-pentanizer system

The hydrocarbon liquid from the bottom of de-butanizer (95-MAF63131) is fed to de-pentanizer (95-MAF63141). The de-pentanizer column overhead is totally condensed in de-pentanizer overhead condenser (95-HFF63141) and collected in de-pentanizer reflux drum (95-MBD63142). A portion of the condensed liquid from the reflux drum is pumped back as reflux to the top of the column by de-pentanizer reflux pump (95-PBA63141/42). The remaining liquid is sent to fuel gas system (system 966) by pentane reinjection pump (95-PBA63143/44). Objectives of the de-pentanizer are to fully stabilize plant condensate so that it meets the condensate specification and stored in the condensate storage atmospheric tanks in condensate storage and loading facility (system 634). The production rate of the condensate (C5+) is determined based on the vapor pressure specification as 68.95 kPaa (10 Psia).

2.32 Condensate Storage and Loading

Condensate storage tank- condensate produced from fractionation system directly to the tanks via single rundown line. Condensate can be routed to any of the two tanks. Condensate produced from fractionation system is stored in condensate storage tank at atmospheric pressure, prior to being loaded into the condensate carrier. Both tanks are single containment floating roof type with 10,500 m^3 working capacity each.

Condensate loading system- condensate loading system is designed for loading rate of 1500 m^3/hr. one loading pump is operated for condensate loading operation. The condensate is transferred to the carrier via the DN 450 loading line.

At the loading berth, condensate loading arm (FAY63420) is provided to accommodate the design loading rate. On the basis of one loading arm operating, the condensate is delivered at a pressure of 400 kPaa at the ship many fold.

An emergency shutdown system is installed to minimize the risk of product spills and mechanical damage to the loading arms.

2.33 Liquefaction system (system no. 695)

Feed gas distillation: dry feed gas from the mercury removal system (system no. 699) is cooled in series propane refrigerant coolers and enters a scrub column where the lighter components go overhead and the heavier components go to the bottom. Overhead gas is fed to main cryogenic heat exchanger tube side. The bottom liquid is sent to the fractionation system.

The feed gas from scrub column overhead is condensed and sub-cooled in the MCHE warm, middle and cold bundles. The Liquefied Natural gas (LNG) leaving the MCHE is fed to LNG hydraulic turbine or its bypass J-T valve. The expanded LNG is sent to the LNG Storage tanks via rundown pipeline.

2.34 LNG storage and loading system

The purpose of LNG storage and loading system is to store the product LNG from 2 LNG processing Trains prior to being loaded into LNG tankers for export. Both tanks are single containment type with 160,000 m^3 capacity each. BOG generated from the tanks is recovered by BOG compression system to be compressed and sent to the fuel gas system.

LNG is loaded from LNG/ condensate combined loading berth, where dedicated loading facilities are provided for each product. The design LNG loading rate is 12,000 m^3/hr. The loading system will be designed to accommodate LNG ship sizes from 125,000 m^3 to 220,000 m^3.emergency shut down system will be installed to minimize the risk of product spill. BOG generated by the loading operation will be sent back from the LNG ship to BOG compression system.

A rundown header is used to transfer the LNG product from two trains to the LNG storage tanks. In order to maintain the temperature of LNG in the loading line below the bubble point, LNG from one tank is circulated using LNG circulation pump. LNG is pumped to the loading line up to the jetty head, then back to the LNG storage tanks via circulation line.

The pressure at the upstream of the FCV located at the end of the circulation line is maintained at a pressure above LNG bubble point.

BOG generated from the LNG due to heat in-leaks from the natural convection/conduction on surface area of liquid-filled piping and equipments, as well as heat transfer from the pumps during operation, flashed vapor from rundown LNG and displacement.

BOG generated is accumulated and sent to the BOG compression system to maintain stable tank pressure. Compressed BOG is sent to HP Fuel gas system after passing through BOG compressor suction drum for quenching at a desired cryogenic temperature and cooled down by an after cooler at a condition required by the HP fuel gas system.

2.35 Refrigeration system (system no. 666)

The system consists of following facilities:

1. The mixed refrigerant (MR) system
2. The propane refrigerant system

The propane refrigeration system is used to satisfy the first level refrigeration needs of the process. The propane refrigerant system utilizes propane evaporating at four pressure levels, to supply refrigeration to the feed gas circuit in the liquefaction system, the MR circuit chillers in the refrigerant system ant the de-ethanizer condenser and LPG re-injection stream from fractionation system. These levels are designated HP, MP, LP and LLP for high pressure, medium pressure, low pressure and low low pressure respectively.

For start-up inventory, propane from propane refrigerant storage is sent into the 91-MBA66661 downstream by propane transfer pump (95-PBA69820). During normal operation propane vapor can be sent from de-propanizer column (95-MAF63121) over head in the fractionation system as make-up to LLP propane compressor suction.

Propane liquid from each of the exchangers can flow into the propane transfer drum, 91-MBA66663. The propane transfer pump, 91-PBA66660 is used foe sending the propane liquid to propane refrigerant storage system in the maintenance period.

The refrigeration to liquefy the feed gas is supplied by a mixture of nitrogen, methane, ethane and propane, know as mixed refrigerant (MR). this mixture is adjusted to provide the optimum cooling and liquefaction in the MCHE.

CHAPTER 3 – CALCULATIONS

3.1 Hydraulics

Hydraulics consist of line sizing for single phase (gas or liquid) and two phase fluids along with the head calculation for the rotating equipment such as pump, compressor, turbine, blower.

Hydraulics divides the plant into various loops out of which some are the loops under process system and some are the loops under utility system. Calculation for each loop requires the details as follows

- Process data
    - PFD, H & MB, Data sheets
    - Start and end point pressure
    - Fluid flow rate and physical properties
    - Equipment and instrument pressure drop
    - Liquid level
- Geometric data
    - Datasheets
        - ✓ Equipment dimensions
    - Plot plans
        - ✓ Isometric design
        - ✓ Critical equipment elevation
        - ✓ Straight pipe length
- Piping information
    - P & ID

- ✓ Pipe size and material class

- PMS
  - ✓ Pipe schedule

Calculation method used gives the result which are diagnosed as follow

Hydraulic calculation

- Single phase lines (liquid or vapor)
  - Velocity m/s
  - Friction loss/100m
- Two phase lines
  - Flow pattern
  - ρv^2
- Rotating equipment (pump)
  - Minimum pressure drop across governing control valve
  - Net Positive Suction Head Available (NPSHA)

Hydraulics calculation can be done using soft ware such as Pro DRAW and also can be done manually by following procedure.

Calculation Procedure (manual calculation)

Single Phase Pressure Drop Calculation (Applicable for Liquid Phase and Incompressible Gas Phase calculations.)

The equivalent length can be calculated using following figure.

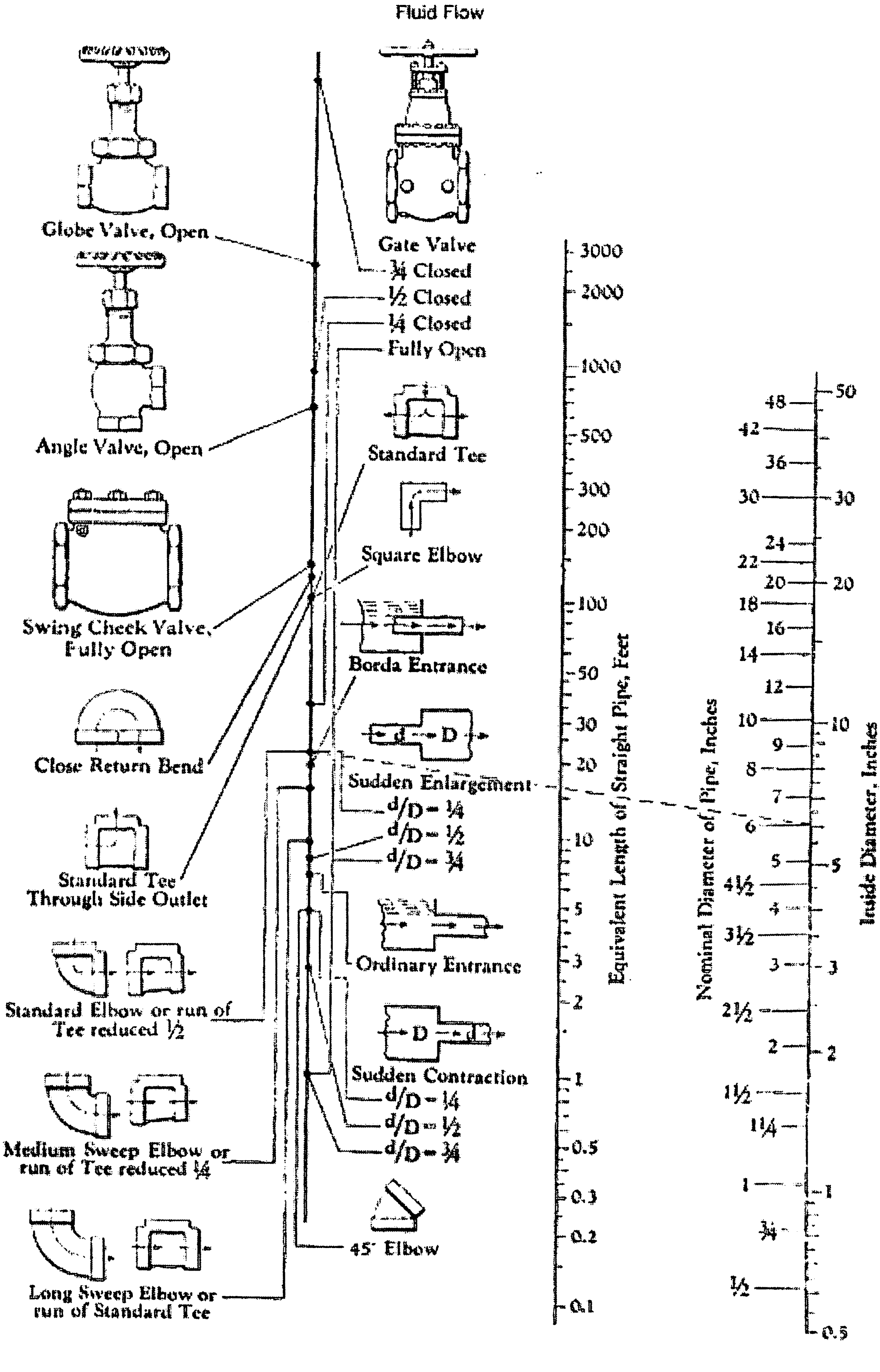


Fig 3.1 Equivalent length calculation

The friction loss in straight piping is calculated as below.

Fanning's equation (for Turbulent Flow; Re>2000)

$$\Delta P_f = \frac{4 \cdot f \cdot L}{D} \times \frac{\rho_\ell \cdot U_\ell^2}{2}$$

$$\text{where: } f = F(\mathrm{Re}, \varepsilon/D)$$

Hagen-Poiseuille's equation (for Laminar Flow ; Re<2000)

$$\Delta P_f = 32 \times \frac{\mu_\ell \cdot U_\ell \cdot L}{D^2}$$

The head loss in straight piping is calculated by following equation;

$$\Delta P = \Delta P_f + \Delta P_h$$

Total Pressure Drop is given as,

$$\Delta P = \Delta P_f + \Delta P_h$$

Friction Factor Calculation

Friction Factor f is derived by convergence of the following equation.

$$\frac{1}{\sqrt{f}} = -4 \cdot \log\left(\frac{\varepsilon/D}{3.71} + \frac{1.26}{\mathrm{Re}\sqrt{f}}\right)$$

Isothermal Gas Phase Pressure Drop Calculation

$$\frac{P_1^2 - P_2^2}{2P_1} = \left[\frac{4 \cdot f \cdot L}{D} + \ell n\left(\frac{P_1}{P_2}\right)^2\right] \cdot \frac{\rho_1 U_1^2}{2} + \left(\frac{P_1 + P_2}{2P_1}\right)^2 \cdot \rho_1 \cdot g \cdot (Z_2 - Z_1)$$

The above equation is derived based on the equations for Material Balance, Equation of Continuity and Ideal gas law.

The conditions of either upstream or downstream are used in the above equation and equation is solved by Newton-Raphson algorithm to get the conditions of the other end.

Two Phase Pressure Drop Calculations

Basic equation

$$\Delta P = \Delta P_f + \Delta P_h + \Delta P_m$$

Where, ΔP_m is derived as $\Delta P_m = (Ma)^2 \cdot \Delta P$. Therefore,

$$\Delta P = (\Delta P_f + \Delta P_h)/(1 - Ma^2)$$

Friction loss in straight pipe is as shown below

$$\Delta P_f = \Delta P_\ell + C \cdot \sqrt{\Delta P_\ell \cdot \Delta P_g} + \Delta P_g$$

Where, $$\mathbf{C} = \mathbf{C}' . \frac{1+10^{-200\left(\frac{\xi}{D}\right)}}{2}$$

$$C' = C_1 \quad (C_1 > C_2 \text{ or } C_2 > C_1 > C_3)$$
$$C' = C_2 \quad (C_3 > C_2 > C_1)$$
$$C' = C_3 \quad (C_2 > C_3 > C_1)$$
$$C_1 = 2 + \frac{32 \cdot \left[1 - 0.16(2.5 + \lambda)^2\right]^3}{1 + 0.005664 \; (\dot{m}_c)^{0.8}}$$

$$C_2 = \left(\frac{\rho_g}{\rho_\ell}\right)^{1/2} + \left(\frac{\rho_\ell}{\rho_g}\right)^{1/2}$$

$$C_3 = \frac{(\rho_g/\rho_l)^{1/8}}{\sqrt{X_g + (1 - X_g)(\rho_g + \rho_l)}}$$

$$\dot{m}c = \dot{m} \qquad (\dot{m} \geq 300)$$

$$\dot{m}c = 300 + \frac{(300 - \dot{m})^2}{40} \qquad (\dot{m} < 300$$

$$\lambda = \log_{10}\left[\left(\frac{\rho_g}{\rho_l}\right) \cdot \left(\frac{\mu_l}{\mu_g}\right)^{0.2}\right]$$

$$K = \frac{Cp}{Cv}$$

Mach= Mdot * (Xg/(press*100000)/Vapor Density)^0.5

Xg = Vapor Density *Vapor Velocity/ Mdot

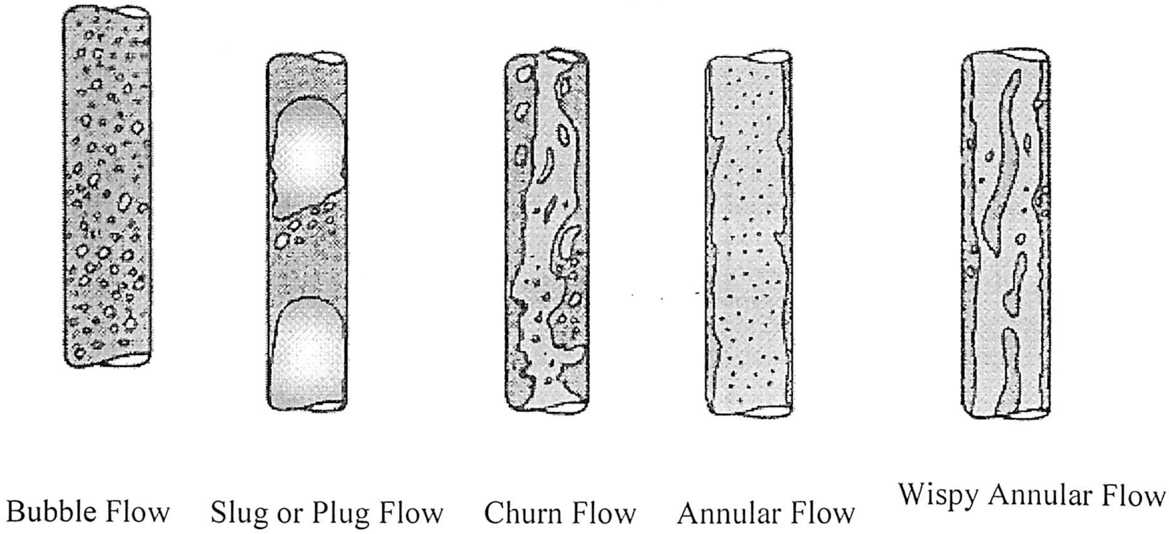
Head loss in two phase flow (Bar)

Head loss for two phase flow= mixed density*G* elevation difference/ 100000/ (1-Mach number^2)

$$Froude'sNo. = U/\sqrt{gD}$$

The flow pattern of the two phase fluid is determined by the Froude's number foe gas and liquid phase individually. The graphs referred are Taitel-duker's and Simpson's map.

Fig 3.2 flow patterns in vertical position for two phase



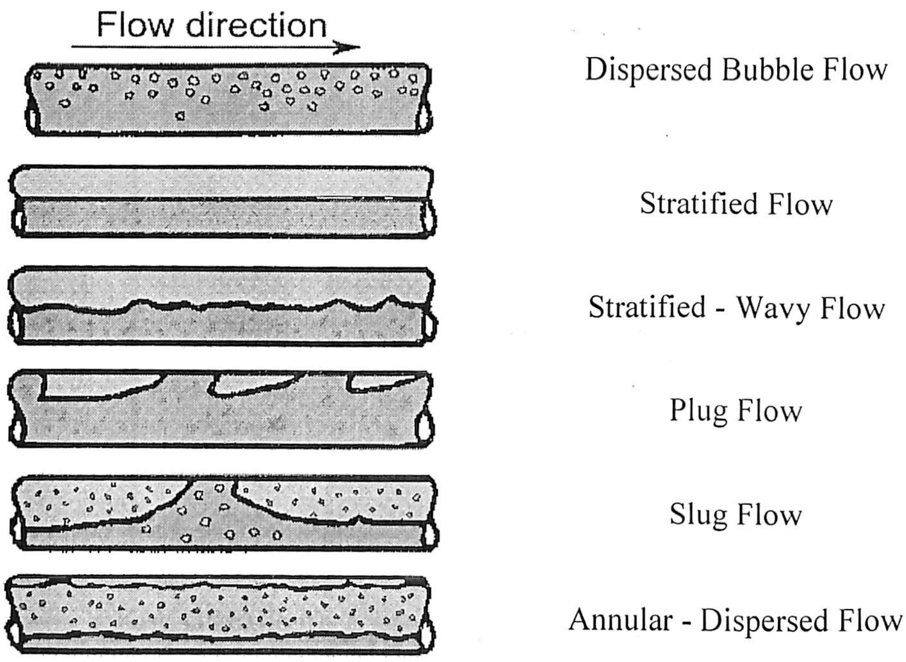


Fig 3.3 Flow patterns in horizontal position for the two phase flow

Refer appendix for the flow pattern for two phase flow graphs.

3.2 Design and Rating of condenser using HTRI

3.21 HTRI

As HTRI can be used for designing and rating of an heat exchanger it also helps in de-bottlenecking the heat exchanger already existing in plant. The condenser of the de- ethanizer system of the fractionation unit is rated to reduce the over designing of the condenser.

The ways to improve Heat exchanger and air coolers are available that too without considerable expenses. Resources are limited to build new grass root plant. Plants can be de-bottlenecked without investing on new equipment and emphasizing on utilizing the existing equipments to its fullest.

During the capacity augmentation of an Hydrocarbon Processing Industries (HPI) plant. Existing Heat exchanger are usually unable to handle the increased throughputs and heat duties.

Usually heat exchangers are overdesign to overcome these kinds of situations. So that the heat exchanger can work even if plant is extended.

Over designing alone cannot be a proper solution, there is other option such as changing from series to parallel operation. On altering the configuration of two shells in series to two shells in parallel reduces the pressure drop and vibration analysis indicated a safe design.

The overdesign for the condenser in this case is reduced to 14% from 25%. This was done to minimize the fouling and hence the losses occurring due to it. The rating of condenser is done on HTRI which supports all kind of changes done to get the excellent design and then to optimize it to the extent it can be proved as total economic heat exchanger.

The input summary given to the HTRI is as follow.

3.3 Separator sizing

Separator sizing is done on the basis of the gas capacity, the liquid capacity, and other parameters like pressure drop through the separator.

Separator Design Using Basic Separation Principles:

Souders-Brown equation for calculating the gas capacity of gas-liquid separators:

$$V_g = K[(\rho_l-\rho_g)/\rho_g]^{0.5}$$

The separation coefficient, K, is an empirical constant given as follows

Separator size	Range of K	Most commonly used K value
Vertical	0.06 to 0.35	0.117 without mist extractor 0.167 with a mist extractor
Horizontal	0.40 to 0.50	0.382 with a mist extractor
Spherical	-	0.35 with a mist extractor

Table 3.1 separation coefficient K

Souder's-Brown equation can also be used for other designs such as bubble cap or trayed towers for dehydration or desulfurization units, and for sizing the mist extractors.

And the following things are given with:

Wire mesh mist eliminators K= 0.35

Bubble cap tray columns K = 0.16 for 24 in. spacing.

Valve tray columns K = 0.18 for 24 in. spacing.

The equation will give the terminal settling velocity, based on this minimum vapor area can be calculated as

$$A_{G=}\frac{Q_G}{V_{ALLOW}}.$$

Choose a vessel diameter in such way that available vapor area is greater than minimum vapor area. The required separator liquid volume can be calculated from

$$V_L = (q_L\ t)/1440$$

Typical retention time for natural gas-oil separation is 1-3 minutes.

Retention time is affected by oil relative density, composition foaming, presence of solids and emulsions.

Liquid Degassing

If where vapor carry-under is not allowed, the vessel diameter shall satisfy the liquid de-gassing criterion: In practice it can be assumed that if bubbles larger than 200 μ_m in size are able to escape, the carry-under will be negligible. This means that the downward velocity of the liquid shall satisfy the following requirement:

$$\frac{4\ Q_{L\max}}{\pi D^2} > \frac{D_P{}^2 g(\rho_V - \rho_L)}{18\mu}$$

Vertical separators are usually selected when the gas-liquid ratio is high or total gas volumes are low. For gas/liquid separation, a vertical vessel should normally be selected for the following reasons:

- A smaller plan area is required (critical on offshore platforms);

- Easier solids removal;
- Liquid removal efficiency does not vary with liquid level (area in vessel available for gas flow remains constant)
- Vessel volume is generally smaller.

3.4 Simulation of De- ethanizer unit

The de- ethanizer unit of the fractionation system is simulated using HYSIS. The flow sheet includes a separator (95-MBD63111), a fractionation column (95-MAF63112), de- ethanizer condenser (95-HBG63111), de- ethanizer reflux pump (95-PBA63111.12) and de-ethanizer reflux drum (95-MBD63113). Process description is as given in the chapter 1.

The in-built critical properties for the specific composition of each stream are considered to simulate the process. The simulated process would be discussed in the result and conclusion section.

CHAPTER 4 – RESULT, CONCLUSION AND RECOMMENDATION

4.1 Work sheets

4.11 Hydraulics work sheet

For Straight Pipe (Suction Side)				
section 607,607A,607B				
Mass Flow Rate	82442.7652	lb/hr	37396	Kg/hr
Diameter	8	inches	0.205	m
Length	63.6056451	feet	19.3870	meters
Density	26.8838544	lb/ft3	430.831	kg/m3
Specific Volume	0.03719705	ft3/lb	0.002	m3/Kg
Temperature			-24.413	*C
Viscosity			0.0680	mpa.s
Fluid Velocity			0.88	m/sec
Reynolds no			**1139073**	
Friction Factor			0.00379	from formula
Pressure Drop Straight Pipe			**237.53745**	Pa
			0.237537448	Kpa

Fittings			
Bend/Tee/Valve	Quantity	L/D	Equivalent L (feets)
180° bend = 5D	0	28	0.000
90° bend = 5D	4	30	80.000
90° bend =1.5D	0	20	0.000
45° bend =1.5D	1	16	10.667
TEE (Straight out)	1	20	13.333
TEE (side outlet)	0	65	0.000
Gate valve (Open)	0	13	0.000
Gate valve (¼-closed)	0	39	0.000
Gate Valve (½-closed)	0	195	0.000
Gate vavle (¾-closed)	0	300	0.000
Membrane valve	0	200	0.000
Ball valve (spherical plug valve)	1	18	12.000
Needle valve	0	1000	0.000
Butterfly valve (larger then 6")	0	20	0.000
Foot valve with strainer(hinged type)	0	75	0.000
Foot valve with strainer(lift type)	0	420	0.000
Globe valve	0	300	0.000
Nozzle (suction nozzle on vessel)	0	32	0.000
Check valve (in-line ball type)	0	150	0.000
Check valve (swing type)	0	135	0.000
Filter (Y-type and bucket type)	0	250	0.000
Equivalent fittings length	116.000		

Total Length	**54.753**	m
Pressure drop due to valves	**5.000**	Kpa
Total Pressure Drop	**5.670854342**	KPa

Pressure Drop Across Equipment

Equipment	Quantity	Pressure Drop	T. P Drop
Reflux Drum	1	2667.189	2667.189
XXXXX	0	0	0
XXXXX	0	0	0
Total Pressure Drop (Suction)		**2667.1890**	**KPa**
Total Pressure Drop L1		**2667.1890**	**Kpa**

Calculation For Suction Head

Height of Liquid in reflux Drum	0.88	m	EQP elevatn	7.675
Suction Static Head	7.555	m	pump elevation	1
suction pressure head	631.07 631.07			
Suction Friction Head	1.34175289	m		
Total Suction Head			**637.28**	m

For Straight Pipe (Discharge Side)				
Section 607C,607D,607E				
Mass Flow Rate	82442.76525	lb/hr	37396	Kg/hr
Diameter	6	inches	0.154	m
Length	46.90728	feet	14	meters
Density	26.8838544	lb/ft3	430.831	kg/m3
Specific Volume	0.037197047	ft3/lb	0.002	m3/Kg
Temperature			-24.413	*C
Viscosity			0.0068	mPa.s
			1.55	m/sec
Reynolds no			**15142949**	
Friction Factor			0.003912	from formula
Pressure Drop For Straight Pipe			**753.988820**	Pa
			0.75398882	Kpa

Fittings

Bend/Tee/Valve	Quantity	L/D	Equivalent L (feets)
180° bend = 5D	0	28	0.000
90° bend = 5D	5	30	75.000
90° bend =1.5D	0	20	0.000
45° bend =1.5D	0	16	0.000
TEE (Straight out)	1	20	10.000
TEE (side outlet)	0	65	0.000
Gate valve (Open)	0	13	0.000
Gate valve (¼-closed)	0	39	0.000
Gate Valve (½-closed)	0	195	0.000
Gate vavle (¾-closed)	0	300	0.000
Membrane valve	0	200	0.000
Ball valve (spherical plug valve)	1	18	9.000
Needle valve	0	1000	0.000
Butterfly valve (larger then 6")	0	20	0.000
Foot valve with strainer(hinged type)	0	75	0.000
Foot valve with strainer(lift type)	0	420	0.000
Globe valve	0	340	0.000
Nozzle (suction nozzle on vessel)	0	32	0.000
Check valve (in-line ball type)	0	150	0.000
Check valve (swing type)	0	135	0.000
Filter (Y-type and bucket type)	0	250	0.000
Equivalent fittings length	94.000		

Total Length			42.960	m
Pressure drop due to valves			467.148	Kpa
Total Pressure Drop			480.225	Kpa
Pressure Drop Across Equipment				
Equipment	Quantity	Pressure Drop		T. P Drop
Fractionation column	1	2690.045		2690.045
XXXXX	0	0		0
XXXXX	0	0		0
Total Pressure Drop (Discharge)		2690.0450		Kpa

Calculation For Discharge Head				
Height of Liquid in column	36.5	m		
Discharge Static Head	35.5	m		
Pressure in column	2690.05	Kpa		
Discharge Pressure Head	636.48	m		
Discharge Friction Head	113.62367	m		
Total Discharge Head			**785.60**	m

TOTAL HEAD OF PUMPS	**148.32**	m
NPSH Calculation		
Vapour Pressure at Process Temp	2667.1890	Kpa
Absolute Pressure in Vessel	2667.1890	Kpa
NPSH Available	6.213247108	m
NPSH Required	5.213247108	m

Power Calculation		
Efficiency	0.5	
Pressure diff	626.86	Kpa
Shut off presure	877.6040729	KPa

For Straight Pipe (Discharge Side)				
section 608.608A,608B				
Mass Flow Rate	82442.76525	lb/hr	37396	Kg/hr
Diameter	4.026	inches	0.154	m
Length	196.3736	feet	60	meters
Density	26.8838544	lb/ft3	430.831	kg/m3
Specific Volume	0.037197047	ft3/lb	0.002	m3/Kg
Temperature			-24.413	*C
Viscosity			0.0068	mPa.s
			1.54	m/sec
Reynolds no			**14996594**	
Friction Factor			0.003913	from formula
Pressure Drop For Straight Pipe			**3096.585393**	Pa
			3.096585393	Kpa

Fittings			
Bend/Tee/Valve	Quantity	L/D	Equivalent L (feets)
180° bend = 5D	0	28	0.000
90° bend = 5D	12	30	120.780
90° bend =1.5D	0	20	0.000
45° bend =1.5D	2	16	10.736
TEE (Straight out)	2	20	13.420
TEE (side outlet)	0	65	0.000
Gate valve (Open)	0	13	0.000
Gate valve (¼-closed)	0	39	0.000
Gate Valve (½-closed)	0	195	0.000
Gate vavle (¾-closed)	0	300	0.000
Membrane valve	0	200	0.000
Ball valve (spherical plug valve)	2	18	12.078
Needle valve	0	1000	0.000
Butterfly valve (larger then 6")	0	20	0.000
Foot valve with strainer(hinged type)	0	75	0.000
Foot valve with strainer(lift type)	0	420	0.000
Globe valve	0	340	0.000
Nozzle (suction nozzle on vessel)	0	32	0.000
Check valve (in-line ball type)	0	150	0.000
Check valve (swing type)	0	135	0.000
Filter (Y-type and bucket type)	0	250	0.000
Equivalent fittings length	157.014		
Total Length		107.740	m
Total Pressure Drop		8.121049923	Kpa

Line Sizing		
L1		
Process Fluid	ethane	
Fluid Phase	Liquid	
Mass Flow Rate	37396	Kg/hr
Density	431	Kg/m3
Volumetric Flow rate	87	m^3/hr
Fluid Velocity	0.877	m/sec
Area of Flow	0.0274924	m^2
Diameter Required	187.142	mm
Line Size Final	205.000	mm
	8.000	inch
L2		
Process Fluid	ethane	
Fluid Phase	Liquid	
Mass Flow Rate	37394	Kg/hr
Density	431	Kg/m3
Volumetric Flow rate	87	m^3/hr
Fluid Velocity	1.552	m/sec
Area of Flow	0.01553533	m^2
Diameter Required	140.678	mm
Line Size Final	154.000	mm
	6.000	inch

University of Petroleum
Dehradun
India
Phone : (+91)
Fax : (+91)

Document Number
P100-DES-TDS-01

Pump Data Sheet

Sheet __1_ of __1_

Project No	XXXX
Date	27/1/2011
Revision	1
Prepared b	XXXX
Verified by	XXXX

PUMPS

LARSEN AND TOUBRO CHIYODA LIMITED

Item Number	P101	Name	HC Transfer Pump	Location	Plant
Service		Type		Operation	hrs/day

Head	Suction		Discharge		Differential	
Static	7.555	m	35.5	m	27.945	m
Pressure	631.07	m	636.48	m	5.40784549	m
Friction	1.3417529	m	113.623669	m	112.281916	m
Total	637.28	m	785.601956	m	148.32	m
NPSH available	6.2132471	m	Rating Head		1.4	

FLUID CHARACTERISTICS

Fluid Name	ETHANE	Fluid Type	liquid	Corrosive Compound		no
Solid Particles	No	Pumping Temp(°C)	-24.413	Normal		Maximum
Density at PT	430.831	Kg/m^3	Vapour Pressure	2667.189	KPa	
Viscosity	0.068	mPa	Frezzing or Pour Point(°C)		NA	
Capacity	m^3/hr	Normal	104	Minimum	Maximum	124.99

PUMP SELECTION

Mgfr		Size & Type	Centrifugal	Modle	Case Type	
Drive		RPM	Design Efficiency			
Rating BHP		Non overload BHP			NPSH Req.	5.2132471
Suction Location	Horizontal	Discharge Location	Horizontal	Curve No.	XXXX	

MATERIAL OF CONSTRUCTION

Case	SS		Impeller	SS	Shaft	Carbon Steel
Base Plate	SS	Coolent	Same process Fluid	Mechanical seal		
Suction Size (inch)	8	Rating	Discharge Siz	6	Rating	

DRIVER

Electric Motor	(Existing or New)	Rating		
Type		Mfg	HP	
RPM	Frame	Volts	Phase	Cycle
Elec. Classification	Connections	Belts	Gear	
Inlet Stream		Psi(g) @	°F	
Exhaust Stream		Psi(g) @	°F	
Steam Rate	Lb/hr			

ASSESSORIES

Safety Valve at Discharge	Yes	No	Priming Nozzle	Yes	No
Pulsating Dampner	Yes	No	Heating Arrangment	Not Required	
Remarks	Tank from where pump takes Suction(meters)		-3 meters	Discharge	
Pump insulation	Hot	Cold	Not Required		
Pump heating Arrangment		Insulation	Type	Material	
Painting	Refer Notes	Temp Above 75°C	Temp Below 75°C		

TWO PHASE CALCULATIONS			
Fluid	ethane		
Total mass flow	51314.524	Kg/Hr	
Total vol flow	439.618	m3/Hr	
Properties	liquid	vapor	
Volumetric flow rate	86.799	352.819	m3/Hr
Mass Flow rate	37395.793	13918.731	Kg/Hr
Density	430.831	39.45	Kg/m3
Mol.wt	-	23.16	
Liquid surface tension	5.791	-	Dyn/cm
Vapor Compressibility factor	0.757	-	
Viscosity	0.068	0.01	mPa.S
Velocity	0.519	2.11199133	m/s
Nre	4265226.900	2655774	
ϵ/D	0.000196078		
Friction factor	0.003480914	0.00351657	
	- 0.000293392	7.6203E-05	
ΔPl(Press. Drop for liquid)	0.124285955	KPa	
Inlet pressure, P1	2668027	Pa	
P2	2668014.866	Pa	
Goal Seek	- 8.63775E+13		
ΔPg(Press. Drop for vapor)	0.012134482	KPa	
mass flux.m	307.0280081	Kg/m3.s	307.028
λ	- 0.871758155		
C1	4.25589267		
C2	3.607285007		
C3	0.040050988	Xg	0.27137
C′	4.25589267		
C	4.072164448		
ΔPf(Frictional loss total)	0.294562339	KPa	

Flow Pattern		
Froude no, liquid	0.328302177	
Froude no, vapor	1.335326715	
Head loss Calculation		
Mach	5.853201307	
λL	0.19744187	
Mixed Density	116.7249965	Kg/m3
head loss	- 0.136851076	Kpa

Temperature	-24.4	oC			
Total velocity	2.64	m/s	inlet elev.Z1	13.7	m
Total length	39.190	m	outlet elev.Z2	9.725	m
Diameter	0.255	m			
SCH	S40S				
Roughness	0.00005	m			
Straight length	13.69	m			

Fittings			
Bend/Tee/Valve	Quantity	L/D	Equivalent length
180° bend = 5D	0	28	0.000
90° bend = 5D	2	30	15.300
90° bend =1.5D	0	20	0.000
45° bend =1.5D	0	16	0.000
TEE (Straight out)	2	20	10.200
TEE (side outlet)	0	65	0.000
Gate valve (Open)	0	13	0.000
Gate valve (¼-closed)	0	39	0.000
Gate Valve (½-closed)	0	195	0.000
Gate vavle (¾-closed)	0	300	0.000
Membrane valve	0	200	0.000
Ball valve (spherical plug valve)	0	18	0.000
Needle valve	0	1000	0.000
Butterfly valve (larger then 6")	0	20	0.000
Foot valve with strainer(hinged type)	0	75	0.000
Foot valve with strainer(lift type)	0	420	0.000
Globe valve	0	300	0.000
Nozzle (suction nozzle on vessel)	0	32	0.000
Check valve (in-line ball type)	0	150	0.000
Check valve (swing type)	0	135	0.000
Filter (Y-type and bucket type)	0	250	0.000
Equivalent fittings length	25.500		

4.12 Separator work sheet

	Separator sizing	
Process Fluid	hydrocarbon mixture	
phase	two phase	
liquid density	510.7	kg/m3
gas density	34.4	kg/m3
total flow rate	59122	kg/hr
liquid flow rate	49914	kg/hr
gas flow rate	9208	kg/hr
liquid viscosity	0.115	m.Pas
gas viscosity	0.011	m.Pas
constant K	0.117	(as per separator)
allowable velosity	0.4353584	m/sec
minimum gas area	0.17078803	m2
Dia from min area	0.46643796	m
Dia selected	0.55	m
retention time	3	min
separator liq vol	4.88682201	m3
height for liq holdup	28.6133749	m

For half open pipe device		
X1	0.165	m
X2	0.2601666	m
X3	0.495	m
total height	29.533541	m

Nozzle sizing		
Ql	48.9	m3/hr
Qg	133.8372093	m3/hr
lambda	0.267597389	
Avg density	161.8566362	kg/m3
Avg vol flow rate	365.2738708	m3/hr
Avg velosity	0.39	m/s
dia	0.260166575	m

4.13 Condenser work sheet

OUTPUT SUMMARY

Process Conditions		Cold Shellside		Hot Tubeside	
Fluid name		propane		ethane	
Flow rate	(kg/s)		15.9091		15.6797
Inlet/Outlet Y	(Wt. frac vap.)	0.175	1.000	0.838 *	0.273
Inlet/Outlet T	(Deg C)	-30.71	-30.79	-3.40	-24.40
Inlet P/Avg	(kPa)	174.003	173.659	2690.04	2685.01
dP/Allow.	(kPa)	0.688	2.000	10.061	30.000
Fouling	(m2-K/W)		0.000100		0.000200

Exchanger Performance

Shell h	(W/m2-K)	2158.17	Actual U	(W/m2-K)	707.66
Tube h	(W/m2-K)	2340.31	Required U	(W/m2-K)	616.41
Hot regime	(--)	Transition	Duty	(MegaWatts)	5.4535
Cold regime	(--)	Flow	Area	(m2)	632.352
EMTD	(Deg C)	14.0	Overdesign	(%)	14.80

Shell Geometry			Baffle Geometry		
TEMA type	(--)	BKU	Baffle type	(--)	Support
Shell ID	(mm)	863.602	Baffle cut	(Pct Dia.)	
Series	(--)	1	Baffle orientation	(--)	
Parallel	(--)	1	Central spacing	(mm)	901.568
Orientation	(deg)	0.00	Crosspasses	(--)	1

Tube Geometry			Nozzles		
Tube type	(--)	Plain	Shell inlet	(mm)	250.000
Tube OD	(mm)	19.050	Shell outlet	(mm)	500.000
Length	(m)	10.000	Inlet height	(mm)	22.501
Pitch ratio	(--)	1.2500	Outlet height	(mm)	774.326
Layout	(deg)	30	Tube inlet	(mm)	350.000
Tubecount	(--)	1036	Tube outlet	(mm)	200.000
Tube Pass	(--)	2			

Thermal Resistance, %		Velocities, m/s		Flow Fractions	
Shell	32.79	Shellside	0.67	A	0.000
Tube	35.76	Tubeside	1.74	B	1.000
Fouling	23.82	Crossflow	0.49	C	0.000
Metal	7.641	Window	0.00	E	0.000
				F	0.000

Supplementary Results

Released to the following HTRI Member Company:

poo

DELL

Xist E Ver. 5.00 2/24/2007 16:38 SN: Friendsl SI Units

Rating - Horizontal Multipass Flow TEMA BKU Shell With No Baffles

Externally Enhanced Tube Geometry			Internally Enhanced Tube Geometry		
Type		Plain	Type		None
Fin density	(fin/meter)		Thickness	(mm)	
Fin height	(mm)		Pitch	(L/D)	
Fin thickness	(mm)				
Root diameter	(mm)				
Area/length	(m2/m)	0.1584			

Mean Metal Temperatures

Mean shell temperature -30.43 (C)

Mean tube metal temperature in each tubepass, (C)

Tube Pass	Inside	Outside	Radial
1	-23.01	-24.47	-23.78
2	-26.13	-26.71	-26.43

Shellside Performance

Nom vel, X-flow/window			0.49 / 0.00		
Kettle Recirculation Ratio (Internal/Feed)				3.55	
Flow fractions for vapor phase					
A=0.0000	B=1.0000	C=1.343e-5	E=0.0000	F=0.0000	

Shellside Heat Transfer Corrections

Total	Beta	Gamma	End	Fin
1.054	1.000	1.054	1.000	1.000

Pressure Drops (Percent of Total)

Cross	Window	Ends	Nozzle	Shell	Tube
0.00	0.00	0.00	Inlet	85.04	2.95
MOMENTUM		-0.00	Outlet	14.96	8.10

Two-Phase Parameters

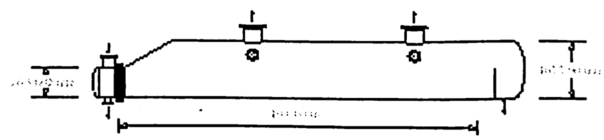
Method	Inlet	Center	Outlet	Mix F
RPM	Shear	Transition	Transition	0.5631
PP/TBR	Flow	Flow	Flow	

Process Data		Cold Shellside		Hot Tubeside	
Fluid name		propane		ethane	
Fluid condition			Boil. Liquid		Cond. Vapor
Total flow rate	(kg/s)		15.9091		15.6797
Weight fraction vapor, In/Out	(--)	0.175	1.000	0.838 *	0.273
Temperature, In/Out	(Deg C)	-30.71	-30.79	-3.40	-24.40
Temperature, Average/Skin	(Deg C)	-30.8	-26.55	-13.9	-22.33
Wall temperature, Min/Max	(Deg C)	-27.60	-22.28	-27.17	-20.20
Pressure, In/Average	(kPa)	174.003	173.659	2690.04	2685.01
Pressure drop, Total/Allowed	(kPa)	0.688	2.000	10.061	30.000
Velocity, Mid/Max allow	(m/s)	0.67		1.74	
Boiling range/Mole fraction inert	(Deg C)		0.000		0.0
Average film coef.	(W/m2-K)		2158.17		2340.31
Heat transfer safety factor	(--)		1.000		1.000
Fouling resistance	(m2-K/W)		0.000100		0.000200

Overall Performance Data

Overall coef., Reqd/Clean/Actual	(W/m2-K)	616.41 /	928.85 /	707.66
Heat duty, Calculated/Specified	(MegaWatts)	5.4535 /		
Effective overall temperature difference	(Deg C)	14.0		
EMTD = (MTD) * (DELTA) * (F/G/H)	(Deg C)	13.97 *	1.0000	* 1.0000

See Runtime Messages Report for warnings.



Exchanger Fluid Volumes

Approximate shellside (L)	12756.5
Approximate tubeside (L)	3403.8

Shell Construction Information

TEMA shell type	BKU		Shell ID	(mm)	863.602
Shells Series	1 Parallel	1	Total area	(m2)	637.476
Passes Shell	1 Tube	2	Eff. area	(m2/shell)	632.352
Shell orientation angle (deg)	0.00		Kettle ID	(mm)	1622.90
Impingement present	No				
Pairs seal strips	0		Passlane seal rods (mm)	0.000	No. 0
Shell expansion joint	No		Full support at U-Bend		No

Weight estimation Wet/Dry/Bundle 33740.4 / 17591.8 / 7221.52 (kg/shell)

Baffle Information

Type		Support	Baffle cut (% dia)	
Crosspasses/shellpass		1	No. (Pct Area)	(mm) to C.L
Central spacing	(mm)	901.568	1	
Inlet spacing	(mm)	0.000	2	
Outlet spacing	(mm)	0.000	Support plates/baffle space	10
Baffle thickness	(mm)	9.525		

Tube Information

Tube type		Plain	Tubecount per shell		1036
Length to tangent	(m)	10.000	Pct tubes removed (none)		
Effective length	(m)	10.199	Outside diameter	(mm)	19.050
Total tubesheet	(mm)	82.646	Wall thickness	(mm)	1.470
Area ratio	(out/in)	1.1825	Pitch (mm) 23.8125	Ratio	1.2500
Tube metal		304 Stainless steel (18 Cr, 8 Ni)	Tube pattern (deg)		30

H. T. Parameters		Shell	Tube	
Overall wall correction				
Midpoint	Prandtl no.			
Midpoint	Reynolds no.	4102	47404	
Bundle inlet	Reynolds no.	4044	98402	
Bundle outlet	Reynolds no.	4886	42071	
Fouling layer	(mm)			

Thermal Resistance

Shell	Tube	Fouling	Metal	Over Des
32.79	35.76	23.82	7.641	14.80

Total fouling resistance	0.00034
Differential resistance	0.00021

Shell Nozzles				Liquid
Inlet at channel end-Yes		Inlet	Outlet	Outlet
Number at each position		2	2	1
Diameter	(mm)	250.000	500.000	50.000
Velocity	(m/s)	7.17	10.09	0.00
Pressure drop	(kPa)	0.585	0.103	0.000
Height under nozzle	(mm)	22.501	774.326	1622.90
Nozzle R-V-SQ	(kg/m-s2)	1169.17	411.31	0.00
Shell ent.	(kg/m-s2)	7.33	10.40	

		Inlet	Outlet	Liquid
Tube Nozzle		RADIAL	RADIAL	Outlet
Diameter	(mm)	350.000	200.000	
Velocity	(m/s)	3.32	4.67	
Pressure drop	(kPa)	0.297	0.815	
Nozzle R-V-SQ	(kg/m-s2)	540.26	2328.94	

Annular Distributor		Inlet	Outlet
Length	(mm)		
Height	(mm)		
Slot area	(mm2)		

Diametral Clearances (mm)

Baffle-to-shell	Bundle-to-shell	Tube-to-baffle
4.7625	15.0249	0.3969

VIBRATION ANALYSIS

Shellside condition		Boil. Liquid	(Level 2.2)	
Axial stress loading	(Mpa)	0.000	Added mass factor	1.761
Beta		4.000		
Position In The Bundle		**Bottom**	**Center**	**Top**
Length for natural frequency	(m)	0.902	0.902	0.902
Length/TEMA maximum span	(--)	0.592	0.592	0.592
Number of spans	(--)	11	11	11
Tube natural frequency	(Hz)	50.4 +	58.8	60.1
Shell acoustic frequency	(Hz)			
Flow Velocities		**Bottom**	**Center**	**Top**
Window parallel velocity	(m/s)	0.00	0.00	0.00
Bundle crossflow velocity	(m/s)	0.15	0.53	4.48
Bundle/shell velocity	(m/s)	8.443e-6	6.265e-5	2.871e-4
Fluidelastic Instability Check		**Bottom**	**Center**	**Top**
Log decrement	HTRI	0.084	0.065	0.046
Critical velocity	(m/s)	2.77	5.68	10.00
Baffle tip cross velocity ratio	(--)	0.000	0.000	0.000
Average crossflow velocity ratio	(--)	0.054	0.093	0.448
Acoustic Vibration Check		**Bottom**	**Center**	**Top**
Vortex shedding ratio	(--)			
Chen number	(--)			
Turbulent buffeting ratio	(--)			
Tube Vibration Check		**Bottom**	**Center**	**Top**
Vortex shedding ratio	(--)	0.028	0.100	0.857
Turbulent buffeting ratio	(--)	0.045	0.163	1.387
Parallel flow amplitude	(mm)	0.000	0.000	0.000
Crossflow amplitude	(mm)	0.002	0.003	0.097
Turbulent buffeting amplitude	(mm)	0.001	0.001	0.027
Tube gap	(mm)	4.763	4.763	4.763
Crossflow RHO-V-SQ	(kg/m-s2)	12.10	21.37	339.23
Bundle Entrance/Exit (analysis at first tube row)			**Entrance**	**Exit**
Fluidelastic instability ratio		(--)		
Vortex shedding ratio		(--)		
Crossflow amplitude		(mm)		
Crossflow velocity		(m/s)		
Turbulent buffeting amplitude		(mm)		
Tubesheet to inlet/outlet support		(mm)	None	None
Shell Entrance/Exit Parameters			**Entrance**	**Exit**
Impingement plate			No	
Flow area		(m2)	0.617	1.212
Velocity		(m/s)	0.57	1.60
RHO-V-SQ		(kg/m-s2)	7.33	10.40

Shell type	BKU	Baffle type	Support
Tube type	Plain	Baffle layout	
Pitch ratio	1.2500	Tube diameter, (mm)	19.050
Layout angle	30	Tube material	304 Stainless steel (18 Cr, 8 Ni)
Number U-Bend supports	0	Supports/baffle space	10

HEAT EXCHANGER RATING DATA SHEET

Service of Unit		Item No.
Type BKU	Orientation Horizontal	Connected In 1 Parallel 1 Series
Surf/Unit (Gross/Eff) 637.48 / 632.35 m2	Shell/Unit 1	Surf/Shell (Gross/Eff) 637.48 / 632.35 m2

PERFORMANCE OF ONE UNIT

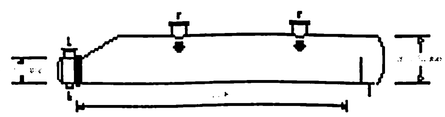
Fluid Allocation		Shell Side		Tube Side	
Fluid Name		propane		ethane	
Fluid Quantity, Total	kg/s	15.9091		15.6797	
Vapor (In/Out)	wt%	17.5	100.0	83.8	27.3
Liquid	wt%	82.5	0.0	16.2	72.7
Temperature (In/Out)	C	-30.71	-30.79	-3.40	-24.40
Density	kg/m3	4.0385 V/L 574.02	4.0244	41.969 V/L 424.96	35.433 V/L 440.98
Viscosity	mN-s/m2	0.0070 V/L 0.1913	0.0070	0.0088 V/L 0.0649	0.0088 V/L 0.0681
Specific Heat	kJ/kg-C	1.5648 V/L 2.3011	1.5647	2.4731 V/L 8.2394	2.0505 V/L 17.295
Thermal Conductivity	W/m-C	0.0110 V/L 0.13	0.0110	0.0229 V/L 0.0974	0.0228 V/L 0.1082
Critical Pressure	kPa				
Inlet Pressure	kPa	174.003		2690.04	
Velocity	m/s		0.67		1.74
Pressure Drop, Allow/Calc	kPa	2.000	0.688	30.000	10.061
Average Film Coefficient	W/m2-K	2158.17		2340.31	
Fouling Resistance (min)	m2-K/W	0.000100		0.000200	

Heat Exchanged 5.4535 MegaWatts	MTD (Corrected) 14.0 C	Overdesign 14.80 %
Transfer Rate, Service 616.41 W/m2-K	Calculated 707.66 W/m2-K	Clean 928.85 W/m2-K

CONSTRUCTION OF ONE SHELL

			Shell Side	Tube Side
Design Pressure		kPaG	1700.03	3250.05
Design Temperature		C	65.00	158.00
No Passes per Shell			1	2
Flow Direction				Downward
Connections Size & Rating	In	mm	2 @ 250.000	1 @ 350.000
	Out	mm	2 @ 500.000	1 @ 200.000
	Liq. Out	mm	1 @ 50.000	@

Sketch (Bundle/Nozzle Orientation)



Tube No. 1036 OD 19.050 mm	Thk(Avg) 1.470 mm	Length 10.000 m Pitch 23.812 mm Layout 30
Tube Type Plain	Material 304 STAINLESS STEEL (18 CR, 8 NI)	Pairs seal strips 0
Shell ID 863.602 mm	Kettle ID 1622.90 mm	Passlane Seal Rod No. 0
Cross Baffle Type SUPPORT	%Cut (Diam)	Impingement Plate None
Spacing(c/c) 901.568 mm	Inlet mm	No. of Crosspasses 1
Rho-V2-Inlet Nozzle 1169.17 kg/m-s2	Shell Entrance 7.33	Shell Exit 10.40 kg/m-s2
	Bundle Entrance	Bundle Exit kg/m-s2
Weight/Shell 17591.8	Filled with Water 33740.4	Bundle 7221.52 kg

Notes: Supports/baffle space = 10.	Thermal Resistance, %		Velocities, m/s		Flow Fractions	
	Shell	32.79	Shellside	0.67	A	0.000
	Tube	35.76	Tubeside	1.74	B	1.000
	Fouling	23.82	Crossflow	0.49	C	0.000
	Metal	7.64	Window	0.00	E	0.000
					F	0.000

HEAT EXCHANGER SPECIFICATION SHEET

			Job No.		
Customer			Reference No.		
Address			Proposal No.		
Plant Location			Date	2/24/2007	Rev
Service of Unit			Item No.		
Size	863.60 - 1622.90 x 9999.88 mr Type BKU	Horz.	Connected In	1 Parallel	1 Series
Surf/Unit (Gross/Eff)	637.48 / 632.35 m2	Shell/Unit 1	Surf/Shell (Gross/Eff)	637.48 / 632.35 m2	

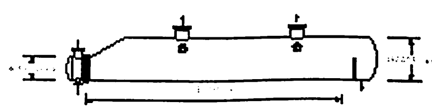
PERFORMANCE OF ONE UNIT

Fluid Allocation		Shell Side		Tube Side	
Fluid Name		propane		ethane	
Fluid Quantity, Total	kg/hr	57272.8		58446.8	
Vapor (In/Out)		10022.7	57272.8	47285.2	15402.6
Liquid		47250.1		9161.68	41044.3
Steam					
Water					
Noncondensables					
Temperature (In/Out)	C	-30.71	-30.79	-3.40	24.40
Specific Gravity		0.5743		0.4252	0.4412
Viscosity	mN-s/m2	0.0070 V/L 0.1913	0.0070	0.0088 V/L 0.0849	0.0088 V/L 0.0681
Molecular Weight, Vapor					
Molecular Weight, Noncondensables					
Specific Heat	kJ/kg-C	1.5648 V/L 2.301	1.5647	2.4731 V/L 8.239	2.0505 V/L 17.295
Thermal Conductivity	W/m-C	0.0110 V/L 0.130	0.0110	0.0229 V/L 0.097	0.0228 V/L 0.108
Latent Heat	kJ/kg	415.678	415.785	322.508	336.706
Inlet Pressure	kPa	174.003		2690.04	
Velocity	m/s	0.67		1.74	
Pressure Drop, Allow/Calc	kPa	2.000	0.688	30.000	10.061
Fouling Resistance (min)	m2-K/W	0.000100		0.000200	
Heat Exchanged W	5453505	MTD (Corrected)		14.0 C	
Transfer Rate, Service	616.41 W/m2-K	Clean	928.85 W/m2-K	Actual	707.66 W/m2-K

CONSTRUCTION OF ONE SHELL

		Shell Side	Tube Side
Design/Test Pressure	kPaG	1700.03 /	3250.05 /
Design Temperature	C	65.00	158.00
No Passes per Shell		1	2
Corrosion Allowance	mm		
Connections Size & Rating — In	mm	2 @ 250.000	1 @ 350.000
Out	mm	2 @ 500.000	1 @ 200.000
Intermediate		@	@

Sketch (Bundle/Nozzle Orientation)



Tube No. 518U	OD 19.050 mm	Thk(Avg) 1.470 mm	Length 10.000 m	Pitch 23.812 mm	Layout 30
Tube Type	Plain		Material 304 STAINLESS STEEL (18 CR, 8 NI)		
Shell	ID 863.602 mm	OD mm	Shell Cover		
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Plate	None	
Baffles-Cross	Type SUPPORT	%Cut (Diam)	Spacing(c/c) 901.568	Inlet mm	
Baffles-Long		Seal Type			
Supports-Tube		U-Bend	Type		
Bypass Seal Arrangement		Tube-Tubesheet Joint			
Expansion Joint		Type			
Rho-V2-Inlet Nozzle	1169.17 kg/m-s2	Bundle Entrance	Bundle Exit	kg/m-s2	
Gaskets-Shell Side		Tube Side			
-Floating Head					
Code Requirements			TEMA Class		
Weight/Shell 17591.8		Filled with Water 33740.4	Bundle 7221.52	kg	

Remarks: Supports/baffle space = 10.

4.2 Simulation of De-ethanizer unit

The process flow diagram as shown in fig 2.2 for the de- ethanizer system describes the equipment in it. The above calculations shown are for the hydraulic loop of de-ethanizer system and the separator sizing is done for the de- ethanizer separator. De- ethanizer condenser is designed and its rating shows the reduction in overdesign reducing the losses due to fouling. Simulation done with the help of the software i.e. HYSIS gives the equipment details which have been also determined by hand calculation or with the help of the Microsoft excel.

4.3 Conclusion

As Pro DRAW is used for calculation of the hydraulics loop in process and in the utility, the manual efforts also provide the same results for the LNG plant; in fact more accuracy can be achieved in certain aspects. HTRI used for the rating of the condenser resulted in reduced overdesign with a increase in length of the tubes along with reduction in number of tubes per pass. This helps in reducing the fouling and to obtain efficient and economic heat exchanger. Separator sizing did not have any different results for manual calculation.

Thus, report consist a brief method to calculate the plant hydraulics along with equipment sizing and the heat exchanger rating with the process simulation.

4.4 Recommendation

Though there are software available to reduce the work of an employee or an student it is essential to understand the importance and the credentials required to calculate manually for a plant. This report is made by using details given for LNG plant already in existence and it required to refer all those credential documents some of which are here and in reference and some were impossible to disclose.

Through this work done path can be developed towards optimization of an heat exchanger present there in one train and its integration with the heat exchanger in the other train so as to utilize the energy and prevent wastage. This can be possible only if the two exchanger are in feasible condition to make them available foe heat integration without any considerable expenses.

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APPENDIX

Taitel- Dukler's map and Simpson's map