A

DISSERTATION REPORT

ON

"LPG EXTRACTION FROM NATURAL GAS"

PROJECT DISSERTATION SUBMITTED IN PARTIAL

FULFILLMENT

OF THE REQUIREMENT

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ENGINEEERING

By

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DEHRADUN

CERTIFICATE

This is to certify that the project report on "LPG EXTRACTION FROM NATURAL GAS." submitted to UPES, Dehradun by Mr. Shyam Sunder Gupta in partial fulfillment of the requirements for the degree of Applied Petroleum Engineering & academic session (2003-2007) is a bonafide work carried out by him under my supervision & guidance.

Date: 08-05-2007

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Contents

Chapter 3

CALCULATION OF A SIMPLE SYSTEM

Chapter 4

ECONOMIZER SYSTEMS

Chapter 5
A CASE STUDY -.

Chapter 6

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I am the Students of 8th Semester of B.Tech (Applied Petroleum Engineering). University of Petroleum and Energy Studies, Dehradun would like to take this opportunity to thank Mentor **Mr. R. MAHAJAN** for giving us valuable inputs without which this project would not have been such a success.

I am really indebted to the constant motivation provided by my respected **Dean Dr. B.P. Pandey** who has been a torchbearer and a driving force in bringing up this uphill task.

I would specially like to thank **Mr. Kamal Bansal** our project coordinator for allowing us to join the major project program.

I am also indebted to all the lecturers of my class whose constant inspiration and unwaning enthusiasm made me intrepid to face all the crests and troughs of life. The university has also been a great help in making of this project especially the library.

Above all, I am indebted to my parents whose moral guidance and support helped me reach to this great position.

1. INTRODUCTION

1.1 <u>GENERAL DESCRIPTION OF LPG:-</u>

Liquid petroleum gases were discovered in 1912 when Dr. Walter Snelling, an American scientist, realized that these gases could be changed into liquids and stored under moderate pressure. From 1912 and 1920, LP-gas uses were developed. The first LPG cook stove was made in 1912, and the first LPG -fueled car was developed in 1913. The LPG industry began sometime shortly before World War I. At that time, a problem in the natural gas distribution process popped up. Gradually facilities were built to cool and compress natural gas, and to separate the gases that could be turned into liquids (including propane and butane). LPG was sold commercially by 1920.

Liquefied petroleum gas (also called liquefied petroleum gas, liquid petroleum gas, LPG, LP Gas, or autogas) is a mixture of hydrocarbon gases used as a fuel in heating appliances and vehicles, and increasingly replacing chlorofluorocarbons as an aerosol propellant and a refrigerant to reduce damage to the ozone layer. Varieties of LPG bought and sold include mixes that are primarily propane, mixes that are primarily butane, and mixes including both propane and butane, depending on the season—in winter more propane, in summer more butane. Propylene and butylenes are usually also present in small concentration. A powerful odorant, ethanethiol, is added so that leaks can be detected easily. LPG is manufactured during the refining of crude oil, or extracted from oil or gas streams as they emerge from the ground.

At normal temperatures and pressures, LPG will evaporate. Because of this, LPG is supplied in pressurised steel bottles. In order to allow for thermal expansion of the contained liquid, these bottles should not be filled completely; typically, they are filled to between 80% and 85% of their capacity. The ratio between the volumes of the vaporised gas and the liquefied gas varies depending on composition, pressure and temperature, but is typically around 250:1. The pressure at which LPG becomes liquid, called its vapor pressure, likewise varies depending on composition and temperature; for example, it is approximately 220 kilopascals (2.2 bar) for pure butane at 20 °C (68 °F), and approximately 2.2 megapascals (22 bar) for pure propane at 55 °C (131 °F). Propane gas is heavier than air, and thus will flow along floors and tend to settle in low spots, such as basements. This should be kept in mind to avoid accidental ignition or suffocation hazards.

While butane and propane are different chemical compounds, their properties are similar enough to be useful in mixtures. Butane and Propane are both saturated hydrocarbons. They do not react with other. Butane is less volatile and boils at 0.6 deg C. Propane is more volatile and boils at - 42 deg C. Both products are liquids at atmospheric pressure when cooled to temperatures lower than their boiling points. Vaporization is rapid at temperatures above the boiling points. The calorific (heat) values of both are almost equal. Both are thus mixed together to attain the vapor pressure that is required by the end user and depending on the ambient conditions. If the ambient temperature is very low propane is preferred to achieve higher vapor pressure at the given temperature

This is because these gases can be liquefied at normal temperature by application of a moderate pressure increase, or at normal pressure by application of cooling using refrigeration.

LPG comes from two sources. It occurs naturally in oil and gas fields and is separated from the other components during the extraction process from the oil or gas field. LPG is also one of the by-products of the oil refining process.

LP Gas usually consists of a mixture of propane and butane for standard heating and cooking purposes. Propane starts vaporising above -45°C, so it is more versatile for general use. Butane starts vaporising above -2°C and requires a propane/butane mixture in cold environments as it will not vaporise as readily as propane. LP Gas can also be used in specialised applications that require a more rigorous specification. Such applications include food processing, aerosol propellants and automotive fuel (autogas).

When natural gas is drawn from the earth, it is a mixture of several gases and liquids. Methane, which is sold by gas utilities as "natural gas" constitutes about 90 percent of this mixture. Of the remaining 10 percent, 5 percent is propane and 5 percent is other gases such as butane and ethane. Before natural gas can be transported or used, the LP Gases (which are slightly heavier than methane, the major component of natural gas) are separated out. Depending on the" wetness" of a producing gas field, gas liquids generally contain 1%-3% of the unprocessed gas stream. Some LP Gases are also trapped in crude oil. In order to stabilize the crude oil for pipeline or tanker distribution, these "associated" or "natural

gases" are further processed into LP Gas. Worldwide, gas processing is a source of approximately 60% of LP Gas produced.

In crude oil refining the LP Gases are the first products produced on the way to making the heavier fuels such as diesel, jet fuel, fuel oil, and gasoline. Roughly 3% of a typical barrel of crude oil is refined into LP Gas although as much as 40% of a barrel could be converted into LP Gas. Worldwide, crude oil refining is the source for the other roughly 40% of LP Gas supplies although the ratio between gas processing and refining varies among regions.

1.2 OBJECTIVE OF STUDY:-

- The term **NGL** (natural gas liquids) is a general term which applies to liquids recovered from natural gas and as such refers to ethane and heavier products (propane, butane, C5+).
- The recovery of light hydrocarbon liquids from natural gas streams can range from simple dew point control to deep Proane extraction.
- The desired degree of liquid recovery has a profound effect on process selection, complexity and cost of the processing facility

2. METHODS OF LPG RECOVERY FROM NG:-

2.1 INTRODUCTION:-

- The term NGL (natural gas liquids) is a general term which applies to liquids recovered from natural gas and as such refers to ethane and heavier products.
- The recovery of light hydrocarbon liquids from natural gas streams can range from simple dew point control to deep ethane extraction.
- The desired degree of liquid recovery has a profound effect on process selection, complexity and cost of the processing facility.

2.1.1 GAS COMPOSITION:-

- The gas composition has a major impact on the economics of NGL recovery and the process selection.
- In general, gas with a greater quantity of liquefiable hydrocarbons produces a greater quantity of products and hence greater revenues for the gas processing facility.
- Richer gas also entails larger refrigeration duties, larger heat exchange surfaces and higher capital cost for a given recovery efficiency.
- Leaner gases generally require more severe processing conditions (lower temperatures) to achieve high recovery efficiencies.

2.1.2 LIQUID RECOVERY:-

- Gases are typically characterized by the gallons per thousand cubic feet of recoverable hydrocarbons in the gas.

- This is commonly expressed as "GPM."

- GPM was traditionally meant to apply to propane and heavier components but is often used to include ethane.

- The other major consideration in the evaluation of NGL recovery options is the specification of the residue sales gas.

- Sales specifications are usually concerned with a minimum Higher Heating Value (HHV) of the gas, but in some instances the maximum HHV can also be a consideration.

- In general, sales gas specifications set the minimum HHV at 950-1000 BTU/scf.

- Thus, if any components such as nitrogen or CO_2 are present in the gas, sufficient ethane and heavier components must remain in the gas to meet the heating value specification.

- If little nitrogen or CO_2 is present in the gas, the recovery level of the ethane and heavier components is then limited by markets, cost of recovery and gas value.

- Removal of liquids results in gas "shrinkage" and reduction of the HHV.

- This shrinkage represents a loss of revenue for the gas sales which must be considered in the economics of an NGL recovery plant.

2.1.3 STABILIZATION

• One of the problems in using dew point control units of both expansion LTS and mechanical refrigeration systems is the disposition of the liquids removed.

• The liquids must be stabilized by flashing to lower pressure or by the use of a stabilization column.

• When the condensate is flashed to a lower pressure, light hydrocarbons are liberated which may be disposed of in a fuel gas system.

• The stabilization column can produce a higher quality and better controlled product. The condensate stabilizer is usually a top feed column which runs at a reduced pressure from the cold separator and has a reboiler to produce a specified vapor pressure product.

• The column contains either trays or packing to provide necessary mass transfer for stabilization of the liquid feed.

• After stabilization, the product is cooled and sent to storage.

2.1.4 DEW POINT CONTROL

• To prevent the formation of liquids in the system, it is necessary to control the hydrocarbon dew point below the pipeline operating conditions.

• Since the pipeline operating conditions are usually fixed by design and environmental considerations, single-phase flow can only be assured by removal of the heavier hydrocarbons from the gas.

• Dew point calculations for a gas stream leaving a separator at 100°F and 1000 psia.

• These dew point curves show that as the pressure is reduced, liquid is formed.

• The heavier the hydrocarbon, the more the dew point temperature increases as the pressure is lowered.

• The cricondentherm of the dew point curve is primarily determined by the nature of the heaviest component in the gas rather than the total quantity of the heavy component in the feed gas.

Typical Low Pressure Retrograde Condensation Dew Point Curves[3]
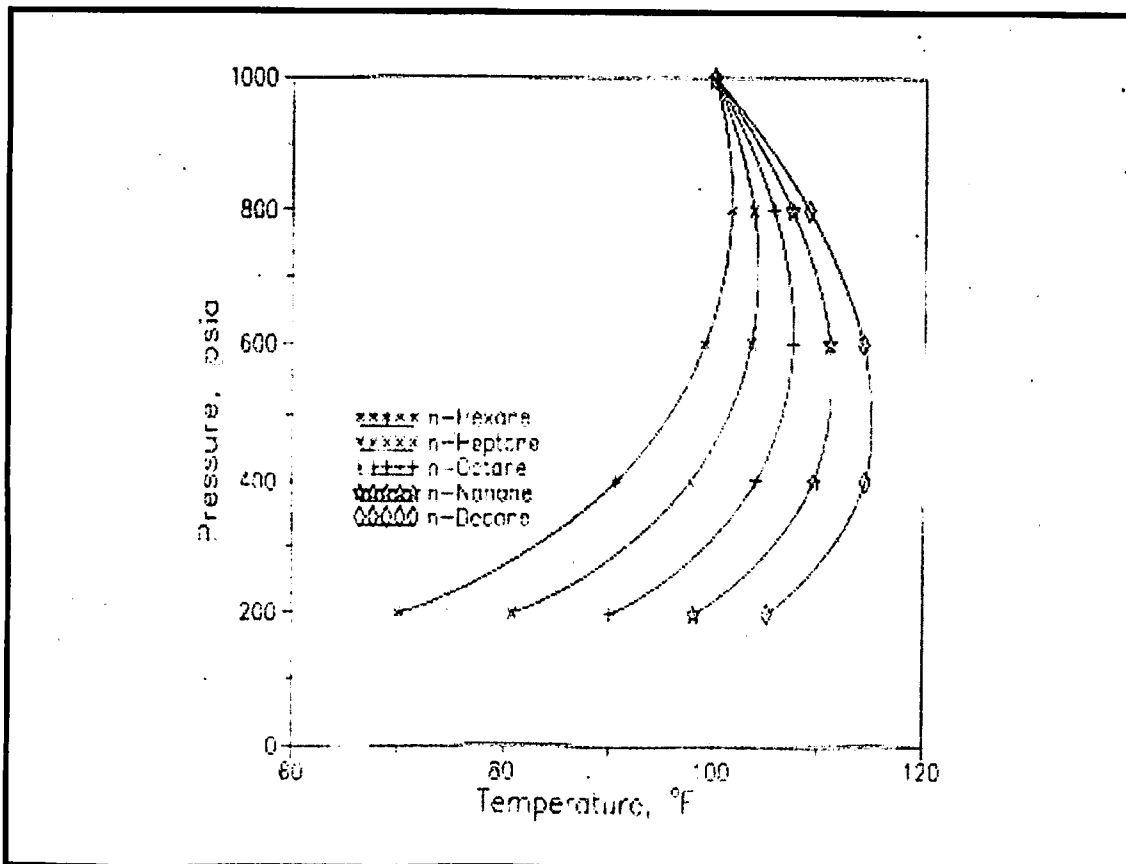
2.2 METHODS OF PROPANE & BUTANE RECOVERY:-

There are three general methods which can be used to achieve the conditions necessary to attain high propane recovery levels:-
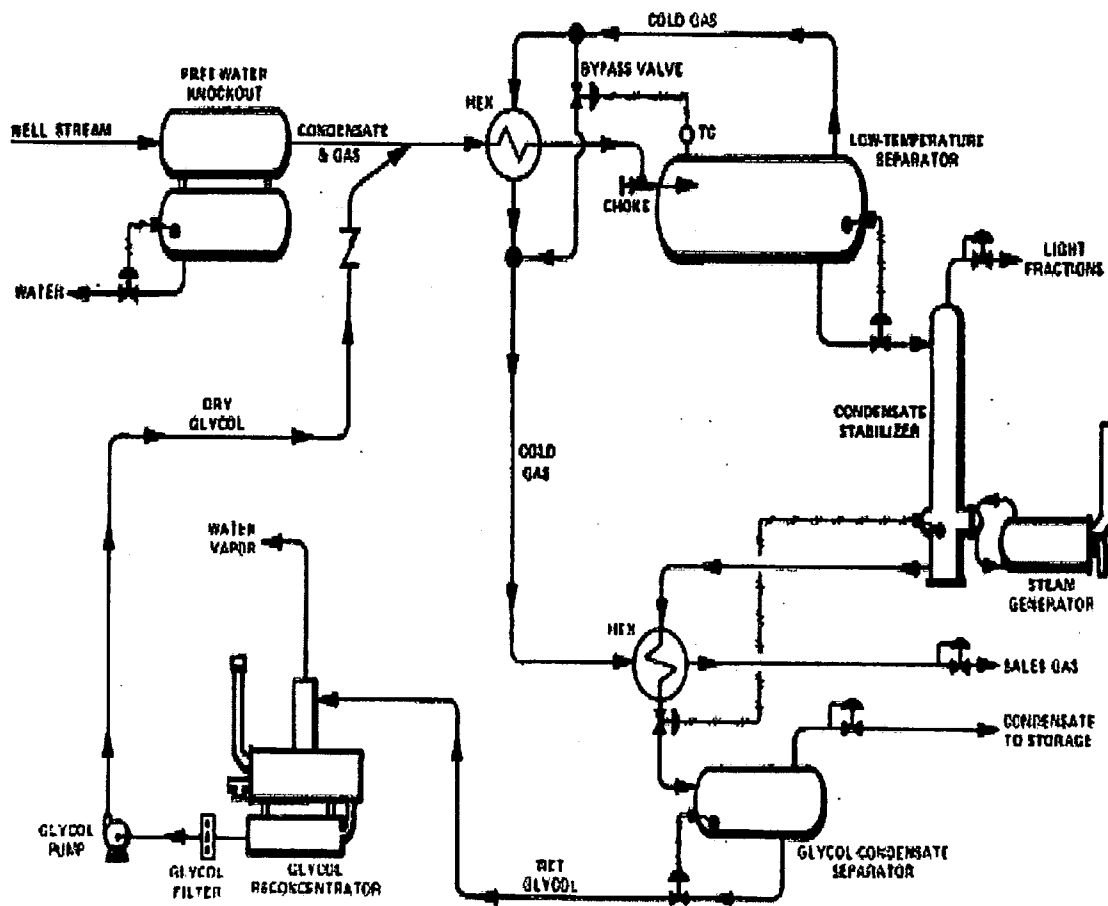
1. LOW TEMPERATURE SEPARATION
2. STRAIGHT REFRIGERATION PROCESS
3. REFRIGERATED LEAN OIL ABSORPTION

2.2.1 LOW TEMPERATURE SEPARATION

- Two methods can be used to reduce the hydrocarbon dew point.

- If sufficient pressure is available, the removal can be accomplished by expansion refrigeration in an LTS (Low Temperature Separation) unit.

- The expansion refrigeration system uses the Joule-Thomson effect to reduce the gas temperature upon expansion.

- This temperature reduction results in not only hydrocarbon liquid condensation but also water condensation.

- The water is generally removed as hydrates in this process, melted and removed.

- Thus, the process can actually accomplish dew point control of both water and hydrocarbon in a single unit.

- The hydrocarbon and water dew points achievable with this process are limited by the pressure differential available as well as the composition of the feed gas.

• It is an attractive process step if sufficient liquid removal can be achieved at the available operating conditions.

• A further modification to this process is to add glycol injection to the high pressure gas to allow the achievement of lower water dew points when available pressure is limited.

• The use of the glycol eliminates the need to heat the LTS liquid phase and helps to ensure that no hydrate formation will block the process equipment upstream of the LTS separator.

LOW-TEMPERATURE SEPARATION SYSTEM WITH
GLYCOL INJECTION AND CONDENSATE STABILIZATION

2.2.2 STRAIGHT REFRIGERATION PROCESS

- Often excess pressure is not available to operate an LTS system.

- An alternative to the expansion refrigeration system is to utilize a mechanical refrigeration system to remove heavy hydrocarbon components and reduce the gas dew point.

- The schematic for a refrigeration dew point control unit is shown in Fig.
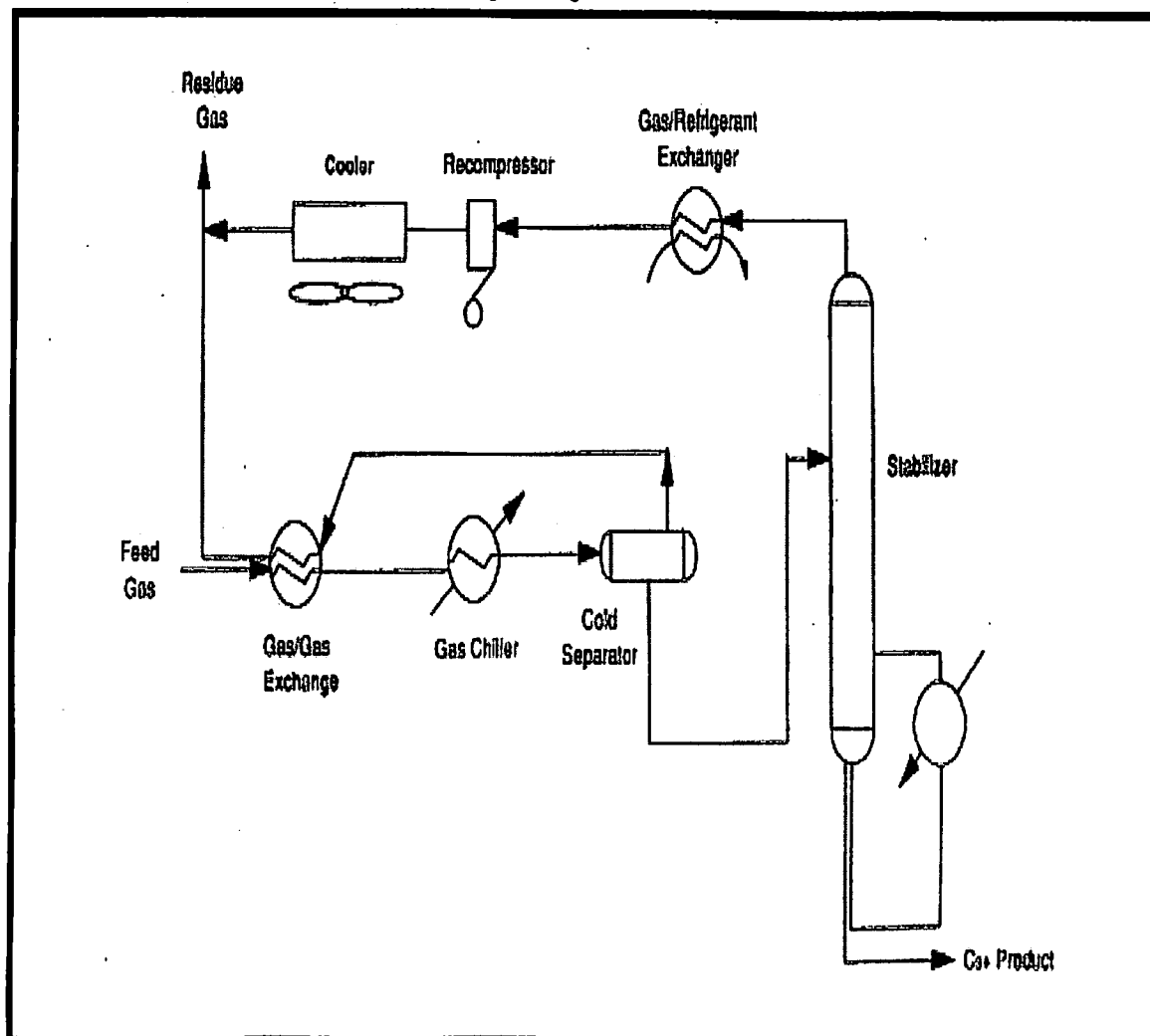
- The gas pressure is generally maintained through the process allowing for equipment pressure drops.

- The gas is heat exchanged and then cooled by the refrigeration chiller to a specified temperature.

- Liquid is separated in the cold separator. The temperature of the separator is set to provide the desired dew point margin for sales gas operations.

- This temperature specification must take into account the gas which is recombined from the liquid stabilization step as well as potential variations in the feed gas pressure.

- Provision must be made in this process for hydrate prevention.

- This can be accomplished by either dehydration upstream of the unit or by integrating the dehydration with the refrigeration unit.

- Use of glycol injection is usually the most cost effective means of controlling water dew points.

- The only drawback is that the refrigeration must be in operation to accomplish the dehydration. If it is desired to operate the dehydration at times independent of the refrigeration, then separate units are used.

- The process can be used for high propane recovery (-40 0F) and in the case of rich gases, for reasonable quantities of ethane recovery.

- The recovery level is a strong function of the feed gas pressure, gas composition and temperature level in the refrigeration chiller.

Recovery Efficiency, Ethane Plus[5]

Process pressure = 600 psig
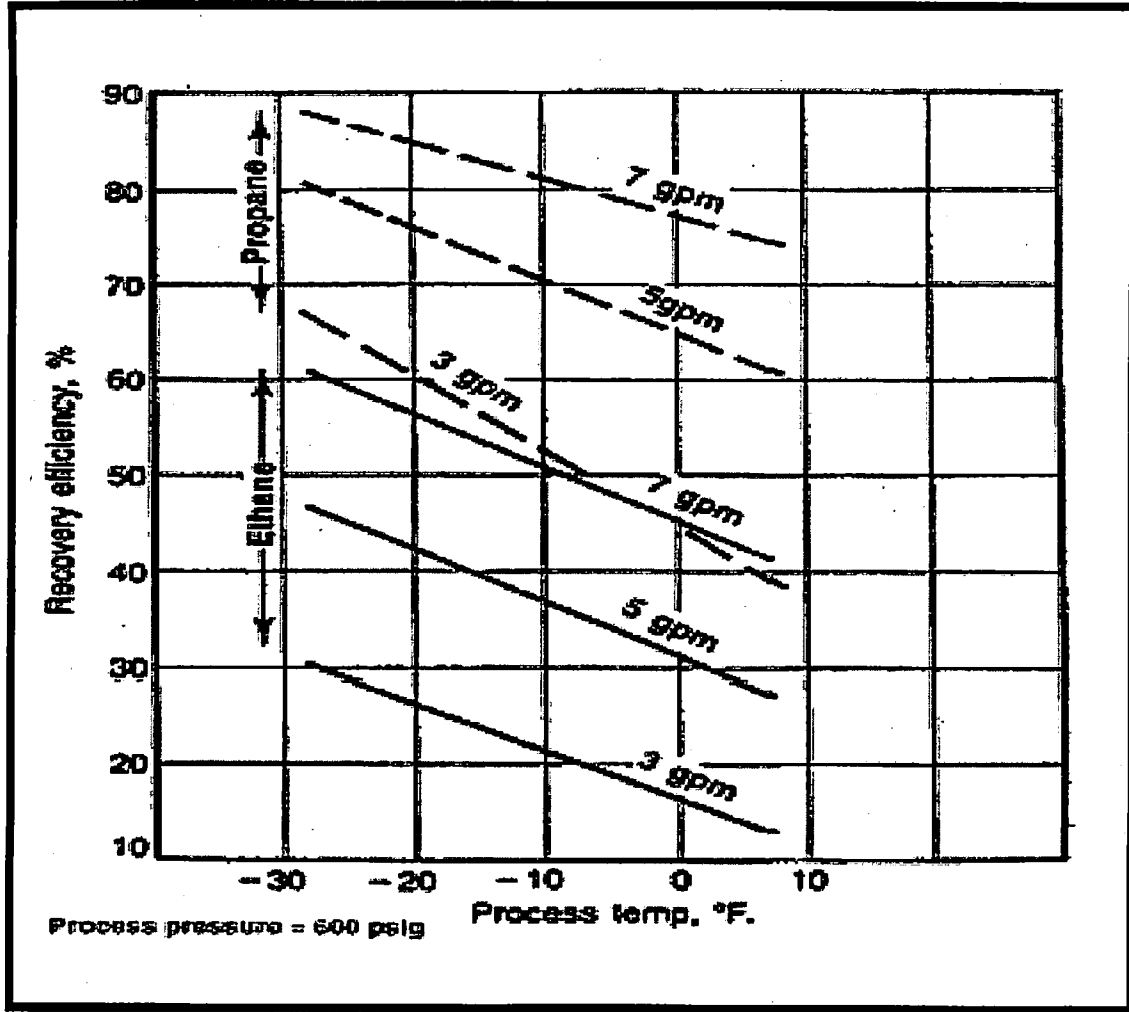
Recovery Efficiency, Propane Plus[5]

Process pressure = 600 psig
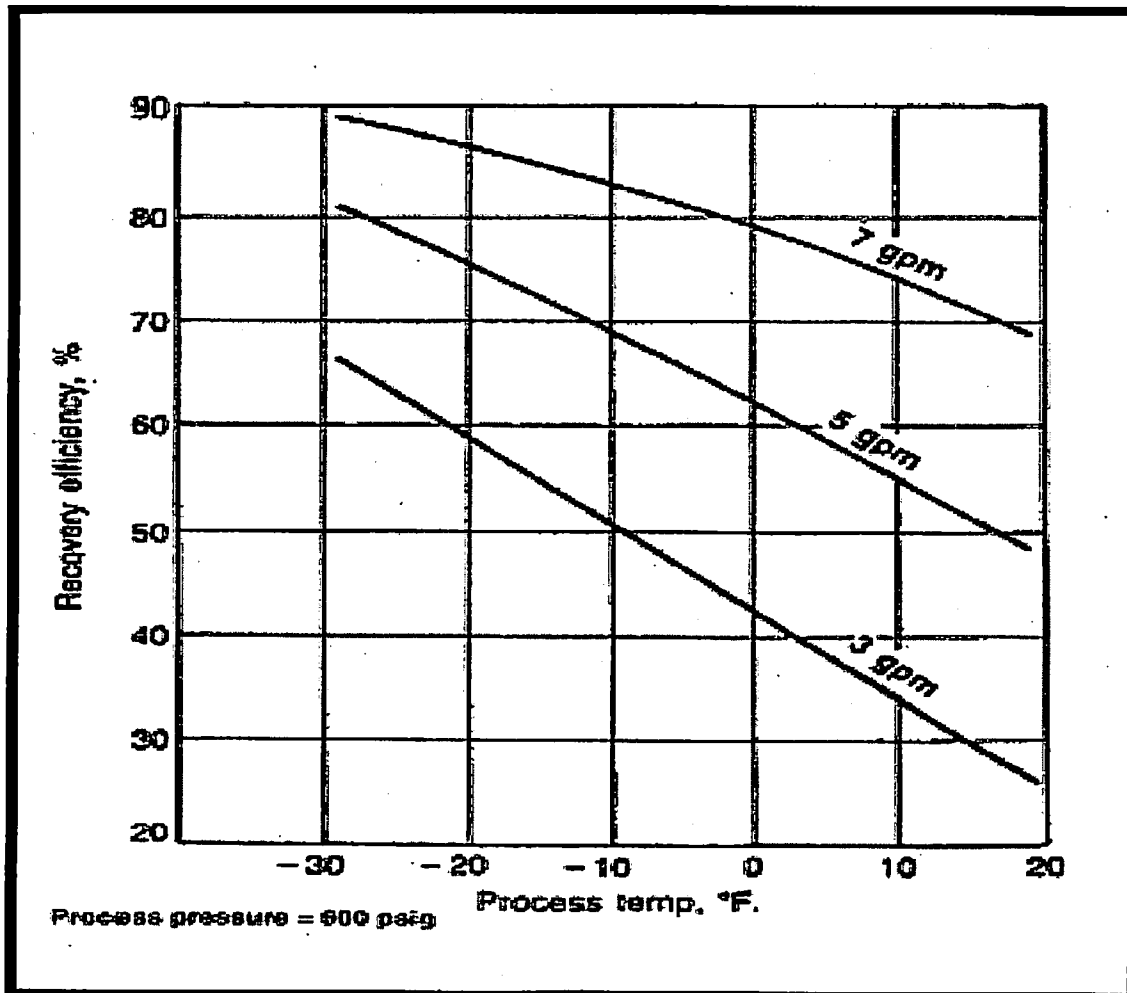
2.2.3 <u>REFRIGERATED LEAN OIL ABSORPTION</u>

•   Lean oil is usually a mixture of paraffinic compounds having a molecular weight between 100 and 200.

•   Raw gas enters the plant inlet separator upstream of the main process where inlet liquids are separated.

•   The gas then enters a series of heat exchangers where cold process gas and the refrigerant reduce the feed gas temperature.

•   This reduction in temperature results in condensation of the heavier hydrocarbons in the inlet gas.
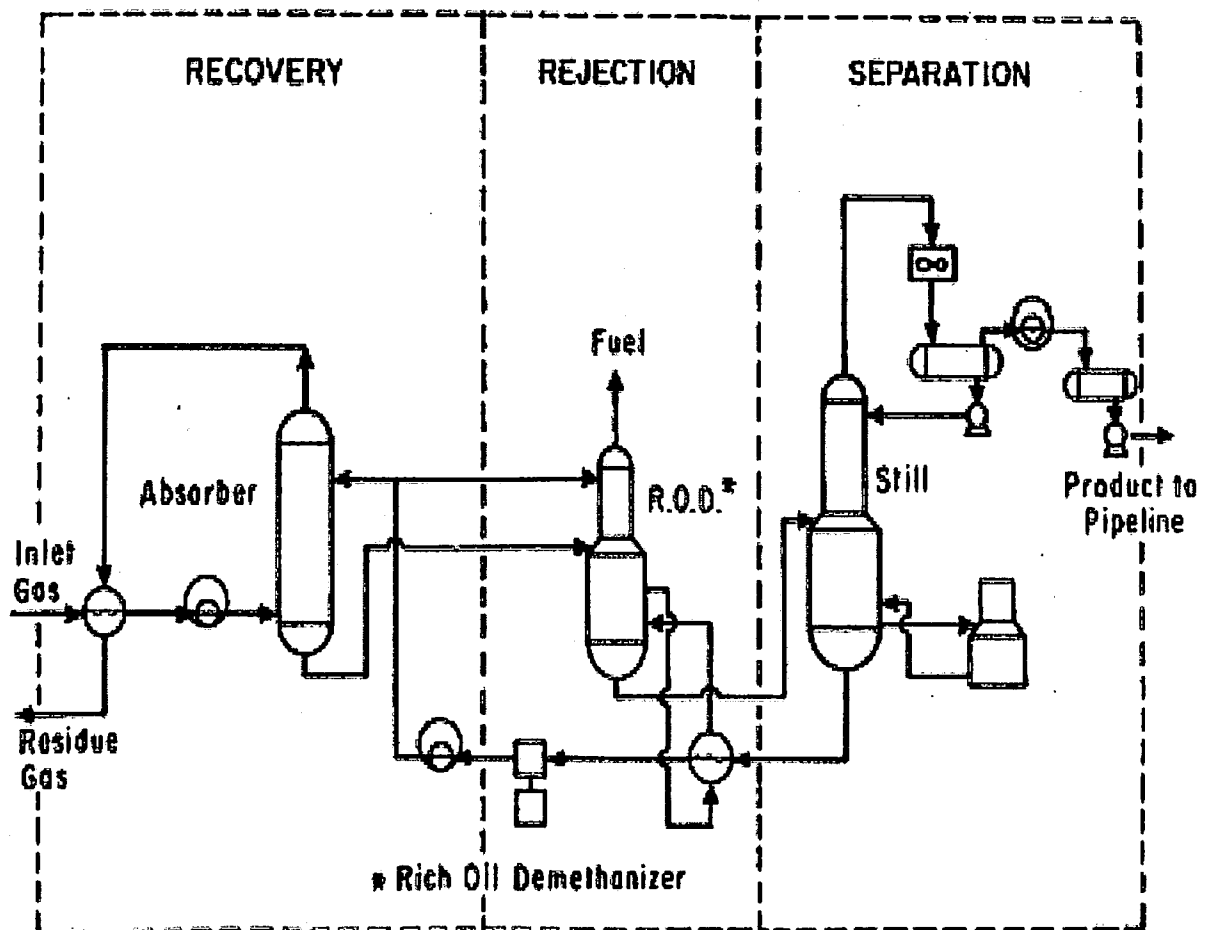
- Countercurrent flow of gas & lean oil in the absorber.

- The lean oil has also been chilled to aid in NGL absorption.

- This column has trays or packing which increase the contact of the gas and lean oil.

- The lean oil physically absorbs the heavier hydrocarbons from the gas.

- The lighter components stay in the gas and leave the top of the absorber.

- The oil and absorbed hydrocarbons leave the bottom of the absorber as "rich oil."

- The rich oil flows to the Rich Oil Demethanizer (ROD) where heat is applied to the rich oil stream to drive out the lighter hydrocarbons which were absorbed.

- Some of the cold lean oil is also fed to the top of the ROD to prevent loss of desirable NGLs from the rich oil.

- The rich oil from the ROD is then fed to a fractionation tower or "still."

- The still is operated at a low pressure and the NGLs are released from the rich oil by the combination of pressure reduction and heat addition in the still.

- The operation of the still is critical to the overall plant operation as this is not only the point where the desired product is produced, but the lean oil quality from the bottom of the column is important in the absorption of NGLs in the absorber.

- The refrigeration required for the oil and gas chilling and the heat inputs to the ROD and still are the key parameters which must be controlled to operate a lean oil plant efficiently.

2.3 ETHANE RECOVERY:-

• Dew Point control and mechanical refrigeration systems are intended for applications where moderate to high propane recoveries are desired.

• In order to achieve higher propane recoveries and ethane recovery, cryogenic temperatures are required.

• Generally, the natural gas processing industry considers cryogenic processing to be processes which operate below –50°F.

• In order to achieve these temperatures, a combination of pressure expansion and chilling is used.

METHODS OF ETHANE RECOVERY:-

There are three general methods which can be used to achieve the conditions necessary to attain high ethane recovery levels.

1. J-T EXPANSION

2. TURBOEXPANDER

3. MECHANICAL REFRIGERATION

Each of these processes has been used successfully, with the turboexpander being the predominant process of choice for ethane recovery facilities.
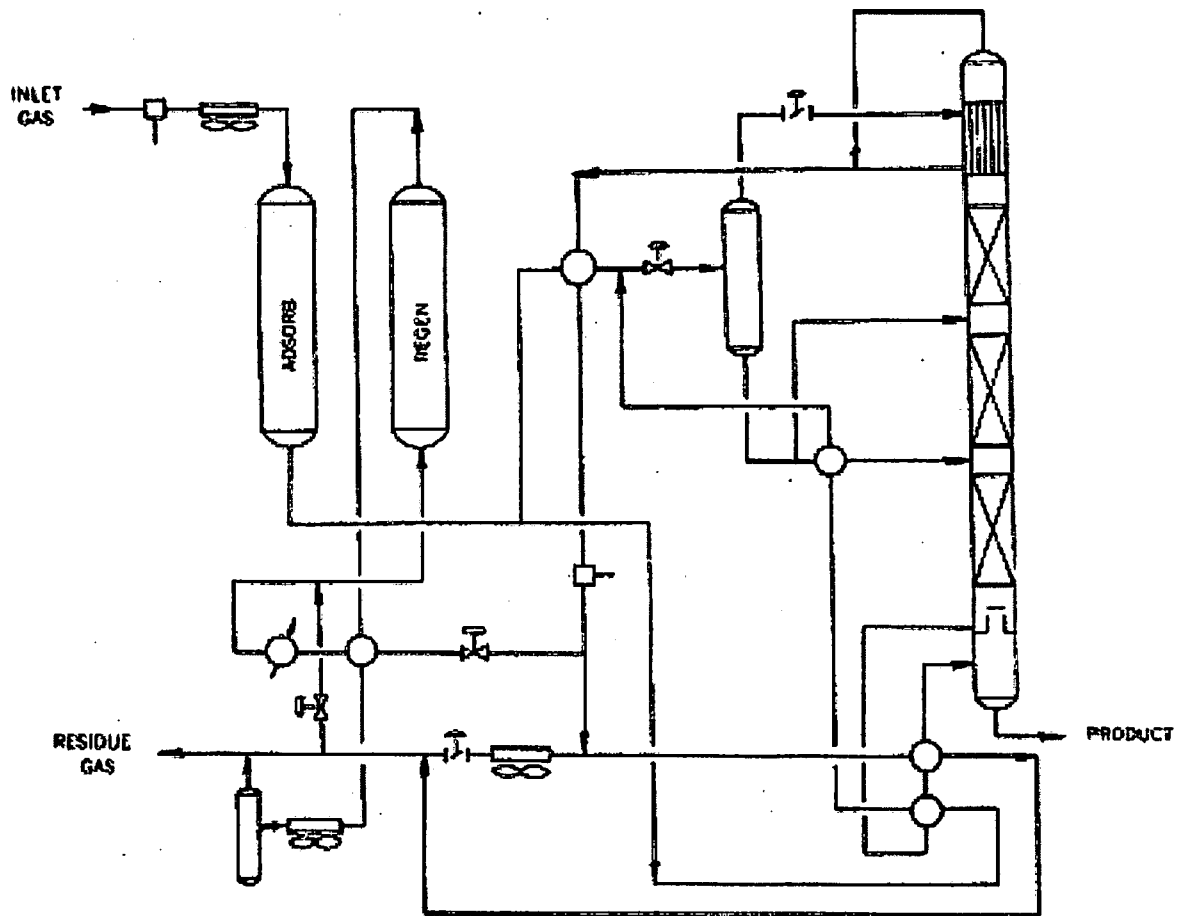
2.3.1 J-T EXPANSION PROCESS:-

• The use of the Joule-Thomson (J-T) effect to recover liquids is an attractive alternative in many applications.

• The general concept is to chill the gas by expanding the gas across a J-T valve. With appropriate heat exchange and large pressure differential across the J-T valve, cryogenic temperatures can be achieved resulting in high extraction efficiencies.

• The main difference between the J-T design and turbo expanders is that the gas expansion is adiabatic across the valve and is nearly isentropic path. Thus the J-T design tends to be less efficient per unit of energy expended than the turbo expander.

• The J-T process does offer some advantages over the turbo expander and refrigeration processes in the following situations:

1. Low gas rates and modest ethane recovery.
2. The process can be designed with no rotating equipment.
3. Broad range of flows.
4. Simplicity of design and operation.

- In order to effectively use the J-T process, the gas must be at a high inlet pressure (over 7000 kPa).

- If the gas pressure is too low, inlet compression is necessary or insufficient expansion chilling will be attained.

- The gas must first be dried to ensure that no water enters the cold portion of the process.

- Molecular sieves or Alumina are used for the drying.

- Methanol injection has been used in a few plants successfully but can be an operating problem.

2.3.2 REFRIGERATED J-T PROCESS

• In some cases the feed gas is not at high enough pressure or the gas is rich in liquefiable hydrocarbons. Then mechanical refrigeration can be added to the J-T process to enhance recovery efficiencies.

• The J-T process with refrigeration added to aid in chilling the feed gas.

• The gas in this design is expanded downstream of the cold separator.
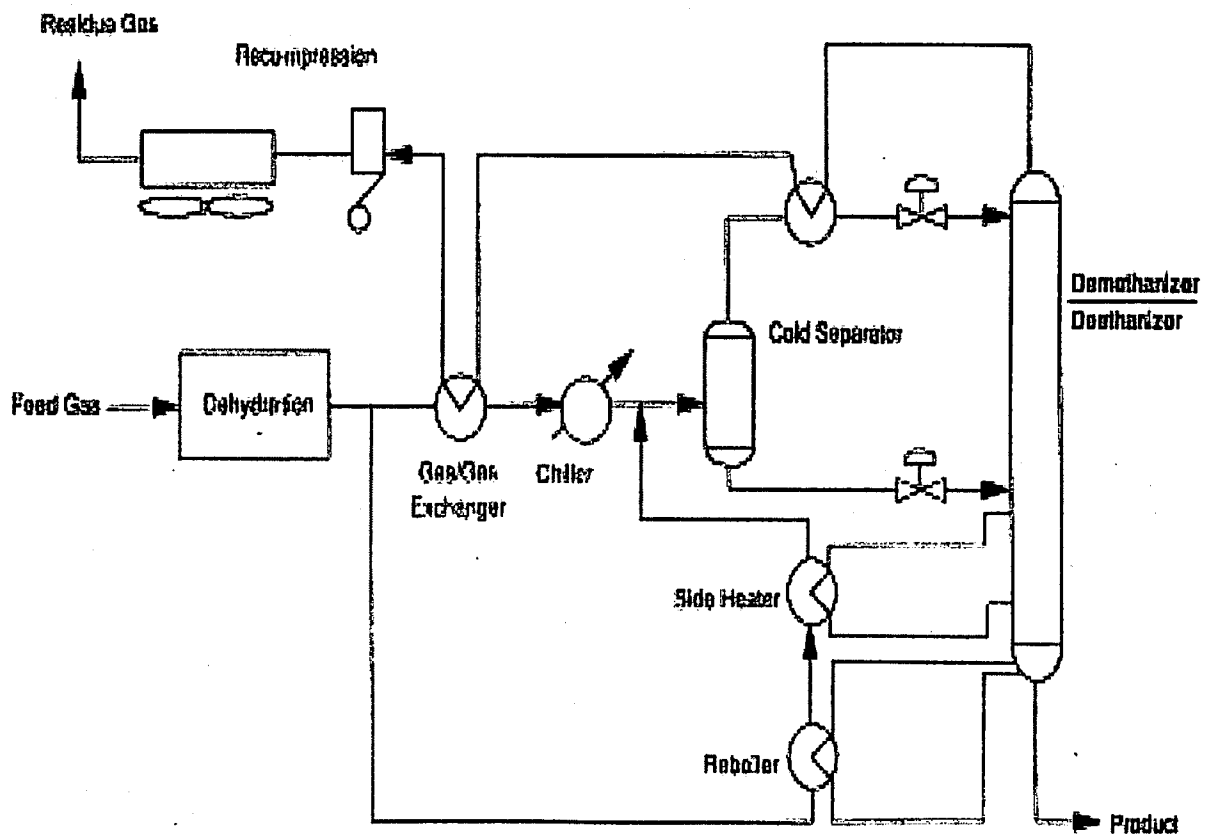
- The location of the J-T valve is dependent on the gas pressure and composition involved.

- The advantage of refrigeration is that lower feed pressure can be used or alternatively, the demethanizer can be operated at a higher pressure thus reducing residue compression.

- The J-T process, whether refrigerated or non-refrigerated, offers a simple, flexible process for moderate ethane recovery.

- It is usually applied to smaller gas flows where some inefficiency can be tolerated for reduction in capital and operating costs.

2.3.3 TURBOEXPANDER PROCESS

- The turboexpander process dominates ethane recovery facility design.

- This process uses the feed gas pressure to produce needed refrigeration by expansion across a turbine (turboexpander). The turboexpander recovers useful work from this gas expansion.

- The expander is linked to a centrifugal compressor to recompress the residue gas from the process. Because the expansion is near isentropic, the turboexpander lowers the gas temperature significantly more than expansion across a J-T valve.
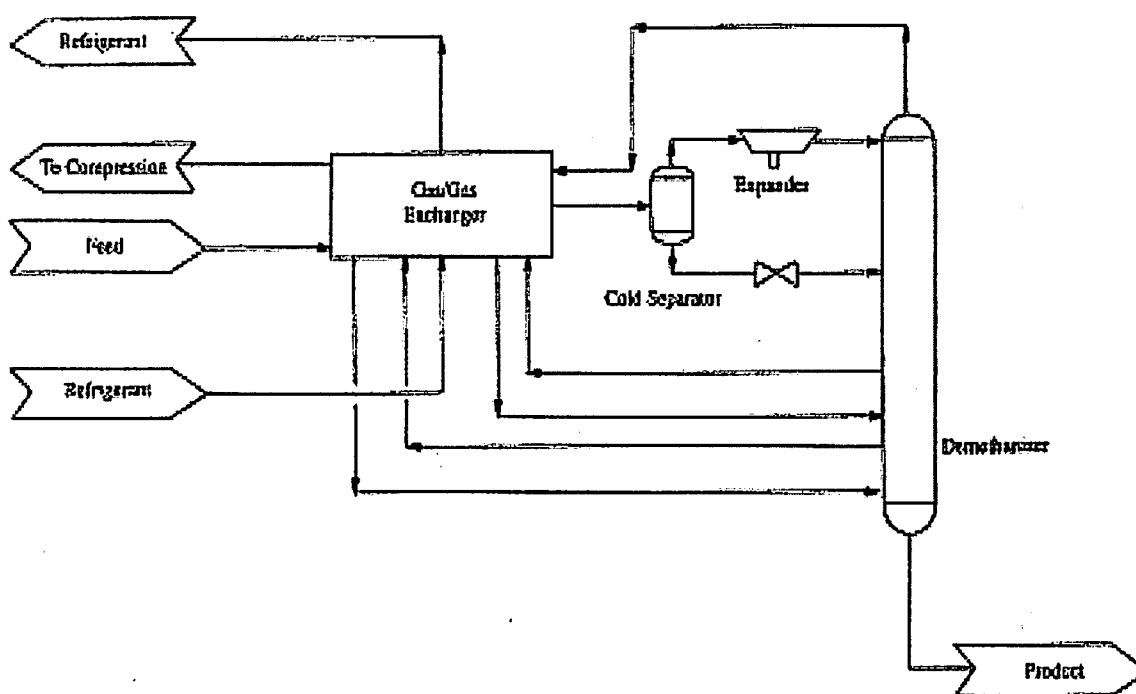
- Dry feed gas is first cooled against the residue gas and used for side heating of the demethanizer.

- Additionally, with richer gas feeds, mechanical refrigeration is often needed to supplement the gas chilling.

- The chilled gas is sent to the cold separator where the condensed liquid is separated, flashed and fed to the middle part of the demethanizer.

- The vapor flows through the turboexpander and feeds the top of the column.

- A J-T valve is installed in parallel with the expander. This valve can be used to handle excess gas flow beyond the design of the expander or can be used for the full flow if the expander is out of service.

- In this configuration the ethane recovery is limited to about 80% or less.

- Also, the cold separator is operated at a low temperature to maximize recovery. Often the high pressure and low temperature conditions are near the critical point of the gas making the operation unstable.

- Another problem with this design is the presence of CO_2, which can solidify at operating temperatures.

- To increase the ethane recovery beyond the 80% achievable with the conventional design, two methods are developed:

Residue Recycle

- In this method, a portion of the residue gas, after recompression is recycled back to the top of the column.

Gas Subcooled Process

A portion of the gas from the cold separator is sent to a heat exchanger where it is totally condensed with the overhead stream. This stream is then flashed to top of the demethanizer providing reflux to the demethanizer.

2.4 FRACTIONATION CONSIDERATIONS

• In all NGL recovery processes, one of the final steps in the plant is the production of the desired liquid product by use of a fractionation column.

• This column produces the specification product as a bottom product with the overhead stream being recycled to the process or sent out of the plant as residue gas product.

• This mixed product then needs to be separated into usable products in a series of one or more fractionation columns.

• The number and arrangement of these columns is dependent on the desired product slate.

• If the NGL stream is an ethane plus stream the first step is to separate the ethane from the propane and heavier components in a deethanizer.

• The propane is then separated from the butane and heavier components in a depropanizer.

• If further processing is desired the butane may be separated in a debutanizer and the butanes further separated in a butane splitter column.

• The butane splitter is only used when a differential value can be realized for the isobutane versus the mixed butane stream.
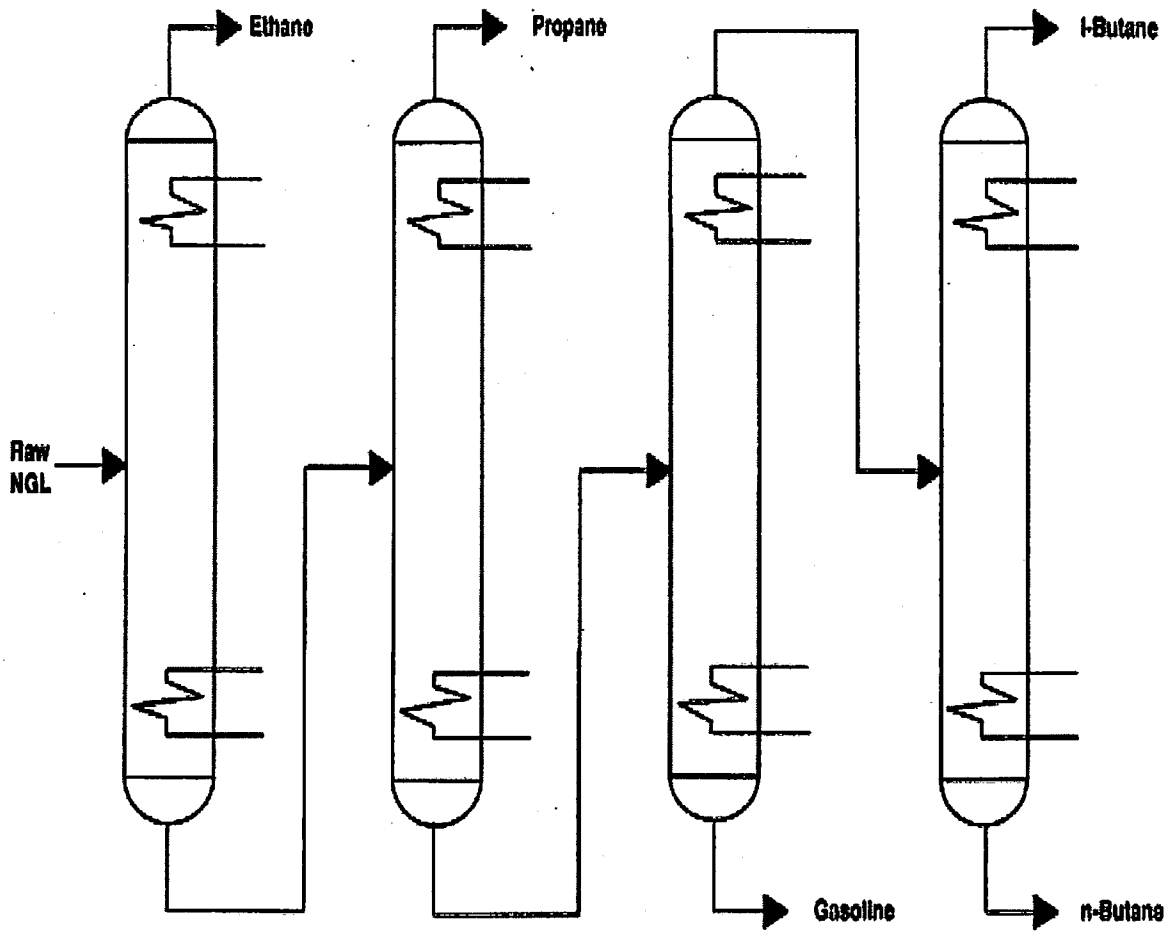
A refrigeration system lowers the temperature of a fluid below that possible when using air or water at ambient conditions. A typical building air conditioner cools air to a temperature of 10-15°c.

At the other end of the scale is the liquefaction of helium at - 268°C. The temperature produced depends on the process objective. If the objective is to recover marketable liquids (NGL's) from a produced gas stream, basic economics controls the temperature specified. If it is to meet a hydrocarbon dewpoint that specification and the processing pressure sets the required temperature.

Several basic processes will be discussed herein.
1. Mechanical Refrigeration
    a. Compression.
    b. Absorption
2. Valve expansion
3. Turbine Expansion

3. CALCULATION OF A SIMPLE SYSTEM

3. 1 MECHANICAL REFRIGERATION :-

A refrigeration system is a heat pump. Low temperature heat is removed from the process fluid and is "pumped "to high temperature (ambient) where it is rejected to. the environment . Energy is required to pump heat. The amount of energy depends on the quantity of heat to be pumped (chiller duty) and how far the heat has to be pumped. (temperature difference between the chiller and the condenser. Energy used to drive a refrigeration process can be in the form of heat or work. The absorption refrigeration system use heat to pump heat.

3.1.1 COMPRESSION REFRIGERATION:-

Compression refrigeration is by far and away the most common mechanical refrigeration process. It has a wide range of applications in the gas processing industry.

- Chilling natural gas for NGL extraction.
- Chilling natural gas for hydrocarbon' dewpoint control.
- LPG product storage.
- Condensation of reflux in deethanizers/demethanizers.
- Natural gas liquefaction '(LNG)

Figure 1 shows a simple single-stage compression refrigeration system Saturated liquid refrigerant at Point A expands across a valve (isenthalpically). On expansion some vaporization occurs. The mixture of refrigerant vapor and liquid enter the chiller typically 3-6°C.lower than temperature to which the process steam is to be cooled. The liquid vaporizes. Leaving at point at C is a saturated vapor refrigerant at Point P

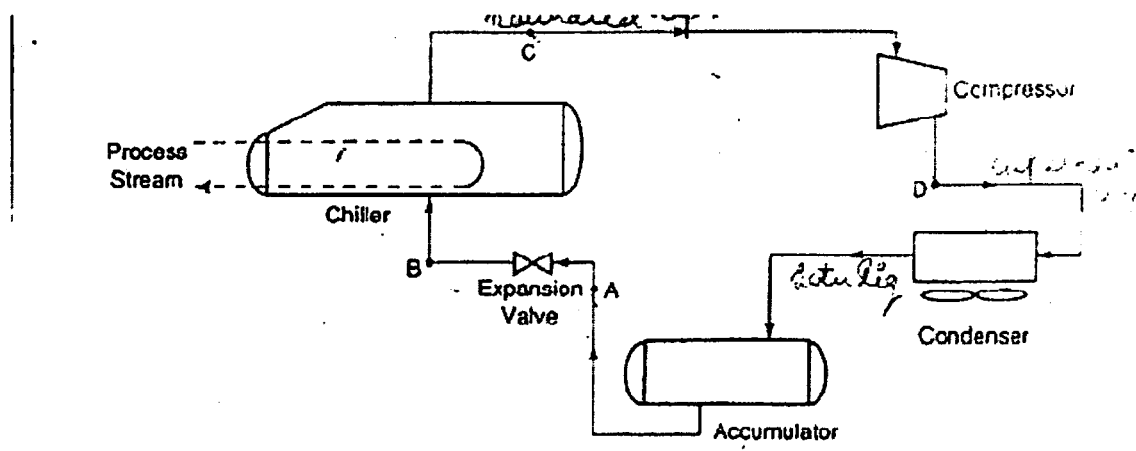and T of the chillier. This vapour is compressed and then enters the condenser as a superheated vapour.

Figure 1 - Flow Sheet of a Simple Refrigeration System

The refrigerant leaves the condenser as a saturated liquid or slightly sub cooled. For air cooling, the condenser temperature will usual be [25-30°F] above the air dry bulb temperature For water cooling the condensing temperature will be 5-IO°C [9-,18 oF] above the water temperature. The accumulator, sometimes called a surge tank or receiver, merely serves as a reservoir for refrigerant as levels vary in the chiller(s) and condenser.

3.2 CALCULATION OF A SIMPLE SYSTEM:-

There are several discrete steps in the sizing of the system shown in Figure. These are summarized below:

3.2. 1. DETERMINATION OF REFRIGERANT CIRCULATION RATE:-

Develop an energy balance around the chiller and expansion valve shown at below. At point A the refrigerant is a saturated liquid (or very close to it). At point C it is a saturated vapor. Qchiller is the chiller duty and is set by the process requirements.
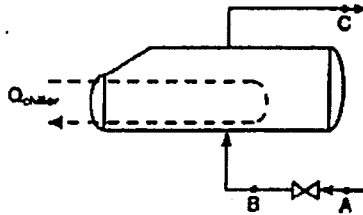
If one writes an energy balance around the system, Qchiller+ mAhA = mchc. But mA = me = m, so

$$m = \frac{Q_{chiller}}{h_C - h_A}$$

Qchiller = chiller duty

he = saturated vapor enthalpy

hA = saturated liquid enthalpy

m = refrigerant circulation rate

3.2.2. DETERMINATION OF COMPRESSOR POWER:-

Calculate theoretical (isentropic) work and use an efficiency to find actual work. The circulation rate from Step (I) is used

$$-W = \frac{m(h_D^{isen} - h_C)}{E}$$
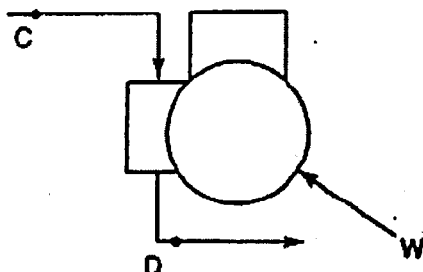
3.3.3. DETERMINATION OF CONDENSER HEAT LOAD (QCOND.):-

There are two ways to do this Knowing Qchiller and W, you can write the overall balance to find Qcond' (Remember, W is negative, so Qcond is the j sum of two negative numbers)

$$Qcond = W - Qchiller$$

If you are performing the calculation manually and wish an independent check of the previous work, write the balance shown below.

$$Qcond = m (hA - hD)$$

3.3 EXPANSION REFRIGERATION:-

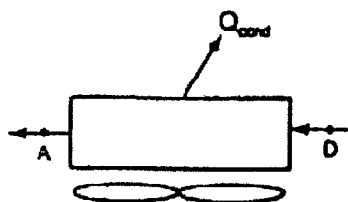Turbine expansion (turbo expanders) refrigeration processes are used in gas processing.The turboexpander was first used in gas processing in the late 196Os. It is now arguably the most popular gas processing method. Turboexpanders can be used for very low temperature NGL extraction as well as for hydrocarbon dewpoint control. Turboexpander are very efficient, reliable and relatively inexpensive.

3.3.1 TURBOEXPANDER:-

Turboexpanders are single-stage radial-inflow turbines. A cutaway of a turboexpander/compressor assembly is shown in Figure 2 Gas enters the expander impeller radially and leaves axially.
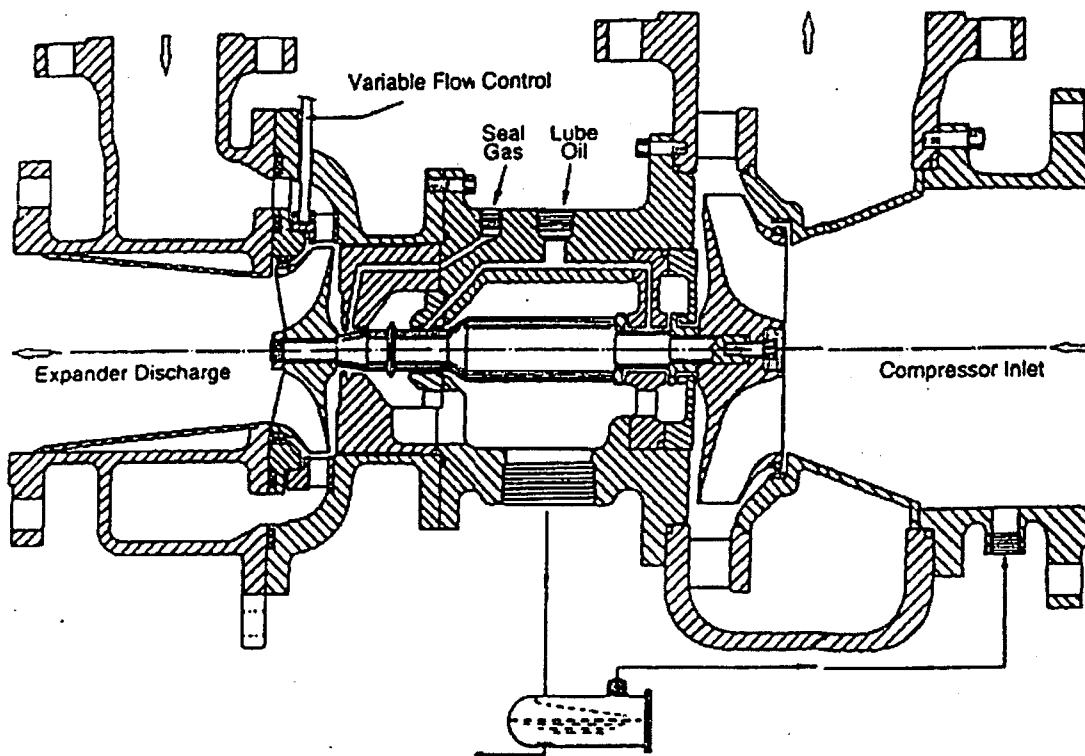
Figure 2 - Cross-Section of an Expander-Compressor

Work generated at the expander wheel is absorbed by a single stage centrifugal compressor in most installations. This is because compression is almost always required downstream of the expander to restore the gas to sales gas pressure. In those cases where compression is not required, the expander can drive a generator to produce electricity.

Expander efficiencies are quite high, approaching 85% (isentropic) for new installations operating at design conditions} Since their introduction in the gas processing industry in the late 1960 s several thousand expanders have been installed and have generally proven reliable and robust.

Figure 3 - Expander Inlet Guide Vanes

Bearings (2-radial and I-thrust) have historically been oil lubricated. The first magnetic bearings were installed m turboexpander In 1991. Several expanders currently use magnetic bearings and this technology is likely to become standard practice in the future.

Seal gas is required to provide a barrier between the wet, cold process gas and the lube oil or magnetic bearings inside the rotor case. Adequate seal gas flow is necessary to prevent process gas from mixing with the lubricating oil and lubricating oil leaking into the process stream. Early seals were labyrinth as shown in Figure 2. Dry gas seals were first used in turboexpanders in 1989 and are commonly used today. They differ somewhat from the design used on centrifugal compressors in that a single seal (rather than tandem seals) is used due to space limitations. Regardless of the seal type, the seal gas must be clean and dry. Several turboexpander failures have been attributed to poor seal gas quality.
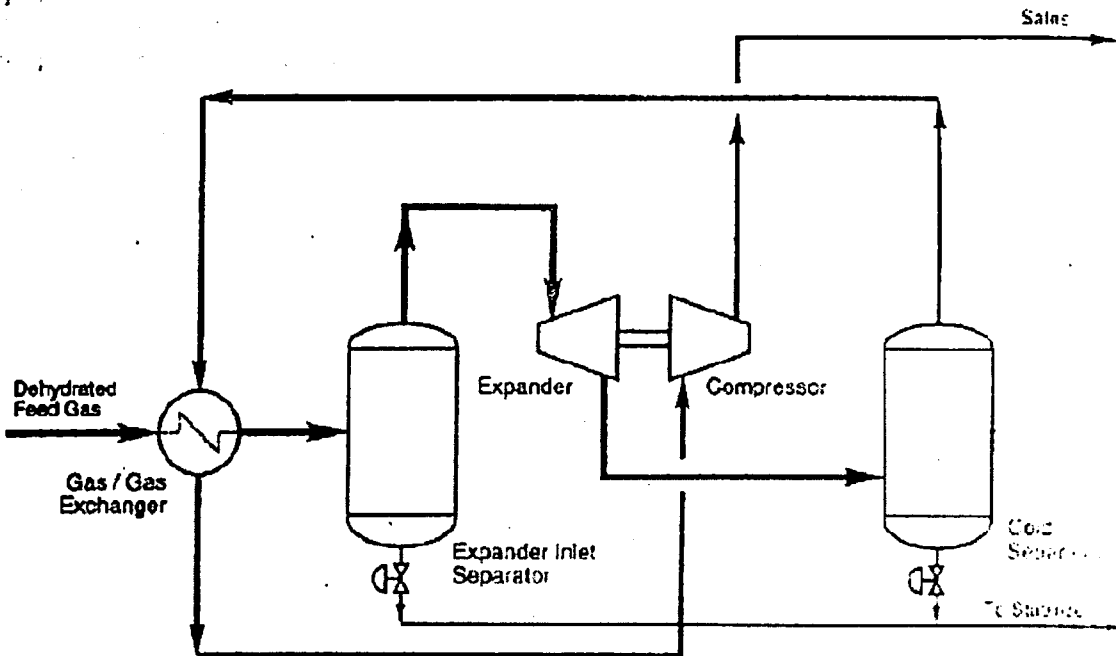
Figure 4 - Expander Process for Hydrocarbon Dewpoint Control

The flow rate through the expander is controlled by inlet guide vanes. These are arranged radially around the expander impeller and their movement is synchronized by rotation of a guide vane ring which is attached to an actuator. The guide vane position, hence expander flow rate, is usually manipulated by a flow or pressure controller figure shows expander inlet guide vanes.Turboexpanders are applied both in deep NGL Extraction (C2+, C3+) plants and hydrocarbon dewpoint control. The primary difference in these applications is the expansion ratio. Figure 3 Expander Inlet Guide Vanes.

For deep NGL extraction, expansion, ratios are typically 3.0-3.5 and for dewpoint control expansion ratios are usually about 1.3-1.5. An example hydrocarbon dewpoint plant using a turboexpander is shown in Figure 4. A simple deep recovery NGL process is shown in Figure 5. Modern deep

recovery processes are more complex than Figure 5, employing significantly more heat integration as well as cooling techniques at the top of the deethanizers/demethanizers to increase NGL recovery. Some of these process schemes will be discussed in more detail in a later section.
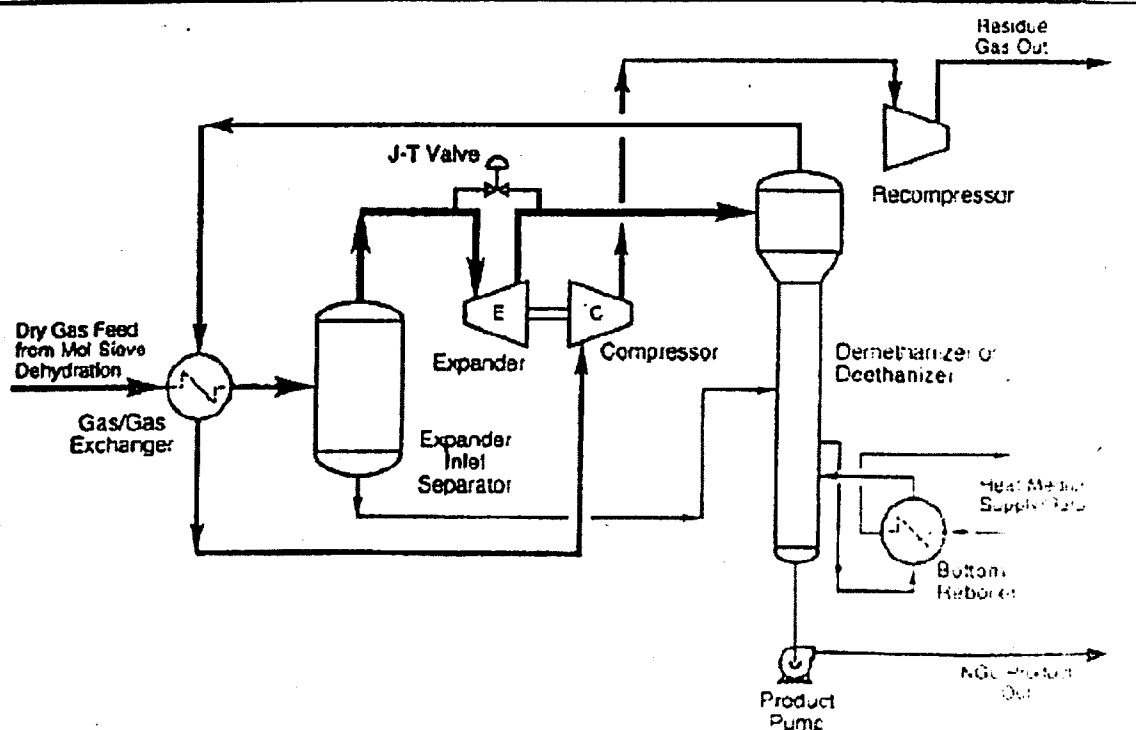
Figure 5 - Simple Expander Process for Deep NGL Extraction

3.3.2 TURBO EXPANDER CALCULATIONS

The theoretical thermodynamic path through an expander is isentropic. The theoretical work is calculated from an isentropic path and corrected to the actual work by use of expander efficiency.

For multicomponent streams and condensation across the expander, the calculation is iterative and tedious, and is almost always done on a computer.

The calculations proceed as follows:

The known variables will typically be feed composition,

T1, P1 and P2'
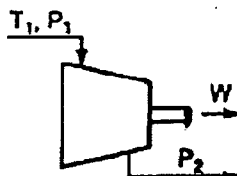
I. from P1 and T1 calculate h1 and s1.

2. Assume a value of T2isen'

3. Run a flash calculation at the assumed T2 and known P2 to establish the phase amounts and compositions.

4. Calculate h2 and S2' (If the outlet is two-phase, these will be total stream values as discussed) .

5, If S2 from (4) equals s1 you have assumed the right temperature. If not, repeat Step, 2-4 untils2=s1

6. When S2= s1 calculated hj = h2 – h1,

7. Calculate expander power, W = m.hisen.Eisen.

8 Calculate h2"act., = h1 + (h2isen- h1)Eisen'

9. Assume a value of T2act. (this will be warmer than T2isen)'

10. Run a flash calculation at the assumed T2act and P2.

11. Calculate h2 (h2"cal.) from the results of the flash.

12. If h2 from (11) equals h2 1you have assumed the correct temperature.

4. ECONOMIZER SYSTEMS:-

4.1 FLASH TANK ECONOMIZER:-

In the simple refrigeration system, a significant amount of vaporization occurs across the expansion valve unlike the multi component phase envelopes discussed.

Therefore, it is relatively easy to calculate the quality or a two-phase mixture.

$$x = \text{liquid fraction} = \frac{h_V - h_F}{h_V - h_L}$$

$$y = \text{vapor fraction} = \frac{h_F - h_L}{h_V - h_L}$$

hF = two-phase enthalpy

hL = saturated liquid enthalpy

hy = saturated vapor enthalpy'

For Example:- hF = hA=hB and hy = he.

The quantity of vapour entering the chiller is 46%.' This vapor provides no refrigeration effect in the chiller, but must be compressed from chiller pressure to condenser pressure increasing the compressor power requirement.

In refrigeration systems the word "economizer" refers generally to any device or process modification which decreases the compressor power requirement for a given chiller duty. Two types of economizers are commonly used - flash tank economizers and heat exchanger economizers)

Figure 6 shows a refrigeration system employing one flash tank economizer. In this system, the saturated liquid refrigerant leaving the accumulator is expanded across a valve to an intermediate pressure where vapor and liquid are separated. The separator liquid is expanded across the. Second valve to chiller pressure while the separator vapor goes to the compressor interstage. The refrigerant entering the chiller now has a higher liquid content. This reduces the refrigerant circulation rate through the chiller as well as the first stage compressor power requirement.
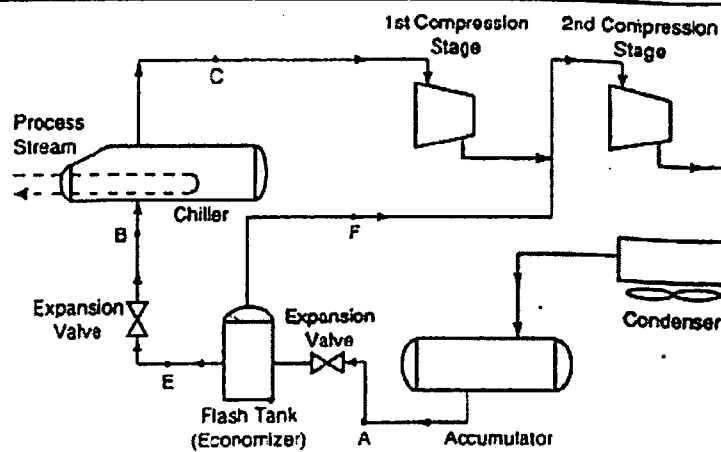
Figure 6 - Flow Sheet of a Refrigeration System with an Economizer

The P-H diagram for the economizer system in Figure 6 is shown .Point E is usually fixed at pressure that results in equal' compression ratios in each stage of compression. Expansion across the economizer. expansion valve is isenthalpic; The vapor formed goes to the second stage of compression as a saturated vapor. The saturated liquid, E, leaving the separator is expanded isenthalpically to pressure B. Notice that the h available from B to C for this system is larger than for the simple system.
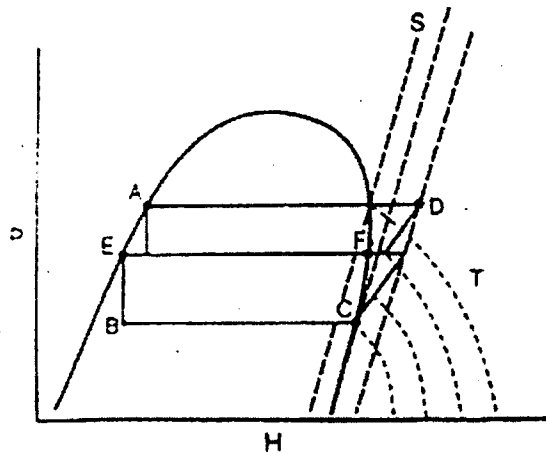
4.2 HEAT EXCHANGER ECONOMIZER:-

A second type of economizer configuration is the heat exchanger economizer shown, in fig. 7 Cold, low-pressure chiller vapor is used to subcool the saturated liquid refrigerant. This decreases the refrigerant circulation rate, and may reduce compressor power.

With regard to the compressor power, two factors offset the reduced circulation rate. The first is exchanger pressure drop. The pressure drop on the low pressure side of the exchanger may be 20-50 kPa [3-7 psi]. This can significantly increase compressor power, particularly when the chiller pressure is near atmospheric. Secondly, the refrigerant vapor entering the compressor is now superheated. Although this reduces the 'likelihood of liquid carryover into the compressor, it results in higher power consumption per unit mass due to the higher suction temperature. Note the constant entropy lines are less steep at higher temperatures.
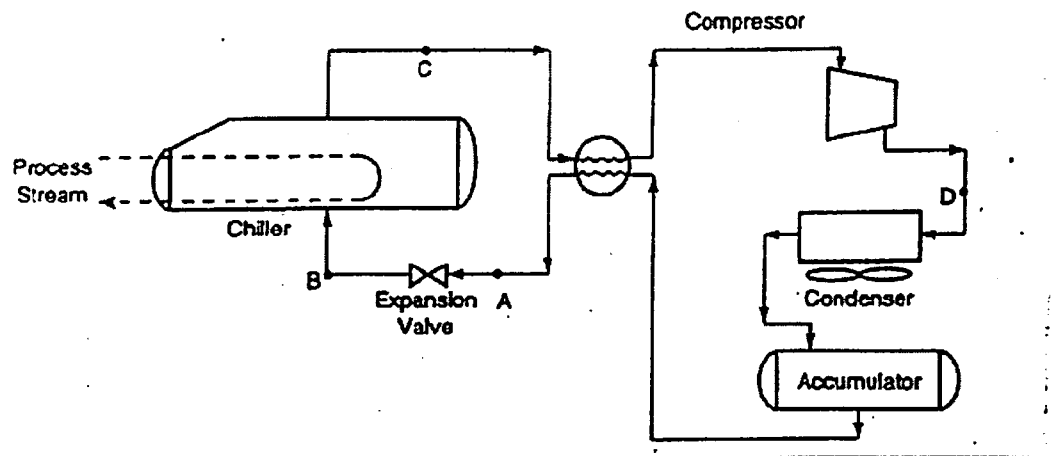
Figure 7 - Flow, Sheet of a Simple Refrigeration System with a Heat Exchanger Economizer'

5. CASE STUDY:-

5.1. INTRODUCTION

Natural gas primarily contains hydrocarbons with some proportion of Carbon Dioxide and very small percentage of other gases. LPG and NGL are value added products. LPG and NGL are recovered from natural gas in the liquid state. Recovery of these products in the liquid form has many distinct advantages over Natural Gas. Typical compositions of dry natural gas and lean gas are given below:

Components	Natural Gas (V/V%)	Components	Lean Gas (V/V%)
Methane	82.34	Methane	88.54
Ethane	6.45	Ethane	4.96
Propane	4.20	Propane	1.34
Iso-Butane	0.85	Iso-butane	0.07
n- Butane	0.95	n-butane	0.05
Iso- Pentane	0.22	Carbon Dioxide	5.04
n- Pentane	0.26		
Hexane +	0.23		
Carbon Dioxide	4.50		

Table: 1 Natural Gas Composition   Table:2 Lean Gas Composition

Recovery of value added products like LPG and NGL is carried out at various Gas Processing Plants. Operation objectives of these plants mainly concentrate on following issues:

- Maximization of recovery of propane and butane (i.e.LPG components)

- Minimization of butane slippage in NGL.

- Minimization of propane loss in Light Ends Fractionation Column top product

- Energy Conservation

- Minimization of Hydrocarbon loss due to flaring

- Quality, Safety, Environment and Occupational Health

To recover LPG and NGL from natural gas, all heavier components are converted to liquid form and then desired components are separated through distillation. It becomes necessary to convert propane and heavier components to liquid form. But in the process, some quantity of ethane remains present in the liquid and at the same time, some quantity of propane also remains in the vapor phase. Through optimization, maximum quantity of propane can be retained in the liquid form whereas loss of propane in to natural gas can be restricted to minimum. After obtaining liquid from the Natural gas, lighter components, primarily methane, ethane, carbon dioxide, are required to be separated because they increase RVP of LPG and in the latter stages, they get liberated as gas and do not get condensed. To facilitate this objective, fractionation of liquid hydrocarbon is carried out. Efforts are put up to minimize propane loss in the column top gases and to minimize slippage of lighter components in the liquid from the column bottom. In the next stage, distillation is carried out to segregate LPG and NGL components. In this stage, efforts are made to minimize carry over of pentane and heavier components in LPG because they increase weathering of LPG and at the same time slippage of butane in the NGL is minimized. Due to slippage of butane in NGL, RVP of NGL increases and during storage, butane gets vaporized and is lost in the atmosphere. Reduction of butane in NGL makes NGL safer to handle. To achieve above-mentioned objectives it is necessary that the whole operation is optimized at each stage. Mainly, following areas can be identified where optimization should be essentially done. 1. Feed gas cooling. 2.Feed gas drying and filtration. 3 Cold box. 4.Feed gas chilling through cryogenics. 5 Separation of condensate. 6.Light Ends Fractionation column. 7. LPG column. 8. Gas dryer

regeneration. 9. Heat Exchangers and steam condensate pots. 10. Minimization of hydrocarbon loss due to flaring

5.2. OBSERVATION AND ANALYSIS :-

An LPG recovery plant requires control of various parameters. Change in a particular value provides desired result for a particular parameter but at the same time it may have negative effect on another parameter. Success of an operation lies with the optimization of the situation. It is important to achieve optimum conditions in important sections of the LPG recovery plant. These sections are broadly classified as:

1. Feed gas cooling

2. Feed gas drying and filtration.

3. Feed gas chilling through cryogenics.

4 Separation of condensate.

5 Light Ends Fractionation column.

6 LPG column.

7 Gas dryer regeneration.

8 Heat Exchangers and steam condensate pots.

9 Steam condensate flash drums.

Optimum inlet pressure, optimum gas flow rate, optimum cooling, optimum turbo expander inlet and outlet pressure, optimum temperature and pressure profiles of LEF and LPG columns, optimum gas dryer regeneration cycle, optimum steam condensate pot pressure and temperature, optimum reflux ratio etc. These features are elaborated for application in an LPG recovery plant. A brief discussion on important instruments and their logical application is also included. Advanced

Process Control system further helps in achieving the targets through minimization of deviation.
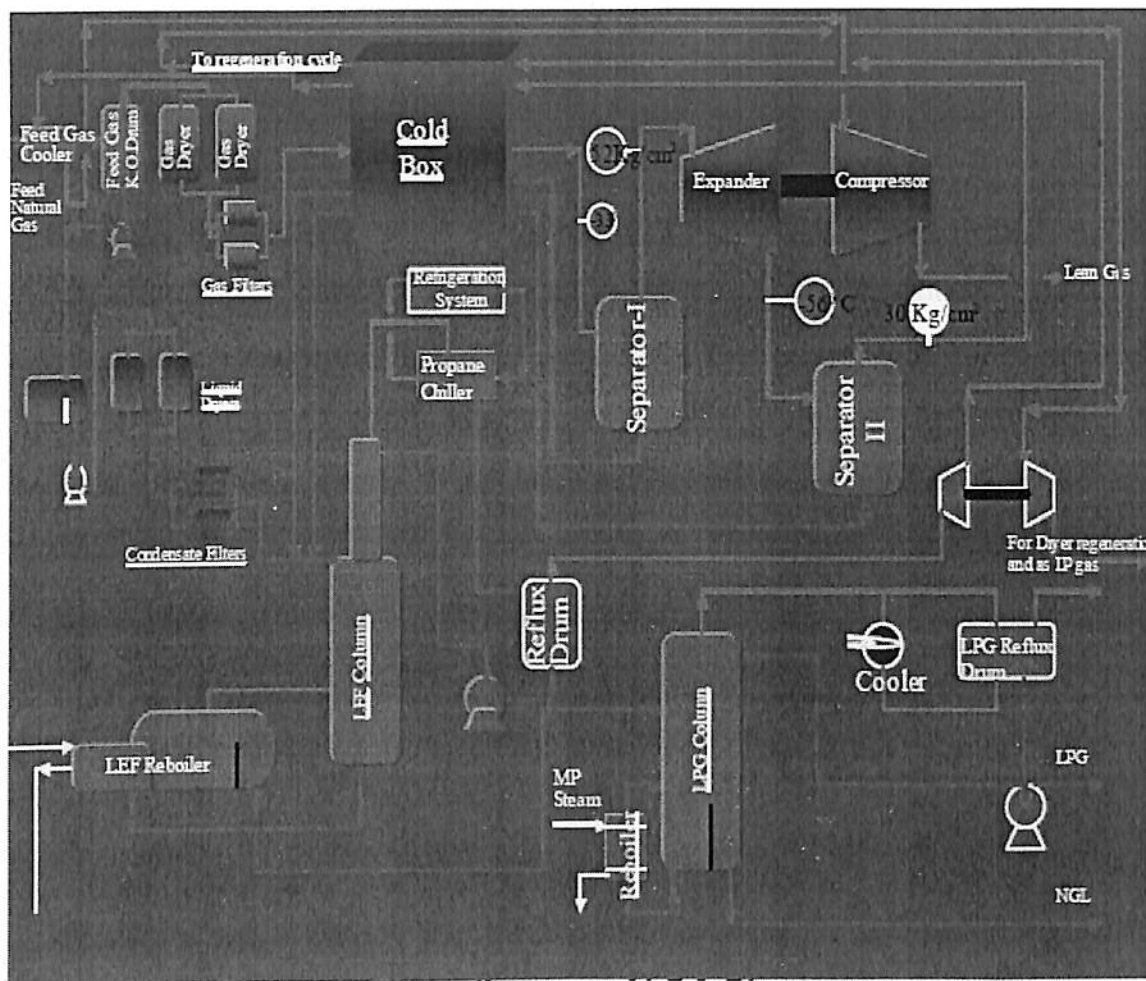
Figure 1: Flow diagram of LPG Plant

5.2.1. FEED GAS COOLING: - To recover maximum possible condensate from natural gas, cooling of natural gas is required. Plants are designed to withdraw maximum heat from the feed gas and is accomplished in stages. In the first stage, feed gas is cooled through separator-II outlet gas. A heat exchanger with suitable design is used for this purpose. In this stage optimum cooling is required. At a pressure of 52 Kg/cm2 downfall of temperature below 18 deg.C can cause hydrate

formation in the heat exchanger. But insufficient cooling can increase load on Gas dryers and can cause insufficient fall in Expander Compressor outlet temperature resulting less liquid recovery from the gas stream and consequently fall in production of VAPs. It is observed that maintaining temperature between 20 deg.C – 22 deg.C provides an optimum situation where good control is available on temperature and fluctuations can be handled. Provision of bypassing of cooling gas is kept in the feed gas cooler and can be used for controlling temperature. Pressure drop across cooler should be monitored in the feed gas stream and any increase in the pressure drop may be due to Hydrate formation. Once hydrate formation starts, it takes only few minutes when Feed Gas Cooler gets filled up by hydrates. In emergency situations, feed gas stream can be bypassed and without cooling, gas can be temporarily taken directly do feed gas dryer. However, a temperature range mentioned above would provide an optimum condition for operation. There should be and normally there are installed dehydrating facilities wherever chances of hydrate formation are present. Methanol is a good, and economical dehydrating agent, which brings down the hydrate formation temperature. Hydrates formation temperature is brought down by methanol according to Hammerschmidt Equation,

$$dT = 2335W/(100M - MW)$$

dT = temperature depression, 0F W = weight per cent inhibitor M = molar mass (molecular weight) of inhibitor (g/mol) = 32.043 for methanol

For example, a 25 wt% solution of methanol will effect the following temperature depression:

$$dT = 2335(25)/[(100)(32.043) - (32.043)(25)]$$

$$= 24.3 \text{ deg F} = 13.5 \text{ deg C}$$

5.2.2. FEED GAS DRYING AND FILTRATION:- Natural Gas contains water, which is required to be removed before its sub zero temperature cooling. Presence of moisture in the feed gas would form hydrates during its cooling. It should be assured that gas dryer outlet dew point is maintained below minus 70 deg.C. While designing gas dryers, if bed height is increased, pressure drop across gas dryer is increased which causes reduction of inlet pressure at Expander inlet. Low Expander inlet pressure means less expansion and less cooling which is not desirable. Normally, Molecular sieves are the drying medium for feed gas. It is important that proper leveling and packing of drying medium is carried out. Loose packing may cause more attrition and damage to drying medium. Feed gas from gas dryer may contain small hard materials. To filter out this material, gas filters are installed. Normally these are cartridge type filters with candle type screens. Pressure drop across these filters should be continuously measured and if increase in the pressure drop is at the higher side, cartridges should be replaced/ cleaned.

5.2.3. Cold Box:- Cold Box is one of the most important equipments of an LPG recovery Plant. It is a complex type of heat exchanger where five different streams enter and go out. They exchange heat amongst themselves. It is better to have an inlet screen with suitable holes and strength in the inlet point of feed gas. It obstructs any solid material. Cold Boxes are designed for continuous duty and for a long life. Main problems encountered are related to increase in pressure drop across Cold Box and hydrate formation. It is necessary to have methanol injection points in the upstream of cold box in all streams. Methanol injection should be used in case of ant obstruction observed. Some times gasket-bursting gives good result for decreasing pressure drop. Feed gas entering at

about 20 deg.C , comes out of Cold Box with temperature about -33 deg.C . If gas dryer is not working properly, hydrate formation can be observed in this stream. If methanol injection is not helping in reduction of pressure drop across cold box, Gasket bursting can be used to remove any mechanical obstruction. At few occasions leakage has been observed in the cold box and in that condition, vendor should be immediately contacted. An arrangement of inlet screen with cleaning facility, continuous pressure drop monitoring, methanol injection facility, good gas drying and filtering provide an optimized situation where operation can be carried out uninterrupted for a long duration.

5.2.4. FEED GAS CHILLING THROUGH CRYOGENICS: - This is the most important section of the LPG recovery plant where cooling of Feed gas is carried out. A screen is installed in the inlet of Expander Compressor. If the pressure drop across Expander inlet screen increases beyond a limit, instrumentation logics trip Expander Compressor. Expander Compressor is energy efficient equipment where energy generated due to expansion of gas in the expander section is used for compression of gas in the compressor section. Various instrument logics safeguard the equipment and eliminate unsafe operation conditions. It should be tried to maintain the expander inlet pressure at around 56 Kg/cm2. Expander outlet pressure should be maintained at around 29.4 Kg/cm2. With these inlets and outlet conditions, Expander outlet temperature is expected to be around –55 deg.C. This is an optimized temperature. If temperature is above –54 deg.C condensation of propane is affected whereas if temperature is below -56 deg.C, no gain in condensation of propane is observed but more ethane gets dissolved in the condensed liquid which causes problem in downstream processing.

An optimized temperature range of –54 deg.C to -56 deg.C at a pressure of about 29.4 Kg/cm2 gives the best result. Cooling to lower temperature causes presence of more ethane in separated liquid. This ethane gets liberated in Light Ends Fractionation Column where it imparts greater load on propane compressor. If ethane crosses LEF Column, then in LPG Condenser it does not condense and is required to be flared out causing direct loss of hydrocarbons. If Expander outlet temperature is at the higher side, this condition does not allow efficient condensation and heavier hydrocarbons are not separated out to the optimal level causing direct effect on VAP production. Expanders are designed to be run up to certain speed in terms of RPM. High RPM may cause damage to equipment while low RPM causes under utilization of equipment. Rotoflo equipments with 20000 RPM give good duty. Different instrumentation logics are provided in Expander Compressors. It is necessary to see that all instruments are functioning correctly and none is by-passed. Expander Compressors have very sensitive lubricating system and Lube Oil quality, pressure and temperature should be properly monitored and optimized. Cooling of natural gas in Expander Compressor is due to Joule Thompson effect.

Joule Thompson through their experiment found a joule thompson coefficient given by,

$$\frac{\Delta T}{\Delta p} \approx \left(\frac{\partial T}{\partial p}\right)_H = \mu_{JT}$$

μJT is not zero for real gas. Due to which temperature variation is seen while expanding a real gas. The positive or negative value of μJT

indicates that whether a gas would get cooled or become hot after expansion. Positive value of coefficient indicates fall in temperature. Expander Compressors are high-speed equipments and besides proper balancing and alignments, they need sufficient and proper instrumentation. Failure of any instrumentation logics can have detrimental effect. In an incidence, due to electrical problem lube oil pumps tripped. Due to instrumentation failure, SDV did not close and as HIC had the passing problem, expander compressor ran without lube oil causing severe damage to shaft, bearing and other parts. Expander Compressors are provided anti surge valves which supply the make up gas for compressor suction. With less flow of gas the anti surge valve would remain constantly open and with fluctuation of feed there would be fluctuation in opening and closing of anti surge valve. Anti surge valves cause a lot of vibration and wild fluctuations provide tremendous stress on expander compressor and anti surge valves along with associated pipelines. Smooth operation can reduce the wear and tear. All PGs, LGs, TGs should be regularly calibrated and checked. Faulty equipments create problem in optimally running the critical machines.

5.2.5. Separation of condensate: - Wherever there is a fall in temperature, condensation takes place. At the inlet of LPG Plant, in the feed gas cooler, condensation takes place. Here, along with liquid hydrocarbons, condensation of free water also takes place. Interface level controllers are installed in the feed gas knock out drum. However, water can be drained in oily water sewage system. Removal of free water at this stage reduces load on liquid dryer. Liquid condensate is pumped and passed through liquid dryer where moisture concentration is brought down to 2ppm. This liquid is sent to LEF Column. Condensate is separated in Separator-I & II.

Majority of the liquid (about 80%) is separated out in Separator-I while rest is accumulated in Separator-II. Expander trip logics are provided in separators where due to high level in separators, Expander Compressor trips. If level comes down significantly, gas would ingress in the bottom outlet causing disturbance in LEF Column. It is essential to optimize separator level. For normal operation maintaining level around 40% provides good margin and flexibility and has greater tolerance for fluctuations. Expander outlet pressure provides energy to Separator-II liquid for pushing it out and feeding to LEF Column. LEF Column operates at 27 Kg/cm2 and Expander outlet pressure is kept around 29.2 Kg/cm2. Further reduction in Expander Compressor outlet pressure reduces energy available for pushing Separator-II liquid. This may cause continuous increase in the Separator –II level. If Expander outlet pressure is increased, Expander outlet temperature starts increasing causing reduction in liquid recovery. Practically, a pressure range of 29 to 29.4 Kg/cm2 has been found as an optimum pressure range.

5.2.6. LIGHT ENDS FRACTIONATION COLUMN: -This is a distillation column where lighter components are removed from the top of the column. Liquid recovered from Natural gas at different stages, is fed to LEF Column. Liquid from the bottom of the column is heated in a U-tube shell and tube heat exchanger up to a temperature of about 107 deg.C . Vapor rises in the column and exchanges heat with the upper plate and vaporizes liquid on the upper plate and so on the principles of distillation lighter components, mainly ethane and lighters are taken out from the top. In the process, some heavier hydrocarbons also vaporize. Column top gases are chilled through propane and heavier hydrocarbons are condensed again. Liquid condensed is totally refluxed via a reflux

drum. The whole system becomes a bit complex and parameters are inter-related. Change in one parameter affects other parameters. Main objective of an optimization effort concentrates on minimization of propane loss in column top gases and minimization of ethane slippage in the bottom product. Column top temp between +1 deg.C and −1 deg.C , with top pressure about 27 Kg/cm2, LEF reboiler outlet vapor temperature about 107 deg.C can be one of the optimum conditions. However the best optimization can be achieved with composition analysis of LEF top vapor, LEF bottom liquid. Incremental adjustments in pressure, temperature, level can reach the optimum condition. Advanced Process Control (APC) can be helpful in operating LEF column continuously with optimized condition. Propane chiller is an important equipment and optimization process is very much dependent on performance of propane chiller. Level of propane in the chiller plays an important role in optimization of LEF Column. Column top temperature is very much dependant on reflux temperature which can be changed through level change of liquid propane in the chiller. But high vapor load on the chiller increases load on propane compressor. However no optimum condition can totally stop propane loss or ethane slippage but an optimum condition can restrict propane loss in the top gas below 3.5% and that should be achieved. Ethane slippage in the bottom product must be minimized because whatever ethane slippage would be there, that would not be re-condensed and would create increase in LPG reflux drum pressure and finally would be required to be flared.

5.2.7. LPG COLUMN:- This is a distillation column where based on boiling point difference, propane and butane are separated out from rest of the heavier hydrocarbons. Propane and butane together known as

LPG, are recovered from the top vapor and bottom product is NGL. Like any other distillation column, this column needs proper control of pressure, temperature, level, feed and reflux rate to achieve an optimized situation. Efforts are made to restrict butane slippage in the bottom product and carry over of heavier hydrocarbons, in the top product. Presence of heavier hydrocarbons in LPG affects volatility and weathering above +2 deg.C is not acceptable by marketing agencies. Presence of lighter components like ethane causes increase in RVP, which also is not acceptable. LPG reflux drum pressure controls the condensation process through level in LPG condenser. Siphon type reboiler is used to heat the bottom liquid in LPG Column. LPG column bottom level is one of the important parameters for controlling pressure in the column. Rate of reflux should be properly optimized to get quality products. LPG column with reboiler outlet temperature around 155 deg.C , top temperature 58 deg.C , reflux drum pressure 10.3 Kg/cm2, reflux level around 50%, can be in one of the optimum conditions. However these parameters depend greatly on feed composition. LPG Column designed with different configurations can have different optimum parameters. However these parameters can be rightly optimized through trial and error method.
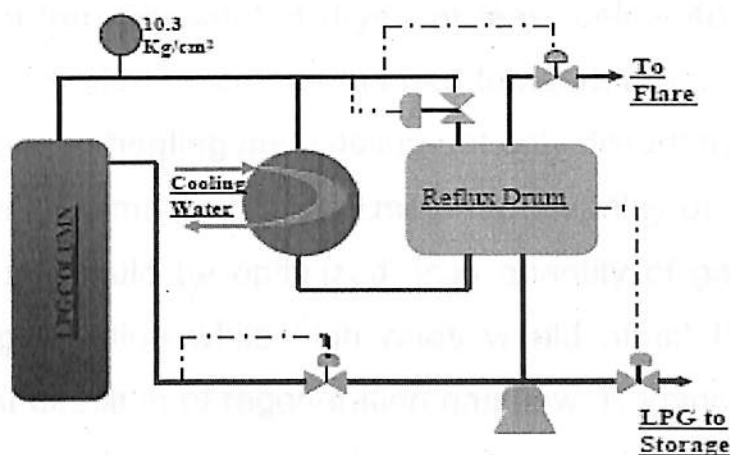
Figure 2: Schematic diagram of LPG Column top section

Advanced Process Control can be helpful in operating LPG column in optimized condition with fast compensation and re-adjustment of parameters. As shown in the Figure2, while stabilizing the parameters, top section of LPG Column produces more complexities. Column top pressure, reflux rate, top temperature, products withdrawal rate, all are inter-related and optimization can be done through incremental changes. through incremental changes.

5.2.8. GAS DRYER REGENERATION: - Regeneration gas at about 265 deg.C and dryer out let gas temperature at 235 deg.C with about one hour regeneration with these parameters provides sufficient drying. However, complete regeneration heating cycle takes about five hours. After regeneration, drying bed should be gradually cooled till the regeneration dryer out let gas temperature falls below 40 deg.C . If bed is not properly cooled, when feed gas is taken in to regenerated dryer, high feed gas outlet temperature causes operational disturbance in the whole downstream process. To minimize such disturbance, it is essential to

sufficiently cool the regenerated dryer, at least below 40 deg.C . Every effort should be made to increase the bed temperature gradually in small increments. Sudden heating and cooling not only disturb the operation but cause permanent damage to drying medium. Quantity of gas to be used for regeneration should be optimized, less quantity of gas would cause insufficient regeneration while high rates would affect the bed due to counter current direction of regeneration gas flow. Regeneration of drying medium is an important aspect and more damage to the drying medium and increase in feed gas outlet moisture dew point is mainly due to improper regeneration. Normally, regeneration of gas dryers is carried out at the dryer pressure around 12 Kg/cm2. After completion of regeneration, the regenerated dryer is pressurized back to operating pressure. This process should be carried out slowly and with dry gas. High gas rate can severely damage the drying bed.

5.2.9. HEAT EXCHANGERS AND STEAM CONDENSATE POTS: -

Heat exchangers are meant for exchange of heat from one fluid to another. Due to various reasons, efficiency of Heat Exchangers decreases and mainly due to scaling of surfaces. It is essential to get heat-exchanging surfaces cleaned periodically. Major problems are encountered in water coolers which get severely affected due to scaling caused by poor quality of cooling water. Periodic cleaning and proper treatment of cooling water can minimize the problem. It is essential to check cooling water in the outlet of heat exchangers. There is always an apprehension of leakage of coolers causing mixing of hydrocarbon in the returning cooling water and posing serious unsafe and hazardous situation. Steam is used in the reboilers at certain pressure. Steam gets condensed and steam condensate is collected in condensate pots. If

steam condensate pot gets flooded, there remains less space left to steam in heat exchanger for exchange of heat and tubes get filled with water, allowing less heat transfer due to condensation of steam. If steam pot gets emptied then level controllers come in direct contact of steam and valve gate and seat become more vulnerable to damage. Further, steam may directly come out of steam pot and cause disturbance in the downstream of condensate pot. It is important that steam is available in heat exchanger at desired pressure. Poor pressure affects heat transfer and poor heat transfer is witnessed. Normally steam condensates are collected in flash drums where steam is flashed out and reused for further heat exchange. This makes the system more energy efficient and heat of steam is used and reused before finally sending the steam condensate to boiler house. Steam pots and other downstream sections are prone to steam leakage primarily due to two phase flow and incorrect pH of steam condensate. While pH can be corrected through additives, effect of two-phase flow can be minimized through sound maintenance practice. All flanges should be properly aligned, steam traps should be of good quality and steam/condensate leakage arrest through injection should be avoided as due to variation of temperature these materials get damaged and get transferred intrinsically causing damage to stream traps.

5.2.10 **MINIMIZATION OF HYDROCARBON LOSS DUE TO FLARING:** Loss of hydrocarbon through flaring is a direct loss of energy and in an energy deficient society it is criminal. Optimization of LPG recovery plant can not only minimize the flaring but in practical terms can achieve hundred percent zero flaring. Although there are many points from where flaring can occur due to high abnormal conditions but in the normal operations there are very few locations from where flaring can

occur. Gas dryers are depressurized for regeneration and the gas is bled in to LP gas header. Fast rate of depressurization can lead to sudden increase in LP gas pressure causing flaring. Slow rate of depressurization can reduce the chances and sufficient action for readjustment of downstream parameters can totally eliminate the chances of flaring. LPG Column is the main source of flaring. This is mainly due to two reasons; carry over of lighter components from the bottom of LEF Column and poor performance of LPG condenser. Disturbance in LPG column operation also can cause flaring. Optimized upstream operations as described above can minimize the chances of carry over of lighter components whereas condensers can be diagnosed for reasons of poor performance. Condensers can have scaling problem or cooling water temperature can be at the higher side. Condensers can be cleaned to remove scaling and cooling water problem can be shorted out through examination and corrective actions. If due to any technical reason, particular equipment is required to be depressurized/emptied out and if there is any chance of presence of liquid hydrocarbons, liquid must be transferred to blow down vessel from where liquid hydrocarbon can be reprocessed. Flare knock out drums should have facility to transfer the collected liquid to reprocessing units or to slop tanks. Control valves are installed at different locations to control the abrupt increase of pressure in a section. This control is done through releasing the hydrocarbon in to the flare header. Proper training, alertness, fast and accurate action by operating engineer can reduce the chances of flaring significantly. Many of these things are gained through experience. Many times situations are unique and an experienced, accomplished process engineers saves a lot through his prompt action.

6. CONCLUSIONS:-

The term NGL (natural gas liquids) is a general term which applies to liquids recovered from natural gas and as such refers to propane, butane (LPG) and other products.

LPG is a gas at atmospheric pressure and normal ambient temperatures, but it can be liquefied when moderate pressure is applied or when the temperature is sufficiently reduced. It can be easily condensed, packaged, stored and utilized, which makes it an ideal energy source for a wide range of applications.

Normally, the gas is stored in liquid form under pressure in a steel container, cylinder or tank. The pressure inside the container will depend on the type of LPG (commercial butane or commercial propane) and the outside temperature.

Butane and Propane are used in combination...

While butane and propane are different chemical compounds, their properties are similar enough to be useful in mixtures. Butane and Propane are both saturated hydrocarbons. They do not react with other. Butane is less volatile and boils at 0.6 deg C. Propane is more volatile and boils at - 42 deg C. Both products are liquids at atmospheric pressure when cooled to temperatures lower than their boiling points. Vaporization is rapid at temperatures above the boiling points. The calorific (heat) values of both are almost equal. Both are thus mixed together to attain

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