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# LPG UNIT
# (ENGINEERING DESIGN GUIDELINE)



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## INTRODUCTION

### Scope

This design guideline covers the basic elements in the field of LPG Units in sufficient detail to allow an engineer to design a LPG unit with the suitable size of diameter, velocity, reflux rate and actual stages. This design guideline includes design of conventional LPG unit, Petlyuk column, condensers and reboilers.

LPG unit is utilized to separate the natural gas liquids into products which can be futher processed. The sizing of the unit and the choice of column is important to give the optimum efficiency and save the energy cost of LPG unit. There are several technologies which have been developed in order to extract LPG from natural gas vapors and liquids (NGL).

The design of LPG unit may be influenced by factors, including process requirements, economics and safety. Include in this guideline is a calculation spreadsheet for the engineering design. All the important parameters use in the guideline are explained in the definition section which help the reader more understand the meaning of the parameters or the term used.

In the application section of this guideline, three case studies are shown and discussed in detail, highlighting the way to apply the theory for the calculation. The theory section explained about conventional LPG sizing, Petlyuk column sizing and other equipment in LPG unit.

## INTRODUCTION

### General Design Consideration

In natural gas processing plants, several stages of separation and fractionation are used to purify the natural gas from the liquid heavier hydrocarbons. This separated liquid is named as natural gas liquids (NGL). The source of Natural Gas Liquids is a natural occurring mixture of gaseous hydrocarbons found in the ground or obtained from specially wells. The composition of natural gas varies in different parts of the world. Its chief component, methane, usually makes up from 80% to 95% of its composition. The balance is composed of varying amounts of ethane, propane, butane, and other hydrocarbon compounds approximately as follows:

1. Ethane (35-55%)
2. Propane (20-30%)
3. Normal Butane (10-15%)
4. Isobutane (4-8%)
5. Pentanes Plus (also called natural gasoline, debutanized natural gasoline) (10-15%)

The raw NGL is sent to LPG recovery plant to separate LPG (i-$C_3$ and i-$C_4$) from stabilized NGL ($C_{5+}$). Both products are very valuable and expensive in the market. Liquefied Petroleum Gas (LPG) is a mixture of hydrocarbon gases, primary propane and butane. The exact composition of LPG varies according to its source, processing principles and depends on the season.

LPG is odorless, colorless and non-toxic. To reduce the danger of an explosion from undetected leaks, commercial LPG usually contains an odorizing agent, such as ethanethiol, which gives it a distinctive pungent odor. LPG has a higher calorific value (94 MJ/$m^3$ equivalents to 26.1 kWh) than natural gas (38 MJ/$m^3$ equivalents to 10.6 kWh).

Table 1 - Properties of LPG

| Name of the property | Value for LPG |
|---|---|
| Freezing point | -187$^{o}$C |
| Specific gravity | 0.588 |
| Vapor pressure at 38$^{o}$C | 1212 kPa |
| Heat content | 50221 kj/kg |

There are several technologies which have been developed in order to extract LPG from natural gas liquids (NGL). Some research and simulations have also been done to find the most optimum and economic technology.

**Conventional Technology**

Natural gas processing begins at the wellhead. The composition of the raw natural gas extracted from producing wells depends on the type, depth, and location of the underground deposits and the geology of the area. The natural gas produced from oil wells is generally classified as “associated-dissolved,” meaning that the natural gas is associated with or dissolved in crude oil. Natural gas production without any association with crude oil is classified as “non-associated.” About 60% of the world supply of LPG comes from associated gas processing, and 40% of the LPG is produced in oil refineries from crude distillation, fluid catalytic cracking units, hydrocrackers, etc.

There are 4 columns used in the conventional process. First stage of LPG extraction from NGL is DeEthanizer. In the DeEthanizer column, methane and ethane are expected to be separated and flow through the top of the column. Since there is no requirement to liquefy methane and ethane, especially in small amounts, these components will be kept in vapor phase and additional condenser is unnecessary.

In this DeEthanizer, 100% methane and ethane in mixed feed can be separated and leaves the top of the column. This product will be used internally as a fuel to generate the steam which is used for reboiler of the column. Heat is supplied to the column by

forced circulation using reboiler pump to reboiler and into the column. The heavier hydrocarbons other than ethane leave the column as bottom product in a liquid phase. Next it is sent to DeButanizer column.

DeButanizer is used instead of DePropanizer because the bottom product of DeEthanizer composed of small amount of propane (2 % of mass fraction) and butane (5 % of mass fraction). So that smaller columns will be used for next extraction. This is for economic reason and the efficiency of the separation as it is easier to extract the product this way and less duty will be required for reboiler.

Before entering DeButanizer column, the DeEthanizer bottom product is expanded from 26 - 17 bar and fed into DeButanizer as mixed-phase feed. This feed is fractionated into mixed propane and butane as overhead product and heavier hydrocarbons ($C_{5+}$) as bottom product. Then the overhead product is totally condensed in the condenser by heat exchange with cooling water, and condensate is collected in reflux drum. The reflux drum should be used in order to prevent cavitation on the pump due to vapor phase. The DeButanizer reflux and product pump serve the dual purpose of supplying reflux to the column and allowing withdrawal of column overhead product butane from the reflux drum. The column heat is supplied by a reboiler and circulation is aided by DeButanizer reboiler circulating pump.

About 100% propane and 99% butane can be recovered from the feed at the overhead column product. This stream leaves the column and is sent to DePropanizer column to separate propane and butane. Meanwhile, the bottom product composed of pentane and heavier hydrocarbons will be stored as natural gasoline. Since the bottom product has high temperature (> 200°C), it will be cooled by heat exchanger before being sent to the storage.

Propane and butane stream is expanded from 16 - 10 bar and enters the DePropanizer as mixed-phase feed. The DePropanizer separates propane as overhead product and butane as bottom product. Condenser is used to totally condense the overhead vapor from DePropanizer. Condensate is collected in DePropanizer column reflux drum. A part of the condensed overhead product is sent back to the column as reflux via pump while the remaining part is withdrawn as a liquid propane product. DePropanizer reboiler heat is supplied by reboiler. Reboiler circulation is aided by reboiler circulation pump.

It is about 99.9% of propane can be recovered as top product and 99.9% of butane is recovered as bottom product. The butane product is field grade butane which is composed of 30% i-butane and 65% n-butane. This is next being sent to butane splitter to get i-butane and n-butane products separately, since it could be sold with higher price than field grade butane.

The field grade butane is expanded from 10 - 5 bar before enter the butane splitter. I-butane is recovered as overhead product and n-butane as bottom product. I-butane is totally condensed by condenser. Condensate is collected in reflux drum. A part of the condensed overhead product is sent back to the column as reflux via pump while the remaining part is withdrawn as a liquid i-butane product. Butane splitter heat is supplied by reboiler. Reboiler circulation is aided by reboiler circulation pump. Butane splitter column separates 96% mole of i-butane in the overhead product and 96% mole of n-butane in the bottom product. Both products have higher price than field grade butane.

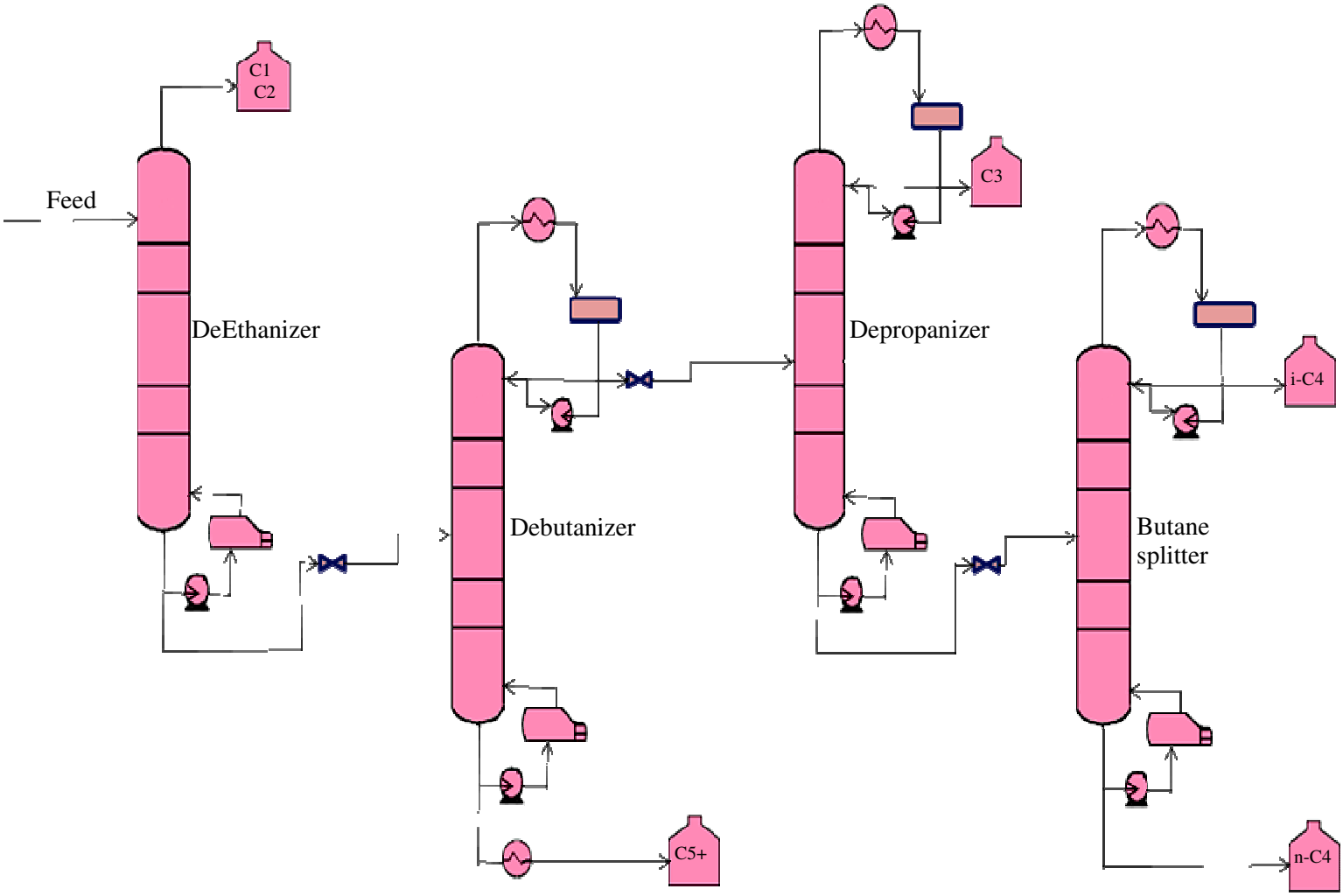


Figure 1: Process flow diagram of LPG unit

There are several stages of gas processing in figure 1. Below is discussing their purpose.

1. Gas-oil separators

A multi-stage gas-oil separation process is needed to separate the gas stream from the crude oil. These gas-oil separators are commonly closed cylindrical shells, horizontally mounted with inlets at one end, an outlet at the top for removal of gas, and an outlet at the bottom for removal of oil.

2. Condensate separator

Condensates are most often removed from the gas stream at the wellhead through the use of mechanical separators. In most instances, the gas flow into the separator comes directly from the wellhead, since the gas-oil separation process is not needed.

3. Dehydration

Dehydration is the removal of this water from the produced natural gas and is accomplished through several methods. Usually using ethylene glycol (glycol injection) system as an absorption mechanism to remove water and other solids from the gas stream. Alternatively, adsorption dehydration may be used, utilizing dry-bed dehydrators towers, which contain desiccants such as silica gel and activated alumina, to perform the extraction.

4. Contaminant removal

Removal of contaminants such as hydrogen sulphide, carbon dioxide, water vapor, helium, and oxygen. The technique is to first direct the flow though a tower containing an amine solution. Amines absorb sulphur compounds from natural gas and can be reused repeatedly. After desulphurization, the gas flow is directed to the next section, which contains a series of filter tubes. As the velocity of the stream reduces in the unit, primary separation of remaining contaminants occurs due to gravity.

5. Methane separation

Cryogenic processing and absorption methods are some of the ways to separate methane from natural gas liquids (NGLs). The cryogenic method is better at extraction of the lighter liquids, such as ethane. Cryogenic processing consists of lowering the temperature of the gas stream to around -120 F. The turbo expander process is the most effective to perform this function, using external refrigerants to chill the gas stream. The quick drop in temperature that the expander is capable of producing, condenses the hydrocarbons in the gas stream, but maintains methane in its gaseous form.

The absorption method, on the other hand, uses a “lean” absorbing oil to separate the methane from the NGLs. While the gas stream is passed through an absorption tower, the absorption oil soaks up a large amount of the NGLs. The “enriched” absorption oil, now containing NGLs, exits the tower at the bottom. The enriched oil is fed into distillers where the blend is heated to above the boiling point of the NGLs, while the oil remains fluid. The oil is recycled while the NGLs are cooled and directed to a fractionator tower.

Two Common Processes for Methane Separation - Gas Subcooled Process (GSP) and Residue Split Vapor (RSV)

Gas subcooled process (GSP) and residue split-vapor (RSV) have been utilized for efficient NGL/LPG recovery from natural gas, particularly for those gases containing significant concentrations of carbon dioxide. To increase the ethane recovery beyond the 80% achievable with the conventional design, a source of reflux must be developed for the DeMethanizer. The GSP was developed to overcome this problem and others encountered with the conventional expander scheme.

A. Gas Subcooled process (GSP)

The GSP was developed to increase the ethane recovery beyond the 80% achievable with the conventional design. a portion of the gas from the cold separator is sent to a heat exchanger where it is totally condensed and subcooled with the overhead stream. This stream is then flashed to the top of DeMethanizer, providing reflux to the DeMethanizer. The expander feed is sent to the tower several stages below the top of the column. Because of this modification, the cold separator operates at much warmer conditions well away from the system critical. Additionally, the residue recompression is less than with the conventional expander process.

The GSP design has several modifications. One is to take a portion of the liquid from the cold separator along with the gas to the overhead exchanger. Generally, this can help further reduce the horsepower required for recompression. Also, the process can be designed to just use a portion of the cold separator liquid for reflux. This modification is typically used for gases richer than 3 GPM (C+2 ).

When CO2 is present in the feed gas, the higher concentrations of C+2 components in the cold liquids help reduce the amount of CO2 concentrating in the upper, colder sections of the tower, allowing higher ethane recovery levels without CO2 freezing. This same process can be operated to reject ethane, but propane recovery efficiency suffers significantly when operated in this mode due mainly to the higher concentration of propane present in the top feed

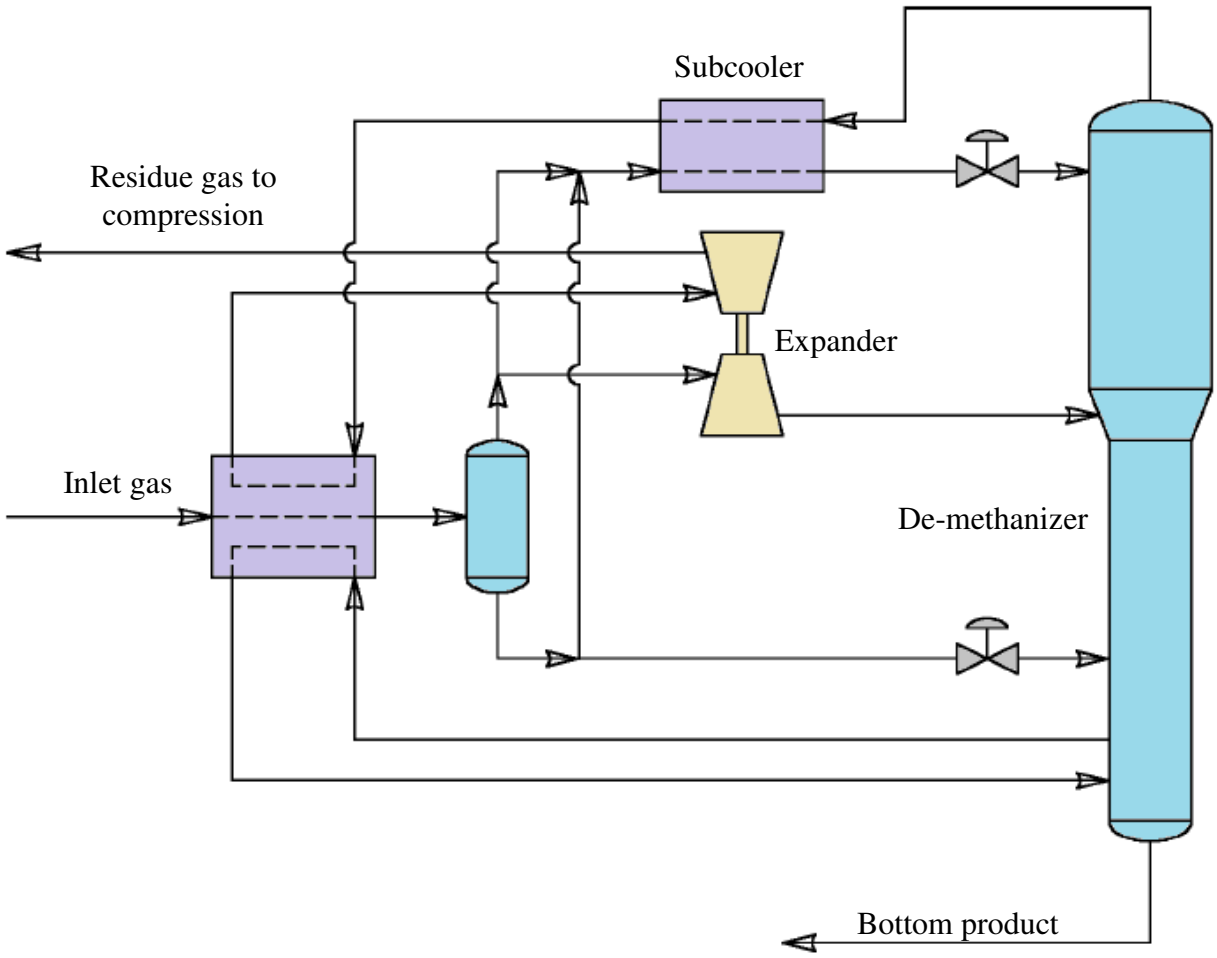


Figure 3: Schematic of GPC (gas subcooled process)

B. Residue split-vapor (RSV)

Another method of producing reflux is to recycle a portion of the residue gas, after recompression, back to the top of the column, the RSV process. The process flow is similar to the GSP design except that a portion of the residue gas is brought back through the inlet heat exchanger. At this point, the stream is totally condensed and is at the residue gas pipeline pressure. The stream is then flashed to the top of the DeMethanizer to provide reflux.

The subcooled inlet gas split and the expander outlet stream are sent lower down in the tower rather than to the top of the column. The reflux provides more refrigeration to the system and allows very high ethane recovery to be realized. The recovery level is a function of the quantity of recycle in the design.

It is $CO_2$ tolerant and the recovery can be adjusted by the quantity of recycle used. The RSV process can be used for very high ethane recoveries limited only by the quantity of horsepower provided.

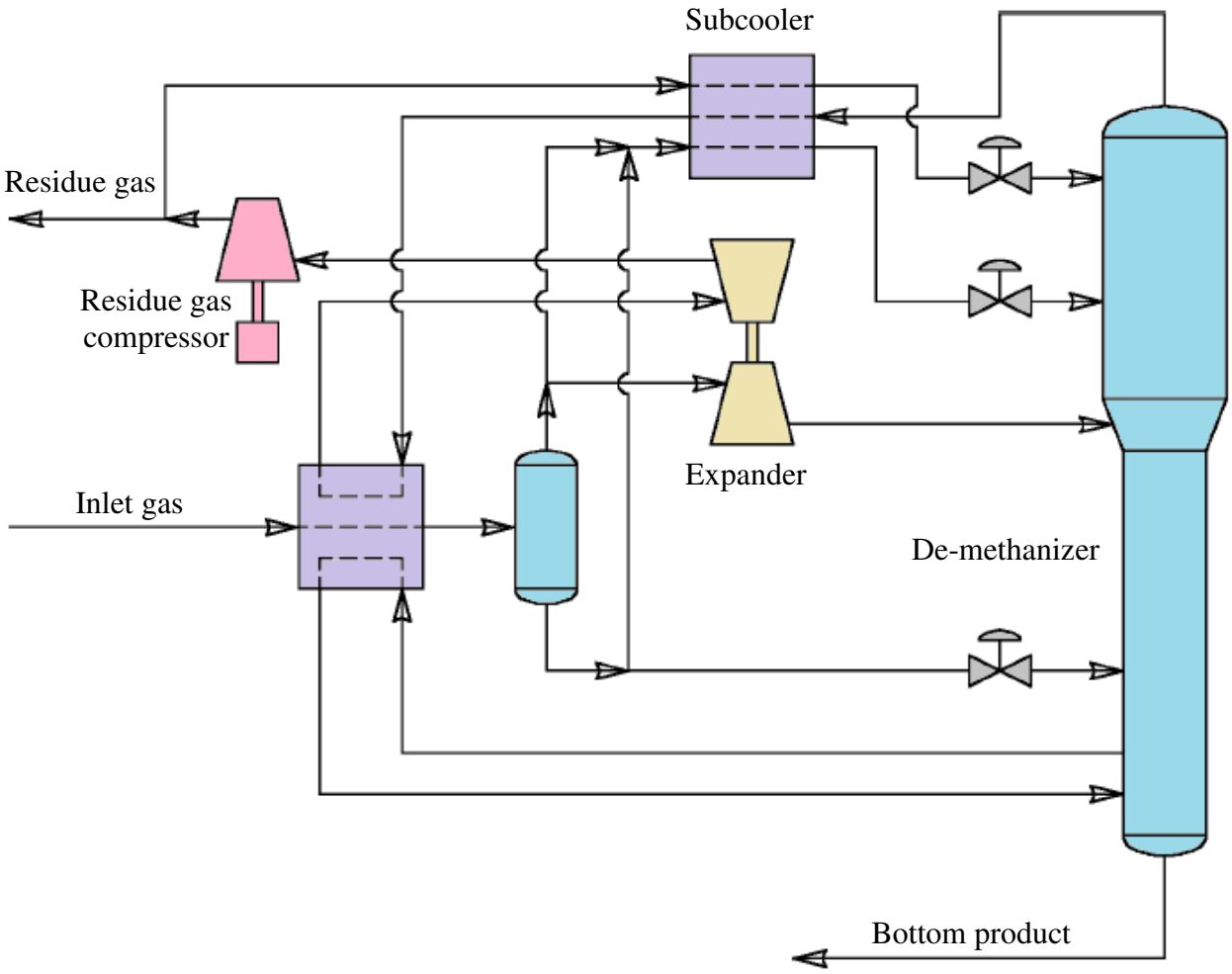


Figure 4: Schematic of residue split-vapor process

6. DeEthanizer

The stripper bottoms from the extraction plant enter DeEthanizer column near the top. The overhead vapor is partially condensed in DeEthanizer condenser by heat exchange with medium-level propane. Condensed overhead product in overhead reflux drum is pumped back to the DeEthanizer by reflux pump. The non-condensed vapor, mainly ethane, leaves the plant to fuel the gas system. Heat is supplied to the column by forced circulation reboiler. The DeEthanizer column operates at approximately 26.9 bar. Approximately 98% of the propane in the DeEthanizer feed is recovered in the bottom product. The residual ethane concentration is reduced to approximately 0.8 mole % in the bottom product. The bottom product from DeEthanizer drains into DePropanizer column.

7. DePropanizer

DeEthanizer bottoms are expanded from 26.9 - 20 bar and enter DePropanizer as mixed-phase feed. The DePropanizer fractionates the feed into a propane-rich product and a bottom product comprised of butane and natural gasoline. Tower overhead vapor is totally condensed in the DePropanizer condenser by cooling water. Condensate is collected in DePropanizer column reflux drum. A part of the condensed overhead product is sent back to the column as reflux via pump while the remaining part is withdrawn as a liquid propane product. Column reboiler heat is supplied by direct-fired heater. Reboiler circulation is aided by reboiler circulation pump. The bottom product is sent to DeButanizer.

8. DeButanizer

The DePropanizer bottoms are expanded from approximately 20 - 7.6 bar and then enter the DeButanizer column as a mixed-phase feed. The column feed is fractionated into a butane-rich overhead product and natural gasoline bottoms. The columns overhead are totally condensed in the DeButanizer condenser by heat exchange with cooling water, and condensate is collected in reflux drum. The DeButanizer reflux and product pump serve the dual purpose of supplying reflux to the column and allowing withdrawal of column overhead product butane from the reflux drum. The column reboil heat is supplied by a direct-fired DeButanizer reboiler, and boiler circulation is aided by DeButanizer reboiler circulating pump. The bottom product leaving the column is cooled

in product cooler. A part of the gasoline product is recycled to the LPG extraction unit and serves as lean oil for the absorber column

**Petlyuk Column**

In order to increase the process efficiency of such distillation processes, the following two alternatives have been proposed both in the literature and by industrial practitioners.

1. Integration of Conventional Distillation Arrangements - Includes sequential arrangement of distillation columns with energy integration between the columns or other parts of the plant.
2. Design of new configurations - Includes Dividing Wall Column; which consists of an ordinary column shell with the feed and side stream product draw divided by a vertical wall through a set of trays. The same configuration is usually denoted as a Petlyuk column

A column arrangement separating three or more components using a single reboiler and a single condenser, in which any degree of separation (purity) can be obtained by increasing the number of stages (provided the reflux is above a certain minimum value). The Petlyuk column consists of a pre-fractionator followed by a main column from which three product streams are obtained and this arrangement has been shown to provide higher energy savings than the systems with side columns, with savings of up to 30% when compared to conventional schemes.

From figure 2, the Petlyuk design consists of a pre-fractionator with reflux and boilup from the downstream column, whose product is fed to a 2-feed, 3-product column, resulting in a setup with only one reboiler and one condenser. A pair of intermediate liquid and vapor streams passes from the prefractionator to the secondary column. The practical implementation of such a column can be accomplished in a single shell by inserting a vertical wall through the middle section of the column, thus separating the feed and the side product draw. Petlyuk's main reason for this design was to avoid thermodynamic losses from mixing different streams at the feed tray location.

Typically, a light key and a heavy key are designated for the whole system. The light key and lighter components will appear in stream Pv while the heavy key and heavier

components will appear in stream P3. The middle components will appear in stream P2. To accomplish such a separation the prefractionator must split non-sharply with the middle components distributing.

The Petlyuk configuration is quite complex. The first step in the analysis is to transform the configuration into something that is easier to handle. To accomplish this goal the secondary column is divided into two separate columns at the sidestream tray.

Petlyuk columns decrease energy expenditures for separation of threecomponent mixtures, on the average, by 30% due to their thermodynamical advantages:

1. In the preliminary column, the composition of flows in feed crosssection is close to feed composition (i.e., thermodynamic losses at mixing of flows are nearly absent);
2. These losses at the mixing of flows at the ends of the columns are nearly absent;
3. Absence of reboiler or condenser at output of component 2 decreases energy expenditures due to the fact that liquid and vapor flows are used twice in the sections located above and below output of component 2
4. Thermodynamic losses for the reason of repeated mixing of flows in the second column at regular separation sequence are absent (the concentration of component 2 at the end of the first column at direct split along distillation trajectory decreases, which requires additional expenditures of energy in the second column for obtaining pure component 2).

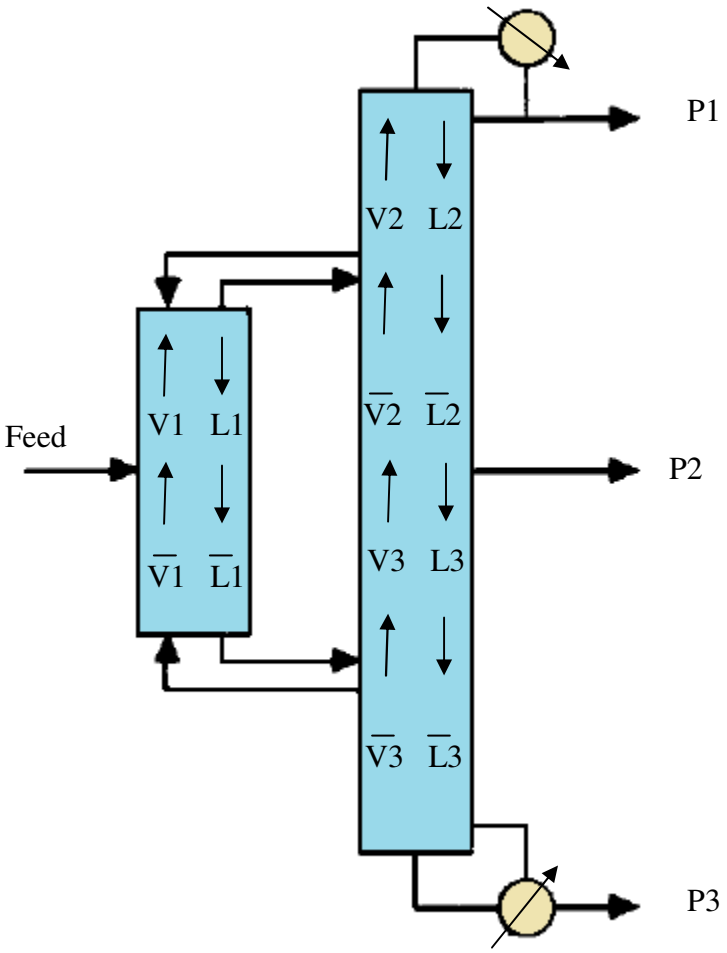


Figure 2: Petlyuk Configuration

## DEFINITIONS

**Bottoms -** The liquid or residual matter which is withdrawn from the bottom of a fractionator or other vessel during processing or while in storage

**Bubble point -** The temperature at a specified pressure at which the first stable vapor forms above a liquid.

**Boiling point** - the temperature corresponding to equilibrium between the liquid and vapor phases at 101.325 kPa (or 14.696 psia).

**Compressibility factor -** A factor, usually expressed as "Z," which gives the ratio of the actual volume of gas at a given temperature and pressure to the volume of gas when calculated by the ideal gas law.

**Condensate -** The liquid formed by the condensation of a vapor or gas; specifically, the hydrocarbon liquid separated from natural gas because of changes in temperature and pressure when the gas from the reservoir was delivered to the surface separators. In a steam system it may be water that is condensed and returned to the boilers.

**DeButanizer -** A fractionator designed to separate butane (and more volatile components if present) from a hydrocarbon mixture.

**Dehydration -** The act or process of removing water from gases or liquids.

**DeMethanizer -** A fractionator designed to separate methane (and more volatile components if present) from a hydrocarbon mixture.

**DePropanizer -** A fractionator designed to separate propane (and more volatile components if present) from a hydrocarbon mixture.

**Dew point -** The temperature at any given pressure, or the pressure at any given temperature, at which liquid initially condenses from a gas or vapor. It is specifically applied to the temperature at which water vapor starts to condense from a gas mixture (water dew point), or at which hydrocarbons start to condense (hydrocarbon dew point).

**Distillation -** The process of separating materials by successively heating to vaporize a portion and then cooling to liquefy a part of the vapor. Materials to be separated must differ in boiling point and/or relative volatility.

**Extraction -** The process of transferring one or more components from one liquid phase to another by virtue of different solubility in the two liquids. It is also used to indicate removal of one or more constituents from a stream.

**Fractionation -** Generally used to describe separation of a mixture of hydrocarbons into individual products based on difference in boiling point and/or relative volatility.

**Gas constant (R) -** The constant multiplier in the Ideal Gas Law. Numerically, R=PV/T, if V is the volume of one mole of an ideal gas at temperature T and pressure P.

**Gas processing -** The separation of constituents from natural gas for the purpose of making salable products and also for treating the residue gas to meet required specifications.

**Heavy ends -** The portion of a hydrocarbon mixture having the highest boiling point. Usually hexanes or heptanes and all heavier hydrocarbons are the heavy ends in a natural gas stream.

**Light hydrocarbons -** The low molecular weight hydrocarbons such as methane, ethane, propane and butanes.

**LP-gas (liquefied petroleum gas) -** Predominately propane or butane, either separately or in mixtures, which is maintained in a liquid state under pressure within the confining vessel.

**Natural gas -** Gaseous form of petroleum. Consisting predominately of mixtures of hydrocarbon gases. The most common component is methane.

**NGL (natural gas liquids) -** Natural gas liquids are those hydrocarbons liquefied at the surface in field facilities or in gas processing plants. Natural gas liquids include ethane, propane, butanes, and natural gasoline.

**Reflux -** In fractionation, the portion of condensed overhead returned to the column to enhance achievable purity of the overhead product.

**Reflux ratio -** A way of giving a relative measurement to the volume of reflux. Usually referred either to the feed or overhead product.

**Recovery -** That percent or fraction of a given component in the plant feed which is recovered as plant product.

**Recycle -** Return of part of a process stream to a point upstream from where it was removed to enhance recovery or control.

**Relative volatility** – the ratio of the concentration of one component in the vapor over the concentration of that component in the liquid divided by the ratio of the concentration of a second component in the vapor over the concentration of that second component in the liquid. For an ideal system, relative volatility is the ratio of vapor pressures i.e. $\alpha = P2/P1$

**Specific gravity -** The ratio of the mass of a given volume of a substance to that of another equal volume of another substance used as standard. Unless otherwise stated, air is used as the standard for gases and water for liquids, with the volumes measured at 60°F and standard atmospheric pressure.

**Splitter -** A name applied to fractionators, particularly those separating isomers (e.g., butane splitter refers to a tower producing most of the isobutane in the feed as overhead and most of the normal butane in the feed as bottoms).

**Stripper -** A column wherein absorbed constituents are stripped from the absorption oil. The term is applicable to columns using a stripping medium, such as steam or gas.

**Traded - column** A vessel wherein gas and liquid, or two essentially immiscible liquids, are contacted, usually counter-currently on trays. Also refer to packed column.

## NOMENCLATURE

| | |
|---|---|
| B | Bottom flow rate, lb/hr |
| $B_{iA}$ | Flow rate of A in the bottoms of column i, lb mol/hr |
| $B_{iB}$ | Flow rate of B in the bottoms of column i, lb mol/hr |
| $B_{iC}$ | Flow rate of C in the bottoms of column i, lb mol/hr |
| D | Distilate flow rate, lb/hr |
| $D_{1C,underwood}$ | Flow rate of C in the distillate of 1 at min. reflux, lb mol/hr |
| $D_{iA}$ | Flow rate of A in the distillate of column i, lb mol/hr |
| $D_{iB}$ | Flow rate of B in the distillate of column i, lb mol/hr |
| $D_{iC}$ | Flow rate of C in the distillate of column i, lb mol/hr |
| $d_T$ | Tower diameter design, ft |
| $DT_2$ | The temperature difference between the hot outlet and the cold inlet stream |
| $E_{OC}$ | Overall efficiency |
| F | Feed flow rate, lb/hr |
| $F_A$ | Flow rate of A in the feed to the dividing wall column, lb/hr |
| $F_B$ | Flow rate of B in the feed to the dividing wall column, lb/hr |
| $F_C$ | Flow rate of C in the feed to the dividing wall column, lb/hr |
| H | Tower height, ft |
| Li | Rectifying section liquid flow rate in column i, lb mol/hr |
| Li' | Stripping section liquid flow rate in column i, lb mol/hr |
| $L_w$ | Weir length, in |
| N | Number of theoretical stages |
| $N_1$ | Number of stages for column 1, Stages |
| $N_{1min}$ | Minimum number of allowable stages for column 1, Stages |
| $N_{act}$ | Actual stages |
| $N_m$ | Minimum stages |
| Q | Heat duty |
| $q_i$ | Quality (saturated liquid fraction) of feed to the column I, |
| $R_i$ | Reflux Ratio in column i |
| $R_{imin}$ | Minimum reflux ratio for column i |
| S | Tray spacing, in |
| $u_V$ | Max. allowable vapor velocity, ft/s |
| Vi | Rectifying section vapor flow rate in column i, lb mol/hr |
| Vi' | Stripping section vapor flow rate in column i, lb mol/hr |
| $V_m$ | Maximum vapor rate, lb/s |

| | |
|---|---|
| $X_B$ | Mole fraction bottom light key |
| $X_D$ | Mole fraction overhead light key |
| $X_F$ | Mole fraction feed light key |

**Greek letters**

| | |
|---|---|
| $\alpha$ | Realtive volatility |
| $\alpha_A$ | Relative volatilities component A |
| $\alpha_B$ | Relative volatilities component B |
| $\alpha_C$ | Relative volatilities component C |
| $\Delta t$ | Log mean temperature difference |
| $\Delta T_1$ | The temperature difference between the hot inlet and cold outlet stream |
| $\mu_L$ | Feed liquid viscocity, Cp |
| $\rho_L$ | Liquid density, lb/ft$^3$ |
| $\rho_V$ | Vapor density, lb/ft$^3$ |
| $\Phi_1$, $\Phi_2$, $\Theta$, $\psi$ | Underwood roots. Use UWMulti to find the underwood roots |

**Superscript**

| | |
|---|---|
| B | Bottom flow rate, lb/hr |
| D | Distilate flow rate, lb/hr |
| F | Feed flow rate, lb/hr |
| H | Tower height, ft |
| N | Number of theoretical stages |
| Q | Heat duty |