International Journal of Engineering Research and Technology. ISSN 0974-3154, Volume 13, Number 12 (2020), pp. 5055-5065

Integrating Dehydration and Natural Gas Liquids Processes for Maximization of Natural Gas Liquids Production

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Abstract

The Growing demand for liquefied petroleum gas (LPG) as fuel is increased in Egypt, in recent decades, this work aims to modify the performance of existing natural gas liquids (NGLs) plants to maximize LPG. This work presents a new technique to the NGLs plant to maximize the LPG plant production capacity and industry profitability. The technique involves integrating the dehydration recovery unit and natural gas liquids unit to obtain the requisite cooling and maximize the LPG production. Besides, a new arrangement for de-ethanizer, de-butanizer, and condensate stabilizer towers in the NGLs recovery unit is presented. These towers achieve some great goal heat recovery and maximize LPG recovery as the highest added value product. HYSYS simulation software Version 11 is used in this work to simulate and analyze the existing and the modified NGLs plant. The results showed that the LPG production is maximized by a 49% percent increase from the original NGLs plant. The increasing production of LPG can share to solve the LPG shortage problem in Egypt. This modification can be taken as guidelines for both new and plants in operation to increase their profits.

Keywords: Natural gas liquids, liquefied petroleum gas, dehydration, integration, maximization, simulation, Egypt

1. INTRODUCTION

In natural gas handling plants, a few phases of separation and fractionation are utilized to decontaminate the natural gas from the liquid heavier hydrocarbons. This separated liquid is referred to as liquids from natural gas (NGLs). To segregate LPG (i-C3 and i-C4) from stabilized condensate (C5+), crude NGLs are sent to the LPG recovery plant [1]. The two items are truly significant and costly in the market. Because of clean consuming qualities and the capacity to meet rigid ecological necessities, the interest in natural gas has expanded impressively in recent years [2].

Recently due to the growing demand for NGLs and specified in Egypt LPG consumption by households increased from 1,963 metric tons, thousand in 1998 to 4,154 metric tons, thousand in 2017 growing at an average annual rate of 4.12% [3].

The gas composition majorly affects the financial aspects of NGL recovery. NG could be normally ordered to lean or dry (for low ethane and heavier C_2+ content) and rich gas (for high C_2^+ content) [4]. As a rule gas with a more noteworthy amount become liquid hydrocarbons produces products a more prominent amount of products subsequently more prominent incomes for the gas processing facility [5].

There are different processes for NGLs recovery that are available, based on alternative cooling methods such as Joule Thomson (JT) expansion, direct external refrigeration using chiller unit (usually propane), and Turbo Expander. More prominent features can recover and improve energy efficiencies if a mix of these alternatives is used [5].

According to Abd El-Ghany et al [6], an outline of the NGL recovery enhancement for GUPCO trans Gulf gas plant by using a new applicable technique is presented. The new technique is based on using a condensate stream to enrich the reflux of the de-ethanizer tower. By applying this technique more energy is recovered, and the efficiency of the plant increased from 38 % to reach 86-90 %. Butane recovery and its LPG production have increased by 170% to 122 tonnes per day instead of 44 tonnes per day.

According to Bhran et al [7], process simulation and performance improvement of a gas plant in operation (El-Wastani Petroleum Company Plant located in Egypt) are studied. The recoveries of butane, propane, or ethane as a final product for sales are maximized. They find that optimizing propane recovery is the optimal route for plant improvements... This is because the choice of propane recovery mode does not rely only on return on investment (ROI) but takes into consideration various objectives such as feed stability, marketing availability, and the recovered NGL quantity.

Also, great retrofitting for LPG plant will appear obviously in the future reality by using a mixed form of refrigerant instead of pure propane refrigerant, butanes and propane recoveries were increased by 13% and 7 % respectively with a 15.95% improvement in total LPG recovery with distinction to the upgraded plant **according to Shehata et al. [8]**.

The modified process has numerous benefits when compared to a traditional gas processing plant.

Many researchers worked on the general elective processes applied for natural gas liquids recovery. But there are limited works investigating treatment with different used techniques and NGLs recovery methods each separately. However, there are limited research studies that focused on improving the NGLs by reducing total capital cost by reducing the number of equipment parallel side by side by enormous production capacity significantly in border heat recovery considerations.

This study focuses a good point in getting out the optimum integrated model for the LPG plant that achieves great benefits in optimized heat recovery and maximized specified products. In this work, an existing natural gas liquids plant is modified with a reduction in the overall superstructure of the plant. The propane pre-cooling and the turbo-expander units used for cooling the natural gas are removed in the modified plant. The cooling duty required for cooling the natural gas for separating the NGLs was obtained by integration between the dehydration unit and LPG recovery unit. Also, a new arrangement of the existing de-ethanizer, de-butanizer, and stabilizer towers was applied and contributed to obtaining the required cooling energy and maximizing the LPG production.

2. SIMULATION RESULTS AND SPECIFICATIONS FOR ORIGINAL NGLs PLANT CASE

The process simulation of the studied case study includes two stages to produce the three products, LPG, sales gas, and condensate as shown in Table 1. These products should be produced to their specifications. Figure 1 presents a block diagram for the original NGLs plant and Figure 2 presents the LPG recovery plant simulation printout.

Table 1. Feed and Production Original Plant Quantities

Feed		Sales Products		
Raw Gas	**Raw Condensate**	**LPG**	**Sales Gas**	**Condensate**
MMscfd	Bbls/day	Ton/day	MMscfd	Bbls/day
161.5	4000	204.4	157.059	5092

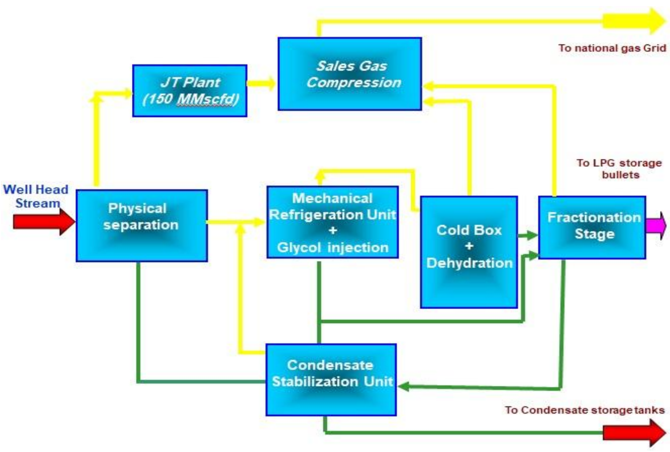

Figure 1. Block Diagram for original NGLs Plant

This study aimed to raise the value of the gas stream feed by adding a new arrangement for LPG recovery unit, using a low-temperature technique, then recovering the LPG from the NGL through an integrated circuit.

After separation of the feed gas streams in the 3-phase separator into gas, liquid, and water bottom stream. The sweet gas enters two exchangers; Gas/Gas and Gas Chiller exchangers. It is cooled to -10 ^oC.

When natural gas is saturated with water flows in a pipeline issues can happen obviously in hydrate formation is prevented by drying the gas chemical addition, the most popular inhibitor today is ethylene glycol (EG) [9].

By directly injecting an 80% ethylene glycol/water mixture upstream of Gas/Gas exchangers and also a 20% glycol/water mixture upstream gas chiller, hydrate formation will be prevented in the gas stream.

The three-phase separator vessel (3-phase LTS) separates the gas, HC liquids, and rich glycol at approximately 18 min residence time. Cold gas from the top LTS used to cool the inlet gas in Gas/Gas Exchangers, where leaving shell side and will meet the sales gas specifications for water and HCDP. The rich glycol is fed to the regeneration unit for regeneration and re-injection.

The HC liquids are fed under control through propane sub-cooler into the condensate stabilization unit passing through Condensate/Condensate Exchanger in Stage-2 for LPG recovery processing.

The condensate stabilization unit is to stabilize the condensate to sales specifications concerning vapor pressure. The condensate from the 3-phase separator passes through the Condensate Stabilizer Feed Drum. The gas separated drum is fed to the suction of the Stabilizer Overhead Compressor (Stab OH Comp) while the condensate is fed to the Condensate Stabilizer tower.

The Condensate Stabilizer tower where the HC liquid flows downward through the 14 trays of the column counter currently contacting with the HC vapor rising through the top of the column when the liquid reaches the bottom of the column flows to the Stabilizer Re-boiler which provides enough heat to vaporize a portion of the HC liquid to produce a C+5 product, cooled by two coolers one after Stabilizer Re-boiler and in Condensate/Condensate inter exchanger received condensate from the separator (3-phase LTS) enters tube side and condensate in shell side from cooler after Stabilizer Re-boiler. Finally, the condensate came stabilized as a final product.

Stage-2 designed simulated to produce LPG product with 80% Butane recovery-based, a turbo expander, and de-Ethanizer, de-Butanizer fractionation towers. With productivity reaches up to **245** ton/day of LPG, **5000** Bbls/day of stabilized condensate, and **153** MMSCF of sales gas.

Turbo Expander is the most efficient machine to control hydrocarbon Dew Point used. The Hydrocarbon Dew Point Control (HCDP) unit consists of Cold Box Exchanger, Expander Suction Vessel, Turbo Expander/Compressor, and Expander Discharge Vessel.

Dry gas flows and enters the Cold Box Exchanger (LNG-100) at -10 ^{0} C and 49.3 bar pressure where it is cooled down to -34.5 ^{0}C by the cold gas from Expander Discharge Vessel (V-100) And cold condensate from Expander Suction Vessel (T/E Suc. V).

Due to cooling of dry gas become condensed and are separated in Expander Suction Vessel (T/E Suc. V) And gas enters to Turbo Expander (T/E) at -34.5 ^{0}C and 49.9 bar and condensate at to -34.5 ^{0}C to LPG production unit through Cold Box Exchanger. The gas achieves a drop-down to - 48.5 ^{0}C and 35.5

bar by Expander operation.

The gas and liquid hydrocarbon that condensed in turbo expander due to temperature drop is delivered to the Expander Discharge Vessel (V-100) for gas and liquid phase separation.

The cold gas from Expander Discharge Vessel (V-100) at - 48.5 ^{0}C and 35.5 bar are used in Cold Box Exchanger (LNG-100) that increases its temperature to -12 ^{0}C and sends to pre-cool the inlet process gas.

At tray# 4, condensate from Turbo Expander Suction Vessel (T/E Suc.V) enters de-ethanizer column after passing through Cold Box Exchanger (LNG-100) at -17 ^{0}C temperature and 48.6 bar pressure.

At tray #7, condensate from Turbo Expander Discharge Vessel (V-100) after passing through Propane Sub-Cooler (E-101) at -31 ^{0}C temperature and 35.28 bar then enters the de-ethanizer column (De-C2).

At tray #10, condensate from Condensate/Condensate Exchanger enters de-ethanizer column at 46 ^{0}C temperature and 8 bar pressure.

Condensate flows down through the 28 valve trays where stripping gas produced in de-ehanizer re-boiler (De-C2) strip off C_2 from the liquid.

Overhead Condenser cools and condenses the gas exiting the de-ethanizer column (De-C2) by C_3 refrigerate in the shell side prior to and disengaged from the liquid hydrocarbon are compressed to the sales gas compressor (De-C2 OH Comp).

De-ethanizer re-boiler heats condensate at 175 ^{0}C temperature and 28.94 bar pressure to produce stripping gas and draw offside flows while the liquid becomes richer with C3+ as it flows downward to the de-butanizer tower for further processing to LPG recovery.

From de-ethanizer re-boiler stream goes to de-ethanizer unit which stream carrying C+3 flows and enter at the top of tray #7 of de- butanizer Tower (De-C4) at 150 ^{0}C temperature and 14 bar pressure.

As the liquid moves down through the 24 valve trays stripping produced consisting of C3/C4 hydrocarbons and strip off C3 and C4 from the liquid contents. At the liquid bottom, column enters in de- botanizer re-boiler where stripping stream that generating at the 200^{0}C temperature which rises up in the tower to stripping C3/C4 from the liquid and leaving the top of the column.

The overheads liquid (Final LPG stream) leaves the top of the column passing through Air Fans Condenser where is condensed and enter the Reflux Drum to condense any liquids that are pumped back at the top of the column as reflux. The bottom liquid (Final Stabilized Condensate stream) at the draw offside of the de- butanizer re-boiler is re-presented as a final product.

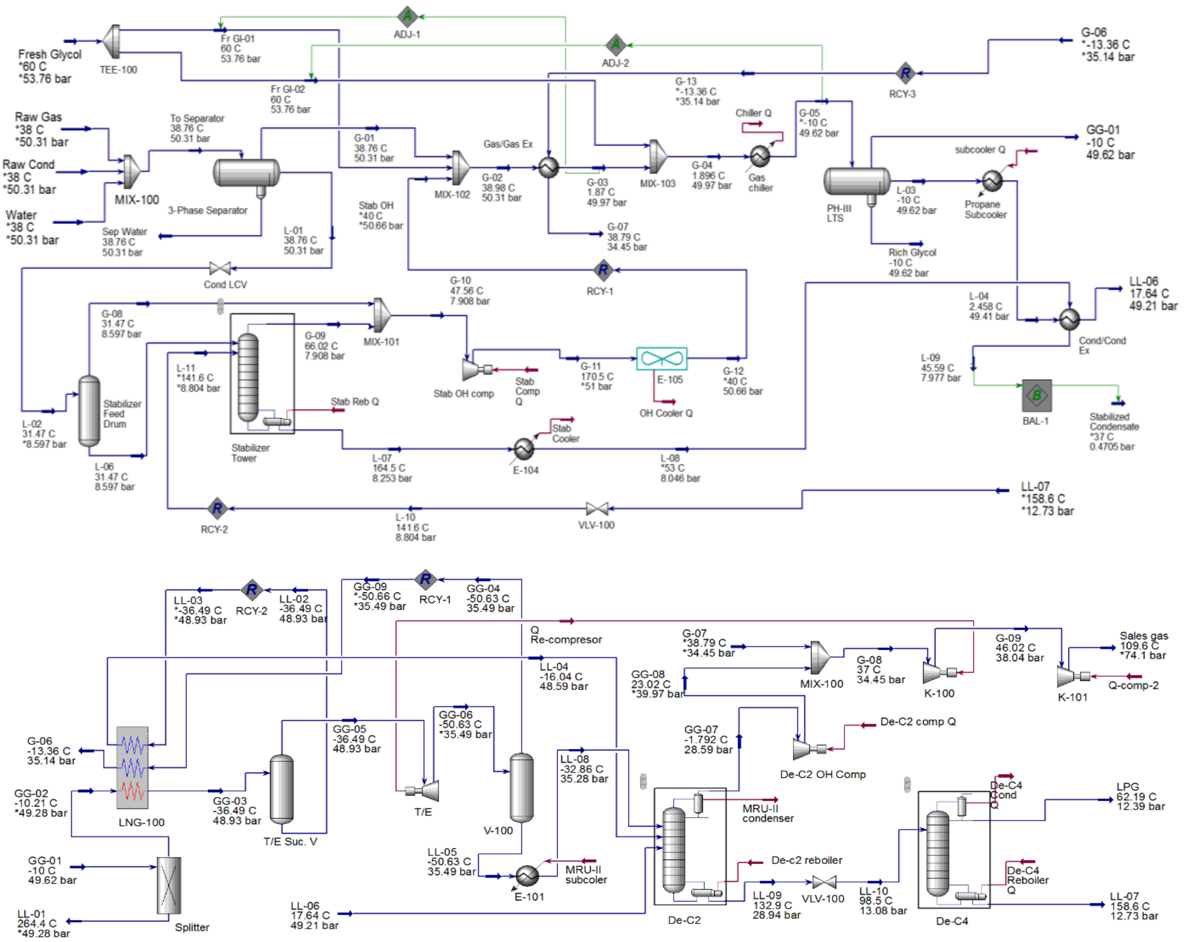

Figure 2. Simulation Printout of Existing NGLs Plant (Original Case Study)

3. SIMULATION RESULTS AND SPECIFICATIONS FOR MODIFIED CASE

The process simulation of Modified design case study, integrated modified model between dehydration unit and NGL recovery (Heavy Component Removal) unit of LPG plant achieve the required target of production sales gas and maximization recovery of propane, n-butane, and iso-butane (LPG) and stabilized condensate (C_5^+) that production according to its specifications that are referring in Table 2 feed and production modified plant quantities and is represented in the simulation following Figure 3 block diagram for Modified NGLs Plant and Figure 4 of Simulation Printout of Existing NGLs Plant (Modified Case Study).

Table 2. Feed and Production Modified Plant Quantities

Feed		Sales Products		
Raw Gas	Raw Condensate	LPG	Sales Gas	Condensate
MMscfd	Bbls/day	Ton/day	MMscfd	Bbls/day
161.5	4000	418.1	152.8	1248

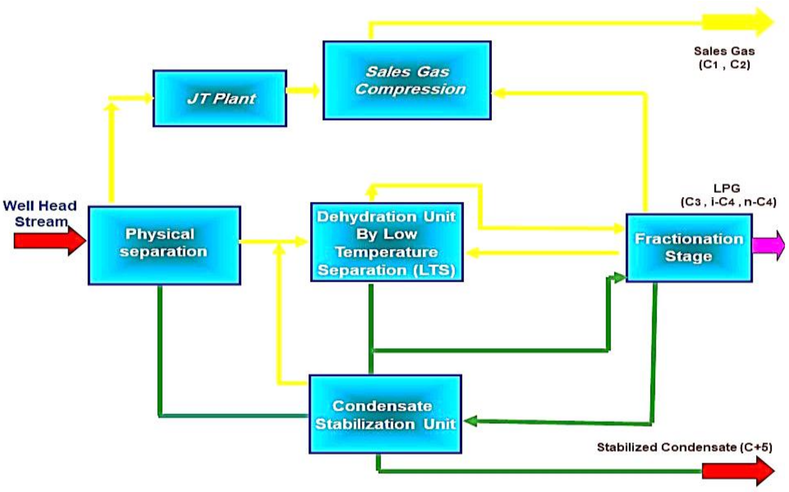

Figure 3. Block Diagram for the modified existing NGLs Plant

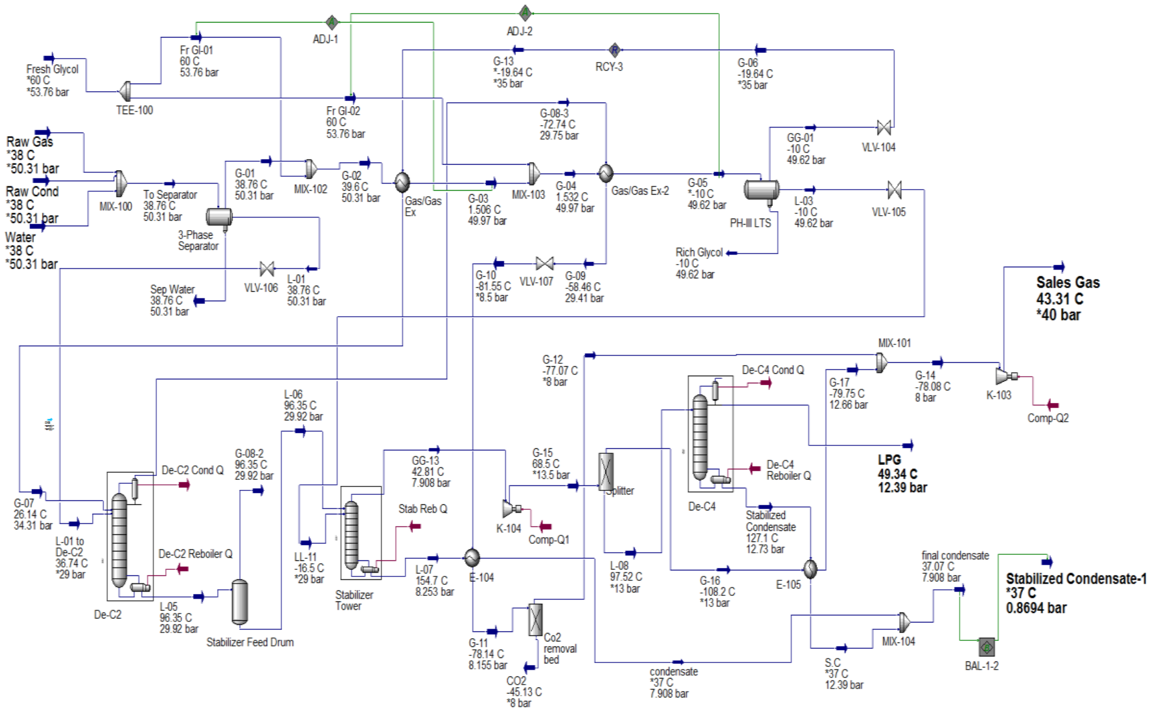

Figure 4. Simulation Printout of the modified existing NGLs Plant

Natural gas feed must be purified before it enters the low-temperature facility. The purpose of gas purification is to separate condensate ($C5^+$), and water from the natural gas to make these fluids suitable for sale or disposal [11]. The LPG plant consists of dehydration and an NGL recovery (Heavy Component Removal) sections Figure 1. The dehydrated gas is sent to a NGL recovery unit to purify from heavy components and represent as sales gas to the Egyptian National Gas Grid with its specifications. This study aimed to raise the value of the gas stream feed by adding a new arrangement for LPG recovery unit, using a low-temperature technique, and then to recover the LPG from the NGL both through an integrated circuit.

3.1 Dehydration Unit by Low-Temperature Separation (LTS)

A Low-Temperature separation (LTS) unit also called straight refrigeration or Low-temperature extraction (LTX) unit used for dew point control or gas conditioning. The process consists of cooling and partial condensation of the gas stream. When inlet pressures are sufficiently high to meet the requirements for discharge pressure to make pressure drop acceptable, cooling is obtained expansion through a J-T valve, otherwise, external refrigeration (but modified by heat recovery interexchange) is required [12]. When water-saturated natural gas flows in a pipeline issues can happen obviously in hydrate formation is prevented by drying the gas chemical addition, the most popular inhibitors today is ethylene glycol (EG) [13]. The feed stream consists of three phases that go through a 3-Phase separator for initial separation into gas and two liquids, gas bubbles rising up, remove free water, and separate condensate. Gas is cooled by two heat exchangers in series (Gas /Gas Ex and Gas/Gas Ex-2) but before hydrate prevention is achieved by directly injecting a total flow rate of 35.928m^3/day an 80%-20% ethyl glycol/water mixtures spilled into the inlet flow rate 34.344m^3/day of the Gas /Gas Ex and into Gas /Gas Ex-2 with flow rate 1.584m^3/day upstream each heat exchanger to absorb any water entrained in the gas stream. Then gas stream sent to 3-phase low-temperature separator (PH-III LTS) to separate the remaining droplets of water with rich glycol solution from condensate that achieves more cooling by J-T valve breaks its pressure for refrigeration Gas/Gas Ex present as a heat recovery stream. A second heat recovery stream is represented by the de-ethanizer column overhead stream (G-08) for cooling Gas/Gas Ex-2.then goes through the CO_2 removal bed to capture carbon dioxide from the gas stream to meet sales gas specifications, which is shown obviously in figure 5.

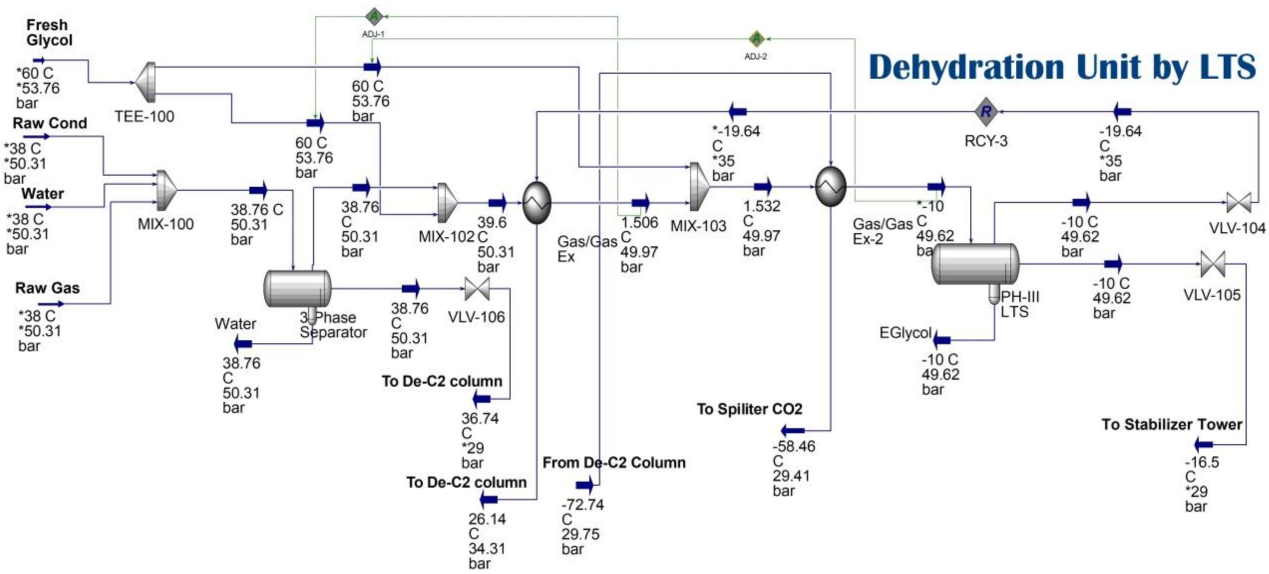

Figure 5. Flow Diagram of Dehydration Unit Integrated with NGLs Recovery Unit.

Table 3. Simulation results for Dehydration Unit

Equipment	Parameter	Unit	Design
3-Phase Separator	Molar Feed	MMscfd	189.1
Gas/Gas-Ex	Heat Duty	MMBtur/hr	15.57
Gas/Gas-Ex-2	Heat Duty	MMBtur/hr	5.142
PH-III LTS	Molar Feed	MMscfd	160.8
Glycol Injection	Std Ideal Liq Vol Flow	m3/day	35.928

International Journal of Engineering Research and Technology. ISSN 0974-3154, Volume 13, Number 12 (2020), pp. 5055-5065

3.2 NGLs Recovery Unit

The LPG recovery unit is added after the dehydration unit, after removing all condensate and water from the gas [14]. There are many proprietary NGL technologies available. They typically use multiple refluxes, column design, and heat exchanger configuration closely integrated to reduce cost. While these innovations may be more efficient and increase recoveries, they may prove to be difficult to operate under off-design conditions [15]. NGL recovery unit consists of a combination of components that must be separated into marketable products [15]. This separation process by a new arrangement occurs in fractionators, which can include, de-ethanizer, de-butanizer, and stabilizer tower in optimized integrated new process model path is displayed in figure 2. The two feed streams to a de-ethanizer is a liquid stream from 3-phase separator injects at tray no.14 by L-01 to De-C2 stream and a gas stream from overhead (PH-III LTS) separator at tray no.12 is injected by G-07 stream, caused by the high pressure and temperature below the initial boiling point, the feed point location is selected where the composition in the tower is similar to the feed composition. A de-ethanizer tower (De-C2) is designed with 28 trays; the condenser pressure is 29.75 bar by full reflux and re-boiler pressure 29.92 bar which separates light hydrocarbon gas mixture from the feed stream. The methane and ethane with traces of carbon dioxide go overhead and the C3+ material goes out the bottom. The de-ethanizer bottom go through the stabilizer feed drum to provide the first feed stream for condensate stabilizer tower injects at the top tray by L-06 and the second liquid feed stream from (PH-III LTS) separator at tray no.12 is injected by L-11 stream. Stabilizer tower is designed with 14 trays - type internal that reduces vapor pressure of condensate by removing lighter components, is typically carried out in the re-boiler and light components go up from the overhead gas GG-13 stream of stabilizer tower, as the liquid falling into the column, becomes linear in light components and richer in the heavy ends at the bottom of the tower represented by 787.7 barrel/day as stabilized condensate after cool down with interexchange heat recovery where E-104, some of the liquid is circulated through the re-boiler to add heat to the tower. Light of overhead gas GG-13 stream flows through K-104 compressor in raise up to 13.5 bar to inlet Splitter where split light hydrocarbon traces (C1 and C2) G-16 stream from heavier component (C3+) in L-08 stream. L-08 stream is de-butanizer feed acts at tray no.10, the de-butanizer tower to recover light distillate (C3, n-C4, iso-C4) through Liquefied Petroleum Gas (LPG) stream by total condensation of overhead distillate from heavy distillate of light naphtha during the refining process through stabilized condensate bottom stream by 467.4 barrel/day. The de-butanizer tower is designed with 24 trays, the LL-10 feed stream acts at tray number 10, the condenser pressure is 12.39 bar and the re-boiler pressure is 12.73 bar. In our integrated simulation production capacity with NGLs extraction by 418 ton/day LPG and 1248 barrel/day as stabilized condensate and 152.8 MMscfd as sales gas.it will be represented in obviously Figure 6 of the Flow Diagram for NGls Train Processes.

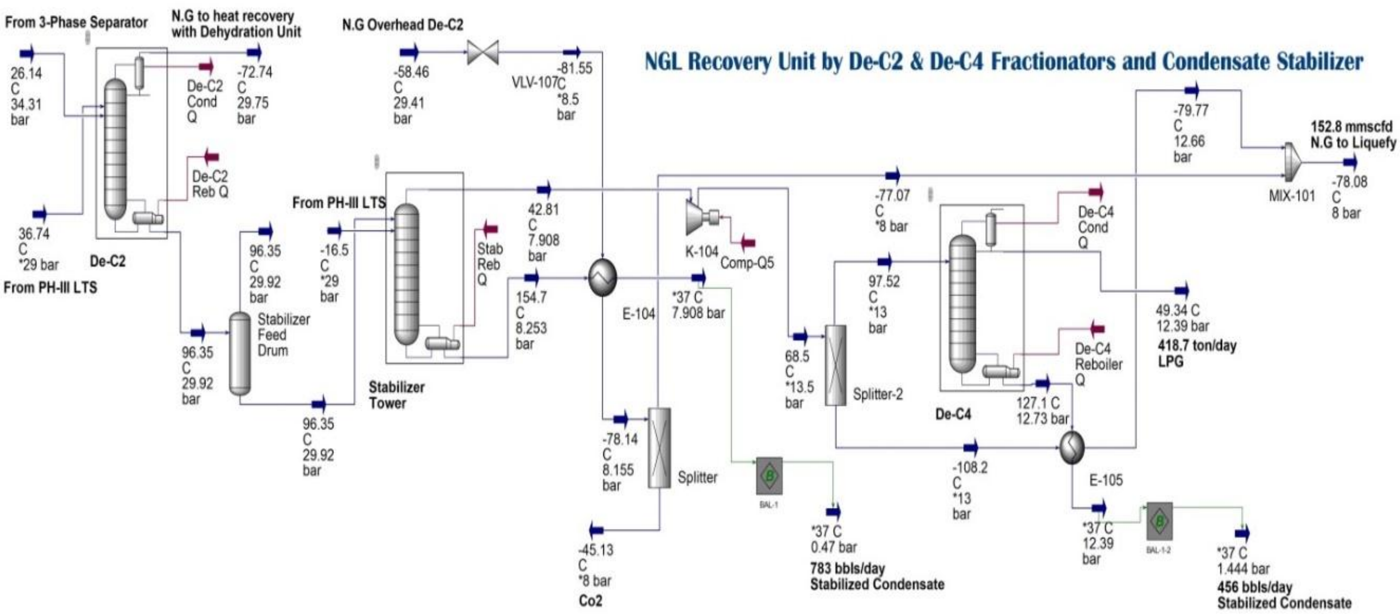

Figure 6. Flow diagram of NGLs recovery unit

There are some general trends common to the typical operation of distillation columns. By knowing temperature and pressure profiles which achieve our recovery target and energy conservation of the distillation process. That will be shown in figures of temperature and pressure profiles of stabilization tower, De-C2, and De-C4 towers and performance summary table 4.

Table 4. Towers Performance Summary

Items		Column		
		Stabilizer	De-Ethanizer	De-Butanizer
Number of Trays		14 sieve tray	28 sieve tray	24 sieve tray
Feed Inlet Tray		Top tray Tray 12	Tray 10 Tray 14	Tray 10
Condenser	Type	——	Full reflux	Total
	Temperature[C]	——	-72.74	49.34
	Pressure[bar]	——	29.75	12.39
	Duty [MMbtu/hr]	——	175.9	9.457
	Reflux Ratio	——	2.5	0.7802
	Reflux Flowrate [MMSCFD]	——	379.9	5.522
Re-boiler	Type	Regular	Regular	Regular
	Temperature[C]	154.7	96.35	127.1
	Pressure[bar]	8.253	29.92	12.73
	Duty [MMbtu/hr]	5.528	145.2	2.516
	Boilup Ratio	5.671	48.44	5.711
	Outlet Flowrate [MMSCFD]	0.7637	6.5	0.541
Product Recovery From Feed Stream [%]		12-100 C5+	100 C1 99.66 C2	100 C3 99.87 i-C4 97.92 n-C4

3.3 The Power and Energy

The energy duties for all heat exchangers are represented by Aspen Energy Analyzer in details as following Table 5 is shown;

Table 5. List of Existing Heat Exchangers Details of Modified NGLs Recovery Case

Heat Exchangers	Type	Base Duty [MBtu/hr]	Hot Inlet Temp [C]	Hot Outlet Temp [C]	Cold Inlet Temp [C]	Cold Outlet Temp [C]	Fluid Type
De-C2-Reboiler	Heater	145.2	125.0	124.0	89.5	96.3	LP Steam
De-C2-Condenser	Cooler	175.9	-36.6	-72.7	20.0	25.0	Cooling Water
Stabilizer-Reboiler	Heater	5.528	125.0	124.0	133.3	154.7	LP Steam
De-C4-Reboiler	Heater	2.515	125.0	124.0	125.9	127.1	LP Steam
De-C4-Condenser	Cooler	9.48	59.4	49.3	20.0	25.0	Cooling Water
E-104	Process Exchanger	0.9987	154.7	37.7	-81.5	-78.2	___
E-105	Process Exchanger	0.4354	127.1	37.5	-108.2	-80.0	___
Gas/Gas Ex	Process Exchanger	15.56	39.6	1.5	-19.6	25.8	___
Gas/Gas Ex-2	Process Exchanger	5.125	1.5	-10.0	-72.7	-58.5	___

4. RESEARCH METHODOLOGY

Today there are a few PC help process reproduction apparatuses dynamic available and regular in a procedure designer's day by day work gear. Various procedure programming exist available today where CHEMCAD, HYSYS, and Aspen Plus are a portion of the significant players available. Aspen HYSYS® programming was utilized to perform process recreations. The recreation was completed as a consistent state reenactment. The product gives stream data to mass and vitality streams that are used in the structure details and cost estimations and for different procedure hardware's, for example, distillations, heat exchangers, and refining segments

The simulation package used in this study is HYSYS V10 which is based on the Peng-Robinson equation of state for calculations. This study focuses on the simulation NGls train processes

The plan of this comparative study is constructed as in the following steps:

- Studying of El-Wastani LPG plant for Wasco company techniques. The determination of the best improvement route depends on an efficient and specialized investigation of these different routes of enhancements.
- Simulation and Studying of the impact of other elective factors, Determining of the optimum LPG recovery mode with the maximum capacity route.
- Determining the optimum recovery mode and comparing it with the increasing capacity route.
- Based on the results, write the pros and cons of each process and give recommendations for the most feasible techniques that should be used for maximization production of NGLs.
- Finally, after performing the cost estimation based on return on investment (ROI) for maximizing train profitability.

5. PROCESS DESIGN

5.1 Plant Feed Gas Chemical Analysis Composition

Natural Gas Feed Analysis composition of the components and process feed condition that is taken from Wasco network wells to process the feed rich gas stream wells, entering is represented in the table 6 as follow:

Table 6. Natural Gas Feed Analysis Composition Process Feed Condition

Component	Gas Mole%	Liquid Mole%
N2	0.04	0.00
CO2	0.32	0.05
C1	86.99	3.72
C2	7.46	2.29
C3	2.72	2.09
i-C4	0.81	2.57
n-C4	0.62	3.46
i-C5	0.30	5.12
n-C5	0.17	3.71
C6	0.19	10.47
C7	0.13	16.21
C8	0.06	19.72
C9	0.01	11.31
C10	0.00	6.80
C11	0.00	3.83
C12	0.00	2.57
C13	0.00	1.92
C14	0.00	1.27
C15	0.00	1.16
C16	0.00	0.58
C17	0.00	0.56
C18	0.00	0.30
C19	0.00	0.10
C20	0.00	0.07
C21	0.00	0.05
C22	0.00	0.03
C23	0.00	0.04
H_2O	0.18	0.00
Total	100.00	100.00
Mole Weight	19.13	107.2
Temperature [^{0}C]	38	38
Pressure [bar]	50.31	50.31
Flow rate [m^3/day]	88632	636

6. ECONOMIC ANALYSIS

6.1 Method Used

The economic assessment for this research is employed using Aspen Process Economic Analyzer (formerly known as Icarus Process Evaluator); Aspen Economic Evaluation is an integrated economics feature in Aspen HYSYS® that enables in their process modeling studies using Aspen HYSYSS, process engineers easily estimate the relative capital and operating costs. [16].

6.2 Equipment Sizing

Equipment size is determined by utilizing the yield information from Aspen Hysys Version 10 simulation program to get the correct detail of the concerned hardware, for example, a region with heat exchanger and force on account of a blower, and so forth.

For example, the exchanger zone is determined by utilizing the duty given by Aspen Hysys. The log mean temperature difference is determined for every re-kettle and condenser as follows:

Duty: Q (given from Aspen HYSYS)

$$\Delta t = LMTD = \Delta T1 - \Delta T2\ (\Delta T1 \Delta T2) \quad (1)$$

Where Δt is the average temperature difference, LMTD is the log mean temperature difference. ΔT1 is the temperature difference between the hot inlet stream and cold outlet steam and ΔT2 is the difference between the hot outlet stream and the cold inlet stream. Area (A) of the equipment (condenser or re-boiler) can be calculated from equation 2.

$$A = QK\Delta t \quad (2)$$

Where k is the thermal conductivity of the equipment material.

6.3 Cost Estimation

A gauge of the capital venture for a procedure may change from a pre-structure gauge dependent on little data to a point by point gauge arranged from complete drawings and details. These assessments are called by an assortment of names; however, five gauge classifications speak to the precision range and assignment typically utilized for configuration purposes. These five acclaimed gauges are significant degree, study, and starter, conclusive, and nitty-gritty evaluations. The exactness of these assessments are going from + or - 30 to + or - 5 percent individually [16].

6.4 Fixed Capital Cost

The fixed capital cost is estimated to get an approximate price for the total plant to be installed and running. In this present work, the calculations are based on a rule of thumb stated that the total fixed capital cost equals the purchased equipment cost multiplied by 2 [16].

Equipment costs are calculated according to cost estimation techniques where, the cost of the new equipment, Cn, is equal to the known equipment cost, Ck, times the ratio of the two plants' capacities raised to a fractional power as indicated in equation 3.

$$Cn = (VnVk)F \quad (3)$$

Where, Vn is the capacity of the new plant, and Vk is the capacity of the known plant. F is a factor; usually takes a value between 0.4 and 0.9, depending on the type of plant. In the use applied estimating technique, a factor value of 0.6 is used according to the literature or historical data [17]. In addition to use of Nelson-Farrar indexes, the most proper approximated calculation could be performed to determine the value of the relevant cost at the current time. This is done by the comparison with equipment prices which are obtained from El-Wastani Company [18].

6.5 Working Capital Cost

The working capital is the measure of capital required to fire up the plant and to fund a principal couple of long periods of working before the plant begins winning. This capital is utilized to cover compensations, crude material inventories, and possibilities. It will be recouped toward the finish of the task and speaks to a buoy of cash to kick the undertaking off. These expenses are important for new businesses and it infers crude materials and intermediates all the while. The working capital expense is thought to be 3% of the fixed capital expense [20].

6.6 Operating Cost

The operating cost includes all the incremental increase in cost due to chemical injection, power, treatment, and utility consumption which is needed to achieve the required target of increasing maximization NGLs and LNG production capacity.

6.7 Return on Investment (ROI)

Return on Investment (ROI) is a success metric that is used to determine an investment's effectiveness or to compare the effectiveness of many different investments. ROI aims to explicitly calculate the amount of return on a given investment compared to the expense of the investment. The profit (or return) of an investment is divided by the cost of the investment to determine the ROI. The result is expressed as a percentage or a ratio [20].

The ROI calculation is a straightforward one, and it can be calculated by using Eq. 4 [21].

$$ROI = \frac{Net\ Return\ on\ Investment}{Cost\ of\ Investment} * 100 \quad (4)$$

The total capital investment includes the fixed capital cost in addition to the working capital cost. The Pay-back period which is the period of time required for the return on an investment to "repay" the sum of the original investment can be calculated as presented in Eq. 5 [22]:

$$payback\ period = \frac{1}{ROI} \quad (5)$$

Notwithstanding the vitality proficiency, the decision of the procedure for a gas handling advancement coastal or seaward would be founded on the two primary boundaries of wellbeing and venture lifecycle cost. Capital expense in all cases would fundamentally impact the lifecycle costs. Lifecycle cost is primarily a function of capital cost and operating cost both of which increase as the number and size of equipment items increase. For the gas processes studied here, these costs are also affected by the complexity of the process arrangement, and its susceptibility to start-up, changes in the composition of the feed gas, and possible errors in thermodynamic modeling that involve higher margins of design.

A process capable of operating under conditions that will yield a profit must be presented as an acceptable plant design. Since annual net profit equals total income (products price) minus all expenses (total annual cost), as we know that the average current price of sales gas is 3 $/ MMbtu, 40 $/bbl condensate, and 850 $/ton LPG [23] [24] [25].

We can conclude from the gained revenue from original and new modified case tables the following table 7.

Table 7. Total Product Sales between Original and Modified NGLs Plants

Products, Unit	Sales Gas, MMscf		Stabilized Condensate, bbl		LPG , ton	
Simulation Base	Original	Modified	Original	Modified	Original	Modified
Quantity	157.059	152.8	5092	1247	204.4	418.7
Selling Price [$/unit]	3		40		850	
Daily Sales Price [$/Day]	491,079	477,653	203,680	49,880	173,740	355,895

The utility costs are taken from the aspen Hysys utility manager V10. The total utility consumption and generation between the Original and Modified LPG Plant are shown in table 8 and table 9.

Table 8. Utility Consumption in Original LPG Plant Case

Items	Fluid	Rate	Units	Rate Units	Cost per Hour	Cost Units
Electricity		479.48	KW	KW	36.248688	USD/H
Air	Air	13230730	BTU	BTU/H	0.013892	USD/H
HP Steam	Steam	17379860	BTU	BTU/H	54.920358	USD/H
MP Steam	Steam	6083012	BTU	BTU/H	12.835155	USD/H
Refrigerant 1	Propane	10640910	BTU	BTU/H	33.625276	USD/H
Refrigerant 1 Generation	Propane	410099.5	BTU	BTU/H	-1.295914	USD/H

Table 9. Utility Consumption in Modified LPG Plant Case

Items	Fluid	Rate	Units	Rate Units	Cost per Hour	Cost Units
Electricity		9370.2	KW	KW	539.723	USD/H
Cooling Water	Water	9.45958E+06	BTU	BTU/H	2.11895	USD/H
MP Steam	Steam	1.50748E+08	BTU	BTU/H	318.079	USD/H

The required capital investment and economic analysis is represented in the following Table 10.

Table 10. Compared Results of Economic Study for Original and Modified LPG Plant

Items	Simulation Base of LPG Plant	
	Original	Modified
Total Project Capital Cost, $	24,304,710	23,593,900
Total Operating Cost, $/Y	6,450,770	10,811,500
Total Utilities Cost, $/Y	2,783,426	7,538,070
Total Product Sales, $/Y	317,219,259	322,672,077
ROI (return on investment),%	13.05	13.68
Pay-Back Time ,Months	0.91	0.88

7. CONCLUSION

Proper integration of NGLs recovery innovation without refrigeration cycle results in noteworthy focal points by lowering overall capital cost requirements and improving NGLs production. Through careful process selection and heat integration, the integrated NGLs train results in lower specific power consumption and increased net profit as compared to each facility unit separately. By optimizing integration between NGLs recovery and dehydration unit achieve great focal points in heat recovery and maximizing improving LPG (increasing 49% from original case plant) sharing to solve the LPG shortage problem in Egypt. Achieve peak of profit from the moderate capacity feed by added value for 152.8 MMscfd sales gas, 418.7 ton/day LPG, and 1248 barrel/day production.

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