International Journal of Recent Technology and Engineering (IJRTE)
ISSN: 2277-3878, Volume-8 Issue-3, September 2019

NGL Recovery Enhancement for GUPCO Trans Gulf Gas Plant by using New Applicable Technique

Ahmed Fathy Abd El-Ghany, M. H. M. Hassanean, Nadia Ali El-Sayed

Abstract: ***It is known that the price of natural gas liquids (NGL) is higher than that of natural gas from which it is derived, so more modifications needed for existing plants to derive more NGL is economically accepted point of view. The main objective of the present work is to present the method applied on Trans gulf (T/G) gas plant to overcome its performance decrease happened after the plant feed gases becoming leaner than its design margin and hence it led to a great decrease in the plant NGL recovery. This achieved by introducing a new simple modification to the existing process scheme obtained by using a condensate stream to enrich the reflux of the de-ethanizer tower so more recovery is obtained. In order to accomplish that goal, some changes in the existing process operating conditions were needed. A simulation is used in this study to examine the existing and the introduced modification utilizing ASPEN-HYSYS software version 8.4 using Peng-Robinson equation of state (EOS). The simulation of the existing plant results in a better understanding of the plant behavior in the different iterations to reach the maximum benefits.***

The plant after suffering from low butane recovery from its feed gas and which considered as a figure to the plant efficiency, it increased by this method from 38 % to reach 86-90 % butane recovery and its LPG production increased by 170% to be ≈ 122 tonne/day instead of ≈ 44 tonne/day while only losing ≈ 16 tonne/day of condensate production. An optimization to the new method is done in this paper so that it doesn't intercept with the existing plant equipment performance for the process safety triggers. Also, the last section of the study describes the economic point of view and the return on investment (ROI) how it was paid back only in 7 days. This modification can be taken as a guideline for both new and existing LPG plants which use only propane refrigeration systems for LPG recovery to increase their profits with the lowest cost possible.

Keywords: Natural Gas, LPG, NGL, Condensate, reflux ratio

I. INTRODUCTION

Recovery from natural gas streams of light hydrocarbon liquids can range from easy control of the dew point to profound ethane recovery. The required degree of liquid recovery has a deep impact on the processing facility's process choice, complexity, and price. The word NGL (natural gas liquids) is a general word that applies to natural gas-recovered liquids and relates to ethane and heavier products as such. The word LPG (liquefied oil gas) defines mixtures of hydrocarbons in which propane, iso, and normal butane, propane, and butane are the primary elements. Typically, olefins are not present in LPG in the manufacturing of natural gas.

The gas composition has a significant effect on the recovery and process choice of natural gas liquid economies.

In general gas with a larger amount of liquefiable hydrocarbon products is a larger product and thus higher profits for the gas processing plant. Richer gas also involves greater cooling tasks, greater heat exchange surfaces and greater capital costs for specified recovery effectiveness. In general, leaner gasses involve more serious handling circumstances (low temperature) to attain handling conditions (low temperatures) for elevated retrieval efficiencies. [1]

The present work aims to increase the production of liquefied petroleum gas unit (LPG) with little decrease in the production of condensate to contribute to the total NGL recovery of the plant by simple modification which is achieved by adding new recycle line to the overhead product of the De-Ethanizer Column and also not to exceed the gas specification limitations to the downstream gas compressors for the offshore platform (October complex)

II. OVERVIEW OF T/G PROCESSING

Referring to the simplified overall process flow diagram (PFD) in Fig.1 Feed gases are combined and compressed in the inlet gas compressors (4 turbine gas Compressors) up to 24 -26 KG/CM², then the gas is dehydrated in the Molecular sieve drying package. The dehydrated gases then cooled in two gas/gas heat exchangers (40-E-1 A/B & 30-E-2 A/B). The cooled gases then enter the fractionation system (the De-Ethanizer and De- Butanizer towers) Where LPG and Condensate are recovered & Methane, Ethane, and excess propane are stripped as Sales Gas. The sales gas is compressed by (4 gas turbine compressors) up to 80 KG/CM² then directed to October gas lift compressors for Well injection and the excess sales gas is *directed to the Gas National Grid (GASCO Network).*

Manuscript published on 30 September 2019
* Correspondence Author
Ahmed Fathy *, Chemical Engineering and petroleum refining department, Suez University, North Sinai, Egypt. Email: ahmed.fathy.07@gmail.com.
M. H. M. Hassanean, Chemical Engineering and Petroleum Refining department, Suez University, Suez, Egypt. Email: mhassanean@yahoo.com.
Nadia Ali, , Chemical Engineering and Petroleum Refining department, Suez University, Suez, Egypt. Email: nadiaali887@gmail.com.

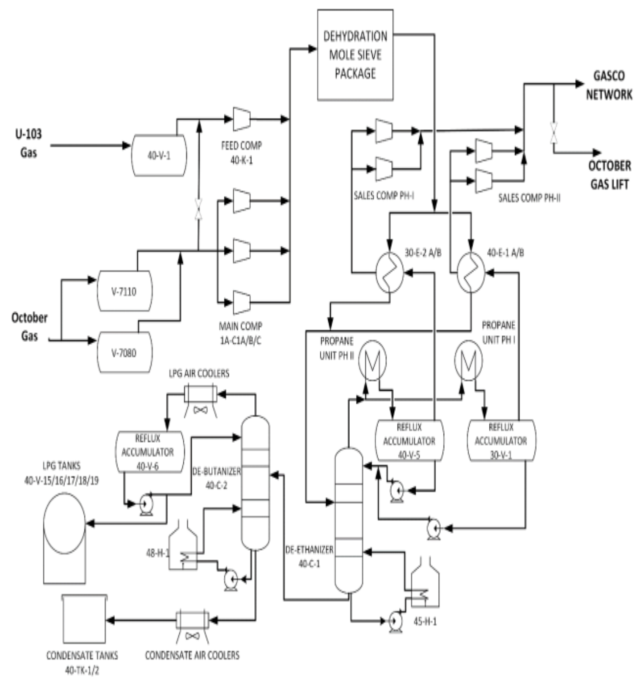

Fig. 1.Trans-gulf gas plant process flow diagram

III. DESIGN BASIS

A. Feed

The feed to the plant was as follows

- Feed gas from both the October production platform and sales gas from Sinai LPG Plants.
- Feed Gas from unit-103

The following table Table-I shows the Feed gases composition of the design basis [2]:

Table- I: Design Feed Gases Specification

Component mole %	OCTOBER		U-103
	Sinai	*October*	
CO_2	0.98	3.43	3.64
N_2	0.58	0.88	0.46
C_1	67.78	58.2	56.95
C_2	19.49	14.67	15.09
C_3	10.31	12.27	12.84
IC_4	0.38	1.8	1.99
NC_4	0.43	4.55	5.34
IC_5	0.03	1.28	1.42
NC_5	0.02	1.46	1.37
C_{6+}	0	1.46	0.9
WATER CONTENT (PPM)	105	SATURATED	SATURATED
H_2S (PPM)	11	10	21
RICH CASE (MMSCFD)	0	100	30
LEAN CASE (MMSCFD)	35	65	30

Total recovery process Depends on one column (De-Ethanizer) which responsible for recovery LPG components from the feed gases and stripping the ethane & lighters components plus the excess propane which provided with OVHD condenser Using (Propane Refrigeration) and Bottom Re-boiling Using a Fired Heater (45-H-1).

The plant facilities are designed to reach Almost 99% of the butane recovery. Several feed tray locations are provided for cold feed gas to insure good column performance for different feed composition. [2]

B. Products

The plant was designed to produce the following:

1. LPG
2. Sales gas
3. Condensate

LPG and condensate products Specs. Are as Per EGPC Specs and the sales gas specs is shown by Table-II, [2]

Table- II: Sales Gas Specification

Component mole %	Rich	Lean
C_1	69.03	68.973
C_2	17.5	18.24
C_3	8.358	8.7442
I-C_4	0.031	0.0243
N-C_4	0.0073	0.00427
CO_2	4.1375	3.21
N_2	0.9344	0.801
Design Rate, kg mole/hr.	5422.95	5671.243
MMSCFD	108.88	113.87
Molecular Weight	22.1277	22.06
Temperature , ^{0}C	55	55
Pressure , kg/cm^2g	90	90

C. Butane Recovery

The plant was designed to recovery 95% of the feed gas butane's in the LPG product as a guaranteed figure for both RICH and LEAN cases

The feed gas treated as the total Feed Gas After inlet compression (at gas molecular sieve inlet) [2]

D. Cooling

Air is used for cooling process streams

E. Storage

- *LPG storage*: Five LPG storage spheres are provided with an approximate storage capacity of 3600 cubic meters for each [2]
- *Condensate storage:* Condensate will be stored in two storage tanks with 28,000 BBL capacity for both. [2]

F. Process design considerations

- *Plant Maximum feed capacity* is 130 MMSCFD
- *Propane Refrigeration:* Propane refrigeration is provided to supply the necessary cooling and reflux generation in the De-ethanizer column. [2]

Retrieval Number C4884098319/2019©BEIESP
DOI: 10.35940/ijrte.C4884.098319

Published By:
Blue Eyes Intelligence Engineering
& Sciences Publication

- *Recovery of butane's from the compressed Dry Gas*: Is achieved in the top section of the De-ethanizer column whereby the gas is counter currently contacted with a propane rich liquid stream from the overhead reflux drum. [2]
- *Turn Down*: The plant turn down is 4:1
- *Dry bulb temperatures*

Normal Average (summer) 38.0°C

Maximum 43.2°C

Minimum 3.8°C [2]

G. Plant Outputs 'Rate:

From the tables above mentioned, it's presumed that the plant outputs should've been as follows:

- *Sales Gas:* 108-110MMSCFD with 22 M.wt.
- *LPG Product Rate:* 951 Ton/Day for Rich Case & 747 Ton/Day for Lean Case.
- *Condensate Product Rate:* 365 Ton/Day for Rich Case & 322 Ton/Day for Lean Case [2]

IV. CURRENT SITUATION

Two main deviations happened from the design basis which are:

- October complex Introducing Gas-Lift Method for oil recovery by using T/G sales gas and sending it back after injecting into the wells so the gas went leaner than the design case
- The low gas feed flow from U-103 due to well depletions.

Table-III represents Current Average Specification and condition for T/G Feed Gases are as follows:

Table- III: Current Average Feed Gases Specifications

Components	October Mole %	U-103 Mole %
N2	0.402	0.59
C1	79.774	74.33
CO2	2.289	1.7
C2	11.081	11.59
C3	4.185	6.66
IC4	0.641	0.94
NC4	0.554	2.07
IC5	0.298	0.8
NC5	0.32	0.56
C6+	0.456	0.75
MMSCFD	65	11
H2S	5	9
H2O	Saturated	Saturated
Molecular Weight	20.63	22.64
Specific Gravity	0.7122	0.7818

The table shows the feed compositions major change from the design basis and how the gas became leaner than the case presumed by the Plant vendor by which the plant can run efficiently to achieve 95+% of total butanes' recovery.

The gradual change in gas specifications led to the search for new applicable modification to the plant operation to defer the efficiency decrease as possible.

V. CONDENSATE RECYCLE TECHNIQUE

The modification was first tested in 2005 and finished by 2018 when the loop completed by adding a Spare pump (40-P-15 A/B) as shown in the following Fig.2.

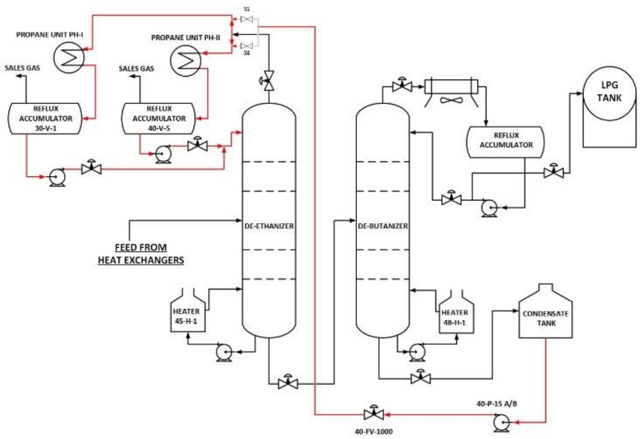

Fig. 2.Condensate Recycle Process Flow Diagram

A. Process Description

Condensate stream is pumped back from Condensate Tanks by 40-P-15 A/B pumps to the flow control valve 40-FV-1000 which controls the quantity sent to the tower.

The stream then split into two-stream with each stream of the De-Ethanizer OVHD gas streams before entering the Propane Units for Chilling.

The streams then are sent back to the reflux accumulators 40-V-5 & 30-V-1 which are low-temperature separators where the Reflux is recovered and the streams are separated into:

a- Enriched reflux with Condensate which is pumped back to the De-ethanizer column.

b- Sales gas out is sent to sales gas compressors Phase I&II where they are compressed and directed to the Gas-Lift system.

The Following HYSYS (Fig.3&4) represent the process scheme and the product Specifications before and after circulating the condensate stream which shows the results of adding the modified loop to the process, the model results are then compared with actual design and production data which shows that the model provides a good match with both design and history production data and so the modified process will be carried on this static model as the best representation of the actual plant

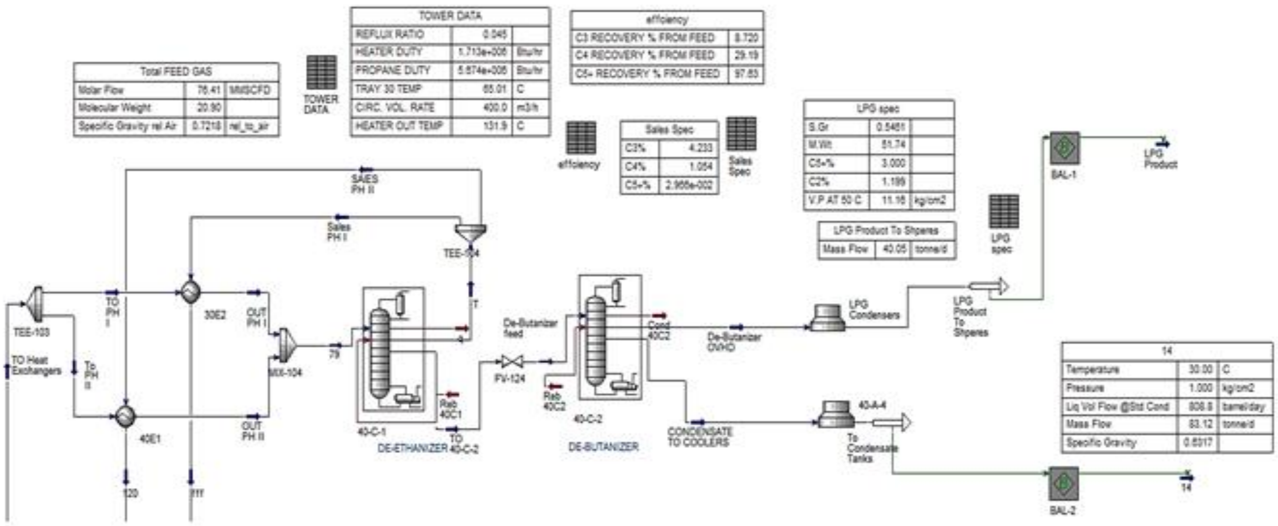

Fig. 3.Towers Performance before Modification

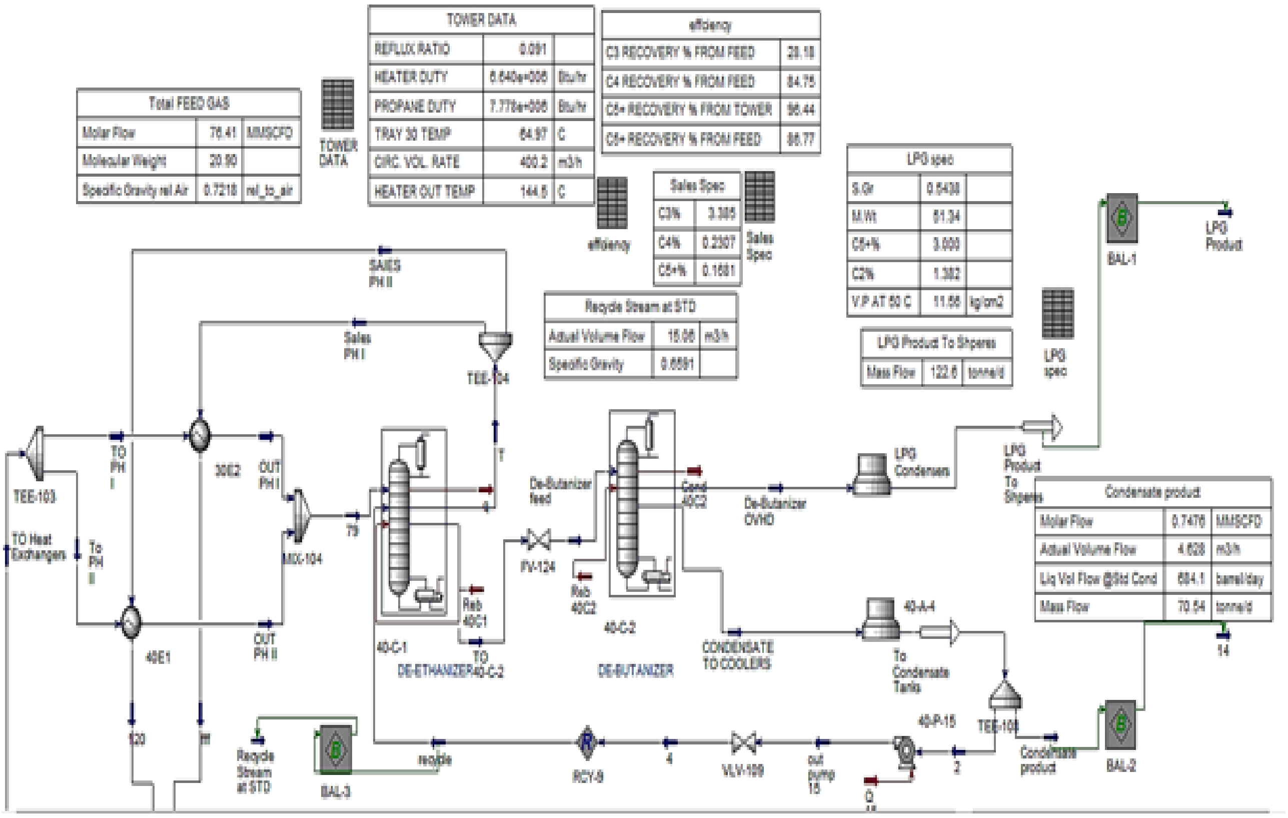

Fig. 4. Towers Performance before Modification

As seen by diagrams, Recycling 15.0 M3/hr. Condensate as shown the results are shown in (Table-IV).

According to the previous table, recycling Condensate Stream with 0.66 Sp.gr that the total NGL recovered increased by 70 Ton/Day.

B. Condensate recycle flow rate effect on Production rate

While putting the LPG product specification with the winter as it has the lowest C5+% recommendation Specs as demanded by EGPC (See Table-V).

Table- IV: Plant performance before & after recycling 15.0 M3/hr. Condensate

Criteria	Without condensate Recycle	With Condensate Recycle
De-Ethanizer Reflux Ratio	0.045	0.91
Propane Chillers Duties (BTU/Hr.)	5.6 x 10^6	7.7 x 10^6
De-ethanizer Heater Duty (BTU/Hr.)	1.7 x 10^6	6.63 x 10^6
Propane Recovery from feed	8.72%	28.00%
Butanes Recovery from the feed	29.19%	84.80%
LPG Product (TON/DAY)	40	122.9
Condensate Product (Bbl. Day/TON. Day)	808/83.12	685.4/70.67
Sales Gas (MMSCFD) / HHV (BTU/SCF)	74.46/1149	73.26/1123
Total NGL (TON)	123.12	193.4

Table- V: LPG Specifications [2]

Property	EGPC LPG Recommended Spec.			ASTM	T/G plant
	winter	*Spring/Autumn*	*Summer*		
C2%(Maximum)	5	5	5		1.4
C5% (Maximum)	3	5	10		3
Relative denisty@60/60 $_o$F (Optional)	Record	record	Record	1657	0.5438
Vapor Pressure@ 50 $_o$C, Kg/Cm2	12.8	11.5	11	1267	11.64
Volatility @ Evaporation of 95% by Volume	4	8	17	1837	NA**
Corrosion degree (Copper bar test)	1	1	1	1838	NA**
Mercaptan %					
(Minimum)	0.003	0.003	0.003	IP	0.003
(Maximum)	0.008	0.008	0.008	272	0.008
H2S%	-	-	-	2420	NIL
Total calorific value, kcal/kg(minimum)	11,800	11,800	11,800	3,588	11,000
**** Not available					

Fig.5. represents the condensate Recycle flow rate increase effect on the production rate of the plant without any other consideration which shows the production increase with the volume flow rate of the new stream.

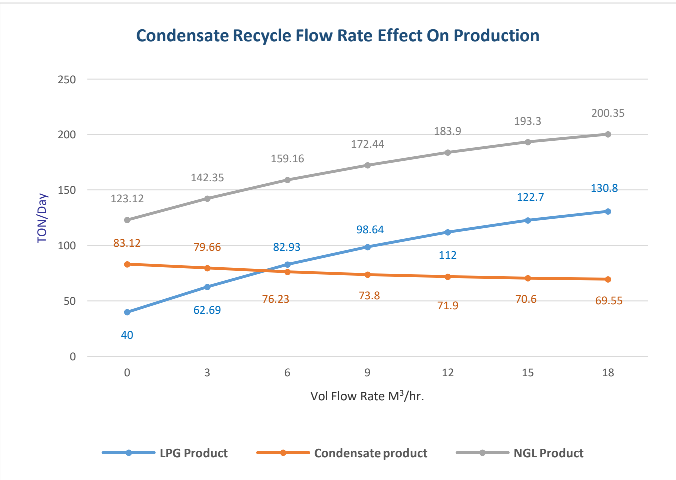

Fig.5. condensate recycle flow rate effect on Production rate

C. The new technique's obstacles

As the limitations found by recycling condensate are the heater & chillers duties increase, fortunately, the current heater and chillers duties available are higher than required by the de-ethanizer column to run with the recycling condensate (see Table-VI) , but the current obstacle is reaching the maximum design temperature for the outlet stream of the de-ethanizer heater (See Table-VII).

Retrieval Number C4884098319/2019©BEIESP
DOI: 10.35940/ijrte.C4884.098319
Journal Website: www.ijrte.org

Published By:
Blue Eyes Intelligence Engineering
& Sciences Publication

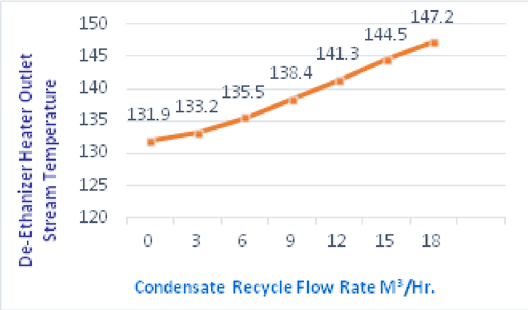

Fig.6. condensate recycle flow rate effect on bottom heater outlet temperature

Fig. 6 shows that increasing the flow rate of Condensate recycle raises the outlet temperature of the outlet stream from the de-ethanizer bottom heater due to the increasing density of the boil-up stream resulting from the increasing condensate in the tower. According to the last feed gas condition and specification, we concluded that the maximum condensate recycle flow rate can be used is <15 M3/Hr. to ensure <145 ^{o}C of the heater stream outlet temperature high limit, this while using 400 M3/hr. of a boil-up stream to the heater.

Table-VI : Equipment Duties

Equipment	Duty MMkcal/Hr.	Duty MMBTU/Hr.
De-ethanizer Bottom Heater	5.9	23.40
Propane Chiller Ph. I	4.12	16.34
Propane Chiller Ph. II	4.74	18.80
De-butanizer Bottom Heater	5.17	20.50
De-butanizer Condensers	6.34	25.14
Condensate Coolers	1.37	5.43

Table-VII: De-ethanizer Bottom heater Process Design Condition [2]

Equipment	
Item Tag No.	45-H-1
Duty	5.9 MMkcal/Hr.
Inlet Temperature	94 ^{o}C
Outlet Temperature	115 ^{o}C
Operating Pressure	28 Kg/Cm^2
Design Pressure	31 Kg/Cm^2
Design Temperature	145 ^{o}C

D. The relation between Condensate recycle and Feed gas condition

Table-VIII represents a case in which the feed gas analysis became richer than the previous case which shows that when the feed gas goes richer how that can affect the efficiency of the plant and the condensate recycle criteria.

The following Table-IX shows the different results before and after recycling 18 M3/Hr. of condensate stream:

Table- VIII: Rich Case Feed Gas Specification

Components	October Mole %	U-103 Mole %
N2	0.585	0.424
C1	77.938	68.559
CO2	2.213	2.219
C2	13.201	12.873
C3	3.915	8.849
IC4	0.290	1.276
NC4	0.598	2.942
IC5	0.338	0.990
NC5	0.407	0.977
C6+	0.515	0.891
MMSCFD	66.8	15
H2S	4	8
H2O	Saturated	Saturated
Molecular Weight	20.84	24.510
Specific Gravity	0.717	0.846

The results in Table-IX indicate the HYSYS case results shown by Fig.8&9 and shows that when the feed molar

Table-IX: Plant performance before & after recycling 18.0 M3/hr. Condensate

Criteria	Without condensate Recycle	With Condensate Recycle
De-Ethanizer Reflux Ratio	0.051	0.10
Propane Chillers Duties (BTU/Hr.)	6.21×10^6	8.74×10^6
De-ethanizer Heater Duty (BTU/Hr.)	2.53×10^6	8.39×10^6
Propane Recovery from the feed	11.37%	31.8 %
Butanes Recovery from the feed	38.3 %	92.16%
LPG Product (TON/DAY)	59	152.8
Condensate Product (Bbl. Day/TON. Day)	1119/114.8	964.3/99
Sales Gas (MMSCFD) / HHV (BTU/SCF)	79.61/1163	78.24/1136
Total NGL (TON)	173.8	251.8

flow raised to 4096 kgmole/hr. Instead of 3806, so this made the 18 M3/hr. of condensate recycle lead to almost the same heater outlet stream temperature as the 15 M3/hr. effect shown in the 1st case (See Fig.7)

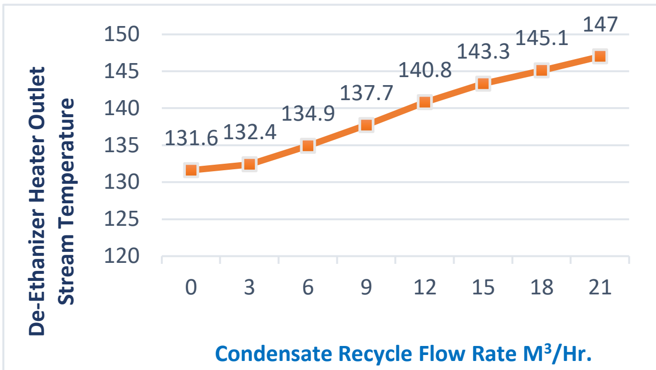

Fig. 7. Rich case Condensate recycle flow rate effect on heater outlet temperature.

Retrieval Number C4884098319/2019©BEIESP
DOI: 10.35940/ijrte.C4884.098319

Published By:
Blue Eyes Intelligence Engineering
& Sciences Publication

International Journal of Recent Technology and Engineering (IJRTE)
ISSN: 2277-3878, Volume-8 Issue-3, September 2019

This means the flow of condensate recycle can't be fixed value, and it needs to be changed according to the feed gas condition (Rich/ Lean) to reach the maximum obtainable total recovery efficiency of the plant.

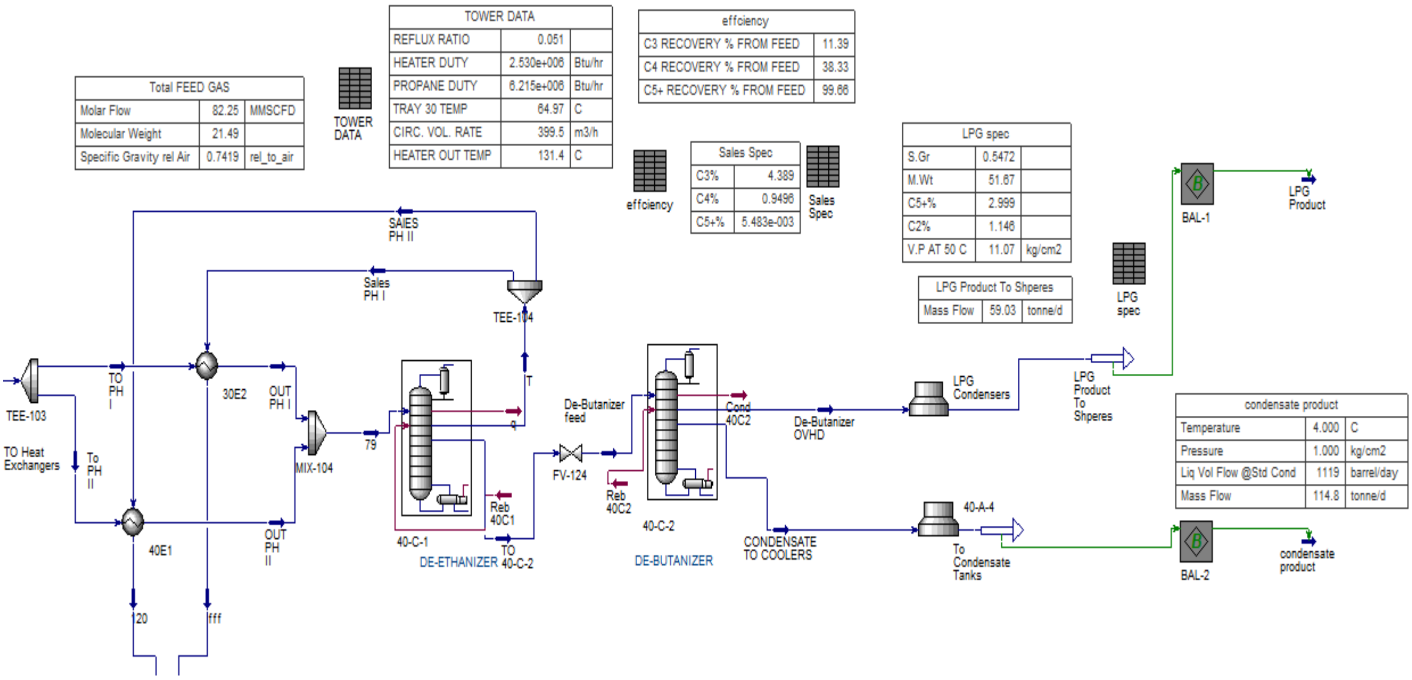

Fig. 8. Rich Case Towers Performance without Condensate Recycle

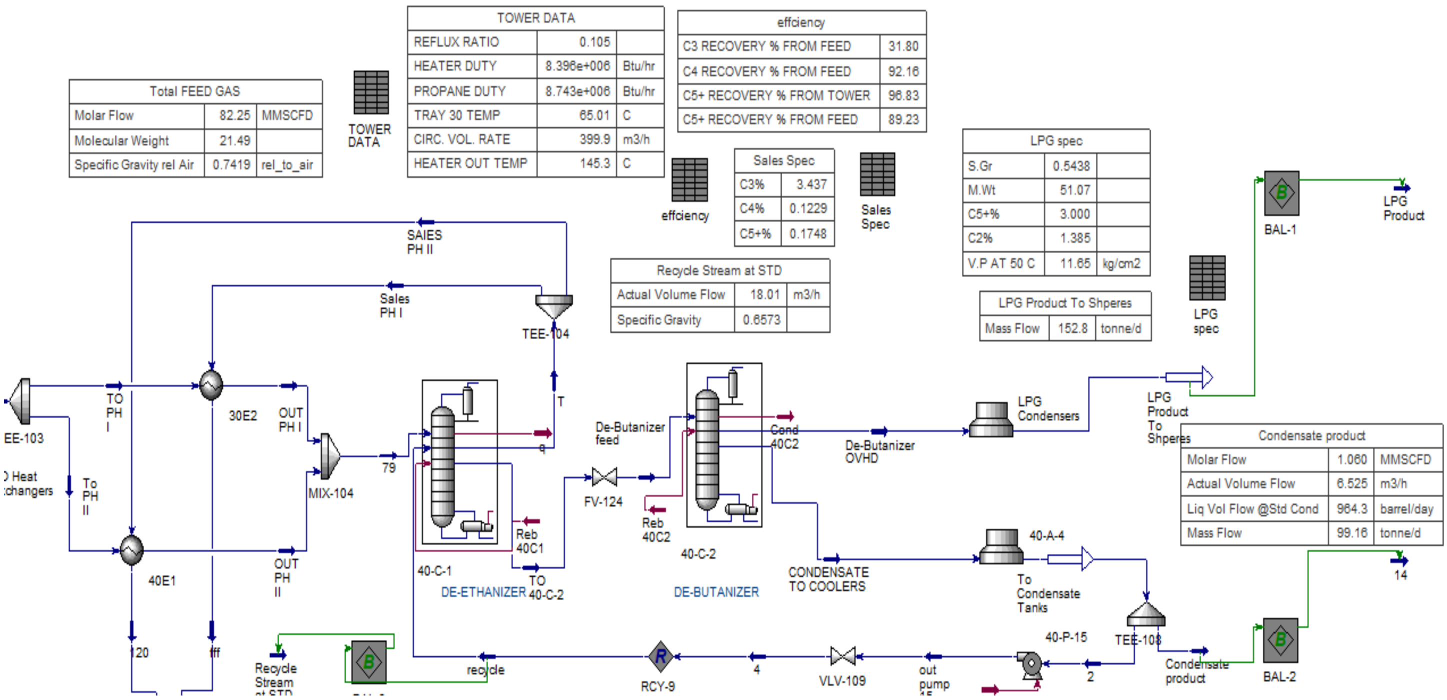

Fig. 9. Rich Case Towers Performance with Condensate Recycling

Retrieval Number C4884098319/2019©BEIESP
DOI: 10.35940/ijrte.C4884.098319
Journal Website: www.ijrte.org

Published By:
Blue Eyes Intelligence Engineering
& Sciences Publication

VI. ECONOMIC EVALUATION

The Total Capital and operating costs until start-up included

- Direct costs which refer to purchased equipment costs and labor costs for equipment installation, piping, civil, structural steel, instrumentation and controls, electrical equipment and material paint
- Indirect field costs such as engineering and supervision, start-up and commissioning, construction expenses, fringe benefits, burdens, insurance, scaffolding, equipment rental, field services, temporary constructions, etc.
- Indirect non-field costs as described below:

a) Freight, taxes and permits, engineering (Basic engineering, detailed engineering, and material procurement).

b) Contingency – allowances for unpredictable events.

c) Other project costs such as general and administrative expenses, contract fees, and home office expenses

As most of job done to install and run the new modification was carried out by our departments and crew, so most of the cost went for the material purchasing and contractions to be

- Direct costs: 54,000 $ in 2005
- 204,000 $ in 2018
- In-Direct Costs: 7,000 $ in 2018

The total Capital investment = 265,000 $ by 2018

The major equipment required is:

- 2 motor-driven centrifugal pumps (pump is spared)
- 5 manual valves (globe & ball)
- 1 flow control valve
- 3” Sch 80 Carbon Steel Piping

B. ROI & Payback Period

The simulated gross heating values for the produced gas is decreased from 1149 BTU/ MMSCF for the original plant to 1023 BTU/MMSCF for the modified plant. This can be attributed to the reduction of heavier hydrocarbons in the gas stream produced from the de-ethanizer in the modified plant which is recovered as LPG in the debutanizer. The gas selling price is 2.65 $/MMBTU which is taken from Egyptian natural gas holding company.

Table- X shows the production rates of LPG, condensate, and gas as well as the revenues from original and modified plants for each train. So the added revenue to the plant after modification equals to 48,668 $/ Day

Table- XI shows the total operating costs calculated by subtracting the power consumption before and after the modification to be = 1135.79 $/Day

ROI= (Total income increment – Total increase in operating Costs) /Total Capital Investment

= (48668-1135.79)/265,000

=0.179

Payback Period in days = 1/ROI

= 5.577 Days

Table- X: Revenues from original and modified Plant

Products	Production		Selling price, $/unit	Total sales price, $	
	Original	*Modified Plant*		*Original*	*Modified Plant*
Sales gas (MMBTU)	85,554	82,259	2.65 $/ MMBTU	226,718	217,986
Stabilized condensate (BBL)	808	684	70 $/ Bbl.	56560	47880
LPG (TON)	40	122.6	800/TON	32000	98080
Daily Revenues $				315,278	363,946

Table- XI: Power consumption of original and modified gas

Equipment type	Power consumption (MMBTU/D)		Selling price, $/unit	Operating Costs $/Day	
	original	*modified plant*		*original*	*modified plant*
De-ethanizer heater	40.8	159.12			
Propane Units	134.4	184.8			
De-butanizer heater	40.8	132			
LPG condensers	50.4	165.6			
Condensate coolers	18.96	72			
Condensate Recycle Pump	0	0.1056			
Total duties for both towers (MMBTU/Day)	285.36	713.625	2.65 $	756.2	1891.1

VII. CONCLUSION

All results obtained from modeling the existing LPG plant and the modified plant using "Aspen HYSYS" process simulation program

Referring to previous tables, the following will be conducts

- The current LPG production rate figure was increased using the modified plant model as the butane recovery was increased from ≈30 to ≈85% for the current average gas feed analysis and it can be more in case of rich gas feed case.
- Also, the condensate product decreased by ≈120 BBL/Day which is ≈13 Ton/day but to gain at least 80 Ton/Day of LPG product which makes the total NGL product gained at least 57 Ton/day.
- The feed gas condition changes daily depending on the current production wells.
- The Total NGL product increase depends on the gas feed specification as the richer feed leads to more NGL gain by the new modification without passing the current equipment design criteria.

- From the economic point of view, new modification has a rapid return on investment (ROI) as the total capital investment (TCI) will be paid back within 7 days maximum if any process troubles encountered and within 6 days if normal operation Running. Also, it has a high net profit compared with the existing LPG plant.

VIII. RECOMMENDATION

Hence, we recommend replacing the current de-ethanizer bottom fired heater pumps (Boil-Up pumps) to other ones have more rated capacity, to lower the heater outlet temperature so we can increase the flow rate of the condensate recycle to obtain more product and profit as well.

REFERENCES

1. Gas Processors supply Association, "Engineering Data Book", 12th edition, Sec 16, Hydrocarbon Recovery, Tuls, OK,2004
2. J.M. Campbell, Gas Conditioning and processing: the basic principle, 7th. 1, Campbell Petroleum Series, Norman, OK, 1992.
3. J.M. Campbell, Gas Conditioning and processing: the basic principle, 7th. 2, Campbell Petroleum Series, Norman, OK, 1992.
4. Egyptian General petroleum Corporation plant "operating & start-up guide for Trans gulf LPG expansion project", Ras Bakr Area, ENPPI, Project No. 2560-200, May 1998.
5. Ahmed Abd El-Kader Bhran, Mohamed Hassan Hassanean, Mohamed Galal Helal. "Maximization of natural gas liquids production from an existing gas plant", Egyptian Journal of Petroleum, 2016.
6. Egyptian Natural Gas holding Company Website: www.egas.com.eg.
7. Aspen Tech., ''Advanced Process Modeling Using HYSYS", 2009.
8. Aspen tech., ''a user guide manual for Aspen physical property V 7.1", 2009

AUTHORS PROFILE

Ahmed F. Abd El-Ghany was born in Egypt. He got his BSc in Chemical Engineering and Petroleum Refining from the Faculty of Petroleum and Mining Engineering (Suez University), Suez, Egypt in 2011. He is working in GUPCO Petroleum Company in Egypt as a Process Engineer. He worked in ALMANSOORI SPECILAZIED Eng. For Oil Production Services. He has some courses petroleum process engineering such as process plant design, surface production operation, pre-commissioning & commissioning start up and some safety courses such as OSHA, Process Safety Management and HAZOP Besides simulation software courses such as HYSYS (Static and Dynamic).

.

M.H.M.hassanean was born in Egypt**,** he has a PhD in Petroleum Engineering from Faculty of Engineering, Cairo University, Egypt in 1990.
He is working as a professor of chemical Engineering in Suez University, Faculty of Petroleum & Mining Engineering, in chemical Engineering and Petroleum Refinery Department.
He is teaching Natural gas processing, Production of natural gas, Conversion process, Unit Operations.
He has supervised on several M.Sc. and PhD Theses, and He is the author of over 20 technical papers in oil & gas.

Nadia Ali was born in Egypt. She got his BSc in Chemical Engineering and Petroleum Refining from the Faculty of Petroleum and Mining Engineering (Suez University), Suez, Egypt. She has PhD in the refinery engineering from Suez University, specialist in the lube oil treatment. She is a faculty member in faculty of petroleum and mining engineering, chemical engineering and petroleum refining department. She has taught several courses in chemical engineering such as specifications of petroleum products and test methods, petroleum refinery engineering, plant design and pollution control. She supervised many scientific theses such as enhancing produced water in oil fields using alternative treatment technologies, flared gas recovery and optimum operating conditions of gas dehydration.