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Process simulation and performance improving of a gas plant in operation

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Abstract

In the first part of this study, the natural gas liquids (NGLs) recovery enhancements techniques particularly by capacity increasing was studied to maximize the plant (El-Wastani Petroleum Company located in Egypt) productivity and profitability. The present work studies new NGL recover enhancements routes based on the recovery maximization of butane, propane, or ethane as a final product for sales. The selection of the best improvement route is based on an economical and technical study of these various routes of improvements. This selection is done using fuzzy logic as an intelligent system for the process of decision making. According to the results of fuzzy logic system, it is noted that in spite of the capacity increasing mode have the highest return on investment (ROI), it is found that the maximization of propane recovery is the optimum route for the plant improvements. This is because the choice of propane recovery mode does not rely only on ROI but takes in consideration various objectives such as feed stability, marketing availability and recovered NGL quantity. The results of this research wok is beneficial for improving the NGL recovery plants in operation as well as the newly ones which should take in consideration these improvements modes before the plant design.

Keywords

Natural gas liquids; NGL recovery improvements; Fuzzy logic.

Introduction

The retrofitting of an existing NGL/LPG (liquefied petroleum gases) processing plant to use the latest technology allows expanding the plant capacity as discussed in details in an earlier research study [1]. Furthermore, this retrofitting allows correcting the limitations and deficiencies inherent in first and second generation process designs at the same time. Generally, this means that plant operations are more stable after the retrofit, making the plant easier to operate and optimize. Third generation NGL/LPG technology can be retrofit into nearly any existing first generation and second generation expander plant. It is proven that most of the existing plants can gain the advantages of the third generation technology, so that capacity is expanded, recovery efficiency is improved, plant operations are simplified, and recurring operating problems are eliminated [2]. A retrofit for increased ethane, propane or butane recovery alone is generally much simpler than one for both increased recovery and throughput because equipment sizing and pressure drop are less of a constraint. However, the economics of most NGL/LPG plants are improved if increases in both recovery and throughput can be achieved [3].

Recovery of NGLs components in gas not only may be required for hydrocarbon dew point control in a natural gas stream, but also yields a source of revenue, e.g., natural gas may include up to about fifty percent by volume of heavier hydrocarbons recovered as NGL. When designing a new plant or upgrading an existing plant it is very important to understand and select the right process to minimize capital and operating expenses. The process selection, complexity, and cost of the processing facility depend on the gas composition, inlet pressure to the gas plant, recovered product specifications and the extent of recovery that is desired. When market demand for ethane, propane, or propane is high, high recovery of ethane, propane or butane needs to be achieved in the gas plants respectively [4, 5].

Lower processing temperatures for higher recoveries of NGLs lead to increasingly complex and expensive refrigeration techniques. Numerous expansion processes are commonly used for hydrocarbon liquids recovery in the gas processing industry, particularly in the recovery of ethane and propane from high pressure feed gas. Recovering 90% and higher of the propane and over 70% of the ethane in natural gas streams typically requires temperatures in the – 125 to – 150 ^{o}F region and lower [6, 7]. NGL recovering processes from Natural Gas are generally based on one of the following or a combination of them: use of external refrigeration, turbo-expansion, Joule-Thompson-expansion, or absorption. As previously referred, each specific NGL recovery problem is characterized by a large number of variables, some of those include inlet pressure, residual gas and the top of the demethanizer column, duty of external refrigeration, power capacity of the turbo-expansion system, temperature and pressure of the cold tank and of the bottom of the demethanizer column [8]. Turbo-expanders are favored in applications where the pressure drop required for process operation is readily available of where recompression costs are of secondary economic importance [6]. Most of the gas plants in operation today use conventional single-stage turbo-expander technology for moderately high ethane recovery [9].

Five major NGLs – ethane, butane, iso-butane, propane and natural gasoline – are used by petrochemical companies as feed stocks and by refineries as blending and processing components. In areas where not enough natural gas processing capacity is available, gas production must be curtailed down to the available capacity but the process could be directed to increase its benefits through increasing the amount and quality of NGLs [10].

El Wastani plant located in Damietta, Egypt, is the plant on which the present study is conducted. El-Wastani Petroleum Company (see Figure 1) operates wells network in order to process the feed gas stream to deliver sales gas, stabilized condensate, and liquefied petroleum gas (LPG) via central processing facilities (CPF) plant.

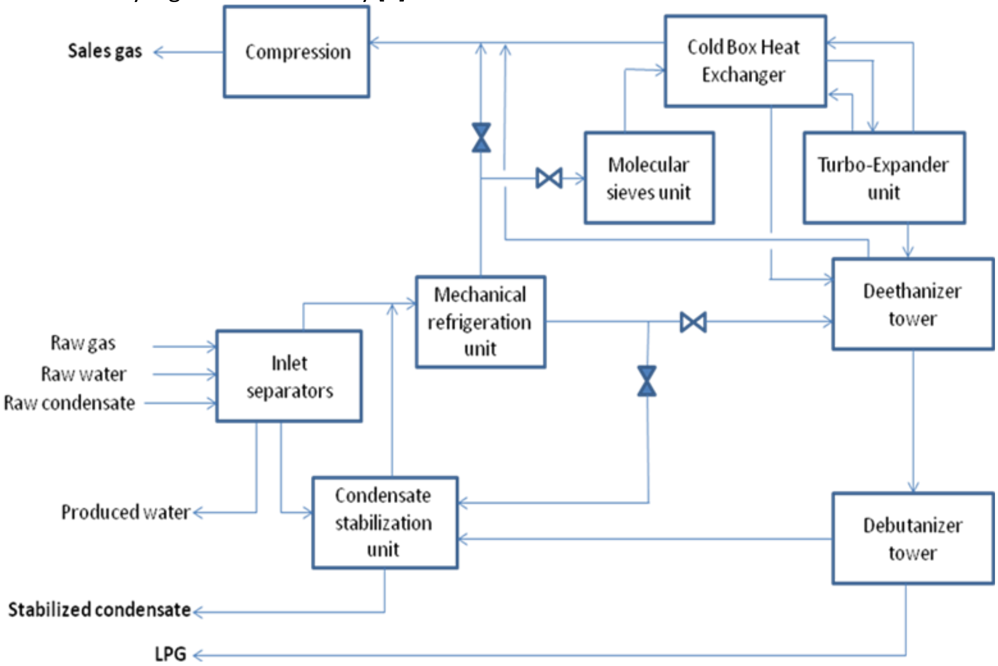

Figure 1 El-Wastani Central Processing Facilities Block Diagram.

Due to the rapid increasing of natural gas and NGL consumption, the need arises to enhance the performance of El-Wastani plant to get higher recovery level of NGLs as well as higher revenue. In the previous article [1] the route of feed gas capacity increase and its effect on performance and profitability of El-Wastani plant were studied. The present work is directed to study other alternative routes for improving the economics and efficiency of the plant. The new NGL recover enhancements routes are based on the recovery maximization of butane, propane, or ethane as a final product for sales. The selection of the best improvement route is based on an economical and technical study of these various routes of improvements. Methods of how to compare different processes are discussed, and optimization of a process to achieve the proper highest return on investment is presented. Each process is then simulated using HYSYS-3.2 which is based on Peng-Robinson equation of state for calculations [11]. The process configurations are developed from process flow diagram (PFD) and or piping and instrumentation diagram (P&ID) information available in-house from El Wastani plant.

Research Methodology

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

- Studying the recovery maximization for ethane, propane, or butane by using HYSYS simulation. Cost estimation is performed and ROI is calculated for each recovery maximization route.
- Studying of the effect of other alternative factors such as feed stability, marketing and total NGL quantity on the decision making process for selecting the optimum route via fuzzy logic technique.
- Determining of the optimum recovery mode and comparing it with the increasing capacity route.

Plant Feed Gas Chemical Composition

As mentioned in the previous article [1], the most critical feed composition was the lean feed (low butanes) composition (see table 1) because it consumes higher energy than other feed composition. So, the improvement of the plant performance operated with the lean feed composition will ensure its performance improvements with other feed compositions.

Table 1 Lean feed gas compositions.

Component	Composition in mole fraction
Nitrogen	0.0011
CO_2	0.0047
Methane	0.8808
Ethane	0.0714
Propane	0.0233
i-Butane	0.0059
n-Butane	0.0046
i-Pentane	0.0021
n-Pentane	0.0011
n-Hexane	0.0014
n-Heptane	0.0011
n-Octane	0.0006
n-Nonane	0.0002
n-Decane	0.0002
H_2O	0.0015
Total	1

Equipment Sizing

Equipment size is calculated by using the output data from Aspen Hysys simulation program to get the proper specification of the concerned equipment such as area with heat exchanger and power in the case of compressor, etc.

For example the exchanger area is calculated by using the duty given from Aspen Hysys. The log mean temperature difference is calculated for each re-boiler and condenser as follows:

Duty: Q (given from Aspen HYSYS)

$$\Delta t = LMTD = \frac{\Delta T_1 - \Delta T_2}{ln\left(\frac{\Delta T_1}{\Delta T_2}\right)} \quad (1)$$

Where Δt is the average temperature difference, LMTD is the log mean temperature difference. ΔT_1 is the temperature difference between the hot inlet stream and cold outlet steam and ΔT_2 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.

$$A = \frac{Q}{K\Delta t} \quad (2)$$

Where k is the thermal conductivity of the equipment material.

Cost Estimation

An estimate of the capital investment for a process may vary from a pre-design estimate based on little information to a detailed estimate prepared from complete drawings and specifications. These estimates are called by a variety of names, but there are five estimate categories represent the accuracy range and designation normally used for design purposes. These five famous estimates are order-of-magnitude, study, preliminary, definitive and detailed estimates. The accuracy of these estimates are ranging from + or - 30 to + or - 5 percent respectively [12].

According to the current status of the plant (El Wastani company) data availability and calculation stage, the preliminary estimate (budget authorization estimate) is applied. This choice is based on sufficient data used to permit the estimate to be budgeted. Probable accuracy of this estimate is within + or - 20 percent.

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 [12].

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

$$C_n = C_k \left(\frac{V_n}{V_k}\right)^F \quad (3)$$

Where, V_n is the capacity of the new plant, and V_k is the capacity of the known plant. F is a factor; usually take a value between 0.4 and 0.9, depending on the type of plant. In the applied estimating technique, a factor value of 0.6 is used according to the literature or historical data [13]. 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 current time. This is done by the comparison with equipment prices which are obtained from El-Wastani Company [14].

Working Capital Cost

The working capital is the amount of capital required to start up the plant and to finance the first couple of months of operating before the plant starts earning. This capital is used to cover salaries, raw material inventories and contingencies. It will be recovered at the end of the project and represents a float of money to get the project started. These costs are necessary at start-ups and it implies raw materials and intermediates in the process. The working capital cost is assumed to be 3% of the fixed capital cost [15].

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 capacity or recovery maximization of ethane, propane, or butane.

Investment Analysis Technique

The results obtained from the simulator for each modification route are analyzed by two different techniques; return on investment and Intelligent System Technique (Fuzzy Logic).

Return on Investment

The determination and analysis of profits obtainable from the total cost of investment and the choice of the best investment among various alternatives are major goals of the investment analysis. Values from 20 to 30 percentages for ROI can be used as a rough guide for judging the plant retrofitting. Consequently, the decision has to be made on whether to install additional equipment to reduce operating costs or not [12, 16].

The calculations of ROI are mainly consisting of the following two major terms:

- Total capital investment which includes the cost of purchased equipment, installation and foundation, instrumentation, piping, and commissioning works.
- The net profit which is resulting via productivity increase after excluding the annual increase in operating cost.

$$ROI = \frac{\text{(Total annual income increment} - \text{annual increase in operating cost)}}{\text{Total capital investement}} \quad (4)$$

The total capital investment includes the fixed capital cost in addition to the working capital cost. 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 bellow:

$$\textbf{Pay back period} = {}^{\mathbf{1}}\!/_{\boldsymbol{ROI}} \quad (5)$$

Intelligent System Technique (Fuzzy Logic)

Fuzzy logic is a tool for transforming human knowledge and its decision-making ability into a mathematical formula. In other words, it provides a meaningful and powerful representation of measurement uncertainties and also a meaningful representation of vague concepts expressed in natural language. In any decision-making process it is necessary to evaluate different alternatives and discard those that do not fit certain previously established criteria [17-19].

For a finite set A of n alternatives and a given system of m evaluation criteria f_i, the multi criteria problem in its general form can be defined as follows:

$$\text{Max } \{f_1(a), f_2(a), \ldots, f_m(a)/a\varepsilon A\} \quad (6)$$

For every criterion f_j of the multi-criteria problem, a membership function x_j is defined and takes values to the interval [0,1]. This transformation is based on the concept of the ideal point. So, the value x_{aj} represented by equation 7 expresses the degree to which the alternative "a" is close to the ideal value f_j^* (the best performance in criterion j) and far from the anti-ideal value f_j^{**} (the worst performance in criterion j) [20-22]:

$$Xaj = \frac{fj(a) - fj**}{fj* - fj**} \quad (7)$$

Many simple decision processes are based on a single objective, such as maximizing profit, minimizing run time, and so forth. Besides the capacity increase route (as an objective), there are other many alternatives which can improve the investigated plant performance such as:

- Feed stability continuity.
- Marketing concerns and shipping availability.
- Return on investment; including cost and profit individually.
- Recovered NGL productivity.

Define universe of "r" alternatives:

$$O = \{O_1, O_2, \ldots, Or\} \quad (8)$$

Let O_i indicate the ith objective. Then the degree of membership of an alternative "a" in O_i, denoted $\mu_{oi}(a)$. The decision function, D, that simultaneously satisfies all of the decision objectives is given by the intersection of all the objectives sets as indicated in the following equation:

$$D = O_1 \cap O_2 \cap \ldots \cap O_r \quad (9)$$

Therefore, the grades of membership that the decision function, D, has for each alternative "a" is given by:

$$\mu_D(a) = \min[\mu_{o1}(a), \mu_{o2}(a), \ldots, \mu_{or}(a)] \quad (10)$$

The optimum decision, "a*", will then be the alternative that satisfies the following equation [20-26]:

$$\mu_D(a^*) = \max_{a\varepsilon A}(\mu_D(a)) \quad (11)$$

Results and Discussion

The performance of El Wastani plant for high recovery of NGL components is studied at 160 MMSCFD as a feed gas capacity with lean feed gas composition (as listed in table 1). High butane, propane or ethane recovery routes are studied for 96 % recovery for each route.

Expansion efficiency is thermodynamically a function of temperature. Therefore, the inlet temperature to the expander is critical to the determination of the expander performance [27]. Expanders are known to be applied at the lowest temperature levels of configurations, as this where they are thermodynamically more efficient [28, 29]. So, the modification and the additional requirement should be arranged for refrigeration and separation in the original scheme.

Butane Recovery Mode

The installation of an additional refrigeration package is required to achieve the high recovery (96%) of butane. The major areas which need revamping are as the following:

- Mechanical refrigeration unit to allow lowering the temperature of the chiller in order to condense the targeted level of butane.
- Condensate stabilization unit to recover all the butanes to maximize the LPG quantity.
- Fractionation train in particular the de-ethanizer tower to handle the additional light gases due to the low temperature process and topped the feed to de-butanizer tower.

Table 2 exhibits the composition analysis of the inlet and outlet gas after and before the plant revamping.

Table 2 Feed and sale gas analysis before and after

Component	Feed Gas		Sale Gas			
			Before		After	
	Mole fraction	Mass flow (kg/h)	Mole fraction	Mass flow (kg/h)	Mole fraction	Mass flow (kg/h)
Nitrogen	0.0011	245	0.001	223	0.0011	245
CO_2	0.0047	1645	0.0042	1426	0.0048	1639
Methane	0.8771	112616	0.9134	114131	0.9083	112616
Ethane	0.0711	17114	0.0659	15435	0.0736	17114
Propane	0.0232	8186	0.0126	4322	0.012	4076
i-Butane	0.0063	2950	0.0019	841	0.0001	43
n-Butane	0.0049	2300	0.0009	411	0	11
i-Pentane	0.0021	1224	0.0001	69	0	0
n-Pentane	0.0011	654	0	20	0	0
n-Hexane	0.0014	967	0	2	0	0
n-Heptane	0.0012	925	0	0	0	0
n-Octane	0.0006	559	0	0	0	0
n-Nonane	0.0002	203	0	0	0	0
n-Decane	0.0002	184	0	0	0	0
H_2O	0.003	436	0	0	0	0
Ethylene Glycol	0.0018	873	0	0	0	0
Total	1	151082	1	136881	1	135745

Table 3 Simulation results for butane recovery mode.

PARAMETER		Unit	Design	Simulated
Gas rate		MMSCFD	160	160
Glycol re-boiler duty		MMBtu/h	1	0.9147
Glycol injection rate		M3/H	1.158	1.037
Gas/gas heat exchanger		MMBtu/h	15.57	18
MRU chiller		MMBtu/h	6.7	15.4
Mole Sieve		MMSCFD	162.000	152.400
Cold Box		MMBtu/h	12.80	12.83
Turbo Expander	Power	hp	Max. 1165	1165.000
	Gas rate	MMSCFD	Max. 190.8	137.000
De-Ethanizer tower	Condenser duty	MMBtu/h	1.435	5.582
	Re-boiler duty	MMBtu/h	8.440	17.85
	Compressor power	hp	200	274.3
	Compressor quantity	MMSCFD	12.01	30.9
De-Butanizer tower	Condenser duty	MMBtu/h	8.458	4.787
	Re-boiler duty	MMBtu/h	6.835	3.565
Condensate Stabilizer tower	Condenser duty	MMBtu/h	5.962	5.676
	Re-boiler duty	MMBtu/h	7.880	5.456
	Compressor power	hp	530.000	219.300
	Compressor quantity	MMSCFD	2.710	1.506
Sale gas boosting compressor	Power	hp	11000	6778
	Quantity	MMSCFD	193.32	155.7

Table 4 Plant Production quantities at butane recovery mode.

Raw Gas Rate MMSCFD	Sales Gas MMSCFD	Sales LPG		Sales Condensate BBLS/day
		Ton/day	Recovery %	
160	155.2	230.1	96.05	4052

Table 3 presents the simulation results of the plant. The incremental percentage and the perspective quantities to be obtained after perform the proposed modification, are tabulated in table 4. Table 5 summarizes all needed economic data to have a viable revamp alternative.

Propane Recovery Mode

In order to achieve 96% propane recovery, an additional refrigeration package as well as a de-propanizer package are required to be installed for sustaining an adequate separation of propane as a commercial product with desired specifications. This is accomplished by calculating and fulfilling all the requirements for each additional equipment.

The major areas which need revamping are as the following:

- Mechanical refrigeration unit to allow lowering the temperature of the chiller in order to condense the targeted level of butane & propane.
- Fractionation train:-
  - In particular the de-ethanizer tower to handle the additional light gases due to the low temperature process and topped the feed to de-butanizer tower.
  - Adding de-propanizer tower to handle the de-ethanizer re-boiler outlet prior to de-butanizer tower and accomplish the required on spec commercial propane.
- Revamp plate and fin heat exchange to handle and utilize the excess energy due to the extra refrigeration power.

Table 6, presents the simulation results of the plant and Table 7 exhibits the composition analysis of the inlet and outlet gas after and before the plan revamping. While the incremental percentage and the perspective quantities to be obtained after performing the proposed modification, are tabulated in Table 8. Table 9 summarizes all needed economic data to have a viable revamp alternative.

Ethane Recovery Mode

For obtaining 96% ethane recovery, it is required to install additional refrigeration package as well as installing the following packages:

- De-methanizer unit package.
- De-ethanizer additional tower to fulfill the increasing of condensed hydrocarbons.
- De-propanizer unit package.

To sustain adequate separation of propane as a commercial product with the desired specifications, this target is accomplished by calculating and fulfilling all the requirements for each additional package. The major areas which will be modified are mechanical refrigeration unit, molecular sieve unit for dehydration, cold box unit, and Fractionation train.

Table 5 Economic study results for butane recovery mode

Fixed Capital Investment, MM$	24.3
Operating Cost, MM$	3.2
Total Capital Investment, MM$	25
Revenue, MM$	366
Increment, MM$	17.11
ROI (return on investment), %	56
Pay-Back Time, Years	1.8

Table 6 Simulation results for propane recovery mode.

PARAMETER		Unit	Design	Simulated
Gas rate		MMSCFD	160	160
Glycol re-boiler duty		MMBtu/	1	0.9513
Glycol injection rate		M3/H	1.158	1.040
Gas/gas heat exchanger		MMBtu/	15.57	18.650
MRU chiller		MMBtu/	6.7	19.98
Mole sieve		MMSCFD	162.000	145.1
Cold Box		MMBtu/	12.80	10.41
Turbo Expander	**Power**	hp	Max.	1165
	Gas rate	MMSCFD	Max.	132.4
De-ethanizer tower	**Condenser duty**	MMBtu/	1.435	11.03
	Re-boiler duty	MMBtu/	8.440	22.47
	Compressor power	hp	200	280
	Compressor	MMSCFD	12.01	36.5
De-propanizer tower	**Condenser duty**	MMBtu/	-	18.84
	Re-boiler duty	MMBtu/	-	18.63
De-butanizer tower	**Condenser duty**	MMBtu/	8.458	5.13
	Re-boiler duty	MMBtu/	6.835	4.616
Condensate Stabilizer tower	**Condenser duty**	MMBtu/	5.962	5.845
	Re-boiler duty	MMBtu/	7.880	5.422
	Compressor power	hp	530.000	218.6
	Compressor	MMSCFD	2.710	1.501
Hot oil system		MMBtu/	32	52.089
Sale gas boosting Compressor	**Power**	hp	11000	6973
	Quantity	MMSCFD	193.32	153.4

Table 10, presents the simulation results of the plant for the ethane recovery mode while Table 11shows the composition and flow rates of the inlet and outlet gas after and before the plant revamp. The incremental percentage and the perspective quantities to be obtained after performing the proposed modification are tabulated in Table 12.

Table 13 summarizes all needed economic data required to have that possible revamp alternative.

Table 7 Feed and sale gas analysis before and after plant revamp for propane recovery mode.

Component	Feed Gas		Sale Gas			
			Before		After	
	Mole fraction	Mass flow (kg/h)	Mole fraction	Mass flow (kg/h)	Mole fraction	Mass flow (kg/h)
Nitrogen	0.0011	245	0.001	223	0.0011	245
CO_2	0.0047	1645	0.0042	1426	0.0049	1638
Methane	0.8771	112616	0.9134	114131	0.9185	112616
Ethane	0.0711	17114	0.0659	15435	0.0745	17114
Propane	0.0232	8186	0.0126	4322	0.0009	310
i-Butane	0.0063	2950	0.0019	841	0	13
n-Butane	0.0049	2300	0.0009	411	0	4
i-Pentane	0.0021	1224	0.0001	69	0	0
n-Pentane	0.0011	654	0	20	0	0
n-Hexane	0.0014	967	0	2	0	0
n-Heptane	0.0012	925	0	0	0	0
n-Octane	0.0006	559	0	0	0	0
n-Nonane	0.0002	203	0	0	0	0
n-Decane	0.0002	184	0	0	0	0
H_2O	0.003	436	0	0	0	0
Ethylene Glycol	0.0018	873	0	0	0	0
Total	1	151082	1	136881	1	131941

Table 8 Plant Production quantities for Propane Recovery Mode.

Raw Gas Rate	Sales Gas	Sales propane		Sales LPG		Sales Condensate
MMSCFD	MMSCFD	Ton/day	Recovery %	Ton/day	Recovery %	BBLS/day
160	153.4	189.9	96.00	128.1	96.00	4080

Table 9 Economic study results for propane recovery mode.

Fixed Capital Investment, MM$	41.14
Operating Cost, MM$	3.38
Total Capital Investment, MM$	42.37
Revenue, MM$	408.02
Increment, MM$	59.67
ROI (return on investment), %	133
Pay-Back Time, Years	0.75

Decision Making Operation

The process of optimization does not rely only on the economic route (ROI), but sometimes it is important to broaden the judgment basis. So, the aim of decision making is to find out the best possible course of action. It is a rational and purposeful activity designed to attain well-defined objectives, such as (feed stability, marketing availability, ROI, and recovered NGL quantity).

In the following section, all-related decision making steps such as data tabulation, fuzzy membership, ..., etc are performed.

In order to perform the procedure of fuzzy logic technique, the first step is the tabulation of the various values of alternatives' objectives. The following sections will exhibit as well as discuss the different parameters and ways of its evaluation. The different parameters are feed stability continuity, Marketing availability, Return on investment, and Recovered quantity of NGL.

Table 10 Simulation results for the ethane recovery mode.

PARAMETER		Unit	Design	Simulated
Gas rate		MMSCFD	160	160
Glycol re-boiler duty		MMBtu/h	1	1.051
Glycol injection rate		M3/H	1.158	1.071
Gas/gas heat exchanger		MMBtu/h	15.57	12.26
MRU chiller		MMBtu/h	6.7	45.04
Mole sieve		MMSCFD	162.000	99.4
Cold Box		MMBtu/h	12.80	5.007
Turbo Expander	**Power**	hp	Max. 1165	1165
	Gas rate	MMSCFD	Max. 190.8	87.93
De-methanizer tower	**Condenser duty**	MMBtu/h	-	17.38
	Re-boiler duty	MMBtu/h	-	11.55
	Compressor power	hp	-	431.6
	Compressor quantity	MMSCFD	-	36.5
	Tower quantity	BPD	-	42357
De-ethanizer tower	**Condenser duty**	MMBtu/h	1.435	18.84
	Re-boiler duty	MMBtu/h	8.440	22.47
	Tower quantity	BPD	9528	12470
De-propanizer tower	**Condenser duty**	MMBtu/h	-	18.84
	Re-boiler duty	MMBtu/h	-	18.63
	Tower quantity	BPD	-	4897
De-butanizer tower	**Condenser duty**	MMBtu/h	8.458	5.13
	Re-boiler duty	MMBtu/h	6.835	4.616
Condensate Stabilizer tower	**Condenser duty**	MMBtu/h	5.962	5.845
	Re-boiler duty	MMBtu/h	7.880	5.456
	Compressor power	hp	530.000	219.3
	Compressor quantity	MMSCFD	2.710	1.501
Hot oil system		MMBtu/h	32	63.773
Sale gas boosting Compressor	**Power**	hp	11000	10250
	Quantity	MMSCFD	193.32	140.8

Table 11 Feed and sale gas analysis before and after plant revamp for ethane recovery mode

Component	**Feed gas**		**Sale gas**			
			Before		**After**	
	Mole fraction	**Mass flow (kg/h)**	**Mole fraction**	**Mass flow (kg/h)**	**Mole fraction**	**Mass flow (kg/h)**
Nitrogen	0.0011	245	0.001	223	0.0012	245
CO_2	0.0047	1645	0.0042	1426	0.0039	1206
Methane	0.8771	112616	0.9134	114131	0.991	112545
Ethane	0.0711	17114	0.0659	15435	0.0039	823
Propane	0.0232	8186	0.0126	4322	0	14
i-Butane	0.0063	2950	0.0019	841	0	0
n-Butane	0.0049	2300	0.0009	411	0	0
i-Pentane	0.0021	1224	0.0001	69	0	0
n-Pentane	0.0011	654	0	20	0	0
n-Hexane	0.0014	967	0	2	0	0
n-Heptane	0.0012	925	0	0	0	0
n-Octane	0.0006	559	0	0	0	0
n-Nonane	0.0002	203	0	0	0	0
n-Decane	0.0002	184	0	0	0	0
H_2O	0.003	436	0	0	0	0
Ethylene Glycol	0.0018	873	0	0	0	0
Total	1	151082	1	136881	1	114834

Table 12 Pant Production quantities for Ethane Recovery Mode.

Raw Gas Rate	Sales Gas	Sales ethane		Sales propane		Sales LPG		Sales Cond.
MMSCFD	MMSCFD	Ton/day	Recovery %	Ton/day	Recovery %	Ton/day	Recovery %	BBLS/day
160	140.8	391.0	95.20	196.1	99.80	128.1	96.00	4080

Table 13 Economic study results for ethane recovery mode.

Fixed Capital Investment, MM$	97.82
Operating Cost, MM$	4.57
Total Capital Investment, MM$	100.76
Revenue, MM$	412.73
Increment, MM$	64.38
ROI (return on investment), %	59
Pay-Back Time, Years	1.68

Feed Stability Continuity

Referring to the annual report of El Wastani company production gas figures and according to Figure 2 which represents the trend of raw gas rate at the inlet of the central processing facilities [30], as well as the average gas rate during the period from July 1, 2012 to March 1, 2013 [31], it is obtained that the average inlet gas rate stability is ranging from 140 to 158 MMscfd.

As shown in Figure 2, it is obvious that the average gas rate production in more conservative terms varies around 149 MMSCFD, so the feed stability term will be determined as a percentage of the required feed gas rate to the expected or average gas rate. According to the fuzzy logic system, the feed rate quantities are revalued to be compatible with the concept of fuzzy membership rule. So, the feed rate quantities are normalized as a fraction of the original capacity (160 MMSCFD), as shown in table 14

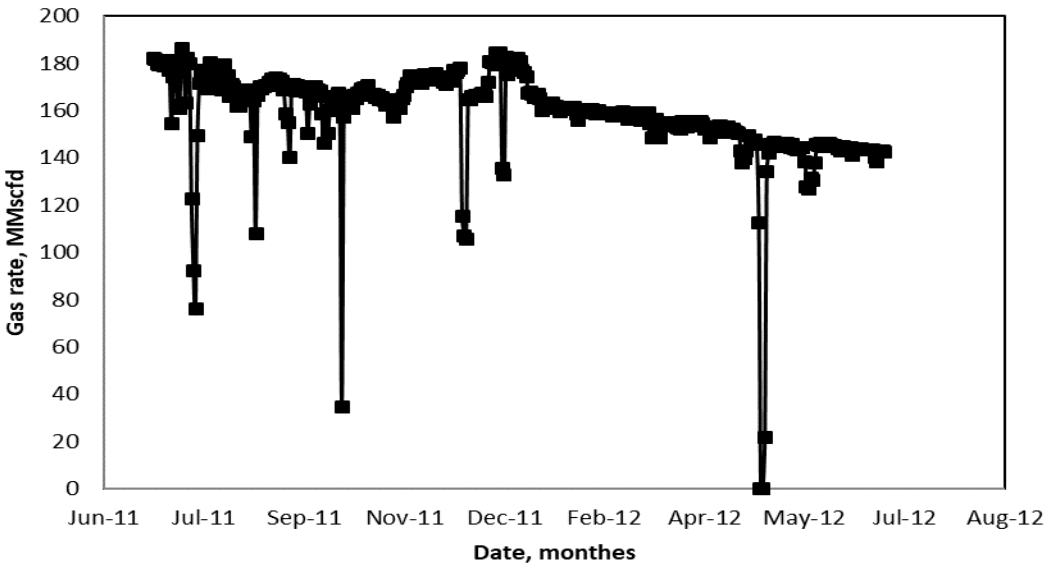

Figure 2 El-Wastani raw gas rate, MMSCFD [30].

Table 14 Normalized value of Feed Stability.

Mode	Ethane	Propane	Butane	200 MMscfd Mode
Feed stability value	(149/160)= 0.9	(149/160)=0.9	(149/160)=0.9	(149/200)=0.75

Marketing Availability

Plant products marketing availability is a very crucial objective and how easily these products could be sold and shipped are important factors in the economics of the plant. Figure 3 represents a promising plan for petrochemical industry in Egypt which is consuming the various natural gas products. Thus, each alternative will be characterized according to the current circumstances as a rough approximation properly.

Where polypropylene industry (propane consuming section) is already established and the UGDC (United Gas Derivatives Company) export terminal is in vicinity to the central processing facility (CPF) of El Wastani plant, so the marketing likelihood of propane mode could be about 75 %. On the other hand, for ethane mode which is evaluated as a new born, it could be estimated to have a value of 50 %. This low value of marketing availability of ethane mode may be attributed to the needs to some construction for pipelines or supply transfer trucks. For both of butane and capacity increase (200 MMSCFD) modes the likelihood is taken as 100 %. This is because all of the production are domestically consumed and transfer infrastructure were already established.

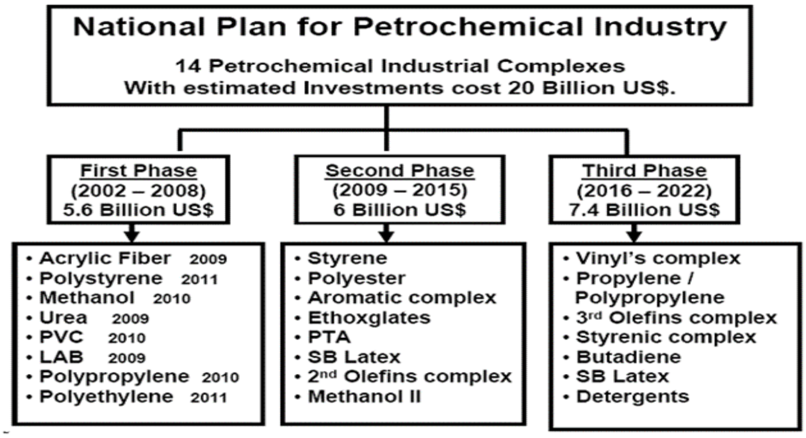

Figure 3 National plan for petrochemical industry in Egypt [32].

Return on Investment

Table 15 summarizes the different values of ROI for the four proposed improvement modes. The ROI values for high recovery of butane, propane or ethane modes are calculated according to the economic data previously presented in sections 3.1, 3.2, and 3.3. From this table, it is clear that operating the plant at 200 MMSCFD is the most profitable route based on its higher value of ROI.

Table 15 ROI Values for the four improvement modes.

Alternative	ROI, %
200 MMSCFD	212
Ethane Mode	59
Propane Mode	133
Butane Mode	56

Recovered Quantity of NGL

NGL products are increasingly requested due to the vast raise in consuming sections such as domestic use, petrochemical plants, …, etc. Thus, the availability and continuity of suppliers should be stable and should fulfill all national sections requirements. Therefore, the productivity improvements do not mainly relate to the profitability in money means, but also it is strategically very important to provide the required raw material and energy sources.

To accomplish the process of decision making, it is important to determine the different values of the various objectives as shown in Table 16. Results showed that ethane mode gives the highest quantity of NGL while the capacity increase route (200 MMSCFD) gives the highest value of ROI and the lowest value of feed stability.

Table 16 Various alternatives and objectives data tabulation.

Alternative/Objective	Feed Stability	Marketing	Total NGL Quantity, Ton/day	ROI, %
200 MMSCFD	0.75	Available	814656	212
Ethane Mode	0.9	Available	1182960	59
Propane Mode	0.9	Available	766272	133
Butane Mode	0.9	Available	681264	56

Data Mapping and Decision Making

According to fuzzy logic methodology mentioned previously in section 2.4.2, the mapped value of each alternative is determined and the results are presented in Table 17.

Table 17 Decision making problem solving results.

Alternative/Objective	Feed Stability	Marketing	Total NGL Quantity, Ton/day	ROI, %	Min.	Max.
200 MMscfd	0	1	0.3	1	0	
Ethane Mode	1	1	1	0.02	0.02	
Propane Mode	1	1	0.2	0.5	0.2	0.2
Butane Mode	1	1	0	0	0	

Due to the results of the decision making process listed in Table17, it is clear that the optimum selection is propane mode route which make the plant more efficient and more profitable.

Conclusions

The present work results showed that the simple decision making methods rely only on ROI only exhibit that the capacity increasing route is more favorable for El Wastanti petroleum Company to increase its profitability and productivity. But by applying an intelligent system such as fuzzy logic technique, the plant can be more profitable by other routes rather than capacity route. From the economic and strategic viewpoints, it’s more profitable to enhance the recovery of NGLs components; specially ethane and propane to assure utilizing of plant full capacity all the time, as well as to optimize the production rate from the wells network based on long productivity plan. So the present work studied new routes for the plant enhancement such as the maximization of butane, propane or ethane recovery.

The process of multi-objective decision making using fuzzy logic system showed that the maximization of propane recovery is the optimum and valuable route for improving El Wastani plant. This choice does not only depend on the higher value of ROI but takes into account other various objectives such as feed stability, marketing availability and recovered NGL quantity.

The efforts done in this present work are helpful and can be applied for improving the existing NGL recovery plants as well as the plants under design.

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