Improving a Process Design for the Condensate Stabilization Unit

Jahanbin, Shahpour; Hosseini, Seyed Nooroldin*+*

Department of Petroleum Engineering, Omidiyeh Branch, Islamic Azad University, Omidiyeh, I.R. IRAN

Sabet, Maziyar

Faculty of Engineering, Petroleum and Chemical Engineering, University Technology Brunei (UTB), Bandar Seri Begawan, BRUNEI

ABSTRACT: *The condensate stabilization process is known as a common operation in gas fields with the aim of diminishing the condensate Reid Vapor Pressure (RVP). Based on feed characteristics and specified products, condensate stabilization units can be optimized by examining different process configurations. In each configuration, there exist some criteria in terms of product quality, energy-saving, and economic recovery, which shall be investigated so that the configuration's performance is evaluated and selected. The aim of this study is to illustrate four condensate stabilization configurations and then to determine the proper column temperature and pressure, feed split ratio, and feed tray location by means of Aspen Hysys in the configurations. In the end, it is shown that the configuration with the lowest operating pressure not only had the lowest fixed capital cost but also consumed the lowest energy due to its high separation efficiency and low energy requirement.*

KEYWORDS: *Condensate Stabilization; Simulation; Aspen HYSYS; Feasibility study; Stripping.*

INTRODUCTION

The condensate coming from a natural gas reservoir must be treated to be safe and environmentally acceptable for storage. Therefore, some mandatory actions such as removing salt and water as well as separating any dissolved gases are needed. All of these are done in the condensate stabilization process [1]. The process is primarily performed so as to reduce the Reid Vapor Pressure (RVP) of condensate. By doing so, the amount of intermediate and heavy components in the condensate is increased, and vaporization does not occur when the condensate is routed to atmospheric storage thanks. Moreover, as gas molecules leave the stabilized condensate, H_2S and mercaptans which are in the raw condensate are totally removed and low-sulfur product is prepared [2].

Literature shows that the condensate stabilization can be accomplished by the means of either fractionation or flash vaporization [1, 3-4]. While flashing vaporization is a simple operation done by employing two or three separators, the fractionation method is a detailed process done using one column. The flash vaporization only uses separators in different pressures in the range of 600 psia to 65 psia to separate vapor from the hydrocarbon condensate. Note that this method is accounted as old technology and is not applied in a modern plant. This technology is not within the scope of the present study.

* *To whom correspondence should be addressed.*
+ *E-mail*: *noor_hoseini@iauo.ac.ir*
1021-9986/2021/6/1960-1972 13/$/6.03

Column stabilizers (fractionators) are popular in the industry contrarily and produce more liquid (stabilized condensate). The final product coming from the bottom of the columns is mainly composed of pentane and heavier hydrocarbons. The bottom product is really free from any gaseous components and is ready to be stored safely at atmospheric pressure. This method is modern and economically attractive. Fixing practical columns in the condensate stabilization units is a difficult task and needs experience and knowledge. There are loads of work guiding engineers on how to set stabilizers' heights, diameters, the number of trays, feed trays, pump around, reflux ratios, and reboiler duties according to the product yields [5-7].

As hydrocarbon condensate is a priceless source of energy, several valuable studies have been conducted to get more clear information about extracting stabilized condensate from the unstabilized condensate. *Moghadam* and *Samadi* [4] compared the performances of multistage flash vaporization and fractionation in two different cases through Hysys and Aspen plus. The first case came from the Khangiran condensation stabilization unit whereas the second from Natural Gas Liquid (NGL) 3100 condensate stabilization unit, both in Iran. In their studies, heat duties and capital costs of units represent the units' performances. RVP of the feed had a great effect on the stabilization units' performances. They showed that the multistage flash vaporization was more practical for the Khangiran case, which had a lower feed RVP than NGL3100. On the other hand, the fractionation unit was far better for NGL3100, which had a higher feed RVP than Khangiran. A simulation study that investigated the performance of an industrial condensate stabilization was conducted using Aspen Hysys by *Rahmanian et al.* [3]. The stabilizer column defined in their work was a non-refluxed type. They evaluated RVP of the final product of the unit with four key parameters in terms of feed flow rate, temperature, pressure, and also reboiler temperature. It was found that the effect of the reboiler temperature over RVP was more tangible than the others. In their other work [8], they investigated the effect of steam temperature and pressure on the quality of products in terms of sulfur and RVP, and they found the optimum steam temperature in their studied unit which benefited the unit more theoretically (not economically). *Uwitonze et al.* [9] proposed four different alternatives for the stabilization process. They introduced a heat integrated method by which the size of the used stabilizer column would decrease with respect to the other alternatives. *Zhu et al.* [10] compared the performances of two concepts, one of which was a non-refluxed stabilizer column and the second of which was a split-flow stabilizer column. It was shown that while having the imposed specs, the split-flow approach could economically benefit the unit more. *Tahouni et al.* [11] focused on the retrofit of the heat exchanger network of the Assaluyeh condensate stabilization unit. The main objective of their study was evaluating the heat recovery of the unit when the throughput of the unit was increased by a 20%. The study showed that after 20% increase in the throughput, the steam consumption could be kept at the existing level if the steam pressures in heat exchangers were changed and 1554 m^2 of additional heat transfer area was installed. *Tavan et al.* [12] compared the performance of an industrial non-refluxed condensate column unit to that of the same unit with an installed water draw rate. To do so, they used HYSYS 3.1 process simulator. Their results showed that the water draws pan/tray increased the performance of compressor and reboiler duties and decreased the off-gas stream impurities. *El Eishy et al.* [13] did a simulation study into the Sannan condensate stabilization unit in Egypt. They investigated the effects of important variables, such as outlet temperature of the process, stabilizer feed drum pressure, stabilizer feed tray location, and reflux ratio of the stabilizer, on the unit. Following that, the optimal process operating conditions with the most achievement were computed. They showed that the condensate productivity of the existing unit could be increased by 26.25%. This culminated in the extra revenue in the gross profit by 1.6 million dollars per year. *Khalili et al.* [14] defined some sequence distillations which could replace conventional distillation columns in condensate stabilization units. It was shown that there are sequences that result in higher recovery in terms of exergy and annual cost than the conventional distillation configuration.

Although the previous studies have investigated different alternatives to the process, it seems they did not go deeply into the sizing procedures. Therefore, their conclusions do not stand on firm ground. The purpose of this manuscript is to propose four different condensate stabilization processes (cases) to produce the desired products with the same specs. This is followed by equipment sizing and economic evaluation of each case.

RVP of 10 psia has been set as the criteria for the off-specification condition of the stabilized condensate product in summer. The first case is on the basis of the existing condensate stabilization unit operating in Assaluyeh in Iran. In the second case, the stabilizer column operates as a non-refluxed tower. In the third case, a split feed flow ratio approach was applied to the unit. Finally, based on the first case, the fourth case at lower feed pressure was used. The effect of operating conditions such as feed temperature, feed pressure, and feed tray on the processes were studied.

THEORETICAL SECTION

Process description

Several gas plants are under construction on the Iranian coast of the Persian Gulf for extracting C5+ products from the associated gases coming from Assaluyeh oil fields. It is assumed that petroleum companies can produce at least 34000 bbl/d (barrel per day) stabilized condensate, together with 25 MMSCD (Million standard cubic feet per day) sweet gas in each phase. Each complex includes offshore facilities (wells, platforms, and undersea pipelines) and onshore facilities for the processing of the reservoir fluid. The complex is fed by the reservoir fluid delivered to the onshore plant via multiphase sea lines. The feed is routed to facilities for High Pressure (HP) separation of raw gas and condensate/water mixture. Its condensate is stabilized for storage and export in the stabilized condensate unit, and light ends are recycled in the HP gas system. The gas is sent to the treatment facilities producing sales gas, gaseous ethane, and NGL. H_2S /CO_2 removal from gas, dehydration, ethane extraction, NGL extraction are done in this unit. Aside from that, there are Monoethylene Glycol (MEG) regeneration and injection unit, Sulphur recovery unit, Utilities, and off-sites required for the operation.

The liquids separated in the slug catchers (receiving facilities) are sent to the Condensate Stabilization unit. The function of this unit is to remove the lightest components from the raw feed and to perform a hydrocarbon liquid product with a RVP of 10 Psia. The pressure and temperature of the entering feed are 30 bara and 23 ℃. Feedstock characteristics and capacity are listed in Table 1. The Process Flow Diagram (PFD) shown in Fig. 1 is the same as what exists in the Assaluyeh oil fields. The raw condensate mixed with glycolated water is first preheated in E-1. The total mixture, then, is routed to a preflash drum

Table 1: The compositions of the condensate stabilization unit's feedstock.

Composition	(%mol)
H2O	21.32
N2	0.26
CO2	0.89
H2S	0.84
C1	20.21
C2	4.77
C3	4.16
iC4	1.5
nC4	3.13
iC5	1.91
nC5	2.14
C6cut	4.13
C7cut	5.52
C8cut	6.52
C9cut	4.68
C10+	11.32
MEG	6.23
Total (kmol/h)	3460

V-1, which needs to operate at 50 ℃ and 28 bara. It is a three-phase separator that directs the flash gas toward the second stage compressor suction drum, V-5. The glycolated water is sent to the MEG regeneration unit. Using a pump, P-1, the hydrocarbon liquid is pumped to the condensate desalter, V-2. Before that, the hydrocarbon liquid is exposed to demulsifying chemicals in order that the desalter drum removes the free aqueous phase from the hydrocarbon liquid as much as possible. To ensure efficient separation of glycolated water from condensate, the operating temperature in the desalter needs to be held at 70 ℃ approximately. This is done by heating the fluid with the condensate desalter preheater, E-2. The water phase-separated in V-2 is sent to the MEG regeneration unit.

The liquid hydrocarbon after being heated through heat-exchanger E-3 is treated in the condensate stabilizer (column) T-1 operating at a pressure of about 10 bara. The pressure is controlled by the means of a control valve shown in the figure. Lighter components in upgoing gas

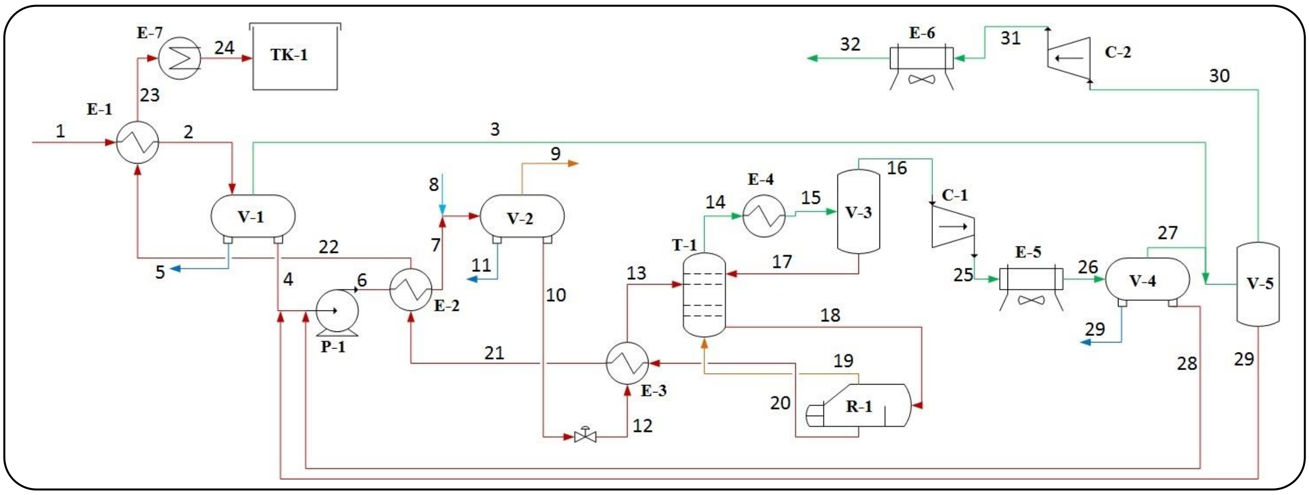

Fig. 1: Condensate stabilization's process flow diagram.

are removed with the condensed liquid serving as reflux. A reboiler, which is served with medium pressure steam (12 bara), is used to fix the column bottom temperature at about 190 °C. In addition, the column is equipped with a partial reflux condenser cooling down the distillate to about 60°C.

The stabilized condensate from the stabilizer bottom column is cooled through condensate stabilizer preheater, condensate desalter preheater, condensate pre-flash heater, and trim cooler E-7 and then is sent to the storage tank TK-1 for exporting. The trim cooler E-7 is designed in order to maintain the stabilized condensate temperature equal to 45 °C.

The stabilizer overhead vapor is compressed by a two-stage motor-driven compressor with cooling and vapor-liquid separation at the inter-stage. In the first stage, the discharge gas is cooled down to 60°C in the air cooler E-5 and sent to the three-phase separator V-4. Water released from the drum, if any, is routed to the Sour Water Stripping (SWS) unit. The hydrocarbon liquid is sent to the raw condensate feed, after the pre-flash drum. The overhead gas is then mixed with gas from pre-flash drum V-1 prior to being routed to the second stage gas compressor suction drum, V-5. Similar to the previous hydrocarbon liquid, the liquid from the suction drum is recycled back to the suction line of P-1. Gas is further compressed by C-2 (as the second compression stage) and cooled in the air cooler E-6 and sent to the gas treatment plant for further processing.

The aim of this work is to evaluate four proposed condensate stabilizer configurations from an economic point of view. The configurations are almost the same, but there are some differences which will be described case by case as follows.

Case 1:

This case is similar to what was mentioned in the process description (Assaluyeh oil fields). Fig. presents the PFD of Case 1. This case is more complex with respect to the others.

Case 2:

The difference between case 2 and case 1 is the absence of the column condenser. It means case 2 does not cool the overhead product at the top of the column, and there is no reflux to be recycled back to the stabilizer column. Indeed, stream 16 in case 2 comes from the stabilizer column directly, while stream 16 in case 1 comes from the column condenser. The configuration of case 2 is shown in Fig. 3.

Case 3:

Fig. 4 is the other alternative taken from case 2. Case 3 can be counted as a modification of Case 2. In order to take advantage of the reflux ratio, stream 15, which is a portion of stream 14, is routed to the top of the column without any preheating through E-3. As stream 15 falls into the column, it makes gas going up leaner in heavy components and richer in the light ends. This cold portion helps the column find its high purity. To achieve the best condition with the lowest heat duty, different split ratios (the ratio of stream 15/stream 14) were examined in this study. It is recommended 15% of the feed be sent to the top tray. The rest of the feed, stream 16, goes to the condensate stabilizer preheater before being sent to the stabilizer column.

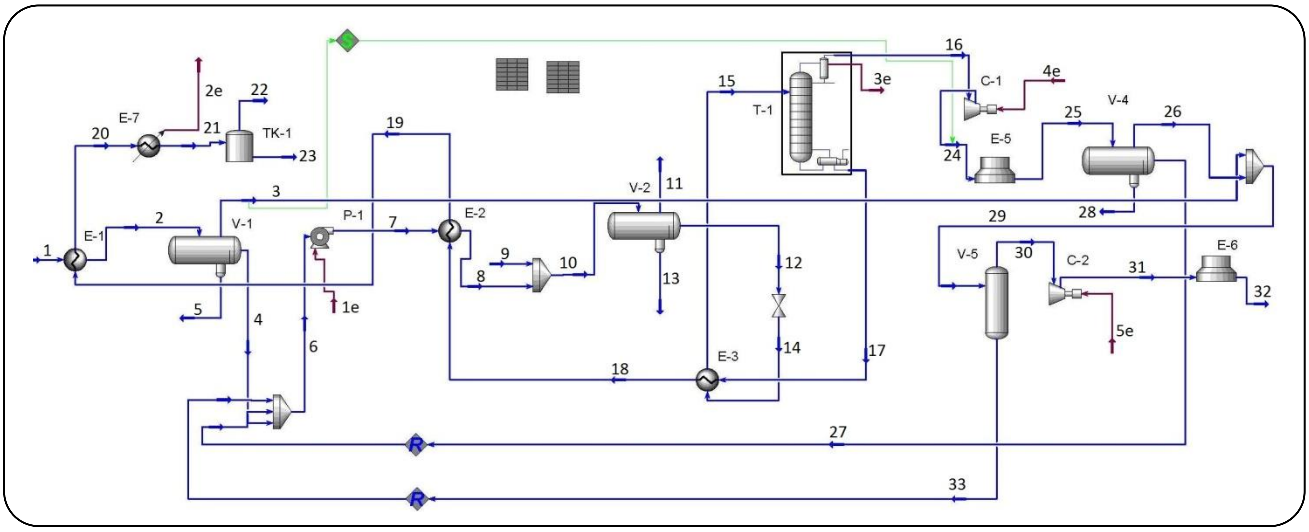

Fig. 2: Case 1 process flow diagram.

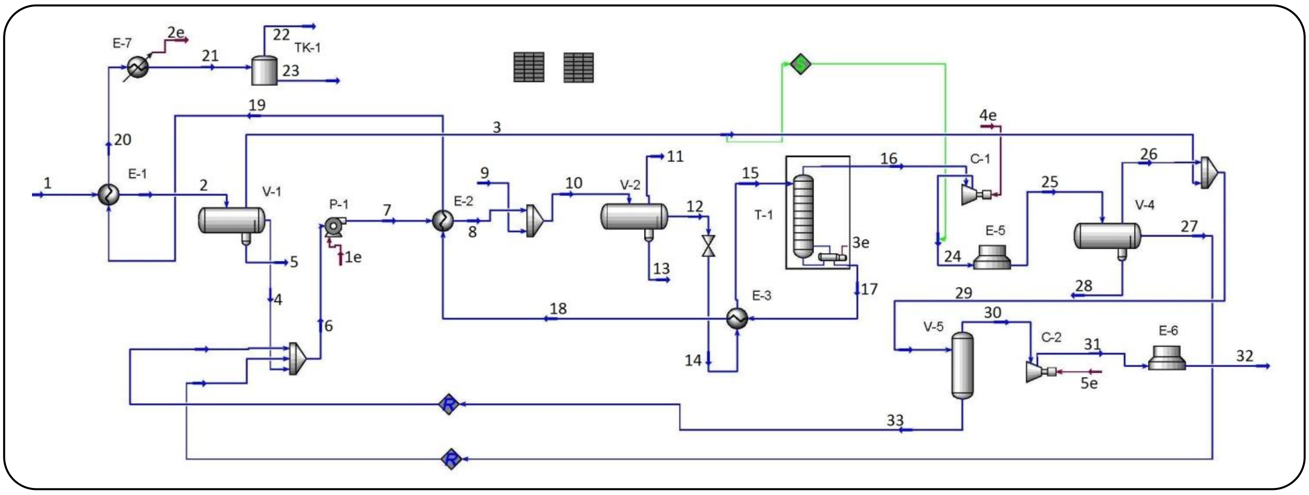

Fig. 3: Case 2 process flow diagram.

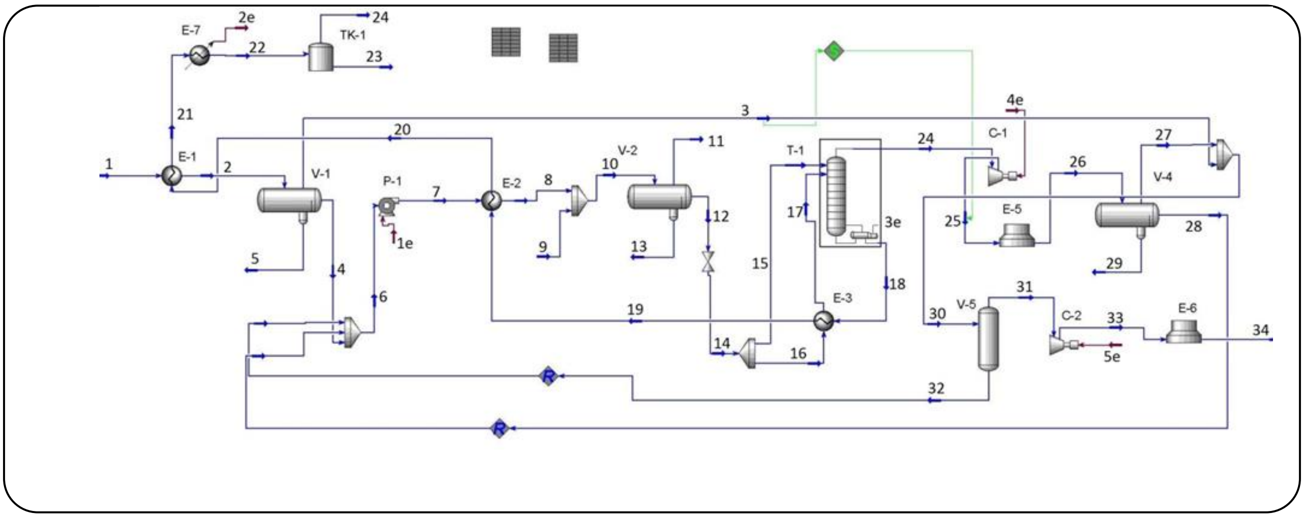

Fig. 4: Case 3 process flow diagram.

Case 4:

Case 4 is similar to Case 1, except that the stabilizer column operates at lower pressure. It is recommended that the operating pressure of the column be about 6 bara. This alternative has its own advantages and disadvantages. As the operating pressure of the stabilizer column is low, the design pressure of the column is decreased. Following that the thickness of the column and related equipment is decreased. This results in a lower capital cost for the unit. However, when the operating pressure is low, the dew point of the overhead product in the condenser is automatically decreased. As a result, it may make it impossible the use of the cheapest cooling system (such as a cooling fan) to condense the overhead product in the condenser. Therefore, the operating cost of the unit is increased.

Simulation Assumptions

To apply the defined pseudo-components in Table 1 to Aspen HYSYS, a hypothetical group in the simulator was defined. The hypothetical group helped the desired condensate pseudo-components in the units be generated. In this way, at least two important properties for each pseudo-component, including molecular weight and density, were needed to complete its definition. Generally, the Equation of State (EOS) is needed to determine the thermophysical properties required for pseudo-components. For the purpose of modeling, Sour Soave-Redlich-Kwong (Sour SRK) EOS was chosen. The sour SRK model combines the Wilson and SRK model and is able to be used for any process containing acid gases, water, and hydrocarbons. The sour SRK uses the SRK equation of state to correctly present the heavy and light ends system. It also uses the Wilson model to account for the ionization of the H_2S and CO_2 in the water phase. Additionally, the sour SRK is applicable for the temperature in the range of 20-140 ℃ and the pressures higher than 50 psi. However, one of the problems in the EOS package is that the liquid density is not estimated accurately for modeling purposes. To deal with this problem, instead of the EOS, COSTALD method was used only for determining liquid density. In addition, the Lee-Kesler correlation was chosen for defining the heat capacity and enthalpy of pseudo-components.

Equipment Sizing

The following assumptions were made when calculating the equipment's properties:

- The vessels were sized according to the approaches defined by Svrcek and Monnery [15-16].

- The sizes of tanks were determined through Gorji's and Jari's methods [17].

- The areas of Heat exchangers were determined through the method presented in *Peters et al.* manuscript [18].

- The power of compressors, fans, and pumps was determined with the help of Aspen Hysys v. 7.3.

- An estimation of the tower diameter and height was made by a method defined in Total practice. The detailed design of the column, such as the type of trays and tray spacing, was conducted through NIOEC-SP-00-50.

- The kettle-type reboiler for the present light hydrocarbon column was used. The reboiler was heated by Medium Pressure Steam (MPS).

- The condenser, if any, was installed on the overhead of the column to recover liquid product and provide internal column reflux. The type of the condenser cooler which was used at the overhead of the column depends on the total heat exchanged in the cooler itself. Usually, a shell and tube heat exchanger is utilized when the dew point of distillate is in the range of 40-50 ℃, while an air cooler covers the range of 60-70 ℃. This rule of thumb dominates throughout this procedure during the preliminary design of the condenser's cooler. It should be mentioned that the design pressure of the column should be set so properly that the dew point of the distillate remains between 40-70 °C.

Economical evaluation

The main part of this article is allocated to the economical evaluation of the different condensate stabilization configurations. The financial analysis plays a significant role in assessing process viability. To know whether a plant loses money or earns money, a primary economical study is needed. By knowing certain factors in the condensate stabilization unit, including the fixed capital cost, cost of manufacture, cost of raw material, and revenue, the economic performance of the unit is determined.

Except for the column cost, which is referred to [18], the cost correlations used for the other equipment in each of the cases come from Smith's manuscript [19]. In this work, not only a preliminary sizing of the equipment, such as vessels, was studied, but also a preliminary assessment of the mechanical design was applied. Materials of construction have a significant influence on the capital cost

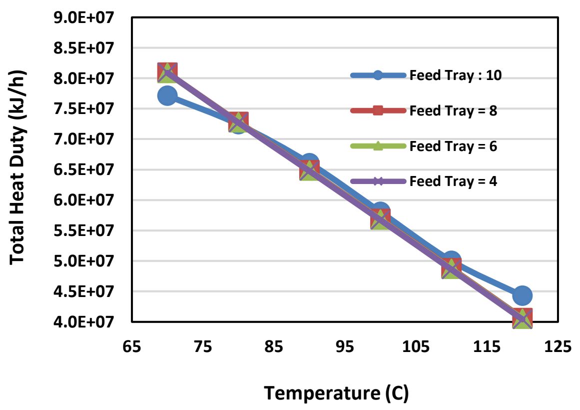

Fig. 5: Total heat duty of the plant in Case 1 at different feed temperatures.

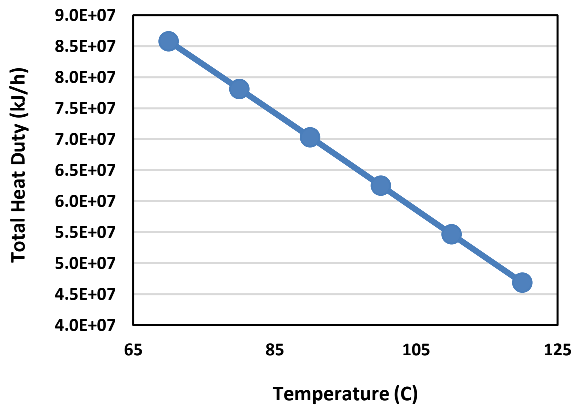

Fig. 6: Total heat duty of the plant in Case 2 at different feed temperatures.

of the equipment, and they were chosen by using the practice documents published by TOTAL practice. Design pressures and temperatures of each equipment, which affect the fixed capital cost, were determined through the Iranian standard of NIOEC-SP-00-50. All defined costs are on the basis of US$ in the present work. They were updated from the year 2001 with the index cost of 150 to 2018 with the index cost of 220.

The prices of materials including unstabilized condensate stabilized condensate, Gas, MPS, and Cooling Water (CW) are $ 50/bbl, $ 80/bbl, $ 3/1000SCF, $ 15.3/ton and $ 0.027/m^3, respectively. It was assumed the construction of each of the plants last two years, and the plants operate for ten years after the construction. A five-year depreciation schedule is assumed. Taxation rate defaults to 40%. The discounted cash flow rate is fixed at 8%. Using these mentioned criteria, each unit was evaluated economically. Net Present Value (NPV) was one of the used approaches to evaluate each of the cases. The greater the positive NPV for a case, the more economically attractive it is.

RESULTS AND DISCUSSION

To find the optimum number of trays, together with the optimum reflux ratio, in the stabilizer column for each case, a short-cut distillation column (available in Aspen Hysys) was used. According to the simulation results, each column needed 20 trays with an average tray efficiency of 60%. In addition, a reflux ratio of 0.1 should be specified for Case 1 and Case 4. Although the feed tray could be indicated through the short-cut distillation column, several runs were conducted so that the optimum feed tray for each case was found. The optimization process was based on the lowest energy consumption by each plant.

To find the best condition in Case 1, two variables were simultaneously examined. The first variable was the feed tray of the stabilizer column, and the second variable was the feed temperature. Fig. 5 shows the examined feed tray number, alongside the range of the feed temperatures. s seen, with increasing the feed temperature, the total heat duty of Case 1 decreases. Herein, the total heat duty has included all the used MPS, air cooler duty, and CW in the unit. Note that the feed temperature increases through the column stabilizer preheater. According to the simulation results, a feed temperature higher than 110 °C is not accessible from an operation point of view. Unlike the feed temperature, the feed tray does not have a big effect on the unit's performance. In this study, feed tray 4 was chosen for more analyses for Case 1.

The feed tray was fixed in Case 2 since it was assumed that the feed entered from the top of the column. As shown in Fig. 6, only the feed temperature was examined in the stabilizer column in Case 2. Similar to before, an increase in the feed temperature reduces the total heat duty of the plant. Due to operation constrain, a temperature higher than 110 °C is not possible, so the feed temperature of 110 °C was regarded as the optimum value in this case.

As mentioned, Case 3 is a modification of Case 2. A portion of the feed is sent to the top of the column. The split ratio, the residual feed tray, and temperature were examined through 144 simulation runs. Fig. 7 shows the specific results related to the residual feed tray 8. According to the figure, one could understand that the higher temperature culminates in lower heat duty in Case 3.

In addition, the figure shows the lowest duty happened when 10% of the flow is routed to the top of the column. In this work, the split ratio of 15% at the residual feed temperature 120 °C entering the eighth tray was chosen as an optimum condition. Note that split ratios lower than 15% with residual feed 120 °C are not reasonable in operation due to physical constraints.

The column in Case 4 operates at lower pressures with respect to that operates in Case 1. Some analyses were done to know in which condition the plant in Case 4 works better. Fig. 8 represents the results of the simulation runs. It is totally clear that the heat duty used by Case 4 is less than the others. The feed tray is not an effective parameter compared with the feed temperature. Normally the feed tray with temperatures higher than 75 °C leads to an undesirable heat cross in the exchangers. Therefore, regarding tray 8 as the feed tray, the optimum input temperature for the feed tray was assumed to be 75 °C.

The column pressure profile for each mentioned case is shown in Fig. 9. From the head to the bottom of the columns, the pressure of the fluid increases tray by tray. The pressure difference between the two sides of the columns for all the cases is roughly 100 kPa. All the cases have the same pressure profile trends. However, as the pressure of the entering feed in case 4's column is lower than In the other cases, its pressure profile throughout the column is lower than the others.

The amount of cold liquid (Reflux and non-reflux liquid) entering the columns, the temperature of the entering liquid and the feed tray of the columns affect the temperature profiles of the columns. The temperature profiles of the fluid inside the columns at every tray are represented in Fig. 10. As seen, the temperature of the column in case 2 is gradually decreased from the bottom tray to the head tray. However, there is no gradual decrease for case 1, case 3, and case 4. The downward trends in the mentioned cases get sharp at the feed trays where cold liquid falls down and decrease the temperature of the hot gas bubbling upward. Moreover, the position of the temperature profile for case 4 is almost 20 °C lower than the three other cases. Due to the low-pressure profile in case 4, the heavy and intermediate components can easily change to vapor and the equilibrium at each tray is more accessible at a lower temperature.

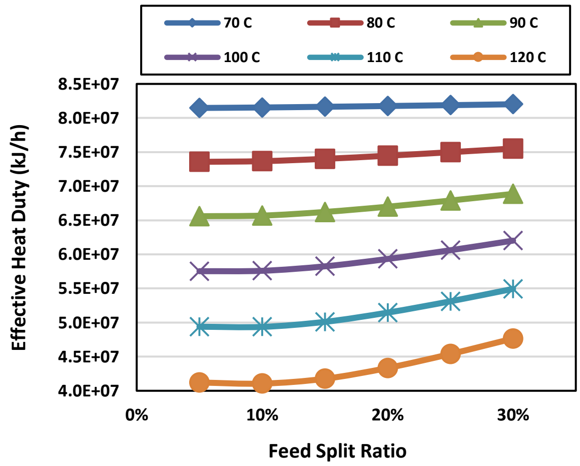

Fig. 7: Total heat duty of the plant in Case 3 at different feed temperatures and feed split ratios.

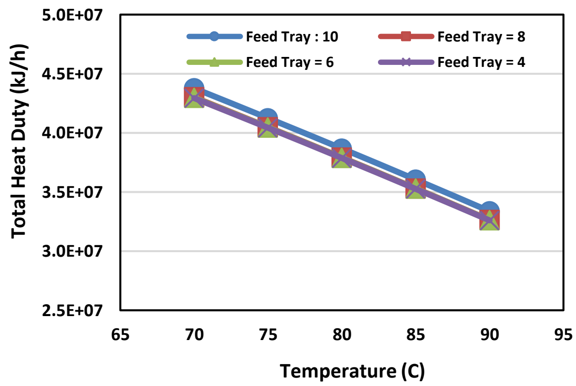

Fig. 8: Total heat duty of the plant in Case 4 at different feed temperatures.

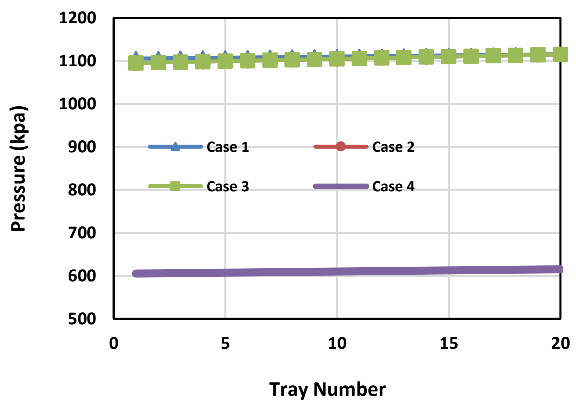

Fig9: Fluid pressure at each tray number in the four cases.

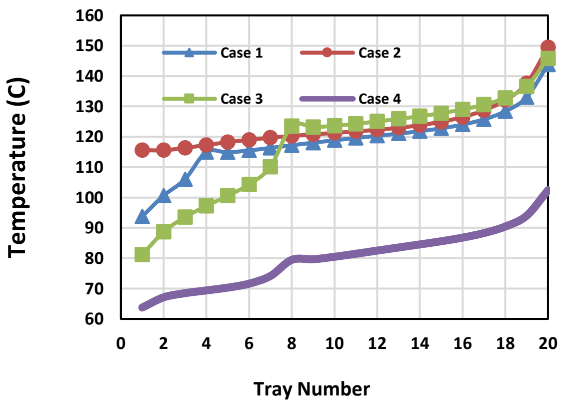

Fig. 10: Fluid temperature at each tray number in the four cases.

It is possible to observe the components' mole fractions at each tray in the columns when using Aspen Hysys. This helps to better understand the separation mechanism by which the columns make the product specification. Unfortunately, according to Table, there are too many components capable of depicting their mole fractions at each tray. Therefore, just light end components which are most important in the condensation stabilization were regarded for more study in this section. Fig. 11 shows vapor component profiles for each case. As can be seen, the mole fractions of Butane and Propane are decreased from the bottom to the head of the column. Unlike, Butane and Propane, Methane and Ethane have upward trends from bottom to head of the column. As shown in the figures connected to each of the cases, the mole fractions of the components are sharply changed at the feed trays. In fact, the mass transfer happening at the feed trays is so critical. For example, the mole fraction of methane in gas is increased by 80% when the gas passes from tray 10^{th} to 8^{th} in case 3. However, Methane is increased only by 10% when the gas passes from tray 8^{th} to the head of the column (the rest of the trays). In fact, the gas hardly gets leaner and leaner from heavy components when passing from the feed tray to the overhead in order that the desired product specification is obtained. Case 2 does not have any input from the middle of the column; therefore, the most of purification for the upgoing gas happens at the top of the column where the cold liquid (the feed) enters the column. When the gas passes from 3rd tray to the overhead, the mole fraction of Methane is increased by 88%. After that, there is no more purification for Methane. In other words, there is no chance of making the gas leaner from heavy components. Therefore, to reach the desired product specification, case 2 needs more reboiler energy consumption. While the required heat duty for case 3 is 2.9×10^7 kJ/h that for case 2 is 3.4×10^7 kJ/h (17% higher than case 3). Please note that distillation is highly energy-intensive and can consume most of the total energy in a typical condensate stabilization unit. In addition, in a stabilizer column, the major operating cost is reboiler energy consumption. Regarding this, Case 2, with high reboiler energy consumption, seems the unit with the most operating cost. This will be proved economically in the proceeding paragraphs.

A preliminary feasibility study for the mentioned processes was conducted. Firstly, typical equipment capacities which affect the capital costs of the units were calculated. Afterward, the cost of the equipment and other capital investments, based on the found capacities, were estimated. Aside from that, it was tried the Operating Expense (OPEX) value for each case was computed. The operating costs cover the raw materials, operating labor costs, maintenance, and utility costs. Here the utility costs cover the heating and cooling systems. Table 2 shows the cost of all equipment used in the cases on the basis of the designed parameters introduced in the previous section. Unlike what was expected previously, case 2 is most expensive than the others. Although case 2 does not have a condenser and seems simpler to operate, the cost of other equipment such as its heat exchangers, compressors, and reboiler is totally higher than the other cases. Case 1 requires more equipment to purchase, install, and operate, but this additional cost is justified by reducing cooling and heating duties. Case 4 is the cheapest plant largely because its column operates at lower pressure. Lower investment belongs to case 4 because its lower design pressure decreases the cost of its equipment.

Fig. 12 shows the cumulative discounted cash flow pattern for the four cases during 12 years. The early stages of the projects which consist of development, design, and other preliminary work take 2 years. This passes without any immediate return. The production starts at the year 2 when revenue from sales begins. The breakeven point of the projects is about in the year 3. Towards the end of the projects, the units continue their operation. The salvage values of the units

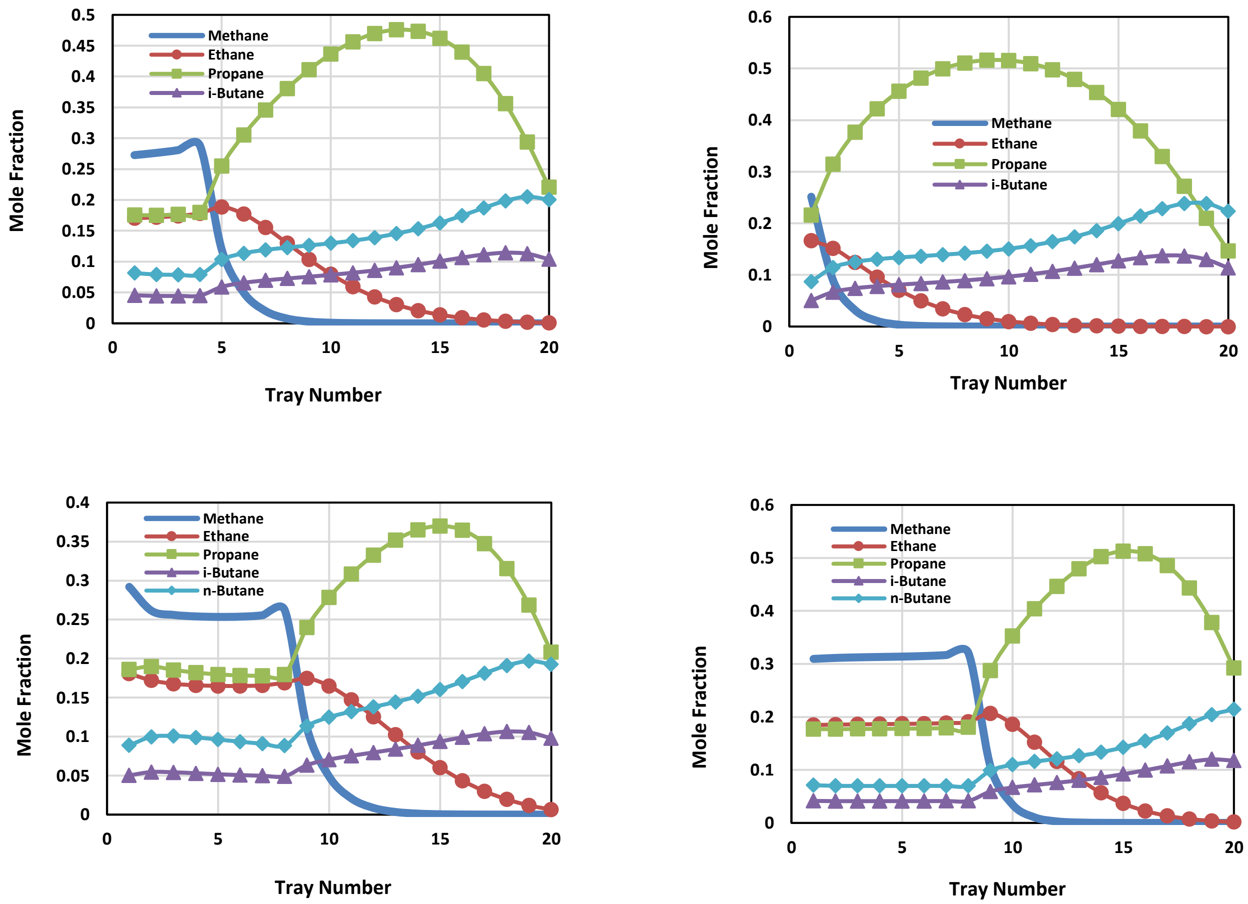

***Fig. 11:** Mole fraction of light end components at each column tray number for (a) Case 1, (b) Case 2, (c) Case 3 and (d) Case 4.*

were added to the final cash flow at the end of the projects. Table 3 compares the amounts of produced stabilized condensate, produced gas, used MPSand used cooling water in these four cases. According to Table 2 and Table 3, one could understand that case 4 not only has the lowest fixed capital costs but also uses the lowest energy (including cooling water and MPS) as compared to the other cases. As the column pressure is lowered, the separation efficiency is improved. This happens due to the fact that the required heat for reboiling reduces, and the separation between light and heavy key components is made better. Case 4 seems be the optimum choice by culminating in the most NPV during 12 years. Fig. 12 confirms this fact by comparing the cumulative discounted cash flows.

Despite the mentioned advantage, case 4 is more susceptible to flooding at its column and tends to increase the cost of reflux cooling and vapor recompression. Therefore, care should be taken when specifying the column pressure.

Nowadays, it is clear that the price of oils is changing. The inconstant price makes any decision about the design of oil production plant hard. To know any possible effects of the product price on NPV, the economic performance of each case was examined at five varying stabilized condensate and gas prices. It was assumed the entering feed in the plants has its constant price of $ 50/bbl throughout the analyses. Fig. 13 and Fig. 14 represent the fact that while case 4 is the superior one under all conditions, case 2 is counted as the weakest one. Furthermore, it is obvious that when the Free On Board (FOB) of stabilized condensate price is $ 70, none of the cases will have positive NPVs. This means that their economical performances are not acceptable.

CONCLUSIONS

An industrial-scale simulation of a condensate stabilization unit has been done using Aspen HYSYS to find the configuration which benefits the most.

Table 2: Capital cost of main equipment defined in each case.

	Cost ($)			
	Case 1	Case 2	Case 3	Case 4
Heat Exchangers	1330302	1356272	1491908	1099325
Separator	4086571	4114479	3988992	3897504
Compressors (*)	4282110	4431725	4176447	4467745
Air Coolers (*)	2014190	2325573	1778361	1587159
Pump (*)	3699201	3718504	3632713	3580045
Tanks	3808346	3808346	3808346	3808346
Tower (**)	4683237	4197626	3697057	3135831
Total	23,903,957	23,952,525	22,573,825	21,575,954

** With stand by.*
*** The whole package.*

Table 1: Products and consumptions defined in each case.

Products and Consumptions	Case 1	Case 2	Case 3	Case 4
Stabilized Condensate (bbl/day)	33990	33890	33940	34020
Gas (scfd)	25.26	25.44	25.33	25.25
MPS (kg/h)	-15921	-17233	-14226	-13314
Cooling Water (kg/h)	-109323	-135546	-75460	-85024

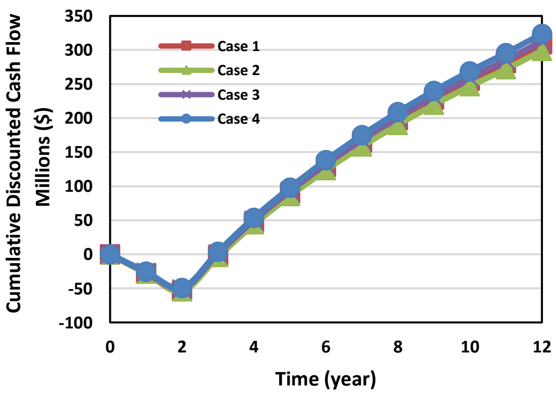

Fig. 12: Cumulative discounted cash flow pattern for the defined cases.

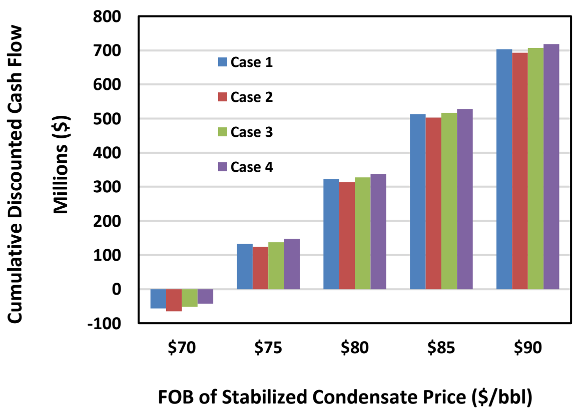

Fig. 13: The variation of NPV with constant raw material prices in the proposed cases for different stabilized condensate prices.

Four different configurations, in which the effect of operating conditions such as stabilizer feed tray, feed temperature, and pressure were examined, were chosen for analyzing the condensate units. It was requested that the RVP of the stabilized condensate be set as 10 psi (summer case). Case 1 is the stabilized configuration unit used in South Pars.

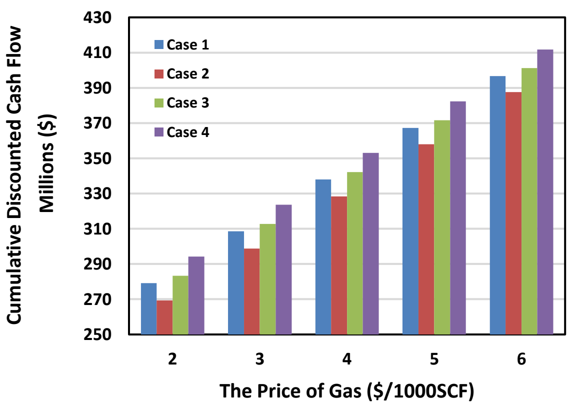

Fig. 14: The variation of NPV with constant raw material prices in the proposed cases for different gas prices.

Results show that there are other alternatives with higher benefits. Case 2 has the highest reboiler and cooler duties in comparison to the other cases, so its high OPEX leads it to the worst case. Case 3, which is a modification of case 2, seems more economical, but Case 4 with the highest cumulative discounted cash flow during ten years of production is regarded as the best configuration.

Received : Aug. 29, 2020 ; Accepted : Dec. 7, 2020

REFERENCES

[1] Stewart, M., "Surface Production Operations: Design of Gas-Handling Systems and Facilities", Gulf Professional Publishing, (2014).

[2] Shirazinia S.R., Semnani A., Nekoeinia M., Hasanpour F., Characterization of South Pars gas condensates by Physicochemical Properties And Multivariate Modeling, *J. Iran Chem. Soc.*, **17**: 663–669 (2020).

[3] Rahmanian N., Sakinah Bt Jusoh L., Homayoonfard M., Nasrifar K., Moshfeghian M., Simulation and Optimization of a Condensate Stabilisation Process, *Journal of Natural Gas Science and Engineering*, **32**: 453-464 (2016).

[4] Moghadam N., Samadi M., Gas Condensate Stabilization Unit: Different Design Approaches, *International Journal of Chemical Engineering and Applications*, **3(6)**: 461 (2012).

[5] Watkins, R.N., "Petroleum Refinery Distillation", (1979).

[6] Salehi G., Amidpour M., Salehi A., Maleki M., Ghorbani, B., Optimization of Distillation Column Operation by Simulated Annealing, *Gas Processing Journal*, **1(2)**: 49-63 (2013).

[7] Sharma R., Jindal A., Mandawala D., Jana S.K., *Industrial & Engineering Chemistry Research*, **38(6)**: 2411-2417 (1999).

[8] Rahmanian N., Ilias I.B., Nasrifar K., Process Simulation and Assessment of a Back-Up Condensate Stabilization Unit, *Journal of Natural Gas Science and Engineering*, **26**: 730-736 (2015).

[9] Uwitonze H., Hwang K.S., Lee I., Modelling and Improving Natural Gas Condensate Process with Stripping and Heat Integration, *Chemical Engineering and Processing: Process Intensification*, **118**: 71-77 (2017).

[10] Zhu L., Liu H., Zhang Z., Pu Y., Simulation Analysis of Stripping Fractionation Process of Gas Condensate Treatment and Practical Application, *Journal of Natural Gas Science and Engineering*, **34**: 216-225 (2016).

[11] Tahouni N., Khoshchehreh R., Panjeshahi M.H., Debottlenecking of Condensate Stabilization Unit in a Gas Refinery, *Energy*, **77**: 742-751 (2014).

[12] Tavan Y., Hosseini S.H., Ahmadi G., Energy and Exergy Analysis of Intensified Condensate Stabilization Unit with Water Draw Pan, *Applied Thermal Engineering*, **155**: 49-58 (2019).

[13] El-Eishy M.M., Abdelalim G.M., Aboul-Fotouh T.M., Simulation Study and Optimization of the Operating Variables Affecting the Performance of an Existing Condensate Stabilization Unit, *Petroleum & Coal*, **61(6)**, (2019).

[14] Khalili-Garakani A., Kasiri N., Ivakpour J., Feasibility Study on Replacement of Atmospheric Distillation Column with New Sequences in a Natural Gas Condensate Refinery, *Computers & Chemical Engineering*, **135**: 106745 (2020).

[15] Monnery W.D., Svrcek W.Y., Successfully Specify Three-Phase Separators, *Chemical Engineering Progress*, **90**: 29-29 (1994).

[16] Svrcek W., Monnery W., Design Two-Phase Separators within the Right Limits, *Chemical Engineering Progress*, **89(10)**: 53-60 (1993).

[17] Gorji A.H., Jari H.K., Improve Selection and Sizing of Storage Tanks: Guidelines Help Engineers Select the Best Tank Design for Plant Applications, *Hydrocarbon Processing*, **85(10)**: 95-101 (2006).

[18] Peters, Max S., Klaus D. Timmerhaus, and Ronald E. West., "Plant Design and Economics for Chemical Engineers", McGraw-Hill New York, (1968).

[19] Smith R., "Chemical Process: Design and Integration", John Wiley & Sons Inc., (2005).