Korean J. Chem. Eng., 40(7), 1575-1586 (2023)
DOI: 10.1007/s11814-023-1411-5

pISSN: 0256-1115
eISSN: 1975-7220

Optimization analysis of the absorption-stabilization process for fluid catalytic cracking unit

Muhammad Saddam Hussain*, Ashfaq Ahmed,†, Liu Yibin*, Muhammad Nadeem Amin***, Tahir Zahoor***, Muhammad Afnan Saleem****, Kosan Roh*****, Murid Hussain******, Muhammad Saifullah Abu Bakar*******, and YoungKwon Park********,†**

*Key Laboratory for Thermal Science and Power Engineering of Ministry of Education, Beijing Key Laboratory of CO2 Utilization and Reduction Technology, Department of Energy and Power Engineering, Tsinghua University, Beijing, 100084, P. R. China
**Institute for Sustainable Industries and Liveable Cities, Victoria University, Melbourne 8001, Australia
***Department of Chemical Engineering, NFC IET, Multan 66000, Pakistan
****State Key Laboratory of Multiphase Flow in Power Engineering, Xi'an Jiaotong University, Xi'an 710049, P. R. China
*****Department of Chemical Engineering and Applied Chemistry, Chungnam National University, Daejeon 34141, Korea
******Department of Chemical Engineering, COMSATS University Islamabad, Lahore Campus, Defence Road, Off Raiwind Road, Lahore 54000, Pakistan
*******Faculty of Integrated Technologies, Universiti Brunei Darussalam, Jalan Tungku Link BE1410, Brunei Darussalam
********School of Environmental Engineering, University of Seoul, Seoul 02504, Korea
(*Received 24 November 2022 • Revised 28 February 2023 • Accepted 2 March 2023*)

Abstract−The absorption-stabilization process (ASP), an important part of petroleum refinery used in the end-use products of petroleum (such as stable gasoline, liquid petroleum gas, and dry gas), is energy-intensive and has low product quality. Aspen Plus process simulator was used for the development of the ASP process model. The developed process model was validated with the actual plant data. The validated model was used to optimize to minimize the cost of the ASP. This work shows that the optimization analysis of the ASP can further improve the product quality and reduce thermal energy consumption. In the new process, changing feeding parameters of supplementary absorption oil, stripping tower intermediate reboiler, and feeding position of stabilization tower reduced the C_3 contents of dry gas considerably and lowered the C_2 and lighter contents of LPG. Additionally, the new process saved 1.32 MW of thermal energy consumption compared with the existing process. The operating cost has been reduced from 10.921 million USD annually to 9.830 million USD per year. Furthermore, the cost-saving effect of this optimization is about 9.99% (1.091 million USD per year).

Keywords: Absorption-stabilization Process, Sensitivity Analysis, Process Optimization, Process Simulation, FCC Process

INTRODUCTION

Currently, there is pressure on refinery operation optimization due to environmental policies and regulations [1,2]. The fluid catalytic cracking unit (FCCU) is the largest unit in the refinery for the production of light hydrocarbons and gasoline [3]. Absorption-stabilization process (ASP) is an important part of all the fluid catalytic cracking units. ASP separates the refinery products. ASP is mainly attached to FCCU, where it separates the mixture from the fractionation tower into dry gas, liquefied petroleum gas (LPG) and stabilized gasoline. A typical ASP consists of four towers (absorption, reabsorption, stripping, and stabilization tower) and two recycle streams. In ASP, because the absorption tower and stripping tower provide feed to each other, they have a very close relationship and play a very important role in the refinery by improving the energy consumption, enhancing product quality and ensuring the smooth work operation of stabilization tower. The rich absorption oil from the absorption tower is used as feed for the stripping tower, and desorbed gas from the stripping tower is recycled for the absorption tower. On the other hand, the operating conditions of the two-tower process are much different, such as temperature, pressure, number of stages and feed stage. To minimize the thermal energy requirements and to increase the purity of products, it is essential to study the effect of feed, and operating parameters of the absorption tower, reabsorption tower, stripping tower, and stabilization tower to improve the performance of the entire ASP.

The ASP has three feeding methods for stripping towers (desorption column): cold feed, hot feed, and hot and cold [4,5]. The temperature of the cold feed is between 30 to 40 °C. Because the temperature of cold feed is relatively low, so the temperature at the top of the stripper is low. It makes the temperature of the desorbed gas low, and the amount of desorbed gas is small. The processing capacity of the absorption tower is reduced, and the low temperature contributes to the absorption. However, this makes the stripping tower bottom reboiler duty increase [6]. The temperature of the hot feed is from 75 to 80 °C. Due to the higher temperature of the hot feed,

†To whom correspondence should be addressed.
E-mail: ashfaq.ahmed@vu.edu.au, ykpark0426@gmail.com

the temperature at the top of the stripping tower is high, which causes the temperature of the desorbed gas to be high. As a result, the amount of desorbed gas becomes large, so the treatment volume of the absorption tower, therefore the load of the compressed rich gas condenser is increased. In addition, due to a large amount of desorbed gas C_3, C_4 components are in the absorption tower. The circulation between the stripper and the compressed rich gas condenser increases the absorption tower's processing capacity and affects the product quality. But the advantage is that the load of the reboiler at the bottom of the stripper is reduced [7]. To effectively use the advantages of hot and cold feeds, many refineries use the hot and cold double-feed method shown in Figs. 3-5. The hot and cold dual-feed divides the condensed oil into two feeds in this method. One feed is the cold feed directly from the oil and gas separator. The other is exchanged with the stable gasoline at the bottom of the stabilizing tower as a hot feed. This method has a significant advantage in refining the product quality and reducing energy consumption [8,9]. Since one of the two feeds is cold feed and the other is hot feed, they have not only the advantages of cold feeding but also of hot feeding. But because of the different temperatures and same composition groups, the same two feeds enter different positions of the stripping tower, which destroys the reasonable concentration distribution of the components in the tower. However, this method has problems with back mixing and axial mass transfer, so it still needs improvement [5]. Therefore, to combine the advantage of both cold and hot feed methods, feed splitting (hot and cold feed method) was adopted to reduce energy consumption and C_3 contents in dry gas. In the hot and cold feeding method for stripping tower [5], there is a problem of back mixing of heavier components that desorb and are found in the dry gas or increase the flowrates to the flash tank and stripping tower; as a result, the equipment size also increases and that affects plant economy.

Bandyopadhyay et al. [10] mentioned that having more than one stream with different compositions and temperatures was beneficial to enhance the mass transfer and separation efficiency of the distillation column. To avoid the back mixing of liquefied gas components and continuous circulation in the equipment, some industries use the secondary condensation process to provide the hot and cold feeds to the stripping column with the different compositions of hot and cold feed. In the secondary condensation process, the gas from the first condensation tank enters the secondary condensation tank, where the gas phase is sent to the primary absorption column, and the liquid phases of both condensation tanks feed to the stripping column, whereas the liquid phase from first condensation tank is used as hot feed while the liquid from secondary condenser is used as cool feed to the stripping tower. The best feed location of the distillation column is where it has the same composition of internal traffic as the feed streams, which can significantly improve the mass transfer in the separation process [11].

After this, many researchers searched about optimizing the absorption-stabilization process by using different feeds and different designs of heat exchanger networks to increase the purity of products and enhance the process economy. Nowadays, industries use APS having two absorbers and two strippers to effectively separate components. Sensitivity analysis is also an effective method to optimize the chemical process [12]. It is common for dry gas to have a high percentage of C_3/C_{3+}. A dry gas that has a C_3/C_{3+} level of less than 2% is required. The C_3/C_{3+} content of dry gas was reduced by altering the supplementary absorption oil flow rate of the existing ASP proposed by Li et al. [13]. A stabilization tower feeds the material into the stripper, which is heated in the middle reboiler. Rather than using stabilized gasoline, light gasoline is removed from the stabilization tower and used as a supplement absorber. A refrigerator (An aluminum bromide refrigerator powered by low-grade hot water) was used to cool the absorber feed to improve the absorption phenomenon. For the absorption column, they used computer simulation to derive the empirical relationship (Eq. (1)) between C_3 and higher component concentrations and temperatures. This method uses 2,432 kW less energy than the original ASP. In addition, the C_3/C_{3+} component of dry gas is reduced to less than 2% of the total dry gas volume. This optimization has resulted in a 17% reduction in energy consumption by operating the absorption tower at a lower temperature.

$$y=0.2282\,T-62.27 \tag{1}$$

Researchers from ETH Zurich optimized and integrated a de-ethanizer into the gas processing unit (GPU) and the FCCU's absorption stabilization process to improve performance and lower C_2 concentration in LPG. When the GPU's de-ethanizer was removed, propylene recovery increased from 94.8 to 99.8%. Extra-absorbent flowrates, stripper feed temperature and C_4 content in stabilizer bottoms as optimization variables were used by Lu et al. in their study. LPG's C_2 content was reduced to less than 0.05%, the concentration of C_3 in off-gas was reduced to less than 1.5%, and the concentration of C_4 in the stabilizer bottom was 3.5%. Thus, the simulation results have shown an increase in propylene production of 1,756 tons per year and plant profit [14]. Zhang et al. developed an algorithm that can improve the end quality products of FCCU. The RVFL structure links the incremental component, mitigating the inaccuracy brought on by the concept drift. Group lasso and L2 regularizations are used to dynamically assign the expansion layer's random parameters based on the incoming data. The RVFLN's universal approximation and rapid convergence qualities allow it to meet the prediction accuracy and real-time demands of online quality prediction, while simultaneously sidestepping the overfitting and catastrophic forgetting issues that plague connectionism models [20]. Both the TE process and the FCC unit modeling tasks benefit more from the WAR-LSTM based technique than the LSTM based approach, making it promising for industrial use. Contrarily, nonuniformly collected data is common in chemical processes, although the model was constructed using consistently sampled data [21].

To the best of our knowledge, maximum work has been done on optimization of ASP. However, there has been no such detailed work on sensitivity analysis of this process. Such as position of stabilization tower feed and intermediate reboiler. In this study, we used the absorption-stabilization process of Changqing Petrochemical's FCCU with 335,000 tons per year rich gas separation capacity. This paper aims to optimize the absorption-stabilization system of Changqing Petrochemical's FCCU by minimizing the thermal energy requirements and increasing the product purity. For this purpose, changing effects of the operational parameters such as temperature, pressure and flowrates of supplementary absorption oil, light die-

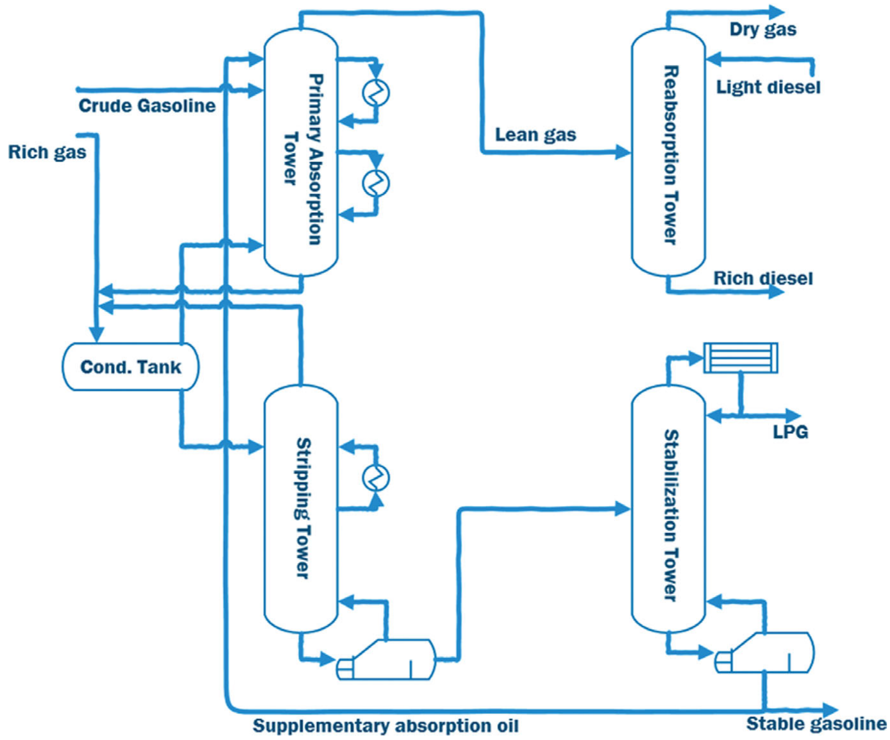

Fig. 1. Typical process flow diagram of the absorption-stabilization process (ASP).

sel, stripping tower feed, the position of intermediate reboiler at stripping tower, and feed position of stabilization tower were studied.

PROCESS DESCRIPTION

The existing ASP is the process of cold feed plus intermediate reboiler, as shown in Fig. 1. It absorbs and stabilizes the crude gasoline and rich gas produced at the top of the fractionation tower and separates a small amount of gas hydrocarbons mixed into the crude gasoline to reduce the gasoline vapor pressure, ensure that it meets the product specifications, and recover the liquefied gas (the main components are C_3, C_4). The absorption stabilization system is the post-treatment section of the FCC unit, which uses the principles of absorption and rectification to separate the rich gas and crude gasoline from the fractionation system into dry gas (C_2 and C_{2-}), liquefied petroleum gas (LPG) includes C_3 and C_4 components, and gasoline products with qualified vapor pressure. The system generally consists of four columns (an absorption tower, stripping tower, reabsorption tower, and stabilization tower).

The rich gas from the fractionation system is compressed by two stages of a gas pressure machine (compressor), injecting washing water into the outlet pipeline to wash the compressed rich gas; after condensing and cooling, it is mixed with the rich absorption oil from the bottom of the absorption tower and the desorption gas from the top of the stripping tower, and then further cooled in a cooler and enters the balance tank (or oil and gas separator/flash tank) which performs balanced vaporization. Non-condensable gas after vapor-liquid equilibrium and condensed oil is sent to the absorption and stripping towers, respectively. To prevent the corrosion of equipment by sulfur, nitrogen, and cyanide, the front and rear pipelines of the cooler and the crude gasoline are poured into softened water for washing, and the sewage is discharged from the balance tank and the crude gasoline washing tank, respectively. The non-condensable gas from the balance tank enters the bottom of the absorption tower, and from bottom to top, it comes into contact with the absorbent crude gasoline; the supplementary absorption oil (stabilized gasoline in reverse), and most of the rich gas C_3 components and a small amount of C_2 components are absorbed by the liquid phase components. Absorption is an exothermic process, and low temperature favors the absorption operation. To make the gas and liquid traffic of the whole tower evenly distributed along with the height of the tower, it is necessary to set up two intermediate pumps around in the absorption tower. This lower temperature will absorb more gases and prevent C_3 and C_+ components from entering lean gas, hence enhancing the quality of dry gas. The lean gas from the top of the absorption tower carries a small amount of absorbent and passes through the pressure control valve to the absorption tower to recover the lean gas with light diesel (fraction of FCC main fractionation column) as the absorbent. The dry gas from the top of the reabsorption tower is rich in ethylene, and the rich liquid at the bottom of the reabsorption tower is returned upstream of the fractionation system.

The role of the stripping tower is to desorb the C_2 in the rich absorption oil components, also known as de-ethanizers. The condensate drawn from the bottom of the balance tank (condensed oil) is pumped into the top of the stripping tower and desorbed under the action of the reboiler at the bottom of the tower (ethane enhances the desorption effect). At present, most refineries generally control the desorption rate at 100%. A large amount of desorption gas at the top of the tower is not acceptable. By avoiding desorption of some C_3, and C_4 components are led out from the top of the tower, mixed with compressed rich gas and rich absorption oil, cooled and entered into the balance tank, and then sent to the absorption tower. The bottom of the stripping tower is de-ethanized gasoline, which exchanges heat with the stable gasoline from the stabilization tower and enters the middle of the stabilization tower. The function of the stabilizing tower is to separate C_4 from the de-ethanized gasoline. De-ethanized gasoline enters under the action of the reboiler at the bottom of the stabilizing tower, the C_4 and light components (known as LPG) are steamed out and obtained at the top of the tower, and the stabilized gasoline at the bottom of the tower split into two parts, one is used as a product output device after heat exchange and cooling, and, the other part is further cooled as a supplementary absorption oil for the absorption tower.

To control the operating pressure of the stabilizing tower, sometimes non-condensable gas must be discharged. Compared with liquefied gas, the saturated vapor pressure of ethane is higher. Excessive ethane content in the de-ethanized gasoline will reduce the stability of the stabilizing tower's operation and increase the load on the condenser at the top of the tower, which is forced to discharge non-condensable gas and the reduction of C_3, and C_4 affect the rectification effect. So, the de-ethanized gasoline C_2 contents affect LPG and gasoline product quality.

PROCESS SIMULATION

ASP is simulated by using Aspen Plus process simulation software V10 [15,16]. In the petroleum refining system, PENG-ROB

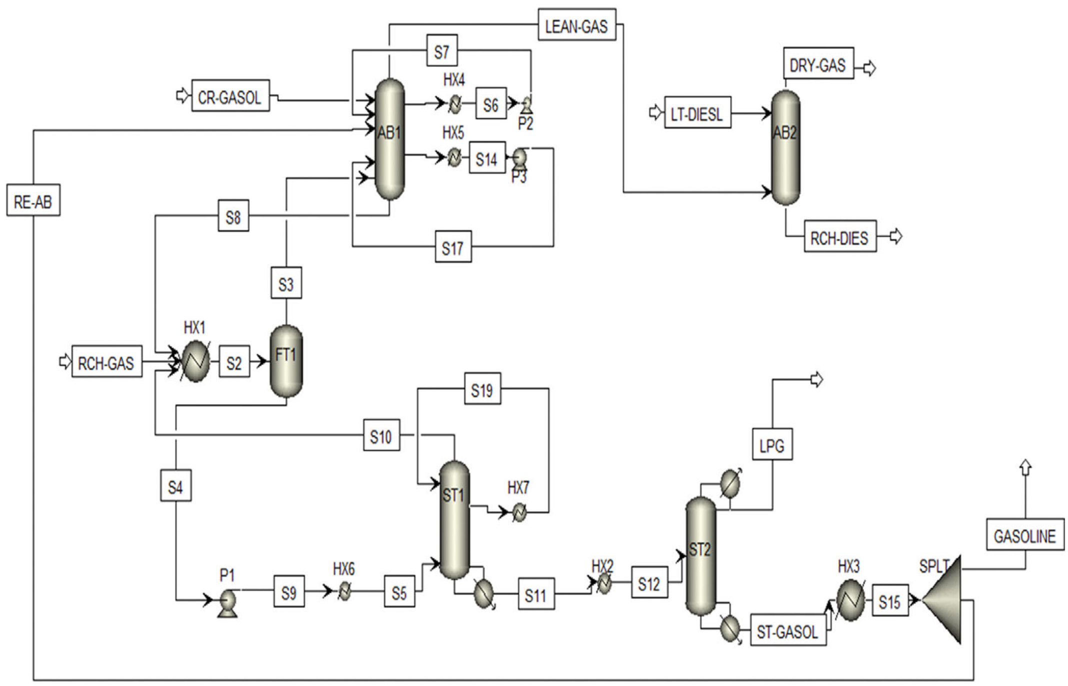

Fig. 2. Simulation flow diagram of the absorption-stabilization process.

Table 1. Parameters of all feed streams of ASP

Stream	P/MPa	T/°C	F/kg·h^{-1}	ρ/Kg·m^{-3}	The TBP distillation curve (%)						
					IBP	10	30	50	70	90	FBP
Rich gas	1.6	42	42,230	-	-	-	-	-	-	-	-
Crude gasoline	1.06	42	86,000	721	41.5	60.0	83.0	106.0	132.0	168.0	187.0
Light diesel	1.03	27.1	14,000	874.2	59.5	202.5	232.5	258.5	295.5	352.0	360.0

Table 2. Composition of rich gas

Component	Vol%	Component	Vol%	Component	Vol%	Component	Vol%
H_2	0.018	CO_2	0.008	Isobutane	0.14	sis-Butene	0.036
O_2	0.002	Ethylene	0.029	n-Butane	0.052	1,3-butadiene	0.004
N_2	0.038	Ethane	0.036	Butene-1	0.036	Pentane	0.002
Methane	0.065	Propylene	0.275	Isobutene	0.046	Isopentane	0.078
CO	0.002	Propane	0.112	Trans-butene-2	0.043	Hexane and above carbon 6	0.002

Table 3. Specifications of absorption tower, reabsorption tower, stripping tower, and stabilization tower

Towers	Pressure/MPa	Temperature/°C	Number of stages	Feed stage
	Top/Bottom	Top/Bottom		
AB1	1.02/1.05	33/41	40	1/6/9/29
AB2	1/1.03	35/42	30	1
ST1	1.12/1.15	40/113	40	1/8
ST2	1.04/1.045	54/164	52	26

Table 4. Comparison of operating parameters

Serial number	Equipment	Simulation parameters of ASP		Existing plant parameters of ASP	
		Pressure (MPa)	Temperature (°C)	Pressure (MPa)	Temperature (°C)
1	Primary absorption tower top	1.02	33	1.02	34.7
2	Primary absorption tower bottom	1.05	41	1.0	41.5
3	Crud gasoline into the Primary absorber	1.06	42	1.06	42.0
4	Supplementary absorption oil	1.065	27.1	1.06	27.9
5	Rich gas into Primary absorber	1.1	36	1.1	35.1
6	Stripping tower top	1.12	40	1.12	40
7	Stripping tower bottom	1.15	113	1.15	110.6
8	Cond. oil into stripping tower	1.7	36	1.7	35.1
9	Reabsorption tower top	1	35	1.03	33.0
10	Reabsorption tower bottom	1.03	42	1.05	43.0
11	Lean gas into re-absorber	1.02	33	1.06	32.3
12	Light diesel into re-absorber	1.03	27.1	1.03	27.1
13	Stabilization tower top	1.04	54	1.04	57.7
14	Stabilization tower bottom	1.045	164	1.065	163
15	De-ethanized gasoline into stabilization tower	1.045	116	1.045	125

and RK-SOAVE are the most accurate thermodynamic method for the simulation of the petroleum refinery [17]. RK-SOAVE is selected as the property method [18], and the RadFrac process model is selected for the absorption tower, reabsorption tower, stripping tower and stabilization tower, Flash-2 is used for the flash tank, and HX-1 is selected for heaters and coolers. This simulation model has been extensively used in literature [13,14,19]. The simulation process flowsheet is shown in Fig. 2. The parameters of all the feed streams, the composition of rich gas, specifications of all towers and comparison of operating parameters of ASP are given in Tables 1, 2, 3, and 4, respectively.

RESULTS AND DISCUSSION

1. Sensitivity Analysis

This section optimizes the main parameters of the absorption stabilization system using the validated process model. Temperature, pressure and flow are the main decision variables in design and production operations. In the following analysis, we also focus on the impact of these important parameters on the product quality and production process.

1-1. Effect of Supplementary Absorption Oil Flow Rate and Temperature

Through the application of sensitivity analysis, the operating parameters of the absorption tower were analyzed. The influence of the supplementary absorbent flow rate and temperature on the quality of dry gas and liquefied gas was obtained, as shown in Fig. 3 (specifying that the molar solubility of C_2 in the bottom of the stripping tower and C_5 in the liquefied gas is constant) and Fig. 4 to Fig. 5 (specifying that the temperature of the stripping tower is constant). For the stripping tower, it is the reverse process of the

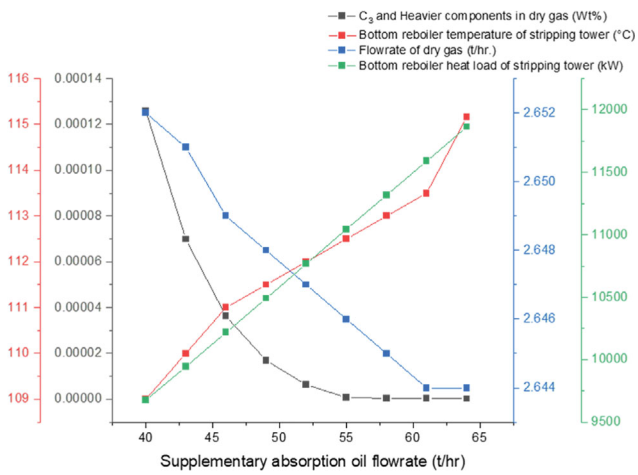

Fig. 3. Effect of supplementary absorption oil on dry gas, bottom reboiler temperature and heat load of stripping tower.

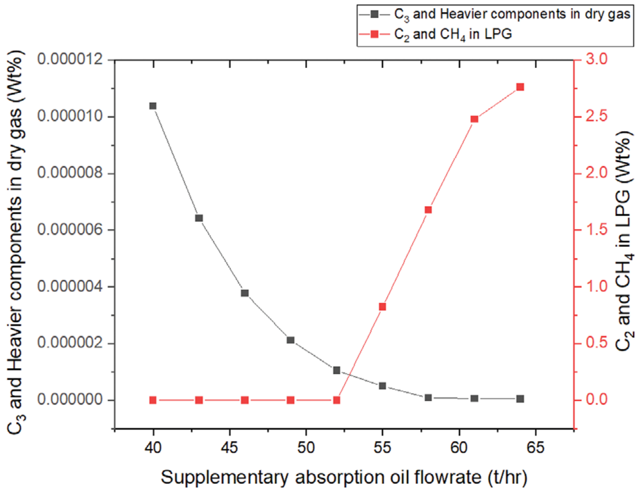

Fig. 4. Effect of supplementary absorption oil flowrate on dry gas and LPG.

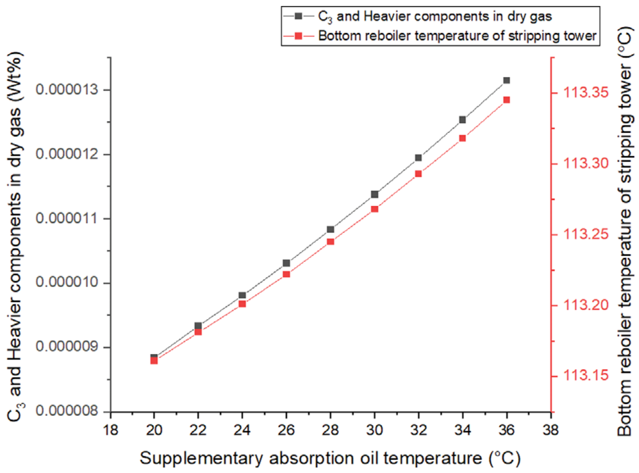

Fig. 5. Effect of supplementary absorption oil temperature on dry gas and stripping tower reboiler temperature.

absorption process, and at the same time, it is equivalent to a rectification process with only a distillation section without a rectification section.

Dry gas is mostly composed of C_1 and C_2 components, whereas the heavier components affect the quality of dry gas, such as C_3 and heavier. As the absorption tower is performing a physical absorption phenomenon, and the quality of dry gas depends on its composition. Fig. 3 shows the influence of supplementary absorbent flow rate on dry gas quality and the effect of supplementary absorption oil where the content of C_3 and above components in the dry gas will first decrease faster. When the supplementary absorbent flow rate reaches 52.5 t/h, it will change slowly, so the supplementary absorbent' flow rate cannot be too high; otherwise, it will significantly increase the operating cost. With increasing the flow rate of the supplementary absorbent while maintaining the same C_2 component at the bottom of the stripping tower, the temperature at the bottom of the stripping tower also needs to be increased, which will increase the heat load of the stripping tower sharply. It can be seen from Fig. 3 that when the flow rate of supplementary absorbent increases, the dry gas production will gradually decrease, which is also after 52.5 t/h. With the supplementary absorbent flow rate increase, dry gas production will no longer be sufficient.

It can be seen from Fig. 4 that when the temperature of the stripping tower is constant at 114 °C, and the supplementary absorbent flow rate is about 61.5 t/h, after that as the supplementary absorbent flow rate further increases, the content of C_3 and above components in dry gas not change. Comparing Fig. 3 and Fig. 4, we can find that when the temperature at the bottom of the tower is controlled at constant, the turning point of the supplementary absorbent flow becomes smaller. When we control the bottom temperature of the stripping tower at 114 °C, the supplementary absorption oil flow rate's influence on dry and liquefied gas is shown in Fig. 4. The content of C_2 and CH_4 in the liquefied gas gradually increases with the increase in supplementary absorption oil flowrate, and when the supplementary absorption oil flowrate reaches 52 t/h, after C_2 and CH_4 content in liquefied gas increases rapidly. This analysis shows that when the operating conditions of other towers remain unchanged, the determination of the supplemental absorption oil flowrate should weigh the quality of dry gas and LPG.

Furthermore, as the temperature of the supplementary absorbent increases, the content of C_3 and heavier components in the dry gas increases monotonically, indicating that the temperature of the supplementary absorbent has a significant impact on the quality of the dry gases, so the temperature of the supplementary absorption should not be too high. On the other hand, as the temperature of the supplementary absorption oil increases, the temperature at the bottom of the stripping tower also gradually increases, and it is also close to a linear shape. It shows that as the temperature of the supplementary absorption oil increases, the heat load of the bottom reboiler of the stripping tower will increase. Fig. 5 also shows that under the most economically optimal conditions, the temperature of the supplementary absorbent should be as low as possible.

1-2. Effect of the Temperature at the Bottom of the Stripping Tower on the Results

In the stripping tower, the operation temperature of the stripping tower has an important influence on the quality of dry gas and liquefied gas. The stripping tower controllable variable is mainly the heat load of the reboiler, that is, the temperature of the bottom of the tower. The temperature of the bottom of the tower has a direct influence on the amount of desorption gas at the top of the tower. Fig. 6 can be obtained by analyzing the temperature sensitivity at the bottom of the stripping tower to the amount of desorption gas produced at the top. Under the condition that other operating parameters remain unchanged, it can be seen from Fig. 6 that as the temperature of the bottom of the stripping tower increases, the amount of desorbed gas at the top of the stripping tower increases slowly, but after 116 °C the slope of the curve increases sharply. Hence, desorbed gas at the top of the tower increases faster. This has a great impact on the load in the tower. If the temperature is too high, the tower's vapor and liquid phase load will be too large. If the maximum load exceeds, the tower will flood, so the temperature at the bottom of the stripping tower should be strictly controlled.

As the temperature at the bottom of the stripping tower increases, the content of C_3 and its heavier components in the dry gas increases,

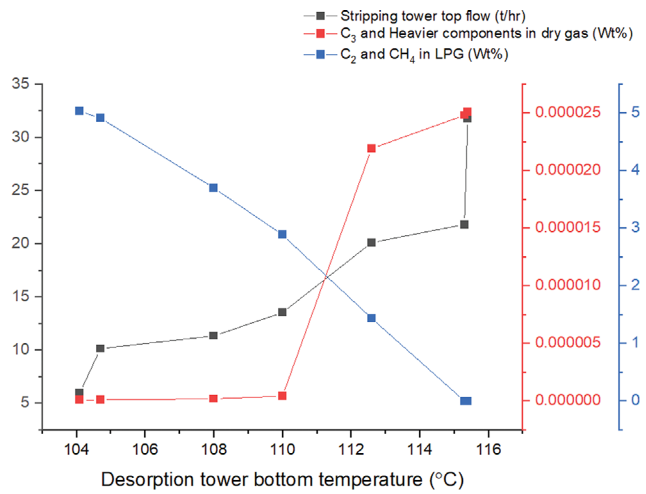

Fig. 6. Effect of the temperature at the bottom of the stripping tower on the results.

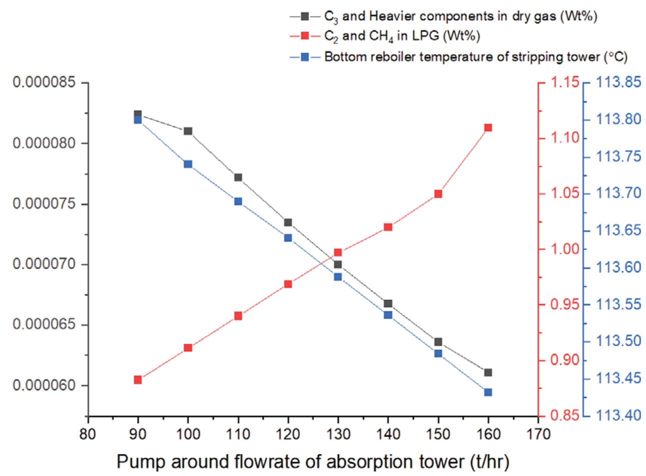

Fig. 7. Pump around flowrate of absorption tower.

because the rise in the temperature at the bottom of the stripping tower means that the desorption gas at the top of the tower increases.

Adding means that more C_3 and C_4 will enter the flash tank, and a cycle is formed between the flash tank and the absorption tower. The increase of C_3 and C_4 in the flash tank will inevitably cause the gas to enter the absorption tower. The content of C_3 and C_4 in the feed increases, and for a certain condition of absorbent under certain operating conditions, the increase in the content of C_3 and C_4 in the gas phase feed means that the range of C_3 and C_4 in the output will also increase. Therefore, the temperature of the bottom of the stripping tower should not be too high; otherwise, the dry gas amount will be low. It can be seen from Fig. 6 that as the temperature of the stripping tower decreases, the C_2 content in the liquefied gas will increase sharply. From the stripping tower temperature simulation results, it can be seen that the bottom temperature of the tower should be controlled between 110 and 115 °C. At this time, dry and liquefied gas quality can be acceptable.

As mentioned, the temperature of the absorption tower has a particularly important impact on the quality of dry gas. The too high temperature will make the dry gas not dry, so we should try to reduce the temperature of the absorption tower. Generally, the temperature in the absorption tower can be reduced by reducing the temperature of the supplementary absorption oil, and the temperature of the crude gasoline feed can be adjusted. And at the same time, a middle section of the absorption tower can be built to extract heat to withstand quality. For non-refining plants, some set up a mid-stage reflux heat extraction (pump around) in the absorption tower, some set up two mid-stage reflux heat extraction, and establish two mid-stage reflux heat extraction, which can optimize the tower's internal heat to a certain extent of gas-liquid load. Still, it needs to add a heat exchanger, and the manufacturing cost will increase. This refinery uses two mid-stage refluxes. The heat extraction load of the middle-stage reflux is mainly adjusted by changing the return temperature of the middle-stage reflux. But this will inevitably put more stringent requirements on the temperature of the condensed stream. Therefore, the reflux flow is generally used to adjust the tower temperature, making the operation easier.

1-3. Effect of Pump Around Flow Rate of the Absorption Tower on the Results

Regarding the specific impact of the pump around flowrate (middle reflux) of the absorption tower on the quality of dry gas and liquefied gas, for a more convenient and intuitive understanding, the desorption gas flow rate controlled and the liquefied gas at the top of the tower is stabilized. The effect of the pump around flowrate in the absorption tower on the quality of dry gas, liquefied gas, and LPG is shown in Fig. 7. When the pump around in the absorption tower's upper section increases, the C_3 and heavier components in dry gas approximately decrease in a linear relation, and C_2 and heavier contents of LPG increase.

So, in the middle, when selecting the flow rate, the quality of dry gas and liquefied gas should be weighed. In the case of this plant, C_2 and CH_4 in the liquefied gas content are lower than the standard, so to reduce the content of C_3 and its heavier components in the dry gas, the flowrate in the pump around section of the absorption tower can between 120 to 140 ton/hr. Sensitivity analysis is used to explore the influence of the pump around flow on the absorption tower, as discussed in Fig. 7. The temperature of the bottom of the stripping tower gradually decreases with the increase of the pump around flow. The decrease in temperature means a decrease in the load of the stripping tower bottom reboiler, so increasing the flow rate in the absorption tower's middle section can reduce the stripping tower's energy consumption. This is mainly because the increase in the flow rate in the pump around the absorption tower will improve the absorption effect of the tower so that more light components enter the rich absorption oil. After the rich absorption oil enters the flash tank, its light component increases, and the light component in the liquid phase after the final flash vaporization will also increase. When the liquid phase enters the stripping tower, it is extracted from the top under the same conditions as the previous operation.

As for the reabsorption tower, its principle is the same as that of the absorption tower. Its role is to make the dry gas drier. Of course, according to the principle of mass balance, it is not difficult to find that after the dry gas dries, the C_3 and above components will be reduced, and will enter the liquefied gas, so that the output of liquefied gas will increase, which will effectively improve economic benefits. Since it is also an absorption tower, the main factors affecting the process are temperature, pressure and flow. Low temperature, high pressure and high flow are favorable to the absorption phenomenon, but selecting these elements requires a balance of operating costs. Too low temperature will put higher requirements on the coolant, and too high pressure and too much flow will put better requirements on the equipment and increase production costs.

Therefore, it is generally possible to establish a zero-standard function according to the actual situation and find the most suitable production operation conditions through optimization to maximize production efficiency. However, due to the complexity of the equipment, we mainly analyzed the absorption and stabilization system from the perspective of product quality and energy but do not specifically analyze the equipment.

1-4. Effect of the Light Diesel Temperature and Flow Rate in the Reabsorption Tower on the Dry Gas

As for how the temperature of the light diesel in the reabsorption tower affects the quality of dry gas, we will discuss it through

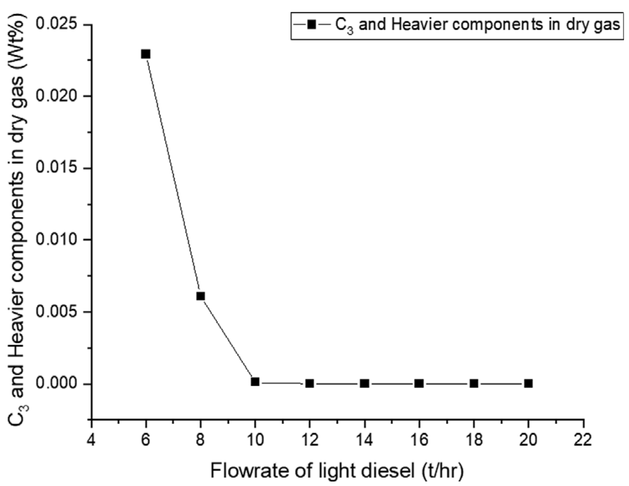

Fig. 8. The effect of light diesel flow rate on C_3 & heavier component content in dry gas.

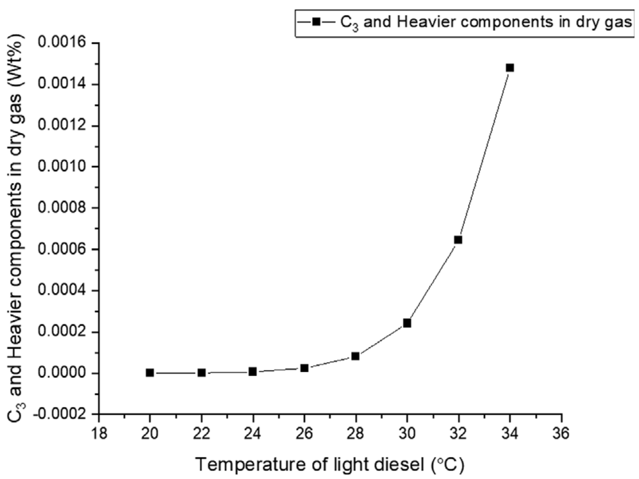

Fig. 9. The effect of light diesel temperature on C_3 & heavier component content in dry gas.

sensitivity analysis. After sensitivity analysis, Fig. 8 and Fig. 9 can be obtained. It can be seen from Fig. 8 and Fig. 9 that as the temperature of the light diesel increases, the content of C_3 and heavier components in the top dry gas also increases gradually. Therefore, under the conditions of other processes, the temperature of the reabsorbent can be minimized or lower. On the other hand, the increased flow rate of light diesel can decrease the C_3 and heavier components in dry gas.

2. Optimization of the Feed Position of the Stabilization Tower

As far as the stabilization tower is concerned, it is a rectification tower with a condenser and a reboiler. Generally, for the design of a rectification tower, the feed position, reflux ratio, number of theoretical plates, the operating pressure in the tower, and the feed temperature are several factors that are often considered. What can be adjusted is the feed position and reflux ratio, feed temperature and the operating pressure in the tower, etc.; the number of theoretical plates cannot be changed. The rectification tower generally has several spare inlets, for the stabilization tower in the absorption stabilization system generally has three feeds, top, middle and bottom respectively, which are selected according to the temperature and the composition of the raw materials.

Choosing the bottom feed can reduce the content of C_5 in the dry gas, but it can make the gasoline vapor pressure of the device too high. Choosing the top feed can cause too much C_5 content in the liquefied gas. Generally, the middle feed is selected. In the rectification process, there are many methods for selecting the best feed position, such as maximizing the separation factor method, minimizing the heat load of the heater and condenser, and the material composition approach method. The separation factor is generally for the light and heavy key components. The obvious components method is more suitable. The material composition approach method is more practical for separating mixtures with fewer components. It is more practical to use when the discharge composition is specified to minimize the heat load of the heater and condenser. The size of the reflux ratio is directly related to the heat load of the condenser and reboiler. As the reflux ratio increases, the heat load of the condenser and reboiler will increase. Of course, an increase in the reflux ratio will improve the separated product's purity. But this is at the expense of energy consumption. The change in reflux ratio will also change the vapor-liquid load in the tower. An excessive flow ratio will increase the vapor-liquid load, and in severe cases, flooding of the tower will occur.

Therefore, when the purity of the top product decreases and the reflux ratio needs to be adjusted, the reflux ratio should be controlled within a certain range. While improving the product quality, various factors should be weighed. Maximizing economic benefit is the most important thing for the company, not the product quality. The influence of feed temperature on rectification mainly affects the heat load of heaters and condensers. Superheated steam feed will reduce the load of the heater but will increase the load of the condenser. Subcooled feed will cause the load of the heater. The increase in feed temperature has no effect on the condenser. Whether the feed temperature should be increased or decreased should be determined based on the actual conditions of the plant. The operating pressure in the tower will also affect the heat load. Under the same design regulations, an increase in pressure will increase the

Table 5. Comparison of results at different feeding positions of stabilization tower

Feed tray	C_3 and C_4 in LPG mole%	C_5 in LPG mole%	C_2 and CH_4 in stable gasoline mole%
10	97.69	0.41	0.26
12	98.02	0.21	0.06
14	98.10	0.17	0.02
16	98.11	0.17	0.01
18	98.10	0.18	0.01
20	98.09	0.19	0.02
22	98.08	0.20	0.03
24	98.06	0.22	0.04
26	98.04	0.25	0.06
28	98.00	0.29	0.08
30	97.95	0.34	0.11
32	97.88	0.41	0.15
34	97.78	0.50	0.21
36	97.65	0.62	0.29

Table 6. The effect of stripping tower feed temperatures on process

Feed temperature °C	Reboiler heat load kW	Pre-feed heater load kW	Dry gas t/h	C_3 & C_{3+} in dry gas (10^{-5}) wt%	C_2 and CH_4 in LPG wt%
30	10,199.7	−775.17	2.64	4.60	0.99
35	9,829.62	−163.42	2.64	4.59	0.99
40	9,463.54	463.67	2.64	4.60	0.99
45	9,102.02	1,108.34	2.64	4.60	0.99
50	8,745.27	1,773.31	2.64	4.60	0.99
55	8,397.6	2,461.98	2.64	4.60	0.99
60	8,057.15	3,178.62	2.64	4.60	0.99
65	7,726.1	3,928.57	2.64	4.60	0.99
70	7,359.82	4,837.88	2.64	4.60	0.99

heat load of the reboiler. It also puts forward better requirements for the material of the equipment.

Due to the existence of virtual components in petroleum simulation calculations, it is difficult to select key components for analysis. This section analyzes the feed plate's position under the conditions of a certain amount of production at the top of the tower and a certain reflux ratio using the working condition study in Table 5. It can be seen from Table 5 that under the conditions of a certain amount of production at the top of the tower and a certain reflux ratio, when the feed position is above 16 plates, as the feed position moves downward, the content of C_3 and C_4 in the liquefied petroleum gas gradually decreases, the content of C_5 in liquefied gas increases.

The content of C_3 and C_4 in the stable gasoline gradually increase before and after the 16th tray. When the feed position is below 16 plates, as the position of the feed plate moves downward, the content of C_3 and C_4 in the liquefied gas gradually increases, and the content of C_3 and C_4 in the liquefied gas gradually decreases. The content of C_5 gradually increased, and the content of C_3 and C_4 in the stable gasoline gradually increased. When the 36th plate was fed, the content of C_5 in the liquefied gas was the highest. The content of C_3 and C_4 in stable gasoline is the smallest, so the best feeder position should be the 16th tray.

3. Optimization of Stripping Tower Conditions

3-1. Optimization of the Stripping Tower Feed

For the stripping tower, the feed temperature has a certain impact on the bottom load and product quality. Through Table 6, we can easily analyze the characteristics of this change. It is obtained under the condition that the C_2 content in the bottom of the stripping tower is not changed.

It can be seen from Table 6 that as the temperature of the feed increases, the heat load of the reboiler at the bottom of the stripping tower gradually decreases, but the total heat load increases. When the C_2 regulation in the bottom of the tower remains unchanged, the dry gas output will increase, which means that when the plant is in a certain time, the control index of the stripping tower is that the C_2 content is constant.

At a certain value, the operation is just at the edge of the operation, where increasing the feed temperature will cause more C_3 and above components to enter the dry gas, turning into low-value-added fuel gas, reducing the refining effect. The temperature used in the feed analysis in this paper is 35 °C.

Table 7. Comparison of stripping tower with or without intermediate reboiler

Equipment	Cold feed with intermediate reboiler heat load kW	Cold feed without intermediate reboiler heat load kW
Desorption reboiler	9,822.2	11,319.7
Stabilization tower reboiler	13,287.1	13,288.4
Stabilization tower condenser	−9,969.14	−9,966.2
Middle reboiler	1,484.86	0

Table 8. The influence of the circulation rate in the middle reboilers of the stripping tower on the process

Stripping tower middle reboiler flowrate t/h	Stripping tower middle reboiler heat load kW	Stripping tower bottom reboiler heat load kW	Dry gas t/h	C_3 and heavier in dry gas (10^{-5}) wt%
20	759.95	10,545.9	2.64	4.59
25	922.13	10,384	2.65	4.67
30	1,075	10,231.4	2.66	4.74
35	1,219.31	10,087.3	2.67	4.88
40	1,355.73	9,951.1	2.69	4.95
45	1,484.89	9,822.19	2.70	5.07
50	1,607.23	9,700.03	2.73	5.19
55	1,723.39	9,584.1	2.74	5.24
60	1,833.75	9,473.96	2.75	5.28

3-2. Optimization of Stripping Tower with and without Intermediate Reboiler

In the desorption operation, to alleviate the heat load of the boil at the bottom of the stripping tower, a reboiler is often installed in the middle of the stripping tower. For the feed, under the condition that the content of C_3 components is in control, dry gas remains unchanged. The results of setting the intermediate reboiler and without the intermediate reboiler are shown in Table 7.

It can be seen from Table 7 that the stripping tower with an intermediate reboiler can effectively reduce the heat load of the bottom reboiler, but it requires a new heat exchanger, which will increase the cost of renovation. Therefore, when the heat source of the reboiler in the stripping tower of the refinery is sufficient and the heat source is no longer used for other purposes, there should be no intermediate reboiler.

When the C_2 content in the top of the tower is specified to be unchanged, the heat extraction load of the intermediate reboiler has a certain impact on the product quality and the heat load of the reboiler. The specific impact is shown in Table 8.

It can be seen from Table 8 that when the heat extraction load of the intermediate reboiler increases if the C_2 content at the bottom of the stripping tower remains unchanged, the dry gas output will increase. Because the C_2 content entering the liquefied gas does not change, the dry gas output will increase. The amount of C_2 component in the gas remains unchanged, so the increased amount will be C_3 and above components, so the heat extraction of the intermediate reboiler should not be too large; otherwise, the dry gas will not dry, and the economic benefits of the entire product will be affected.

With an intermediate reboiler, it is equivalent to increasing the stripping tower bottom reboiler temperature. For the desorption operation, the increase in temperature will facilitate the desorption. When the C_2 content in the bottom of the stripping tower is very low, by controlling the temperature of the bottom of the stripping tower, the quality of the dry gas can be effectively adjusted. However, when the C_3 content in the bottom of the stripping tower is low, and the adjustable variable in operation is only the temperature of the bottom, reduce the temperature. The content of C_3 and above components in dry gas can indeed be reduced to make the dry gas quality meet standards, but the lowering of the bottom temperature of the tower will cause the liquefied gas to contain more C_2 components resulting in the undesired quality of the liquefied gas.

Therefore, when the content of the C_2 component at the bottom of the stripping tower is low, the intermediate reboiler can be selected, but when the content of the C_2 component is high, it should be carefully considered, because in that case it will be necessary to adjust the supplementary absorption oil or the lean absorption oil. The conditions for entering the tower will increase production costs to a certain extent. As mentioned, the use of an intermediate reboiler is beneficial to the desorption, mainly due to the increase in the temperature in the tower.

The position of the intermediate reboiler also has an impact on the energy consumption of the whole process. Table 9 analyzes the position of the intermediate reboiler (this analysis stipulates that the intermediate reboiler heat load remains unchanged and the C_2 content in the bottom of the tower) under stable conditions.

It can be seen from Table 9 that as the position of the intermediate reboiler moves down, the heat load of the reboiler at the bottom of the stripping tower gradually increases up to 14 trays and

Table 9. The effect of middle reboiler location on results

Liquid out tray number	Liquid in tray number	Stripping tower bottom reboiler heat load kW	Dry gas t/h	C_3 & C_{3+} in dry gas (10^{-5}) wt%
8	7	9,823.06	2.63	6.64
10	9	9,831.99	2.64	7.12
12	11	9,833.94	2.65	7.23
14	13	9,834.18	2.65	7.24
16	15	9,833.99	2.65	7.24
18	17	9,833.72	2.65	7.24
20	19	9,833.42	2.65	7.24
22	21	9,833.10	2.65	7.23
24	23	9,832.75	2.65	7.23

then decreases, the output of dry gas gradually increases, and the content of C_3 and above components shows fluctuation. Thus, when the heating load of the intermediate reboiler is constant, the lower the intermediate reboiler, the more economical the product can be. However, as the intermediate reboiler moves down, the same heat and the required temperature are obtained. The position is gradually increased, and the addition of an intermediate reboiler is mainly to use the low-temperature heat source to improve the economy of the entire energy. When the intermediate reboiler moves down, the economic advantage in terms of energy will gradually decrease, so comprehensive consideration is required when selecting the location of the intermediate reboiler.

Additionally, after setting all the parameters at optimized conditions (obtained from the above optimization analysis), the new process successfully saved 1.32 MW of thermal energy consumption (2.59%) compared with the existing ASP. The operating cost was reduced from 10.921 million USD annually to 9.830 million USD per year. Furthermore, the cost-saving effect of this optimization was about 9.99% (1.091 million USD per year).

CONCLUSION

A new ASP for petroleum refinery is presented in this article. A process optimization study was done to improve the product quality and energy efficiency of the absorption-stabilization process. We performed an optimization analysis on the ASP. The following are the main findings:

The analysis of the absorption tower of the ASP shows that, under the condition that the product quality control remains unchanged, the heat load at the bottom of the stripping tower will also increase. These quantitative analysis results have important guiding significance for adjusting the dry gas and liquefied gas problem. So, the suitable supplementary absorption oil flow value is adjusted at 52.2 t/h. The analysis of the reabsorption tower of the ASP shows that the temperature of the light diesel oil is reduced, and the flow rate lean absorption oil is increased, then the flow rate can effectively reduce the contents of C_3 heavier in the dry gas. Under the same flow rate of light diesel oil, with the increase in temperature of light diesel oil, the amount of C_3 and heavier increased in dry gas. So, for suitable conditions, temperature and flow rate analysis of light diesel against C_3 and heavier contents in dry gas is very important and selected as 30 °C and 10 t/h, respectively.

The optimization analysis of the stripping tower of the ASP shows that under different feeding temperatures, thermal efficiency is greater at 30 °C than that of 35 °C and greater temperatures. The comparison shows that the use of an intermediate reboiler can significantly improve the thermal efficiency of the stripper, but the sensitivity of the composition without reboiler is also higher. When the C_2 content in the feed changes greatly, it is easier to make the product unqualified. The optimization analysis of the temperature at the bottom of the stripping tower shows that the temperature controlled at 110 to 115 °C is the more suitable; the too high temperature will significantly increase the vapor and liquid load in the tower. The location sensitivity of the intermediate reboiler has shown that as the position goes to the bottom side, the C_3 and heavier contents in dry gas increase, but between the 7^{th} and 8^{th} plate, the heat load of the stripping tower is minimum.

Further, more optimization analysis of stabilization tower shows that feed at 16^{th} plate can reduce C_2 contents of LPG. Additionally, compared with the existing ASP, the new process successfully saved 1.32 MW of energy consumption (2.59%). The operating cost has been reduced from 10.921 million USD annually to 9.830 million USD per year. Furthermore, the cost-saving effect of this optimization is about 9.99% (1.091 million USD per year). Hence, this research opens new ways for the future optimization of the ASP further according to energy efficiency and product quality.

REFERENCES

1. A. Marafi, H. Albazzaz and M. S. Rana, *Catal. Today*, **329**, 125 (2019).
2. R. Sadeghbeigi, Butterworth-Heinemann (2020).
3. Dr. P. McDonald, Survey, *Oil Energy Trends*, **10** (2017).
4. D. Xiang, W. Lidong and D. Yingsheng, *J. Chem. Eng.*, **26**, 46 (1998).
5. X. Xu, *Oil Refining Des.*, **23**, 14 (1993).
6. Z. Luhong, W. Lu and S. Jinsheng, *Pet. Ref. Chem. Ind.*, **31**, 44 (2000).
7. D. Xiang, W. Shaomin and L. Changgeng, *Chem. Eng. Process*, **30**, 16 (2002).
8. G. Soave and J. A. Feliu, *Appl. Therm. Eng.*, **22**, 889 (2002).

9. G. S. Soave, S. Gamba, L. A. Pellegrini and S. Bonomi, *Ind. Eng. Chem. Res.*, **45**, 5761 (2006).
10. S. Bandyopadhyay, M. Mishra and U. V. Shenoy, *AIChE J.*, **50**, 1837 (2004).
11. S. H. Lee and M. J. Binkley, *Hydrocarb. Process. (International ed.)*, **90**, 101 (2011).
12. P. Seferlis and A. N. Hrymak, *Comput. Chem. Eng.*, **20**, 1177 (1996).
13. G. Li, X. Yong, L. Yushu and H. Ben, *IEEE*, 825 (2011).
14. E. Lu, Q. Pan and H. Zhang, ESCAPE-15 (2005).
15. Aspen plus user guide[M], *Aspen Tech. lnc.* (2009).
16. I. D. G. Chaves, J. R. G. Lopez, J. L. G. Zapata, A. L. Robayo and G. R. Nino, Case Studies. In: *Process. Analy. and Sim. in Chem. Eng. Springer, Cham.* (2016).
17. J. P. Gutierrez, L. A. Benítez, J. Martínez, L. A. Ruiz and E. Erdmann, *Int. J. Eng. Res.* (2014).
18. S. Lanyi, *Chem. Ind. Press* (2012).
19. S. K. Wasylkiewicz, L. C. Kobylka and F. J. L. Castillo, *J. Chem. Eng.*, **92**, 201 (2003).
20. X. Zhang, Y. Zou and S. Li, *Info. Sci.*, **530**, 95 (2020).
21. X. Zhang, Y. Zou and S. Li, *Neurocomputing*, **367**, 64 (2019).