# Gas Condensate Stabilization Unit: Different Design Approaches

Navid Moghadam and Masoud Samadi

Abstract—Gas condensate or natural gasoline is a valuable liquid hydrocarbon mixture which could be utilized as fuel or converted to different petroleum products like gasoline, light oil, jet fuel etc. It should be processed so that its water, salt, and acid contents meets the required standards for storage and transportation. Its vapor pressure, determined as Reid Vapor Pressure (RVP), must lie within a range that light components don't evolve as a separate gas phase in transport pipelines or storage tanks. The optimum value is usually 10 psia in summer and 12 psia in winter. In the present study, we investigated condensate stabilization using two methods of multistage flash vaporization and distillation (fractionation) within two case studies. The results show that depending on the properties of feed to be stabilized, one or both of the two methods could be practical and/or economic.

*Index Terms*—Gas condensate, HYSYS, ASPEN PLUS, optimization, simulation, stabilization.

### I. INTRODUCTION

The process of increasing the quantity of intermediate  $(C_3)$ to  $C_5$ ) and heavy ( $C_{6+}$ ) components in gas condensate is called condensate stabilization. The main purpose of this process is to reduce the vapor pressure of the condensate liquids to prevent the production of vapor phase upon flashing the liquid to atmospheric storage tanks. On the other hand, the scope of this process is to separate very light hydrocarbon gases, methane and ethane in particular, from the heavier hydrocarbon components  $(C_{3+})$ . Stabilized liquid has a vapor pressure specification, as the product will be injected into a pipeline or transport pressure vessel, which has definite pressure limitations [1]. It is usually 10 psia in summer and 12 psia in winter case. It should be noted that acidic contents (i.e., H<sub>2</sub>S, CO<sub>2</sub>, mercaptanes etc), glycol, free water, and possible salts must be also removed to have a gas condensate suitable for end-use applications. There are two methods for stabilizing gas condensates: flash vaporization and fractionation (distillation). Fractionation is more efficient and provides better control over vapor pressure for tanker shipment [2]. Non-refluxed distillation column is simpler but less efficient than refluxed one. Since it doesn't require any external cooling source, it is particularly applicable to remote locations [2]. Benoy and Kale [3] investigated gas condensate stabilization through three stage flashing, non-refluxed, and refluxed distillation with and without hydrocarbon recovery from the flashed gas by compression. They found that heat duty in multistage flashing is much higher than distillation while liquid recovery was less. Nevertheless, liquid recovery was independent of the technique applied when intermediate hydrocarbons were recovered from the flashed gas.

Herein, we undertook two case studies in which two-stage flash vaporization and distillation were investigated. The main objective was to adjust the RVP of product at 10 psia (good for summer) while keeping the duties of heat exchangers as low as possible and obtaining as much as stabilized condensate as possible. The process simulations were performed by means of ASPEN HYSYS and ASPEN PLUS software.

## II. PROCESS DESCRIPTION

NGL-3100 gas plant is under construction near the oil production unit of Cheshmeh-khosh located at the southwest of Iran for optimal recovery of  $C_{2+}$  product from the associated gases of North Dezful oil fields. It's planned to produce 46,000 bbl.d<sup>-1</sup>  $C_{2+}$ , 1180 bbl.d<sup>-1</sup> Stabilized Sour Condensate, 6.12 MSCMD Sweet Gas, and 0.62 MSCMD Sour Gas.

Shahid Hashemi-nejad (Khangiran) Gas Refinery is located at the northeast of Iran. The feed to the refinery comes from sweet gas reservoirs of Khangiran and sour gas reservoirs of Mozdoran. The treatment of 41.5 MSCMD of sour gas, dehydration of 7.4 MSCMD of sweet gas, distillation of 360000 L.d<sup>-1</sup> of gas condensates, and recovery of 2400 ton.d<sup>-1</sup>sulfur are the most important operations undergone in this plant. The feed to the stabilization units of Khangiran and NGL 3100 plants are mixtures of filter separators, slug catcher, and regeneration gas separators of which the main specifications are listed in tables I and II. Peng Robinson equation of state was applied for thermodynamic calculations of both gas and liquid phases. Since there was no data available on the boiling point curves (including ASTM and TBP distillation curves), the assessment of C10+ and C6+ oil cuts specifications was carried out by using their respective specific gravities (Sp.Grs.) and molecular weights (MWs) [4].

The final temperature and pressure of the stabilized condensate are governed by the storage conditions which are 35.3°C (95.5°F) and 2 bar in NGL3100 and 37.8°C (100°F) and 0.96 bar in Khangiran case.

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TABLE I: FEED PROPERTIES OF KHANGIRAN GAS PROCESSING PLANT

Comp.	H <sub>2</sub> O	N <sub>2</sub>	CO <sub>2</sub>	$H_2S$	COS	C <sub>1</sub>
(mol% )	0	0.02	7.47	11.20	0	11.43
Comp.	C <sub>2</sub>	C <sub>3</sub>	i-C <sub>4</sub>	n-C <sub>4</sub>	i-C <sub>5</sub>	n-C <sub>5</sub>
(mol% )	0.97	0.59	0.91	1.27	1.12	1.12
Comp.	C <sub>6</sub>	C <sub>7</sub>	C <sub>8</sub>	C <sub>9</sub>	C <sub>10+</sub>	
(mol% )	1.83	10.41	6.90	6.90	37.86	
Temp. (°F)	Press. (psia)	Molar Flow (lbmole.hr <sup>-1</sup> )	M.W.	RVP (psia)	Sp. Gr.	GCR <sup>1</sup>
74	139	293.6	119.4	155.7	0.715	179

TABLE II: FEED PROPERTIES OF NGL 3100 GAS PROCESSING PLANT STABILIZATION UNIT

Comp.	H <sub>2</sub> O	N <sub>2</sub>	CO <sub>2</sub>	$H_2S$	COS	C <sub>1</sub>
(mol%)	0.00	0.05	2.07	5.31	0	9.07
Comp.	C <sub>2</sub>	C <sub>3</sub>	i-C <sub>4</sub>	n-C <sub>4</sub>	i-C <sub>5</sub>	n-C <sub>5</sub>
(mol%)	13.76	20.5	5.67	15.84	6.49	6.23
Comp.	C <sub>6+</sub>	$CS_2$	Merca ptanes			
(mol%)	14.92	6e-4	0.0452			_
Temp. (°F)	Press. (psia)	Molar Flow (lbmole.hr <sup>-1</sup> )	M.W.	Sp. Gr.	RVP (psia)	GCR
77	413	595.8	53.1	0.569	235.9	251

## III. MULTISTAGE FLASH VAPORIZATION

Our main objectives were to recover as much as liquid condensate as possible while keeping its RVP at 10 psia for the summer case. However, heat duties of steam heaters and shell and tube heat exchangers (so as their surface areas) should be minimized to reduce the operating and capital costs to the least possible amount. The liquid stream leaving the second separator (L3 stream in Fig. 1) is recycled to preheat the liquid stream leaving the first one (L1 stream in Fig. 1) aiming at lowering the heat duty of the following steam heater (E-102 in Fig. 1). A pressure drop of 5 psi was designated to both shell and tube sides. The heat duty of E-102 is adjusted aiming at bringing the RVP of the final liquid product to 10 psia. The pressure of the first stage is set to the flow line pressure i.e. 139 psia (in Khangiran case). The storage tank pressure is 13.92 psia. The rule of thumb proposed by Campbell (2004) suggests that an equal pressure ratio between the stages would be considered. Here, it gives a pressure of 44 psia for the second separator (V-102 in Fig. 1). However, the trend of the total heat duty (the sum of heat duties of E-100, E-101, and E-102 heat exchangers) and the recovered stabilized condensate shown in Figs. 2 and 3

 $^1\,\text{Gas}$  To Condensate ratio, STD  $m^3.h^{\text{-}1}$  gas flow per STD  $m^3.h^{\text{-}1}$  liquid flow

suggest that a lower pressure may be more economic. Hence, the second separator pressure was assumed 25 psia. In practice, a higher pressure should be selected to compensate for the pressure losses in the transport pipeline and to allow recovering the flashed gas in the second separator and sending it to the fuel gas system on demand.



Fig. 1. Multistage flash vaporization scheme with the recycle of hot stabilized condensate.



Fig. 2. Total heat duty versus second separator pressure in Khangiran case



Fig. 3. Stabilized condensate flow rate versus second separator pressure in Khangiran case

The RVP and  $H_2S$  concentration of stabilized condensate, the stabilized condensate to feed ratio, and the total heat duty of heat exchangers are listed in table III. The dimensions of separators could be considered as 4.5 ft in diameter and 15.75 ft in length using horizontal vessels with length to diameter ratio of 3 and retention time of 12 minutes.

In NGL3100 case, however, use of two flash separators doesn't yield the required RVP of the condensate and cannot be considered as an option. We added one separator and investigated the flash vaporization scheme. Despite the high heat duty applied, the stabilized condensate recovery was very low (3.5%) and the  $H_2S$  content was much higher than that of fractionation. We can conclude that flash vaporization is not practical in NGL3100 case.

TABLE III: STABILIZED CONDENSATE SPECIFICATIONS IN KHANGIRAN AND NGL3100 CASES

Khangiran	RVP (psia)	H <sub>2</sub> S (ppm)	CFR <sup>2</sup> (%)	THD <sup>3</sup> (btu.h <sup>-1</sup> )
Two Flash Tanks	10	4467	87	5.730 x10 <sup>5</sup>
Fractionation	10	6635	70	8.484 x10 <sup>6</sup>
NGL3100				
Three Flash Tanks	10	206	3.5	5.438x10 <sup>6</sup>
Fractionation	10	<1	30	5.639x10 <sup>6</sup>

### IV. FRACTIONATION

A distillation column with a reboiler and without any condenser (non-refluxed column) was used (Figs. 4 and 5). The pressure of top tray and reboiler are primarily considered as 130 and 135 psia in Khangiran case and 140 and 145 psia in NGL3100 case respectively. Boil-up ratio was adjusted aiming at having the bottom product RVP at 10 psia. The feed is preheated to an extent that the preheater, reboiler, and cooler (E-101 in Fig. 4) heat duties are lowered as much as possible in Khangiran case. It is notable that the stabilized condensate flow rate and the total heat duty variations have the same trend indicating that there's a trade-off between the heating cost and the stabilized condensate profit.

The criterion of preheating extent is different in NGL3100 case. It is determined in such a way that the temperature of cooled stabilized condensate (DL2 in Fig. 5) reaches to 95 F (which is almost the stock tank temperature). Since the feed pressure of NGL3100 is high (413.4 psia) it passes through a Joule-Thomson valve (VLV-101 in Fig. 5) and reaches to 130°F and 159.6 psia.



Fig. 4. Fractionation (Distillation) scheme with the recycle of hot stabilized condensate (bottom product) in Khangiran case

For a definite separation, the reflux and boil-up ratios (consequently the total heat duty of condenser and reboiler and their respective surface area) decrease as the number of stages increases. However, increase in the number of stages results in higher capital costs due to higher heights of column and trays price. Therefore, there's a trade-off between the number of stages and the total heat duty. On the other hand, the flow rate of stabilized condensate rises with respect to the number of stages. The changes in reboiler duty and flow rate

<sup>2</sup> Condensate To Feed Ratio (lbmol cond./lbmol feed)

of stabilized condensate in NGL3100 case are depicted in figures 6 and 7. It can be seen that the addition of stages further than 26 has small effect on the reboiler duty and the condensate flow rate. Hence, this number may be primarily considered as the optimized one. If the overall tray efficiency is considered as 0.85, the actual number of stages will be 31 which equals to 30 trays. The valve tray type with one pass, flooding fractional approach of 0.8, and tray spacing of 1.5 ft were chosen. The calculated values for diameter and column pressure drop were 3.07 ft and 1.57 psi respectively. A similar procedure in Khangiran case indicates 10 theoretical and 13 actual stages. The column diameter and pressure drop were calculated as 2.17 ft and 0.76 psi respectively (at 1 ft tray spacing). Considering the diameter of column (2.17 ft), choosing a packed column would be better. If we use 1 in. (25 mm) pall ring packing with approximate HETP<sup>4</sup> of 1.48 ft (0.45 m), the column diameter and pressure are calculated as 2.02 ft and 0.04 psi respectively. The properties of product and the total heat duty of reboiler and heat exchangers are listed in table III.



Fig. 5. Fractionation (Distillation) scheme with the recycle of hot stabilized condensate (bottom product) in NGL3100 case



Fig. 6. Reboiler heat duty versus number of stages of distillation column in NGL3100 case



Fig. 7. Stabilized condensate flow rate versus number of stages of distillation column in NGL3100 case

<sup>4</sup> Height Equivalent to Theoretical Plate

<sup>&</sup>lt;sup>3</sup> Total Heat Duty

More rigorous designs should take economic evaluations into account. The capital cost of column is related to the vessel, trays or packings, internals, and the reboiler (a kettle type shell and tube heat exchanger). This cost may be annualized via dividing by payback period in years. The operating costs account for the energy required for generation of high pressure steam passing through the reboiler tubes to partially vaporize the bottom liquid. Total annual cost (TAC) may be then calculated from (1). The number of trays which yields the minimum TAC may be considered as the optimized number of trays.

$$TAC = \frac{\text{capital cost}}{\text{payback period}} + \text{energy cost}$$
(1)

The f.o.b. purchase cost of the distillation column can be calculated by (2) to (11) [5].

$$C_p = C_s + C_T \tag{2}$$

$$C_{S} = F_{M} \cdot C_{V} + C_{PL} \tag{3}$$

 $C_{V} = \exp(7.0374 + 0.18255[\ln(W)] + 0.02297[\ln(W)^{2}])$ (4)

$$W = \pi (D_i + t_s) (L + 0.8D_i) t_s . \rho \tag{5}$$

$$C_{PL} = 285.1(D_i)^{0.73960} (D)^{0.70684}$$
(6)

$$C_T = N_{TN} . F_{NT} . F_{TT} . F_{TM} . C_{BT}$$
(7)

$$F_{NT} = \frac{2.25}{1.0414^{NT}} \quad N_T < 20 \tag{8}$$

$$F_{NT} = 1 \quad N_T < 20 \tag{9}$$

$$F_{TM} = 1.401 + 0.0724 D_i \quad (316 \ S.S.) \tag{10}$$

$$C_{BT} = 369 \exp(0.1739 \, D_i) \tag{11}$$

where,  $C_P$  is the distillation column cost (\$),  $C_S$  is the shell cost (\$),  $C_T$  is the tray cost (\$),  $F_M$  is the material factor which is 2.1 for 316 stainless steel (S.S.),  $C_V$  is the cost of empty vessel including nozzles, manholes, and supports (\$),  $C_{PL}$  is the platform and ladder cost (\$),W is the weight of shell and two heads (lb),  $D_i$  is the inside diameter (in),  $t_s$  is the shell thickness (in), L is the column height (in),  $\rho$  is the construction material density (lb.in<sup>-3</sup>),  $N_T$  is the number of trays,  $F_{TT}$  accounts for the tray type which is 1.18 for valve trays,  $F_{TM}$  is the correction factor for the material of construction, and  $C_{BT}$  is the base cost of trays.

The f.o.b. purchase cost of the reboiler which is a kettle type shell and tube heat exchanger is calculated by (12) to (15) [5].

$$C_P = F_P \cdot F_M \cdot F_L \cdot C_R \tag{12}$$

$$F_P = 0.8510 + 0.1292 \left(\frac{P}{600}\right) + 0.0198 \left(\frac{P}{600}\right)^2 \qquad (13)$$

$$F_M = a + \left(\frac{A}{100}\right)^b \tag{14}$$

$$C_B = \exp(11.967 - 0.8709[\ln(A)] + 0.09005[\ln(A)]^2)$$
(15)

where,  $F_P$  is the pressure factor, P is the shell side pressure (psig),  $F_M$  is the material factor, a and b are 2.70 and 0.07 for S.S./S.S. material, A is the surface area (ft<sup>2</sup>), and  $C_B$  is the base cost (\$). Energy cost may be considered as 6 \$.GJ<sup>-1</sup>. The aforementioned equipment cost formulas are indexed to

the year 2000 and should be updated to the year 2011. It can be done by multiplying the base cost by the cost index ratio as stated in (16). We used Chemical Engineering (CE) index which is 394 and 585.7 for the years 2000 and 2011 respectively.

Current cost = base cost 
$$\left(\frac{I_{2011}}{I_{2000}}\right)$$
 (16)

The diagram of TAC with respect to the number of trays is shown in Fig.8. It implies that the number of trays at which the minimum TAC occurs is 9 (or 10 stages).



Fig. 8. Total annual cost versus number of trays of distillation column in NGL3100 case

Since the stabilized condensate flow rate increases with respect to the increase in number of stages, it is better to choose the optimized number of trays by maximizing the annual profit stated in (17). Annual sale is calculated by multiplication of gas condensate price (in \$.bbl<sup>-1</sup>) by gas condensate flow rate (in bbl.y<sup>-1</sup>). The stabilized condensate price was taken as 100 \$.bbl<sup>-1</sup>. For simplicity, we calculated the annual profit obtained by the addition of each tray (profit per tray) instead, by assuming the number of two trays as a base. Since the duty and area of the preheater and the associated equipments do not change significantly with respect to the addition of column trays, the aforementioned trick would be rational. Fig. 9 shows the annual profit obtained by the addition of each tray. It's obvious that profit earned by using more than 9 up to about 30 trays is considerable. Hence, the number of 30 trays (or 31 stages) which was suggested earlier sounds reasonable.

Annual profit=Annual stabilized condensate sale-TAC (17)



Fig. 9. Annual profit per tray versus number of trays of distillation column in NGL3100 case

### V. PROCESS SELECTION

We saw that the multistage flash vaporization is a practical method to stabilize the condensate in Khangiran case, as it is already utilized in the aforementioned gas processing plant. However, in NGL3100 case, it doesn't sound promising considering the low condensate recovery and high total heat duty. Capital costs of the process equipments aside, multistage flash vaporization is more suitable than the fractionation in terms of total heat duty and the flow rate of stabilized condensate (or condensate to feed ratio) in Khangiran case. However, in NGL3100 case, it is clear from table III that fractionation is far better than flashing since it provides more condensate recovery, less heat duty, and H<sub>2</sub>S content within the standard range. It seems that RVP of the feed has a great effect on the performance of stabilization unit which, in these cases, were 155.7 psia and 235.9 psia. We should take this fact into account that distillation column is further capable of handling variations in the feed composition (which affects its RVP). H<sub>2</sub>S content is high in both methods in Khangiran case and must be reduced to standard values (e.g. less than 50 ppm).

## VI. CONCLUSION

In the present study, two methods of multistage flash vaporization and non-refluxed fractionation were investigated to stabilize gas condensate by reducing its RVP to 10 psia in summer. Gas condensates of two gas processing plants were studied. Results show that depending on the feed properties (especially its RVP) one or both of the aforementioned techniques is/are applicable and/or economic. In Khangiran case (RVP and GCR of the feed were 155.7 psia and 179 respectively), multistage flashing provides less total heat duty, more condensate recovery and less H<sub>2</sub>S content which makes it more interesting. In NGL3100 case (RVP and GCR of the feed were 235.9 psia and 251 respectively), fractionation yields less total heat duty and much higher condensate recovery. It is also capable of lowering H<sub>2</sub>S concentration to less than 1 ppm. Therefore, fractionation is highly preferred. Calculations made to determine the specifications of distillation columns and flash separators as follows: a non-refluxed trayed distillation column with 31 stages (30 trays), 3.07 ft in diameter, and 54 ft in height in NGL3100 case; and a similar type of column with 13 stages

(12 trays), 2.17 ft in diameter, and 14.4 ft in height or a packed column with the same number of stages but 2.02 ft in diameter and 21.3 ft in height in Khangiran case; and two cylindrical horizontal separators with dimensions of 4.5 ft in diameter and 15.75 ft in length in Khangiran case. Finally, detailed cost benefit analysis together with safety considerations and environmental concerns would determine the final selection of the right technique.

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