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Process Simulation and Assessment of a Back-up Condensate Stabilization Unit

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Abstract

A simulation was conducted using Aspen HYSYS[®] software for an industrial scale condensate stabilization unit and the results of the product composition from the simulation were compared with the plant data. The results were also compared to the results obtained using PRO/II software. The results show that the simulation is in good agreement with the plant data, especially for medium range hydrocarbons. For hydrocarbons lighter than C₅, the simulation results over predict the plant data while for hydrocarbons heavier than C₉ this trend is reversed. The influences of steam temperature and pressure, as well as feed conditions (flow rate, temperature and pressure) for the product specification (RVP and sulphur content) were also investigated. It was reported that the operating conditions gave rise to the production of off-specification condensate and it was also found that the unit could be utilized within 40 to 110% of its normal throughput without altering equipment sizing and by the operating parameters.

Keywords: Condensate Stabilization Unit; Aspen HYSYS[®]; PRO/II[®]; RVP; Sulphur Content

1. Introduction

Condensate stabilisation refers to stripping of light hydrocarbons (methane and ethane) and removal of acidic components from a liquid hydrocarbon to meet the marketing standards. Hydrocarbon condensates recovered from a natural gas, especially in remote offshore platforms, sometimes do not undergo further processing but are simply stabilized for blending with crude oil streams and then exported as crude oil. For the case of raw

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condensate, there are no strict specific requirements for the product other than the process specifications. In general, the process of increasing the amount of intermediates (C_3 to C_5) and heavy fractions (C_{6+}) in the condensate is called condensate stabilization (Mokhatab et al., 2006). The hydrocarbon condensate stabilization is also required to minimize the hydrocarbon losses from the storage tank (Benoy and Kale, 2010). This process is performed because a vapour phase must not be produced upon flashing in the atmospheric storage tank. Besides, the purpose of this process is to separate light hydrocarbon gases such as methane and ethane from the heavier hydrocarbon components such as propane and the others. Heavier components can be used for oil refinery cracking processes which allow the production of light products such as Liquefied Petroleum Gas (LPG) and gasoline (Gary, and Handwerk, 2001). Nevertheless, the stabilized liquid has some vapour pressure specifications as it is transferred into pipelines (Mokhatab et al., 2006) and therefore the raw condensate must be processed at certain pressure and temperature so as not to allow to release of light gas in the condensate export pipeline or tanker.

In general, condensate stabilization accomplishes several goals, the foremost of which are:

- a) To increase the recovery of methane-ethane and LPG products.
- b) To lower the vapour pressure of the condensate which makes it more suitable for blending and reducing the evaporation losses while the product is stored or shipped.
- c) To sweeten the raw liquid entering the downstream plant (if any) by removing the acid gases such as hydrogen sulphide and carbon dioxide contents in order to meet the required specifications.
- d) To maintain the purity and molecular weight of the lean absorption oil free of certain components such as pentanes and heavier hydrocarbons.

The vapour pressure of condensate is measured by the Reid Vapour Pressure (RVP) test, (ASTM D323-99a, 2012). . The impact of RVP is often referred to as the gasoline volatility. RVP can also be estimated without performing the actual test by using an algorithm (Esparragoza et al., 1992; Benoy and Kale, 2010; www.intertech.com). In this study, RVP has been set as a criterion for off-spec conditions of the product - that is, a maximum of 10 psia in summer and 12 psia in winter. In actual plant conditions, any condensate produced from this range is called off-spec product and is sent to an off-specification storage tank for temporary storage and further processing at a suitable time. The off-spec tank has the capacity to store 24 hours off-spec production.

Process simulation software packages are extensively used nowadays to estimate the product efficiency and enhance the performance of the system by optimizing operating parameters (Bao et al., 2002; Šoóš et al., 2003; Ye et al., 2009; Peters et al., 2011; Lastari et al., 2012; Tavan et al, 2013). There have been few simulating software packages such as Aspen Plus[®], Aspen HYSYS[®] and PRO/II[®] for use in the oil and gas industries. For example, the hydrogen production with steam methane reforming in a fluidized bed membrane has been simulated by Aspen Plus (Ye et al., 2009). This simulation demonstrates considerable responses against the change in pressure, temperature, steam-to-carbon ratio and permeates the side partial pressure of the reactor. Besides, the result was compared with a pilot scale experimental study and not at real industrial scale. Carbon dioxide capture by MEA absorbent was studied and simulated by Aspen Plus and Aspen HYSYS[®] (ErikØi, 2012). Aspen Hysys[®] was also used to simulate azeotropic separation of ethane and CO₂ using reactive absorption (Tavan and Hosseini, 2013).

PRO/II[®] is a commercial process simulator widely used in the oil, gas and petroleum industries (Liao et al., 2001, Leet et al., 2013, Kim et al., 2013), for instance, in the production of methanol from natural gas, CO₂ absorption has been simulated for a FPSO

(floating production, storage, off-loading) system (Kim et al., 2013). In another example, the CO₂ reforming of methane has been modelled with PRO/II[®] to consider the effect of Ni-based catalyst (Lee et al., 2013). The conversion of CH₄ versus the change in concentration of H₂O and CO₂ has been studied and the Ni/MgO was chosen as a desirable catalyst in order to produce synthesis gas; the result of modelling was validated by experimental data not plant data.

The objective of this study is: i) to find the right operational window and optimum conditions for a current operational BCSU in terms of producing on-specification product and ii) to compare the simulation software packages PRO/II[®] and Aspen HYSYS[®] for this specific unit operation.

1.1 Block flow diagram of Condensate Stabilization Unit

Figure 1 shows the block flow diagram of a gas plant consisting of a Condensate Stabilization Unit (CSU) and a back-up CSU (BCSU) located at Asaluyeh port in the southern part of Iran. BCSU in this plant is the subject of this study.

Firstly, reservoir fluids which consist of gas, water and condensate are produced and primarily processed at the offshore platforms. Then, some free water is removed from the mixture and the rest is transported to the on-shore plant. The transportation of the treated reservoir fluids is transported through a 32 inch pipeline about 120 km from the off-shore processing platform plant to the on-shore plant. In the presence of water, the gas mixture can form gas hydrates, which hampers the smooth flow of gas in the pipeline. Hence, monoethylene glycol (MEG) is injected via a 4 inch piggy backline to the exit stream from the offshore platform in order to prevent the formation of gas hydrates (see Figure 1).

Once the gas mixture arrives at the onshore plant, it will be separated into two streams; a gas stream and a liquid stream in the slug catcher. The gas stream is transferred to the gas plant and the liquid stream that consists of condensate, MEG and water is further separated to form a condensate stream and a mixture of MEG and water stream. The mixture of MEG and water is treated in the MEG regeneration unit where MEG is recycled to the off-shore via a 4 inch piggy back line. Then the condensate stream is fed to the CSU. A BCSU is designed to run the plant during CSU failure. After treating in CSU or BCSU, the stabilized condensate is transferred to storage tanks for exporting purposes to local plants or overseas.

1.2 Process description of BCSU

The BCSU process is similar to stage separation utilizing the equilibrium principles between vapour and condensate phases. Equilibrium vaporization occurs when the vapour and condensate phases are in equilibrium at the temperature and pressure of separation (Mokhatab et al., 2006).

Figure 2 shows a typical flash vaporization process for condensate stabilization with the same concept as BCSU in this study. The main feed which is a condensate produced from the inlet separator (slug catcher) passes through a heat exchanger and then enters the high-pressure (HP) flash tank where the pressure is maintained at 600 psia. A pressure drop of 300 psia helps flash of large amounts of light ends which are discharged as sour gas stream after recompression. The sour gas can be sent to further units or recycled into a reservoir for enhanced oil recovery purposes. After that, the bottom liquid from the HP tank enters the middle pressure (MP) flash tank where the additional methane and ethane are released. Then, the bottom product re-enters the low-pressure (LP) tank and they are fed to a condensate stripper for purification before transferring to the storage tank. This reduces excess flashing of condensate in the storage tank and the amount of inert gas, such as nitrogen, for blanketing

purposes (Mokhatab et al., 2006). Multi-stage flashing is based on the principle of progressively lowering the pressure of condensate during each stage (Esparagoza et al., 1992). This is enhanced for the flashing of lighter components from the condensate.

In BCSU, only a simple heating and cooling process is considered as the main objective is to reduce the capital costs and, more importantly, BCSU is not a continuous operation. Hence, the back-up unit prefers to use the flash vaporization method to run its operation. This method imposes some pressure only to stabilize the condensate before being sent to the storage tank.

2. Methodology

2.1 Feed Condition

In analysing the performance of the BCSU system, simulation was carried out using Aspen-HYSYS (ver. 2006). It is essential to have a model that is reliable in representing the BCSU system as some of the data are unavailable from the plant and are only available via calculations from the HYSYS model. To achieve this objective, the simulation results were compared to the actual operating values gained from real plant data available from the South Pars gas field (Asaluyeh, Iran).

Figure 3 shows the envelope curve of the feed to the BCSU. The feed consists of 0.57 liquid hydrocarbon phase fraction, 0.26 vapour phase fraction and 0.18 aqueous phase.

2.2 Simulation Method

The simulation is performed based on a reference BCSU in operation (Behbehani and Atashrouz, 2011). The Peng-Robinson (PR) equation of state (1976) was used for modelling. Figure 4 shows the process flow diagram of the real BCSU used in this work. The purpose of this process is separation of aqueous phase and gaseous hydrocarbon from the condensate and

then to stabilize it for export by adjusting RVP as an indication of the volatility of the condensate. This is because the quality of the product depends on the composition and also RVP before marketing.

In the process, firstly, the main feed from the on-shore plant enters a pre-flash drum to remove light hydrocarbons; however, most of the acid gases and lighter paraffins are also removed in this step. Next, the condensate temperature is increased in two sequential heat exchangers and a High Pressure (HP) heater up to 80°C and 143°C, respectively. Finally, this fluid crosses a shell and tube heat exchanger and degassing in the last flash drum is transferred to the condensate storage tanks which are equipped with an external floating roof.

The off-gas includes light hydrocarbons such as methane, ethane, propane and hydrogen sulphide. The aqueous phase containing MEG is sent for further processing to the MEG regeneration unit. Besides that, components with a sulphur element, e.g. mercaptans and also water are sent to off to the specification tank and finally are transferred to waste treatment.

2.3 Influence of Process Parameters

Process parameters which affect the final product specification include steam conditions such as temperature, pressure and feed conditions such as flow rate, temperature and pressure. To study the effect of these parameters on the product specification such as RVP and sulphur content, each parameter varies while the others are kept constant. All of these scenarios are listed in Table 2. This helps to find the right operational window for producing the on-specification product.

3. Results and Discussion

In order to validate the simulation, the compositions of the stabilised condensate from Aspen HYSYS are compared with the plant data. Figure 5 shows the comparison for 24 components of the product. It can be seen that the trends of compositions between the plant data and HYSYS are the same. However, compositions of *n*-C₄, *i*-C₅, *n*-C₅, *n*-C₆ and *n*-C₇ are under-predicted in the case of HYSYS. In other words, the mole fraction of light hydrocarbon components obtained through the simulation is lower than that of the plant data. This indicates that the unwanted hydrocarbons are already flashed before being sent to the storage tank.

Furthermore, the compositions of the hydrocarbons heavier than *n*-C₉ obtained using HYSYS are slightly higher than those of plant data. In general, the quality of the simulated product is the same as the plant data because their difference is about 5% and does not affect the overall product specification.

An attempt was made to run the simulation with PRO/II[®] software version 7.1 from Invensys Ltd. [www.invensys.com, 2012]. Figure 6 compares the compositions of the stabilised condensate obtained by PRO/II, HYSYS and plant data. Clearly, both PRO/II and HYSYS predictions follow the trend of the plant data. The trend was expected because the same thermodynamic package, i.e. PR equation of state was used for the both software packages. The difference between plant data and simulation results may be attributed to the fact that PR is generally very accurate in predicting the liquid densities especially non-polar ones as compared to polar materials, while this is not the case here as there was water (13%) and MEG (4.8%) as polar components in the feed. The major difference is that for the light hydrocarbons, PRO/II shows a better agreement with the plant data than HYSYS while for the heavy hydrocarbons the reverse is true. However, the results obtained from both software packages are very close (maximum difference is 8.5% for C₁₁₊). This low difference was

expected as again the same thermodynamic package, PR, was used in this work. The difference may come from the point that these software packages use their own methods of simulation which are well protected from public access due to the commercial impact.

A close look at the sulphur containing compounds, i.e., M-mercaptan, E-mercaptans, etc., proves that the simulation matched the plant data exactly. As the sulphur content in the product affects the condensate price, it can be concluded that as far as the price is concerned, the simulation can be used as a strong tool to predict the compositions of sulphur components.

3.1 Effect of Steam Temperature

Figure 7 shows the influence of steam temperature on the RVP and sulphur content of the final product in BCSU under the operating conditions given in Table 1. The higher temperature results in a lower RVP value. This indicates that as predicted the higher steam temperature removes more acid gases and light hydrocarbons. In the range of steam temperatures selected, RVP reduces from 8.4 psia to 6.4 psia. From this range, the best steam temperature is 143°C in terms of preventing more loss of propane and butane as well as stripping corrosive and sour components. At 143°C the RVP of the product would be 7.9 psia.

Figure 7 also shows that the concentration of sulphur components decreases as the steam temperature increases. This is because the components which contain sulphur elements are removed rapidly at higher temperatures and acidic components are flashed. The highest sulphur concentration is 2500 ppm which is at the lowest temperature of 138°C. In the plant under study, sulphur content has not been set as a means of quality control. This may be regulated in the near future. It means that stabilized condensate with any sulphur content is exported to the international market. For the time being, there is no condensate treating unit

under operation. However, there is a promising plan to design and install a mercaptan removal unit for the purpose of condensate treatment.

3.2 Effect of Steam Pressure

Figure 8 shows the effect of steam pressure on product specifications of the condensate. The RVP is decreased as the steam pressure increased. The lowest steam pressure is 10 bar and the highest pressure is 65 bar. Steam pressures lower than 10 bar and higher than 65 bar resulted in temperature cross in the heat exchanger which is not valid to proceed the simulation. Further, due to limited steam supply from the acting units, steam pressures higher than 65 bar were not chosen. Under the range of selected steam pressure, the RVP changes from 7.94 to 7.92 psia. The optimum condition is 35 bar to remove the unwanted hydrocarbons and also stripping sour components which cause a RVP of 7.93 psia. It means that the higher steam pressure increases the heat exchanger duty. As a result of high heat exchanger duty, there was more flashing of acidic gases.

Figure 8 also shows that the sulphur concentration decreased as the steam pressure increased. From the trend, it can be seen that the high steam pressure causes the removal of sulphur contents faster. Comparing Figures 7 and 8 makes it clear that the effect of steam temperature is more pronounced than the effect of steam pressure on the RVP and sulphur concentration of the final product. For pressure increase from 10 to 65 bars, sulphur concentration decreases around 3 ppm while for a temperature increase from 140 to 160°C the sulphur concentration decreases around 100 ppm.

3.3 Effect of Feed Flow Rate

Figure 9 shows the influence of change of feed flow rate in percentage compared to the normal flow rate (4645 kmol/hr) on both RVP and the sulphur content of the product. Shown in Figure 9, increase in feed flow rate increases the RVP of the product. This can be explained by the knowledge that for the highest feed flow rate which needs to be stabilized, a greater amount of heat is required. However, the heat transfer was kept constant. As a result, RVP is increased because there is insufficient heat to maintain the RVP of the product. At a molar flow rate of 1848 kmol/hr, which is equal to a turn down of 40%, there will be a temperature cross in the heat exchanger. Furthermore, at the feed flow rate of 5574 kmol/hr (120%), temperature cross also occurred in the heat exchanger. Therefore, the optimum condition for feed flow rate is in the range of 40% to 110%.

It can be seen from Figure 9 that sulphur concentration is increased as feed flow rate is increased. The reason is that the higher feed flow rate into the process results in higher sulphur content to the unit and if no adjustment of heat transfer is carried out, more sulphur is produced in the product. It can be concluded that to decrease the sulphur concentration in the final product, the feed flow rate should be kept low or the heat supply should be adjusted. The lowest sulphur concentration in the product is 1494 ppm at the 50% of feed flow rate and the highest value is 2502 ppm at 110%.

3.4 Effect of Feed Temperature

Figure 10 shows the influence of feed temperature on the RVP and sulphur content. RVP is decreased as the feed temperature is increased indicating the flash-off of light hydrocarbons. A sub-zero feed temperature is not practical due to the plant location and the environmental conditions. However, for the purpose of the study, a minimum of -5°C was investigated. To keep a product specification of 10 psia in the summer, the feed temperature

should be from -10°C to 20°C. The normal feed temperature in summer is 17.7°C which gives a RVP of 7.9 psia.

Figure 10 also shows that sulphur concentration is decreased as the feed temperature is increased. This is because the mercaptan components which are the source of the sulphur elements are removed at high temperature. Therefore, a higher feed temperature is favourable for the process. The minimum sulphur concentration in the product is 2375 ppm at a temperature of 45°C and the maximum sulphur concentration is slightly higher than 3000 ppm at -5°C.

3.5 Effect of Feed Pressure

Figure 11 shows the effect of feed pressure on the RVP and sulphur content. RVP increases as feed pressure increases. This is because at high feed pressure the feed tends to change to the liquid phase while in the three-phase separator the pressure should be as low as possible to flash-off the acidic gases. The lowest feed pressure is intentionally set to 1200 kPa because in actual conditions a pressure lower than this causes the automatic shut-down of the compressor as a result of low suction pressure to protect it from potential vibration damage. The RVP changes from 6.9 psia to 8.9 psia corresponding to a feed pressure of 1200 and 1300 kPa, respectively.

Figure 11 also shows that the sulphur concentration is increased as the feed pressure is increased. Higher pressure is not favoured for removing sulphur content. The lowest sulphur concentration is 2281 ppm at feed pressure of 1200 kPa.

4. Conclusions

Simulation of a Back-up condensate stabilization unit has been conducted to examine the conditions which give rise to produce off-specification product. RVP has been set as the

criteria for the off-specification conditions of the product - that is, a maximum of 10 psia in summer and 12 psia in winter.

To validate the simulation, the data have been compared with the plant data. A comparison has also been made with the simulation results of the PRO/II software. The comparison showed that the model was valid and very closely follows the trend of the plant data with a maximum discrepancy of -5%. It can be used as the prediction tool for the plant under operation.

The effect of operating conditions such as steam pressure and temperature, feed conditions such as pressure, temperature and flow rate on the quality of product in terms of RVP and sulphur content have been studied. The effect of steam temperature on both RVP and sulphur content is more pronounced than the effect of pressure. It has been found that the optimum steam temperature is 143°C which gives a RVP of 7.9 psia that is still in the range of on-specification product.

The effect of feed flow rate is also very significant compared to the effect of feed temperature and pressure. It has been found that to meet the RVP requirement, a tolerance of 40 to 110% feed flow rate can be used without manipulating the steam conditions.

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References

- ASTM D323-99a, 2012. Historical Standard: ASTM D323-99a Standard Test Method for Vapour Pressure of Petroleum Products (Reid Method), www.astm.org.
- Bao, J., Gao, b., Wu, X., Yoshimoto, M., Nakao, K. Simulation of industrial catalytic-distillation process for production of methyl tert-butyl ether by developing user's model on Aspen plus platform. *Chemical Engineering Journal*, 2002. 90(3): p. 253-266.
- Benoy, J. and Kale, R.N. 2010, "Condensate Stabilization" in *Offshore World, India*; Chemtech Foundation.
- ErikØi, L., Comparison of Aspen HYSYS and Aspen Plus simulation of CO₂ Absorption into MEA from Atmospheric Gas. *Energy Procedia*, 2012. 23: p. 360-369.
- Esparragoza, J.J. V, Iglesias-Silva, G.A., Michael W. Hlavinka, M.W., Jerry A. Bullin, J.A., 1992. "How To Estimate Reid Vapour Pressure (RVP) of Blends", *Hydrocrabon Processing*, 1992, p. 135-138.USA, Bryan Research & Engineering, Inc.
- Gary, J.H. and Handwerk, G.E. (2001). *Petroleum Refining: Technology and Economics* (4th ed.). CRC Press. ISBN 0-8247-0482-7.

- Kim, W.S., Yangb, D.R., Moona, D.J., Ahna, B.S., The process design and simulation for the methanol production on the FPSO (floating production, storage and off-loading) system. *Chemical Engineering Research and Design*, 2013.
- Lastari, F., Pareeka, V., Trebblea, M., Tadea, M.O., Chinnb, D., Tsaic, N.C., Extractive distillation for CO₂–ethane azeotrope separation. *Chemical Engineering and Processing: Process Intensification*, 2012. 52: p. 155-161.
- Lee, Y.J., S.-I. Hong, and D.J. Moon, Studies on the steam and CO₂ reforming of methane for GTL-FPSO applications. *Catalysis Today*, 2011. 174(1): p. 31-36.
- Liao, B., Lei, Z., Xu, Z., Zhou, R., Duan, Z., New process for separating propylene and propane by extractive distillation with aqueous acetonitrile. *Chemical Engineering Journal*, 2001. 84(3): p. 581-586.
- Mokhatab, S., Poe, W.A. and Speight, J.G. 2006, *Handbook of Natural Gas Transmission and Processing*, UK, Gulf Professional Publishing, Page 281.
- Behbehani, R.M., Atashrouz, E., 2011, *Fundamentals of Natural Gas Processing & Transmission*, Ayej; Tehran, Iran, Publisher.
- Oil and Industry Jargon De-coded, 1 November 2012
www.naturalhub.com/slweb/defin_oil_and_Natural_Gas.html
- Peng, D.Y., Robinson, D.B., 1976, A new two-constant equation of state, *Ind. Eng. Chem. Fundamental*, 15(1), 59-64.
- Peters, L., et al., CO₂ removal from natural gas by employing amine absorption and membrane technology—A technical and economical analysis. *Chemical Engineering Journal*, 2011. 172(2): p. 952-960.

Soo's, M., Graczova', E., Markos, J., Molna'r, A., Steltenpohl, P. Design and simulation of a distillation column for separation of dichloropropane from a multicomponent mixture. Chemical Engineering and Processing: Process Intensification, 2003. 42(4): p. 273-284.

Tavan, Y. and S.H. Hosseini, A novel application of reactive absorption to break the CO₂-ethane azeotrope with low energy requirement. Energy Conversion and Management, 2013. 75: p. 407-417.

www.intertek.com accessed on 5 Jan. 2014.

www.Intertek.com Accessed on 6 January 2014.Discovery Channel. 20 June 2012
<<http://curiosity.discovery.com/question/what-is-reid-vapor-pressure>>.

www.invensys.com, PRO/II Comprehensive Process Simulation, accessed on Dec. 2012.

Yea, G., Xiea, D., Qiaoa, W., Graceb, J.R., Limb, C.J., Modelling of fluidized bed membrane reactors for hydrogen production from steam methane reforming with Aspen Plus. International Journal of Hydrogen Energy, 2009. 34(11): p. 4755-4762.

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