



# **Process Simulation for Crude Oil Stabilization by Using Aspen Hysys**

**Hussein Al-Ali<sup>1\*</sup>**

<sup>1</sup>*Basra Oil Company, Basra, 61003, Iraq.*

## **Author's contribution**

*The sole author designed, analysed, interpreted and prepared the manuscript.*

## **Article Information**

DOI: 10.9734/JERR/2020/v16i417174

Editor(s):

(1) Dr. Syamsul Bahari Bin Abdullah, Universiti Malaysia Pahang, Malaysia.

(2) Dr. Guang Yih Sheu, Chang-Jung Christian University, Taiwan.

(3) Dr. Djordje Cica, University of Banja Luka, Bosnia and Herzegovina.

Reviewers:

(1) Andrii Torpakov, Institute of Pulse Processes and Technologies of NAS of Ukraine, Ukraine.

(2) Maria Dolores Robustillo Fuentes, Brazil.

Complete Peer review History: <http://www.sdiarticle4.com/review-history/59532>

**Original Research Article**

**Received 06 June 2020**  
**Accepted 12 August 2020**  
**Published 05 September 2020**

## **ABSTRACT**

A light fluid from different reservoir formation started recently to associate the production of the crude oil stabilization plant which is unfortunately not enough to release off all light components and as a results the true vapor pressure increased in the storage tanks more than 12 psi. From the results in Aspen Hysys, it was found that manipulating of working parameters for the existing plant likewise the inlet temperature, dry fluid flow rate, water flow rate and the temperature of the outlet fluid from Fired heater have no great effect on the true vapor pressure (TVP). The TVP at normal feed conditions of 50.5 C and for the plant with third and fourth stages are 14.96 Kg/Cm<sup>2</sup>. a and 10.23 Kg/Cm<sup>2</sup>. a, respectively. It was found that for the third stage, the changing in feed flow rates for both dry and water have no effect on the reducing TVP, while to stabilized the TVP for the exported crude oil within range of (68947.6 – 82737.1) Pa/(10 – 12) psia when the the fourth separator used in the process plant, the feed dry fluid flow rate (26.4 – 105.6) KBD, the minimum base sediment and water cut in the feed stream 4 Vol%, the inlet fluid temperature (43-51.5)°C and the differential temperature across the fired heater in range of (16-24)°C with feed temperature range (40-55)°C.

**Keywords:** *Aspen Hysys; True Vapor Pressure (TVP); crude oil stabilization process; light components.*

\*Corresponding author: Email: [alali.hsh@gmail.com](mailto:alali.hsh@gmail.com);

## 1. INTRODUCTION

### 1.1 Background

The X field is one of three segments of one giant oil field complex located in Southern Iraq lies approximately 65 km Northwest of Basra city.

The field currently has three existing degassing stations (DS6, DS7, and DS8) with a total of 10 oil/gas separation trains and three test trains; with a total nameplate capacity of approximately 600 kbd

Full well stream production from individual wells is combined with other wells in a system of field manifolds designed to connect from 4-18 wells each, generally from the same well row.

Production from each manifold is then routed to one of the three Degassing Stations (DSs).

All incoming lines from the manifolds tie into the degassing station inlet manifolds where production can be manually directed to any one of the separation trains, including the test train. The phase separation (oil, gas, water) into separated streams represents the first step in the plant processing of the product.

Once the separation is done, each stream undergoes the proper processing for further field

treatment. [1-4]. Crude oil consists of complicated mixtures which can be very tough to handle, meter, or transport. Further to the problem, it's also unsafe and uneconomical to ship or delivery these combinations to refineries and gas plants for the processing [5,6].

Crude oil stabilization is a pre-treatment process which involve the elimination of light hydrocarbons to reduce the vapor pressure to meet pipeline, storage or tanker Reid Vapor Pressure (RVP) specification [6,7].

The main target of this work is stabilized the exported crude oil from the process plant that handling different compositions of live oil and to minimize the TVP within range of (68947.6 – 82737.1) Pa/(10 – 12) psia and propose a solution for the existing plant to decrease the true vapor pressure.

### 1.2 Problem Statement

The first structure of the present crude stabilization plant depends on Mishrif formation fluid which has a medium density fluid, however as of late, a Zubair layer is associated with the plant which has fluid with higher API and totally content all the more light components that cause high true vapor pressure in crude oil from the third stage separator.

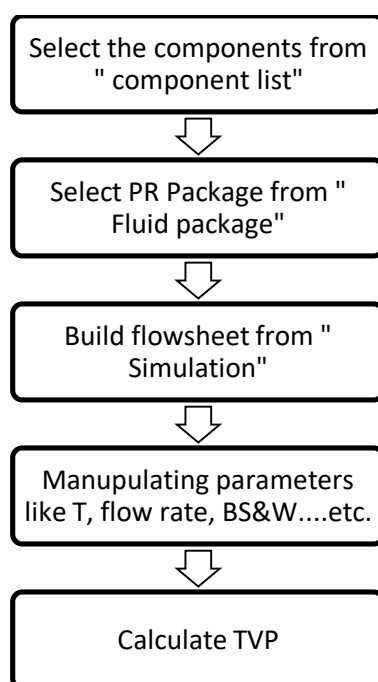


Fig. 1. A flow chart for the simulation procedure by Aspen Hysy

The unrefined oil from the third stage separators in the present plant flashes in the rough tanks ensuing the release of vapors to build up the true vapor pressure TVP in storage tanks. To avert existing rough tank overpressure, vapors are being discharged through select nozzles on the arch rooftop which requires extraordinary safety measures to get to the tanks zone where vent gases from the degassing drum and from the new capacity tanks are directed to the flare. As result of stabilization of unrefined petroleum from the process has a TVP go higher than the accepted limits of (68947.6 – 82737.1) Pa/(10 – 12) psia.

### 1.3 Objective of Work

The principle point of this project is to lessen the true vapor pressure (TVP) for the stabilized crude plant within the limits of 10-12 psi for storage/trade by adding a fourth stage separator to release the rest of gas with oil in the third stage that make TVP high. Along these lines, to achieve the main target, the following objectives should be accomplished:

1. Use the process plant data for the crude stabilization and compare them with the results of Aspen Hysys simulation.
2. Manipulate different parameters (Temperature, pressure, flow rate, etc.) to study the effect on the crude stabilization.
3. Optimize the inlet crude specification to maximize oil and gas production at the process plant.

Fig. 1 shows the flow chart of Aspen Hysys simulation procedure.

## 2. METHODOLOGY

### 2.1 Basis of Study

Aspen Hysys process simulation will use the following feed fluid composition as a basis for this work.

### 2.2 Process Description

Fig. 2 shows the process flow diagram (PFD) of the simulation of the main crude stabilization unit using Aspen HYSYS software.

Incoming gathering lines from the field manifolds typically tie into all station inlet manifolds. This allows production from any field manifold to be routed to any production train within the station.

The inlet reservoir fluid at a temperature and pressure of 50.5°C and 13.03 Kg/Cm<sup>2</sup>.a, respectively and Bs&WC of 15 Vol%.

The 1<sup>st</sup> stage separator operates at 13.03 kg/cm<sup>2</sup>.a. The first stage separator is a two-phase vessel with associated gas going to flare and liquid dumping to the 2<sup>nd</sup> stage separator across a level control valve. The second stage separator operates at a normal operating pressure of 4.58 kg/cm<sup>2</sup>.a. This separator is also a two phase vessel, with associated gas going to flare and liquid dumping either to the Desalter system across a level control valve.

Wet crude from the second stage separator flows into a balance vessel operating at 2.122 kg/cm<sup>2</sup>.a. The balance vessel de-gases the crude oil prior to being pumped into the liquid filled dehydrator. This wet crude is pumped into the dehydration vessel (electrostatic treater) operating at 7.178 kg/cm<sup>2</sup>.a. The electrostatic grid inside the desalting vessel improves oil/water separation in the liquid filled vessel. The water flows to a Coalescer and the crude flows to the desalting vessel after it is mixed with fresh wash water to dilute the salt with water and ensure a minimum water cut in the Desalter. The Desalter vessel operates at 5.981 kg/cm<sup>2</sup>.a to prevent gas break out in the liquid filled vessel. Electrostatic grids in Desalter and dehydrator vessels, improve water/oil separation by increasing the coalescing phenomena between salty water droplets. Per design, the fresh wash water is pumped to upstream of the desalting vessel, and the water stream separated from the desalting vessel is pumped into the dehydration vessel as the dilution water for the dehydrator.

Two heat exchangers are used to economize the heat losses, one for crude oil located before the Fired heater and in the downstream of balance vessel and the second heat exchanger is before the Desalter to increase the temperature of wash water.

Fired heater is used to accomplish the required temperature of fluid to separate the emulsion from crude oil in the Desalter and Dehydrator vessels where this temperature shall be 90°C.

The oily water is discharged from dehydrator to Coalescer. The heavy fluid in the three phases Dehydrator is fed to a Coalescer vessel at a pressure of 2.081 kg/Cm<sup>2</sup>.a where it skims oil from produced water and recycle it to the system

Table 1. The reservoir fluid composition for feed

Component	Reservoir Composition		M.wt ( g/mol )	Density (g/Cm <sup>3</sup> )	Component	Reservoir Composition		M.wt (g/mol)	Density (g/Cm <sup>3</sup> )
	Wt%	Mol%				Wt%	Mol%		
N <sub>2</sub>	0.061	0.316	-	-	C14	2.098	1.546	-	-
H <sub>2</sub> S	0.025	0.108	-	-	C15	1.990	1.370	-	-
CO <sub>2</sub>	0.340	1.128	-	-	C16	1.924	1.242	-	-
C1	3.445	31.402	-	-	C17	1.995	1.213	-	-
C2	1.809	8.798	-	-	C18	1.797	1.033	-	-
C3	1.882	6.241	-	-	C19	1.484	0.808	-	-
iC4	0.439	1.104	-	-	C20	1.689	0.874	-	-
nC4	1.467	3.691	-	-	C21	1.609	0.793	-	-
neo-C5	0.004	0.008	-	-	C22	1.490	0.701	-	-
iC5	0.811	1.644	-	-	C23	1.419	0.639	-	-
nC5	1.062	2.152	-	-	C24	1.328	0.573	-	-
C6	1.984	3.366	-	-	C25	1.269	0.526	-	-
Benzene	0.057	0.106	-	-	C26	1.228	0.490	-	-
C7	2.155	3.145	-	-	C27	1.188	0.456	-	-
Toluene	0.217	0.344	-	-	C28	1.173	0.434	-	-
C8	2.282	2.921	-	-	C29	1.154	0.413	-	-
Ethylbenzene	0.101	0.138	-	-	C30	1.099	0.380	-	-
m-andp- Xylenes	0.254	0.350	-	-	C+31	1.098	0.368	436.84	0.906
o- Xylene	0.151	0.208	-	-	C+32	1.003	0.325	450.87	0.909
C9	2.033	2.317	-	-	C+33	0.957	0.301	464.89	0.912
C10	2.508	2.577	-	-	C+34	0.919	0.281	478.92	0.914
C11	2.332	2.181	-	-	C+35	0.877	0.260	492.94	0.917
C12	2.208	1.895	-	-	C+36	39.421	7.118	810	0.955
C13	2.164	1.716	-	-	Total	100%	100%	-	-

Note: 1. The fluid composition analysis is based on dry basis mole fractions, plant Incoming Fluid Composition

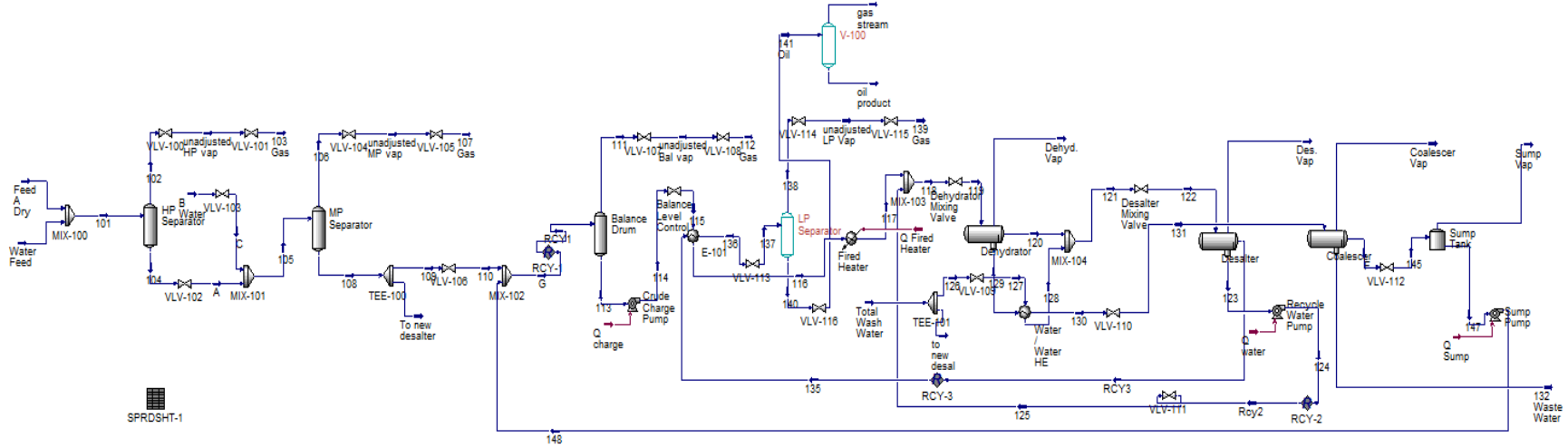


Fig. 2. Process flow diagram for the crude stabilization unit

again before the last one is sent to wastewater treatment unit. Finally, a two phase separator operates at 2.147 Kg/Cm<sup>2</sup>a is used to ensure the crude oil stabilization before the oil is fed into storage tanks.

### 2.3 Aspen Hysys Crude Stabilization Model (Pseudo Component)

For oil, gas and petrochemical applications, the Peng-Robinson EOS (PR) is generally the recommended property package to be used [8]. Where, PR Fluid Packages are most enhanced in HYSYS, which has highest T & P range, and

has special treatment for key components, largest binary interaction database: good standards for hydrocarbons.

Detailed inlet composition and feed properties used for this simulation are tabulated as per Table 2 and Table 3 respectively.

The phase envelope in Fig. 3 was calculated by Aspen Hysys on dry basis. The feed conditions of fluid is 50.5°C and 13.03 Kg/Cm<sup>2</sup>.a which is lie between the bubble curve and dew curve from the phase envelope. This means the approaching feed will be in two phase state.

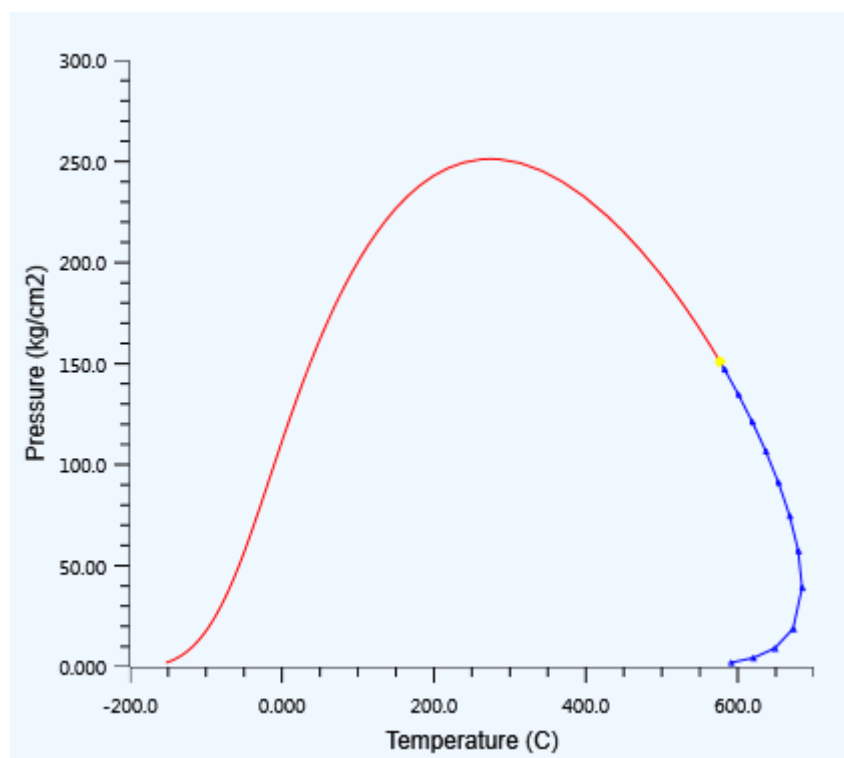
**Table 2. Feed composition**

Component	Mol %	Component	Mol %
N <sub>2</sub>	0.00109	C14	0.00536
H <sub>2</sub> S	0.000374	C15	0.00475
CO <sub>2</sub>	0.0039	C16	0.0043
C1	0.1088	C17	0.0042
C2	0.0305	C18	0.00357
C3	0.0216	C19	0.0028
iC4	0.0038	C20	0.00303
nC4	0.0128	C21	0.00275
neo-C5	0.000028	C22	0.0024
iC5	0.0057	C23	0.0022
nC5	0.00746	C24	0.001986
C6	0.0117	C25	0.00182
Benzene	0.00037	C26	0.0017
C7	0.0109	C27	0.00158
Toluene	0.0012	C28	0.0015
C8	0.01012	C29	0.00143
Ethylbenzene	0.00048	C30	0.00131
m-and p- Xylenes	0.00121	C+31	0.001275
o- Xylene	0.00072	C+32	0.001126
C9	0.00803	C+33	0.00104
C10	0.0089	C+34	0.000973
C11	0.0076	C+35	0.0009
C12	0.00656	C+36	0.02466
C13	0.0059	H <sub>2</sub> O	0.65356

*Important notes: 1. The simulation used by Aspen Hysys in this paper, used the water as a separate stream to mix it with the inlet oil stream (dry basis mole fraction) ranging of 15% of water to the inlet crude flow*

**Table 3. Feed stream properties**

Properties	
Vapor/Phase Fraction	0.15945
Temperature, °C	50.5
Pressure, Kg/Cm <sup>2</sup> .a	13.03
Molar Flow, kgmole/h	9612.53
Mass Flow, kg/h	585136.74
Std Liquid Volume Flow, barrel/day	105118
Molecular Weight	60.87
Liquid Mass Density @ std cond., kg/m <sup>3</sup>	899.1



**Fig. 3. Phase envelope curve for inlet feed**

### 3. RESULTS AND DISCUSSION

#### 3.1 Hysys Simulation (Pseudo Components) Validation

To guarantee the approval of the simulation that is done by this task, the stabilized crude oil composition taken from the plant is compared with the results of a simulation from Aspen Hysys as appeared in the figure below. The mole fraction is finished by dry basis where the water content is overlooked in the crude oil. In view of the information in Fig. 4, it very well may be seen that the HYSYS results are slightly match with plant data.

The chart appears, the two components have a high deviation in the simulation as compare with plant data which are:

- Pentane
- Heptane

In view of the rough compositional investigation, these segments were unfit to be measured because of co-eluting with different components. Along these lines, the sums were bunched together as lumped components.

The lumped components are tabulated as Table 4.

The composition of paraffinic components and heavy crude components from C11 to C36+\* are nearly equal for both plant data and the simulation. In addition, there are a trace amounts of CO<sub>2</sub>, H<sub>2</sub>S and N<sub>2</sub>.

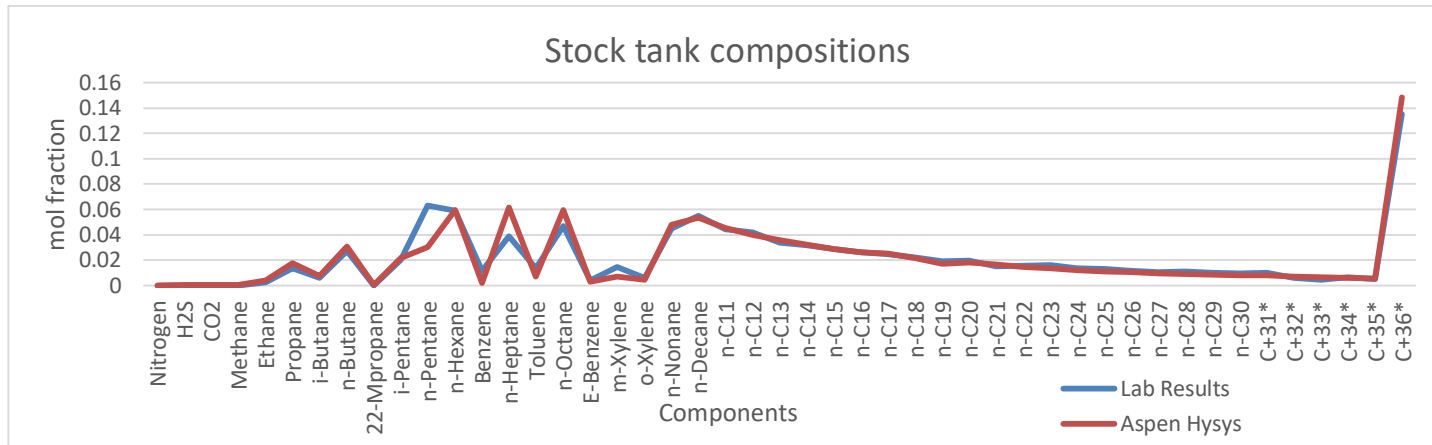
There are no significant contrasts and in this way, it is demonstrated that the simulation done by utilizing the HYSYS software is substantial and can be the basis of predicting tools for operational reason.

#### 3.2 Hysys Simulations (Pseudo Components)

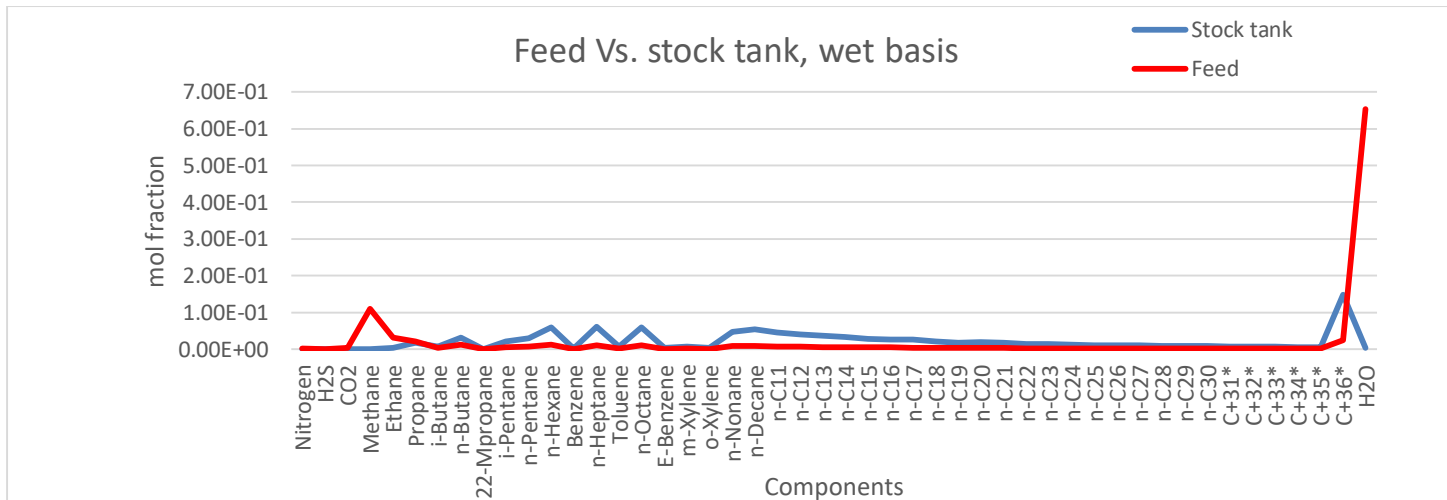
Fig. 5 shows a comparison between feed compositions for the Aspen Hysys with the stabilized crude compositions with free water content.

Based on the analysis data in Fig. 5, it's very well can be seen that the water mole fraction in the feed stream approaches to around 0.7. Aspen Hysys simulation can illuminate approximately 99% of the water content in the crude stabilization system to lessen the water content to 0.4 Vol% in the product stream.

Additionally, from Fig. 5, it's far seen that the intermediate components in the stabilized crude



**Fig. 4. Comparison between the Hysys simulation and lab results for the stock tank composition**



**Fig. 5. Inlet and outlet comparisons for Hysys simulation (pseudo components) wet basis**



**Table 4. Detailed compositions of lumped components for plant data**

C5*	i-pentane n-pentane cyclo pentane 2-methyl pentane 3-methyl pentane
C7*	iso-heptane n-heptane Methyl cyclo hexane Toluene

from Hysys simulation are higher than the feed flow, this is basically because of excessive pressure of vessels which are trap these components within fluid stream.

Fig. 6 demonstrate the distribution of components in the dry basis, without considering the water composition, in the inlet and outlet of Aspen Hysys simulation.

In view of essential analysis in Fig. 6, it is seen that the dry feed has high amounts of volatile components (C1 — C4) as compare with the stock tank crude. This is because the high majority of these components are flashed off from the stabilized crude.

On the other hand, the stabilized crude generated by Aspen Hysys contains intermediate – heavy hydrocarbons to results in the quality of the crude with medium API gravity where the stabilized crude is assessed to have an API gravity of 22 °.

### 3.3 Adding Fourth Stage Separator

The existing process stabilization plant sending the crude oil to storage tanks and then exported via a booster pump through exportation pipeline. The true vapor pressure for exporting crude oil has a value reach up to 143824.6 Pa / 20.86 psia for feed temperature 25°C in winter and around 14.96 psia at 50.5°C in summer when Fired heaters in service, but in some cases where heaters are out of service for maintenance, the TVP found to be 169473.1 Pa / 24.58 psia in summer and 267654.5 Pa / 38.82 psia in winter. Therefore, a fourth stage vessel (V-100) is suggested to add in the simulation of Aspen Hysys in the downstream of third separator where the discharge stream of 141 oil is fed into the new vessel as shown in Fig. 7, to evaluate the TVP of the produced stabilized crude from the process plant.

In this work, different parameters are manipulated to study the effect of the operating conditions on the crude stabilization process for the fourth stage and compare the results with the existing plant of 3<sup>rd</sup> stage only.

A normal incoming crude inlet to the terminal is at 88 Kbd at 50.5 °C, 13.03 kg/Cm<sup>2</sup>. a with BS &W of 17.64 Vol%. There are four major inlet properties studied in this paper of flow rate, temperature, free water content, and temperature difference around the Fired where set as manipulated variables with True Vapor Pressure of the stabilized crude.

Table 5 shows the properties of the stabilized crude obtained from the HYSYS simulation (Pseudo Components):

### 3.4 Effect of Different Operating Conditions

In order to study the effects of the manipulated parameters, all other values, except the parameter being studied need to be kept constant and compare results between the exist process stabilization plant with a new vessel.

#### 3.4.1 Effect of inlet dry fluid flow rate

The normal current feed flow rate used for the base case study is 88.0 kbd. The flow rate is then decreased to 20% and expanded to 230% with 10% interims.

From the diagram, the TVP reaches to consistent estimation of around 96526.6 Pa / 14 psia and increasing the flow rate has no impact on the TVP change for third stage separator.

As compared with the 4<sup>th</sup> stage separator the adjustment in feed flow rate impacts the TVP of the stabilized crude as the flow rate increases, the TVP is decreased to reach a value below 68947.6 Pa/ 68947.57 Pa / 10 psia when the flow rate exceeds 105 kbd. This reduction in TVP is

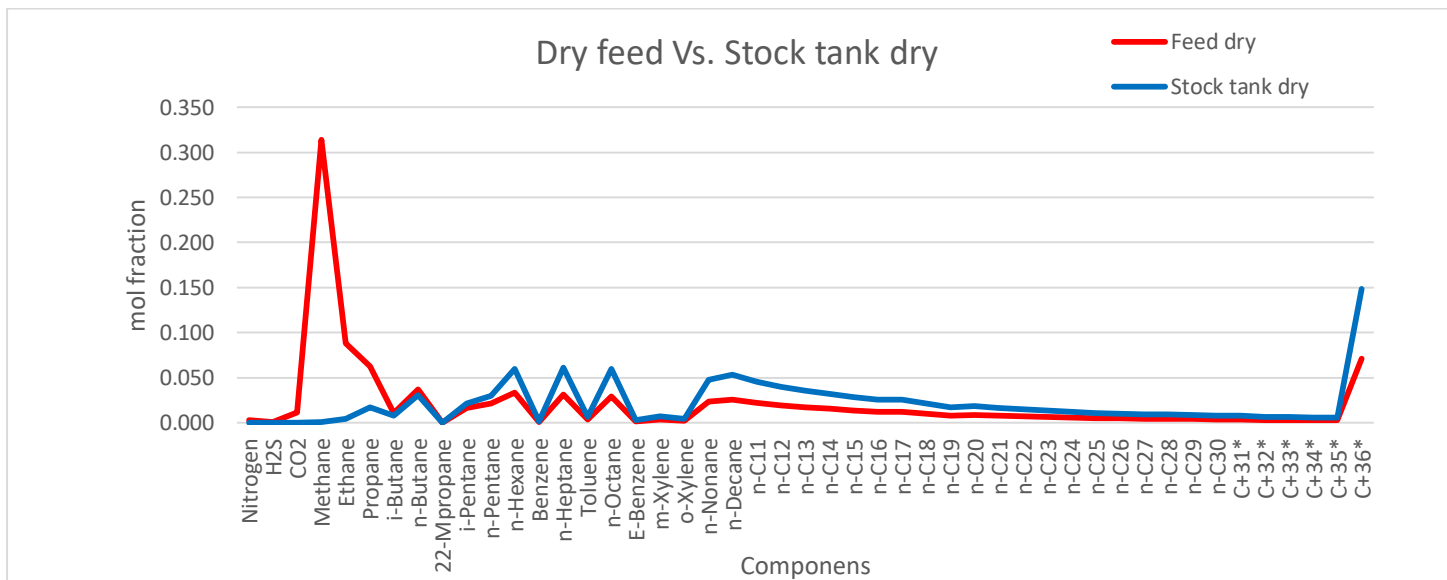
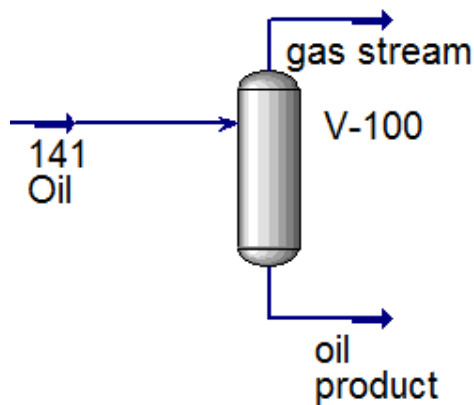


Fig. 6. Inlet vs. outlet composition of Hysys simulations (pseudo components) dry basis



**Fig. 7. Suggested fourth stage vessel**

when the flow rate increased, the Bs&W % will diminish in the feed stream which means the light components mole fraction will increment where the light components is effectively discharged through the separators. In this manner, the TVP would bit by bit diminishes with the increasing of feed flow rate.

When the 4<sup>th</sup> stage separator is used the maximum and minimum flow rate fed to the stabilized process plant should be 26.4 — 105.6 kBD to achieve the TVP within (68947.6 – 82737.1) Pa/(10 – 12) psia.

**3.4.2 Effect of base sediment and water (BS.W %)**

The current facility is fitted for processing 15700 barrels per day of free water content in the feed

crude in which proportional to 22.5 Vol% for 69880 barrels of crude oil production per day. Fig. 9 demonstrates the effect of the water inlet flow rate towards the TVP of the stabilized crude.

The TVP still higher than 82737.1 Pa/ 12 psia with the base BS&W% of 2% utilized in the simulation for the existing crude stabilization plant.

From the results of simulation shown in Fig. 9, as the Bs&W increased the TVP of stabilized crude will increase also. This because increasing the water content in the fluid stream means the total mixture density and mass flow of this steam will increase, which require a high duty of Fired heater to warm the process fluid to the required temperature before entering the LP vessel and the 4<sup>th</sup> stage vessel which it will affect the separation inside the pressure vessels along these lines to reduce the amount of volatile component that should be flashed off.

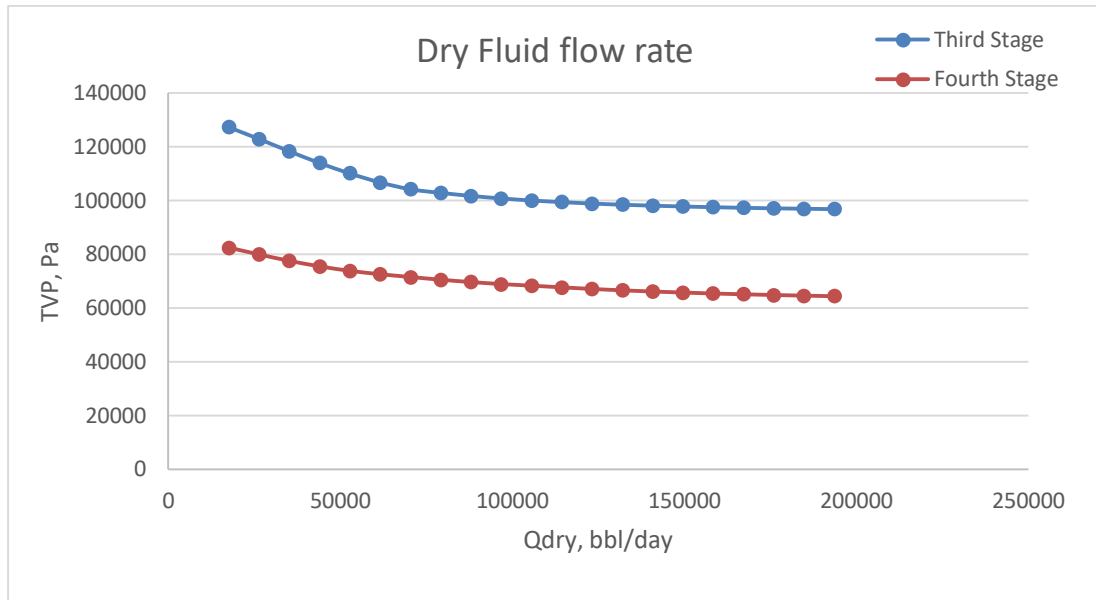
From results obtained from 4<sup>th</sup> stage separator the minimum Bs&W% that accomplished the TVP with in limit is 4% and the stabilized process plant can deal with over 40% of a Bs&W% in the feed stream.

**3.4.3 Effect of inlet temperature**

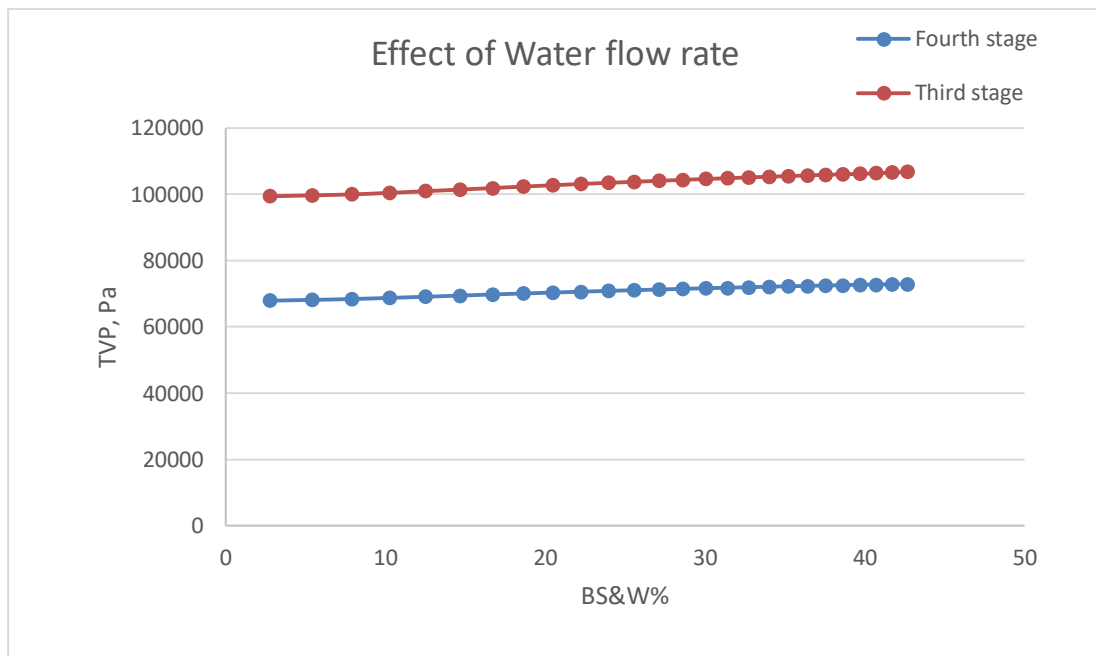
The inlet fluid temperature is typically 50.5°C in summer at station manifold. So to determine the impacts of feed temperature against TVP of stabilized crude, the temperature is diminished to 20°C and after that expanded to 70°C at 5°C intervals. Fig. 10 demonstrates how the adjustment in feed temperature influences the TVP of the stabilized crude.

**Table 5. Properties of fluid from 3rd and 4th stage separators**

Properties	Inlet Flow	Outlet Flow	
		3rd stage	4th stage
Molecular Weight	60.87	270	275.9
Mass Density, kg/m <sup>3</sup>	899.1	917.4	917.4
Operating Pressure, kg/Cm <sup>2</sup> .a	13.03	2.147	1.381
Temperature, °C	50.5	69.88	70.35
Total Mass Flow, kg/h	585136.74	424100	421900
Oil flow rate , Kbd	105.118	69.661	69.270
Water Flowrate, bbl/day	15.7	0.1886	62.34
BS&W, % Vol	15	0.207	0.09
Total Production, Kbd	120.39	69.85	69.207
GOR , m <sup>3</sup> /m <sup>3</sup>	-	1.4562	2.107
True vapor pressure , Pa	-	103145.57	70533.367



**Fig. 8. Effect of dry feed flow rate towards product TVP**



**Fig. 9. Effects of water flow rate towards product TVP**

In fact the inlet fluid temperature does not exceed the 60°C as maximum in summer and the lowest temperature is 25°C in winter, therefore the temperatures out of this range which meets the limit of TVP are not applying for the existing 3rd stage separator.

As can be found in the diagram, as the temperature of the feed is increased, the product TVP steadily decreased. This is because

increasing the inlet temperature will flash off the light components in vessels to decrease the TVP of the stabilized crude.

On the other hand, for the proposed new separator the minimum and maximum temperatures that cause crude stabilization plant accomplish the TVP (68947.6 – 82737.1) Pa/(10 – 12) psia is in a range of 43°C to 51.5°C. Below this minimum temperature means more duty

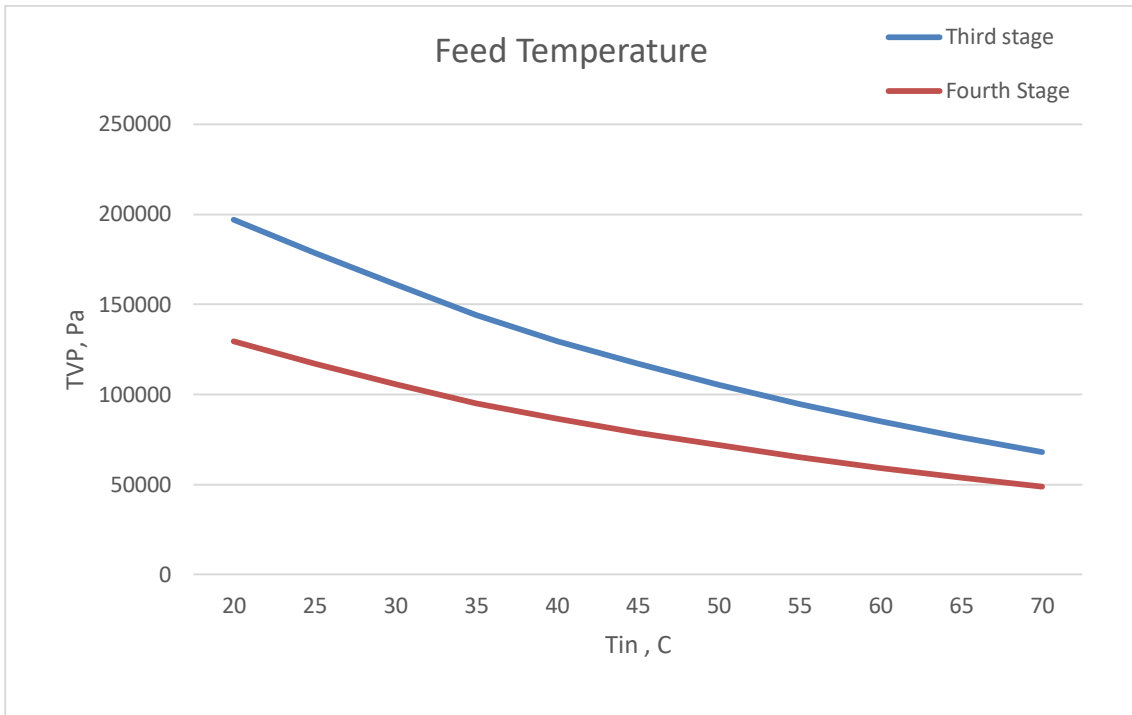


Fig. 10. Effect of feed temperature towards product TVP

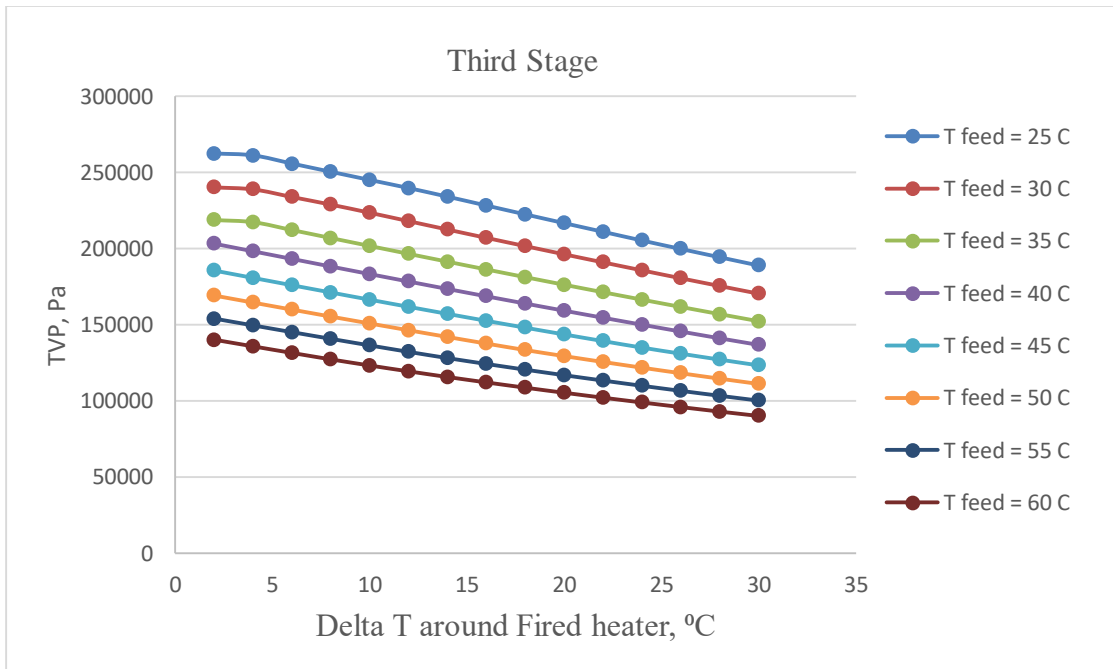


Fig. 11. Effect of fired heater temperature difference on TVP from 3<sup>rd</sup> stage

required to stabilize the crude and flashed off the light gases, while temperature above 51.5°C will release more light components and make oil in stored tanks heavier. In general, as the feed temperature increased, the TVP decreased. This is because the vapor fraction in the field

increased by increasing the inlet temperature and light components has greater chance to be released through separators before the fluid entered to the fired heater, which frees the rest of gas in the third and fourth stages to decrease TVP.

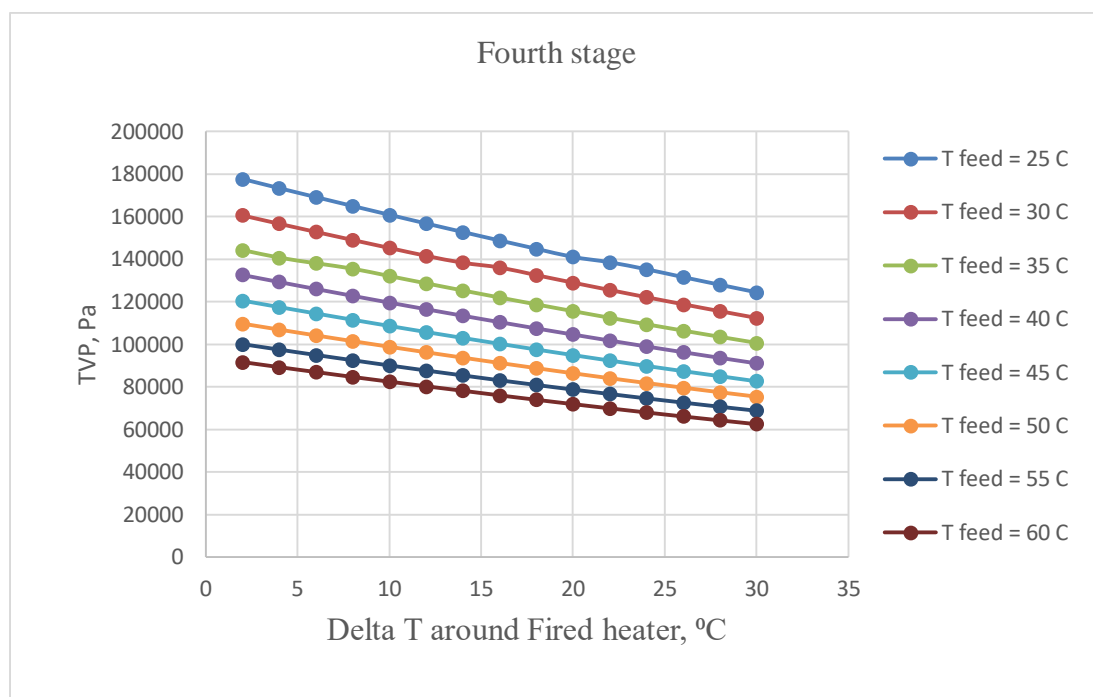


Fig. 12. Effect of fired heater temperature difference on TVP from 4<sup>th</sup> stage

### 3.4.4 Effect of Fired heater's temperature difference and feed temperature on TVP

To study the effect of the temperature difference around the Fired heater on TVP, all other conditions should be kept constant except the feed temperature manipulated in range of (25 – 60)°C.

Figs. 11 & 12 show as the temperature difference across the Fired heater increased the TVP decreased for each specific feed temperature, this is because increasing the heat duty in the Fired heater increased the temperature of the outlet fluid to gas off the light components through the next vessels.

As compare between the two above charts, Fig. 12 shows the proposed process of adding a fourth separator to meet the requirement of TVP within range of (68947.6 – 82737.1) Pa/(10 – 12) psia for temperature difference a round the Fired heater of ( 10 – 30 )°C and the fluid feed temperature to the plant is (45 – 60)°C, respectively.

The charts demonstrate the ideal temperature difference that required to achieve the TVP within range of (68947.6 – 82737.1) Pa/(10 – 12) psia is 10-30°C for the suggested fourth stage separator with feed temperature range 60°C – 45°C,

respectively. But the existing plant with 3<sup>rd</sup> stage vessel is cannot handle the fluid to achieve the TVP within limits for the produced oil as shown in Fig. 12.

## 4. CONCLUSION

Process simulations of a crude oil stabilization unit are conducted using Aspen HYSYS software to study the effect of the operating conditions which cause the production of off-specification product. TVP has been set as the criteria for the off-specification conditions of the stabilized crude oil in the range of (68947.6 – 82737.1) Pa/(10 – 12) psia.

A comparison has been made between the actual results of plant data and that is obtained from Aspen Hysys to validate the process simulation which shown that the model was valid and very closely follow the trend of the plant data and can be used it as a tool to predict the plant operation.

The effect of operating parameters, for the existing plant and for the proposed fourth stage vessel, such as feed flow rate, feed temperature, free water flow rate, and fired heater outlet temperature on the quality of crude oil in terms of TVP and have been studied.

It has been found that the TVP requirement for the existing plant cannot be met under present operating conditions.

But by adding a fourth stage separator, it has been found that the inlet dry flow rate should be around (26.4 – 105.6) kbd, whereas the BS.W% should be in the range of 4% and less than 40%, the inlet temperature between (43 – 51.5)°C.

### COMPETING INTERESTS

Author has declared that no competing interests exist.

### REFERENCES

1. Manish VS. Practical selection and design guide: For gas liquid separators with optimization; 2000.
2. Rahmanian N, Aqar DY, Dainure BMF, Mujtaba IM. Process simulation and assessment of crude oil stabilization unit. Asia-Pac J Chem Eng. 2018;e2219. Available:<https://doi.org/10.1002/apj.2219>
3. Abdel-Aal HK, Aggour M, Fahim MA. Petroleum and Gas Field processing, by Marcel Dekker Inc; 2003.
4. Arnould K, Stewart M. Surface Production Operations, Volume 1: Design of Oil Handling Systems and Facilities. Gulf Professional Publishing; 3 edition. September 13; 2007.
5. Al-Mhanna NM, Simulation of high pressure separator used in crude oil processing. Processes. 2018;6:219. DOI:10.3390/pr6110219
6. Edwin M, Abdulsalam S, Muhammad IM. Process simulation and optimization of crude oil stabilization scheme using Aspen-HYSYS Software. International Journal of Recent Trends in Engineering & Research. DOI: 10.23883/IJRTER.2017.3230.MIIUW
7. Manning FS, Thompson RE. Oilfield processing of petroleum: Crude oil. Pennwell Books: Tulsa, OK, USA. 1991;2.
8. Towler G, Sinnott RK. Chemical Engineering Design: Principles, Practice and Economics of Plant and Process Design; Elsevier: Waltham, MA, USA; 2012.

© 2020 Al-Ali; This is an Open Access article distributed under the terms of the Creative Commons Attribution License (<http://creativecommons.org/licenses/by/4.0>), which permits unrestricted use, distribution, and reproduction in any medium, provided the original work is properly cited.

*Peer-review history:*  
*The peer review history for this paper can be accessed here:*  
<http://www.sdiarticle4.com/review-history/59532>