

Simulation and Optimization of H₂S Expulsion from Crude Oil with the Use of Equilibrium Model

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Abstract. Equilibrium model approach has not been used for simulation of stripping towers as the trays of those towers are non- Equilibrium in real service conditions. This paper pursues an effective approach for simulating a cold stripping tower with new trays of centrifuges by applying Equilibrium model through Aspen plus software. In this paper, Peng Robinson (PR) equation has been used as equation of state (EOS) and vaporization efficiency of components in order to change the conditions of trays from Equilibrium to non-Equilibrium.

Experimental results arisen by cold stripping tower situated at Lavan Island in Iran were compared with the results of simulation in order to validate efficiency of the proposed simulation approach. Topics like process convergence and analyzing the effective parameters on reduction of H₂S and optimizing of the process have been considered in this paper. The performance of the tower improved considerably by fulfillment of operational optimized conditions and reducing amount of H₂S in oil from 175 to 136 ppm .

Keywords: Simulation, Stripping tower, Hydrogen sulfide (H₂S), Crude oil, Aspen Plus.

1. Introduction

Some essential processing such as vapor pressure reduction, desalination, sweetening and promotion of quality (e.g. API), etc. have been performed on crude oil to use in refinery and transportation. Such processes are known as crude oil stabilization [1]. Due to the market & environment requirements, the crude oil should be produced with level of 15 ppm of H₂S. Some chemical methods was used for improving efficiency of H₂S expulsion from crude oil (e.g. DMC&DMD) [2] but for mechanical method crude oil sweetening is usually performed by towers furnished by trays with the use of different approaches such as cold and hot stripping via making contact between sour oil and sweet gas which results in mass transfer between liquid & gas phases in the related trays as a result H₂S gas transformed from oil to gas [3,4].

There are two ways for simulation by Aspen plus: Equilibrium Stage Method and Rate Base Method. Non Equilibrium shall be applied when the type and characteristics of tray are available & definable by software [5]. As the used centrifugal tray had novel design and correspondingly there are limitation on selection of type and efficiency of the tray, RadFrac model has been employed in this research work. By comparing the results of simulation and real experimental data acquired from stripping tower, it was demonstrated that it is possible to simulate a tower with negligence of tray model and efficiency.

To determine the effect of different processing factors over H₂S decrement & API increment, the sensitivity analysis was carried out and final by applying of those parameters as variables, the process optimization was performed.

2. Methods

In this research work, a novel type of tray (centrifugal tray) has been designed and utilized instead of the previous valve tray design to promote efficiency and reduction of total pressure drop of stripping tower. Also,

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the numbers of the used trays were decreased from 21 to 20. The experimental results of crude oil analysis, sour & sweet gas have been achieved by ASTM standard & Research Institute of Petroleum Industry (RIPI).

Two flows of residual oil & associated gas which reported at the laboratory have been mixed under temperature of 76 F and pressure of 62.7 Psia with gas oil ratio (GOR) of 46.82 to generate entrance oil flow in the tower.

As mentioned earlier, the number of the new trays applied in stripping tower was 20 and the total pressure drop of the tower is 2 psig. Other operational parameters of the tower have been depicted in result and discussion section (Table 3).

Process simulation was performed by Aspen Plus software and the Radfrac was selected for column. Peng Robinson equation was applied as there are some components like H₂S & CO₂ which have low polarity. Fig. 1 shows simulation procedures schematically.

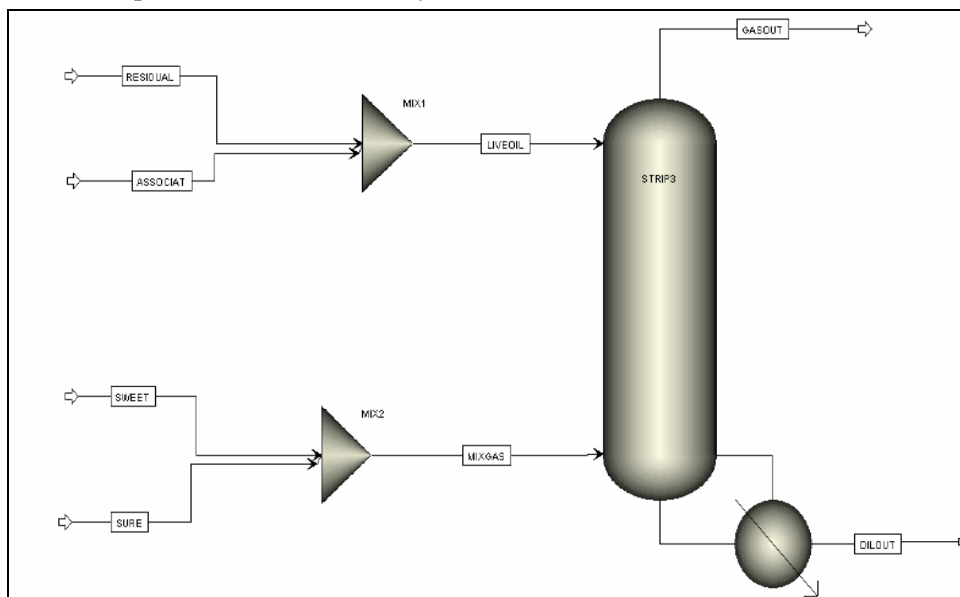


Fig.1: Schematic of simulation procedures

Initially the process persisted against convergence. So, estimation of temperature of each tray was used as an efficient approach to convergence the process. Such estimation of temperature was performed considering temperature of entrance oil and gas in the tower and with regard to the initial software calculation (table 1).

Table.1: Estimated temperature of trays

Tray No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Temp. (F)	99	98.5	98	97.5	97	96.5	96	95.5	95	94.5	94	93.5	93	92.5	92	91.5	91	90	88	87

Then the proposed simulation approach was tested against real data obtained from the tower by changing vaporizing efficiency for each component in the software (Table 2).

Table.2: Component vaporization efficiency.

Component	H2S	C1	C2	C3	iC4	nC4	iC5	nC5	C6	C7	C8	C9	C10	C11	C12†
Vaporization Efficiency	0.7	-	0.58	0.56	0.8	0.93	2.3	3.7	-	1.4	1.4	0.8	1	1.05	1.3

Simulation procedure and operation steps have been demonstrated schematically in figure 2.

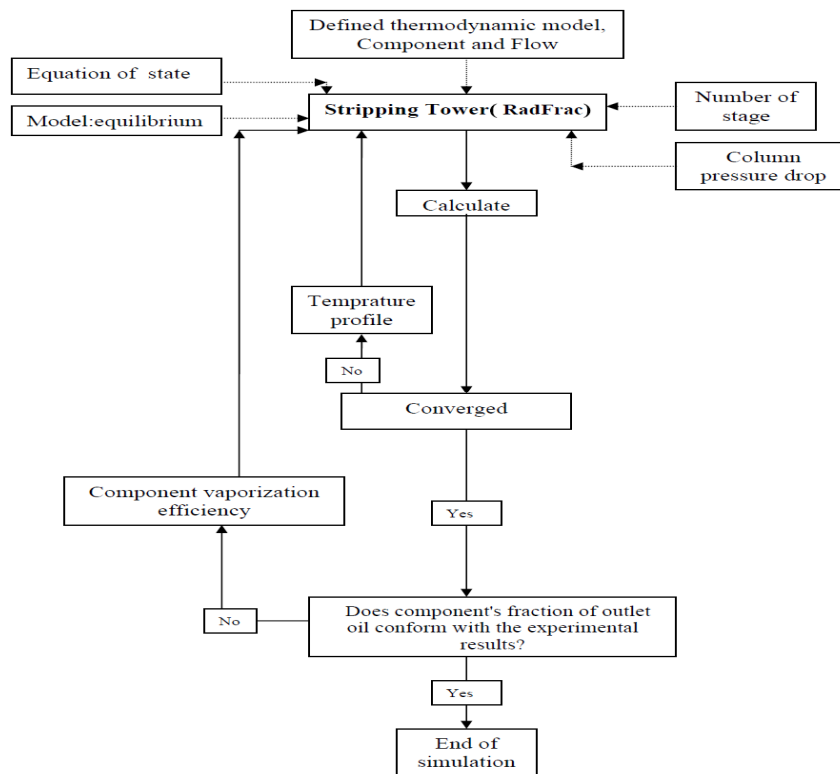


Fig. 2: Flowchart of simulation in the stripping tower

3. Results & Discussion

Process parameters of stripping tower and comparison of the experimental and simulation results of output oil have been demonstrated in Table 3 which it proves accordance of experiments and the simulation.

Table.3: The results of experiments & simulation plus comparison of output oil characteristics.

Parameter	Lab.Result					Simulation Result		
	Sweet Gas	Sure Gas	Residual Oil	Associat Gas	Live oil	Oil out	Oil out	Error
Temp(F)	115	130	69.8	69.8	76	120	123.9	-3.9
Pressure (Psia)	54.7	39.7	14.696	14.696	62.7	29.5	30.7	-1.2
Flow (Cu.ft/hr)	3.06×10^6	9.67×10^6	7055.54	61834.66	7306.98	7550	7914.24	-364
M.W (lb _m /lbmol)	31.2	53.85	-	-	210	198	197	1
Density (lb _m /Cu.ft)	-	-	-	-	51.35	51	52.1	-1.1
%mol								
H2S	1472 ppb	0.23	0.00	3.68	0.34	175ppm	165 ppm	10ppm
N2	0.32	0.04	0.00	0.01	0	0	6 ppm	-
CO2	0.57	0.16	0.00	2.63	0.24	0.01	90 ppm	-
C1	42.24	1.07	0.00	13.58	1.24	0.1	0.1	0
C2	25.55	6.98	0.83	20.87	2.66	1.62	1.0	0.62
C3	19.08	41.89	3.66	28.60	5.93	9.03	9.1	-0.07
iC4	2.92	10.23	1.76	6.65	2.21	3.39	3.3	0.09
nC4	5.33	20.89	4.49	13.46	5.31	7.56	7.6	-0.04
iC5	1.21	5.44	1.57	3.43	1.74	1.77	1.7	0.07
nC5	1.21	5.70	1.32	3.45	1.51	1.24	1.5	-0.26
C6	0.89	4.42	12.02	2.40	11.14	11.94	11.8	0.14
C7	0.45	2.13	9.42	0.96	8.65	8.34	11.1	-2.76
C8	0.2	0.77	9.20	0.26	8.39	8.77	11.3	-2.53
C9	0.02	0.05	7.55	0.01	6.86	6.93	4.6	2.33
C10	0	0	7.07	0.00	6.43	5.81	4.2	1.61
C11	0	0	6.29	0.00	5.72	5.04	4.5	0.54
C12+	0	0	34.82	0.00	31.65	28.43	28.1	0.33
GOR(Cu.ft/bbl)			46.82					

3.1. The analysis of sour/sweet gas flows and tower pressure effects on the H₂S proportion of output oil

The analysis of sensitivity was carried out with the use of both experimental and simulation data and the following inputs: change of sour gas from 0.2 to 3.5 (MMscfd), the sweet gas injection with constant flow of 0.8 (MMscfd) with crude oil flow rate of 30000 BPD. The results have been depicted in figures (3a) & (3b).

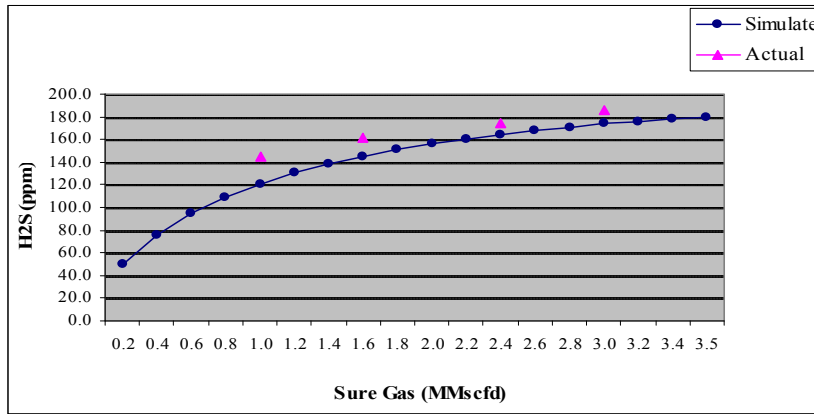


Fig. 3a: Effect of sour gas change on H₂S in output oil

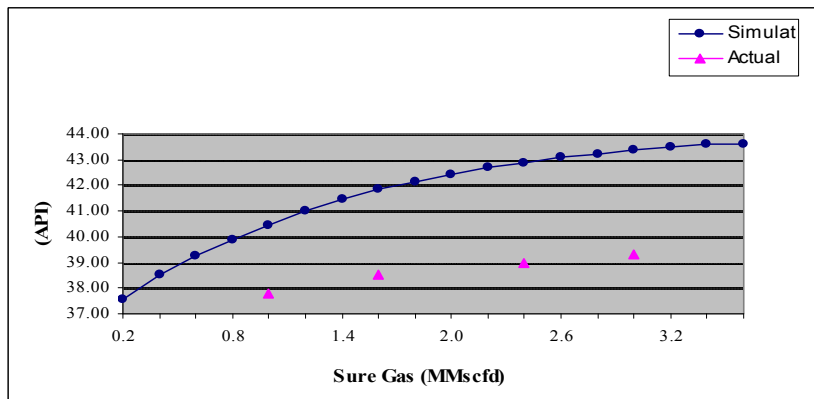


Fig. 3b: Effect of sour gas change on API in output oil

As seen in figures (3a) & (3b), the experimental and simulation results has a fine conformity and also H₂S & API of the output oil increased by more sour gas proportion.

Diagrams in Figures 4a, b have been obtained by applying the following changes in the process: Changing the rate of sweet gas from 0.2 to 0.8 (MMscfd) and injection of sour gas in the constant rate of 2.4 (MMscfd) with barrel per day (BPD) of 30000.

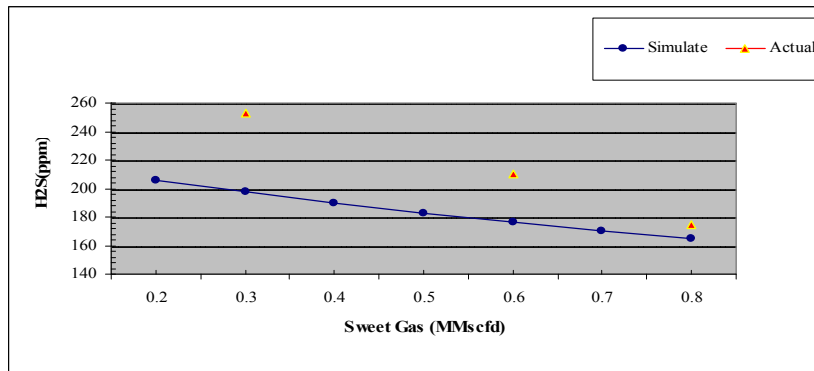


Figure.4a: The effect of sweet gas change in output oil H₂S

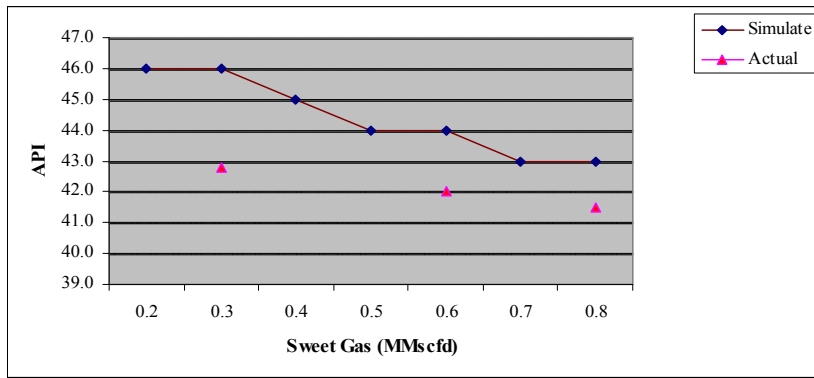


Figure.4b: The effect of sweet gas change in output oil API

As seen in the figures (4a) & (4b), the content of output oil H₂S decreased by applying more sweet gas with a fairly fast slope in a linear manner. Also the content of API has decreased by applying more sweet gas. As seen, the experimental results depicts descending curve.

There is a curve in figure (5a, b) following change of tower pressure in the range of 28-33 (Psia) under sweet gas injection with rate of 0.8 (MMscfd), sour gas within 2.4 (MMscfd) and input oil of 30000 BPD.

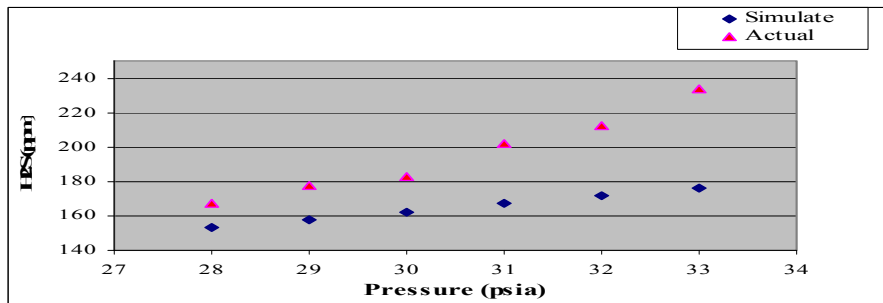


Figure.5a: The effect of stripping tower pressure change on output oil H₂S

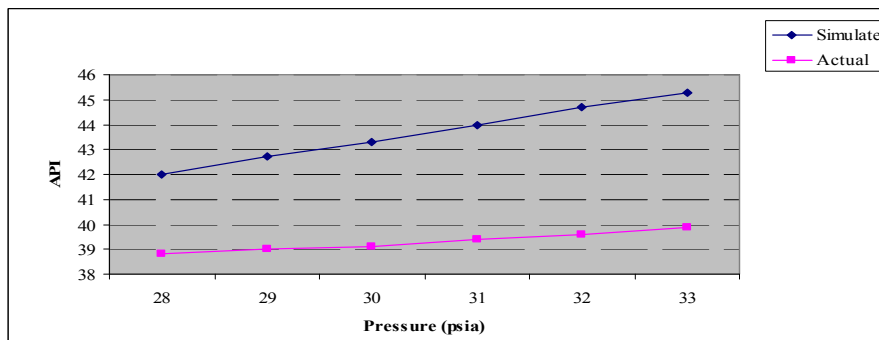


Figure.5b: The effect of stripping tower pressure change on output oil API

The content of output oil H₂S with partially steep & linear slope increased by pressure increment according to the figures (5a) & (5b) as a result the API rate increased.

3.2. Stripping Tower Optimization

Considering the sensitivity analysis of process parameters which was described in the previous section and with consideration of this issue where API increased while H₂S decreased, we produce more premium oil then we optimize the process via changing the sour & sweet gas , pressure tower and although determination of constraints for inlet oil and density of oil output. Finally the optimized parameters were validated by checking the stripping tower experimentally via optimized conditions. The results of optimization and conditions have been demonstrated in Table 4.

Table.4: Results of optimized stripping tower by simulation

Variable	Variable Range	Constraint	Optimum
H ₂ S (ppm)	-	-	128
Live Oil (BPD)	-	≤34200	30962
Sour Gas (MMscfd)	0.2-3.5	≤3.5	1.36
Sweet Gas (MMscfd)	0.2-1.2	≤1.2	1.19
Pressure (Psia)	28-32	-	32
Sour Gas Temperature (F)	-	-	130
Sweet Gas Temperature (F)	-	-	115
Density of Oilout (lb _m / ft ³)	-	≤51	51

It is evident from Table 4 that the amount of H₂S shall reach optimally 128 (ppm) under tower pressure of 32 (Psia), injection of sour gas in rate of 1.36 (MMscfd) and sweet gas in the rate of 1.19 (MMscfd), with no reduction of oil density (API).

4. Conclusion

The following results can be inferred from this research work:

- The present paper represents a modern method for real stripping tower simulation without caring the form of tray & total efficiency of the relevant tower (Fig. 2).
- With more sweet gas flow the content of output oil H₂S shall be decreased in the tower but in contrast the quality is reduced (Figs. 4a & 4b).
- The content of output oil H₂S shall be increased by increment of sour gas flow, the rate of quality shall be increased too (Figs. 3a & 3b).
- The quality of oil and the amount of H₂S would increase by increment of tower pressure (5a & 5b).
- The stripping tower was subjected in optimal conditions, therefore it was established that the amount of API & H₂S of the output oil from stripping tower acquired within average of 136 (ppm) & 38.6 by increment of sweet gas injection from 0.8 (MMscfd) to 1.2 (MMscfd), while the sour gas injection decreased from 2.4 to 1.4 (MMscfd) and increment of tower pressure from 28.7 to 32 (Psia) with input oil of about 30000 BPD. The rate of H₂S decreased to 39 (ppm) but API increased 0.7 degrees in contrasting to the primary indications which it demonstrates oil quality improvement of tower output.
- If the content of H₂S should be decreased more by available equipments, it is better to apply alternative methods like hot strapping or DSC which were investigated economically and by process.

5. Acknowledgements

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6. References

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