

Simulation of CO₂ removal in a split-flow gas sweetening process

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Abstract—Split-flow gas sweetening is known to consume less energy than a conventional gas sweetening process when the inlet sour gas contains a high concentration of acid gases. In this work, a computer simulation of a split-flow natural gas sweetening process based on absorption/stripping process with alkanoamine (MEA and DGA) solutions, using Aspen plus, was performed. The input of parameters such as the concentration of sour gases (CO₂, H₂S) in the feed gas has been examined. Simulation results show that the split-flow gas sweetening process can reduce the reboiler duty of a stripping tower better than the conventional gas sweetening process according to the concentration of CO₂ in the feed gas.

Key words: Split-flow, CO₂ Removal, Gas Sweetening, Aspen Plus

INTRODUCTION

The removal of carbon dioxide and hydrogen sulfide from natural gas is a process of substantial industrial importance because of health and environmental problems and lower transmission costs. Since the 1930s various amine solutions have been used widely in the natural gas sweetening process [1]. The conventional gas sweetening processes consist of the absorption/stripping of CO₂ and H₂S with aqueous amine solutions. Gas sweetening processes constitute about 70% of the processes used for treating natural gas [2]. Since acid gas removal from natural gas requires regeneration of alkanoamines, the energy requirement for the stripping section will be much greater when the natural gas contains a high concentration of acid gases. Many researchers have studied and modified the conventional gas sweetening process with the various alkanoamine solutions such as monoethanolamine (MEA), diethanolamine (DEA), di-2-propanolamine (DIPA), methyldiethanolamine (MDEA), and diglycolamine (DGA). The MEA and DEA are very reactive and thus are able to absorb acid gases at a fast rate [3]. However, the maximum CO₂ loading capacities of MEA and DEA solutions are less than that of other amine solutions, such as DGA. The stripping of CO₂ from MEA or DEA during solvent regeneration requires a large amount of energy input as compared to DGA. The advantage of DGA solutions was that it could be operated at high concentrations, which resulted in lower circulation rates and reduced energy requirements for regeneration [4].

Mathematical modeling and computer simulation of gas sweetening have been performed to understand how the design variables affect each other in a process. Weiland et al. [5,6] and Kvamsdal et al. [7] modeled the conventional amine process and predicted the effects of the MEA concentration on the overall mass transfer coefficient of various gas liquid rates and operating pressure. Idem et al. [3], Lee et al. [8] and Mofarahi et al. [9] simulated CO₂ removal with other solutions such as DEA, MDEA and DGA in a conven-

tional gas sweetening process. Problems of the conventional amine process such as a large amount of energy consumption and increasing amine loss were greatly improved by Ball and Veldman [4]. Ko et al. [10] investigated the modeling and simulation of the Ball and Veldman process, which showed more CO₂ removal and the lower energy consumption than the previous amine process.

When sour natural gas contains a high concentration of acid gases, the modified amine process which operates with two solvent streams called “a split-flow process” can reduce energy consumption [1,11]. Many researchers have already studied the advantages of the split-flow amine process in a gas sweetening process. Rahimpour and Kashkooli [11] developed the mathematical modeling of CO₂ absorption in DEA solution and examined the performance of a split-flow absorber. Mohebbi and Moshfeghian [1] devised a new method of calculating the solvent stream rates in a split-flow amine process with various amine solutions such as MEA, DEA and MDEA.

In this study, the computer modeling and simulation of the split-flow amine process treating the natural gases which contain various CO₂ concentration (1.7 mole%, 13.6 mole%, and 42.7 mole%) were performed. The CO₂ removal from a natural gas and the reboiler duty in the stripping section was compared with that of the conventional amine process. The effect of different amine solutions (MEA and DGA) on the performance of a split-flow absorber/stripper was investigated and the consequences of the different operation conditions discussed. A modeling and simulation of those processes were performed using the commercial software, Aspen Plus [12].

PROCESS DESCRIPTION

A schematic diagram of a conventional natural gas sweetening process designed by Ball and Veldman is shown in Fig. 1 [4]. The sour natural gas is contacted counter-currently with the lean amine solvent in the absorber. The effluent gas which contains only a very small amount of carbon dioxide leaves overhead from the column. The rich amine solution loaded with CO₂ is heated against the returning hot regenerated lean amine solution in a heat exchanger and

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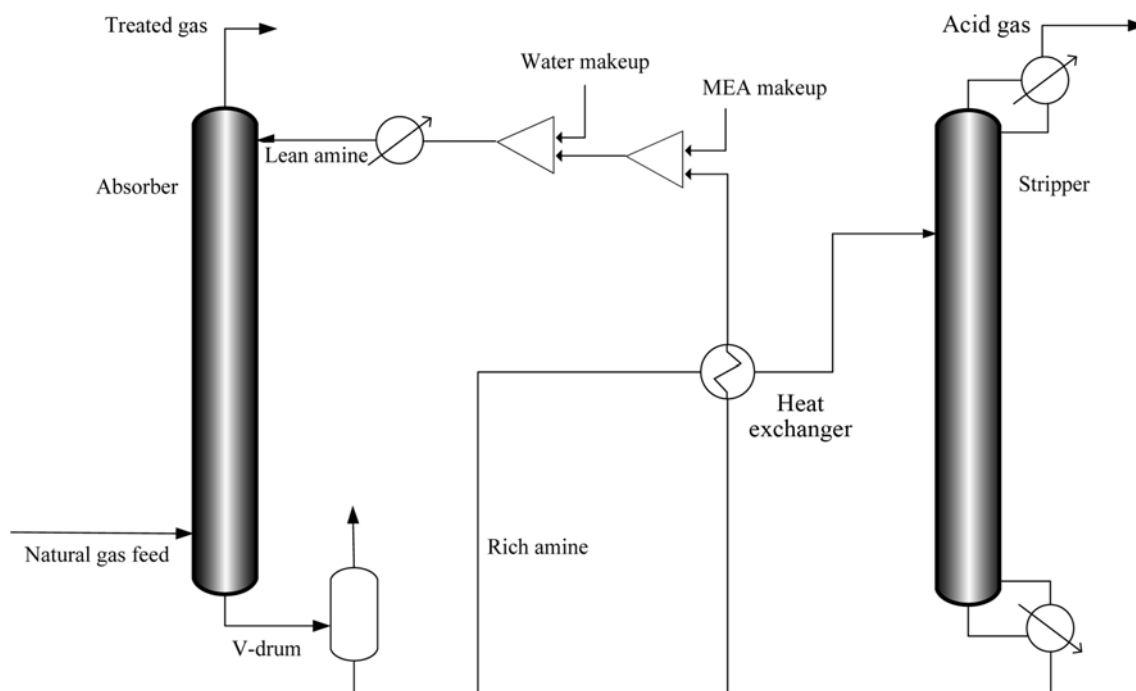


Fig. 1. The conventional amine process by Ball and Veldman [4].

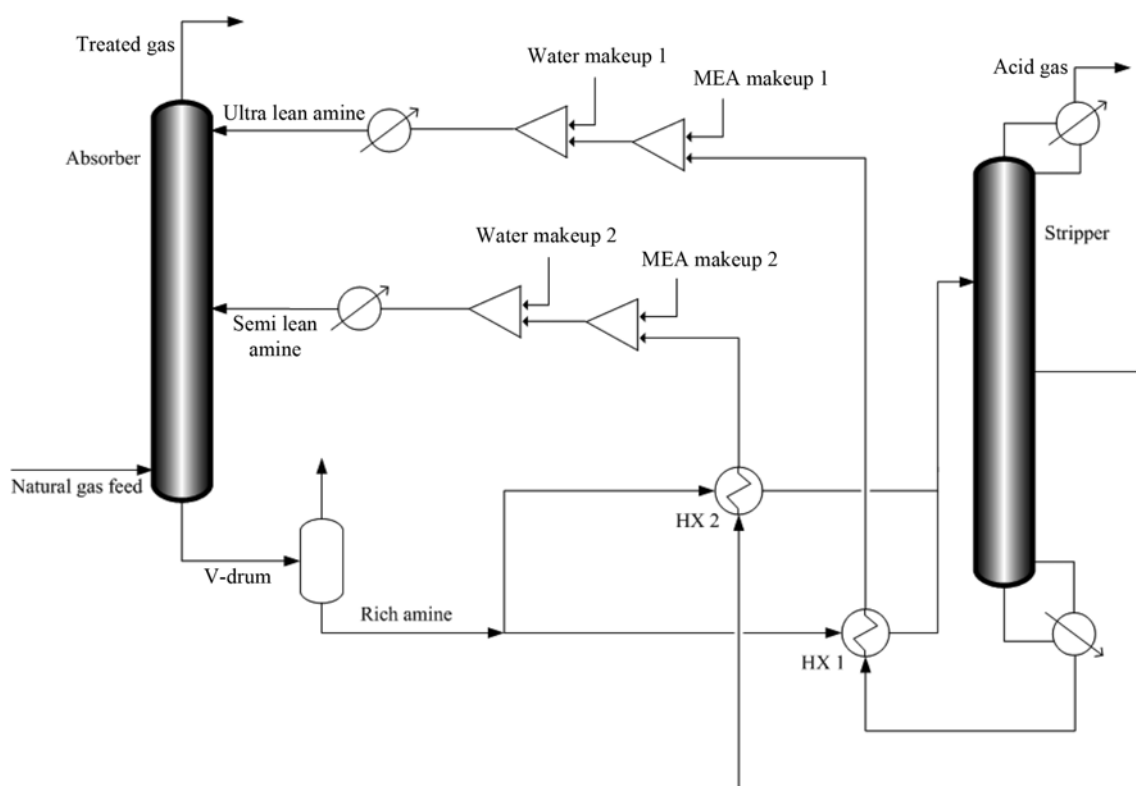


Fig. 2. The split-flow amine process [1].

enters the stripper where the remaining dissolved CO_2 and H_2S are stripped by counter-current contacting with a steam generated by a reboiler. The lean amine solution from the bottom of the stripper is cooled and fed into the top of the absorber.

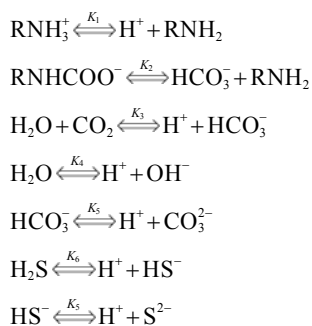
Fig. 2 shows a schematic diagram of a split-flow amine process

[1]. A portion of the lean amine solution is taken from the stripper as a side stream, cooled, and fed into the tray some distance below the top of the absorber. Whereas, the major portion of the lean amine solution from the bottom of the stripper is cooled and fed into the top of the absorber. Using a split-flow configuration in which the

semi-lean amine is fed to some midpoints of the absorber while the ultra lean amine is fed at the top, the split-flow amine process has an advantage of energy reduction in the stripper [13,14]. The amine in the side draw is not completely stripped and has some residual CO₂ and H₂S. However, when it is contacted with the incoming sour natural gas, it can still pick up the majority of the CO₂ and H₂S in the absorber [1].

SIMULATION

The equilibrium between acid gases, amines and H₂O is assumed to be governed by multiple equilibrium reactions. The equilibrium reactions included in the design model are as follows [10]:



where RNH₂ represents alkanoamines, such as MEA and DGA.

The thermodynamic and transport properties were modeled using the so-called “amine property method,” which describes the amines-acid gases-H₂O system with the Kent-Eisenberg method for K-values

Table 1. Conditions required for using the “Amines Property Method” describing the amine systems [12]

	MEA	DEA	DIPA	DGA
Temperature (°F)	90-280	90-275	90-260	90-280
Maximum H ₂ S or CO ₂ loading (mole gas/mole amine)	0.5	0.8	0.75	0.5
Amine concentration in solution (mass percent)	5-40	10-50	10-50	30-75

Table 2. Feed stream information [16]

	Case 1	Case 2	Case 3
	A natural gas from Canada (Alberta)	A natural gas from Miskar Field Tunisia	A natural gas from Western Colorado
Components	Mole percent	Mole percent	Mole percent
Helium	0	0	0
Nitrogen	3.2	16.903	26.1
Carbon dioxide	1.7	13.588	42.66
Hydrogen sulfide	3.3	0.092	0
Methane	77.1	63.901	29.98
Ethane	6.6	3.349	0.55
Propane	3.1	0.96	0.28
Butane	2	0.544	0.21
Pentane	3	0.63	0.25

Table 3. Column specification and operating conditions used in the simulation [10]

Absorber	
Trays	6 stages
Feed stage	6 stage
Pressure	
Stage 1	250 psia
Column pressure drop	1.0 psi
Stripper	
Trays	12 stages
Feed stage	5 stage
Condenser type	Partial vapor
Reboiler type	kettle
Pressure	
Condenser	11.5 psia
Stage 1	15.5 psia
Column pressure drop	2 psia

and enthalpy [12,15]. Table 1 represents the conditions required for using amine property method [12]. If the amine concentration is outside the recommended range, the Chao-Seader method should be used for calculating K-values [12].

A modeling and simulation of the modified conventional amine process by Ball and Veldman and the split-flow amine process were performed using Aspen Plus. The information of the natural gases as the feed streams to the absorber is reported in Table 2 [16]. The specification and operating conditions of the absorber/stripper columns used in the simulation are shown in Table 3 [10]. The removal of CO₂ and H₂S in natural gas from Canada and Tunisia was simulated by using 15 wt% MEA aqueous solution [10]. Natural gas from Western Colorado was treated by 40 wt% DGA aqueous solution [10]. Even though the MEA is able to absorb acid gases at a fast rate, the use of DGA can result in significant energy savings due to the high CO₂ loading capacity when a high CO₂ is contained in a natural gas such as that from Western Colorado [17]. Natural gas containing CO₂ and H₂S enters the absorber on 6th stage as a feed stream. Two lean amines enter the 1st stage and 4th stage of the absorber, respectively, in the split-flow amine process. The rich amine solution containing a large amount of acid gases from the bottom of the absorber column passes through a pressure reducing valve and flash drum to remove hydrocarbons at atmospheric pressure. After being heat exchanged through the heat exchanger, the rich amine solution enters the 5th stage of the stripper column. As shown in Fig. 2, two lean amine solutions recovered from the stripper were cooled in the heat exchangers by the rich amine solutions. A small amount of amines and water leaves overhead from the absorber/stripper column. To maintain the operating concentration of the amine solutions, the amines (MEA and DGA) and the water should be added to make up for a loss [10].

RESULT AND DISCUSSION

The conventional and the split-flow amine processes of the gas sweetening process were simulated using the different amine solutions (MEA and DGA). The performances of the CO₂ and H₂S re-

moval have been studied under the optimized operating conditions, which were obtained until the minimum reboiler duty in the stripper

Table 4. The optimized operating condition (case 1) used in the simulation

The conventional amine process (case 1)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	136.91	85.14
Distillate rate (lbmol/hr)	9613	520
Reflux ratio		1.21
Bottom		
Temperature (°F)	123.3	222.8
Bottom rate (lbmol/hr)	13317	12792
Boilup ratio (Basis: mole)	0.76	0.12
The split-flow amine process (case 1)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	139.2	88
Distillate rate (lbmol/hr)	9592	520
Reflux ratio		1.21
Bottom		
Temperature (°F)	88.19	220.8
Bottom rate (lbmol/hr)	13205	11976
Boilup ratio (Basis: mole)	0.74	0.12

Table 5. The optimized operating condition (case 2) used in the simulation

The conventional amine process (case 2)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	114.99	90.1
Distillate rate (lbmol/hr)	8687.7	1450
Reflux ratio		3
Bottom		
Temperature (°F)	148.5	222.9
Bottom rate (lbmol/hr)	70061	68611
Boilup ratio (Basis: mole)	0.13	0.17
The split-flow amine process (case 2)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	99.99	90.19
Distillate rate (lbmol/hr)	8668	1450
Reflux ratio		3
Bottom		
Temperature (°F)	135.36	218.76
Bottom rate (lbmol/hr)	70022	63631
Boilup ratio (Basis: mole)	0.12	0.18

column was attained. Tables 4-6 show the optimized operating conditions for the absorber and stripper. The temperature of the reboiler

Table 6. The optimized operating condition (case 3) used in the simulation

The conventional amine process (case 3)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	100.06	140.6
Distillate rate (lbmol/hr)	5756.47	4266.8
Reflux ratio		1.3
Bottom		
Temperature (°F)	138.04	224.9
Bottom rate (lbmol/hr)	69037	64770
Boilup ratio (Basis: mole)	0.15	0.29
The split-flow amine process (case 3)		
Stage performance	Absorber	Stripper
Top		
Temperature (°F)	100.22	138.53
Distillate rate (lbmol/hr)	4211	4266
Reflux ratio		1.3
Bottom		
Temperature (°F)	100.3	218.8
Bottom rate (lbmol/hr)	70604	57482
Boilup ratio (Basis: mole)	0.12	0.27

Table 7. Result of the CO₂ and H₂S removal in the absorber

Case 1			
	Mole flow [lbmol/hr]	Input	Output
The conventional amine process	CO ₂	170	2.37E-06
	H ₂ S	330	0.379
The split-flow amine process	CO ₂	170	3.61E-06
	H ₂ S	330	0.0085
Total flow [lbmol/hr]		10000	
Case 2			
	Mole flow [lbmol/hr]	Input	Output
The conventional amine process	CO ₂	1359	2.69E-29
	H ₂ S	9.2	1.95E-21
The split-flow amine process	CO ₂	1359	5.60E-07
	H ₂ S	9.2	5.58E-06
Total flow [lbmol/hr]		10000	
Case 3			
	Mole flow [lbmol/hr]	Input	Output
The conventional amine process	CO ₂	4266	8.87E-16
	H ₂ S	0	0
The split-flow amine process	CO ₂	4266	9.31E-02
	H ₂ S	0	0
Total flow [lbmol/hr]		10000	

of the stripper in the split-flow amine process was 2-6 °F lower than the conventional amine process. The temperature differences are caused by side cut stream from the middle of the stripper in the split-flow amine process. The lower amine concentration in the bottom of the stripper caused by the side cut in the split-flow amine process decreases the boiling point of the amine solution in the reboiler of the stripper.

Table 7 shows the simulation results of the CO₂ and H₂S removal in the absorber. The effluent gases from the absorber contain a very small amount of CO₂ and H₂S in the conventional amine process and the split-flow amine process, respectively. The removal of CO₂ in

Table 8. Solvent rate of the absorber

Case 1				
Solvent rate (MEA) to the absorber [lbmol/hr]				
The conventional amine process			639.7	
The split-flow amine process	Ultra lean amine	595.9	637.3	
	Semi lean amine	41.38		
Case 2				
Solvent rate (MEA) to the absorber [lbmol/hr]				
The conventional amine process			3401	
The split-flow amine process	Ultra lean amine	3122	3395	
	Semi lean amine	273.5		
Case 3				
Solvent rate (DGA) to the absorber [lbmol/hr]				
The conventional amine process			6643	
The split-flow amine process	Ultra lean amine	6573	6612	
	Semi lean amine	38.88		

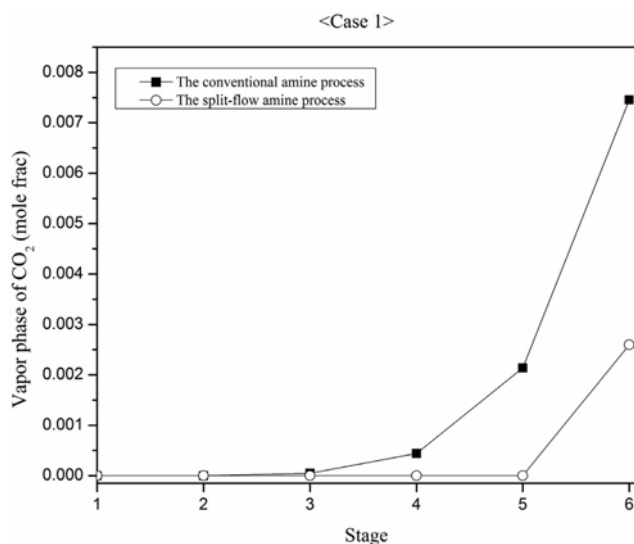


Fig. 3. Profiles of vapor phase mole fraction of CO₂ in the absorber for case 1.

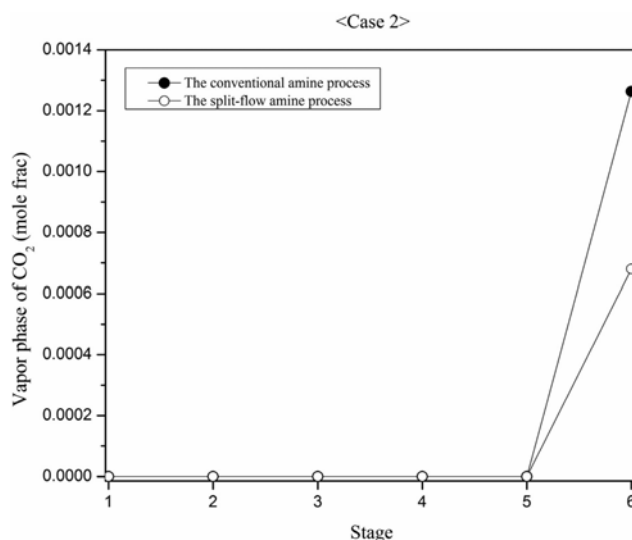


Fig. 4. Profiles of vapor phase mole fraction of CO₂ in the absorber for case 2.

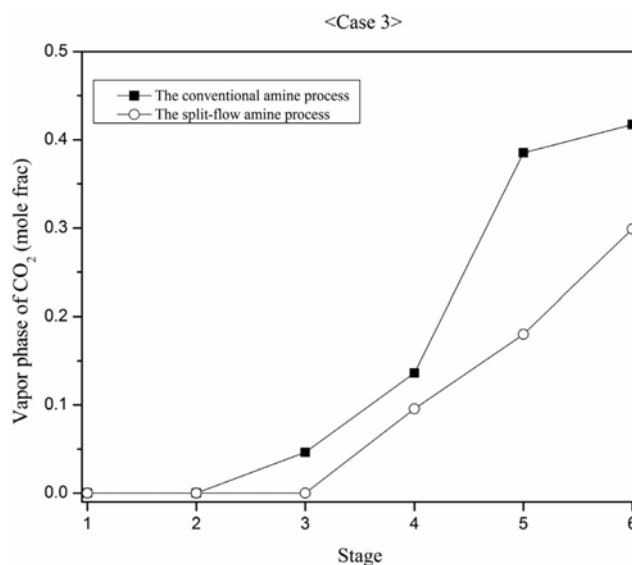


Fig. 5. Profiles of vapor phase mole fraction of CO₂ in the absorber for case 3.

the split-flow amine process was achieved with a slightly lower solvent rate than the conventional amine process as shown in Table 8.

Figs. 3-5 present the profiles of vapor phase mole fraction of CO₂ in the absorber in the conventional amine process and the split-flow amine process, respectively. Even though the CO₂ concentrations in the feed gas are the same, the CO₂ removal in the split-flow amine process was started in the middle stage in the absorber. This is associated with CO₂ absorption in the absorber by semi-lean solvent streams in the split-flow amine process. The semi-lean amine solvents were fed into the 4th stages of the absorber, which absorbed CO₂ gas effectively. However, the outputs of CO₂ in the conventional process are much lower than the split-flow amine process as shown in Table 7. The split-flow process has an advantage in the sense of the energy consumption, not the sweetening performance.

In the split-flow amine process, the reboiler duty of the stripper

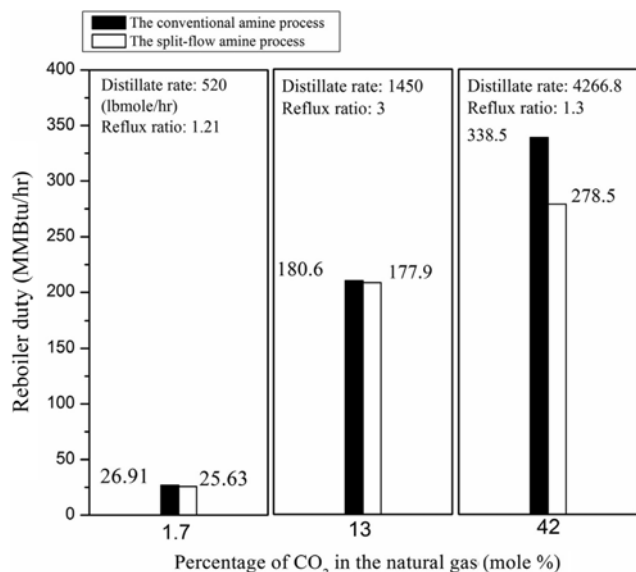


Fig. 6. Reboiler duty of the conventional amine process and the split-flow amine process.

was 5-18% lower than the reboiler duty of the stripper in the conventional amine process. Since a portion of the amine solution is taken from the stripper as a side stream, the amount of the amine solution in the bottom of the stripper in the split-flow amine process is decreased in comparison with the conventional amine process. Fig. 6 provides the results of energy consumption in the conventional amine process and the split-flow amine process, respectively. The reboiler duties of the stripper in the split-flow amine process were 25.63 MMBtu, 177.9 MMBtu and 278.5 MMBtu in cases 1-3, respectively. The reboiler duty of the split-flow amine process was 5-18% lower than the conventional amine process in cases 1, 2 and 3, respectively. When a high concentration of CO₂ was contained in the natural gas, the split-flow amine process showed the removal of CO₂ in the feed gas at a lower reboiler duty than the conventional amine process.

CONCLUSION

The simulation of a natural gas sweetening process with the split-flow amine process was performed with various CO₂ concentrations in a natural gas. The simulated results were compared with the results of the conventional amine process using Aspen Plus. Considering the reactivity and the loading capacity of the amine solvents for CO₂ gas, the split-flow amine process has been simulated with different amine solvents (MEA for the natural gases from Canada (1.7% CO₂) and Tunisia (13.6% CO₂), DGA for natural gas from Western Colorado (42% CO₂)). With the optimized solvent rate at

the same operating conditions, the use of DGA gas treating solvent resulted in significant energy savings over MEA solvent for natural gas containing high CO₂ concentration. According to the simulation results obtained by Aspen Plus, it is confirmed that the split-flow amine process can reduce the energy consumption over the conventional amine process in the stripping sections, which is much greater when the inlet natural gas contains a high concentration of CO₂. Simulation results will be useful for the estimations of lean and semi-lean amine stream rates at the design stage of the split-flow amine process as well as during the optimization procedure.

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