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Simulation of Khangiran gas treating units for various cooling scenarios

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ABSTRACT

Khangiran refinery sour gas feed has been acquired from Mozdouran reservoir in the past twenty eight years (1982–2010). In this period, the reservoir pressure has been dropped from 6600 psia to less than 5000 psia. It is anticipated that the pressure decrease will continue (even more rapidly due to higher production rates and sharper decrease in gas compressibility factor) in the near future. This considerable reduction in reservoir pressure has lead to appreciable increase in sour gas temperature entering the gas treating unit (GTU) due to Joule–Thompson effect. Such temperature increase had various adverse effects on the performance of the GTU process (e.g. higher contactors temperature peaks and larger antifoam consumptions).

A variety of cooling scenarios have been considered in the present article. The required cooling facilities were designed for lowering the temperature of sour gas or lean solvent by air or water. The corresponding entire GTU processes were then simulated using both HYSYS and Aspen software's. The simulation results indicate that the sour gas cooling scenario is not sufficiently effective because of the temperature peak encountered in the absorption towers. Furthermore, although the solvent cooling strategy was more effective from both economical point of view and its impact on the contactor temperature profile, however it may lead to some operational difficulties resulting from heavy hydrocarbons condensation. To avoid such predicaments, both sour gas and lean solvent streams should be cooled simultaneously via air coolers.

1. Introduction

Khangiran sour gas refinery was originally founded in late 1970 decade and commissioned in early 80s. The original plant consisted of three gas treating units (GTU) refining around 30 MMSCMD Mozdouran reservoir sour gas at the peak capacity. The refinery production rate was increased to 50 MMSCMD at the beginning of new millennium by constructing two additional GTUs. All sweetening units were designed using 34% DEA[1] in water as the solvent. The amine was replaced by 45% MDEA[2] solution, recently.[3]

Table 1 illustrates the dry sour gas analysis (prepared by Amine Expert Company on 9th Feb. 2004) for the contactor feed of the Khangiran GTUs (Shahsavand, 2007). For all simulations, the dry gas was initially saturated with water before entering the GTU process. The average molecular weight of C_6^+ was computed about 156, based on RIPI report (Shahsavand, 2007). The sour gas contains about 10% carbon dioxide and hydrogen sulfide which should be separated using suitable amine solution in an absorption tower. The performance of amine contactor column is highly sensitive to the entering temperature of both sour gas and lean solvent. The optimal absorption temperature of H_2S and CO_2 by DEA solution is about 38 °C and 49 °C, respectively (Campbell, 1976). The operational temperature for MDEA solvent is usually about 50 °C (IPS, 2005). Fig. 1 shows that the gas temperature goes through a maximum inside the contactor and then drops to the outlet temperature (Al-Baghli et al., 2001). Evidently, increasing the entering gas and liquid temperatures boost the maximum peak inside the absorber to a value, which is more close to the stripping temperature of hydrogen sulfide (116 °C for H_2S and 149 °C for CO_2).

Mozdouran reservoir pressure has been dropped from 6600 psia to less than 5400 psia during the past twenty eight years (1982–2010). The reservoir gas temperature was always around 250 °F (121 °C). The sour gas cools down to a temperature of 55 °C (in winter) or 65 °C (in summer) when entering the refinery at the pressure of about 1070 psia. Both Joule–Thompson effect and heat transfer to the environment are responsible for this large temperature decrease ($\Delta T \cong 60$ °C). It is shown elsewhere (Shahsavand,

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[1] Di-Ethanol Amine ($C_4H_{10}O_2NH$).
[2] Methyl Di-Ethanol Amine ($C_5H_{13}O_2N$).
[3] GTUs number 1 and 5 were initially tested using MDEA solution in 2005. All five GTUs have been utilizing MDEA as solvent since mid 2006. It is anticipated that some GTUs will return to DEA solution again in near future due to some operational difficulties (mainly high CO_2 concentration of the sweet gas).

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Table 1
Dry sour gas analysis (mole %) of Khangiran refinery GTUs feed.

Components	C_1	C_2	C_3	iC_4	nC_4	iC_5	nC_5
mole %	88.901	0.509	0.058	0.008	0.028	0.01	0.014
Component	C_6^+	H_2S	N_2	CO_2	C_6H_6	C_7H_8	C_8H_{10}
mole %	0.035	3.588	0.368	6.459	0.015	0.005	0.002

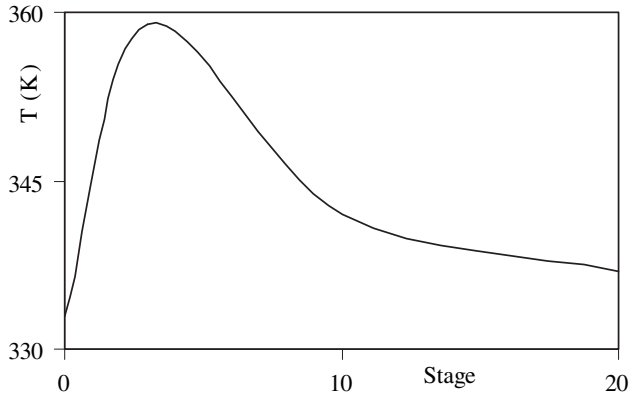

Fig. 1. Temperature profile across the amine absorption column.

2007) that about 80 percent of the overall temperature drop is due to Joule–Thompson effect for a pressure decrease of $\Delta P \cong 4300$ psi. The heat transfer phenomenon plays a less essential role because all the pipelines from well-head to the gathering centers and then to central metering facility (CMF) are buried underground[4]. Fig. 2 shows the average annual sour gas temperature increases from 59 °C in 1984 to about 66 °C in 2004 (Shahsavand, 2007).

The design temperature of Khangiran refinery absorption columns was considered about 52 °C. To prevent heavy hydrocarbons condensation (which leads to larger antifoam consumptions), the lean amine temperature should be sufficiently (≈2 °C) higher than the entering sour gas. Evidently, increasing the feed gas temperature from 52 °C to 66 °C will raise the maximum peak temperature inside the contactor (up to 95 °C which is more close to 116 °C stripping temperature) and hence drastically reduces the performance of absorption column.

The preliminary design of the required cooling facilities (air coolers or heat exchangers) for decreasing the sour gas and lean amine temperatures were performed. The next section reviews the final results of such meticulous calculations. All design calculations for solvent cooling facilities were carried out for 34% DEA solution. Evidently, such cooling facilities can be successfully used to cool down the 45% MDEA solution, because the MDEA rate is always lower than the corresponding DEA solution for the same sour gas capacity.

Furthermore, two powerful chemical engineering simulation programs (HYSYS and Aspen) were employed to simulate the entire GTU process under the actual operating conditions[5], worst case (WC) when both sour gas and lean amine enter the contactor at 70 °C and also for the following cooling scenarios:

- Scenario 1 (S1): cooling down the sour gas to 50 °C while maintaining the lean solvent temperature at the worst case situation (70 °C).

[4] The underground temperature (above 1 m deep) is usually considered constant (≈4 °C) and does not change with season.

[5] As will be presented in Table 5 and Figs. 4–8.

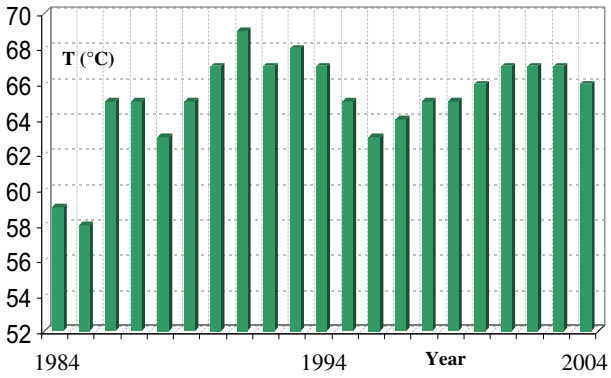

Fig. 2. Sour gas temperature increase during period of 1984–2004.

Table 2
Design specifications of sour gas air coolers for two different capacities.

Parameter	30 MMSCMD	50 MMSCMD
No. of fans in parallel	24	38
No. of fans in series	2	3
Total no. of fans	48	114
Total fans power (HP)	1810	4332
Air intake for each fan (ACFM[a])	455,280	458,900
Total installed cost ($, 2009)	5,000,000	11,500,000

[a] Actual cubic feet per minute.

- Scenario 2 (S2): decreasing the lean solvent temperature to 50 °C while preserving the sour gas temperature at maximum anticipated value of 70 °C.
- Scenario 3 (S3): cooling both the sour gas and lean solvent to 50 °C.

2. Design results of the required cooling facilities

Various scenarios were investigated for cooling sour gas and/or lean solvent solution using air cooler, heat exchanger and a combination of heat exchanger and cooling tower. The final results indicated that using the air coolers is the optimal procedure for cooling both the sour gas (from 70 °C to 50 °C) and the lean amine (from 64 °C to 50 °C) streams. To be on the safe side, the

Table 3
Air coolers design specifications for cooling lean DEA.

Parameter	Each tower	Overall (5 GTU)
No. of fans in parallel	2	20
No. of fans in series	2	2
Total no. of fans	4	40
Total fans power (HP)	4 × 56	40 × 56
Total installed cost ($, 2009)	250,000	2,500,000

Table 4
Heat exchangers design specifications for cooling lean DEA.

Parameter	Each tower	Overall (5 GTU)
Shell ID (m)	2	–
Tube length (m)	6	–
No. of tube passes	2	–
Tube OD (cm)	2.54	–
Pitch (in)	1.25	–
Number of tubes	2400	10 × 2400
Bare surface area (ft^2)	12310	10 × 12310
Total installed cost ($, 2009)	1,200,000	12,000,000

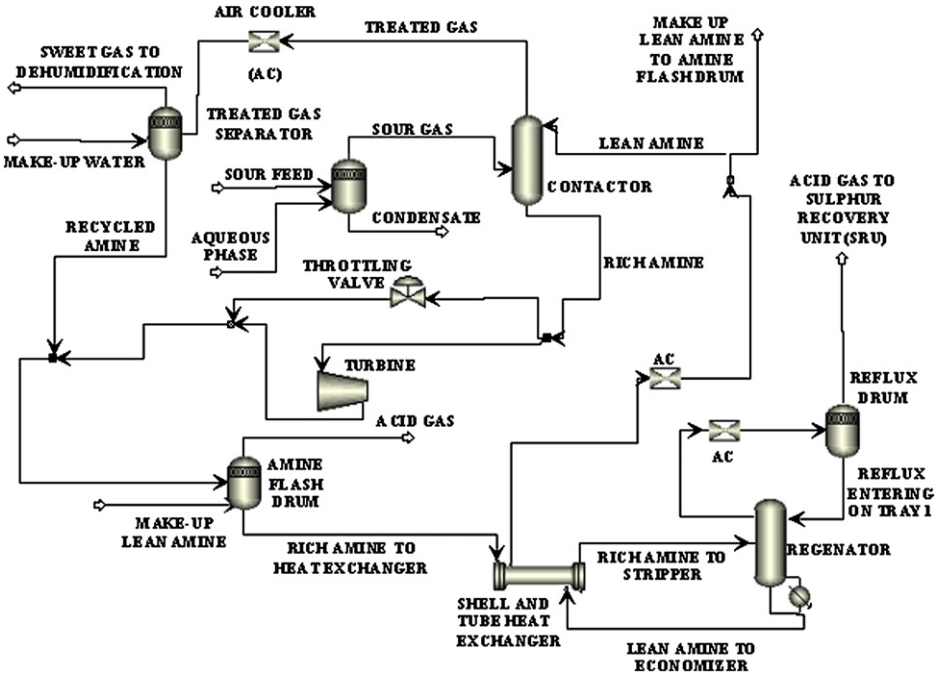

Fig. 3. Simplified schematic diagram of a Khangiran refinery GTU train.

ambient air temperature was considered 40 °C at the worst possible situation. Two different cooling facilities were designed for working capacities of 30 MMSCMD (at the minimum demand) and 50 MMSCMD (at the maximum capacity) sour gas flow rates. Table 2 reviews the final specifications of such thoroughly designed air coolers. The design procedures of Ludwig (1981); Kern (1977); Walas (1988); Mccabe et al. (2001); Peters and Timmerhaus (2002); Chemical engineering news (2009) were used to calculate the desired specifications of the required air coolers. The final results were then checked using Aspen B-JAC software.

In a similar approach, two different cooling procedures were considered to cool down 500 m^3/h lean solvent, entering each contactor. In the first scheme, the lean solvent was cooled from 64 °C to 50 °C using 40 °C ambient air. As the second strategy, the hot solvent was cooled down in a conventional heat exchanger again form 64 °C to 50 °C utilizing 25 °C cooling water. Tables 3 and 4 present the final design specifications of air and water cooled exchangers for all five gas treating units of Khangiran refinery. Each GTU consist of two absorption columns (total number of contactors = 10). As before, the design procedures of references Ludwig (1981); Kern (1977); Walas (1988); Mccabe et al. (2001); Peters and Timmerhaus (2002); Chemical engineering news (2009) were employed to compute the desired specifications of the required air coolers or shell and tube heat exchangers and the final results were checked using Aspen B-JAC software. Evidently, the capital investment for shell and tube heat exchangers scheme is about five times of the corresponding values of air coolers. On the other hand, the operating costs of air coolers are much greater than water heat exchangers (Walas, 1988).[6]

3. Simulation results for the entire GTU process

Fig. 3 illustrates the simplified schematic diagram of a typical gas treating unit (GTU) train of Khangiran refinery. Each GTU consists of two parallel trains with two absorbers and two strippers. The entire refinery has five parallel GTUs with 10 contactors and 10 strippers. Although, both trains of each GTU share the same amine and gas flash drums, however, it is always assumed that each train performs independently and there is no interaction between two adjacent parallel contactors or strippers. Hence, the simulation results of the simplified GTU (depicted in Fig. 1) can be easily extended to the whole sweetening process of Khangiran refinery.

The entire process of Fig. 3 was simulated using both HYSYS and Aspen programs with the data presented in Table 5 for the design operating variables of a typical GTU train. All 5 GTUs of the Khangiran refinery sweetening processes were originally designed using 34% DEA in water as the solvent. The amine was replaced by 45% MDEA solution, at the present condition. For this reason, all simulations were performed for both DEA and MDEA. The computed results at various operating conditions and different cooling scenarios were compared together.

Both HYSYS and Aspen programs have their own amine property packages. Aspen simulation software has the additional capability of employing the so-called "Electrolyte Insert" thermodynamic package which considers only the equilibrium model for predicting

Table 5
Design operating conditions of the each GTU train of Khangiran refinery.

Process Variable	T(°C)	P(psia)	Rate (unit)
Sour gas (to contactor)	52	1050	173000 (SCMH)
Lean amine[a] (to contactor)	53	1050	18654[b] (kg mol/h)
Sweet gas (form air cooler)	38	1050	153000 (SCMH)
Rich amine (from contactor)	77	1050	19400 (kg mol/h)
Rich amine (from amine flash drum)	71	88.9	19854 (kg mol/h)
Acid gas (from amine flash drum)	71	88.9	3091 (SCMH)
Rich amine (to heat exchanger)	71	88.9	19854 (kgmol/h)
Rich amine (from heat exchanger)	99	88.9	19854 (kg mole/hr)
Lean amine (to heat exchanger)	121	27.9	19932.8 (kg mol/h)
Lean amine (from heat exchanger)	93	27.9	19932.8 (kg mol/h)
Acid gas (to air cooler)	120	27.9	2266 (kg mol/h)
Acid gas (form air cooler)	52	27.9	2266 (kg mol/h)
Stripper reflux stream	52	24.7	1436.4 (kg mol/h)

[a] Lean amine CO_2 loading (0.02–0.05 mol CO_2/mol amine).
[b] Lean amine volumetric flow (455 m^3/h for 34 wt% DEA solution).

[6] Shell and tube heat exchangers do not use moving parts.

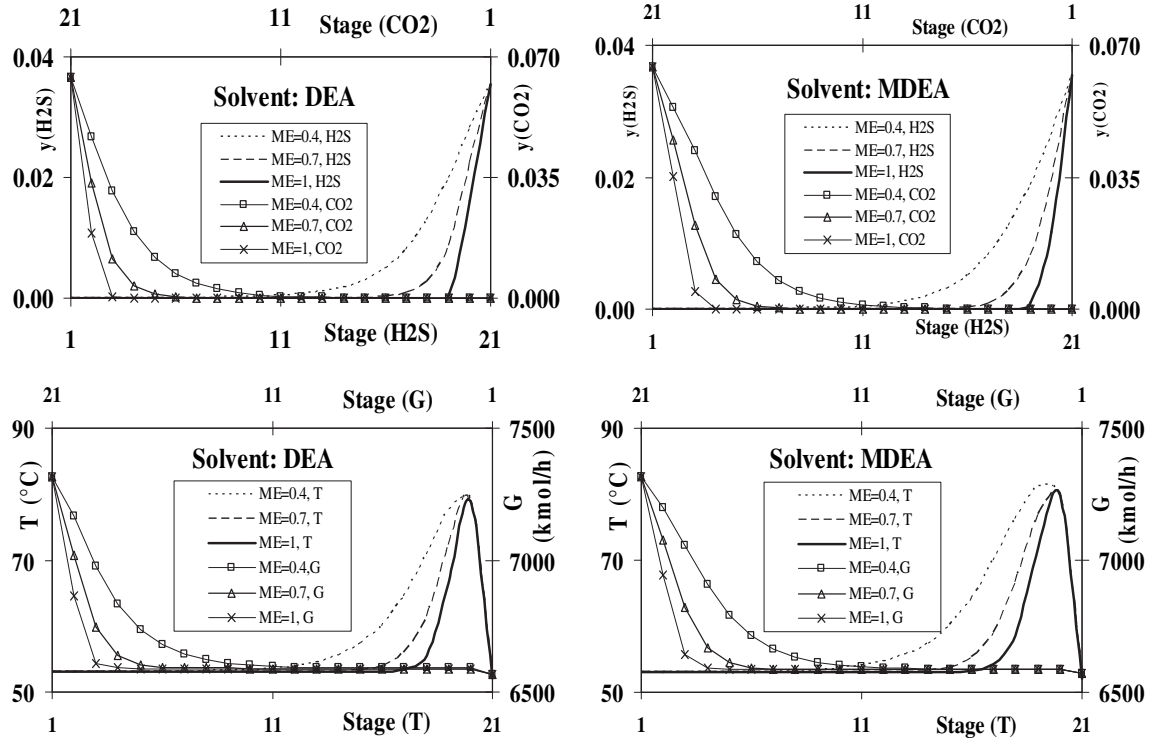

Fig. 4. Temperature, gas flow rate and vapor compositions distributions for contactor using HYSYS at normal operating conditions for various efficiencies and different solvents.

the rate of chemical reactions involved. The original Aspen "amine package" which uses both equilibrium and kinetic models usually does not provide sufficiently accurate predictions for MDEA solutions. For the present case study, the "amine package" was used for both HYSYS and Aspen simulations except otherwise stated.

Figs. 4 and 5 illustrate the predicted distributions for gas phase CO_2 and H_2S mole fractions, temperatures and gas flow rates across the entire contactor using HYSYS and Aspen software's at normal operating conditions as depicted in Table 5 for various assumed Murphree efficiencies with different solvents (DEA 34% and MDEA

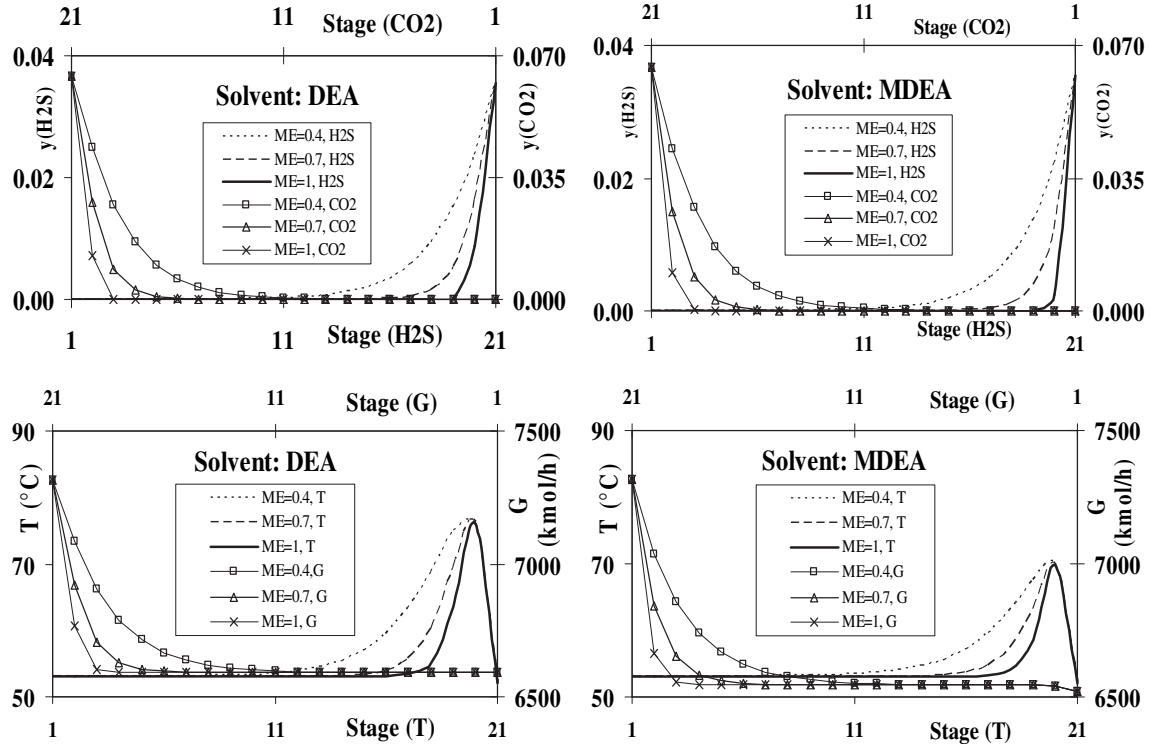

Fig. 5. Temperature, gas flow rate and vapor compositions distributions for contactor using **Aspen** at normal operating conditions for various efficiencies and different solvents.

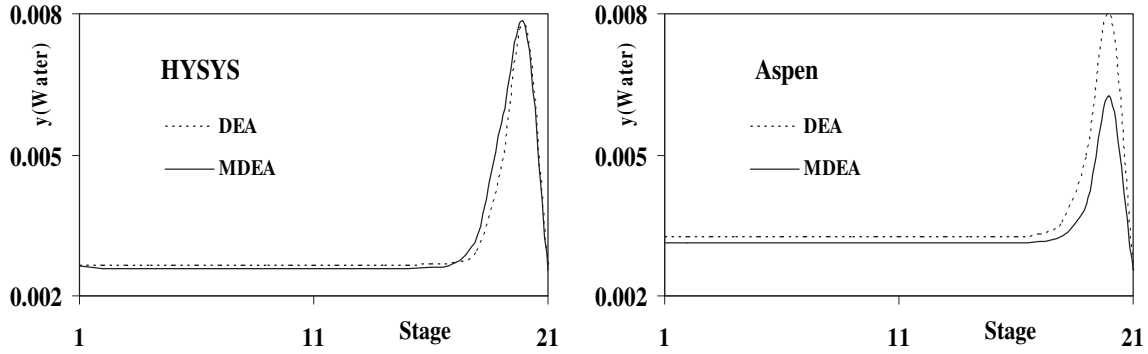

Fig. 6. Gas phase water mole fraction distributions across contactor using HYSYS and **Aspen** at normal operating conditions for various solvents at unit efficiencies.

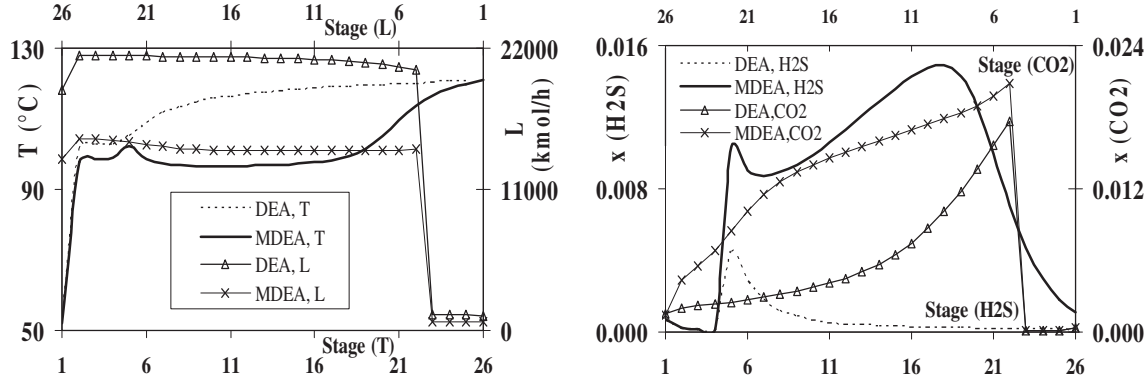

Fig. 7. Temperature, liquid flow rate and compositions distributions for stripper using HYSYS at normal operating conditions with different solvents.

45% in water). The stages were numbered from the top of absorber column and stage 21 represents the entering sour gas conditions. Amine data packages were used in HYSYS simulations for both DEA and MDEA solvents and Aspen simulations of DEA solutions. This data package was not able to produce reliable predictions for MDEA simulations of Aspen program. The Electro-Insert data package was imported to simulate the later case.

Evidently, both software's predict monotonic decrease in predicted gas compositions and the corresponding gas flow rates. The rate of descend increases more rapidly for larger Murphree efficiencies. As mentioned before in Fig. 1, the contactor stage temperatures go through maximum and then drop approximately to the entering solvent temperatures. This phenomenon is due to excessive rate of exothermic absorption occurring in the few bottom trays. The dilute DEA solutions react more rapidly and vigorously with both carbon dioxide and hydrogen sulfide and release lots of heats as gas enters the contactor. The more concentrated MDEA solutions react rapidly and selectively with hydrogen sulfide. Evidently, the temperature pick would be lower for 45% MDEA solution than 34% DEA solution because of less reactivity of MDEA with carbon dioxide and lower water content of the corresponding solvent. This point has been clearly demonstrated in predictions of Aspen software as depicted in Figs. 5 and 6. Higher temperatures lead to larger water vapor pressures and hence higher water mole fractions in gas phase.

Figs. 7 and 8 demonstrate the simulation results for corresponding stripping columns with automatically computed Murphree efficiencies. For better comparison of simulation results, the HYSYS software calculated efficiencies were also used by Aspen program. Furthermore, bottom to feed ratio (or reboiler

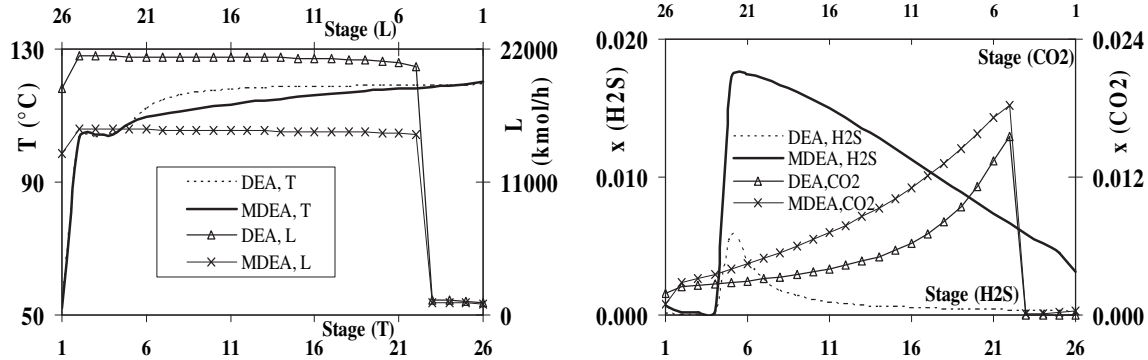

Fig. 8. Temperature, liquid flow rate and compositions distributions for stripper using **Aspen** at normal operating conditions with different solvents.

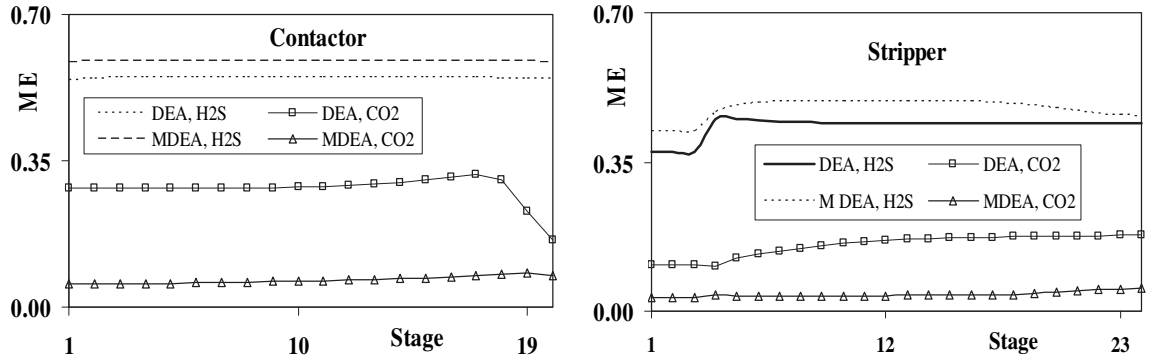

Fig. 9. Variation of automatically computed efficiencies across contactor and stripper.

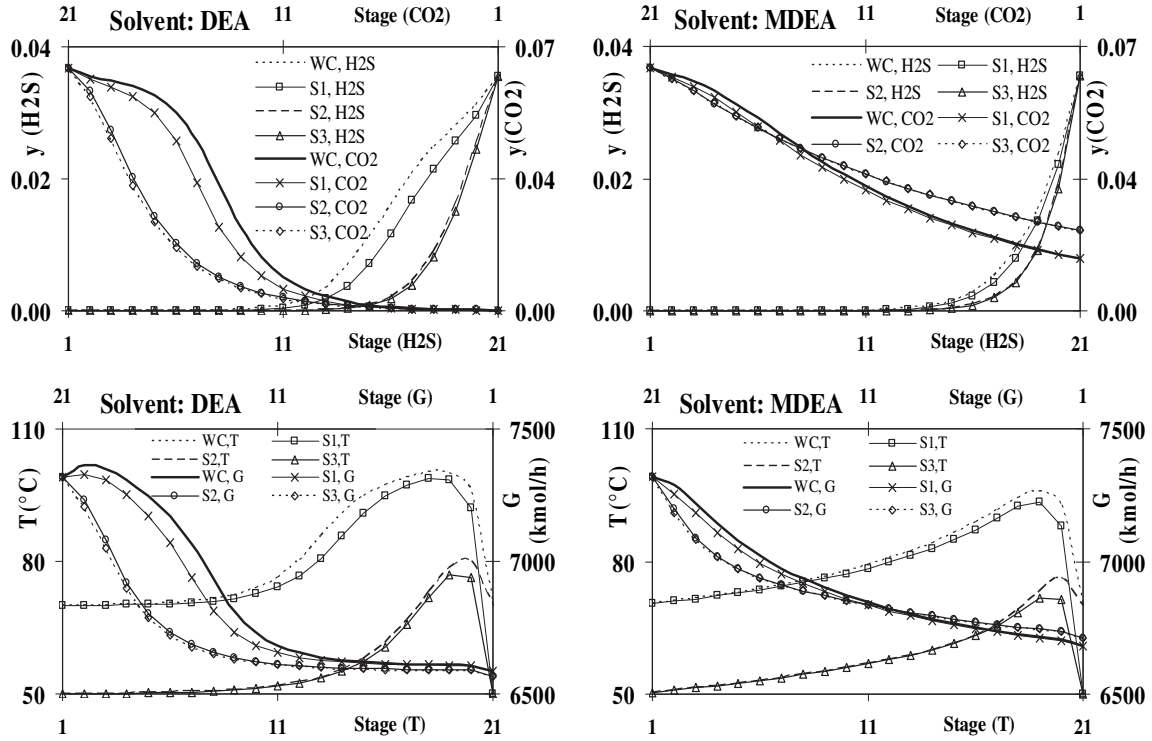

Fig. 10. Temperature, gas flow rate and vapor compositions distributions across the contactor using HYSYS program for various solvents with different cooling scenarios.

temperature) and condenser temperature were considered equal as the design specs of both simulation programs. Liquid flow rate and the corresponding CO_2 and H_2S mole fractions make more sense for stripping process. Therefore, the above parameters are plotted in Figs. 7 and 8 for the entire columns.

It can be seen that both simulation programs (using amine data package) produce similar reasonable results for DEA solutions. HYSYS program predicts unreasonable hydrogen sulfide composition profile across the stripping column for MDEA solution. Evidently, none of mole fractions in liquid phase can increase across the stripping tower. Aspen simulation program also performs inadequately for MDEA solution when using the amine data package. Fortunately, importing Electro-Insert (EI) data package highly improves the estimated composition profiles emphasizing large H2S selectivity of MDEA solvent.

As mentioned before, HYSYS software is capable of computing Murphree efficiencies for both absorber and stripper columns. Fig. 9 shows the variation of such calculated Murphree efficiencies for both absorber and stripping stages. The stripping column condenser and reboiler Murphree efficiencies are practically 100%. These stages were considered for temperature, compositions and flow rate plots of Figs. 7 and 8 and are not included in Fig. 9. The Murphree efficiencies of the stripping column encounter a jump at feed stage because an increase in liquid flow rate produces more turbulence and results higher efficiencies.

Figs. 10 and 11 compare the predictions of HYSYS and Aspen software's for worst case operating conditions (WC[7]) with various cooling scenarios as mentioned in section one, using different solvents (DEA 34% and MDEA 45%). Again, the required Murphree efficiencies for carbon dioxide and hydrogen sulfide were automatically computed by HYSYS software. Other components efficiencies were taken as unity. For comparison purposes, the HYSYS software calculated efficiencies were also used by Aspen program.

Both Aspen and HYSYS simulation programs provide quite similar predictions for temperature, gas flow rate and gas phase

[7] WC: Both sour gas and lean amine entering contactor at 70 °C.

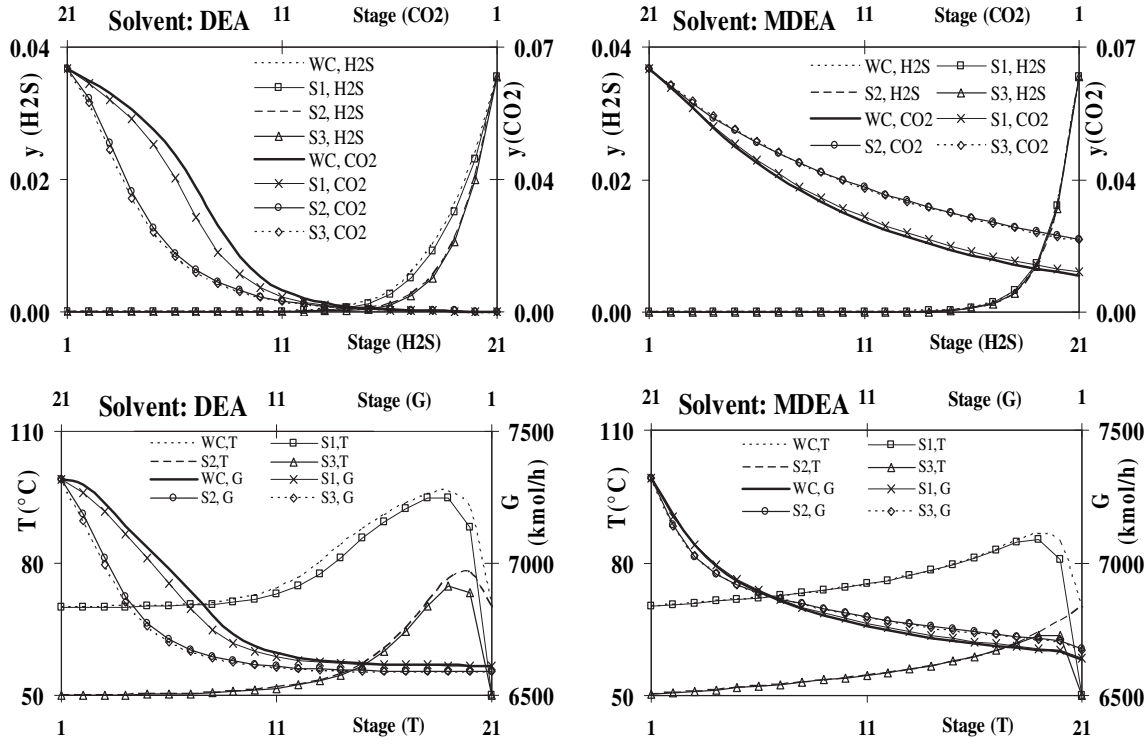

Fig. 11. Temperature, gas flow rate and vapor compositions distributions across the contactor using Aspen program for various solvents with different cooling scenarios.

composition of MDEA and DEA solutions as shown in Figs. 10 and 11. It should be noted that the temperature changes of both gas and liquid feeds have significant effect on absorption capacity of DEA solution. Even at worst case condition with DEA as solvent, both carbon dioxide and hydrogen sulfide concentrations are within the acceptable limits. Decreasing the sour gas feed temperature does not improve the contactor performance because of its low heat capacity. On the other hand, any temperature drop in DEA solvent stream drastically improves the absorber performance as shown in Figs. 10 and 11. Since MDEA solution is very hydrogen sulfide selective, therefore its H2S absorption capacity is always extremely high and can not be affected by moderate changes in sour gas or solvent temperatures. Furthermore, such temperature changes have minor effect on the distribution of carbon dioxide concentration across the entire contactor. In all cases, the mole fraction of carbon in sweet gas stream is higher than the allowable standard values.

4. Conclusion

Mozdouran sour gas temperature has been escalated in the past two decades due to reservoir pressure drop and the corresponding Joule–Thompson effect. Various cooling scenarios have been considered in this article. The required cooling facilities were designed and the contactor absorption performance was thoroughly studied for each cooling scenario using HYSYS and Aspen simulation programs. It was clearly shown that despite of preliminary anticipations, the sour gas cooling scenario has no practical effect on the performance of contactor. The solvent cooling scenario was very effective for DEA solution but it does not affect the absorption tower performance for MDEA solution.

Furthermore, the entering solvent can not be cooled below the entering sour gas temperature, because such cooling leads to excessive foaming in contactor top trays due to considerable considerations of C_3^+ hydrocarbons. To achieve desirable results, both sour gas and lean solvent streams should be cooled simultaneously using previously designed air coolers.

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