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Introducing a Novel Method for Heat Integration of Natural Gas Sweetening Process Based on Reducing TAC, Energy Consumption, CO_2 Emission, and Exergy Loss

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ABSTRACT

The main problem is the high energy consumption in the gas sweetening process, especially in the distillation column, which makes it necessary to optimize the energy consumption of this process. A new method was used in this study for heat integration of natural gas sweetening process. For this purpose, the liquid and vapor streams leave the absorption column entered the upper and lower section of the stripper column, respectively. The feed to the stripper column was preheated by the bottom product the stripper column. Liquid and vapor streams reduce hot and cold utilities of the distillation column, and feed preheating reduces the heating duty of the reboiler. The recovered solvent was then precooled by sweet gas. According to the results, the proposed method reduced energy consumption of the basic process by 67%. Based on the results obtained from calculating the amount of CO_2 emission, the amount of emission of this poisonous gas decreased by approximately 70%. Moreover, proposed process could reduce total annual cost (TAC) of the basic process in the maximum value of 66.76%. This reduction in TAC is due to the reduction of operating cost and capital cost by 66.77 and 66.66% respectively. Also, in comparison with the basic process, exergy loss of the novel process decreased by 66.5%.

1. Introduction

Fossil fuels are the main source of energy worldwide [1]. Combustion of fossil fuels accounts for 81% of the world's commercial energy supply[2]. Since natural gas exists in deep underground reservoirs, natural gas may contain several non-hydrocarbon components, such as carbon dioxide (CO_2) and hydrogen sulfide (H_2S) [3]. These impurities are undesirable compounds that may cause several environmental and technical problems like corrosion, global warming,

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climate change and acid rain [4]. Accordingly, natural gas must be sweetened through various methods such as physical absorption, membrane process which uses selective membranes to separate CO_2 and H_2S from natural gas, offering an energy-efficient solution for treating gas with moderate sour content, hydrate which removes CO_2 and H_2S by forming crystalline structures under low temperatures and high pressures, effectively separating sour gases from natural gas, cryogenic distillation which is a gas sweetening method that separates sour gases, such as CO_2 and H_2S, from natural gas by cooling the mixture to low temperatures, allowing the removal of unwanted acidic components, and absorption using chemical solvents such as amines [5-8].

Gas sweetening with amine solvents is one of the most mature technologies to remove acid gases [1]. Amine gas treatment utilizes primary, secondary, or tertiary amines, such as Monoethanolamine (MEA), Diethanolamine (DEA), and Methyldiethanolamine (MDEA), to react with acidic gases and form stable compounds, enabling their removal from the gas stream. Sour natural gas enters an absorber column where it contacts the amine solution, which absorbs CO_2 and H_2S. This "rich" amine solution, now containing these gases, moves to a regeneration column where CO_2 and H_2S are released, allowing the amine to be recycled in the process. This cyclic method is both efficient and economical, though it requires considerable energy, particularly during regeneration [1]. The choice of amine depends on factors like the concentration of acid gases, operating conditions, and the required purity of the treated gas. For instance, MEA reacts quickly but has a high energy demand for regeneration and is prone to degradation when handling high levels of acid gases [1].

Therefore, to reduce the energy consumption of the gas sweetening process using amines, heat integration methods such as Vapor Recompression (VRC) which is a technique commonly applied in processes requiring low-grade heat. By compressing vapor, usually steam, VRC increases its temperature and pressure, allowing it to be reused within the process, thereby reducing the need for external heat sources and cutting energy costs. It is especially useful in processes such as distillation, where it helps lower fresh steam consumption by recycling vapor back into the system. Effective VRC relies on optimizing the compression to meet process requirements without excessive energy use, Heat pumps which move thermal energy from a cooler area to a warmer one through mechanical work, making them a versatile option for heat integration. They can substitute conventional heating and cooling systems, especially when combined with Pinch Analysis to enhance integration within a plant. By utilizing electricity to recover waste heat, they support lower-emission operations.

Heat pumps are well-suited for applications needing ongoing heating and cooling, such as in the food and chemical industries. Key challenges involve achieving an ideal temperature lift (the difference between input and output temperatures) to maximize efficiency while keeping operational costs manageable and Self-Heat Recuperative Technology (SHRT) which combines a heat pump with internal heat exchangers to capture and recycle energy within a system. It collects low-temperature waste heat from different stages of the process and elevates it to higher temperatures for reuse. This self-sustaining loop greatly minimizes the need for external energy input. SHRT is especially advantageous for processes with variable heat requirements, such as chemical reactions or batch operations. Its main benefit is the reduced demand for hot and cold utilities, making it a highly efficient choice for energy-saving plants, are employed [9]. The advantages of amine based absorption processes are cost effectivness, easily to be retrofited in an existing industrial plant. In addition, these processes can handle large amounts of feed stream [4].

Mohamadi-Baghmolaei et al. [5] studied different combinations of MDEA and DEA to reduce exergy destruction, energy loss and CO_2 emission in gas sweetening process. According to their results, absorption column and stripper

were responsible for 37% of total exergy destruction and 29% of total energy loss, respectively. DEA concentration plays an important role in exergy and energy management of natural gas sweetening process. The optimal amine concentration was 40 wt%, which reduced the total exergy destruction from 6.6 MW to 4.8 MW and the energy loss by 36.7%. CO_2 emission have reached 2893.6 tons per year.

Babaei and Esfahany [10] proposed a new combined RDWC-PV-VP process equipped with a solar heating system for ETBE production, which can provide more than 70% of the heat utility required for feed preheating, RDWC reboiler and PV heaters from solar energy. Based on the results obtained from this study, the capital cost and utility cost were decreased, and this decrease in utility cost, in addition to reducing the cost of energy, was also affected the reduction of CO_2 emission, and the amount of CO_2 emission was decreased to 94%.

Gutierrez et al. [11] used MDEA solution to compare a conventional process of natural gas sweetening with an absorption process using recompressed vapor and a membrane process in terms of compression and pump power, total specific heat, CO_2 removal, CH_4 wastes, and capital cost. The process which used a membrane system consumed lower energy than the other two processes by 77% and 72%, respectively. It also emitted less pollution by 80% and 76%, respectively. However, the overall cost of the membrane system was respectively 12% and 5% more than the capital cost of the other two methods. Moreover, the CH_4 waste in this system was higher than the two other processes.

Ahmadi et al. [12] examined energy optimization in the natural gas sweetening process using amine solvent. The process is widely used due to the high ability of amines to remove acidic compounds from natural gas, but it has a high energy consumption. To reduce this energy consumption, the study investigates two methods for thermal integration of the amine recovery column: Vapor Recompression (VRC) and Bottom Flashing (BF). The results show that the Bottom Flashing method performs better than the Vapor Recompression method. The BF process reduces energy consumption by up to 90% compared to the conventional process, while VRC reduces it by around 88%. Additionally, the BF method reduces the consumption of hot and cold utilities by 33% compared to VRC, which leads to a 96% reduction in the total annual cost (TAC) of the process. Therefore, the article concludes that the Bottom Flashing method is the best choice for thermal integration in the natural gas sweetening process due to its superior efficiency in reducing both energy consumption and costs.

He et al. [13] outlined the use of water-selective and CO_2-selective membranes in a multi-stage process. Pebax® membranes were employed for dehydration, and carbon membranes were used for CO_2 removal. Simulation results identified the optimal operating conditions for minimizing natural gas processing costs while achieving a high methane purity of over 97.5% and a dew point of below -40°C. The economic feasibility of the membrane process was evaluated, showing a potential minimum cost of $\$2.55\times10^{-3}$ per cubic meter of sweet natural gas, depending on various factors like permeate pressure and recycling of captured CO_2.

Ababneh et al. [14] developed and simulated a Separation Vessel (SV) unit using a custom Aspen Plus model. This unit works by throttling the feed gas, which cools and solidifies CO_2 and H_2S, allowing high-purity methane to remain in the vapor phase. The solid acid gases are melted and collected separately. Various feed compositions were tested to optimize the process, and the simulation results showed that the SV unit could produce methane with over 99% purity while removing over 90% of CO_2 and H_2S. The study highlighted that the SV unit uses only 27% of the energy compared to a traditional amine sweetening unit, offering improved energy efficiency, fewer equipment requirements, and avoiding the use of corrosive chemicals. Additionally, unlike the amine process, the SV method does not cause

water contamination. Although the amine process achieved slightly higher acid gas removal rates, the SV unit proved to be more energy-efficient and had lower operational costs.

Berchiche et al. [15] aimed to improve the energy efficiency of a natural gas sweetening unit by integrating a bi-level heat recovery process. They focused on using waste heat from upstream natural gas production to reduce energy consumption in the amine-based sweetening unit. The approach included direct heat integration (DHI) of high-temperature waste heat and the use of an Organic Rankine Cycle (ORC) to recover low-temperature waste heat. The study utilized data from three natural gas production facilities and simulated various scenarios using Aspen HYSYS software. It was shown that direct heat integration reduced the heating requirements of the amine unit by up to 100%, significantly cutting energy demand. ORC integration further provided net electricity production, covering between 30% and 190% of the facility's power needs, depending on the conditions. One key result is that propane (R290) was found to be the most efficient working fluid for the ORC, although higher carbon fluids provided a better balance between energy output and operating pressure. The study concludes that the integration of DHI and ORC in natural gas processing can significantly improve energy efficiency and reduce operating costs.

Song et al. [1] presented a novel approach to natural gas purification using heat pump-assisted MEA absorption to reduce energy consumption. The conventional MEA process consumes significant energy, especially during solvent regeneration, but the proposed method recovers waste heat and pressure through vapor recompression, drastically cutting energy costs. Simulations showed that the new system reduced energy consumption by 78.9%, from 41.2 MW to 7.2 MW, and improved methane purity in the treated gas. The study concludes that this advanced process is more energy-efficient and cost-effective, making it a better option for natural gas purification.

Natural gas sweetening process is an energy intensive process so, very high energy consumption is the main problem in this process. In this article a new method of heat integration was proposed, the proposed method allows us to use an accessible solvent while reducing energy consumption and emissions of various pollutants without the need for equipment such as a compressor. In this method, a liquid stream and a vapor stream leave the absorption column and enter the top and bottom of the stripper column, respectively. In this way, part of the heat required in the stripper column is provided by these two strseams and reduces the heat duties of the reboiler and condenser. The stripper column feed is then heated by the bottom product and the solvent recycled to the absorption column is precooled by sweet gas. The preponderance of this method in the natural gas sweetening process, compared to other heat integration techniques such as VRC, heat pump and SHRT, is that it does not rely on the compressor, which actuates to a reduction in process costs. Hence, the proposed method is expected to outperform other heat integration strategies. In addition, CO_2 emission, exergy loss and TAC were calculated for the basic process and the proposed process to further investigate the effects of this novel method on the basic process.

2. Case study

This study focused on the natural gas sweetening process through monoethanolamine solvent. Table 1 indicates the feed/solvent qualifications. It is worth mentioning that feed/solvent qualifications are identical in both processes.

Table 1. Feed/Solvent conditions [16]

conditions	Feed	Solvent
Temperature [°C]	30	30
Pressure [kPa]	3000	5000
Mass flow rate [kg/h]	75600	100000

3. Methods

In this research, the Aspen HYSYS V.11 software was utilized for process simulation, and the Modified HYSIM Inside-Out solver was used to run the solver to enhance the precision. In addition, the Acid Gas-Chemical Solvents equation was applied for the fluid package [16].

3.1. Process simulation

This paper evaluated the natural gas sweetening by monoethanolamine solvent with and without heat integration.

3.1.1. The process of natural gas sweetening by monoethanolamine solvent without heat integration

Fig. 1 illustrates the simulation of the natural gas sweetening process using a monoethanolamine solvent in Aspen HYSYS software. In this setup, the feed and solvent enter the absorption column from the bottom and top, respectively. This column, with 20 trays, operates at pressures of 4800 kPa at the top and 4900 kPa at the bottom. The sweet gas exits as the top product, while the bottom product, a mixture of solvent and sour gas, exits the first column, reaches 100°C, and enters the fifth tray of the next column. The second column also has 20 trays, with pressures of 101.3 kPa at the top and 130 kPa at the bottom. In this column, the top product is H_2S and CO_2, while the bottom product, containing monoethanolamine and water, is restored to the solvent's initial pressure and temperature after passing through a pump and cooler, then returns to the beginning of the process [16].

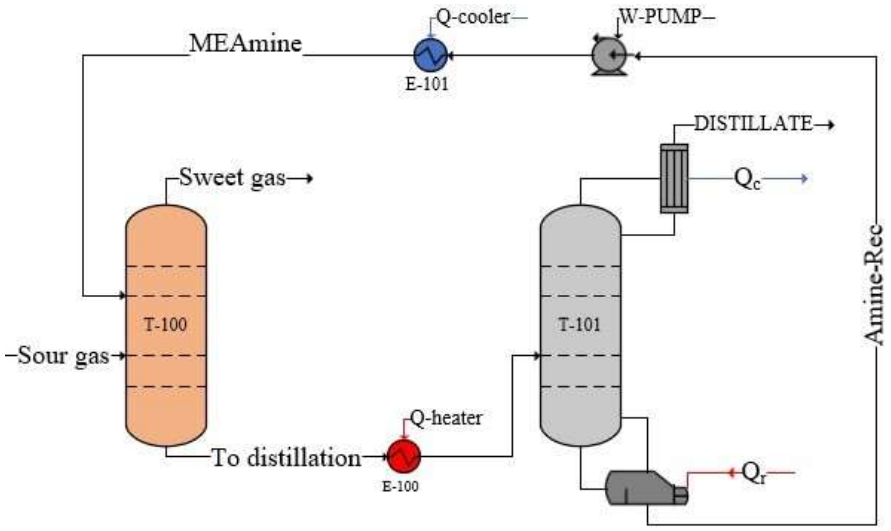


Fig. 1. Process of natural gas sweetening using monoethanolamine solvent [16]

3.1.2. The process of natural gas sweetening by monoethanolamine solvent with heat integration

Fig. 2 provides a detailed view of the natural gas sweetening process using monoethanolamine (MEA) solvent with integrated heat recovery, simulated in Aspen HYSYS software. In this process, natural gas enters at the bottom of the absorption column, while MEA solvent enters from the top. The 20-tray column operates with a pressure of 4900 kPa at the bottom and 4800 kPa at the top. The top product of this column is sweet gas, while the bottom product, a mixture of MEA and sour gas, is preheated by the bottom output of a second column. After preheating, this mixture is further heated to 100 °C and then fed into the fifth tray of the second column. To optimize the heat duties of the reboiler and condenser, two streams exit the first column-one in liquid form and the other as steam-entering trays 2

and 19 of the stripper column, respectively. Similar to the first column, the second column contains 20 trays, with pressure settings of 101 kPa at the top and 130 kPa at the bottom. The top product of this second column consists of H_2S and CO_2 gases, while the bottom product, containing MEA and water, is initially cooled by sweet gas from the first column, reaching 75 °C at 5000 kPa. This stream is then further cooled to 30°C by a cooler before being cycled back to the beginning of the process [16].

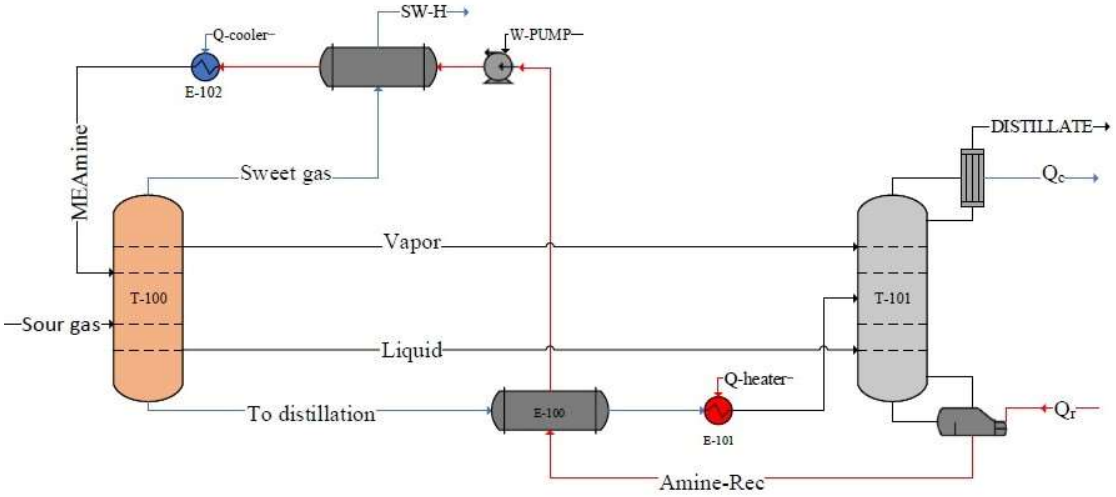


Fig. 2. Process of natural gas sweetening using monoethanolamine solvent with heat integration [16]

3.2. Environmental evaluation

CO_2 emission is a key parameter in the environmental assessment of chemical processes. The increasing emission of CO_2 and the subsequent warming of the earth due to the greenhouse effect have raised global concerns. Therefore, reducing CO_2 emission is considered an effective factor in the design of chemical processes. CO_2 emission in distillation processes are measured using Eq. 1 [10, 17].

$$CO_{2|emission} = \frac{Q_{fuel}}{NHV} \times \frac{C\%}{100} \times \alpha \quad (1)$$

where α is the molar mass ratio of CO_2 and equals to 3.67, NHV is the net heating value for heavy fuel oil and equals to 39771 kJ/kg, C% is the carbon content, which is 86.5%, and Q_{fuel} is the duty of fuel and calculated from Eq. 2 [6].

$$Q_{fuel} = \frac{Q_{Proc}}{\lambda_{Proc}}(h_{Proc} - 419) \times \frac{(T_{FTB} - T_0)}{(T_{FTB} - T_{Stack})} \quad (2)$$

where Q_{Proc} (kJ/h) is the heat duty of the reboiler, λ_{Proc} (kJ/kg) is the latent heat of low pressure steam (LPS) or medium pressure steam (MPS), h_{Proc} (kJ/kg) is the enthalpy of LPS or MPS, T_0 is the reference temperature (T_0=25°C), T_{FTB} is the flame temperature (T_{FTB}=1800°C), and T_{Stack} is the stack temperature (T_{Stack}=160°C). CO_2 emission for electrical power is equal to 51.1×10^{-6} $kgCO_2/kJ$.

3.3. TAC calculation

In this research, in order to compare the economic cost of the two simulated processes, the parameter of Total Annual Cost (TAC) has been used. The Total Annual Cost has been calculated through Eq. 3 [17, 18].

$$TAC = \text{Operating Cost} + \left(\frac{\text{Capital Cost}}{\text{Payback Period}}\right) \quad (3)$$

This paper considers a payback period of 5 years and a man-hour value of 2480 hours [16]. Further information regarding the economic analysis can be found in Appendix B [18].

3.4. Exergy

Unlike energy analysis, exergy accounts for both quantity and quality of various energy forms, so exergy has been recommended as a measure of system performance [19]. Exergy is the maximum available work obtained by bringing a system to equilibrium with its environment [19]. Exergy of a material stream which flows through a system can be divided into chemical exergy and physical exergy. Total exergy (Ex_{total}) of a material stream, physical exergy (Ex_{ph}), and chemical exergy (Ex_{ch} for ideal gas and ideal mixture) are measured using Eqs. 4 to 6 [19].

$$Ex_{total} = Ex_{ph} + Ex_{ch} \tag{4}$$

$$Ex_{ph} = (H - H_0) - T_0(S - S_0) \tag{5}$$

$$Ex_{ch} = RT_0 \sum_i n_i \ln x_i \tag{6}$$

Except the exergy that enters or leaves the system through a material stream, the heat that enters or leaves the system can also have exergy and calculated from Eq. 7 [18, 19]

$$E_Q = Q \times \left(1 - {}^{T_0}/_{T}\right) \tag{7}$$

Exergy loss can be calculated from Eq. 8[19].

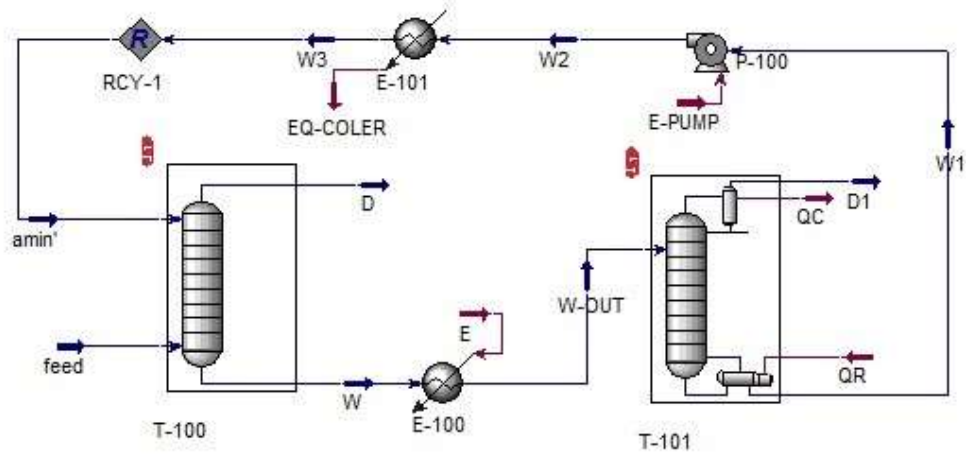
$$Ex_{loss} = \sum Ex_{in} - \sum Ex_{out} - \Delta Ex_{system} \tag{9}$$

In steady state condition, ΔEx_{system} is equal to zero.

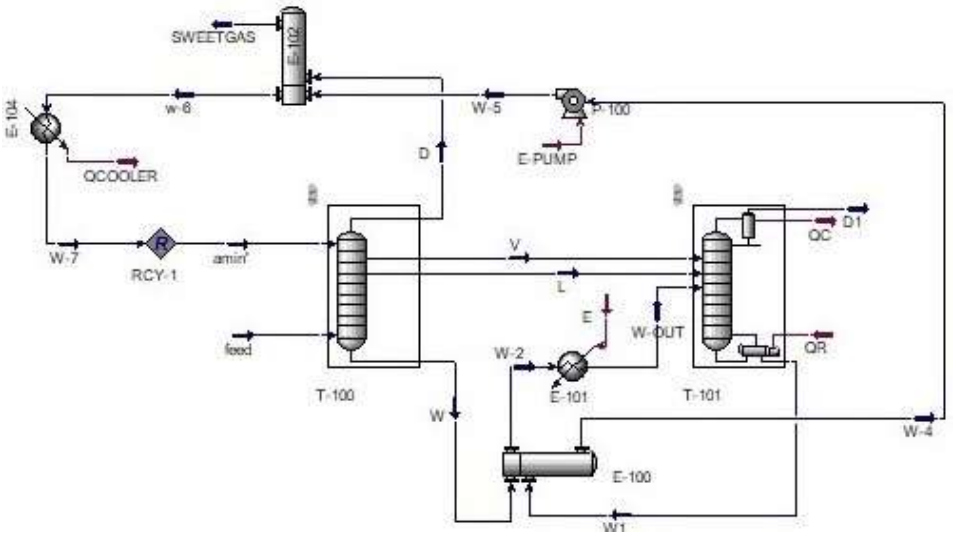
4. Results and discussion

As shown in Fig. 3 two side streams from the absorption column are utilized to supply the heat duty required for the distillation column. A liquid stream exits the bottom of the absorption column and enters the top of the distillation column, while a vapor stream leaves the top of the absorption column and enters the bottom of the distillation column. The feed entering the distillation column is preheated by the bottom product, thereby reducing the heat duty required by the reboiler. The bottom product of the distillation column, which contains amine, is recycled back to the absorption column after being precooled by the sweet gas stream. By introducing steam and liquid into the second column, the amount of steam generated at the bottom and the liquid produced at the top are reduced, leading to a decrease in utility consumption for this column.

Fig. 4 compares energy consumption, CO_2 emission, exergy loss, and total annual cost (TAC) in the natural gas sweetening process using a monoethanolamine (MEA) solvent, both with heat integration (P2) and without it (P1). As shown, the gas sweetening process with heat integration consumes less energy than the process without it. To further investigate the basic and proposed processes, exergy loss and CO_2 emission were calculated. Fig. 4 illustrates that due to reduced energy consumption, CO_2 emission also decrease significantly, from 30,408,648.15 kg/h in the basic process to 10,099,327.51 kg/h in the proposed process. The exergy loss results follow a similar trend, showing a reduction from 11,827,041.31 kW in the basic process to 3,955,668.3 kW in the proposed process. This reduction in exergy loss is primarily attributed to the decreased hot and cold utility requirements in the proposed process, leading to lower exergy of heat. Table 2 shows exergy of heat of the basic and the proposed process.



(a)



(b)

Fig. 3. Simulation the processes in Aspen HYSYS software (a) Without heat integration, (b) With heat integration

Table 2. Exergy of Heat of the Processes

	Exergy of heat [kW]	
Energy stream	**Without heat integration**	**With heat integration**
Q-Heater	517.34	157.50
Q-Cooler	1132.02	391.48
W-Pump	188.2	185.2
Q-Condencer	7876470.30	2589889.89
Q-Reboiler	19703917.18	6545586.66

The reduction in energy consumption results in a 66.77% decrease in the operating cost of the proposed process. The requirements for vapor and liquid reflux in the column are also reduced, leading to a decrease in column diameter and the heat transfer area of reboilers, condensers. Consequently, the capital cost of the proposed process is reduced by approximately 66.66%.

Table 3 indicates both processes's operating and capital cost.

Table 3. Operating and Capital cost of the Processes

Process	**Operatng cost [$/year]**	**Capital cost [$]**
Basic	10.7×10^9	5.1×10^9
Proposed	3.5×10^9	1.7×10^9

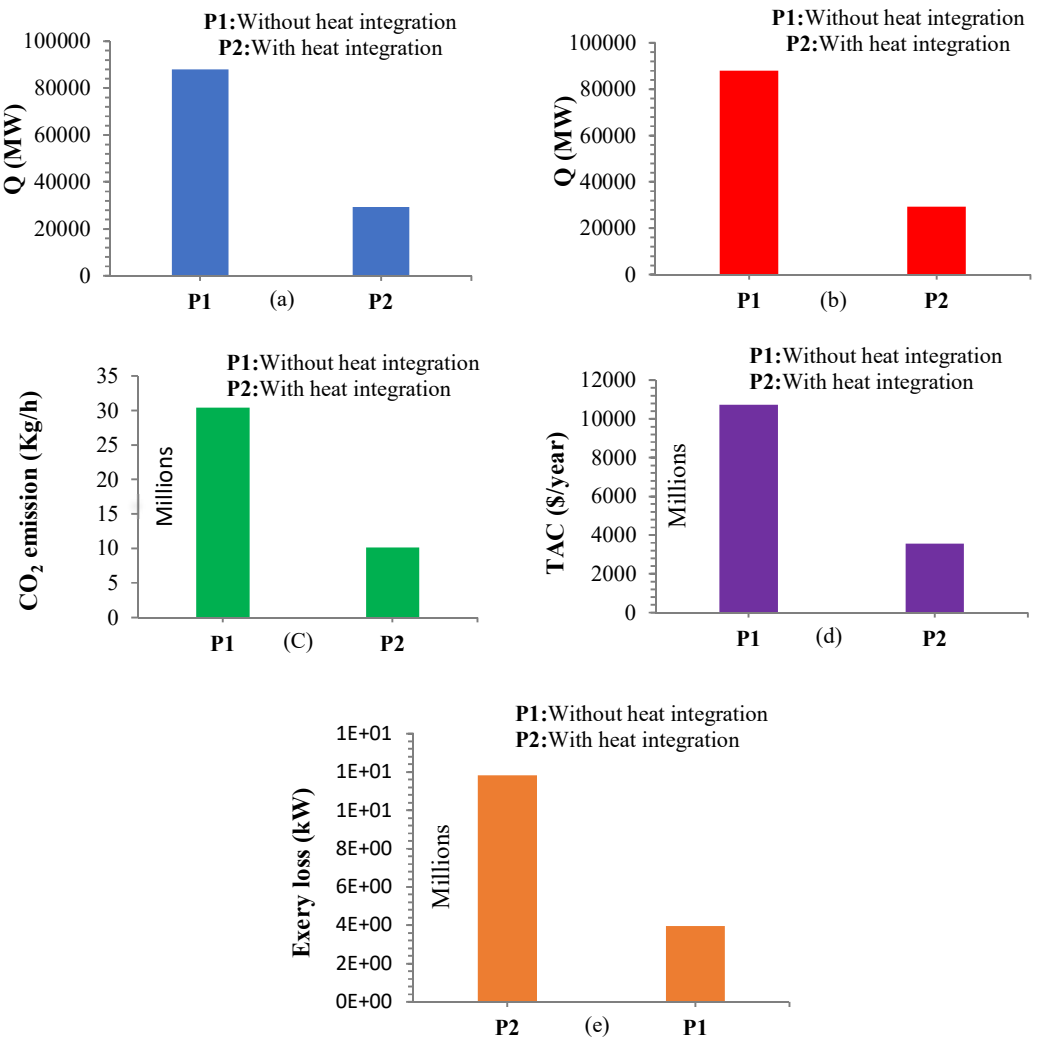


Fig. 4. Comparing the processes (a) Energy Consumed by Cold Utility, (b) Energy Consumed by Hot Utility, (c) CO_2 emission, (d) TAC, (e) Exery loss

5. Conclusion

The main problem is high energy consumption in the gas sweetening process so that large part of energy is consumed in the distillation column for amine recovery. Accordingly, energy optimization of this process seems necessary given the global energy crisis in recent years. A novel method was used in this study for heat integration of the natural gas sweetening process. Two side streams from absorption column were used to supply the heat duty of the distillation column. A liquid column leaves the bottom of the absorption column and enters the top of the distillation column. A vapor stream leaves the top of the absorption column and enters the bottom of the distillation column. The presence of liquid and vapor in the distillation column reduces the heat duties of the reboiler and condenser. The feed entering the distillation column is preheated by its bottom product, reducing the heat duty of the reboiler. The bottom product of the distillation column containing amine is recycled to the absorption column after being precooled by the sweet gas stream, significantly reducing the hot and cold utilities of the process. The most important outcomes are listed below.

1. By adding steam and liquid to the second column, the amount of steam produced at the bottom and the liquid produced at the top of the column are reduced, leading to a decrease in utility consumption for this column. Consequently, the proposed method reduced the total energy consumption by 67%.
2. Due to the reduction in hot and cold utility consumption in the proposed process compared to the basic process, the primary exergy changes in the proposed process, relative to the basic process, are related to exergy of heat. Based on the obtained results, exergy loss in the proposed process is reduced by 66.5%.
3. Due to the reduction in energy consumption, the operating cost of the proposed process dropped by 66.77%. Additionally, the required amounts of reflux vapor and liquid in the column decreased. This reduction led to a decrease in the column diameter and the heat transfer areas of the reboilers, condensers, and coolers. Consequently, the capital cost of the proposed process was reduced by approximately 66.66%. As a result, the proposed method reduced the TAC by 66.7%.

Nomenclature

Abbreviation			
BF	Bottom Flashing	PV	PerVaporation
DEA	Di Ethanol Amine	RDWC	Reactive Dividing Wall Column
DHI	Direct Heat Integration	SV	Separation Vessel
MDEA	Methyl Di Ethanol Amine	SHRT	Self-Heat Recuperation Technology
MEA	Mono Ethanol Amine	TAC	Total Annual Cost
ORC	Organic Rankine Cycle	VRC	Vapor Recompression
		VP	Vapor Permeation

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Appendix A. Energy and Mass balance

Table A.1. Material Streams of the Process Without heat integration

Stream	Pressure [kPa]	Temperature [C]	Mass Flow [kg/h]
feed	3000	30	75600
amin'	5000	30	100000
D	4800	30.1	67062.4
W	4900	71.6	108537.5
W-OUT	4800	100	108537.5
D1	101.3	54.1	8538.6
W1	130	110.7	99998.8
W2	5000	111.5	99998.8
W3	5000	30	99998.8

Table A.2. Material Streams of the Process With heat integration

Stream	Pressure [kPa]	Temperature [C]	Mass Flow [kg/h]
feed	3000	30	75600
amin'	5000	30	100000
D	4800	30.1	67026.9
W	4900	71.6	108489.9
W-OUT	4800	100	108489.9
V	4805.2	30.1	36.7
L	4894.7	70.2	46.3
D1	101.3	53.8	8570.1
W1	130	110.7	100002.9
W-2	4900	92.6	108489.9
W-4	130	90.9	100002.9
W-5	5000	91.6	100002.9
w-6	5000	75	100002.9
SWEETGAS	4800	69.5	67026.9
W-7	5000	30	100002.9

Table A.3. Energy Streams of the Processes

	Heat Flow [kJ/h]	
Equipment	**Without heat integration**	**With heat integration**
Heater	1.099×10^7	2.932×10^6
Cooler	3.149×10^7	1.688×10^7
Pump	6.775×10^5	6.669×10^5
Reboiler	3.169×10^{11}	1.053×10^{11}
Condencer	3.169×10^{11}	1.053×10^{11}

Appendix B. Total annual cost calculation

In this section, the relationships related to cost calculations are presented[10, 18].

B.1. Cost of column

$$\text{Column cost[\$]} = \frac{M\&S}{280} \times 101.9 \times D_c^{1.066} \times L_c^{0.802} \times (2.18+3.67) \quad \text{(B.1)}$$

where Dc [ft] is the diameter of the column and Lc [ft] is the height of column.

B.2. Cost of tray

$$\text{Tray cost[\$]} = \frac{M\&S}{280} \times 4.7 \times D_c^{1.55} \times L_c \times (1+1.8+1.7) \quad \text{(B.2)}$$

where L_c [ft] is the height of column.

B.3. Cost of heat exchanger

$$\text{Heat exchanger[\$]} = \frac{M\&S}{280} \times 101.3 \times A^{0.65} \times (2.29+F_c) \quad \text{(B.3)}$$

where A is heat transfer area and FC is 5.0625 for the reboiler and 3.75 for the condenser.

B.4. Cost of feed pump

$$\text{Pump cost[\$]} = 26700 \times \left(\frac{24 \times F_F \times 3600}{50000}\right)^{0.53} \quad \text{(B.4)}$$

where FF is the total feed rate to the pump [m3/s].

B.5. Utility costs

Cooling water = 0.354 ($/GJ)

Low pressure steam (LP) (5 barg, 160°C) = 13.28 ($/GJ)

High pressure steam (HP) (41 barg, 254°C) = 17.70 ($/GJ)