ALKANOLAMINE SOLUTION PERFORMANCE IN CARBON-DIOXIDE CHEMICAL ABSORPTION PROCESSES: INFLUENCE OF LEAN ALKANOLAMINE DEGREE

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Abstract: Eliminating acid gases, the most prevalent method in natural gas sweetening, is through the application of alkanolamine reactions. Numerous alkanolamine approaches every one of these have been extensively studied processes can be used to perform acid gas removal depending on the both sweet gas and sour gas parameters., Never the less one of the remaining lingering concerns is the financial implications on the process. In the present investigation, (DEA+MDEA) And mixed DGA, DEA Procedures are used to natural gas sweetening that is generated in one of the natural gas fields with significantsH2S/CO2 Ratio. A total of seven cases are created for each procedure in order to examined how the coolers operational setting affect the process performance. To ensure that lean amine hitting the absorber is a specified temperature ,the cooler function is changed for each cases .ASPEN HYSYS is used to simulate each scenario ,and ASPEN Fiscal analysis is used to access its economic viability .According to the finding of this investigation ,as the temperature of lean-amine rises so do the appropriate solution flow rate .Nevertheless ,performing both the DGA and DEA operations at greater quantities of lean amine temperature resulted in lower process capital cost and cooling duty .In the scenario of the DEA +MDEA the proceeding running at decrease lean amine temperature also led to reduced hydrocarbon collection.

I. INTRODUCTION

The methods that include the use of alkanolamine solutions for the removal of acid gases from natural gas are the most popular techniques that are employed for this purpose. Each of the well-developed alkanolamine procedures is suited for sweetening the natural gas that meets specified standards for sour and sweet gas [1-9]. Nevertheless, one of the most significant problems is the high expense that is connected with carrying out these procedures [10-12]. The expenses that are involved with these procedures have been the subject of a great deal of research and investigation.

Polasek et al and Bae et al studied alternate flow methods for natural gas sweetening [11], Warudkar et al studied the effects of stripper operating parameters [10], Cousins et al studied modifications on the process flow sheet [14], Sohbi et al and Fouad et al studied the effects of using mixed alkanolamines [6, 7], and Kazemi et al and Ghanbarabad Mores et al applied optimization techniques [12, 17, 18] Both Freeman et al. [8] and Banat et al. [19] recommended employing concentrated piperazine combinations as a strategy for lowering costs and energy needs associated with the sweetening procedures, respectively.
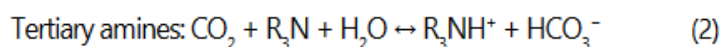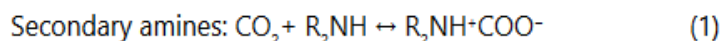
Multiple techniques have the potential to be applicable for the sweetening of natural gas to specified standards. When faced with circumstances such as these, one of the queries that naturally arises is, "Which procedure is the most cost-effective technique to utilize for sweetening the natural gas given these specifications?" Additionally, the temperature of the lean solution as it enters the absorber is an essential factor that plays a role in determining the overall cost of the sweetening process. Changing the temperature of the lean amine might have an effect on the flow rate of the solution that is required in order to achieve the desired characteristics of the sweet natural gas. This has a significant influence on the expenses involved with the operations of natural gas sweetening. On the other hand, the selection of a certain temperature for the lean amine will have an effect on the amount of force that must be supplied to the cooler. Another question that has to be answered is at what temperature the lean solution should be introduced into the absorber in order to get the highest possible level of economic performance. In this work, I attempted to find answers to these two issues for the scenario of natural gas that is generated in a gas field that has a high CO2/H2S ratio, has relatively high CO2 contents but low H2S contents, and has low pressure. Within the scope of this investigation, the consequences of the cooler's operating settings are looked at. The DEA process, the DGA process, and the combined process (MDEA+DEA) are all developed for sweetening the natural gas that is generated in a gas field that has a high CO2/H2S ratio. In order to explore the influence that the cooler's operational settings have on the performance of the operations, seven different scenarios have been devised. Aspen HYSYS is utilized to do the simulations, and Aspen economic assessment is implemented to carry out the economic analysis. Following this, the results of the simulation and the economic evaluation are analyzed in order to identify the optimal operating settings for the cooler that is used in the process. Although there have been some studies on the impact of lean amine parameters on the performance of the sweetening processes [20], I was unable to discover any thorough research that studied the indicated target parameters for the various sweetening procedures.

II.   GAS FEED SPECIFICATIONS

It is clear from the data that has been presented that the natural gas produced in this gas field has a high CO2/H2S ratio, high CO2 content, low H2S content, and low pressure. As a result, it is anticipated that the findings of this study will be applicable for sweetening for the natural gas produced in other gas fields that have a similar ratio of CO2 to H2S. The sour gas that is produced in this field meets these specifications.

A SUMMARY OF THE THREE PROCESSES

Alkalonomines are utilized on a large scale for the removal of acid gas from natural gas [1, 15, 21-26]. Based on the amount of alkyl groups that form bonds with the N atom in the structure of the amino group, Alkalonomines are categorized as amines that are primary, secondary amines, or tertiary amines. The most widely utilized alkanolamines are monoethanolamine (primary), diethanolamine (secondary), with methyldiethanolamine (tertiary) [15, 25-29]. The selection of an alkanolamine process for the sweetening of natural gas has an effect on the cost of capital and operational expenses, the energy needs, the size of the equipment, and in certain cases the kind of equipment that is required for the sweetening [25, 27]. Through reactions (1-2) [17, 30], alkanolamines are able to take up the acid gases that are present in natural gas.

$$\text{Secondary amines: } CO_2 + R_2NH \leftrightarrow R_2NH^+COO^- \qquad (1)$$

$$\text{Tertiary amines: } CO_2 + R_3N + H_2O \leftrightarrow R_3NH^+ + HCO_3^- \qquad (2)$$

DEA

Diethanolamine, also known by its abbreviation DEA, is a kind of secondary amine. Aqueous solutions of DEA are utilized in the process of removing hydrogen sulfide and carbon dioxide from natural gas [25, 26, 31]. There are a variety of compounds, including COS, CS2, SO3, and SO2, that are capable of catalyzing the breakdown or deactivation of alkanolamine solutions [2, 32]. Due to the low reaction rate with CS2 and COS, DEA and other secondary amines are the preferable choice for sweetening natural gas [25, 26]. This is because significant levels of CS2 and COS are present in the sour gas, and DEA has a low reaction rate with both of these components. Because DEA solutions are not very discriminating, they are suitable for the absorption of either H2S or CO2 from natural gas [26]. The concentration of DEA solutions used in industry ranges from 25 to 40 weight percent [27, 33]. The DEA sweetening process is simulated by using the Aspen HYSYS simulator, and the different scenarios of simulation are economically assessed by using the aspen economic evaluation (Icarous) program. The findings are compared to those obtained by the DGA and MDEA+DEA procedures. The DBR-Amine property package has been utilized in order to do simulations of this process. Figure 1 presents the simulation flow sheet for your perusal. We utilized a tray absorber that had a total of 20 theoretical steps. In addition to this, the model of the regenerator column was based on a tray column that had 18 theoretical steps. The pressure within the regenerator can range anywhere from 27.5 psia (when it's acting as a condenser) to 31.5 psia (when it's acting as a boiler). The pressure of the rich DEA is lowered to 90 psia in the valve, and it was expected that there would be no reduction in pressure in a two-phase separator.
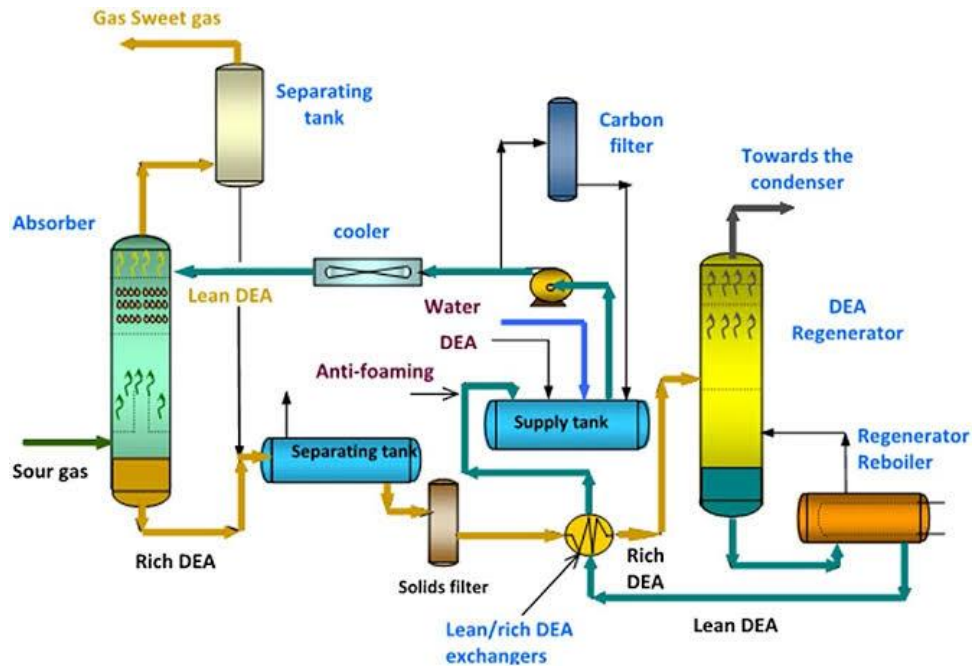
Fig. 1: Simulation flow sheet for the DEA sweetening process

DGA

The sweetening of natural gas often makes use of Diglycolamine, which is a primary amine. Due to the low vapor pressure of DGA, it is possible to use aqueous solutions of this amine at quite large concentrations (40-70 wt%) for the sweetening of natural gas. This results in low amine circulation rates, which are required for the sweetening of natural gas [25, 33]. When it comes to the treatment of natural gas at low pressure, DGA solutions are among the most successful. DGA has a propensity to preferentially absorb CO2 when H2S is present [33]. On the other hand, DGA absorbs aromatic compounds, which makes the sulfur recovery unit more problematic [34]. As a result, DGA is an excellent candidate for sweetening natural gas with a relatively high CO2 content. In light of these assertions, DGA has been chosen to represent one of the solutions for the sweetening of natural gas in accordance with the requirements. In this particular investigation, the natural gas was sweetened with an aqueous solution containing 65 weight percent of DGA. Both the simulation and the economic assessment of this process have been carried out with the assistance of Aspen Hysys and Aspen economic evaluation. The simulation of this process made use of the DBR-Amine property package. Figure 2 depicts the simulated flow chart for this procedure. We utilized a tray absorber that had a total of 20 theoretical steps. In addition to this, a tray column consisting of 20 theoretical phases was utilized in the process of designing the regenerator column. It has been determined that a pressure of 24 psia should be applied to the regenerator. The pressure of the rich DEA is lowered to 25 psia in the valve, and it was believed that there would be no pressure drop in the two-phase separator.
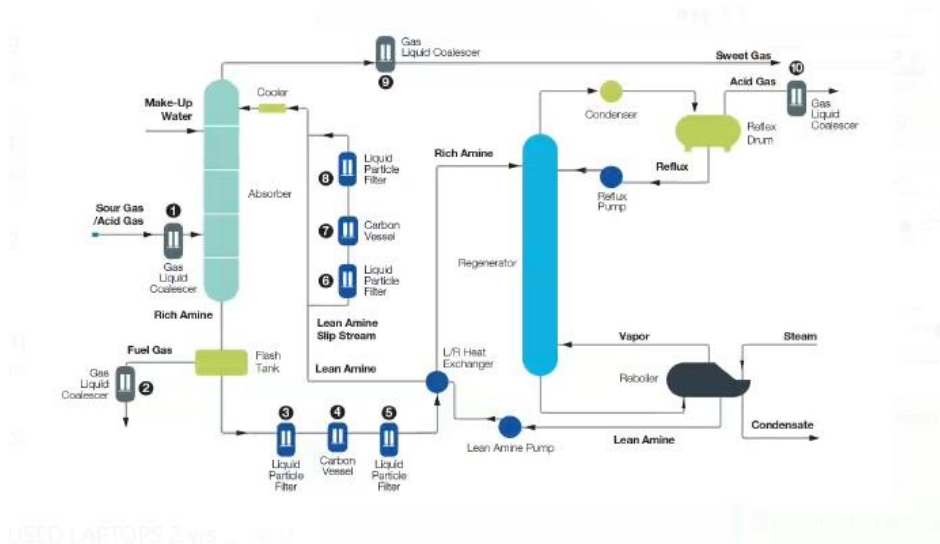

Fig 2: Simulation flow sheet for the DGA sweetening process

MDEA+DEA

In the presence of carbon dioxide, the tertiary amine known as methyldiethanolamine (MDEA) is reported to have a better selectivity for absorbing hydrogen sulfide gas [27]. MDEA's interaction with H2S takes place at a nearly instantaneous rate, but its reaction with CO2 takes place at a more gradual pace. However, a large number of studies have shown that increasing the overall CO2 absorption rate of a process by adding trace quantities of primary or secondary amines to a tertiary amine increases the rate of CO2 absorption to rise [6, 25, 27, 33, 35-37]. Because of the comparatively high amount of carbon dioxide (CO2) in the sour gas, I decided to improve the CO2 absorption rate of the MDEA process by adding 10 weight percent of a secondary amine (DEA) to the solution. This can make this process a promising process for sweetening of the natural gas described in section 2, and it was my decision to do so because of the relatively high CO2 content in the sour gas. The second rationale for combining the recommended amine solutions is to combine the high level of reactivity of the secondary amine with the comparatively low level of regeneration energy demands of the tertiary amine. In industrial settings, the usual concentration of MDEA in aqueous solutions is between 30 and 50 weight percent. For the purpose of this investigation, an aqueous solution consisting of 40 weight percent MDEA and 10 weight percent DEA has been chosen for the purpose of sweetening the natural gas presented in section 2. This solution was chosen because it is one of the examples that has the best performance in terms of the absorption of CO2 [6]. For the purpose of simulating this process, Aspen HYSYS is utilized, and For the purpose of economically analyzing this process, Aspen economic assessment is utilized. In order to simulate this process, the DBR-Amine property package is typically utilized. Figure 3 provides a depiction of the flow chart for the simulation. We utilized a tray absorber that had a total of 20 theoretical steps. In addition to this, a tray column consisting of 20 hypothetical phases was utilized in the process of designing the regenerator column. It has been determined that a pressure of 24 psia should be applied to the regenerator. The pressure of the rich DEA is lowered to 25 psia through the valve, and it was believed that there would be no reduction in pressure in the two phase separator.

III.   SWEETENING AT ADVANCED LEVELS

Applications: Advanced sweetening is a comprehensive portfolio of amine-based methods for sweetening natural gases (DEA, MDEA, and activated MDEA). The Advanced Amines processes address all forms of acid gas removal applications, for any feed gas composition and product criteria as low as 1 ppm H2S and 50 ppm CO2.

Description: TOTAL's considerable industrial and operational experience in developing these technologies (formerly SNEA-(P), ELF Group) is the foundation of the Advanced Amines portfolio. It consists of the following procedures:

➤ **High load DEA:** Based on the use of high concentration (4 mol DEA/l) and high packing (mol acid gas/mol DEA) DEA for exceptionally efficient full deacidification.
➤ **Selective MDEA:** Based on the use of pure MDEA aqueous solution for selective H2S elimination or H2S enrichment.
➤ **Activated MDEA:** With a variety of patented activators, for all applications involving bulk CO2 removal and full deacidification. Particular benefits of activated MDEA include partial or complete flash regeneration from the solvent for CO2 removal operations.
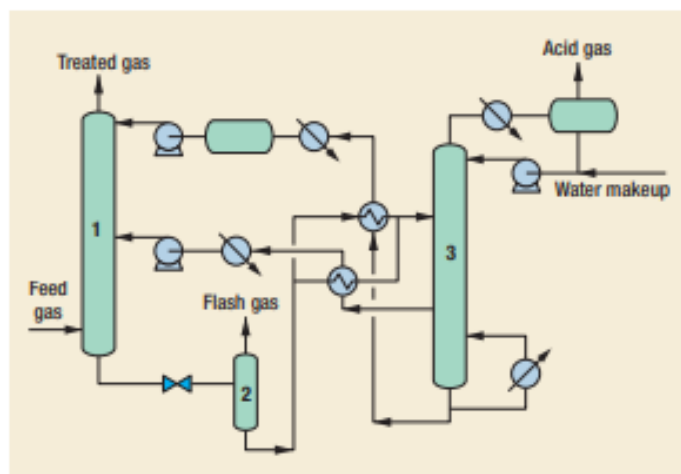

Fig 3: Simulation flow sheet for the MDEA+DEA sweetening process

Different flow schemes, ranging from the traditional absorber/thermal regenerator method to more complex flow schemes, are possible for all of these processes. The Double Split Flow method, for instance, maximises acid gas removal while requiring the least amount of energy. It works by sending a flow of semi-lean amine back from the intermediate level (3) of the thermal regenerator to the intermediate level (1) of the absorber. Utilising activated MDEA and flash solvent regeneration can further reduce energy needs. High COS hydrolysis levels (up to 95% COS elimination) are also possible with high load DEA thanks to a patented absorber design. Installations: There are about 120 units, roughly a quarter of which are run by TOTAL, with unit capacities ranging from 0.3 to 25.2 Nm3 per day.

IV.   MDEA SWEETENING PROCESS

Application: Removal of CO2, H2S, COS, RSH, and other gas-related gases from synthesis gas, natural gas, or other gases.

Products: Treated gas that satisfies the requirements of pipelines, LNG plants, ammonia plants, or petrochemical plants. Contains relatively little inert gas and emits an acidic gas. It is feasible to produce food-grade CO2.

Description: Absorption using an Alkanolamine MDEA solution and an activator system removes the acid components contained in the input gas. Flashing and/or stripping via one or more regeneration phases regenerates the rich solution that exits the absorber. To fulfil the needs of specific applications, various solvent types and concentrations can be coupled with various process designs. It is feasible to adapt gas treatment to the customers' financial priorities.
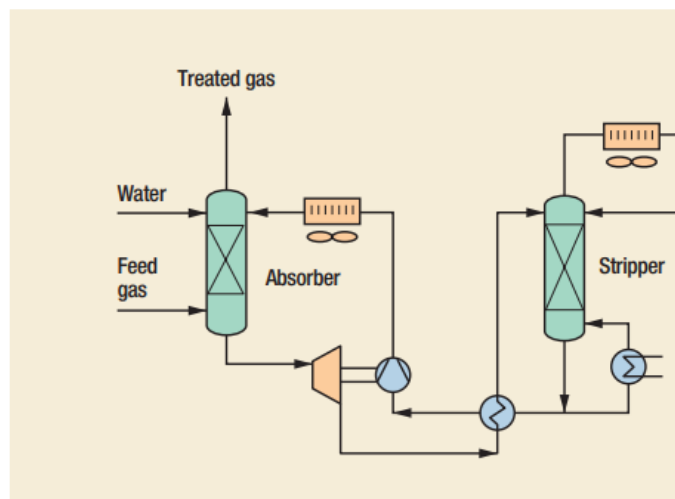

Fig 4: Simulation flow sheet for the MDEA sweetening process

Operating conditions: Reference plants include feed gas capacities ranging from 3,000 to 810,000 Nm3/hr, absorber temperatures ranging from 30°C to 90°C, absorber pressures ranging from ambient to 120 bar, and feed gas compositions ranging from 0.5 to 25% CO2 and 0 to 15% H2S. It is possible to treat at greater pressures with increased CO2/H2S in the input gas.

Economics: Because of the higher acid-gas loadings achieved with the solvent, the process is very energy efficient; this allows for low circulation rates and lower energy consumption, as well as a reduction in necessary equipment size. CO2 removal from aalkanolamine gas requires 1 kWh/kmol CO2 electrical power and 32 MJ/kmol CO2 thermal energy. Natural gas treatment thermal energy consumption: 15-20 MJ/kmol CO2 and H2S eliminated (flash regeneration). Other benefits include: extremely little hydrocarbon co-absorption, no degradation products, no corrosion (mostly carbon steel equipment may be used), minimal foaming tendency, no reclaimer operation is required, and the solvent is harmless and biodegradable.

V.   RESULTS AND DISCUSSION

The results of the simulation and the operational circumstances

Studying the impact of cooler's operating parameters on the performance of the three sweetening processes led to the development of seven distinct scenarios, one for each of the three processes. Each of these situations illustrates the properties of the system under a certain operating condition of the cooler. In each of these scenarios, the duty of the cooler is adjusted until the temperature of the lean solution reaches the value that was planned for. In every one of these

hypothetical situations, the parameters of the process are altered in such a way as to achieve concentrations of sweet natural gas that are lower than 1mol% CO2 and lower than 4ppm H2S.

During the modeling of these processes, the minimum temperature approach for each of the heat exchangers was assumed to be 10 degrees Celsius, and the adiabatic efficiency of the pump was set at 75%.

Following the conclusion of the simulation of three processes, the following seven scenarios are implemented for each process, and the process is then economically assessed using aspen economic assessment version 7.3.

The circulation rate (gpm) of the solution is one of the most essential aspects of a sweetening process [15, 38]. Increasing the solution flow rate results in a rise in the capital and operational expenses of the process, as well as the size of the equipment and the energy needs [15, 25, 39]. Figure 4 presents the findings on the flow rates of the solutions produced by the processes under various conditions. The statistics shown in Figure 4 make it abundantly evident that the amine circulation rate for the mixed amine process is significantly greater than that of the DGA and DEA in all seven of the considered situations. Figure 4 further demonstrates that when the temperature of the lean amine rises, there is a corresponding modest increase in the solution flow rate that is required for each procedure, with the lowest temperature of the lean amine exhibiting the lowest minimum necessary solution circulation rate. As was said previously, raising the solution flow rate in a sweetening plant produces a rise in the facility's capital and operational expenses as well as the size of the equipment required for the process, as well as the energy needs. On the other side, lowering the temperature of the lean solution calls on the cooler to perform a greater amount of work. This bigger duty might be attained by either expanding the area where contact occurs of the heat exchanger or switching the cooling material; nevertheless, regardless of which option is used, this modification will result in an increase in the cost of operating the plant. On the basis of these assertions, it would appear to indicate that there should be a point of operation that is considered to be optimal for the cooler in a sweetening plant. In the present study, I attempted to determine this threshold for each of the three distinct methods of sweetening. According to the data shown in Figure 4, the resultant solution circulation rate for the mixed process is noticeably greater compared to the solution circulation rate for the DEA process. This discovery is thought to be caused by the fact that methyldiethanolamine absorbs H2S preferentially while having lesser capabilities for absorption of carbon dioxide (CO2) [25, 26, 40].
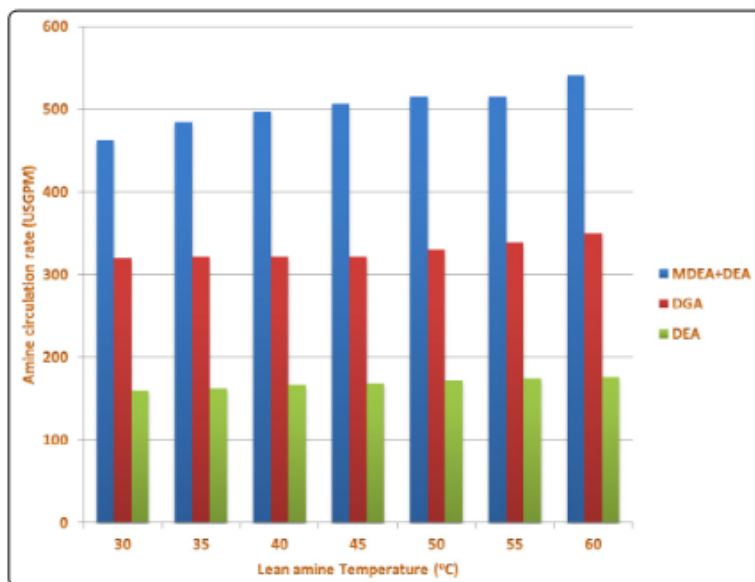

Figure 4. Effects of the lean amine temperature on the required solution circulation rates

The number of hydrocarbons that are taken in by the solution as it is passing through the contactor is yet another crucial feature of the functioning of sweetening procedures. According to prior research, the co-absorption of hydrocarbons is primarily an issue that arises with physical and physical-chemical solutions [9, 25, 27, 41, 42]. Nevertheless, I investigated this parameter on the three chemical absorption systems in order to validate the findings from the simulations. As can be seen in Figure 5, even while the rate of hydrocarbon picks up by the solution is increased at temperatures that are lower when using the mixed amine process, the rate of hydrocarbon picks up by the solution stays at a very low rate regardless of the different cooler's operating parameters used in the three processes. For the mixed amine procedure, the solution was able to pull up a maximum of 0.0004 percent more hydrocarbon than any of the other

21 simulated situations. Figure 5 clearly reveals that at temperatures greater than 45 degrees Celsius, the MDEA+DEA mechanism takes up less hydrocarbon than it does at lower temperatures. The temperature of the lean amine has little effect on the hydrocarbon that the DEA and DGA systems pick up on, however.
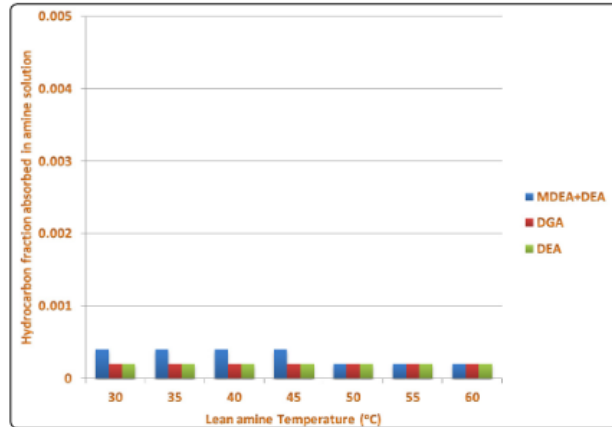

Figure 5. Effects of the lean amine temperature on the rich amine hydrocarbon pick up

Because the chemistry processes that result in the absorption of acid gases through the alkanolamine solutions are exothermic [25, 43-45], it is reasonable to anticipate that the ambient temperature of the rich amine will be higher than that of the lean amine entering the contactor. The temperature difference between these streams can be a parameter that shows the intensity of the mechanism of absorption in the contactor. Figure 6 and Figure 7 demonstrate the temperature difference between the rich amine stream and the less concentrated amine stream, as well as the rich amine temperature. According to the results presented in Figure 7, the temperature of the reach amine rises whenever the temperature of the lean amine that is entering the contactor is made higher. On the other hand, regardless of which of the three procedures one uses, the temperature distinction between the two streams narrows as the temperature of the lean amine rises. In the DEA process, the temperature of the rich amine is even lower than the temperature of the lean amine for temperatures of the lean amine that are higher than 35 degrees Celsius. The phenomenon is ascribed to the fact that there was a greater temperature differential between the streams of feed gas and lean amine, which resulted in a stronger transfer of heat across the cold feed gas (which was at 21 degrees Celsius) and the lean amine.

Because the rich amine that is at the bottom of the contactor needs to be removed and regenerated at high temperatures, this is another significant concern that needs to be handled here. The temperature of the rich amine has a direct bearing on the amount of energy that the system requires to function properly. Therefore, a rise in the temperature of the rich amine will result in a reduction in the amount of energy that the system needs (or the contact area of the heat exchanger).
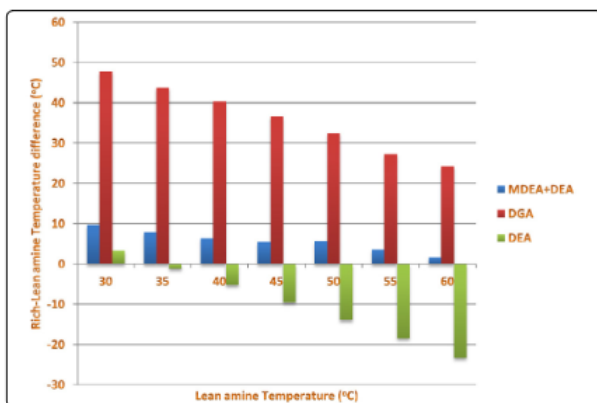

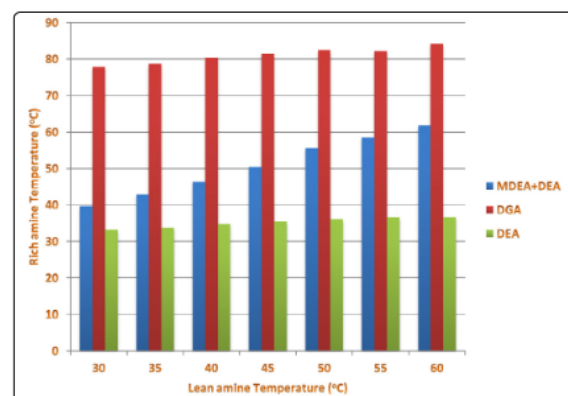Figure 6. Effects of lean amine temperature on the Rich-Lean amine temperature difference

Figure 7. Effects of the lean amine temperature on the rich amine temperature

The energy demands of the processes involved in sweetening are yet another essential aspect of these procedures. The temperature of the lean amine has a direct influence on the amount of duty that has to be applied in the cooler. The temperature of the lean amine also has an effect on the energy needs of the stripper as well as the duty of the heat exchanger. Figure 8 demonstrates that the cooler's duty rises for all three operations once the temperature of the lean

solution drops. This is an expected result due to the fact that the temperature difference surrounding the cooler grows while the output temperature drops. At the lean amine temperature that is the greatest, the minimal cooler duty that should be observed is observed. This finding is consistent with the pattern that was anticipated. Figure 9 demonstrates, in addition, that when the temperature of the lean amine rises, the heat exchanger duty gradually declines as a result of this tendency. Figure 9 reveals another significant finding, which is that the heat exchanger duty for the DGA process is significantly lower than that of the DEA process and the MDEA+DEA method. This observation is as a result of the fact that the operating temperature of the rich amine in the DGA process is significantly greater than that of the rich amine in the remaining two processes. In comparison to the DGA and MDEA+DEA processes, the DEA process has a lower solution circulation rate, which is another factor that contributes to the reduced cooler and heat exchanger employment of the DEA process.
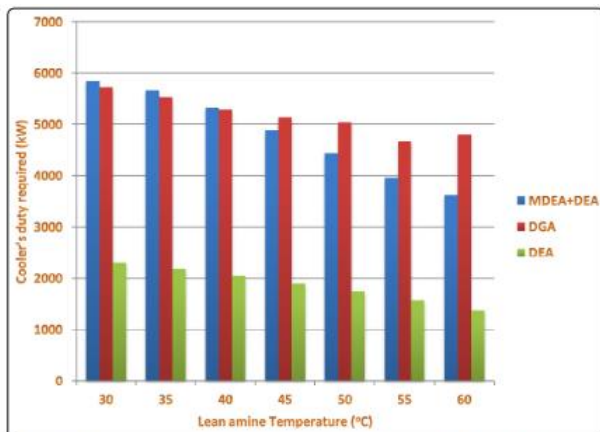

Figure 8. Effects of the lean amine temperature on the cooler's duty
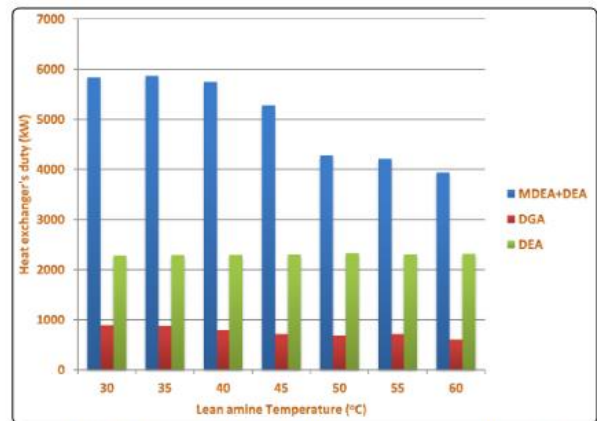

Figure 9. Effects of the lean amine temperature on the heat exchanger's duty

Following the completion of the modeling of seven different scenarios for each of the processes, an overall economic evaluation is performed on every scenario to determine its financial viability. It has been speculated that the construction of the projects will begin in the year 2014. The findings are shown in US dollars or US dollars per year depending on the situation. When estimating the capital and operational costs of the processes, certain parameters, such as the degree of difficulty of the processes, the starting date, and the level of instrumentation, are taken into consideration. When the temperature of the lean amine hits 40 degrees Celsius, the MDEA+DEA process reaches its lowest possible capital costs, as seen in Figure 10. These findings are based on the findings of an economic evaluation. In addition, it is evident that the capital expenses of the DEA and DGA processes follow a pattern of lowering when the temperature of the lean amine rises from 30 oC to 60 oC. When the DEA process is employed, the lean amine temperature reached during this process is the highest temperature that was evaluated. The capital expenses associated with the DGA process are only marginally higher than those associated with the DEA process. The DEA method results in a lower overall procedure capital expenditures.
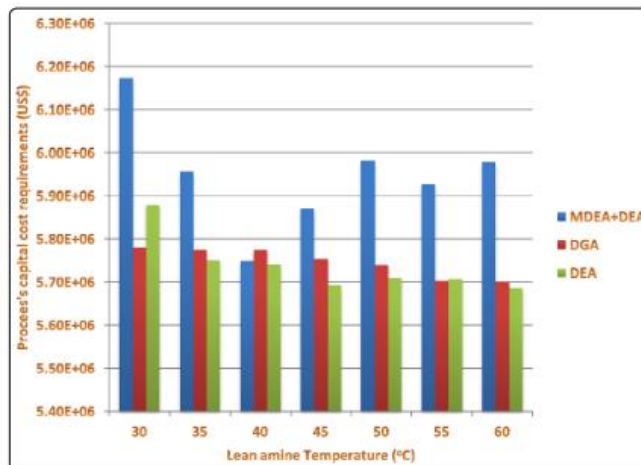

Figure 10. Effects of the lean amine temperature on the capital costs of the processes

Figure 11 displays the yearly cost of operation figures obtained from simulating seven different scenarios for each of the operations. The yearly operating costs of the three processes do not appear to be strong functions of the operating temperature of the lean amine, based to the findings presented in Figure 11. It is possible to provide a justification for these observations by casting doubt on the data shown in figures 8 and 9. It was said previously that altering the temperature of the lean amine does not affect the duty of the strippers reboiler. Additionally, it was stated that lowering the temperature of the lean amine has a favourable effect on the duty of the heat exchanger but a negative effect on the duty of the cooler. On the basis of this information, one may draw the conclusion that the adverse and beneficial impacts of this modification are not particularly significant or that these effects cancel each other out.

This is the reason why there has not been any obvious change in the costs of utilities and, as a consequence, the yearly operating expenses of the system that have been recorded. Figure 11 makes it abundantly evident that the yearly operating expenditures and utility costs associated with the DEA methodology are significantly cheaper than those associated with the DGA process as well as the MDEA+DEA process. Taking into account that each of the three processes has a life cycle of 25 years, the primary expenses connected with the processes are the yearly operating costs and the utility costs.

According to the findings presented in Figure 11, the yearly operating expenses and utility costs of the DGA and DEA processes are unaffected by the choice of lean amine temperature. As a result, the lean amine temperature does not play a significant role in the costs associated with these two processes. However, the results of the MDEA+DEA process are more convoluted, and the yearly operating costs of the process do not follow a straightforward pattern. The lean amine temperature of 30 degrees Celsius is where the annual operating costs are the lowest. Taking into account that this process has a life cycle of 25 years, this level of temperature demonstrates the highest return on investment for it.
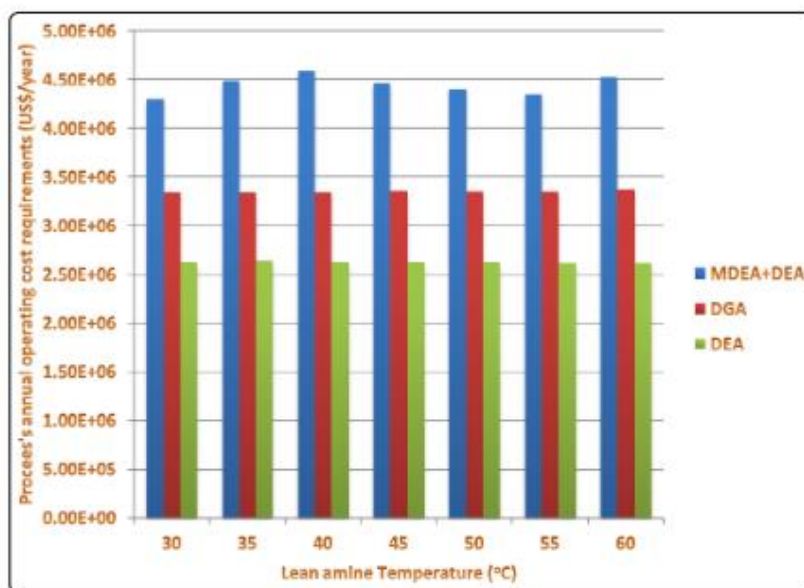

Figure 11. Effects of the lean amine temperature on the annual operating costs of the processes

According to the previous talks, there are a number of benefits associated with running the DGA, DEA, and MDEA+DEA sweetening processes at higher lean amine temperatures. Operating the process at higher lean amine temperatures results in cheaper capital costs for the process, lower rich amine hydrocarbon picks up in the case of MDEA+DEA process, and lower cooler duty.

An inquiry of the price and energy consumption of several additional sweetening techniques that are suitable might be a way to enhance the outcomes of this research. The prices and energy needs of alternative acceptable methods for the sweetening of natural gas with specifications near to those of the natural gas that I have discussed might be the subject of future studies, and this could be an area of focus for those research projects.
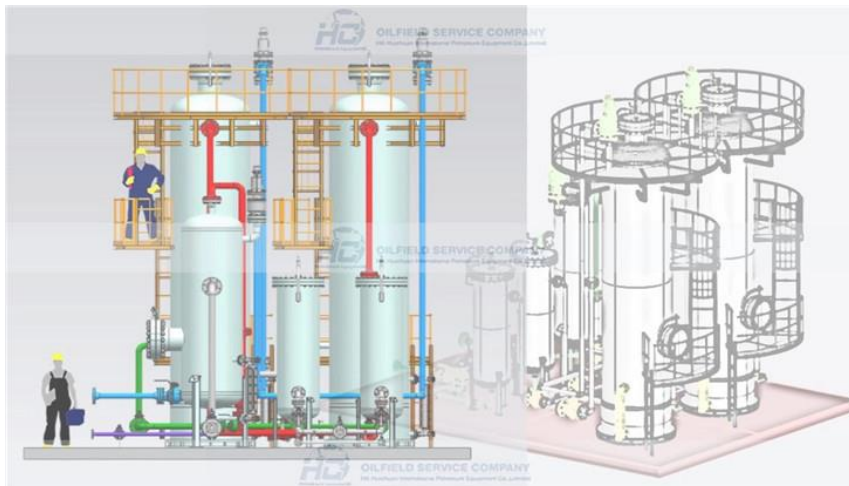
Fig 12: Sweetening process unit

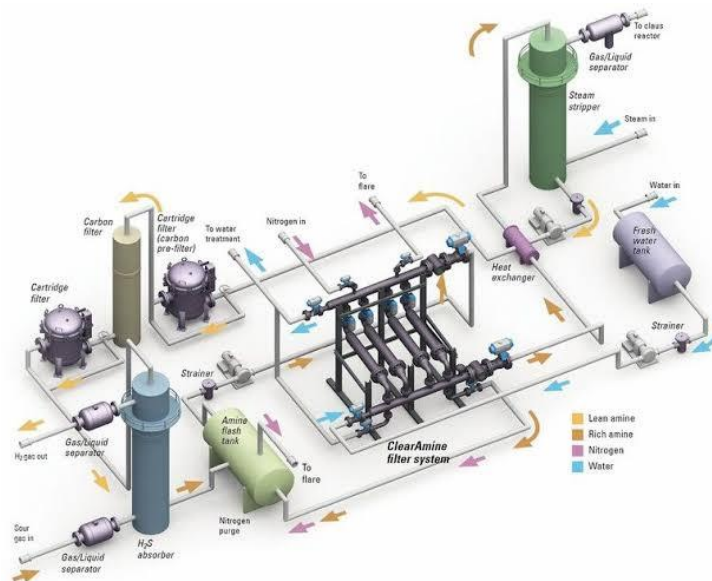**Fig 13:** High efficiency of natural gas sweetening

Fig 14: Modernizing amine sweetening unit of natural gas

Modernising a natural gas amine sweetening unit entails improving the present system to increase efficiency, dependability, and safety. Amine sweetening is a typical method for removing pollutants from natural gas streams, notably hydrogen sulphide (H2S) and carbon dioxide (CO2).

Various enhancements are possible during the modernisation process. Here are several examples:

➢ **Upgraded Equipment:** It is possible to upgrade or replace older, less efficient pieces of machinery such the heat exchangers, absorbers, and strippers. This improves the unit's overall performance.

➢ **Advanced Automation:** The sweetening process may be better monitored and optimised by putting modern automation and control technologies in place. This ensures optimum performance by regulating the chemical consumption, temperature, pressure, and other crucial variables.

➢ **Energy Efficiency Enhancements**: Modernization sometimes incorporates energy-saving techniques like heat recovery systems, which warm up incoming gas or water using waste heat from the operation. This lowers energy use and running expenses.

➢ **Safety Improvements:** The plant and its operators will be in a safer environment thanks to upgraded safety features such gas leak detection systems, emergency shutdown systems, and better fire prevention measures.

➢ **Environmental Considerations**: The goal of modernisation is to have less of an influence on the environment. The inclusion of amine degradation control devices, which reduce the atmospheric emissions of amine compounds, is one example of such upgrade.

A contemporary amine sweetening equipment is more effective, more cost-effective, safer, and compliant with environmental regulations all around. It enables improved control, boosted output, and a more dependable natural gas purification process.

VI. CONCLUSION

The benefits and disadvantages of cooler operation settings on the functionality of three sweetening processes intended for sweetening natural gas generated in an oil or gas field with a high CO2/H2S ratio (with the requirements stated in section 2) have been explored. The aforementioned procedures were developed for sweetening natural gas produced in a gas field with a high ratio of CO2 to H2S. The natural gas that is extracted from this gas field will be sweetened using a combination of the DEA, DGA, and MDEA+DEA procedures. Each of these methods was devised in such a manner that it would be possible to get concentrations for the sweet gas that were lower than 1mol% CO2 and lower than 4ppm H2S.

According to the findings of this research, when it comes to the DEA and DGA processes, running the processes with a higher lean amine temperature exhibits numerous advantages in the range of temperature for the lean amine that is between 30 and 60 degrees Celsius. By running the processes at higher values of lean amine temperature, it was possible to get lower values of the following expenses: the capital costs of the process, the rich amine hydrocarbon picks up in the case of the MDEA+DEA process, and the cooler's duty. Despite the fact that an increase in the temperature of the lean amine caused a slight increase in the required circulation rate of the solution in order to achieve concentrations of less than 1mol% CO2 and less than 4ppm H2S for the sweet gas, it is still recommended that the DGA along with DEA sweetening processes be carried out at higher temperatures for the lean amine.

An inquiry of the price and energy needs of several additional sweetening techniques that are suitable might be an opportunity to enhance the outcomes of this research. The prices and energy needs of alternative acceptable methods to facilitate the sweetening of natural gas using specifications near to those of the natural gas that I have discussed might be the subject of future studies, and this could be an area of focus for these research projects.

RECOMMENDATION:

➢ Optimize the lean alkanolamine degree: The degree of lean alkanolamine impacts the overall performance of the solution. Experiment with various diethanolamine (DEA), Diglycolamine (DGA), and methyldiethanolamine (MDEA+DEA) ratios to determine the optimal composition for successful CO2 absorption.

➢ Conduct thorough testing: Conduct extensive laboratory testing to evaluate the performance of alkanolamine solutions before using them in large-scale absorption operations. This involves examining potential CO2 absorption capability, solubility, equilibrium, overall degradation potential.

➢ Monitor degradation and byproduct formation: Alkanolamine solutions can degrade and generate byproducts over time, affecting their effectiveness. Implement frequent monitoring and analysis to detect any deterioration or byproduct development, allowing for solution regeneration or replacement in a timely manner.

➢ Optimize operating conditions: Changing operating variables such as temperature, pressure, and solution flow rate can have a considerable influence on the performance of an alkanolamine solution. Experiment with various situations to determine the best set of parameters for effective CO2 absorption.

➢ Improve heat management: Proper heat control is critical for preserving alkanolamine solution performance. To achieve ideal ambient temperatures throughout the absorption process, use efficient heat transfer mechanisms or temperature control devices.

➢ Regular maintenance and solution regeneration: It is critical to plan frequent maintenance as well as regeneration of alkanolamine solutions in order to ensure constant performance. This entails removing deteriorated solution, purifying it, and refilling any required additions or components.

CONFLICTS OF INTEREST:
The Authors declare that they have no conflict of interest.

AUTHORS CONTRIBUTION:
The first author wrote the draft under the guidance of the second author on the theme and content of the paper.

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