



Case report

Dependence of operational properties on reboiler duty in the removal of CO₂ from natural gas with aspen Hysys

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ARTICLE INFO

Keywords:

Absorption
Carbon dioxide
Natural gas
Circulation rate
Reboiler duty
CO₂ specification

ABSTRACT

Obtaining CO₂ specification for cooking-gas requires an equilibrium stage in the absorption column of an existing plant. The column temperature influences the stage efficiency, 25 %, which must be kept constant for all stages to optimize the absorption temperatures. Additionally, the procedure employed the consistent daily plant data which were simulated directly. In this method the recycled 29 % monoethanol amine (MEA) start-up was imposed, so that 10 stages within 20–60 °C at 1.1×10^2 kPa were subjected to a circulation rates of 6000–16,000 kg.mol/h. These parameter were used to evaluate the reboiler duty. The Kent–Eisenberg thermodynamic model renders circulation rate the most effective operational property. The reboiler duty at 1.19×10^9 W yielded 81.08 % grade of natural gas with negligible amount of CO₂ retained in the natural gas; which implies that a high stage efficiency was attained. Cleaner aqueous amine was obtained by raising the reboiler temperature to yield higher absorption efficiency, but the temperature was restricted to a maximum value of 120 °C due to amine degradation at higher temperatures. Thus, by adding the loss due to old circulation pump, the overall efficiency is significantly improved and the energy requirement is minimized.

1. Introduction

When burning natural gas for energy usage, less carbon dioxide (CO₂) and almost all other air pollutants are released into the atmosphere [1] compared to when burning coal (200 pounds of CO₂ per million British thermal units (MMBtu) or petroleum 160 (MMBtu) products for the same amount of energy. In contrast, 117 (MMBtu) CO₂ equivalent is released by natural gas [2]. Therefore natural gas is now used more frequently in Nigeria as transportation fuel for carts and for the production of farm and industrial power due to its lesser pollution.

Domestic homes and restaurants use natural gas as cooking gas because of the effectiveness, the availability and the comparatively clean-burning quality. However, there are some environmental and safety concerns with the untreated natural gas. The presence of carbon monosulfide (CS), CO₂, H₂S, elemental sulfur, mercaptans and contaminated water decreases the quality of sales gas and their presence is not acceptable, since they corrodes pipelines and production equipment. More often than not, the gas containing CO₂ when release into the atmosphere leads to acid rains. Therefore, the contractors are legislated to demand for safe specification of CO₂ content before any sales are made by producers. This has necessitated the need to bring down the CO₂ component to a safe measure in the natural gas.

Although, some aqueous solutions containing piperazine [3], are used as washing agent for removing H₂S, CO₂ and Carbonyl sulfide (COS) from the natural gas, coke-oven gases, and synthesis gases. Continuous investigations include invention of new and efficient means of CO₂ removal by equilibrium conditions and other associated procedure. Apart from continuous adsorption [4], a means of modeling [5] and experiments deduced by [6] presented a procedure for the removal of carbon-dioxide where piperazine was among methyl diethanol amine (MDEA), methyl ethanol, diethanol amine (DEA) and their blends [7–9]. The mass transfer flux of the CO₂ has been optimized using artificial neural networks (ANNs) [9] and some other combination of amine has been reported by Ref. [10]. The kinetic studies for these amine was carried out by Refs. [11,12]. A high pressure absorption procedure was done by Ref. [13]. Subsequently the kinetic study for the procedure that included an ionic liquid amine blends have been reported by Refs. [14, 15].

Although in current studies the monoethanol amine (MEA) has been confirmed as suitable for the removal of CO₂ [16]. The work of [17] reported that CO₂ was removed from a mixture of gases by amine blends of monoethanol amine and piperazine, their thermodynamic analysis included the use of an improved Kent Eisenberg model. An absorption studies by Ref. [18] adopted the experimental, simulation and

E-mail address: grace.olugbenga@uniabuja.edu.ng.<https://doi.org/10.1016/j.rineng.2024.101755>

Received 3 October 2023; Received in revised form 3 January 2024; Accepted 3 January 2024

Available online 17 January 2024

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thermodynamic modeling of an acid gas removal which was validated using data from pilot plant to record the extent in the capturing of CO₂ with mono-ethanol amine solution. Following this research [19], simulated the removal of CO₂ using mono-ethanol amine with computational fluid dynamics and a high percentage of absorption was achieved. The MEA alone and in combination with methyl di-ethanol amine has exhibited improved absorption rates of CO₂ from aqueous solutions, in both cases the MEA is prevalent in effecting the absorption process. Evidently, the potential of MEA is known as an absorbent for CO₂ [20] but the process optimization remains a stiff challenge because of the varying circulation rate which differs with the plant in focus. Additionally, the associated convergence involved in simulating the removal of CO₂ using ASPEN Hysys is yet to be reported for associated natural gas with the constraint of a maximum number of stages available in the physical plant even when a higher grade of CO₂ retained in the outlet gas depends among other factor on higher number of plates. There are hindrances to the convergence of simulated process by reason of the warning response which are not within the convergence capability of the software. Therefore exploring within many versions of software, one of the available tools for the simulation process is Aspen HYSYS which consists of an acid gas removal model that supports various amines in the data base [21,22]. Some absorption facts are found in the extensive work done by Refs. [23,24].

The carbon dioxide content of natural gas must be reduced in order to raise the heating value of the gas, stop pipeline and gas processing equipment from corroding, and keep the CO₂ from crystallizing during liquefaction. Over time, many procedures and their improvements have been created to treat different types of gas to avoid contributing to environmental pollution, but to meet gas standards and optimize costs [25].

Among the new treatment route for commercial purposes is the utilized acid gas removal technologies which are chemical absorption processes that uses amine solutions. The weakly bound intermediate compounds that are formed by the CO₂ and sorbent interactions during chemical absorption processes can be reversed by applying a significant amount of heat in order to regain the CO₂ and restore the sorbent [26, 27]. Henry's rule states that physical solvents, like the polyethylene glycol-dimethyl ether mixture employed in the Selexols process, preferentially absorb CO₂ from the natural gas source, this phenomenon increase the absorption capacity at significant temperature.

For the gas-liquid systems, the cost variables are the required circulation rate, which is dependent on the loading capacity of the sorbent, the amount of CO₂ that needs to be removed and the stripping cost [28]. In absorption process where low or moderate CO₂ partial pressures are found, the physical solvents often require a lower acid gas loading capacity than the chemical absorbents. However, because the heat of absorption for physical solvents is far lower than that of chemical solvents, the former procedure require less energy for stripping. Noting that the application of monoethanolamine (MEA), which has been thoroughly explained in a variety of procedures has been widely employed as a solvent for the non-selective removal of acid gas; the industry is currently working to commercialize MEA-based CO₂ collecting systems [29,30] for example 30% MEA solvent is used in the Fluor procedures. The solvent is combined with inhibitors to stop equipment corrosion and solvent deterioration. Amine concentrations of up to 35% can be used and reported to be used in conjunction with a corrosion inhibitor if CO₂ is the only acidic gas in the feed [15,31]. The evolving thermochemistry of MEA [19,32] are reported in which some results are included in the work of [20,33]. It is necessary to heat the rich amine solution from the absorber before adding it to the desorption column. It is also necessary to withdraw the heat in the lean amine to be regenerated before re-entry into the absorber. For this reason, cross flow heat exchangers could be used where both streams are constructed to exchange heat [34]. This optimize the heat to the lean amine cooler and the reboiler (in the desorption column) which are in charge of lowering the temperature of the lean amine to the absorber [35].

It is therefore evident that the groups of amines suitable for removing a small amount of CO₂ is the MEA. While the procedure is simple, MEA has a strong reactivity with CO₂ and a high solution capacity. However, one of its drawbacks, is that, it requires high energy to regenerate the MEA. An additional complication is that the high pressure gas streams are not a good fit for the MEA [36]. It is therefore of importance that the CO₂ removal process must be targeted towards lowering the regeneration energy. Because regeneration energy need makes up a significant portion of the operating costs. According to Ref. [37] it is about 80% of operating expense. Similarly [38] claimed that this value is about 60% of total operating expense.

The aim here is to improve the duty of the reboiler where the CO₂ removal occurring in an existing plant must be efficient and the retained CO₂ grade must meet standard requirement. The process was modelled in Aspen Hysys with equilibrium based methods by including a cross flow, reconstruction of the flow of MEA, which was not in the existing plant. The interaction effects of operational parameters in a plant with 10 specified stages is considered and the absorber temperature is monitored, MEA circulation rate affects the process and a reduced specific heat duty is expected. The parameter variation are based on experiments carried out in Aspen to explore the significance of each operational parameter and the interaction effect among parameter.

Additionally, the absorption of CO₂ occurs in the absorber and the CO₂ retained in the natural gas are calculated as a high grade which implies a very small quantity of CO₂ remaining in the gas phase. More often than not, the trays of an absorption column may not function at equilibrium in the physical plant. Therefore the attempts to model these columns with the physical operating conditions, required that the deviations from equilibrium in multistage systems should be corrected with stage efficiencies. Efficiencies are often used alongside the equilibrium stage to match operational data (of the physical column), along with the number of equilibrium stages in each section of the column. The knowledge of this equilibrium stage is necessary to push through solving the challenges of prediction of product specification which must be accompanied with reduction in the value of the reboiler duty.

2. Modelling absorption and desorption process

The absorption of CO₂ involve the basic calculations of tray efficiency requiring the knowledge of tray column in which a single tray column has an overall column efficiency (E_o) evaluated as a ratio of the equilibrium number of plate (N_{EQ}) to the actual number of plate (N_{actual}) in Equation (1)

$$E_o = \frac{N_{EQ}}{N_{actual}} \quad (1)$$

There are many definition of tray efficiencies in which the Murphery's efficiency, Equation (2), is the applicable option to the removal of a small amount of CO₂.

$$E_{ij}^{MV} = \frac{y_{ij} - y_{i,j+1}}{y_{ij}^* - y_{i,j+1}} \quad (2)$$

Where E_{ij}^{MV} is the Murphery efficiency, the component is 'i', the stage is j, y_{ij}^* is the composition of vapor in equilibrium with the liquid leaving the tray.

A few software has been used to model CO₂ absorption and desorption dynamically, for example [38]. With the aid of the commercial simulator, convergence of the process flowsheet (with a full compilation of flow streams and unit operations) was obtained. For the simulation of most unit operations, the sequential modular that employs forward sequence is a common example in Aspen Plus and Pro/II. Recycle convergence blocks are used in the evaluation of up-to-date stream and are used to compare the previously calculated streams. In this work, Hysys applications has expressed the capacity to calculate the flowsheet to obtain desorption, overcoming the short comings in the calculating

steps for the physical column, the convergence of the flowsheet were usually completed as though the system was sequential modular. Also the tool has the capability of adapting the recycle blocks to converge flowsheets [39]. Similarly different acceleration techniques (e.g. dominant eigenvalue techniques or Wegstein) and nested (simultaneous calculation sequence) are used for flowsheet calculations.

In the past, column convergence is peculiar and was the main approach to absorption processes [40,41]. It involves steady-state process simulation procedures. This is because the CO₂/amine/water combination is not an ideal system which makes an efficient column solver significant. Generally, column solver techniques are available and are similar to the default Hysim Inside-out approach in the Aspen HYSYS application. They include: Modified Hysim Inside-out, Newton Raphson Inside-out, sparse continuation solver, Simultaneous - correction, OLI solver, Fixed or adaptive damping factor [41,42].

(Column stages are associated to production cost of natural gas) thus, the flowsheet calculation were carried out by using a known model called Kent Eisenberg equilibrium [43], this simulation can be improved by using Aspen HYSYS similar to the former experiment [44] where their results revealed the major challenge with flowsheet calculation in which too many column-stages are tendencies to diverge convergence. This has been a challenge in most procedure during column -stage calculations. In the several attempt to improve flowsheet calculations, it was found that the Modified Hysim Inside-Out approach with adaptive damping offers an exceptional convergence [41]. In the base case calculated grade of CO₂ removed with a specific heat intake, their results suggests that the calculated CO₂ removal efficiency can be observed concurrently with the heat duty while varying the amine circulation rate. Thus a clear indication of minimum heat consumption at a specified rate of circulation was attained. The findings by these reports are appreciable. However the direct physical plant were not simulated in their report. In this current research, this procedure serve adequately to obtain process conditions and standard product specification based on available parameter in the physical gas plant. By following these procedure in ASPEN HYSYS, comparing the possible product quality, under a concurrently derived absorbent aimed at varying circulation rate, the reboiler duty was evaluated. This is because of the currency shift in inflation in Nigeria and thus an urgent need to determine an efficient operating plant is required.

The extensive property databanks, rigorous equation solvers and accurate representations of electrolytes found in Aspen hysys V9 makes it a preference application in the current research. The Electrolyte-Marguke property method was used for absorption process involving the CO₂- H₂O-MEA links. In this, the equilibrium-controlled reactions and reaction kinetics were carried out using [45–47] absorption procedures [48], also developed means, so that the equilibrium conditions are expanded in the method used. As soon as the appropriate equilibrium is attained, the sensitivity analysis of the temperature of MEA, feed temperature, circulation rate and the number of stages were done to expresses the results for clarity of purpose. And that the result will reveal the efficient interaction of these sensitivity items during plant operations. Overall, the CO₂ remaining in the natural gas meets the contractors specification when about 80 % grade of CO₂ is retained. While fixing all the parameters collected from the plant, the circulation rate is adjusted and until the least re-boiler duty is attained.

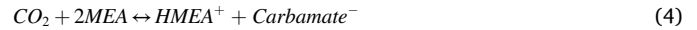
2.1. Modelling the absorption process

The molar concentration of the amine solution required for the absorption of CO₂ is one parameter used to the evaluation of an absorption process. The moles of amine required in the absorption process is obtained from the chemical reactions of the process. The first step in the absorption of CO₂ into the MEA is the mass transfer of CO₂ from the natural gas to the MEA, this is expressed in Equation (3).



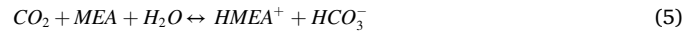
Once the CO₂ enters the amine, the protonated amine ion (HMEA⁺) and a carbamate

ion are formed as product of the reacting CO₂ and 2 mol of the MEA - Equation (4)



Where the carbamate is HN(C₂H₄OH)COO⁻)

Concurrently the bicarbonate ion is equally formed, the reaction is presented in Equation (5)



Therefore, the total molar concentration of CO₂ given in Equation (6) [49]

is the summation of the concentrations of the bicarbonate ions, the carbonate ions (HCO⁻) and the free CO₂ in the systems. These are the three forms in which the CO₂ occurs in the reacting system.

$$\sum Conc.CO_2 = Conc.HCO_3^- + Conc.CO_3^{2-} + Conc.Carbamate^- + Conc.CO_2 \quad (6)$$

On the other hand, the total concentration of the MEA Equation (7) [49] is the summation of the MEA portion of the carbamate, the protonated MEA (HMEA⁺) and the free MEA, which are the forms in which the MEA exist in the system.

$$\sum MEA = Conc.Carbamate^- + Conc.HMEA^+ \quad (7)$$

It is important to note that the reaction kinetics are vital to the description of the equilibrium stages, hence the kinetics are written in the form of the removed concentration of CO₂ which is equivalent to the rate at which the amine solution absorb the CO₂. Equation (8)

$$\frac{d(Conc.CO_2)}{dt} = k_2 Conc.CO_2 Conc.MEA \quad (8)$$

Where the left hand side of Equation (8) is the rate of absorption of the CO₂ (each mole of the CO₂ reacted per volume and time), k_2 is the temperature-dependent rate constant $Conc.MEA$ and $Conc.CO_2$ are the concentrations of free MEA and CO₂ respectively.

Next is, to validate the complicated reaction that occurs between the CO₂ and the amine (MEA). This has been explained in the work done by Ref. [50] that is; the rate constant of the reaction is a temperature-dependent value. This was based on the theory provided by Ref. [51] where the kinetic of the reaction was introduced based on Zwitterion mechanism. It is accepted as the base mechanism for the reaction presented in this work because of the absorptive nature of the MEA.

By taking advantage of the known concentration of the CO₂ in the absorption system, the equilibrium stage can be expressed with confidence of an adequate calculation, hence the equilibrium description is expressed in a gas/liquid system - Equation (9). Here, the constant "H_e" provided by Henry [52,53] is applied to the equilibrium state of the CO₂ content in the gas and the corresponding concentration of the free CO₂ in the amine phase

$$pCO_2 = H_e \cdot Conc.CO_2 \quad (9)$$

here pCO_2 is the partial pressure of CO₂ and $Conc.CO_2$ is the molar concentration of free CO₂ in the amine solution. During the absorption process the state of the mass transfer of CO₂ must be monitored in order to note the exact state of equilibrium. Hence the Peng Robinson's equation is chosen [54,55], because it can appropriate both the partial pressure of the gas in the gas and the liquid phases. Although the state of CO₂/amine absorption process is associated with the minor gas with non-ideal challenges, which are responsible for the difficulty in the achievement of the equilibrium state. It was taught that; replacing the partial pressure of the CO₂ with fugacity of the CO₂ will simplify the complications that are responsible for the equilibrium identity. Therefore this research included a procedure where the aqueous amine/CO₂

mixture was specified by the summation of the concentration of both the amine and the CO₂. Subsequently, the amine/CO₂ equilibrium was defined by the equilibrium that existed between the $p\text{CO}_2$ above a well defined solution having a specific temperature - Equation (10)

$$p\text{CO}_2 = f\left(\sum \text{Conc.}_{\text{CO}_2}, \sum \text{Conc.}_{\text{MEA}}, T\right) \quad (10)$$

Where f is the fugacity. Equation (8) is applied as an underlining principle in the CO₂ absorption process of the Nigerian associated natural gas. The application of this equation is validated by the plant operational system in which the plant data are directly recorded during the absorption process. In practice/plant operations, the $p\text{CO}_2$ are the measurement data that are traditionally used to identify the amine/CO₂ equilibrium state. This is done by recording the $p\text{CO}_2$ at a particular temperature where the summation of the concentrations of the CO₂ and the amine are known. The direct interpretation is that, the minor non-ideal problems are solved numerically and the data from the physical plant can be used to validate the simulated work.

2.2. General chemical equilibrium models

The expression of the partial pressure in Equation (8) fits the plant data empirically which is regarded as the plant model. This procedure is important because there may be models which do not give exact calculations of the component in the liquid phase.

By using the water/carbonate/bicarbonate system [43], have evaluated both Henry's constant and equilibrium constant. Their reference procedure was based on the findings of [56]. The simulation in Aspen Hysys is a run of a model obtained by modifying their equilibrium constant [44] for the amine/carbamate equilibrium and the amine/-protonated amine which were fitted for plant data.

Activity-based model in Aspen Hysys uses the chemical potentials of both of the molecular and ionic constituent. These components are functions of their concentrations in the liquid bearing systems of such component. The absorption procedure is equally affected by the temperature of the medium.

In this work, we have used the plant data available to fix the parameters for the aqueous MEA/CO₂ system from the equilibrium model available in Aspen Hysys to enable absorption to be more accurate [57]. Margules property package was selected. There are however other available activity model in Aspen which are Wilson, van Laar, UNiVersal-QUAsi-Chemical (UNIQUAC) model, NRTL, general-NRTL, extended-NRTL and Chien Null's model-NRTL. Noting that the evaluation of the ionic components are within the subroutines in Aspen Hysys [41,57], also noting that the model [58,59] applies as procedure in this absorption, the absorption of CO₂ into the amine phase was evaluated as the summation of the concentrations of the constituent which are written as $\sum \text{Conc.}_{\text{CO}_2}$ and $\sum \text{Conc.}_{\text{MEA}}$ at specified temperature. Following the theoretical background from the basic absorption processes from Equations (1)–(9), the absorption of CO₂ by MEA was simulated in ASPEN Hysys and the convergence was appropriate for the available plant data. The simulation was repeated for sensitivity analysis where the simulation was run at five different temperatures and five varying circulation rates. The results enabled the evaluation of the reboiler duty and the compilation of the parameter-guideline for operation of the absorption column in the industry.

2.3. Adoption of the modelled absorption process for associated natural gas

The removal of CO₂ from a Nigerian associated natural gas was carried out to meet the standard requirement of CO₂ in cooking gas (5000 ppm). The gas composition was obtained from AGIP. The gas is to be stripped of CO₂ using existing plant. The plant has been designed for operation in a local community in Nigeria. The plant is feed with MEA having a column unit which consist of 10 trays. The associated natural

gas is usually passed through the bottom of the absorber. In this column, the natural gas mixture were counter-currently contacted with the MEA where the chemical reaction occurs. The MEA bearing the CO₂ are output of the bottom of the absorption unit next to the preheater which uses heat exchanger. This output usually move to the stripper where MEA is regenerated and are recycled to the absorption column. It is required to evaluate this process.

2.4. Methodology and development of ASPEN model

A steady-state condition was developed in Aspen HYSYS V9 (Iservrc). By using the new technology in Aspen HYSYS®, the column was analyzed. The test operating parameters was improved, yielding a significant improvement in saving energy, throughput and product quality. The aim is to reduce the amount of CO₂ in the associated natural gas to the specification required in cooking gas. The base case operating condition was forced to initiate the process so that about 1 % of error rendering the model valid. In order to maintain several numerical run for research purposes, a rigorous absorption process was simulated.

2.5. Description of the absorption process

A process flowsheet of the amine sweetening unit is presented in Fig. 1. Raw natural gas (Table 1) flows into the system and is cleaned of free fluids to avoid the passage of impurities by an inlet separator. The gas from the separator flows into the absorber through the bottom stream. Here, the gas flows to the top counter-currently contacting the MEA, which descends from the top of the amine absorber. An exothermic reaction occurs which raises the temperature of the gas. This reaction enriches the amine solution which contains the absorbed acid gases and it flows out of the column through the bottom whereas, the treated gas exits from the top of the absorption column. The enriched amine passes through a pump where the pressure is raised from $1.1 \times 10^2 \text{ kPa}$ to $2 \times 10^2 \text{ kPa}$ followed by a flow into a heater which elevates the temperature from 40 °C to 60 °C, heated further in the heat exchanger to 78 °C for regeneration. Subsequently, the amine flows counter currently against the gases. The acid gases flows out as the top product stream and stored in a tank. The amine flows out through the bottom and cooled from 123.5 °C to 101.8 °C. The lean amine flows from the heat exchanger into a cooler that reduces the temperature by 61.8 °C. The lean amine which flows from the cooler with a temperature of 40 °C enters the make up where water and amine are mixed with the initial lean amine. The outlet stream is fed into a recycle for the MEA.

2.6. Input process parameter for CO2 removal

The CO₂ content in the associated natural gas is 2.59 %, this constitute the CO₂ feed into the simulation. The other components were added according to the corresponding proportion presented in Table 1. The specification for input data are provided in Table 2. Noting that the capacity of the physical plant is 5153 m³/h or 7.779×10^5 barrel/day, a plant design with this capacity was developed in the ASPEN Hysys. The associated natural gas was made to enter the design at 40 °C, $1.1 \times 10^2 \text{ kPa}$ pressure through the bottom of the 10-plate-absorber. A 29 wt% concentration of MEA was made up with water (Table 2), having a temperature higher than the temperature of the MEA-CO₂ solution. The process flow was first made to converge the absorption of CO₂, next the regeneration of MEA occurred and finally the MEA was recycled back to the absorption column. The actual gas composition was recalculated. In the makeup unit, the operation was adjusted so that the circulation rate of lean amine was adjusted (Fig. 1).

2.7. Available absorber and technique

The absorber of the physical plant was sized in Aspen Hysys V9. The process parameter was specified (Table 3). The column diameter was

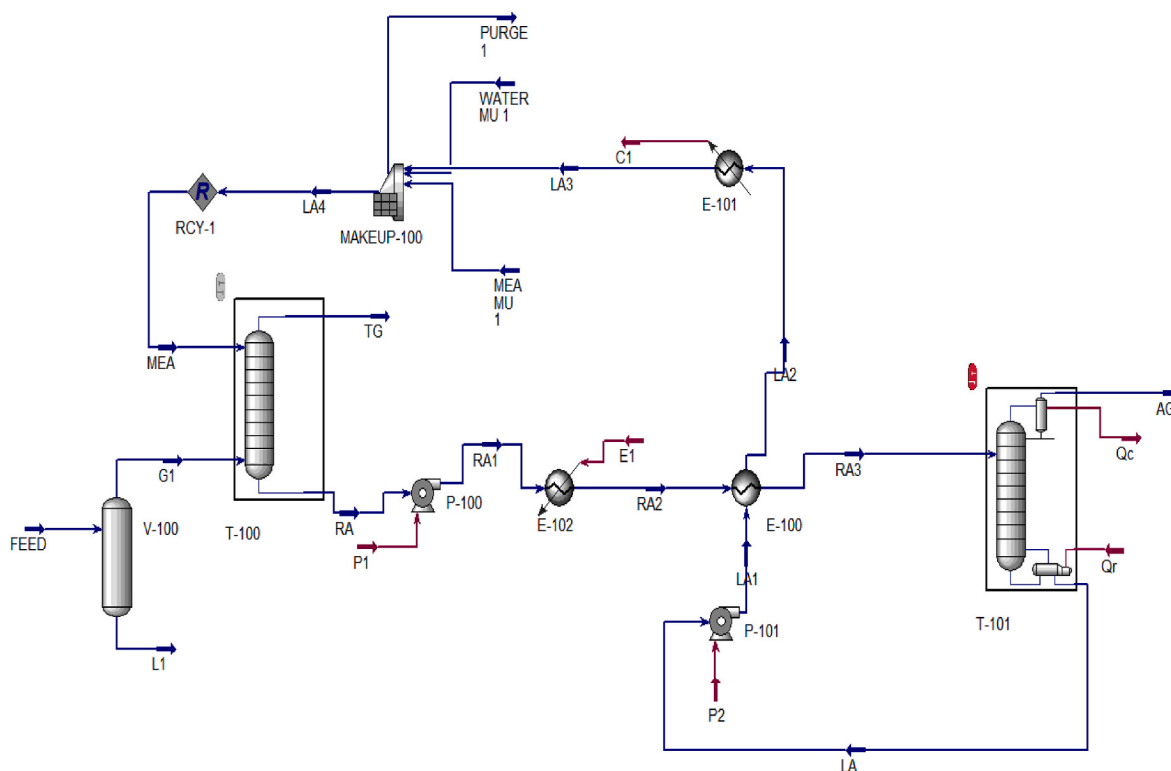


Fig. 1. Process flowsheet of the removal of CO₂ and stripper unit including auxiliary units.

Table 1

Composition of raw associated natural gas (source: AGIP, Nigeria).

Constituents	Chemical Formula	Volumetric fractions
Methane	CH ₄	0.7881
Ethane	C ₂ H ₆	0.1046
Propane	C ₃ H ₈	0.0462
Iso-Butane	C ₄ H ₁₀	0.0079
N-Butane	C ₄ H ₁₀	0.0097
Iso-Pentane	C ₅ H ₁₂	0.0031
N-Pentane	C ₅ H ₁₂	0.0027
N-Hexane	C ₆ H ₁₄	0.0021
N- Heptane	C ₇ H ₁₆	0.0010
Carbon dioxide	CO ₂	0.0259
Nitrogen	N ₂	0.0061
Water	H ₂ O	0.0026
Hydrogen sulphide	H ₂ S	0

1.5 m (with 10 trays and 0.6 m tray space). The equilibrium and kinetic reactions which were obtained were solved with the rigorous calculations for the removal of CO₂ using the acid removal package. The calculation [24] was an optimum reduction of CO₂ content in the natural gas where values in Tables 3 and 4 were manipulative and the returning results rely on thermodynamic model which are non-random of the liquid electrolyte system [59]. Also by taking advantage of the procedure given by Ref. [60]. The configuration in Hysys V9 was used to simulate the extent of removal for values as high as 0.45 mol of CO₂/mole of amine where the assumption of 1/100 of CO₂ loading was achieved by altering the duty specification. Therefore, by using a rate-based technique, the rigorous calculations were used for the removal of CO₂ using MEA which advanced the modeling process. In this technique, an adequate representation of acid gas removal was simulated for the physical plant.

The manipulated circulation rate of 6000kg.mol/h to 16,000kg.mol/h was used, this is because the least value of circulation rate that could run the physical plant which serves as constraint to the evaluation of reboiler duty if the grade of the CO₂ retained must meet specifications.

Table 2

Recorded specification- Average Daily Plant data.

Variables	Value
Absorber pressure	210kPa
Absorber temperature	40 °C
No. of stages (Absorber)	10
Lean amine temperature	60 °C
Ean amine pressure	211kPa
Lean amine flow rate	16,000kg.mol/h
Amine to water ratio	7.0:3.0
Gas temperature	50 °C
Gas pressure	310kPa
Gas flow rate	95,000 kg mol/h
No. of stages (Stripper)	20
Stripper pressure	220.6 kPa
Stripper inlet stream temperature	98.89 °C
Reboiler temperature	365 °C
Condenser tempertaure	333.91 °C
Reboiler duty	2.73 × 10 ⁸ W
Parameter for plant sizing in ASPEN HYSYS	
Gas density	17g/mol
Molar flow rate	25,000kg.mol/h
Space velocity	1 h ⁻¹
L/D ratio	4.5
Amine solution	29 wt%

Next, the number of stages were varied from 6 to 10, the absorber temperatures were varied between 20 and 60 °C, the absorber pressure was varied from 1×10^2 kPa to 1.8×10^2 kPa, the temperature of the reboiler were varied from 119.5 °C to 123.5 °C so that a less energy can run the physical plant. The energy required for regenerating the MEA was minimize correspondingly. Table 4 presents the variables that were used.

The temperature and pressure values are reported in Table 4. Table 5 are the referenced value and the optimized value attained preceding the data obtained for the experiment done in Aspen Hysys. The specification for the first set of data plotted are presented in Table 6.

Table 3

Input data used for simulation of the physical plant.

Description of variables	Value
Absorber pressure	100 kPa
Absorber temperature	40 °C
Lean amine temperature	40 °C
Lean amine flow rate	6000 kg mol/h
natural gas temperature	40 °C
Amine to water ratio	2.9:7.1%mol/%mol
Stripper pressure	150kPa
Stripper inlet stream temperature	48.89 °C
Heat duty of reboiler	1.95×10^9 W

Table 4

Manipulative data used for simulation of the physical plant.

Measured variables (output)	Varied variables (input)	Range of Variation (unit)
Objective 1: (absorber unit)	Lean amine flow rate	6000–16,000 kg mol/h
Acid gas (CO ₂) removal rate.	natural gas temperature	10–30 °C
	Lean amine temperature	15–35 °C
	Absorber pressure	100–180kPa
Objective 2: (stripper unit)	Stripper pressure	150kPa
Amine regeneration rate	Stripper temperature	125 °C

Table 5

Existing physical plant value and Value obtained for improved process parameters.

Varied variables	Existing physical plant value	Value obtained for improved process
Lean amine flow rate	160,000kg.mol/h	120,000 kg mol/h
Amine to water ratio	7:3	1.5:8.5
Lean amine temperature	60 °C	35 °C
Absorber pressure	690kPa	110 kPa
Stripper pressure	209kPa	100 kPa
Stripper temperature	187 °C	123 °C
Reboiler Duty	3.665e+006hp	1.20×10^9 W

Table 6

Specification returned after simulation.

Variables	Value
Absorber pressure	100kPa
Absorber temperature	35 °C
No. of stages (Absorber)	10
Lean amine temperature	35 °C
Lean amine pressure	150kPa
Lean amine flow rate	120,000 kg mol/h
Amine to water ratio	1.5:8.5
Gas temperature	25 °C
Gas pressure	100kPa
Gas flow rate	68 kg mol/h
No. of stages (Stripper)	10
Stripper pressure	150kPa
Stripper inlet stream temperature	68 °C
Reboiler temperature	123 °C
Condenser temperature	111 °C
Reboiler duty	1.611×10^5 hp

3. Results

The results presented are extracted from the inbuilt program with the specification of the physical plant. Here, the grade of CO₂ retained in the natural gas are provided in all the plotted curves. This first set of data are those of the reboiler duty and the circulation rate required to which the number of trays in the column are significant.

The feed data were plotted first at 6,000kg.mol/h for 5.89×10^8 W to observe the onset of absorption. This data is included in Fig. 2 which shows that the gradual increase in the amine circulation rate from 6000 to 16, 000kg.mol/h can be used to remove 22.78 % of CO₂ hence 5.89×10^8 W to 1.62×10^9 W are values of power required for the physical reboiler duty. This is based on the available capability of the physical reboiler duty. At 16, 000kg.mol/h, the percentage absorption is about 22.78 % (77.22 % of CO₂ grade retained) which brings the amount of CO₂ remaining in the natural gas to 0.0059 from 0.0259. This grade of CO₂ retained is within the specification of the contractor. Any circulation rate above 12,000kg.mol/h (80.31 % of CO₂ grade retained) is probably not necessary because the removal grade at a higher circulation rate is no longer significant to the specification required by the contractor. Although absorption is steady up to the 22.78 % removed, the reboiler duty increases significantly and the grade of CO₂ is even lesser. Therefore it is adequate to work at amine circulation rate of 6000 to 12,000kg.mol/h based on the composition of natural gas (Table 1) and input specification (which are listed in Table 2).

The higher the circulation rate, the better the grade of the CO₂ retained. The five different run converges which implies that; the available physical plant must be energized with a significant reboiler duty to use 1.2×10^9 W of power to obtain 80.31 % of CO₂ grade retained. These values are for 10 staged-column. However, if the number of stages are varied the absorption are affected and attaining convergence could be challenging.

For the data in Fig. 3, the reboiler duty was brought to 1.2×10^9 W for the 10 stages, 19.69 % of the CO₂ remained in the natural gas. That is, a CO₂ grade of 80.31 % is retained. Achieving this brings the amount of CO₂ remaining in the natural gas to 0.005, this implies that, the energy requirement is reduced. Using 1.2×10^9 W for stages less than the 10stages, a further reduction in the CO₂ content of the natural gas was attained in the absorber by fixing value for all the parameter in Tables 1 and 2, the circulation rate was at 12,000kg.mol/h, the number of stages were varied from 6 to 10 in order to obtain lesser reboiler duty which minimize the operational cost. While the lower number of stages yielded better grades of CO₂ retained, it was clear that the reboiler duty is the same value at lower values of the number of stages. In practice when running the physical plant after several recirculation of amine, the grade of CO₂ are low for less number of plates, the 9 stages gave the least reboiler duty, so that the CO₂ removed is 18.15 % (81.85 % grade of CO₂ retained) at 1.2×10^9 W. Therefore to work at a reboiler duty of 1.2×10^9 W is to minimize cost in an extended plant operation. It is important to note that higher number of stages will require a high column height which will add to the capital cost. For this composition of natural gas, the stage efficiency is maintained at 0.25 (Murphree stage efficiency) while the column height can also be varied and as the stage efficiency can be changed substantially.

Clearly, the removal grade increases after increasing the number of stages beyond 9 but the reboiler duty increases significantly. Therefore the calculation of reboiler duty results into a high divergence if more than 10 stages are in the column. The Aspen Hysys V9 calculates efficiency using 25 % in equation (6) and the calculated plate efficiency is from 0.08 to 0.14.

The temperature carried by the feed into the absorber, influences the grade of the CO₂ removal which is better off because the high temperature reduces the heat requirement with higher column height. The progression of the simulation is however constraint to less than 12 stages because only a small percentage of CO₂ is remaining in the natural gas even at the ninth and tenth stages. Beyond the twelfth stage, the simulation diverges due to the relatively small amount of CO₂ remaining in the natural gas. Hysys calculations were done for the estimated 25 % stage efficiency too.

The specification of the contractor was attained as absorption is at a maintained equilibrium, when the feed temperature was within 20 °C–60 °C (Fig. 4), the absorption grade slightly altered largely due to increased rate of reaction, even with a constant stage efficiency

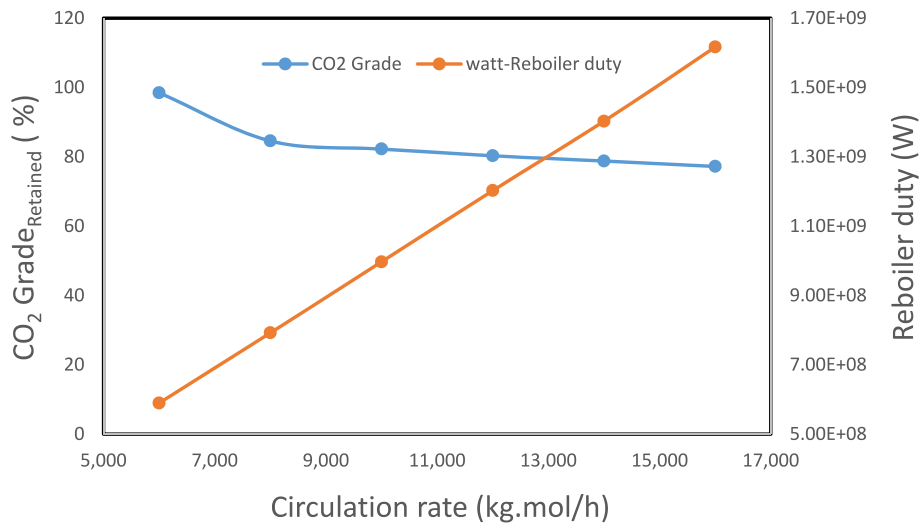


Fig. 2. Amine circulation rate of 6000 to 16, 000kg.mol/h and reboiler duty.

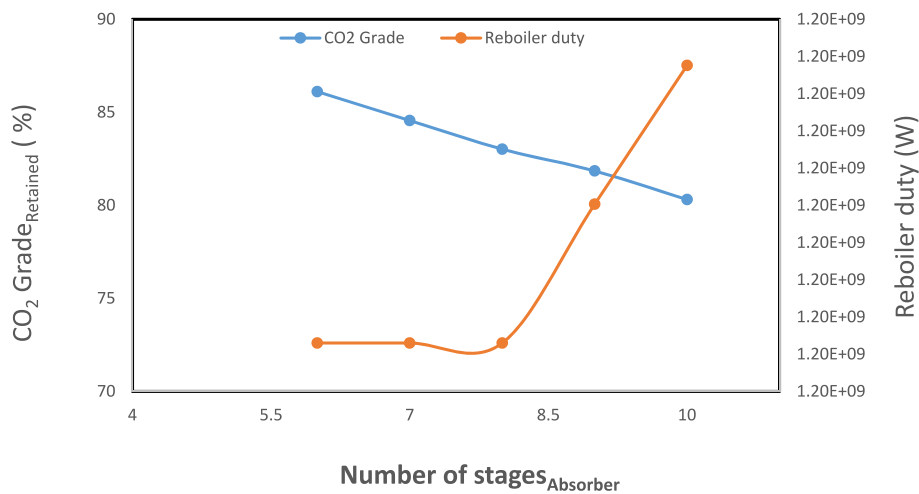


Fig. 3. Values obtained for stages less than 10 and reboiler duty at 1.61E+06hp.

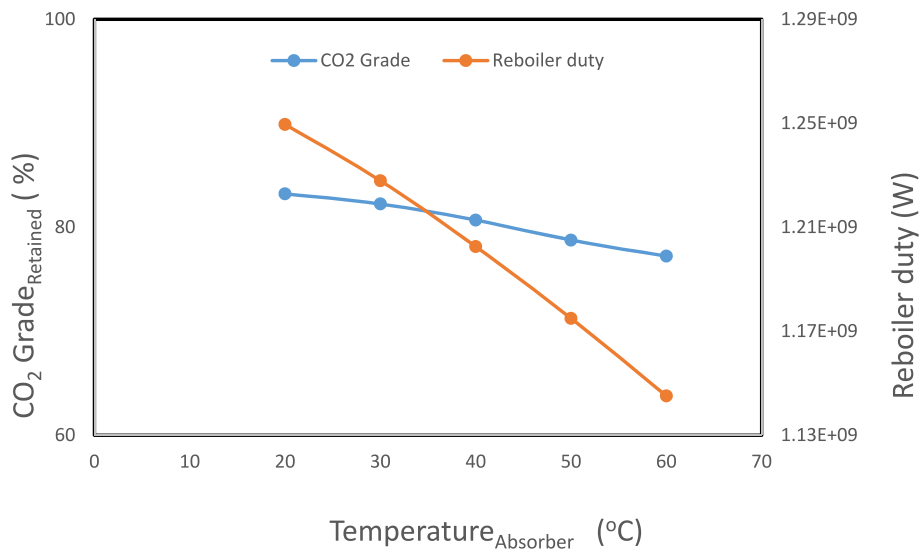


Fig. 4. Data for Reboiler duty and corresponding temperature carried by the feed in the absorber.

maintained for both the gas and liquid feed (Fig. 4) the reboiler duty of 1.2×10^9 W with which 19.31 % CO₂ was absorbed is at 40 °C of the feed temperature yielding about 80.69 % grade of CO₂ retained. Any value of the reboiler duty higher than this value will only yield insignificant increase in the absorption of CO₂ because of the relatively small amount that is retained.

The intersection point occurs where the reboiler duty is 1.22×10^9 W at 1.6×10^2 kPa (Fig. 5) this is where the grade of CO₂ is 77.22 %. Below the point of intersection the grade of CO₂ reduced significantly. Additionally a high value of pressure and reboiler duty is needed because of the high pressure in the absorber. Therefore, 100kPa value of pressure is maintained for the outlet of absorber with a maximum pressure drop of 0.6×10^2 kPa at the inlet so that the grade of CO₂ is maintained and up to 81.08 % of CO₂ grade is retained thus reducing the energy back to 1.19×10^9 W.

The reboiler duty is 1.19×10^9 W at 1×10^2 kPa with a high grade of CO₂ retained up to 81.08 % (in Fig. 5) however in Fig. 6, the reboiler duty is 1.04×10^9 W giving an unstable response of CO₂ grade which is insignificantly changing with temperature when the temperature exceeds 120 °C. The absorption decrease with increased temperature of the reboiler. Although the reboiler duty reduces the grade of the CO₂, renders it unstable so that an increased temperature beyond 120 °C is not adequate for the absorption process. (with this large margin in the reboiler duty only a little change in absorption was observed) this means the shift in the CO₂ absorption of 80.3 %–80.7 % is erratic if the process is repeated at these conditions, the amine absorption will most likely degrade. Thus at about 122.5 °C there was no convergence in the simulation because amine degrades and does not work well at above 120 °C. Which are responsible for most challenges with absorption processes involving amine. The gradual rise in temperature from a low value up to 120 °C yields purer amine and better CO₂ grade. However, lower values of power to the reboiler means only less amount of CO₂ will be absorbed for a prolonged operating temperature which is not cost effective.

The stripper pressure was specified to 2×10^2 kPa (a) in the calculation this is due to the difficulty to get a converged solution at other pressures having a solution warning (the warning indicator reads “outside range”).

4. Discussion

The use of the Kent-Eisenberg model was limited when it was applied for the removal of CO₂ using DEA for the data specification of the

physical plant in this work. The constrained results are presented in Figs. 1–6. The data specification are presented in Tables 1–6 This finding is in agreement with the work of [61]. Therefore MEA was employed for the absorption process so that the specification required by the contractors can be obtained.

Further, there is cost effectiveness by reduction of heat requirement [62–64] while adhering to the adequate reboiler temperature of about 120 °C: Thus the simulation involving the variation of the circulation rate, the number of stages, the reboiler temperature, the absorber temperature and pressure was targeted at reducing the heat required to the barest minimum. A minimization program [65,66] could be cumbersome to accommodate the list of specifications and complicated operational procedure involving the physical plant. Hysys was made to simulate optimally the retained restriction of the physical plant and each unit. Thus the results from the evaluation of heat required as per the CO₂ grade retained is 4MW/kg of CO₂ removal. With this calculated value, 0.25 corresponds to the value for the efficiency of the CO₂ absorption where the removal is about 22.75 %. This simulation reduced energy requirement at an absorber temperature of 35 °C, with 10 plates in the absorption column. The pressure of the stripper is 1.0×10^2 kPa. The temperature of the reboiler is 123 °C. It is obvious that, these set of values were adequate to minimize the reboiler duty and can serve as a parameter guideline.

Although the CO₂ removal is about 80 % (Fig. 2), and the grade retained increased (Fig. 2), the number of plates were changed to vary the column height so that the efficiency of the stage is varied, it was observed that the heat duty was reduced and the retained grade increased with a stable amine in the column throughout the process (Figs. 2 and 3), beyond 11 and up to 12 stages there was no convergence in the simulation because of the low amount of CO₂ in the feed. The simulation proceeds with the calculation of efficiency for each plate which was found between 0.08 and 0.13.

4.1. Heating source to the reboiler

In the physical plant, the heat source is directed to the reboiler which generates quantifiable energy; the simulation was done to reduce the overall electric power supply to the reboiler and hence influence the efficiency, therefore heating at a pressure of 1×10^2 kPa and up to 1.6×10^2 kPa was generated to obtain 123 °C in the absorber. The heat that would be supplied to the physical plant are under these conditions of temperature and pressure which makes it feasible to optimized the absorption process expressed at the stated stage efficiency. With the

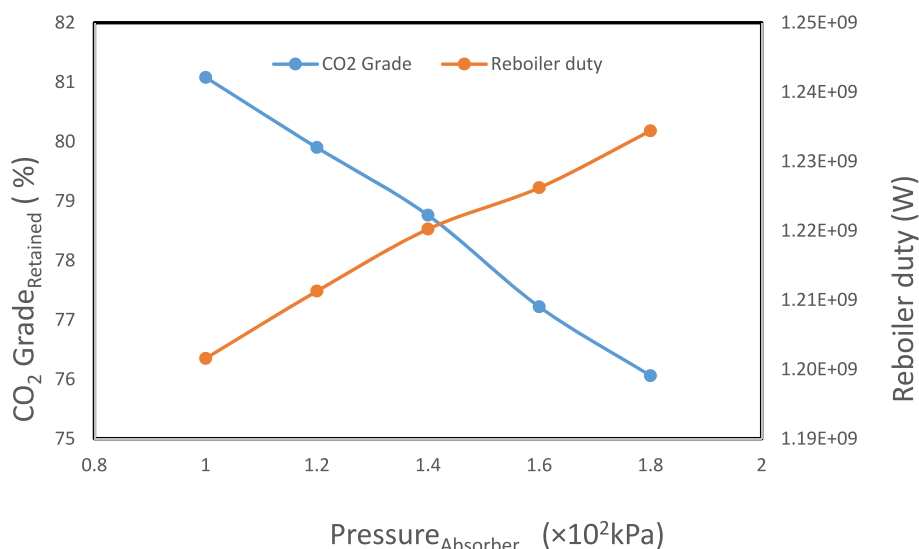


Fig. 5. Absorption pressure in column and reboiler duty.

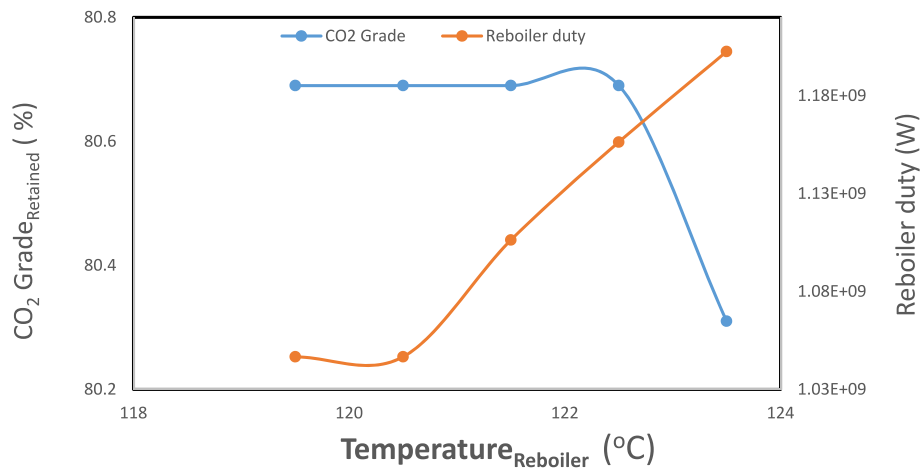


Fig. 6. Temperature controlled reboiler duty at 1.19×10^9 W at 100 kPa and grade of CO₂ retained.

addition of the value of the stage efficiency the total thermal efficiency improved compared to the previous plant operation.

By considering the delivered circulation rate of the physical circulation pump, an overall efficiency is lowered in the plant. An additional process outside the absorption desorption units included the compression of the CO₂ which was energized by combustion of methane gas and the total efficiency drops to about 50 % this brings the evaluated heat duty to 3.8.MJ/kg of CO₂ removed. The temperature of the stripper can be lowered to reduce heat duty, the stripping pressure can be medium and further reduced. The effect can be studied on the overall heat requirement if the calculation is able to converge because it may be marginal.

Based on the same initial data, a small change in variables yield a slight difference in the amount of CO₂ removed at an accuracy lower than that which was obtained in this research. However a better accuracy can be achieved when a tight convergence limit is feasible. Also, due to the absorption equilibrium, the uncertainty may be high. This is because CO₂ removed increased when the equilibrium changed to Li-Mather model from Kent Eisenberg model. The uncertainty in amine models is less compared to Peng Robinsin in this calculation.

4.2. Analysis of the absorption process

The evaluation of the process included neglecting the heat loss and pressure drops due to the flow of materials from one equipment to another, this is common in ASPEN simulations eg Ref. [67]. The link between the equipment are the auxiliary units of pipes, valves, and pumps. This heat and pressure values are also encountered in the physical plant and the water wash unit is included to reduce the flow of MEA which precedes the reclaimed section for the recovery of the degraded MEA. These physical units contributes to the amount of energy consumed by the process. Therefore the idea of conserving the heat consumed by configuring the stream efficiently is important if the overall energy requirement will be brought to a low value, although an initial additional capital may be necessary. In this data combination discussed in the former, the heat consumption is estimated to be 3.8. MJ/kg the value is comparable to the work done by Refs. [41,68]. An optimum value calculated is also with the range of the estimated value in similar models [41]. Perhaps the value can be regarded as an indication of an improved version of the simulations carried out.

4.3. Challenges in the absorption of CO₂ into MEA

There are less challenges associated with convergence, although in this research, the stripping section were not easily converged. The section required properly specified temperature, stage efficiency and the

number of stages before convergence is obtained. Also it was observed that Modified Hysim Inside out Algorithm with adaptive damping was employed and the best convergence was obtained. However above 12plates, simulated processes diverged which was the drawback within the range of all specifications. Equally it is challenging to device an optimization tool which will accommodate unforeseen operational challenges that are physical. By using Kent Eisenberg model to attain convergence the percentage of MEA charged was less than 30 % with a corresponding temperature of about 120.5 °C.

4.4. Number of stages, circulation rate, absorption temperatures and pressure

The effect of the three parameter have been presented in Figs. 2–6 hence the parameter variations are pathway to obtaining an improved process stream. With the recirculation rate varied within 6000 to 16, 000kg.mol/h. The process was improved when the amount of heat consumed is about 3.8.MJ/CO₂ removed (Fig. 7). The detailed calculation of energy are expressed in Ref. [69]. This value is the minimum heat consumption for the process (the corresponding lean amine circulation rate is 120,000kg.mol/h. The physical plant is not a software, so the plant will work even with increased number of plates, yielding a better quality of CO₂ removed. This implies that more heat will be consumed. But in the simulation package, the high temperature diverged the simulation, so that the convergence of the stripper is negatively affected. At lower temperatures however, the CO₂ removal process occurred, while in the simulation, 40 °C is ideal for the absorption of CO₂ to proceed which means that, at temperatures lower than 40 °C the reaction rate was low (Fig. 4). Noting that the evaluation of efficiency in Aspen Hysys V9 uses a temperature dependent function, the optimum absorption is realistic in the physical plant with the inclusion of temperature dependent stage efficiency. With these conditions established in the absorption process, it is easier to evaluate the absorption process.

A gradual and low heating makes the solution of the amine to become purer. Next, the optimum temperature is reached by the improvement in CO₂ removal which invariably is the used amine. At close to 120.5 °C the operation of the stripper is restricted and the temperature in the column cannot increase. Therefore the amount of CO₂ removed for each mole of amine loading is determinable and can be used for cost estimation. The result here is in agreement with the work of Oi done in 2007. However at the same temperature, the pressure in the stripper has to be maintained at $1 \pm 0.8 \times 10^2$ kPa for the stripper to converge. This is also in agreement with the procedure presented by Ref. [70]. If limitations are encountered, it could be due to the input data obtained from the physical plant because it is necessary for the pressure of desorption to be uniform all through the stripper up to the exit of the

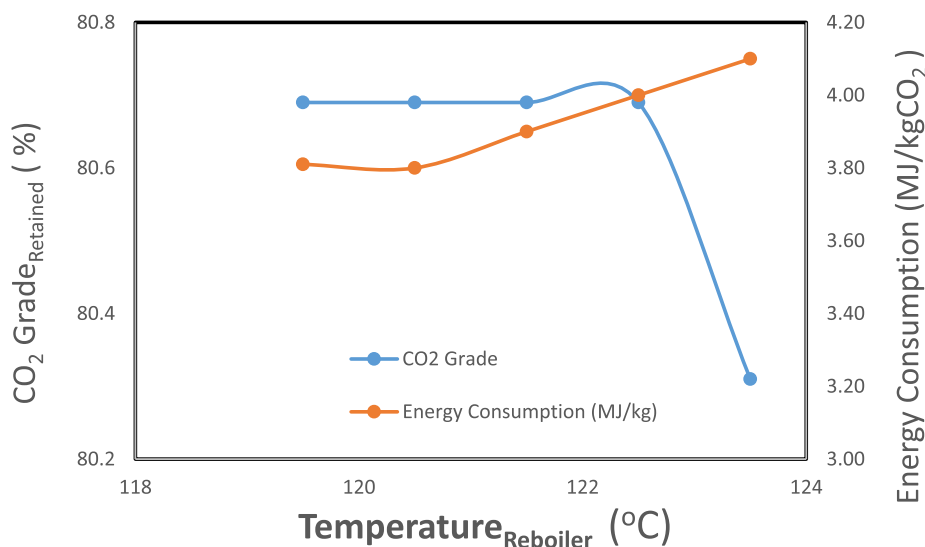


Fig. 7. Energy consumption calculated for the physical plant and the grade of CO₂ retained.

stripper. In a case where it is feasible to calculate the reaction rate of the CO₂ removed without the use of special software packages (e.g. Aspen plus), the removal of CO₂ may be given an alternative evaluation. That is; both software (e.g. ProMax and ProVision) and theoretical calculation can be used for the simulation of CO₂ removal [41,71]. The result can be compared and conclusions can be made on absorption model. In the same vein, a mixing stage is important and an equilibrium model should be a robust one.

It should be noted that the stage efficiency in the first stage of the process (absorption) is not easily achieved when a physical plant is a case study. MEA was chosen as a preferred absorbent because other amine solution have reported challenges with the parameters of equilibrium models for associated natural gas having a small amount of starting CO₂, for instance mixed amine require careful consideration of reaction rate and the results are expected from a rate based model. An example of a rate based model is RatFrac.

5. Conclusions

The grade of CO₂ retained in the natural gas can be determined at any stage of the simulation process. By fixing the specification of the physical plant for simulation, an optimum CO₂ removal is attained. A better process flow sheet can also be created because the quality of CO₂ removal is based on the processing conation which is controlled by the physical plant. All parameters can therefore be altered until the optimum process is attained to suit the targeted plant. Therefore the response factor are highly dependent on the circulation rate, amine concentrations, pressure and temperatures at the absorption stage which are all dependent on the reboiler duty.

For the grade of the CO₂ retained, the manipulation of the simulated work can enhance the removal of CO₂ using the amine solution. This is applicable for reduction in energy requirement thus, there is a direct link

between stripper pressure and temperature from the reboiler.

The Murphery efficiency can be calculated using physical plant data and the efficiency is significant in the simulated work. The amount of energy consumed is directly connected to the required pressure in the absorption column. Usually the energy consumed in this unit can be reduced by some MJ per kg of CO₂ removed. This require a fixed absorption pressure to an atmospheric pressure value including only a fraction of one in the value of pressure drop for inlet while the atmospheric pressure is fixed for the outlet which correspond to the reduction in energy requirement.

Legend

CRediT authorship contribution statement

Adeola Grace Olugbenga: Writing – review & editing, Writing – original draft, Visualization, Validation, Supervision, Project administration, Methodology, Investigation, Funding acquisition, Formal analysis, Data curation, Conceptualization.

Declaration of competing interest

The author declare that she has no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

Data will be made available on request.

Abbreviations

(E _o)	ratio of the equilibrium number of plate to the actual number of plate
NEQ	equilibrium number of plate
N _{actual}	actual number of plate
Where E _{ij} ^{MV}	Murphery efficiency
'i'	each component
j	each stage
y _{ij} [*]	composition of vapor in equilibrium for the liquid leaving the tray

- FEED - Natural gas (To Separator V-100).
Temperature: 40°C, Pressure: 1.1×10^2 kPa, Molar flow: $8.500e+004$ kgmol/hr.
 - G1 - Natural gas, (From Separator V-100 to Absorber T-100).
Temperature: 40°C, Pressure: 1.1×10^2 kPa, Molar flow: $8.500e+004$ kgmol/hr.
 - L1 - Entrained liquid - Entrainment solid and liquid, (From Separator V-100).
Temperature: 40°C, Pressure: 1.1×10^2 kPa, Molar flow: 0
 - MEA – Mono ethanolamine (To Absorber T-100)
Temperature: 43°C, Pressure: 1.1×10^2 kPa, Molar flow: $1.200e+005$ kgmol/hr.
 - TG – Treated gas, Stream 1 (From Absorber to Separator)
Temperature: 40.60°C, Pressure: 1.1×10^2 kPa, Molar flow: $8.987e+004$ kgmol/hr.
 - RA - Rich Mono ethanolamine, Stream 1 (From Absorber T-100 to Pump)
Temperature: 23.99°C, Pressure: 1.1×10^2 kPa, Molar flow: $1.151e+005$ kgmol/hr.
 - RA1 – Rich Mono ethanolamine, Stream 2 (From Pump to Heater)
Temperature: 24°C, Pressure: 2×10^2 kPa, Molar flow: $1.151e+005$ kgmol/hr.
 - RA2 – Rich Mono ethanolamine, Stream 3 (From Heater to Heat Exchanger)
Temperature: 68°C, Pressure: 1.5×10^2 kPa, Molar flow: $1.151e+005$ kgmol/hr.
 - RA3 – Rich Mono ethanolamine, Stream 4 (From Heat Exchanger to Stripper)
Temperature: 78°C, Pressure: 1 bar, Molar flow: $1.15e+005$ kgmol/hr.
 - AG – Acid Gas (From Stripper)
Temperature: 111.3°C, Pressure: 1.5×10^2 kPa, Molar flow: $7.414e+004$ kgmol/hr.
 - Qc - Condenser energy (Energy from Stripper)
Heat flow: $8.461e+004$ Btu/h, Power: $3.325e+005$ hp
 - Qr – Reboiler energy (Energy to Stripper)
Heat flow: $4.096e+009$, Power: $1.610e+006$ hp
 - LA – Lean Mono ethanolamine, Stream 1 (From Stripper to Pump P-101)
Temperature: 123.5°C, Pressure: 1.5×10^2 kPa, Molar flow: $4.099e004$ kgmol/h.
 - LA1 – Lean Mono ethanolamine Stream 2 (From Pump to Heat Exchanger)
Temperature: 123.5°C, Pressure: 2×10^2 kPa, Molar flow: $4.099e004$ kgmol/hr
 - LA2 - Lean Mono ethanolamine, Stream 3 (From Heat Exchanger to Cooler)
Temperature: 101.8°C, Pressure: 1.8×10^2 kPa, Molar flow: $4.099e+004$ kgmol/hr.
 - LA3 – Lean Mono ethanolamine, Stream 4 (From Cooler to Makeup)
Temperature: 40°C, Pressure: 1.5×10^2 kPa, Molar flow: $4.099e+004$ kgmol/hr.
 - LA4 – Lean Mono ethanolamine, Stream 6 (From Makeup to Recycle)
Temperature: 43.84°C, Pressure: 1.5×10^2 kPa, Molar flow: $1.200e+005$ kgmol/hr.
 - WATER MU1 – Water makeup (To Makeup)
Temperature: 40°C, Pressure: 1.5×10^2 kPa, Molar flow: $7.901e+004$ kgmol/hr.
 - MEA MU1 – Methyl ethanolamine makeup (To Makeup)
Temperature: 40°C, Pressure: 1.5×10^2 kPa, Molar flow: 5.781 kgmol/hr.
 - PURGE – Water (From Makeup)
Temperature: 43.84°C, Pressure: 1.5 bar, Molar flow: 0
 - E1 – Energy Stream (For Heater)
Heat flow: $4.185e+008$, Power: $1.645e+006$ hp
- Equipment
- V-100 – Separator
 - T-100 – Absorber
 - P-100 – Pump
 - E-102 - Heater
 - E-100 – Heat Exchanger
 - T-101 – Stripper
 - P-101 – Pump
 - E-101 - Cooler
 - MAKEUP-100 – Makeup
 - RCY-1 – Recycle

k_2 temperature-dependent rate constant
 $\text{Conc.}_{\text{MEA}}$ concentrations of free MEA
 $\text{Conc.}_{\text{CO}_2}$ concentrations of free CO₂
 p_{CO_2} partial pressure of CO₂
 f fugacity
 $y_{i,j} - y_{i,j+1}$ difference in the composition of vapor in successive stage with the liquid leaving
 $E_{i,j}^{\text{MV}}$ Murphery efficiency at composition of vapor in equilibrium with the liquid leaving
 $\text{Conc.}_{\text{Carbamate}^-}$ concentration of the carbamate ion
 HMEA protonated MEA
 MEA Monoethanol amine
 $\text{Conc.}_{\text{HMEA}^-}$ concentration of the protonated MEA ion
 H_e Henry's constant

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