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# Environmental analysis of Acid Gas Sweetening with DEA

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***Abstract***
Natural gas has wide applications throughout many sectors, but it needs to comply with specific standards to ensure efficient utilization. Therefore, it is required to reduce the impurities in natural gas through the process of purification, also known as acid gas sweetening. Acid gas sweetening involves the usage of amines, which are basic compounds that bind with acidic components such as hydrogen sulfide and carbon dioxide, forming weak bonds. The acid gas sweetening process is based on several separation processes, where this report will mainly target the distillation column. In this report, a detailed study has been carried out on the installation column using process simulation Aspen HYSYS V10) and the manual steps of the solution column design using the McCabe-Thiele method. The results were then compared with each other, in which it has been found that the comparison is not fair at the slightest. This is because of the fact that the distillation column design on Aspen HYSYS was based on a multi-component system while the manual design calculations were based on the binary system. Therefore, a list of recommendations and advice were made to even out the comparison and improve the accuracy of the study. A safety assessment has been conducted on the process, in which the natural gas stream and the whole process has been analyzed. The severity of natural gas in the acid gas sweetening process has been deduced using the DOW index. The process units and equipment were then reviewed for safety considerations using the hazard operability (HAZOP) study, utilizing a qualitative analysis technique to highlight the main risks which can be endured during operations and their effects on the equipment and its surroundings, with actions which can be implemented on how to mitigate the risks of such events to come true. The actions are presented in modified P&IDs to visualize the control systems, alarms and valve systems which can be installed onto the equipment. The results showed a distillation column design requirements of 18 stages from the simulation model and the safety section reveals the severity of natural gas component, possessing a fire and explosion (F&EI) index of 165.1125. The material hazards and risks are mitigated through the HAZOp study guidelines and actions. Collectively, the design and safety considerations in this study provide huge potential economic and energy wise, as it has the potential to increase energy savings by $4.5*10^5$ kJ/h and economic profitability increase by 16%, proving that applying the guidelines mentioned in this study shall mitigate risks, reduce costs, increase profitability and improve environmental awareness.

***Keywords***: Acid gas sweetening; simulation; design; HAZOP; DOW.

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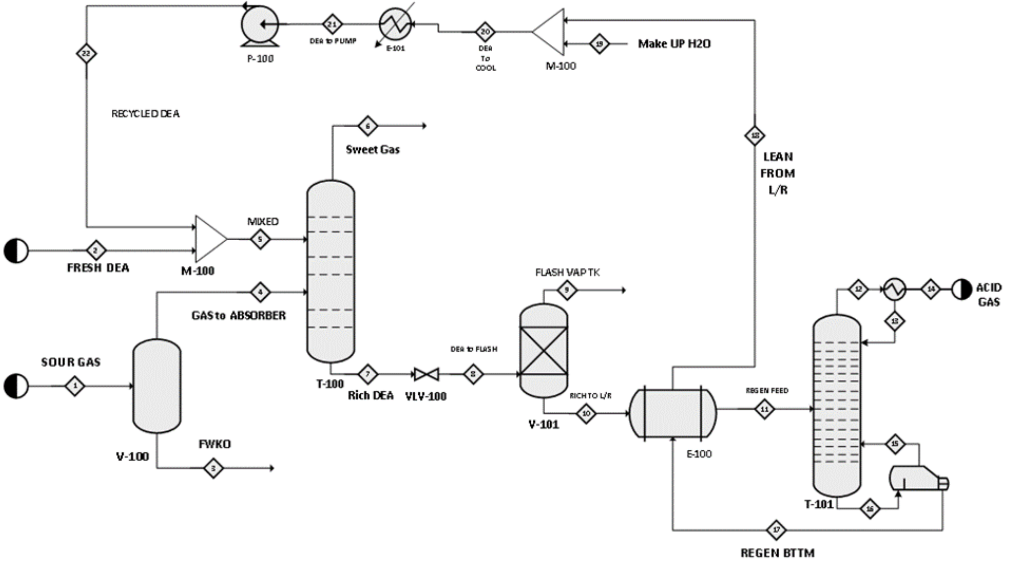
***Graphical Abstract***



## I. INTRODUCTION

Natural gas (NG) has been used for decades over wide applications [1]–[3]. It is a naturally occurring hydrocarbon whose main constituent is methane with traces of nitrogen, hydrogen sulfide and carbon dioxide [4]–[7]. As seen in ***figure 1***, natural gas is utilized in several sectors, such as power generation, industrial heating, heating, and cooking in the residential sector, commercial heating, chemical manufacturing, and can also be used as an exporting asset in the liquefied form alongside other uses, with the main contributor to power generation [8]–[12].

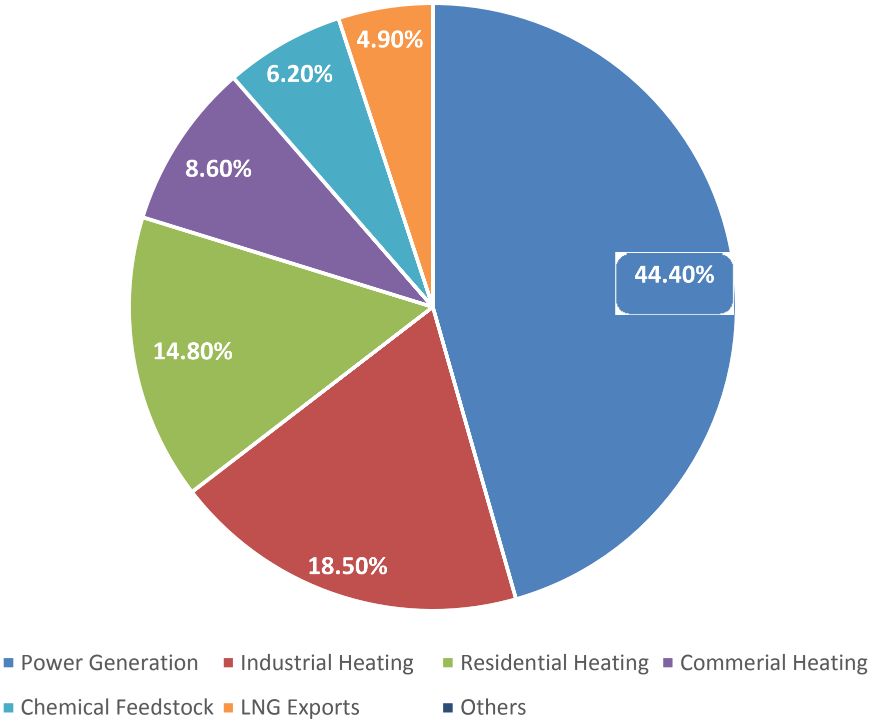


***Figure 1 Global consumption of natural gas.***

One must take into consideration that the natural gas fed into the infrastructure for any sector must meet certain requirements. Specific standards are set to standardize the natural gas product being pumped into homes, industries, and other sectors in a said region. As for the ISO 13686, natural gas quality must reach at least 70% methane (CH4), must not exceed 20% in carbon dioxide (CO2), and only small unnoticeable traces of hydrogen sulfide (H2S) are allowed [13]. The reason why these specific standards are set are based on various explanations. CO2 is preferably reduced for two main reasons; that is to mitigate corrosion rate of pipelines and equipment as well as improving the heating value of natural gas. In the case of liquifying natural gas (LNG), the crystallization of CO2 through the cryogenic process can be avoided through reducing the CO2 content prior to the liquefaction process [14]. As for the other acidic gas, hydrogen sulfide, it is a common acidic gas, with a powerful odor close to egg rot, which is usually released due to the metabolic activity of sulphate reducing bacteria who are anaerobic microorganisms who reduce sulphates and sulfur compounds found in crude oil into hydrogen sulfide [15]. Hydrogen sulfide needs to be handled, alongside its issues in the oil and gas exploration and production, it induces the corrosion of engines, boilers as well as reducing the calorific value of natural gas [16].

Consequently, the purification of natural gas is a critical step to achieve the required standards to use natural gas as efficiently as possible. Therefore, the extraction of high-quality natural gas can be accomplished through acid gas sweetening. Acid gas sweetening, also known as acid gas removal, is the process in which an amine is commonly used to remove acidic gases, most commonly H2S and CO2 [4], [8], [13]. One of the main processes required is distillation. Distillation is a separation process which separates a mixture of components based on their volatilities. There are several types of distillation, such as continuous, batch, azeotropic, extractive, and many more [17].

Fractional distillation is a prevalent method of distillation primarily employed for the separation of mixtures depending on their boiling points. The distillation column is equipped with a series of trays, and the selection of the number of trays is determined by design and calculations. The fractional distillation column is commonly employed in the process of petroleum fractionation. Reactive distillation is a process where separation and reaction take place at the same time. The column houses a catalyst that facilitates the reaction, occurring as the mixture passes through the column. It offers significant benefits since it integrates both response and separation into a single unit. Another critical type is azeotropic distillation which involves the addition of an "entrainer" to a combination. This is done when the components of the mixture have identical boiling temperatures at a certain composition, resulting in an azeotropic mixture. Consequently, an entrainer is introduced into the mixture to decrease the boiling point of one of the components, so disrupting the azeotropic behavior. This facilitates the separation of the components based on their boiling points. Additionally, Extractive distillation is a process similar to azeotropic distillation, except instead of employing an entrainer, a distinct solvent is introduced. The solvent is deemed selective, enabling it to remove a certain component from the mixture. Batch distillation involves the complete separation process taking place within a single vessel. The mixture is initially introduced into the column and subsequently heated by the boiler connected to the column. The vaporized stream that is produced ascends to the upper part of the column, where it encounters the condenser. The condenser then transforms the stream into liquid form, which is then collected in a drum vessel. This method can be iterated at regular intervals. Continuous distillation is the term used to describe the preceding forms of distillation, whereas batch distillation is referred to as a discontinuous distillation process. As for the regeneration stage in the acid gas sweetening process, it is most convenient to use the first type of distillation, which is fractional distillation [18]–[20]. Therefore, fractional distillation has been selected as the project's focus of study.

This report will forecast the study made on acid gas sweetening for natural gas purification. The methodology will describe the process description and the steps taken to perform the two main objectives of the whole study: the process simulation for the acid gas sweeting of natural gas and the manual design of the fractional distillation column. Moreover, the results of the simulation model and the manual design have been analyzed based on the final composition, number of stages in the distillation process and other parameters, where they are collected in material stream summary tables. Both tabulated results have been put into comparison to assess the accuracy and point out any deviations. In the final concluding step, all the aspects throughout the report will be summarized with a few recommendations to ensure more accurate results and decrease the margin of error and deviancy between the manual calculations and the simulation results. The process is then assessed from an environmental and safety perspective using the DOW index and the HAZOP study [21], [22], followed by P&ID modifications as per the actions required to conduct from the HAZOP study.

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## II. METHODOLOGY

The methodology pursued mainly depends on three aspects, which are the aim of conducting this study: the process design and safety analysis, as seen in ***figure 2***. Following this methodology, the simulation and manual design of the process as a whole and with a focus on distillation has been accomplished, alongside a detailed safety and environmental analysis using DOW index and HAZOP study. The first step was to understand and review the acid gas sweetening process, looking into previous works conducted previously, analyze points of strength and weakness as well as improving on previous assessments. The literature review led to the project to be divided into two main constituents, process design and safety analysis. Starting with process design, in which the process is first designed by constructing a process flow diagram (PFD) in which all the equipment, parameters and process description are mentioned. The process is then simulated using chemical process simulation software Aspen HYSYS V10, followed by manual calculation done on the mass balance and design of distillation calculations. The manual calculations have been carried out in order to assess the predictiveness and accuracy of the distillation column in the simulation in reference to the manual design, with recommendations and advice given on how to improve the process. The second stage is the safety analysis stage, in which two crucial techniques were used, DOW index and HAZOP study. The DOW index study was carried out on the natural gas component, calculating its fire and explosion index (F&EI) to assess the severity of the material within the process. The equipment safety was analyzed using HAZOP study, constructing tables for possible scenarios, the consequences and how to mitigate them. The mitigation measures are figuratively represented in modified P&ID diagrams, demonstration process control instrumentation, control loops, alarms and valve selection and placement.

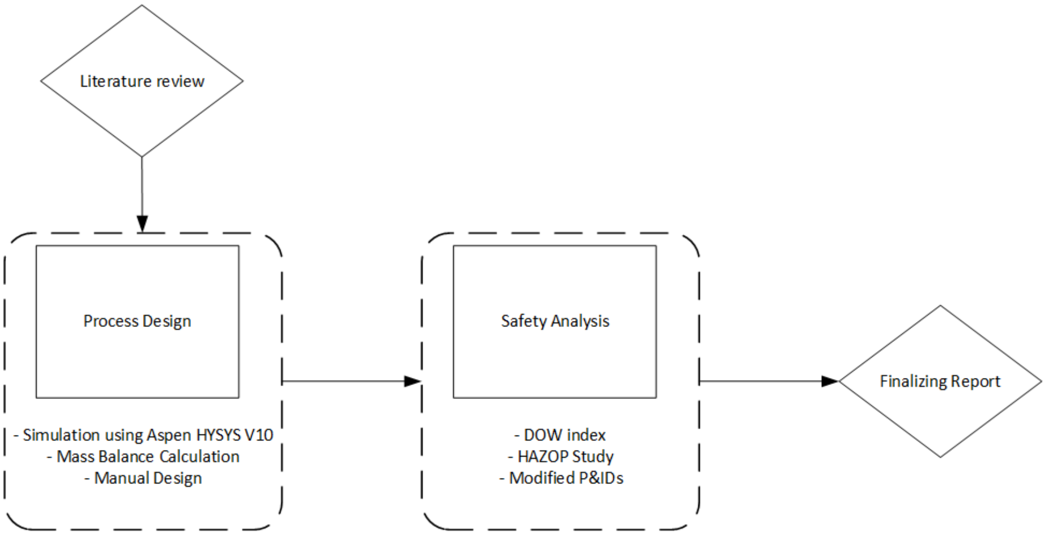

***Figure 2 Project methodology steps.***

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## III. PROCESS DESCRIPTION

Analyzing the information from ***figure 3***, The process consists of three main stages: contraction, regeneration, and recirculation. The first stage is the contraction, in which the natural gas comes in contact with the amine. This provokes an acid-base reaction, in which the amine base bonds with acidic gases ($H_2S$ and $CO_2$), where the intermolecular bonds are of weak nature. This process releases two streams: purified NG stream and rich amine stream. The purified stream is the desired stream which can then be transported to the NG distribution network. However, the process does not end at this stage, because the rich amine stream can be treated where the acid gas constituents are removed so that the amine stream can be re-used or re-circulated back into the contractor tower. The rich amine stream is then passed into the regeneration stage, which consists of a separation process (usually distillation) in which the amine is separated from the acid gas mixture. The acid gas stream is sent to be treated later on, while the amine stream or regenerated bottom stream is recycled in the recirculation stage in order to re-use the amine in the contractor tower, thus decreasing cost of purchasing new amine and increasing cost-effectiveness and efficiency [23]. The regeneration stage is a crucial process in which distillation is utilized.

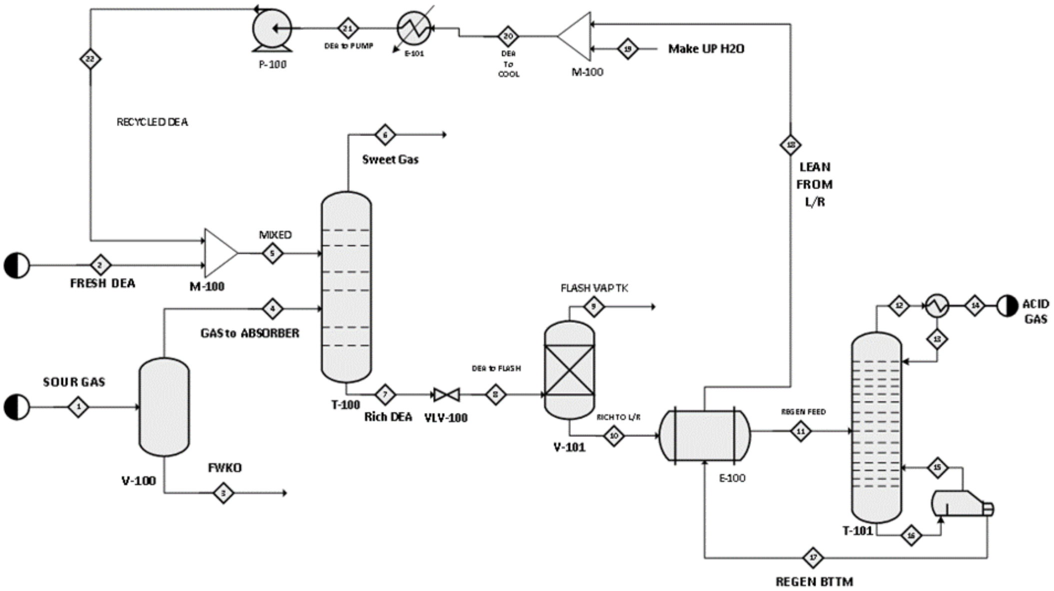

***Figure 3 Process flow diagram of acid gas sweetening.***

Sour components in natural gas can lower the calorific value and lower the overall quality of combustion as mentioned before. The most used to preserve the gas quality while maintaining the assets integrity is using amines as a solvent in an absorber system to extract the acid gases; the most prevailing ones are $H_2S$ and $CO_2$. The lean amine in the case study used is Di-ethanolamine (DEAmine) when in contact with sour gas after water removal from the free water knock out (FWKO) the amine extracts the sour gas via bonding with the $H_2S$. This is also known as $H_2S$ Scavenging. The bonds created by the amines to remove the $H_2S$ are considerably weaker compared to the ones of the sour gas with the natural gas. With some average heating the bonds break, releasing the acid gas and freeing the amine for recycling and further reuse. The heat subjected to the lean amine will cause desorption of the acid gases from the amine solution; the lean state is then achieved (lower concentration of acid gases). This is considered a very polluting procedure due to the release of $CO_2$ and $H_2S$ that can cause global warming and $SO_x$ formation in the atmosphere; therefore, adding a further treatment unit for the acid gas discharged will ensure environmental compliance. Whether it involves physical adsorption or chemical conversion, in both cases the aim is to minimize the harmful environmental impact and remove impurities. The purified or treated gas can be utilized in other industrial processes, according to the project specific consideration.

The lean amine is not directly recycled, further purification is needed to ensure all the acid gases are removed. With proper further treatment and recycling the amine absorption closed loop is completed. This recirculating loop ensures the continuous and efficient removal of acid gases from the gas stream, making the amine absorption

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process a sustainable and effective method for treating natural gas or other industrial gas streams contaminated with acidic components. The design and operation of such systems are critical to achieving optimal performance in terms of both gas purification and energy efficiency.

The removal of acid gases can be referenced to couple of factors; the usage of Di-ethanolamine in the amine contractor tower as an absorbent and the actual design of the absorption and regeneration stages are the two main parameters to watch and control. The pressure control on the amine is crucial to obtain better absorption. Before the initial stage in the case of Di-ethanolamine, reducing the pressure will ensure the release of hydrocarbon gas more efficiently. Amine contracting towers vary according to size ranging from 5 up to 20 or more stages according to the amount of feed inlet. In the design steps one must find a fitting combination of stage numbers taking into consideration both the energy consideration and how efficient the separation is.

Affective acid desorption requires a certain desired temperature; to reach the maximum separation from the distillation a heat lean/rich amine exchanger is essential. The heating step allows to increase the overall efficiency (heat) of the system and furthermore reduces the temperature of the lean amine coming of the regenerator. The heat transfer between the hot lean amine produced from the regenerator coupled with the colder stream of the released rich gas is crucial to get better separation. The lean gas is then kept in a recycling closed loop; this ensures no losses from economic. Although the process is considered a heavy carbon contaminating process but assuring chemical sustainability of reusing the amine is a step towards the right direction of decarbonizing natural gas, at least in the processing part.

## IV. PLANT DESIGN

### *A. Manual Design*

Manual calculations play a crucial role in confirming the accuracy and reliability of simulation results in various engineering processes, including those related to gas processing systems such as amine absorption. Manual calculations play a crucial role in confirming the accuracy and reliability of simulation results in various engineering processes, including those related to gas processing systems such as amine absorption. Furthermore, this can be a benchmark for quality assurance or validation of newly established work.

### *B. Mass Balance*

Conservation of mass during chemical reactions is one of the main things studied in chemistry and known as “Mass balance” [24]–[28]. The principle is that mass isn't produced or destroyed in a chemical reaction, but simply transferred and equation 1 has been used to refer to the overall mass balance used in equation 1, where the rate of accumulation has been specified to be zero as the process is in steady state.

$$\begin{aligned} Rate\ of\ Accumalation \\ &= input\ flowrate - output\ flowrate + generation\ rate \\ &- depletion\ rate \qquad (1) \end{aligned}$$

That is, the sum of masses for reactants must be equal to that of products in a chemical equation. This idea is very significant, because chemists can use it to calculate how far a reaction has gone and what proportion of products have resulted from given amounts of reactants.

Chemistry is no stranger to mass balance, but if you want chemicals to be handled together under the same principle, then first pay attention to stoichiometry (the quantitative proportion between reactants and products). The ratio of moles for the reactants and products is derived from the balanced chemical equation, which thus specifies this relationship quite exactly. One can calculate the quantities of each reactant consumed and product formed by using stoichiometric numbers from a balanced chemical equation. Chemists can then use this to determine the theoretical yield of a reaction, meaning how much product is possible.

It is possible to calculate the percentage yield of a reaction using mass balance. Percent yield, as its name implies, is the ratio between actual and theoretical yields (expressed %). The percent yield tells you something about the efficiency of a chemical reaction and can also indicate how good your process is.

Example on mass balance for a distillation process that has no consumption or formation, is done on V-100:

1. List the givens as shown in ***table 1***.
2. Sour gas = 24,230 kg/hr.
3. Gas to CONT = 24,140 kg/hr.
4. FREE WATER KNOCK OUT = 90 kg/hr.

***Table 1 Material Composition.***

| | | |
|---|---|---|
| Nitrogen | 0.023 | 2.3% |
| Carbon dioxide | 0.0934 | 9.3% |
| Hydrogen sulfide | 0.0301 | 3% |
| Methane | 0.7194 | 70% |
| Ethane | 0.0607 | 6% |
| Propane | 0.0211 | 2% |
| I-butane | 0.0078 | 0.7% |
| N-butane | 0.0087 | 0.8% |
| I-pentane | 0.0057 | 0.56% |
| N-pentane | 0.0044 | 0.4% |
| N-hexane | 0.0080 | 0.8% |
| N- heptane | 0.0371 | 3% |
| $H_20$ | 0.0046 | 0.4% |

The mass balance of the components in V-100 has been made to determine the overall flowrates as well as the components flowrates of Nitrogen, $CO_2$, $H_2S$, $CH_4$, Ethane, Propane, I-butane, N-butane, I-pentane, N-pentane, N-hexane, N-heptane and $H_2O$, as demonstrated in equations 2 to 15. The input mass flow rate of the sour gas equals the output mass flow of both the gas and the knocked-out water. This means that the process has no consumption or formation. Apply the overall mass balance and component mass balance equations on each unit in the process.

$$Overall\ mass\ balance: 24{,}230\frac{kg}{hr} = 24.140\frac{kg}{hr} + 90\frac{kg}{hr} \tag{2}$$

$$Nitrogen\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.023 = (24{,}140\frac{kg}{hr} * 0.023) + 0 \tag{3}$$

$$Carbon\ Dioxide\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0934 = \left(24{,}140\frac{kg}{hr} * 0.0938\right) + \left(90\frac{kg}{hr} * 0.0024\right) \tag{4}$$

$$Hydrogen\ Sulfide\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0301 = \left(24{,}410\frac{kg}{hr} * 0.0302\right) + \left(90\frac{kg}{hr} * 0.022\right) \tag{5}$$

$$Methane\ mass\ balance:\ 24{,}230\frac{kg}{hr} * 0.70 = \left(24{,}140\frac{kg}{hr} * 0.7914\right) + \left(90\frac{kg}{hr} * 0.0010\right) \tag{6}$$

$$Ethane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.067 = \left(24{,}140\frac{kg}{hr} * 0.0610\right) + \left(90\frac{kg}{hr} * 0.0001\right) \tag{7}$$

$$Propane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0211 = \left(24{,}140\frac{kg}{hr} * 0.0212\right) + 0 \tag{8}$$

$$I - butane\ mass\ balance: 24{,}230\frac{kg}{hr} *\ 0.0078 = \left(24{,}140\frac{kg}{hr} *\ 0.0078\right) + 0 \tag{9}$$

$$N - butane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0087$$
$$= \left(24{,}140\frac{kg}{hr} * 0.0087\right) + 0 \qquad (10)$$

$$I - pentane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0057$$
$$= \left(24{,}140\frac{kg}{hr} * 0.0057\right) + 0 \qquad (11)$$

$$N - pentane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0044$$
$$= \left(24{,}140\frac{kg}{hr} * 0.0045\right) + 0 \qquad (12)$$

$$N - hexane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0080$$
$$= \left(24{,}140\frac{kg}{hr} * 0.0080\right) + 0 \qquad (13)$$

$$N - heptane\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0371$$
$$= \left(24{,}140\frac{kg}{hr} * 0.0372\right) + 0 \qquad (14)$$

$$Water\ mass\ balance: 24{,}230\frac{kg}{hr} * 0.0046$$
$$= \left(24{,}140\frac{kg}{hr} * 0.008\right) + \left(90\frac{kg}{hr} * 0.9943\right) \qquad (15)$$

### *C. Distillation*

The conditions of equilibrium between a substance's liquid and vapor phases at a specific temperature and pressure are described by vapor-liquid equilibrium (VLE), a basic concept in thermodynamics [14]. A common graphic illustration of the ideas of VLE is an equilibrium diagram, also known as a phase diagram or XY plot. In the context of an XY plot, the composition of the vapor and liquid phases is depicted. For binary systems, the x-axis typically represents the mole fraction of one component, usually the more volatile one, while the y-axis represents the mole fraction in liquid phase. The slope of the tie line connecting the vapor and liquid compositions at equilibrium indicates the relative volatility of the components. This relative volatility, a ratio of vapor pressures, plays a crucial role in processes like distillation, influencing separation efficiency [15].

In the McCabe-Thiele method, the equilibrium curve, representing the vapor-liquid equilibrium for the given binary system, is plotted on the same graph as the operating line, which represents the conditions in the distillation column. The intersection of these two lines provides key information about the number of trays or equilibrium stages required for the desired separation [16].

For DEA and water, the formation of an Azeotropic mixture is dependent on the concentrations of each component. The presence of an Azeotropic is particularly relevant in processes such as amine gas treating, where DEA is commonly used to remove acidic gases like carbon dioxide (CO2) and hydrogen sulfide (H2S) from natural gas streams. Fortunately, in this case the composition is only 28% of DEAmine in water which is far from the Azeotropic point. Equation 16 has been used, Where:

$\alpha$ is the relative volatility, which is the ratio of the vapor pressure of the more volatile component to the vapor pressure of the less volatile component at a given temperature.

$\alpha > 1$ is indicated that component A is more volatile.

$\alpha < 1$ is indicated that component B is more volatile.

$\alpha = 1$ is indicated that there is no separation, the 2 components have the same volatility.

$$y = \frac{\alpha x}{1 + (\alpha - 1)x} \tag{16}$$

The vapor liquid equilibrium is then plotted according to the given values of the vapor and liquid fractions of water as seen in ***table 2***.

***Table 2 $H_2O$ vapor and liquid molar fractions.***

| X $H_2O$ (mole) | Y $H_2O$ (mole) |
|---|---|
| 0 | 0 |
| 0.1000 | 0.42834 |
| 0.2000 | 0.73471 |
| 0.3000 | 0.86303 |
| 0.4000 | 0.92315 |
| 0.5000 | 0.95499 |
| 0.6000 | 0.97339 |
| 0.7000 | 0.98471 |
| 0.8000 | 0.99197 |
| 0.9000 | 0.90677 |
| 1.000 | 1.000 |

The following steps were followed in order to perform the equilibrium curve as demonstrated in ***figure 4***. First, determine process operating conditions by setting the desired.

(Xf): Mole fraction of the light component in the feed mixture =0.35,

(Xd): mole fraction of the light component in the distillate =0.975, and

(Xb): mole fraction of the light component in Bottom product= 0.025.

Then draw the diagonal line, feed line(q-Line), q=1 (saturated liquid), and operating line of rectifying section:

The rectifying section is found by drawing a horizontal line from (Xd) until intersect with forty-five-degree line. Then find the Y- intercept using the given reflux ratio, XD/(R+1). The rectifying line is drawn between the corresponding point of (Xd) and corresponding point of (Xf). The operating line of stripping section has been drawn. The stripping section is found by drawing horizontal line from the bottom product (Xw) until intersect with forty-five-degree line. Then The stripping line is drawn between the critical point of (the end point of the rectifying section) and corresponding point of (Xw). Determine the Number of Trays(N): Analyze the number of stages by counting the intersected points between the stripping and rectifying sections in the vapor liquid equilibrium line and the operating line.

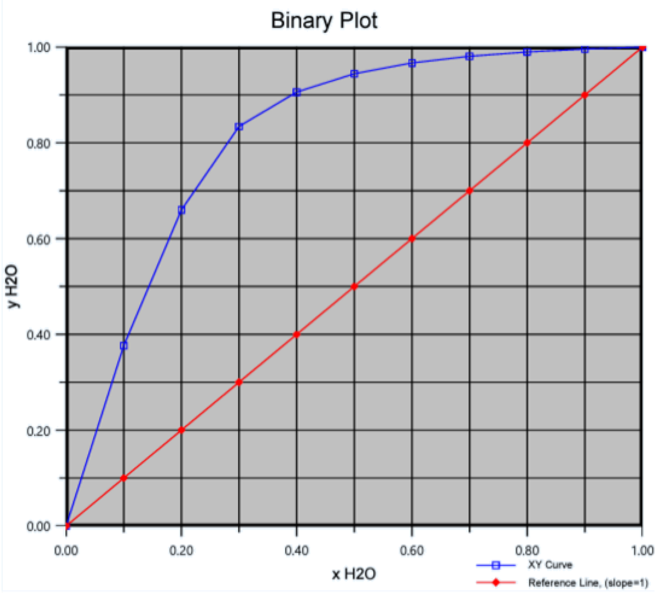

***Figure 4 Binary plot of water phases equilibrium according to Aspen HYSYS V10.***

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All of the previous steps were done as per referenced distillation column design steps [4], [29]–[32]. This has been done to compare and analyze the number of stages required as per the simulation model as well as the manual calculations as shown in ***figure 5***.

**Xf= 0.35 Xw= 0.025 Q=1 Saturated Liquid R= 0.5**

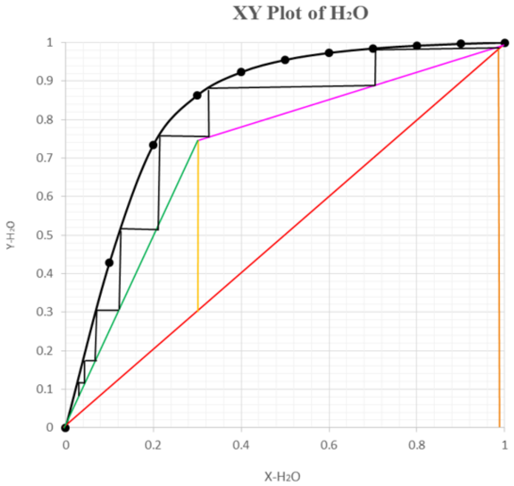

***Figure 5 Determined number of stages required in distillation column using binary phase equilibrium system.***

*D. Simulation Using Aspen HYSYS*

The simulation model was developed as displayed in ***figure 6.*** To achieve the requirements for gas quality, acid gases like H2S and/or CO2 are extracted from sour natural gas using amines or specialized solvents. In the contactor tower, lean amine meets the sour gas. The acid gases form a weak connection with the amine because of an acid-base reaction. The rich amine next passes via a regenerator, where heat dissolves the amine's weak link with the acid gas to release the gas and regenerate the amine for future use. The lean amine is returned to the contactor in a recirculating loop as the acid gases leave the regenerator's top for additional treatment. An amine contactor receives a natural gas stream that is saturated with water. In this instance, the absorbing medium is di-ethanolamine (DEA), diluted to a level of 28% in water. There are twenty actual phases in the contactor. Before the rich amine reaches the lean/rich amine exchanger, most of the absorbed hydrocarbon gas is released by flashing it from the contactor pressure of 1000 psia to 90 psia. The rich amine in the lean/rich exchanger is heated to 200 °F, which is the regenerator feed temperature. There are 20 actual steps in the regenerator as well. Lean amine is generated at about 255°F, while acid gas is rejected from the regenerator at 120°F. After cooling, the lean amine is recycled back to the contactor.

First, the component list is needed according to the process requirements is (Carbon Dioxide $CO_2$ - Hydrogen Sulfide - Methane C1 - Ethane C2 - Propane C3- Isobutane i-C4- Normal Butane n-C4 - Iso-Pentane i-C5 - Normal Pentane n-C5 - Hexane C6 - Heptane C7- Water $H_2O$ - Di-Ethanol Amine). The fluid package is acid gas- chemical solvent. The process diagram starts with sour gas as the feed stream entering the vessel V-100 at 0 delta P, with two output streams FWKO (downstream) and GAS to CONT (upstream), feed stream is at Temperature of 86 F, Pressure 1000 psia and Molar Flowrate of 25 MMSCFD, with components and compositions as following: (Nitrogen 0.0016 - Carbon Dioxide 0.0413 - Hydrogen Sulfide 0.0172 - Methane 0.8692 - Ethane 0.0393 - Propane 0.0093 - Iso-Butane 0.0026 - Normal Butane 0.0029 - Iso-Pentane 0.0014 - Normal Pentane 0.0012 - Hexane 0.0018- Heptane 0.0072 - Water 0.005 - Di-Ethanol Amine 0.00).

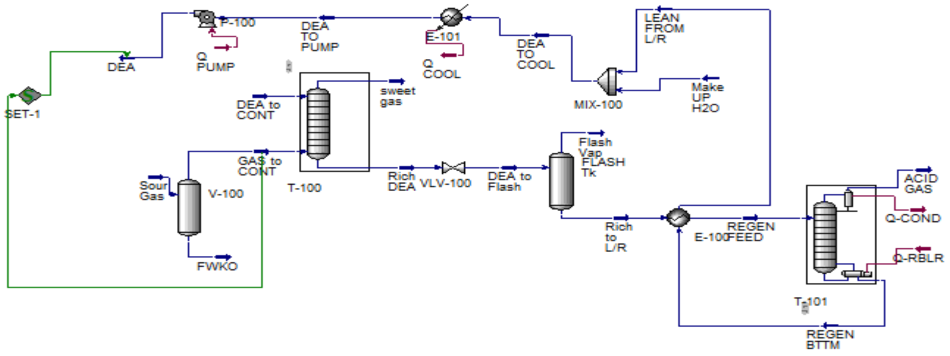

*Figure 6 Aspen HYSYS simulation of acid gas sweetening process.*

Gas TO CONT enters the absorber column producing SWEET GAS stream(upstream) and RICH DEA (downstream) under certain conditions. The absorber consisted of 20 stages. With top pressure of 995 psia, top temperature of 100 $^{o}$F, bottom pressure of 1000 psia bottom temperature of 160 $^{o}$F. RICH DEA enters a valve to reach a pressure of 90 psia and then to the flash tower to release most of the absorbed hydrocarbon gas before it enters the lean/rich amine exchanger. Next, RICH TO L\R produced from the flasher is sent to the heat exchanger with tube side and shell side pressure difference of 10 psia, and temperature of REGEN FEED is 200 $^{o}$F. The heat exchanger is producing REGEN FEED stream goes to the distillation column and LEAN FROM L\R stream is sent to a mixer with MAKEUP H2O stream.

The distillation column conditions are very detailed. 18 stages with condenser pressure of 27.5 psia and 2.5 psia pressure drop, alongside reboiler pressure of 31.5 psia and damping factor of 0.4. The top temperature in the condenser is specified to be 179.6 $^{o}$F with a reflux ratio of 0.5. the overhead vapor flowrate is valued at 2.0 MMSCFD, and the reboiler' s duty is 1.356e7 BTU/hr.ACID GAS stream is produced from the distillation column (upstream), REGEN BTTMS stream as (downstream) is sent to the heat exchanger. LEAN from L\R is sent with MAKEUP H2O stream mixed through a mixer at temperature of 70 $^{o}$F, pressure 21.5 psia, molar flowrate of 190 MMSCFD and a saturated ideal liquid volumetric flow of 190 USGPM. The mixer is designed to meet the same specifications of MAKEUP H2O. DEA to COOLER produced from the mixer is sent to cooler at pressure drop =5 psia. DEA TO PUMP produced from cooler is sent to the pump to increase the stream's pressure up to and temperature at 95 F. DEA stream finally as a product from the pump which is containing 0.9485 water and 0.0499 Di Ethanol Amine. Set unit is installed at the streams (DEA and GAS TO CONT), DEA as the target and GAS TO CONT as the source, by maintaining the pressure as the target value and set the number of the multiplier at 1 with offset is -5.

***Table 3*** below is illustrating the stream names with their important composition values according to the process diagram to show some of end product produced from specific units as following:

***Table 3 Resulted streams data.***

| Unit name | Stream name | Component | Composition |
|---|---|---|---|
| Vessel | FWKO | H2O | 0.9943 |
| | | H2S | 0.0022 |
| | GAS TO CONT | H2S | 0.0302 |
| Distillation column | REGEN BTTM | H2O | 0.0004 |
| | | H2S | 0.18013 |
| | | Di Ethanol Amine | 0.0626 |
| Pump | DEA | H2O | 0.94854 |
| | | Di Ethanol Amine | 0.0499 |

Looking at the simulation results, it is quite an effective separation composition of approximately 95% separation when using 18 trays in the distillation column. For the manual design results, water and DEA can be separated at a desired aqueous water composition of 0.975 in the distillate where the number of trays required are 7 trays.

However, it would be unfair to put both results in a direct comparison due to the nature of the system and the mixture of components. In the simulation case, it has been based on the multi-phase system where all the components in the rich amine stream have been taken into consideration as per the calculation mechanism of Aspen HYSYS V10. Therefore, it would be recommended to use a multi-phase system calculation scheme when doing the manual distillation design to make it a fair and even comparison between the manual and simulation results.

## V. PROCESS SAFETY

Hazard and safety considerations is a crucial component of any operations especially in the Petro-chemical field [33]. Process safety considerations are crucial not only for preserving the assets but also to ensure the protection of on-field workers, engineering, and environment surrounding. Special considerations must take place in the Petro-chemical field due to the sensitivity and high risks of the operations taking place. While dealing with hazardous, flammable, and reactive fluids, even minor incidents can lead or escalate into catastrophic accidents. The process safety analysis conducted focused on two main categories (Dow F&EI and HAZOP) that give the most possible detailed process safety analysis with the simulation data provided [34]–[37]. ***Figure 7*** gives a guideline on the method used to determine F&EI, which has been applied to the natural gas dominant fraction in the contractor tower, in which the process unit has been selected and the material factor of methane was obtained. The factor was later on utilized in the calculation of the general and special process hazards in order to determine the process unit hazards factor, which was multiplied by the methane's material factor to deduce the fire and explosion index of methane.

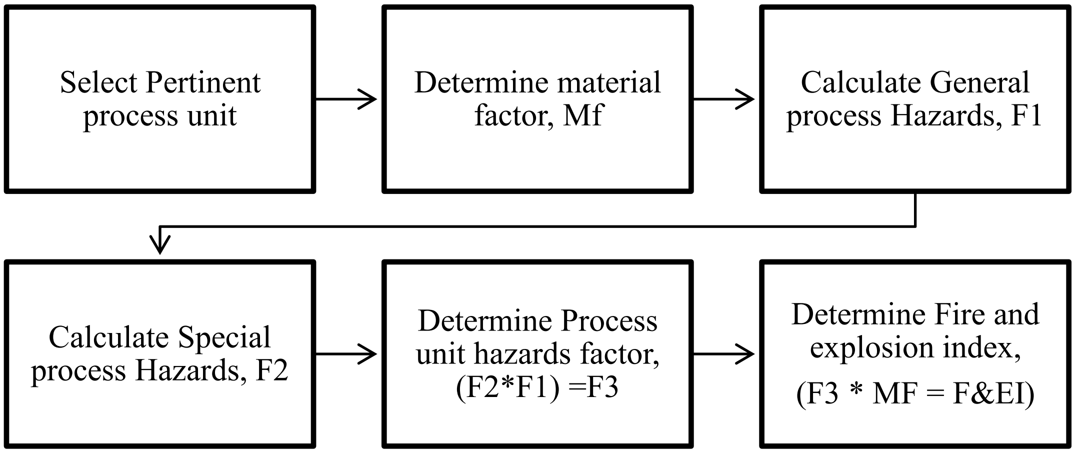

***Figure 7 Fire and explosion index calculation flowchart.***

### *A. DOW Index*

The Dow Fire and Explosion Index was developed to quantify the hazards relative to fire and explosion risks in a set unit. Both the data provided by the American Institute of Chemical Engineers (AIChE), and the simulated results from ASPEN HYSYS assisted in the criteria of choice regarding the acid gas sweetening process. As demonstrated in ***table 4***, the first step is selecting the process unit providing all the specifications needed to determine the index. The unit in use is the DEA Contractor Tower (T-100); where Acid gas enters from a bottom gas stream with a Diethanolamine (DEA) to separate the corrosive gases (CO2 & H2S) to elevate the quality of natural gas produced. A simple yet crucial step; providing the proper naming and details of the unit and process in use will ease the process for later routine inspections and reduce the amount of human error while auditing for future maintenance. The next step of the plan is to determine the material factor (MF) according to a set criterion.

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*Table 4 Calculating the fire &explosion index step wise procedure.*

| **Process unit** | **DEA Contractor Tower T-100** | |
|---|---|---|
| **Materials in unit process** | **Diethanolamine (DEA) – Methane (NG) – Water** | |
| **Basic material for material factor** | **Methane (NG)** | |
| **Material factor** (Note requirement when until temperature over 140 F (60°C)) | **21** | |
| **1. General process Hazards** | **Penalty factor for range** | **Penalty factor for used** |
| **Base factor………………….** | 1.00 | **1.00** |
| **A. Exothermic chemical reactions** | 0.30 to 1.25 | **0.5** |
| **B. Endothermic processes** | 0.20 to 0.40 | |
| **C. Material Handling and transfer** | 0.25 to 1.05 | |
| **D. Enclosed or indoor process units** | 0.25 to 0.90 | |
| **E. Access** | 0.25 to 0.35 | **0.35** |
| **F. Drainage and spill control** | 0.25 to 0.50 | |
| **General process Hazards factor(F1) …….** | | **1.85** |
| **2. Special process Hazards** | | |
| **Base factor………………….** | 1.00 | **1.00** |
| **A. toxic Materials** | 0.20 to 0.80 | **0.80** |
| **B. Sub-Atmospheric pressure (500 mm Hg)** | 0.50 | |
| **C. Operation in or near flammable range** | | |
| **1. Tank farms storage flammable liquids** | 0.50 | |
| **2. process upset or purge failure** | 0.30 | |
| **3. Always in flammable range** | 0.80 | **0.80** |
| **D. Dust Explosion** | 0.25 to 2.00 | |
| **E. Pressure**<br>Operating pressure **995 psig**<br>Relief setting **1500 psig** | | **0.80** |
| **F. Low temperature** | 0.20 to 0.30 | |
| **G. Quantity of flammable / Unstable material:**<br>Quantity **37830.0711 Ib**<br>Hc= **$21.5* 10^3$ BTU/Ib** | | |
| **1. Liquids or gases in process** | | **1.35** |
| **2. Liquids or gases in storage** | | |
| **3. Combustible solids in storage, dust in process** | | |
| **H. Corrosion and erosion** | 0.10 to 0.75 | **0.10** |
| **I. Leakage – Joints and packing** | 0.10 to 1.50 | **0.4** |
| **J. Use of fired equipment** | | |
| **H. Hot oil heat exchange system** | 0.15 to 1.15 | |
| **L. Rotating equipment** | 0.50 | |
| **Special process Hazards factor(F2) ……………….** | | **4.25** |
| **Process until hazards factor(F2*F1) =F3 …………** | | **7.8625** |
| **Fire and explosion index (F3 * MF = F&EI)** | | **165.1125** |

As mentioned in the technical manual published by the AIChE, the material choice in the case of mixture or multiple materials, should be according to the safety considerations and the material factor. If one or more material is of significant concentration, 5% or more; the choice of material must correspond with the highest MF. The material of concerns that exceed the 5% threshold are DEA and Methane that is commonly referred to as natural gas (NG); although DEA is essential in the sweeting process, the MF of methane is much higher (21) compared to

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DEA's material factor of (4). Such a difference is significant to focus on methane rather than any other material in the process. The next step after choosing the right MF, is calculating the F1 and F2 that are related to the general process hazards and special process hazards respectively. Combining both info given from simulation and the index penalty system, finding the applicable hazards that correlate to the process is key to achieve highest accuracy. Starting off with general process calculations (F1), the reactions taking place taking place inside the

$$2R_2NH + H_2S \rightleftharpoons (R_2NH_2)_2 \cdot S$$
$$(R_2NH_2)_2S + H_2S \rightleftharpoons 2(R_2NH_2)HS$$
$$2R_2NH + CO_2 \rightleftharpoons R_2NCOONH_2R_2$$

absorber where acid gas is absorbed by DEA is a slightly exothermic process that favors low temperature and high pressure. According to the Index, those specifications fall under the moderate exotherms condensation reaction with a penalty factor of 0.5. The next two specs, material handling and enclosed process units are both not applicable in the acid gas sweeting process due to the use of piping and the full automation of the process where no handling is required when dealing with NG. The next applicable spec is related to access from and toward the unit operation at least from two sides with no obstructions. The estimated process area of a DEA contractor tower that produces 32.5 million standard cubic feet per day MMSCFF is more than the estimated threshold of 925 m2, therefore receiving a penalty of 0.35. Drainage and spill control is the final spec in the F1 benchmark that is not considered applicable due to absence of any combustible/flammable liquids. Adding up the penalty of each spec with a bas factor of 1 will equate to an F1 of 1.85 that is considered relatively high given the fact only two out of the six specifications are considered applicable.

Moving to calculating the Special Process Hazards Factor F2 which includes, firstly, Toxic Material(s), when responding to an event, emergency responders may find it more difficult to investigate or minimize damage due to the presence of toxic compounds. Apply a penalty of 0.20 x NH. The component with the greatest NH should be used in mixes. As stated in NFPA 704, NFPA 325M, or NFPA 49, NH is a material's health factor. You can locate the NH for several materials in Appendix A. Determining an NH for novel materials might be aided by consulting an industrial hygiene specialist. As the materials that on very short exposure could cause death or major residual injury, so for the methane case NH = 4 is selected. Therefore, the toxicity penalty of methane = 0.2 x 4 = 0.8. Then, Sub-Atmospheric Pressure, this section deals with a process circumstance in which a system's air leakage might provide a risk. Air contact with materials that are susceptible to moisture or oxygen, or the creation of combustible mixtures when air is introduced, can all pose a risk. This penalty is applied only if the absolute pressure is less than 500 mm Hg (equivalent to 10 in Hg vacuum) whereas the penalty is 0.50. As all pressure values in this process are above 500 mmHg, this penalty is neglected. As well as Operation in or Near Flammable Range, there are certain operating conditions which can cause air to enter and be entrained into the system. The introduction or entry of air could lead to the formation of a flammable mixture and create a hazard. This section is intended to cover the following conditions:

NF = 3 or 4 flammable liquid storage tanks allow for the inhalation of air during pump-out or abrupt tank cooling. There is a 0.50 penalty. A penalty of 0.50 would be necessary for an open vent or non-inert gas padded functioning pressure-vacuum relief system. A 0.50 penalty would also apply for the storage of flammable liquids over their closed cup flash points without inerting. There is no penalty if a closed, intake vapor recovery system is utilized, and its airtightness is guaranteed. A penalty of 0.30 would be applied to process equipment or process storage tanks that could be in or close to the flammable range only in the case of an instrument or equipment failure. A penalty of 0.30 is required for each process unit that depends on inert purging to keep it out of the flammable range. This penalizes tank vehicles and cushioned barges as well. If the penalty listed under "Sub-Atmospheric Pressure" has already been taken, this penalty is not imposed. Processes or operations that are by nature always in or near the flammable range, either because purging is not practical or because it was elected not to purge. receive a penalty of 0.80. Therefore, as the nature of methane is flammable itself or near an external source, the penalty range is 0.8.

Following on the list of special hazards, Dust explosion is not relevant due it its involvement with solid particles that is not available in the sweetening process; furthermore, no penalties were given in the low temperature section due to the usage of stainless steel with a high critical temperature that is out of range of the index provided. Concerning the pressure specifications, the operating pressure of the absorber is a relatively high pressure of 995 psig with a design & relief device set pressure of 1500 psig. According to the graphs provided by the index, the penalties of the operating and relief set pressure are 0.86 and 0.92 respectively. The operating set penalty divided by the pressure set penalty will provide an adjustment factor of:

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$$0.86 * \left(\frac{0.86}{0.92}\right) \cong 0.8$$

The quantity of methane was based on the simulated process, where 37830.0711 Ib, which has been assigned as the quantity of flammable material. The heat of combustion (Hc) is 21.5 x $10^3$ BTU/Ib as per the Dow's Fire and Explosion Index Hazard Classification Guide. To get the penalty factor, the graph for flammable and combustible penalty factor calculation from the classification guide has been referred to as means of calculation and interpreting the value of the penalty factor of point G. The quantity of flammable material is multiplied by Hc to get total BTU in process, which is then divided by $10^9$ and substituted into the classification guide's interpreted graph equation The penalty factor has then been deduced to be 1.35.

For point H, it has been determined that corrosion in the process unit is less than 1.27 mm/year, and may reach a predicted value of 1 mm/year [23], thus the penalty factor is 0.10. The next point (I), represents leakage inside joints and packings, methane is considered to have a penetrative nature, and the absorber's main internal constituent contains packing [17]. Thus, the penalty factor is 0.4. The absorber unit does not use a hot oil heat exchange system, nor does it use fired heat equipment due to lack of combustion reactions within the unit in addition to the fact that there no rotating equipment. Such observations of the process unit resulted in the absence of consequences of points J, K and L.

The penalty factors have been summed up to give a value of 4.25, which is F2, the special process hazards factor. Consequently, the process unit hazards factor F3 is then calculated by multiplying F1 by F2 to give an F3 value of 7.8625. F3 can be then multiplied by methane's material factor (MF), giving a resulted value of 165.1125, which is referred to as the fire and explosion index (F&EI).

### *B. HAZOP Study*

The importance of conducting a Hazard and Operability (HAZOP) study cannot be overstated. This systematic process safety management tool serves as a cornerstone for risk assessment and mitigation in industries prone to hazardous operations. Through a HAZOP study, potential hazards and deviations from normal operation are meticulously unearthed, enabling organizations to take initiative-taking measures to minimize or eliminate these risks. Such preventive actions not only safeguard against accidents and injuries but also prevent environmental harm and property damage. Furthermore, HAZOP studies are often a legal requirement, ensuring compliance with regulatory standards and reducing the likelihood of legal repercussions. Beyond compliance, these studies offer the opportunity for process optimization, enhancing operational efficiency and cost-effectiveness. Therefore, HAZOP study results have been tabulated in this section. Starting with ***table 5***, which is V-100, taking into consideration several scenarios regarding pressure drops and exceeding pressure as well as flow deviations. Such deviancies may result in the drop in the separation efficiency of the FWKO from the sour gas.

***Table 5 HAZOP Study on V-100.***

| Guide word | Variable | Deviation | Possible cause | Possible consequences | Action |
|---|---|---|---|---|---|
| None | Sour gas feed | Loss of sour gas feed. | Pipeline leakage/blockage | Operation is halted | - Proper maintenance<br>- Check clogging<br>- Revise pipe sizing |
| Less | Vessel pressure | Pressure is less than designed pressure range | - Sour gas flowrate lower than design specs<br>- Lower temperature<br>- Leakage in vessel or feeding pipeline | - Insufficient separation<br>- Trapped sour gas droplets | - Install pressure regulating valve (PRV)<br>- Install pressure indicator and/or alarm |
| More | Vessel pressure | Pressure exceeds intended pressure range | - Higher feed or vessel temperature<br>- Blockage in pipes | - High presence of FWKO in vapor GAS to CONT stream<br>- Risk of vessel explosion | - Regular maintenance<br>- Install pressure control systems with cascading temperature control systems |
| Reverse | Direction of | Sour gas | Insufficient overhead | Drop in flowrates of | - Correct pump |

| | | | | | |
|---|---|---|---|---|---|
| | flow | | pressure | GAS to CONT and FWKO streams | design.<br>- Install non-return valve (NRV) |
| As well as | Liquid and vapor flow | Improper separation of FWKO from sour gas stream | Misinterpretation in vessel design regarding contact time, pressure, and vessel dimensions | Insufficient separation | - Reconsider design specifications,<br>- regular maintenance of vessel,<br>- Installation of flow control systems<br>- Installation of PRV |
| Other than | Composition | Tracing of $CO_2$ and $H_2S$ | Nature of sour gas | - Vessel corrosion<br>- Contamination of GAS to CONT stream | - Corrosion inhibitors<br>- Regular maintenance |

Studying the next unit which involves the absorber in which the main process of acid gas sweetening happens. It is crucial that the absorber is studied intensively as demonstrated in **table 6**. There are several deviations which may happen, including disruptions in pressure, composition, and risks of contamination. Such issues urge the need to take actions such as installation of alarms, control systems, creation of bypass systems and regular maintenance of the pipelines.

***Table 6 HAZOP study on T-100.***

| Guide word | Variable | Deviation | Possible cause | Possible consequences | Action |
|---|---|---|---|---|---|
| No | Rich DEA stream | No flow | - Turn off flow control valve<br>- Blocked line | Absorber flooding | Turn off the feed, then check that the flow indicator and control. Clear line. |
| Low | Flow of feed gas | Decrease feed gas flow | - Blockages in the line<br>- Failure on the pump | - Insufficient absorption of acid gas in a solvent<br>- Poor solvent circulation | - Check for blockages<br>- Check pump operation and replace it |
| High | Flow of feed gas | Increase feed gas flow | Failure of flow control valve | - Off-spec product<br>- Foaming in the system. | - Reduce flow by closing or replacing flow control valve<br>- Regular maintenance |
| High | Pressure | Over pressure | Relief valve failure open | Pipe Leaks | Install high pressure alarm |
| More | Composition | Higher $H_2S/CO_2$ content >5% | Failure of analyzer | - Corrosion<br>- Off-spec product.<br>- Wrong absorption | - Add high composition alarm at 5%.<br>- Check analyzer. |
| Other than | Contamination | Impurities in feed stream | Failure of filter | - Plugging,<br>- Corrosion<br>- Fouling<br>- Off-spec product | Replace inlet filters |
| Reverse | Flow of DEA stream | Flow reversal | Control valve failure | Contamination | Install check valves (non-return valve) |
| As well as | DEA stream | High in water content | Dilution | Flooding | Monitor composition |
| Part of | DEA stream | Bypass stream | Failure in the valve | Incorrect absorption | Check the bypass valve |

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In ***table 7***, the HAZOP study of V-101, the flash vaporizer, is assessed. As for any flash vaporizer, it is quite essential to set the scenarios revolving around temperature and pressure changes, flow deviations, and contamination, in which these scenarios may lead later on to leakages, blockages, disruption in the separation efficiency and other severe possible consequences.

***Table 7 HAZOP study on V-101.***

| Guide word | Variable | Deviation | Possible cause | Possible consequences | Action |
|---|---|---|---|---|---|
| None | Liquid feed | Loss of Liquid feed | Rapture/ leakage in the line | No separation occurs | Install backup supply |
| Less | Flash vapor flow | Flow rate decreases below the minimal | Blockages in line | - Insufficient separation<br>- Vapor and liquid not fully separated | - Check for blockages<br>- Install low flow alarm |
| More | Flash vapor flow | Flow rate increases beyond the design | Failure of flow control valve | Off-spec product (high vapor level) | - Install high flow alarm.<br>- Repair and maintenance of flow control valve |
| Reverse | Flow reversal | Disorder of the liquid/vapor phase | Place incorrect inlet or place it inverted | Off-spec product due to the carryovers | Install check valves (non-return valve) |
| As well as | Liquid and vapor flow | Liquid phase carryover the vapor phase | Leakage in the pipe | Insufficient separation | - Regular pipe maintenance<br>- Install backup pipe |
| Other than | Composition | Presences of $CO_2$ and $H_2S$ | Leakage in the line | - Increase the corrosion rate<br>- Shut down of process | Acid gas detector |
| Part of | Vapor flow | Decrease vapor outlet flow | Blockage in the vapor outlet | Reduce the vapor flow through outlets | Proper maintenance and cleaning |

The heat exchanger's HAZOP study mainly revolves around two main components, the shell and tube since it is a shell and tube exchanger. The deviancies in the shell and the tube regarding flowrate direction, pressure, temperature, and contamination are recorded in ***table 8***, with actions which can be taken to mitigate and prevent these deviances to avoid possibilities of temperature and pressure drop as well as risks of explosion.

***Table 8 HAZOP on E-100***

| Guide word | Deviation | Possible Cause | Possible Consequences | Action |
|---|---|---|---|---|
| More | Increase in temperature | Failure of heating water valve | - Temperature of process fluid increases<br>- Possibility of explosion | High temperature alarm |
| More of | Increase in tube side pressure | Failure of process fluid valve | Damage of tube | Install high pressure alarm. |
| Less | Less flow of heating water | Pipe blockage | Temperature of process fluid decreases | Low Temperature alarm |
| As well as | Contamination of process fluid line | - Leakage of tube and cooling water<br>- Loss of control system on heat exchanger | Contamination of process fluid | Proper maintenance and operator alert |

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| | | | | |
|---|---|---|---|---|
| Also | Tube corrosion | - Hardness of heating water<br>- Corrosive impurities in water<br>- Old tube<br>- Continuous direct contact with water | - Less heating<br>- Bursting of tube | - Repeatedly observation<br>- Tube Panting<br>- Cathodic protection |

T-101 is a distillation column which consists of the main column coupled with a condenser and a reboiler. Therefore, the HAZOP study on the T-101 covered these three constituents as depicted from ***table 9***. The main deviancies which may happen are regarding temperature, pressure, and pressure, therefore it would be recommended to install several cascading control systems and alarms to maintain such issues.

***Table 9 HAZOP study on T-101.***

| Guide word | Variable | Deviation | Possible Cause | Possible Consequence | Action |
|---|---|---|---|---|---|
| NO | Flow of | No flow | - Pipe blockages<br>- Control valve shut<br>- Valves fail.<br>- Tube leakages and blocking<br>- Pump failure | - Column dries out<br>- Possible dangerous concentration<br>- No operation | - Install low level alarm<br>- Check maintenance procedure and schedule<br>- Make a bypass<br>- Emergency plant shut down |
| LESS | | Less flow | - Pipe blockages.<br>- Control valve shut<br>- Valves fail.<br>- Tube leakages and blocking<br>- Pump failure | - Column dries out<br>- Changes in product quality | - Install low level alarm<br>- Check maintenance procedure and schedule<br>- Make a bypass |
| More | | More flow | - The control valve fully opened<br>- Increase pumping capacity<br>- Control valve failure | - Column flooding<br>- Changes in product quality<br>- Temperature decrease<br>- Rise in bottom | - Install high level alarm<br>- Install control valve.<br>- Regular maintenance |
| HIGH | | High level | Output blockages | - Over pressure reflux drum<br>- Condensed liquid flow back to distillation | - Install high level alarm<br>- Check maintenance procedure and schedule. |
| Low | | Low level | Pipe partial clogged and leakage | - Level decrease in the level<br>- The valve closed<br>- Back flow of material | - Install low level alarm<br>- Check maintenance procedure and schedule<br>- Install valve |
| High | High in temperature | Increase temperature above target | Possible failure of condenser or reboiler | Explosion hazard | - Install high temperature alarm<br>- Check maintenance |

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| | | | | | |
|---|---|---|---|---|---|
| | | | | | procedure and schedule |
| Low | Low in temperature | Decrease temperature above target | Possible failure of condenser or reboiler | - Temperature decreases<br>- Separation rate decreases | - Install low temperature alarm.<br>- Check maintenance procedure and schedule. |

The final HAZOP study has been implemented on P-100 pump, as seen in ***table 10***. The pump's main deviation which may happen is regarding flow's rate, pressure, and direction, which may be result of failures in pump's operations. Therefore, it would be wise to consider adding the necessary sensors, valves, and control systems in order to mitigate the risks of increase in pressure which may lead to explosions as well as the aftermath effects later on the process by decreasing the over sweeting process productivity due to low DEA flowrate.

***Table 10 HAZOP on P-100.***

| Guide word | Deviation | Cause | Consequences | Action |
|---|---|---|---|---|
| Reverse | Flow | High pressure level from the mixer outlet | - Potential of pressure and temperature increase led to<br>- Explosion hazard | - Non return valve installation<br>- Pressure and temperature alarm or sensor |
| No/none | Flow | Compressor failure | Non-sufficient DEA for the sweetening process | Low concentration alarm |
| More | Flow | Failure of Pump control | Pressure increasing | - Temperature sensor<br>- Pressure sensor |

## VI. MODIFIED P&IDS

The HAZOP studies on the equipment led to the suggestion of several actions which are to be taken into consideration and implemented onto the process equipment. The actions through control systems, valves, alarms, and bypasses can be visualized through the construction of necessary P&ID diagrams as shown in ***figure 8 to 14***, which give detailed visual representation on the placement and connection of each of the control elements with tis respective constituent and how they are connected to the equipment simultaneously.

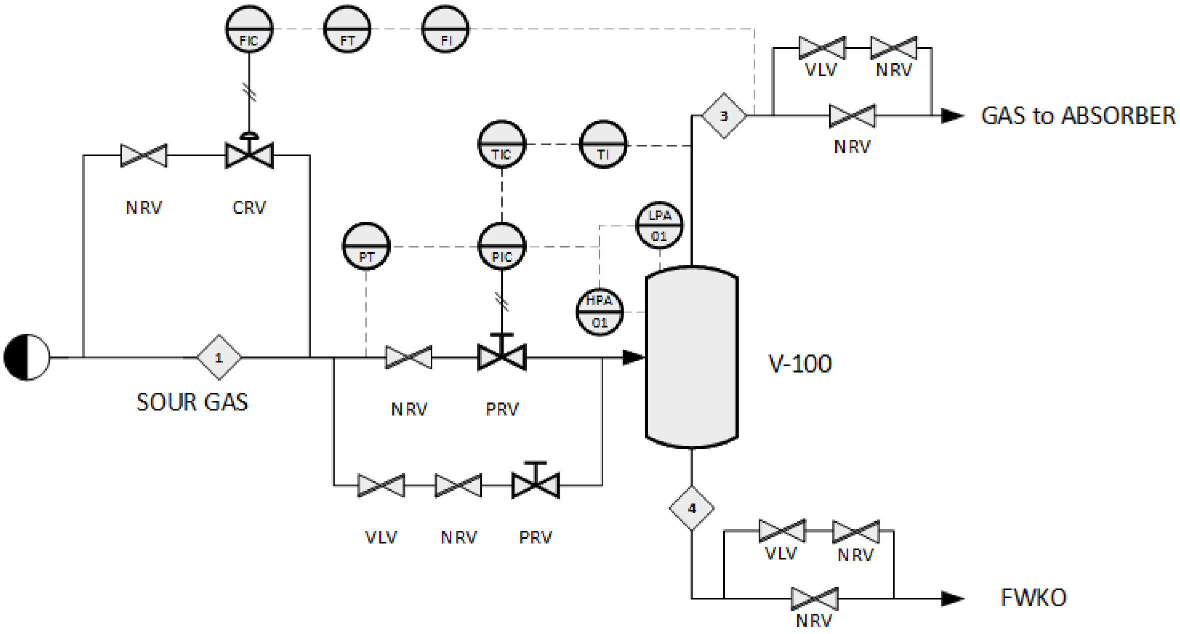

***Figure 8 Modified P&ID for V-100.***

Analysing ***figure 8***, Sour gas stream 1 is going into a vessel and has 2 bypass streams, one automated by a flow control loop and the other bypass stream has a check valve, a nonreturn valve, and a manual pressure relief valve passing through A non-return valve followed by a pressure relief valve that is automated by temperature and pressure cascade control loops Upper product of the vessel (gas) goes through a non-return valve and has a bypass stream consisting of a check valve and a non-return valve. The bottom product (FWKO) goes through a non-return valve and has a bypass stream consisting of a check valve and a non-return valve.

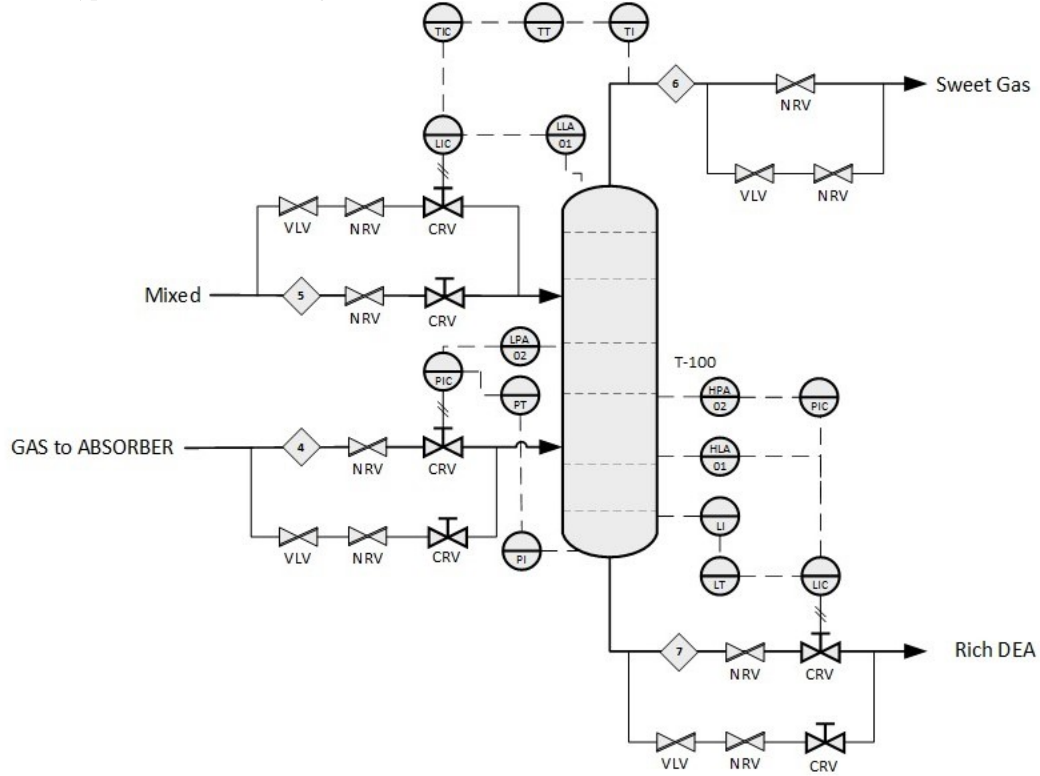

***Figure 9 Modified P&ID for T-100.***

Two streams entering a distillation column each with a bypass stream as seen in ***figure 9***. Mixed stream enters a distillation column through a non-return valve and a control valve. Stream no. 5 has a bypass stream containing a check valve, non-return valve, and a control valve. The control valve is automated by a cascade control loop containing temperature indicator (TI), temperature transmitter (TT), and temperature indicating control (TIC) in cascade with the level control system. Stream 4 is going to a distillation through a non-return valve and control valve with a bypass stream containing an automated flow control loop. Stream 6 contains sweet gas and contains a NRV and CRV with a bypass stream containing a check valve and a non-return valve. Stream no. 7 contains the rich DEA and contains a NRV and CRV with a bypass stream containing a check valve and a non-return valve. Stream no. 7 has a control loop of pressure indicating control (PIC), cascaded with level control systems and pressure alarms.

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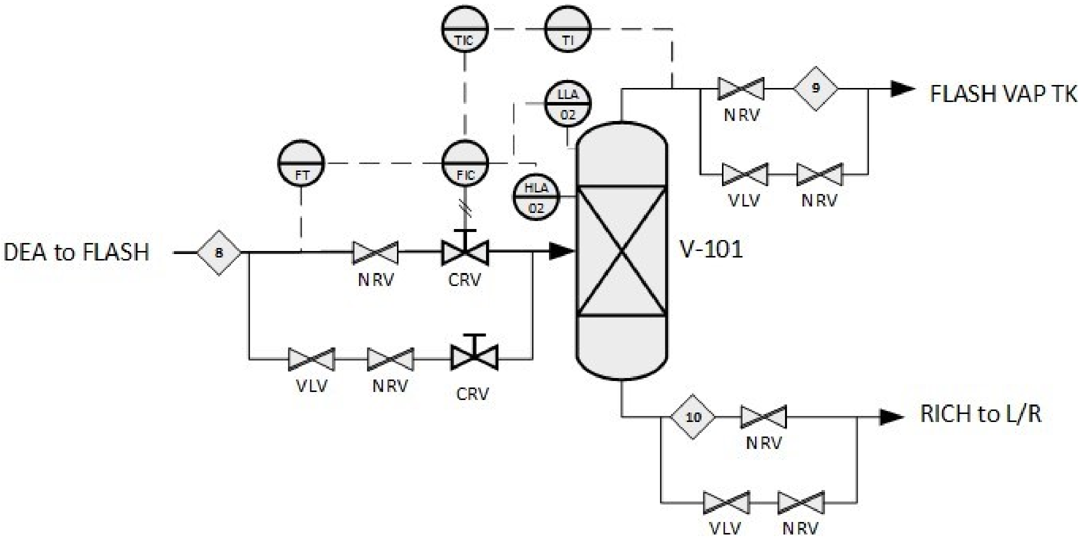

***Figure 10 Modified P&ID for V-101.***

In ***figure 10***, DEA in stream 8 is going into a vessel through a non-return valve and a control valve automated by flow control loop and temperature control loop and a bypass stream containing a manual control valve  Upper stream of the vessel which is stream 9 has a non-return valve and a bypass stream containing a check valve and a non-return valve  While downstream which is stream 10 has a non-return valve and a bypass stream containing a check valve and a non-return valve

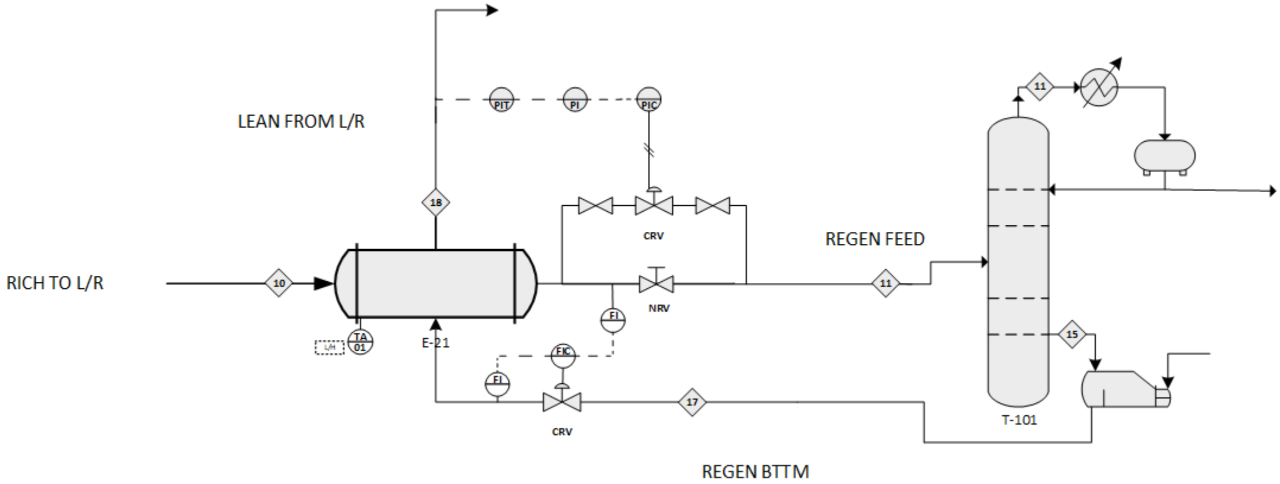

***Figure 11 Modified P&ID for E-100.***

It is critical to instrument the control system for the shell and tube heat exchanger E-101 as means of not only safety measures for the heat exchanger, but the distillation column as well, as depicted from ***figure 11***. Consequently, the heat exchanger is equipped with pressure controlling systems and temperature controlling systems, with an addition of temperature alarms (TA) for detecting low and high (L/H) temperature disruptions as a call for manual intervention.

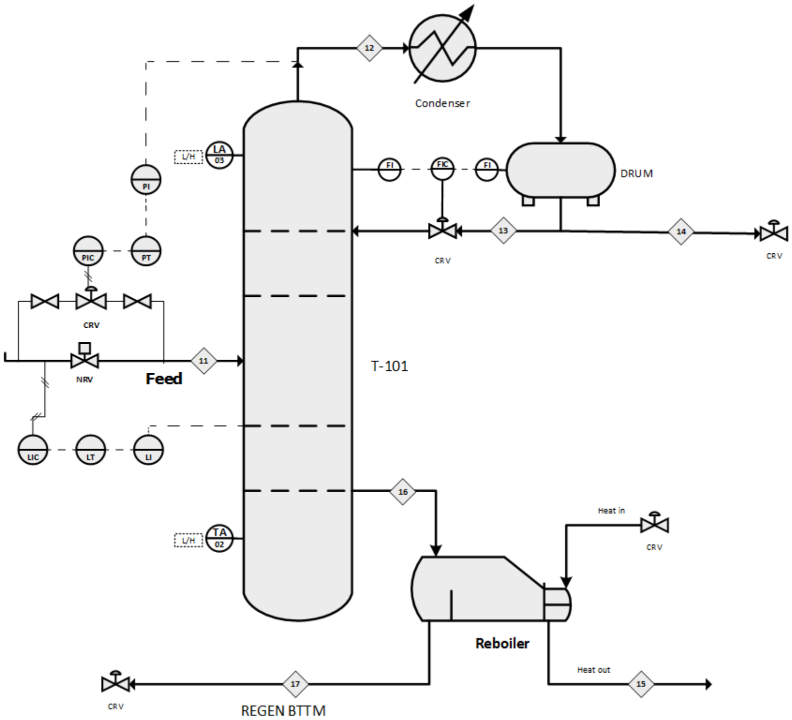

*Figure 12 Modified P&ID for T-101.*

Two streams entering a distillation column, each with a bypass stream as shown in ***figure 12***. Mixed stream enters a distillation column through a non-return valve and a control valve. Stream 5 has a bypass stream containing a check valve, non-return valve, and a control valve. The control valve is automated by a cascade control loop containing temperature indicator (TI), temperature transmitter (TT), and temperature indicating control (TIC) in cascade with the level control system. Stream 4 is going to a distillation through a non-return valve and control valve with a bypass stream containing an automated flow control loop. Stream 6 contains sweet gas and contains a NRV and CRV with a bypass stream containing a check valve and a non-return valve. Stream no. 7 contains the rich DEA and contains a NRV and CRV with a bypass stream containing a check valve and a non-return valve. Stream no. 7 has a control loop of pressure indicating control (PIC), cascaded with level control systems and pressure alarms.

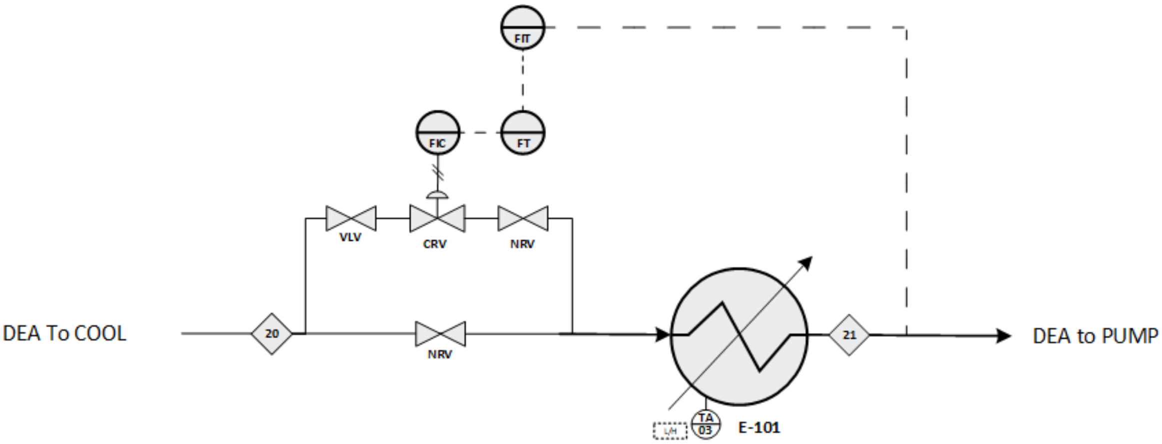

*Figure 13 Modified P&ID for E-101.*

Cool DEA in stream 20 goes into a reboiler as seen in ***figure 13***, through a non- return valve with a bypass stream containing several valves. Check valve (VLV) is placed so that if the mainstream is shut down or blocked, check valve is responsible for flowing bypass stream. A non-return valve (NRV) allows DEA to flow in only one direction and to make sure there is no return of flow. A control valve (CRV) containing a pressure relief valve which is placed between the reboiler and pump. If the pressure increases to a point above the preferred tank pressure, the relief valve will pop open quickly and relieve the flow to the pump. The control valve is automated by a control loop containing flow indicating transmitter (FIT), flow transmitter (FT) flow indicating control (FIC) which instructs to open/close the bypass control valve. Heated DEA comes out of the Re-boiler in stream 21.

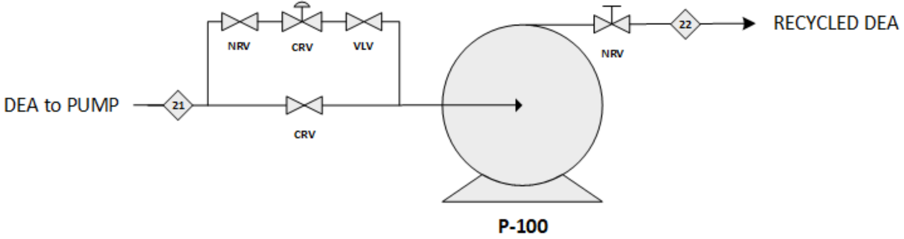

***Figure 14 Modified P&ID for P-100.***

In ***figure 14***, DEA enters the pump through stream 21, passing a control loop containing a control valve containing a pressure relief valve which is placed between the reboiler and pump. If the pressure increases to a point above the preferred tank pressure, the relief valve will open quickly and relieve the flow to the pump. A non-return valve allows DEA to flow in only one direction and to make sure there is no return of flow. Check valve is responsible for flowing bypass stream. DEA is pumped to a non- return valve in stream 22.

## VII. ECONOMIC AND INDUSTRIAL POTENTIAL

Safety considerations implementation is vital as they have various benefits, especially economically and in the oil and gas industry [38]–[40]. As the amount of contaminants, and most commonly $H_2S$ increases in the feeding stream, the damage index (DI) increases, which then results in losses which consequently and negatively affect the generated annual profit. In a studied industrial case study, the acid gas sweetening process plant for natural gas has been studied regarding the relation between the safety, economic, energy and environmental impacts, in which it has been proven that the safety considerations caused positive significant improvements in all of the interconnected aspects. The case study has proven that energy savings of $4.5*10^5$ kJ/h and the profitability increased by 16% under normal feed quality. However, with lower quality, some of the profits may be compromised to be invested into safety precautions such as those that have been previously mentioned, through the installation of valves, control instrumentation systems and modifications, safety losses could be mitigated. Consequently, this may result in a potential increase in annual profit from $ 115 million to $ 160 million per year [15], [41].

## VIII. CONCLUSION

The study has provided guidelines for design and safety considerations for the acid gas sweetening process for natural gas, in which the design made using Aspen HYSYS V10 resulted in distillation tower which required 18 stages for separation of DEA for recycle. The safety section which used DOW index and HAZOP study, highlighted that the natural gas stream had an F&EI factor of 165.1125, requiring careful handling of material and guidelines for mitigating risks was presented in HAZOP study. As a whole, this report has discussed the process of acid gas sweetening which is mainly used for the purification of natural gas from acidic gases such as hydrogen sulfide and carbon dioxide. The report's methodology has been divided into two main sections, the simulation, and the manual design of the distillation column. However, it has been deduced that the comparison was not fairly reasonable due to the fact of the different types of component mixture systems used, where the simulation calculations were based on a multi component system while the manual calculations were based on a binary system, showing great deviations between both methods. As a result, it would be recommended to assess both methods based on the multi component system to ensure a more accurate and fair comparison. Another recommendation can be made in which the distillation column can be replaced with a phase separator. This is due to the fact that water, DEA and acidic gas have boiling points far away from each other, making it easy to separate DEA from the rich amine stream and thus easier to recycle back. Thus, assuming the same composition of the rich amine stream, a two-phase separator would be a more effective alternative to the distillation column. As for the safety aspect, after analyzing the F&EI of NG, it is fair to say that one must consider safety measures and take the handling of such material quite seriously to avoid the risks of any fatal accidents, especially the various types of explosions. The HAZOP study with the modified P&IDs has shown that there are a vast number of scenarios and possibilities which can happen, and alongside these scenarios, one should take the measures required, being the installation of necessary control loops, instruments, alarms and selecting the right and effective types of valves and bypasses. The study shows great potential in being applied in the natural gas industry, in which it showed to increase profitability by 16% and contributed to energy savings of $4.5*10^5$ kJ/h.

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