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SCHOOL OF MINERAL RESOURCES ENGINEERING

Master Thesis

**BASIC DESIGN OF GAS PROCESS TRAIN IN UPSTREAM
FACILITIES**

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Abstract

Natural gas is considered as the most promising fossil fuel of the future. However; its route from the production to the end use requires a substantial processing. The scope of the thesis is the study of the available established technologies targeted to a product of commercial value. Basic design considerations of an acid gas removal unit and a dehydration are presented; where the target is the removal of H_2S , CO_2 and water content respectively.

Simulation of an acid gas removal unit using DEA and a dehydration unit using T.E.G. was conducted in collaboration with Asprofos Engineering company, by using the Aspen HYSYS Software. A North Pars Iranian gas is introduced as a feed case study. This type of gas is considered to be lean and sour. The simulation of the designed units succeeds to produce 30 MMSCFD of natural gas rich in methane and free of H_2S and water content.

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1. INTRODUCTION

1.1 Natural Gas, History and Importance

Natural gas is considered today as the most energy efficient fossil fuel. It can be described as an environmentally friendly, clean fuel when compared with other fossil fuels like coal and crude oil. Likewise, it burns more efficient and produces lower levels of potential emitted harmful byproducts. During natural gas combustion, the emissions of sulfur dioxide (SO₂) are negligible, as well as the emissions of nitrogen oxides (NO_x) and carbon dioxide (CO₂); something which consequently contributes to reduce problems associated with acid rain, the ozone layer or greenhouse gases. Therefore, a global shift from other carbon fossil fuels to natural gas -especially as far as power applications are concerned-is expected to improve energy efficiency and seems promising for a future of lower carbon emissions. At present, natural gas provides the 23,4 % of the total global energy supply, with a share that would certainly increase in the upcoming years [1,2,3].

Although the starting point of the global natural gas industry is considered as the nineteenth century; there are evidences that natural gas was primarily discovered in the Middle East between 6000 and 2000 BC. However, the first natural gas company was officially established in 1858 in Fredonia, New York (Fredonia Gas Light Company). Accordingly, the first gas pipeline-covering a distance of approximately 200 miles- was built in 1925, after the introduction of seamless steel pipes which allowed the construction of long and strong sections [1].

Turning point in the growth of natural gas applications and usages, is considered to be the quest of energy alternatives, a consequence of the crude oil shortages in late 1960's and early 1970's. Since then, the interest for natural gas exploration and production has been steadily increased, making natural gas a considerable fossil fuel energy source. The natural gas proven reserves increased from 19 trillion cubic meters in 1960 to about 190 trillion cubic meters in 2010. The uprising production rate of natural gas during the last decades can be seen in Figure 1, strengthening natural gas position in the world energy market, although certain constraints still remain, like the substantial capital costs required in production, transportation and regasification [1].

1.2 Natural Gas Reserves Distribution

In 2017 proved reserves slightly rose by 0,4 trillion cubic meters and reached a total of 193,5 trillion cubic meters with almost 50% of them distributed and controlled by

OPEC countries. Middle East holds the largest share of the global distribution; approximately 79 trillion cubic meters of proved reserves, almost 41% of the global sum. Similarly to the reserves, the production also increased by 131 billion cubic meters reaching an annual total of 3.680 billion cubic meters. In the production sector Russian Federation holds the largest share 17,3 % or 635 billion cubic meters per year.

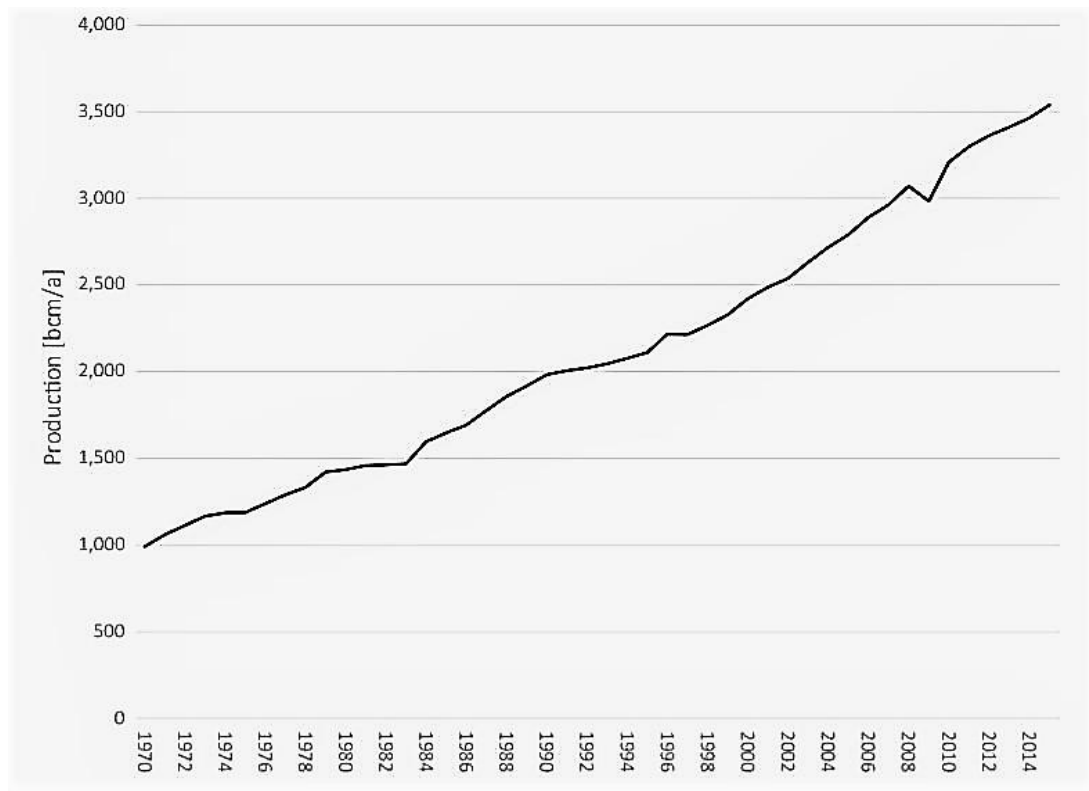


Figure 1.1 Development of worldwide natural gas production since 1970 [4]

Natural gas consumption followed the same pattern, reaching a total of 3.670 billion cubic meters per year. Although US is by far the first consumer (739,5 billion cubic meters), the uprising pattern is also strongly supported by the Chinese gas demand, where consumption increased by over 15% [2].

Table 1.1 Natural gas numbers in 2017 [2]

Proved reserves	Production	Consumption
193,5 tcm	3,680 tcm	3,670 tcm

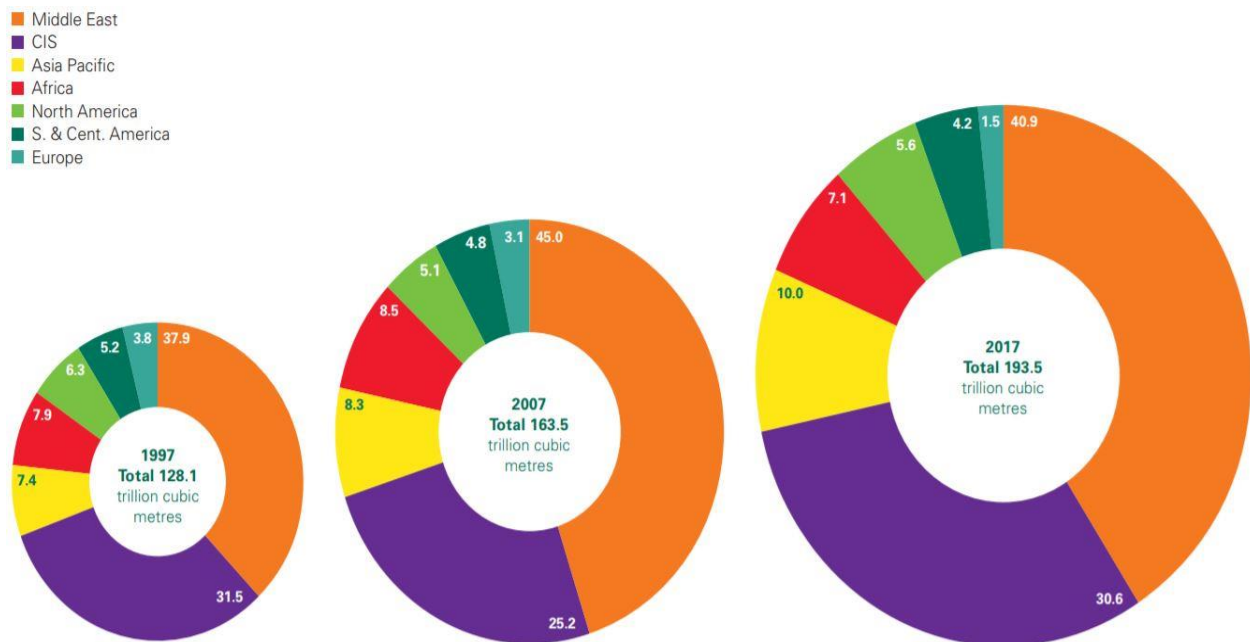


Figure 1.2 Distribution of natural gas proved reserves in 1997, 2007, 2017 [2]

1.3 Natural Gas Generation and Reservoir Classification

Natural gas is formed from organic matter that was trapped within sediments and transformed over time into their present form. Two main mechanisms are known to be responsible for the degradation of the organic matter to natural gas: biogenic and thermogenic.

- 1) Biogenic gas: is formed at low temperatures and shallow depths mainly due to the action of bacteria on organic matter deposited and accumulated in the sediments.
- 2) Thermogenic gas: is formed at deeper depths due to degradation of kerogen which was resulted as a combined effect of high pressures and temperatures. Thermogenic gas formation can be further distinguished to direct thermal cracking of sedimentary organic matter (primary thermogenic gas) and to thermal cracking of oil that has been formatted in the first stage (secondary thermogenic gas).
- 3) Abiogenic gas: In this case, the starting material is the volcanic gases (not the organic matter). Methane is formed by the reduction of carbon dioxide during magma cooling, commonly in hydrothermal systems, due to water rock interactions.

In order the gas to be accumulated three things are necessary:

- The source rock where the compaction of organic matter is taking place
- The porous media which accommodates the created gas
- The caprock which acts as an impermeable vertical barrier, a trap, that keeps the gas inside the porous media and prohibiting it to escape to the atmosphere.

Pressure, temperature and geological history have a crucial affect to the evolution of a hydrocarbons mixture and may contribute to the generation of different type of gases; something which has been observed in some cases even during the production from the same reservoir and at nearby wells [3,6].

Natural gas can be classified according to its origin and chemical composition. Conventional and unconventional gas formations is a first distinction used for the geological reserves and rock type, as well as for the production methods. Conventional gas term is used for the “free gas” trapped in multiple small porous zones in rock formations such as carbonates, sandstones etc., and it generally occurs in deep reservoirs with crude oil or not. Unconventional gas on the contrary refers to natural gas inside tight sands, shales, geo-pressurized aquifers and gas hydrates, where typically large quantities of natural gas exist but it is difficult to extract. [1,6].

As far as the coexistence of gas and oil is concerned; there is a further classification of the associated gas which coexists with crude oil and the non-associated gas which exists solely in the reservoir. Greater depths usually host lighter oils. Therefore, it is expected to discover associated gas reservoirs as the depth increases. At depth values higher than 20.000 ft, only gas is expected. The coexistence of gas with oil changes the thermodynamic behavior of the system and accordingly the reservoir production and the gas processing method. Associated gas (also called wet gas) is usually leaner in methane and richer in higher molecular weight paraffinic constituents. The gas stream traditionally (rich gas) has high proportions of natural gas liquids (NGLs). The non-associated gas (also called gas or dry gas) is produced from a geological formation that typically does not contain much, if any, crude oil or heavier than methane hydrocarbons (gas liquids). This gas is usually rich in methane and lean in higher molecular weight hydrocarbons and condensate components. It may also contain non-hydrocarbon gases, such as carbon dioxide and hydrogen sulfide. It should be noted that, after gas

processing, methane will be produced in a pure state regardless of whether the raw gas state was associated or non-associated [1].

Another classification is according to the liquid content of the gas and is characterized as rich or lean gas. Accordingly based on its sulfur content it can be characterized as sweet (negligible amounts of H_2S -less than 4ppmv) or sour (H_2S in non-acceptable levels). Although these terms can refer to both acid gases - H_2S and CO_2 - they are usually applied only for the H_2S content, as the CO_2 can be acceptable even at higher levels, as long as the heating value of the gas is satisfactory [1,3,5].

1.4 Natural Gas Composition and Phase Behavior

Natural gas can be considered as a complex mixture of hydrocarbon and non-hydrocarbon components that exist at gaseous phase under atmospheric conditions. Hundreds of different compounds can be present in the natural gas. Even wells from the same reservoir can produce gases of different compositions as the reservoir is depleted. Primarily, natural gas contains methane (CH_4), significant amount of ethane (C_2H_6), propane (C_3H_8), butane (C_4H_{10}), pentane (C_5H_{12}), hexane (C_6H_{12}) mostly in traces as well as heavier hydrocarbons. Nitrogen (N_2), carbon dioxide (CO_2), hydrogen sulfide (H_2S) and other sulfide components such as mercaptans (R-SH), carbonyl sulfide (COS), carbon disulfide (CS_2) can be present. Traces of argon (Ar), hydrogen (H_2) and helium (He) can be also found. Furthermore, traces of metallic substances are known to be present in many natural gases; such as arsenic (As), selenium (Se), mercury (Hg), lead (Pb), thallium (Th), cadmium (Cd) and uranium (U). The composition of natural gas is commonly expressed in mole fractions or mole percentages of its constituents. The heavier hydrocarbons usually are lumped together in a heptane-plus (C_7+) fraction. Natural gas composition and especially the portion of the heavier hydrocarbons in the mixture; along with the contribution of the pressure and temperature conditions under which the gas is found and produced, lead to different phase behavior of the gaseous systems such as dry, wet and condensate [1, 3,5,6].

- 1) Dry gas: this term is attributed to gas mixtures which contain mostly pure methane and heavier hydrocarbons are almost absent. This type of gas exhibits no condensation neither in reservoir pressure path during production nor in the wellbore pressure path.

- 2) Wet gas: this term is attributed to gaseous systems which produce liquid phase during production at surface conditions. It is believed that no condensation occurs in the reservoir pressure path and whatever condensates exist, they are formulated inside the wellbore pressure path.
- 3) Condensate gas: this term is attributed to gaseous systems with complex thermodynamic behavior. Condensation occurs both in the reservoir pressure path and inside the wellbore.

Table 1.2 Mole composition and characteristics of single-phase reservoir fluids [6,7]

Component	Volatile Oil	Gas Condensate	Dry Gas
C1	64,36	87,07	95,85
C2	7,52	4,39	2,67
C3	4,74	2,29	0,34
C4	4,12	1,74	0,52
C5	2,97	0,83	0,008
C6	1,38	0,6	0,12
C7+	14,91	3,8	0,42
MW of C7+	181	112	157
GOR scf/bbl	2000	18200	105000
API	50	60,8	54,7
Fluid Color	Orange	Light straw	Water white

This behavior causes deliverability and transportation problems and it is a function of composition and temperature. The knowledge of criconderm is essential in such cases, because liquid dropout occurs in the temperature region between critical point and the criconderm. Dry gas (unassociated gas) consists of methane as the major constituent with little or no C₂+component, whereas wet gas (associated gas) contains C₂+ constituents higher than 10 vol% [7]. The location of the two-phase locus in a P-T diagram depends on the composition. It can be seen respectively in figure 4 that a lean gas encloses a relatively small area in the low pressure and temperature region, while a rich gas generally exhibits a much larger two-phase region which extends to higher pressures and temperatures. For comparison, mole compositions of different reservoir fluid systems are presented. Production with gas-oil ratios greater than

100,000 scf/bbl is commonly called lean or dry gas, although there is no universally acceptable rule for distinction between the two categories.

Wet gas as a term, sometimes is used to describe what is known as gas-condensate.. Gas to oil ratios are considered to be indicative of the fluid composition. Therefore, high GOR is connected with low concentrations of heavier HCs while low GOR usually states a gas rich in heavier HCs. Under normal conditions CH_4 , C_2H_6 , C_3H_8 and butanes are in gaseous state, whereas pentanes and higher are liquid. Therefore, pentanes are also known as condensates. Propane and butanes are known as Liquefied Petroleum Gases (LPG) as they can be liquified at ambient temperatures and they are also referred as “bottle gas”. Furthermore; compressed natural gas (CNG) is produced from natural gas by compression approximately to 200 bar and has the same chemical composition with Liquefied Natural Gas (LNG) which is liquefied gas cooled down to approximately $-162\text{ }^\circ\text{C}$ [1,4]

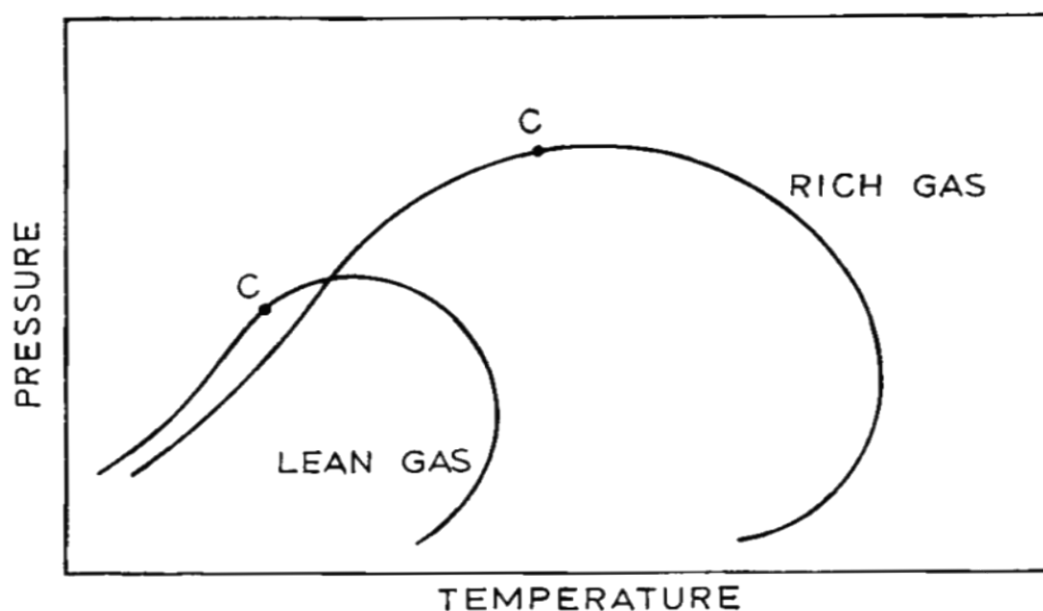


Figure1.3 P-T Diagram of lean vs rich gas [7]

Table 1.3. Typical Composition of Natural Gas Systems [1,7, 8]

Constituents	Compostion (vol%)			
Hydrocarbons	Wet	Dry	Terminology	
CH ₄	84,6	96,0	Methane	
C ₂ H ₆	6,4	2,00	NGL	
C ₃ H ₈	5,3	0,60	NGL	LPG
i-C ₄ H ₁₀	1,2	0,18		
C ₄ H ₁₀	1,4	0,12		
i-C ₅ H ₁₂	0,4	0,14	NGL	Natural Gasoline
C ₅ H ₁₂	0,2	0,06		
Hexanes	0,4	0,10		
Heptanes	0,1	0,80		
Non-hydrocarbons				
CO ₂	≤5			
He	≤0,5			
H ₂ S	≤5			
N ₂	≤10			
Argon	≤0,05			
Radon, Krypton, Xenon, etc	traces			

1.5 Natural Gas Commercialization and Usage

Transportation of natural gas is usually a challenging task due the fact that most of the time gas reserves are quite distant from the final consumer. Gas is primarily transported to markets either via pipelines or LNG ships.

Pipelines still provide a safe and economic way of transporting natural gas mainly onshore, although they are vulnerable to economic and political turmoil when they are overcrossing borders. Apart from that they also present lack of flexibility because in case of a shut-in event, gas processing plants should also stop operating. Furthermore; recompression procedures due to friction losses are mandatory, especially for long pipelines, something which can be costly.

Liquefied natural gas (LNG) technology eliminates these uncertainties and drawbacks by providing a reliable and safe, although technically challenging and costly way to transport natural gas to the consumption. Natural gas is cooled down to -162 °C at atmospheric pressure and converted to a condensed liquid which can be transported by

specially designed ships. This technology which has been developing over the last 30 years offers the opportunity to store natural gas and transfer it over oceans connecting long distance production sites and market centers, especially due the recent advances in LNG floating processing sites. On the other hand, liquification requires special cryogenic materials and regasification procedures which increase project costs. Nowadays approximately 25% of the natural gas production is being transported via LNG ships.

Compressed natural gas (CNG) technology allows the compression, storage and transportation of natural gas by using high pressure operating containers. This technology can satisfy moderate gas productions, especially those of offshore associated gas reservoirs, by providing reliable and easy market access. Low cost of processing (40% less than a LNG plant) is a strong benefit.

Gas to Liquid (GTL) technology offers the opportunity to convert natural gas into liquid products and transport it as such. GTL process are consisted of three main stages. The first one includes the partial oxidization of natural gas and conversion of it into syn gas (a mixture of hydrogen and carbon monoxide). Then catalyst procedures turn the syn gas into liquid HCs which look like wax at ambient temperatures. Isomerization and cracking are following to create liquid products like LPG, kerosen, ammonia, lubricants and other liquid products that naturally occur from crude oil. Shell opened first GTL processing plant in Malaysia in 1993. The world's largest GTL plant also running by Shell and Qatar Petroleum consortium; opened in Qatar in 2011 and covers about 8% of Shell's global production.

Gas to Solid (GTS) is another alternative for natural gas commercialization. This technology combines low temperature water and gas in order to create solid gas hydrate slurries which can be transported easily at low pressures reducing the cost of storage and transportation.

A serious drawback is the small quantity of natural gas that can be stored by this technology; when compared with LNG. BG Group, NTNU and other laboratory teams are working on this technology for over a decade and even though some positive results have been generated for commercialization of this method, no relevant project is scheduled until today [1].

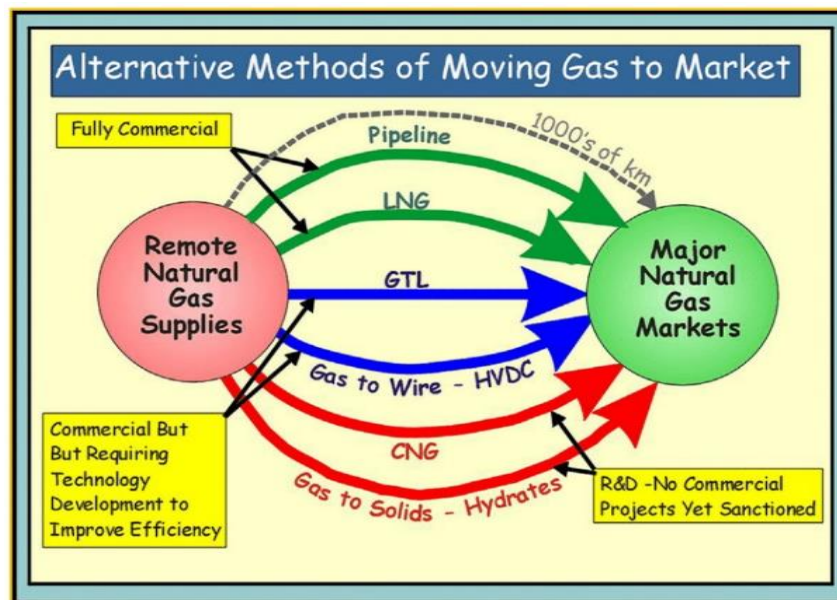


Figure 1.4. Options for natural gas commercialization [1]

Gas to wire (GTW) is a promising commercialization alternative which converts natural gas into electricity. This process allows the easy and direct transportation of the produced electricity to the demand site and introduction to the electrical grid by high voltage current cables. Serious drawbacks are the start-up and shut-down operation which lack the necessary flexibility and introduce safety risks [1].

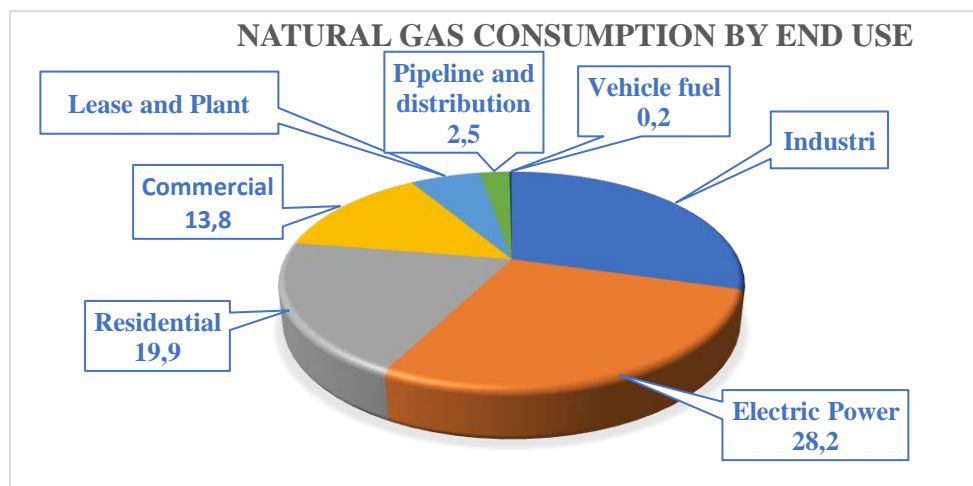


Figure 1.5. Natural gas global consumption% by end use in 2017 [8]

Natural gas provides a wide range of applications. The main sectors are:

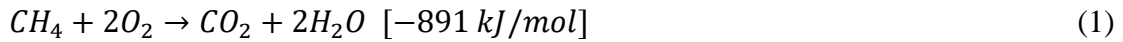
- Residential and commercial: heating
- Industrial: heating, feedstock

- Electricity generation
- Transportation fuel: limited use

1.6 Natural Gas Properties and Specifications-Gas pipeline quality

Natural gas properties strongly dependent on its composition, pressure and temperature. Therefore, parameters like gas specific gravity, molecular mass, molar volume, viscosity, deviation factor and isothermal compressibility are crucial to describe its quality. Natural gas properties can be either measured directly from laboratory tests or estimated from its chemical composition [4]. Commercial natural gas is most commonly odorless, colorless, tasteless and lighter than air.

Combustion of gas equals oxidization, which is an exothermic reaction, during which energy is released as the reaction occurs. For pure methane and oxygen, the oxidization reaction equation is the following:



The gas industry has introduced gross calorific value (GCV) of natural gas, which takes into account the condensation of all water content that have been produced during combustion. On the contrary, in net calorific value, water remains in vapor phase. Calorific value is expressed in energy per volume Btu/scf, where “s” in scf stands for reference conditions which commonly refer to the American Oilfield System Units, i.e. 60 °F and 14,7 psia. 1 Btu is the energy required to raise the temperature of 1 pound of water by 1 °F. When natural gas burns it releases approximately 1000 Btu/scf. Therefore, this number is an easy way to convert the volume of gas to its heat equivalent and thus calculate the actual energy consumption. Obviously natural gas mixture’s GCV can be easily calculated from the combination of the molar composition of each component and their specific calorific values [1,4].

Wobbe Index is defined by the following equation:

$$W_G = \frac{GCV}{\sqrt{SG}} \quad (2)$$

Where $SG = \frac{\rho_{gas}}{\rho_{air}}$, which is dimensionless and so W_G has the same units with GCV.

Table 1.4. Natural Gas Properties [1,5]

Property	Value
Relative molar mass	17-20
Carbon content, weight %	73,3
Hydrogen content, weigh %	23,39
Oxygen content, weight %	0,4
Hydrogen/carbon atomic ratio	3,0-4,0
Relative density, 15 °C	0,72-0,81
Boiling point, °C	-162
Autoignition temperature, °C	540-560
Octane number	120-130
Methane number	69-99
Stoichiometric air/fuel ratio, weight	17,2
Vapor flammability limits, volume %	5-15
Flammability limits	0,7-2,1
Lower heating/calorific value MJ/kg	38-50
Methane concentration, volume %	80-99
Ethane concentration, volume %	2,7-4,6
Nitrogen concentration, volume %	0,1-15
Carbon dioxide concentration, volume %	1-5
Sulfur concentration, weight % ppm	<5
Specific CO ₂ formation g/MJ	38-50

What actually Wobbe Index provides, is a measure of the heat input to an appliance at a given gas pressure. Natural gases of different compositions and thus different calorific values may produce the same amount of heat per unit of time if their Wobbe Indices are equal.

Dew point is also an important property which indicates the lowest acceptable gas temperature, in order to avoid the co-existence of gas with water or HCs liquid phases. Once again composition and pressure conditions hold vital role in dew points fluctuations. Elevated values of water concentration and pressure lead to elevated water dew point values. The same pattern is followed for hydrocarbons dew point.

Methane number is another important parameter, which applies for the knock resistance of a gaseous fuel. A safe operation of a given engine requires a minimum value of methane number. Pure methane has by definition a value of 100, while natural gas mixtures are affected once again by their compositions. Inert gases like CO₂ and N₂ lead to an increase of methane number, whereas heavier HCs existence results to lower methane number.

Natural gas specifications are dictated by the final use of the product. It is essential that the quality of the gas remains constant over time. For instance, gas power equipment require a stable fuel quality for equipment's efficiency and longevity. Industrial utilizations of natural gas, like in steel, aluminum, glass, pulp and paper production, also pose strict specifications to natural gas feedstock, because any variations might lead to reduced quality of the final product.

Impurities like H₂S, SO₂ which are considered as toxic, harmful and environmentally unfriendly, must be removed, as well as other corrosive sulphur components, like carbonyl Sulphur, mercaptanes and elementary sulphur, which might lead to pipe plugging. N₂ and CO₂ inert gases which do not provide any heat content to the gas stream must be removed from the final product. Water also is a challenge due to its corrosiveness and its tendency to cause blockages when it is combined with CO₂ and sulphur. Metals like mercury, which can sometimes be present in traces, react with the aluminum alloys of some heat exchangers and cause problems in further processing units of gas.

Odorization of natural gas is also something that has to be taken into account, when natural gas is about to be delivered by pipelines. The addition of an odorant is imposed by most regulators so that the gas is easily detectable in case of leakages.

Although these are the most basic challenges of gas processing, countries define different specifications for natural gas product. Neither the parameters themselves nor their values are harmonized internationally. As a result, both partners in a contract are

free to set and agree on the specifications. Such non-harmonization although it provides some flexibility, it also introduces contractual risks, exposures and safety issues.

Natural gas producers can affect the quality of a natural gas stream by:

- blending different qualities of gas
- reducing the Wobbe Index and calorific value by injecting inert gases or by removing higher HC's (derichment process)
- increasing Wobbe Index by injecting LPG's or by removing inert gases like nitrogen and carbon dioxide

In general, a commercially acceptable natural gas should be:

- A methane rich mixture
- Within a specific Btu content range.
- Delivered at a specific hydrocarbon dew point temperature level to prevent condensation
- Safe for combustion equipment and environmentally friendly, thus free of trace amounts of compounds or elements such as hydrogen sulfide, carbon dioxide, mercaptans, nitrogen, water vapor, oxygen.
- Free of water in order to prevent corrosion and gas hydrates formation
- Stable in composition
- Interchangeable and easy to be mixed with pre-existing gas in pipeline

Table 1.5. Typical pipeline gas specifications

Characteristic	Specification
Water content	4-7 lbm H ₂ O/MMscf of gas
Hydrogen sulfide content	0,25-1,0 grain/100 scf
Gross heating value	950-100 Btu/scf
Hydrocarbon dew point	14-40 °F at specified pressure
Mercaptans content	0,25-1,0 grain/100 scf
Total sulfur content	0,5-20 grain/100 scf
Carbon dioxide content	2-4 mol%
Oxygen content	0,001mol% (max)
Nitrogen content	4-5 mol%
Total inerts content	4-5 mol%
Sand,dust,gums, free liquid	None
Typical delivery temperature	Ambient
Typical delivery pressure	400-1200 psig

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2. GAS PROCESSING UNITS

2.1 Natural gas processing objectives

Most of the time, the main objective of a gas processing plant is the achievement of the adequate removal of acid gases and the production of an enriched methane mixture , free of water content, entrained solids, heavier HCs, neutral gases and impurities, which satisfies gas pipeline specifications, heating value and safety and operation requirements of the equipment that is going to be used, as well as environmental concerns and legislation limits that have to do with its combustion products. Sometimes slight changes in feedstock/raw gas properties or final product requirements can affect crucially not only the plant design, organization and technology that are going to be used, but also the cost and feasibility and a revamp might be needed. Gas processing plants are considered to be “fit for purpose” designed and therefore detailed gas well data must be available. [1,2]. A typical flow chart related to the main gas processing route from a gas well to the final consumer can be seen in figure 2.1.

Typical well streams are defined from high velocities, solid and liquid contaminants. Therefore; the gas from a well; should pass through an inlet separator which cleans the gas flow from well impurities and liquids like HCs and free water, before entering the main gas processing plant [3].

From a technical point of view, the amount of acid gases and liquid hydrocarbons that a gas contains can provide rough guidelines about the process design that must be followed. A dry gas for example requires mainly the removal of acid gases, a procedure known as sweetening, dehydration to remove water content in order to avoid hydrate formation and hydrocarbon dewpoint control in order to meet sales specifications.

A sulfur recovery unit, which is considered as an off-site support unit for sulfur production, or a nitrogen removal unit could be also a requirement depending on the gas streams composition.

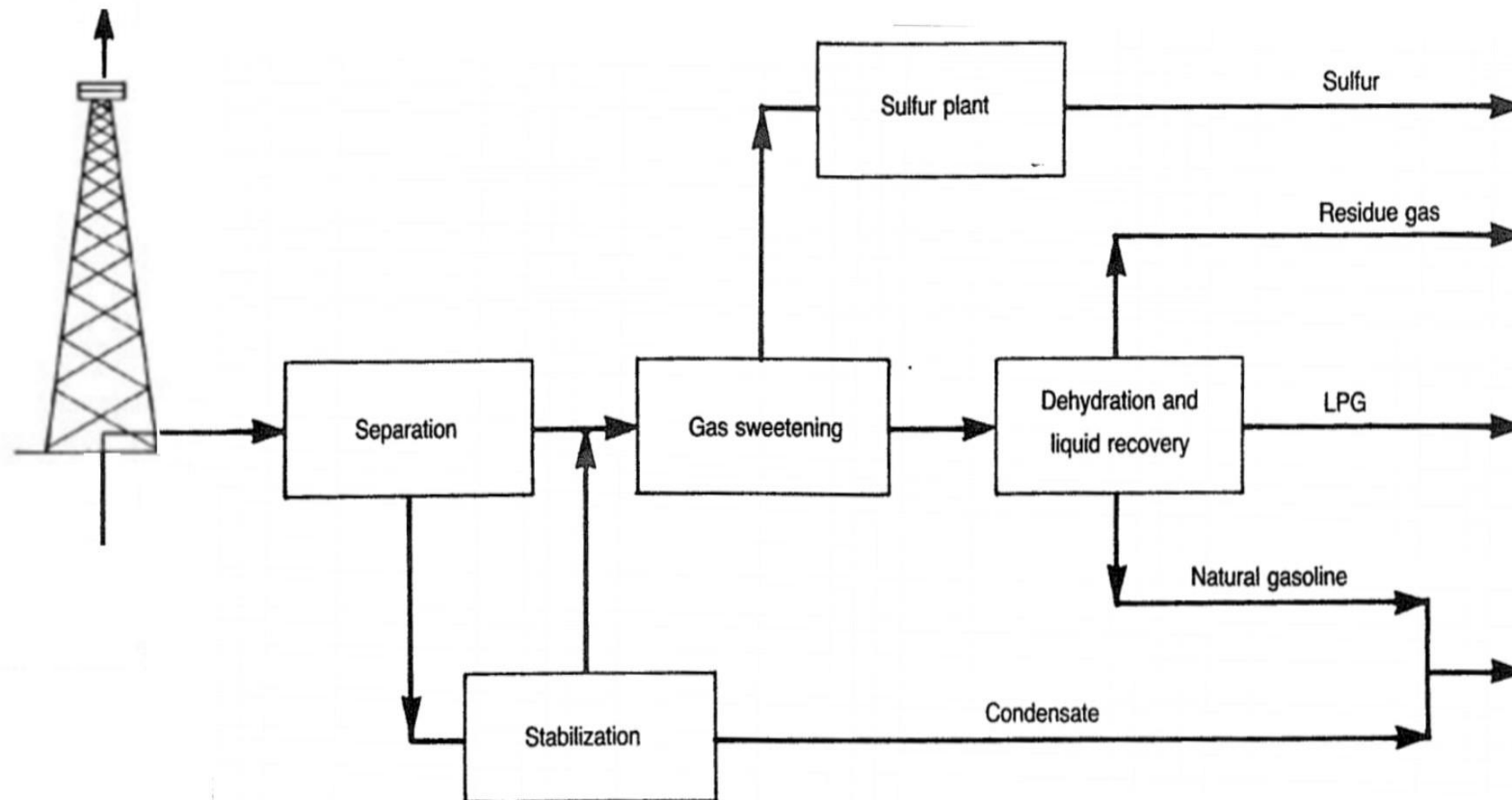


Figure 2.1 Main gas processing route units [3]

A rich in heavier HCs gas could be treated further for LPG production, natural gasoline and liquid ethane recovery (NGL recovery) in order to increase plant revenues. Most of the NGL's units are designed for 99% propane recovery. But an NGL unit which runs optimally today might not be able to treat future gas feeds with different compositions. Therefore, this lack of flexibility could mean further re-investment or renovation. Taking into account the complexity of these units and their high costs (significantly higher than a simple dew point control unit) the decision of an NGL unit construction should always be well studied. Pre-fabricated units that can be bought and transferred on site are preferable; always taking into account NGL recovery capacity [1].

2.2 Inlet facilities

Well streams are usually defined of turbulent velocities, containing free water and water vapor, HCs liquids, waxes, solids etc. A combination of foam, bubbles mist, free water and liquids may reach the entrance of the inlet facilities. Slugs that have been created due to pipeline terrain alternation, sands that occur due to pipeline erosion, aerosols, emulsions and scale that arise from the interaction between formation water and production media need to be removed or eliminated. Inlet facilities may include slug catchers, separators and pressure control equipment in order to ensure flow stabilization and a first free liquid and solid particles separation as well as to establish a protective pre-treatment unit for the rest processing sections following. Good knowledge of the liquid contaminants as well as process conditions is a requirement for the design of a high efficiency inlet facility. Slug catchers can provide a steady flow, reducing flow variations, while separators protect the plant from pressure variations and execute a first separation between different phases [1,3].

2.2.1 Slug Catchers

Slug catchers are usually located at the end of raw gas transmission pipelines and in safe distance from the rest of the processing facilities. They can be described as devices than can handle irregular incoming liquid volumes and at the same time can facilitate a initial bulk separation between liquid and gas phases. They are designed with adequate volume capacity to catch potential large liquid slugs and hold them temporarily before letting them enter the separation facilities in a manageable rate. Two types of slug catchers are the most common, the vessel type and the manifold pipe type. Vessel types are used for low expected liquid volumes. Pipe type slug catchers consist of a series of pipe fingers which create the storage volume for potential large liquid slugs. Pipe type

slug catchers have the advantage of an easy potential revamping by simply adding more pipes when large slugs are expected and at the same time for the same storage liquid capacity are less expensive than vessel types [1,4,5].

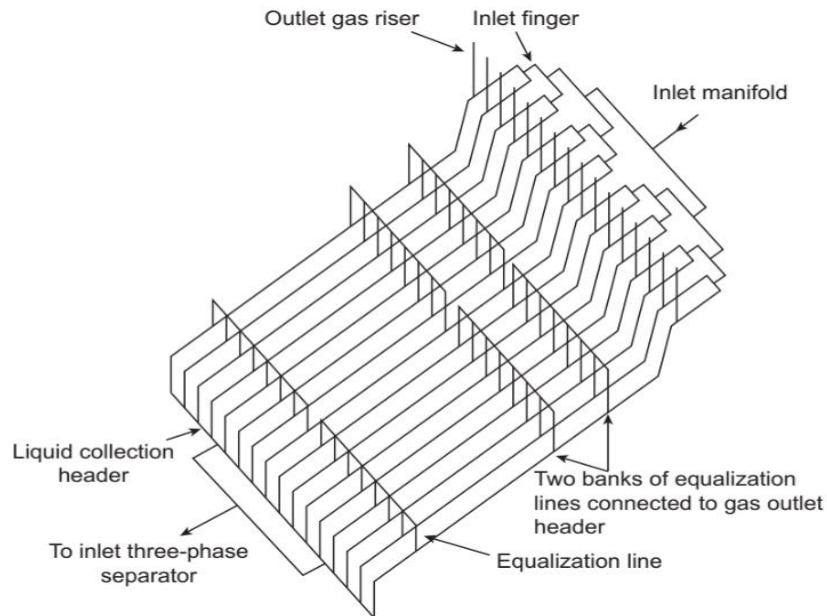


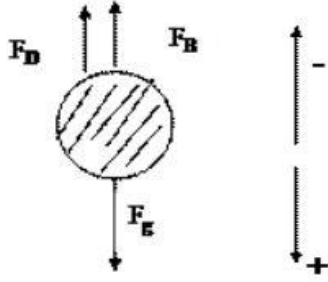
Figure 2.2 Typical configuration of a pipe type slug catcher [1]

2.2.2 Separators

There are three main separation principles:

- 1) Gravity: Density difference between fluids determines the interaction of gravity and buoyancy which leads the droplet to follow a certain direction and finally to settled down.
- 2) Introduction of additional forces: By taking advantage of the fluid momentum, a change in flow direction can lead to phase separation. Centrifugal forces are applied and density separation of fine particles occur. Electrical forces can also be applied to promote separation between charged particles.
- 3) Adhesion forces: Increase in droplet size by using devices with geometry that favors impingement and coalescing and subsequently leads to separation.

Usually more than one of the pre-described principles are combined in order to achieve the desired extent of separation. Main equations that describe gravity separation (which is considered as the dominant one) can be seen below by considering a spherical droplet settling under gravity :



Drag Force

$$F_D = C_D \times A \times \frac{1}{2} \times \rho_{fluid} \times u_{fluid}^2 \quad (2.1)$$

Where

F_D : is the drag force

A: the cross-sectional area of the droplet

$\frac{1}{2} \times \rho_{fluid} \times u_{fluid}^2$: the kinetic energy of the fluid

C_D : Drag force coefficient, a correlation factor that computed graphically

Gravity Force

$$F_G = \rho_{droplet} \times g \times \frac{4}{3} \times \pi \times \frac{d^3}{8} \quad (2.2)$$

Where

F_G : is the gravity force exceeded at the droplet which is considered spherical

$\rho_{droplet}$: droplet density

d: droplet diameter

Buoyancy Force

$$F_B = \rho_{fluid} \times g \times \frac{4}{3} \times \pi \times \frac{d^3}{8} \quad (2.3)$$

Where :

ρ_{fluid} : fluid density
d: droplet diameter

Table 2.1 Settling velocity laws

Re value	Settling Law	Settling Velocity Equation	Separation Appliance
500-200,000	Newton's	$u_t = 1.74 \times \sqrt{\frac{\Delta\rho \times g \times d}{\rho_{fluid}}}$	large droplets
2-500	Intermediate	$u_t = \frac{3.49 \times g^{0.71} \times d^{1.14} \times \Delta\rho^{0.71}}{\rho_g^{0.29} \times \mu^{0.43}}$	gas-liquid, liquid-liquid
<2	Stoke's	$u_t = \frac{1.488 \times g \times d^2 \times \Delta\rho}{18 \times \mu}$	Small droplets-high viscosity liquids

When the drag force is equal to the difference between gravity and buoyancy force, then droplet acceleration takes value equal to zero. Subsequently that means the droplet moves with constant velocity which is known as terminal or free settling velocity:

$$\frac{du}{dt} = 0 \Rightarrow F_D = F_G - F_B \quad (2.4)$$

$$u_t = \sqrt{\frac{4 \times \Delta\rho \times g \times d}{3 \times \rho_{fluid} \times C_D}} \quad (2.5)$$

Terminal flow velocity depends highly on drag force coefficient and therefore on flow pattern which can be described from Reynolds Number

$$Re = \frac{d \times U_{drop} \times \rho_{fluid}}{\mu_{fluid}} \quad (2.6)$$

Where:

d: is droplet diameter

μ_{fluid} : is the fluid viscosity

ρ_{fluid} : fluid density

Different flow patterns and therefore different Reynolds Numbers can lead to different settling velocities that are determined from three different laws which can be seen in table 2.1

As far as the settling of droplets that have been formulated due to adhesion; Souders and Brown equation provides the droplet velocity with respect to the vapor upward velocity of the gas phase:

$$Vt = KSB \times \sqrt{\frac{\rho_{droplet} - \rho_{vapor}}{\rho_{vapor}}} \quad (2.7)$$

Where

KSB: is the Souders Brown equation dependent on separator geometry, additional adhesion forces equipment design, flow rates and fluid properties.

ρ_{vapor} : the gas phase density

$\rho_{droplet}$: liquid droplets resulted from adhesion or additional forces equipment density

Main targets of separation devices are:

- Inflow energy control
- Velocity reduction which promotes and enhances gravitational separation
- Foam control
- Stabilization of flow-Elimination of turbulent fluid flow
- Assurance of solid particles removal
- Pressure handling if it is required
- Cleanout ports availability
- Provided outlet for gases and liquids and eliminate re-entrainment of the separated phases

Separators can be categorized depending on different criteria:

- 1) Geometry: a) Vertical, b) horizontal c) spherical
- 2) Phase separation: a) Two phase separators where gas phase is separated from a liquid phase b) Three phase separators where gas phase and two discrete liquid phases (oil and water) are separated
- 3) Operating Pressure: a) Low pressure separators which can operate between 10-180 psi b) Medium pressure separators which operate between 230-700 psi and high-pressure separators which operate between 975-1500 psi.

Regardless of shape or orientation; most commonly used separation devices usually contain three main sections:

- 1) Inlet section: Inlet section reduces momentum and facilitates a rough initial separation between gas and liquid phase while ensuring gas flow distribution and prevents re-entrainment of liquid phase. Fluids are usually subjected to a centrifugal force up to 500 times bigger than the force of gravity. The options of inlet section that are commonly involved are:

- No inlet
- Diverter plate
- Half-pipe
- Vane type
- Cyclonic

Vane type and cyclonic inlet devices are considered to be the most efficient although diverter plate is the most commonly used.

- 2) Main separation department: Sufficient length and height are provided in order gas-liquid gravity separation to take place. Retention time holds key role and allows the liquid droplets to settle down and be separated from each other and from the gas phase respectively. Surge volume is important in order to smooth the flow. Defoaming plates and weirs also promote phase separation.
- 3) Mist Extraction-Outlet section: Mist extraction section facilitates gas-liquid separation that did not take place in previous sections by removing very small droplets which coalesce together on a knitted mesh pad or through a series of vanes or cyclone tubes [1,2,4,5].

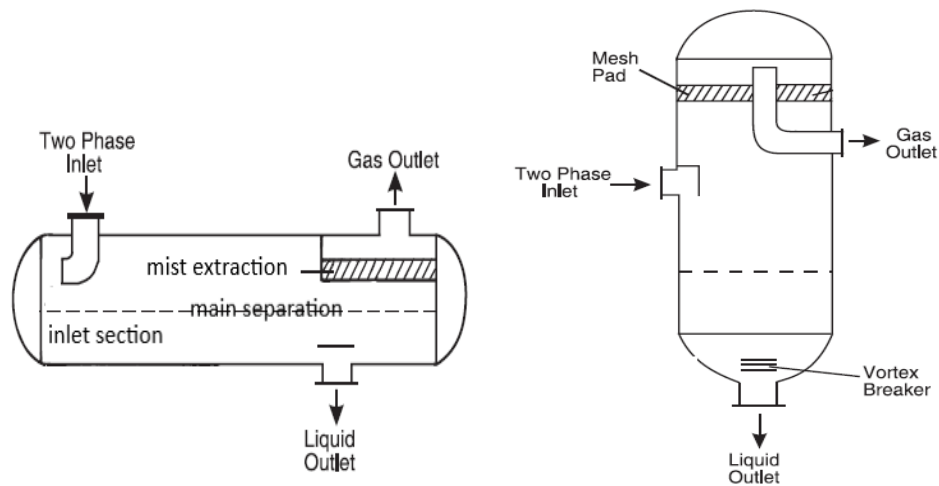


Figure 2.3 Main sections of most common gravity separation configuration

Table 2.2 Separation technology efficiency [1,2,4]

Technology	Droplet size
Gravity Separation	300 μm
Centrifugal Separation	8-10 μm
Mist pads	10 μm
Vane separation	10 μm
Liquid-Gas Coalescing	0.1 μm

2.2.3 Considerations of Separation Facilities

Key design considerations that must be taken into account are the following [5]:

- Liquid inlet and rough estimation of expected flow variations
- Gas flow rate
- Specific gravity and density of each phase
- Settling velocities and retention time
- Temperature and pressure conditions
- Solid content
- Foaming tendencies

Table 2.3. Separator Configurations Operational Conditions [1]

Separator	Pressure	Gas Throughput	Phases	Solid Content
Horizontal	High-Medium-Low	Moderate	2-3	-
Horizontal Double barrel	Moderate	High	2	-
Horizontal with a boot	Moderate	High	2	-
Vertical	Moderate	Moderate-Low	2	High
Spherical	High	High	2	-

2.3 Acid Gas Removal Unit

Acid gas removal, also known as gas sweetening process, removes the acid gases H_2S , CO_2 , which may be present in natural gas, as well as other sulfur contaminants such as $R-SH$, COS and CS_2 . The main target of the process is to produce an output almost free of H_2S , leaving a concentration of 4 ppmv or $\frac{1}{4}$ grain sulfur per 100 scf. This means that the outlet stream must contain approximately maximum 0,0004% mol of H_2S , to avoid toxicity effects and corrosiveness. Accordingly, CO_2 , a corrosive gas that lowers final product's Btu content, must take values inside the range of pipeline specifications limits (2-4% mol). Further reduction down to 2 ppm is needed if the treated gas is going to be processed in an LNG plant. Both of gases are considered to be harmful for the environment, especially H_2S which presents high toxicity [1].

Many options are available for gas sweetening. These options have their basis on one of the following mechanisms [1,2,4,6]:

- Absorption: It occurs when a liquid phase absorbs a soluble to it component of a gas stream. It is considered as the most important mechanism in gas treatment.
- Adsorption: Very similar to the absorption. The separation of one of the gas constituents (adsorbate) can be achieved due to its adherence on the surface pores of a solid (adsorbent). This procedure is considered reversible and desorption can be achieved by simply increasing the temperature and breaking the weaker bonds between the adsorbate and the adsorbent. Chemical adsorption on the other hand is irreversible

- Membrane permeation: A permeable membrane allows only the selective passage of gas through it, holding at the same time certain components which previously were dissolved in the gas stream
- Chemical conversion: The chemical transformation of an undesirable component to another substance which is easy to be removed [1,7,8,9,10].

The sweetening processes can be also specified as:

- 1) Chemical solvent processes: During these procedures chemical absorbers absorb the acid gas and some COS. These absorbers are applicable when low concentrations of acid gas are required to the final treated stream or the partial pressures of these components are relatively low in the feed. Although these solvents cannot absorb mercaptans due to low solubility, the relatively low co-absorption of hydrocarbons is a considerable advantage. Alkanoamines and hot potassium carbonate solution K_2CO_3 are the main options; although the last one is not favorable anymore due to reduced effectiveness and co-production of hazardous materials difficult to dispose [1,2,7,8,9,10].
- 2) Physical solvent processes: In this procedure there is absence of chemical reaction. The solvents are adequate to treat high pressure streams but require relatively low temperatures and more processing stages. Non-corrosiveness is an advantage when compared to chemical solvents. On the contrary, they present high co-absorption with hydrocarbons. Most common options here are fluor solvents like propylene carbonate (PC), Dimethyl ether of polyethylene glycol (DPG) which is used in SELEXOL Process, methanol (MeOH) which is involved in Rectisol Process (Lurgi Company) and N-Methyl-2-pyrrolidone (NMP).
- 3) Mixed Solvents: These hybrid solvents can combine high acid gas selectivity at elevated pressures and efficient removal of R-SH and COS. Most of these solvents are trademarked like Sulfinol UCARSOL (Dow Chemical). Although they considered to be adequately effective for the process targets and sometimes comparable to alkanolamines, they are ineffective when treating low pressure streams and can present elevated hydrocarbon losses. Complexity is also another possible drawback as more process steps must be added. MDEA is the most popular chemical solvent that is used in mixed solvents.

- 4) Solid bed absorption and adsorption process: A solid fixed bed is used as the absorbent mean in order to strip the gas effluent of its H_2S , either through a chemical reaction or ionic bonding. These solvents are effective in removing R-SH and other impurities like mercury and chlorides and they also present good H_2S selectivity. It is important to note that some solid bed absorption processes can produce treated gas streams with only 1 ppm of H_2S . On the contrary most of them are restricted by tight ranges of operating conditions like pressure and temperature, while those who are not heavily depended on conditions, may produce high quantities of hazardous materials which need special disposal. PH control also sets considerable constraints, but the greatest drawback is the fact that the system must shut down in order to replace the solid bed when it is spent. Most popular solid bed processes are the iron sponge and zinc oxide methods. PURASPEC which developed by Johnson Matthey is also gaining points, but its high cost is major disadvantage. Other process are the iron oxide slurry processes like Chemsweet and Sulfa-check both established by NATCO, but they are considered to be less effective in CO_2 removal and can possibly cause foaming, plus the fact that the unit must shut down in order to replace the used slurry. Molecular sieves based on zeolite, activated alumina and silica are the most common adsorption processes with the advantage of a wide pressure range, but the narrow range of pore size selection and possible hydrocarbon losses set limits.
- 5) Membrane based processes: These processes have their base on diffusion. Semi-permeable membranes ensure the removal of H_2S , CO_2 and other impurities. They are easy to install and do not need any specific supporting utilities or supervising during their operation. On the other hand, a pre-treatment unit is needed because these membranes are vulnerable to HCs corrosiveness, and precipitations, uplifting the cost of the overall process. Apart from that only few reliable manufacturers are available globally.
- 6) Cryogenic distillation: It applies extended cooling to remove acid gases by reducing their temperature low enough to cause condensation. Main target of this technology is the CO_2 removal and it can be beneficial where high pressure CO_2 is needed in neighbor wells for gas injection. This separation processes requires dehydration pretreatment in order to protect the system from hydrate

formation. Ryan Holmes and FZ (ExxonMobil Upstream Research Company) have been developed.

- 7) Gas treatment with microbes: A new perspective in gas processing has been established from Shell Global Solutions THIOPAQ process, in order the H_2S to be removed. Besides that, this technology seems to have the capability to replace sulfur recovery unit as well as tail gas treatment [1,2,3,5,7,8,9,10,11].

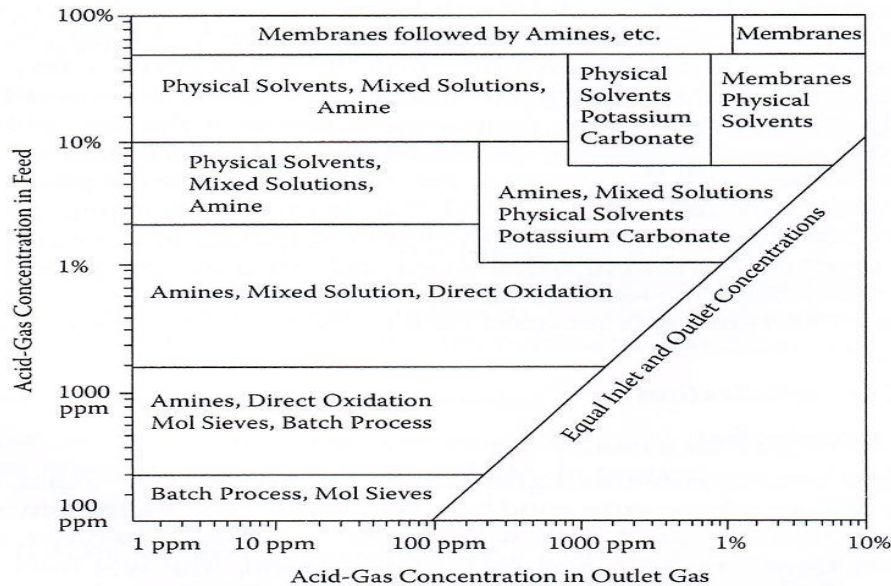


Figure 2.4 Sweetening processes. Feed and outlet acid gas concentration[11]

The selection of the gas sweetening process should be based on the process objectives and therefore:

- Gas flow (volume, composition, P&T conditions)
- Site location and ambient onshore/offshore temperatures
- Ease of water content control
- Type and concentration of impurities, purity requirements and product specifications
- Process capabilities
- Feasibility and desirability of a sulfur recovery unit
- Capital and operating costs
- Environmental restrictions
- Safety and reliability

Table 2.4 Gas Sweetening Processes Efficiency [1,11,2,6]

Process	Gases Removed			
Chemical Solvents	H ₂ S	CO ₂	RHS	COS-CS ₂
MEA	✓	✓		✓
DEA	✓	✓		✓
TEA				
DGA	✓	✓		✓
MDEA	✓			✓
DIPA	✓	✓		✓
K ₂ CO ₃	✓	✓		✓
Physical Solvents				
PC	✓	✓	✓	✓
Selexol	✓	✓	✓	✓
Rectisol	✓			
NMP	✓	✓	✓	✓
Hybrid Solvents				
Sulfinol	✓	✓		✓
UCARSOL	✓			
Solid Bed				
Molecular Sieves	✓	✓	✓	✓
Zinic Oxide	✓			
PURASPEC	✓			
Slurries	✓			
Chemsweet	✓			
Sulfa-Check	✓			
Membrane Based				
Polymers	✓	✓		
Cryogenic				
Ryan Holmes	✓	✓		
FZ	✓	✓		
Microbial				
THIOPAQ	✓			

2.3.1 Acid gas removal with chemical solvents absorption process-Alkanolamines as an option

Alkanolamines are the most efficient chemical solvents used in chemical solvent absorption of H₂S and CO₂. They are widely preferred in acid gas removal because they provide an adequate gas purification, high thermal durability, stability and high H₂S selectivity. Besides that, their availability and moderate cost make them commercially popular. Alkanolamines are amine aqueous solutions that have not particular color, with an intense and sharp odor. They are generated from NH₃ molecule with substitution of one or more hydrogen atoms by a hydrocarbon group that also contains a hydroxyl. This magnitude of substitution classifies them in primary, secondary or tertiary alkanolamines. Structural details can be seen in figure 2.4

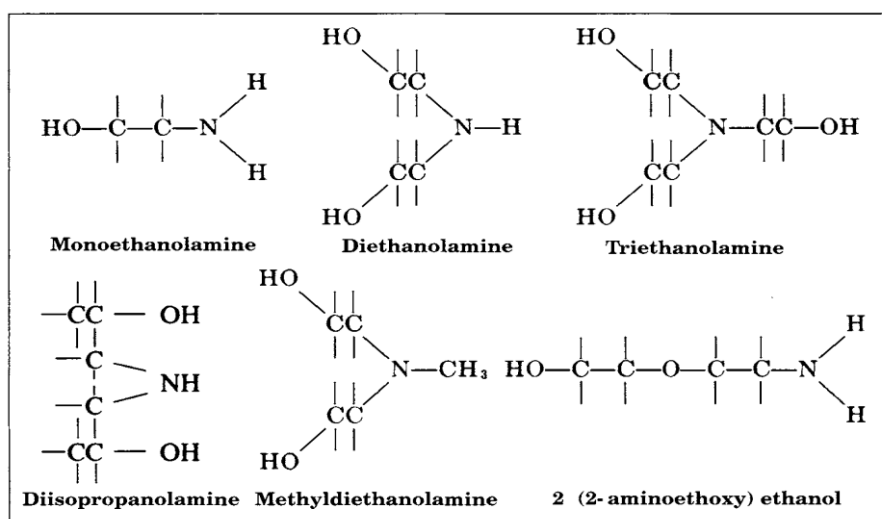
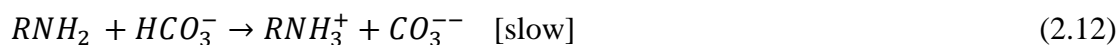
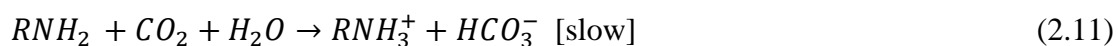


Figure 2.5 Alkanolamines Structural Details [6]

The amino groups increase the alkalinity of the aqueous solution which promotes acid gas removal where the hydroxyls decrease vapor pressures, fact that empowers water solubility. Simple elevation of temperature increases the vapor pressures of absorbed gases and helps to recover them from the aqueous solutions[4,6,10]. The basic reactions that take place are the protonation of RNH₂, the hydrolysis of CO₂ and carbamate formation which can be seen below:





It can be clearly seen that H₂S reacts directly and fast with the amine solutions whereas CO₂ can follow two paths, the fast and the slow one[10].

2.3.2 Primary Amines-Monoethanolamine (MEA) and Diglycolamine (DGA)

Primary amines like M.E.A and D.E.A. are considered to form strong bases and present strong alkalinity. Therefore, they behave more aggressively and react fast enough with both acid gases leading to a stable carbamate through a rapid path [10].

M.E.A. can be very effective when low pressure inlet streams are present and low concentrations of acid gas are needed in the treated stream. Due to its high alkalinity it is very efficient and can remove both gases down to very low levels. On the contrary M.E.A. reacts irreversibly with COS and CS₂ and a regeneration process stage might be needed. Even with regeneration, reversal of the chemical process is not always achievable due to solvent degradation and the occurring formation of heat stable salt products and corrosive compounds. Consequently, some solvent loss in such cases must be expected. Besides that, M.E.A solution concentration cannot reach values more than 20% due to possible corrosion. Gas pick can reach 0,35 mol of acid per 1 mol of M.E.A. It is considered as a good option for low acid gas loadings [1,6,10].

D.G.A. can be used in a solution strength of 50 up to 60 mass percentage and therefore can remove much more acid gas than M.E.A. This higher concentration reduces the need for circulation and therefore energy consumption. It presents also higher alkalinity than M.E.A. and better selectivity against CO₂ as well as efficiency against COS and CS₂. Urea and thiourea can be produced from degradation and a thermal regeneration is needed. It's lower freezing point creates an advantage in cooler climates against M.E.A.

2.3.3 Secondary Amines- Diethanolamine (D.E.A.) and Diisopropanolamine (D.I.P.A.)

D.E.A. is a secondary amine which is considered to be not only effective against CO₂ and H₂S but also against COS and CS₂. Although it can manage almost all types of impurities, due to low vapor pressure and heat of reaction it is preferable for high pressure operations and for high concentrations of H₂S in raw gas stream, but on the other hand sometimes it is not able to provide pipeline gas specifications at low

pressures. Reduced heat of reaction compared with M.E.A. means less cost. Besides that, a reclaimer is not a requirement and degradation byproducts are not that much corrosive. It can remove 0,35 to 0,85 mole of acid gas for each mole of its aqueous solution which has typical 25-30 mass percentage of D.E.A. Higher loadings may require the usage of inhibitors unless stainless steel equipment is engaged.

D.I.P.A. is another secondary amine with many reactivity similarities with D.E.A. It is very popular in Europe and Asia processing plants as it is very effective against H_2S and CO_2 when it is used in higher pressures. Furthermore, it can treat adequately COS and CS without producing harmful byproducts. Its aqueous solution is not corrosive and it is also combined with physical solvents in Sulfinol process[1,6,10].

2.3.4 Tertiary amines Methyldiethanolamine (M.D.E.A.) and Triethanolamine (T.E.A.)

Tertiary amines have no free hydrogen connected to nitrogen and therefore carbamate cannot be rapidly formatted. So, CO_2 removal follows a slow route through bicarbonate formation reaction, which means that these amines are effective against H_2S but not fully selective against CO_2 , except at elevated pressure operations where theoretically a 1 mole loading of CO_2 per mole of amine can be achieved.

M.D.E.A. is the most widely used amine in gas treating nowadays when H_2S elimination is considered. Low solution losses and high H_2S selectivity in the presence of CO_2 , low corrosiveness, low energy consumption and smaller reclaiming unit as well as degradation resistance are the most important advantages of M.D.E.A. M.D.E.A. can also be effective against CO_2 when boosted and combined with chemical activators, requires low absorption temperature and the mass percentage in its aqueous solution is 30 up to 50%. M.D.E.A can remove 0,8 mol of acid gas per mole of aqueous solution.

T.E.A. was the first amine used in gas sweetening and was also a popular solvent for CO_2 removal in ammonia production plants, where Haber-Bosch process is used, but over the years was almost totally replaced from other amines like MEA because it was not able to produce a pipeline gas within limits specifications [1,6,10].

2.3.5 Description of amine sweetening processing

Sour gas stream enters the inlet scrubber/separator where solids and liquids which might have been carried out through the pipes are removed. Then the clean sour gas

stream enters the bottom of the contactor/amine absorber and flows countercurrent to the lean amine solution which has been introduced to the absorber from the top entrance. The temperature of the lean amine stream entering the contactor must be at least 10-15 °F above the temperature of the feed to avoid any hydrocarbon condensation. Inside the contactor, which usually operates within a temperature range of 120 to 200 °F, acid gases are stripped out from the sour gas stream due to their contact with the amine solution [1,4,6].

The temperature starts to increase usually four trays from the bottom. Avoidance of increased temperatures is highly advised, to avoid possible amine losses, otherwise an extra separator for the sweet gas stream must be considered. In hot climate regions, especially in desert plants, cooling of the feed gas stream is highly advised to avoid possible temperature equilibrium across the height of the absorber which could downgrade the absorber functionality. A typical absorber consists of 20 trays where absorption occur. The number of the trays as well as the location of the feed entrance affect significantly the separation. Higher feed location means more contacting stages but also less CO₂ slippage. Lower feed location results to more H₂S absorption and less CO₂ removal. Multiple feed points could also provide a possible CO₂ removal improvement. The overall result of the absorber operation is a rich (H₂S, CO₂) amine solution leaving the contactor from its bottom and a sweet gas stream exiting the top of the absorber. The rich amine solution containing high amounts of acid gases, passes through a flash drum/tank where skim off and HC's recovery is executed. Then it heads to a heat exchanger where it absorbs heat from the lean amine solution in order the heat duty both on the reboiler stripper and the solvent cooler to be reduced. Afterwards the rich amine solution enters the amine stripper where bond breaking is taking place through elevated temperatures and heat provided from the reboiler. Acid gases are separated from the rich amine solution which has been introduced from the mid portion of the stripper. The stripped gases move to a condenser and separator which cools them down and recovers water which is typically returned to the stripper as a reflux. On the other hand, the acid gases are either vented if they satisfy regulation limits or sent to a sulfur recovery unit. Sometimes if the gas processing is taking place near injection well, these gases might be used in enhanced oil recovery and gas injection purposes. Heated lean amine exits the bottom of the stripper and goes through the lean/rich heat

exchanger and a solvent cooler before it enters the absorber as it was pre-described. Heat exchanger and cooler help to reduce temperature and keep the lean amine solution temperature approximately 10°F warmer than the inlet gas temperature in order the amine to stay above HC's dew point. A surge tank/amine storage tank is usually placed before the absorber in order to maintain a steady circulation and prevent flow fluctuations. Typically, this amine storage tank is designed to adequately manage the total volume of the amine inventory. This tank also helps to control the amine solution quality when is needed. A filtration system is also used after the storage tank and helps maintaining the amine solution within purity and efficiency specifications. Besides that, removal of solid degradation products and other absorbed contaminants ensures the minimum possible corrosion and prevents equipment fouling. Sometimes a reclaimer might be needed for the lean amine stream leaving the bottom of the stripper, especially when solutions of MEA and DGA are used, due to high occurrence of degradation. This reclaimer typically operates as a semi-continuous batch operation treating heat stable salts and degradation byproducts. Reclaiming equipment is usually designed and provided from the amine suppliers or from a reclaiming company which can be contracted[1,4,6,10,11,12,13].

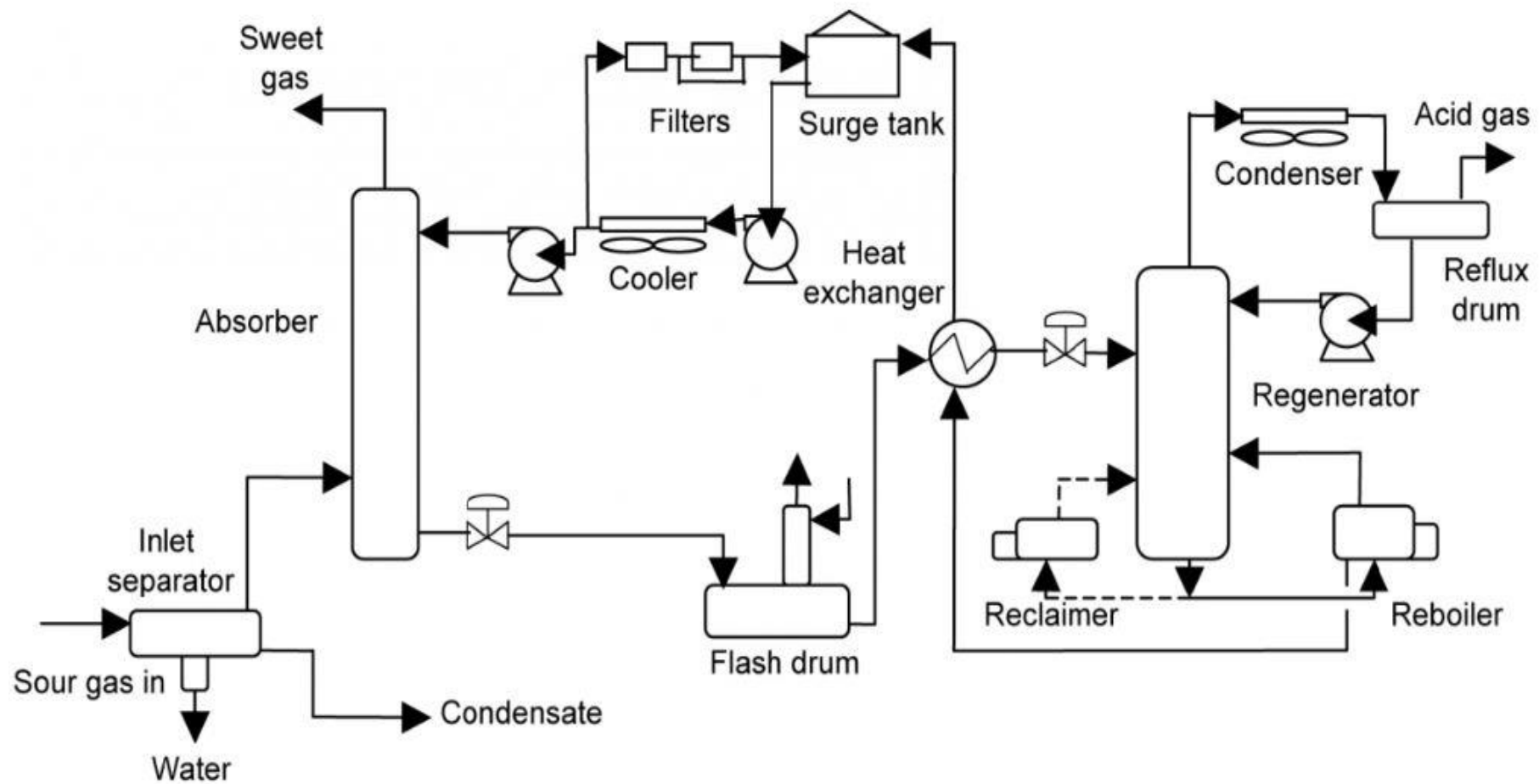


Figure 2.6 General flowchart of a gas sweetening processes using amine solution [1]

2.3.6 Amine Gas Sweetening Unit Design Considerations

1. Inlet separation: Although inlet facilities proceed with a first separation and contaminants removal, any undesirable content left could affect negatively the amine sweetening unit. Foaming, corrosion, blockages would set in danger the operation of the unit. Therefore; emphasis must be given to the appropriate separation device design for preventing all these potential dangers. Real inlet gas data over adequate periods of time are necessary to ensure proper designing [1,10].
2. Contactor: Pressure control is crucial. Any variation in pressure difference or liquid level could cause potential foaming problems and reduce contact between countercurrent fluids and therefore H₂S removal. Antifoaming inhibitors should be used if such situation occurs[1,10].
3. Amine losses: Amine solution losses could significantly affect the operational costs of an amine sweetening unit. Any amine quantities that have been swept away in sweet gas stream should be recovered by a separator[1,10].
4. Filtering: A two stage filtration that can adequately remove particles down to 5µm should be taken into account. The filter system usually must be capable of treating approximately 20% of the main streams circulation rate and it can be placed either before the lean amine solution enters the contactor or after the rich amine solution leaves the bottom of the contactor. A first filtration stage holds solid particles down to 10µm and a second filter of activated carbon removes particles down to 5µm as well as any possible HC's. Certain techniques such as gas chromatography and mass spectrometry can identify the contaminants which may occur inside the processing streams and help to adjust the filtration. Most of these filters can be regenerated when their beds are saturated. This regeneration is usually taking place by using steam and foaming formation is a good indication to apply filter regeneration[1.10].
5. Amine regeneration: Low pressures and moderate temperatures should be taken into account as far as steam streams are concerned in the amine stripping procedure to protect the equipment and prevent amine degradation. Different amine types require different handling although an average pressure of 350 kPa and 140oC of steam is considered optimal. Condensate controller should be set on the steam inlet to prevent condensed liquids form entering the reboiler tube; a situation that may fire up flooding [1,4,10].

6. Lean-Rich amine heat exchanger: High operational pressure would increase the cost of the heat exchanger, but at the same time it could minimize the escape of acid gases from the rich solution and prevent corrosion effects. Low fluid velocities (about 0.8 m/s) although may increase capital investment due to large surface area required and reduces the operational and maintenance costs.
7. Corrosion: Generally high operating temperatures and pressures enhance corrosion in an amine sweetening unit. Reboilers and stripping columns tend to be the most sensitive areas. H_2S to CO_2 ratios also affect the corrosive behavior of the streams. It is believed that acid gas streams containing more H_2S than CO_2 tend to be less corrosive. H_2S in the presence of free water forms a weak acid which attacks iron surfaces and leaves behind iron sulfide (FeS). A filter cake of created FeS could offer some permanent protection to the metal surfaces but it can be swept away easily and expose new untouched surface to acids. CO_2 with water results to carbonic acid (H_2CO_3) which also attacks iron to create either the soluble ferrous bicarbonate ($\text{C}_2\text{H}_2\text{FeO}_6$) or the insoluble ferrous carbonate (FeCO_3) which can be hydrolyzed further into iron oxide (Fe_2O_3). H_2S could potentially react with Fe_2O_3 and create precipitation of FeS . The amine solution concentration also affects general corrosion of the system. Stronger solutions tend to be more corrosive especially when high acid gas loadings are treated [1,4,6,10].

2.4 Dehydration Unit

The main purpose and target of this unit is to adequately remove water content from the gas stream. Although most of the water content (free water) associated with gas is removed at the inlet facilities by separation devices; water vapor still remains. Dew point of natural gas is indicative of its water content. At dew point pressure-temperature, the mixture of natural gas and water presents an equilibrium between gas and liquid water phase. Therefore; a slight pressure incline or a temperature reduction could ignite water condensation. A dew point depression of natural gas mixture and subsequently the removal of the water vapor content is crucial, as water through condensation could create hydrates [$\text{C}_n\text{H}_{2n} \cdot 18\text{H}_2\text{O}$]. Favorable conditions for hydrate formation are low temperatures and high pressures and of course the co-existence of moisture and natural gas. Gas hydrates can be described as trapped HCs molecules inside a water crystal lattice. The presence of gas makes the lattice stable. Pipeline

plugging is a common problem as far as gas hydrates are concerned. Gas hydrates agglomeration could increase significantly the pressure differential across a formed plug; setting equipment and the pipeline network itself in danger of malfunction or even destruction. Besides that, remnant water combined with H_2S and CO_2 could cause corrosion problems and even erosion due to acid formation. The products of erosion subsequently could create further problems when introduced in the downstream process equipment; reducing their efficiency. Furthermore; liquid plug created by water vapor condensation could reduce pipeline capacity and increase friction. Another consideration that has to be taken into account is that natural gas heating value is reduced when contains water content above the desirable limits [1,4,6,10,11,12,13].

2.4.1 Principles of Dehydration

Three main principles are used to remove water content.

1. Absorption: Natural gas is stripped of its water content by flowing countercurrent with respect to a hygroscopic liquid (usually a glycol solution). The two streams are brought in contact inside an absorber consisted of trays and due to the glycol solution water affinity, water vapor is removed from the gas which travels upwards inside the contactor. Usually a number of trays up to 10 is enough for a complete stripping. The rich glycol after exiting the contactor flows to a regenerator in order to become free water and ready for another cycle[1,4,5,6,7,8,9,10].

The absorbent solution as an must present the following properties:

- Absorption efficiency with respect to the water-good dehydration ability
- Inert with respect to the hydrocarbons and absence of absorption HCs efficiency
- Non-corrosiveness
- Low viscosity in operating temperatures
- Simple and low cost recovery procedures

Calcium chloride ($CaCl_2$) and glycerol solutions were also used as dehydrating agents in gas industry. Nowadays the most commonly used glycols in gas industry are: Monoethylene Glycol (MEG), Diethylene glycol (D.E.G.) and Triethylene Glycol (T.E.G.).

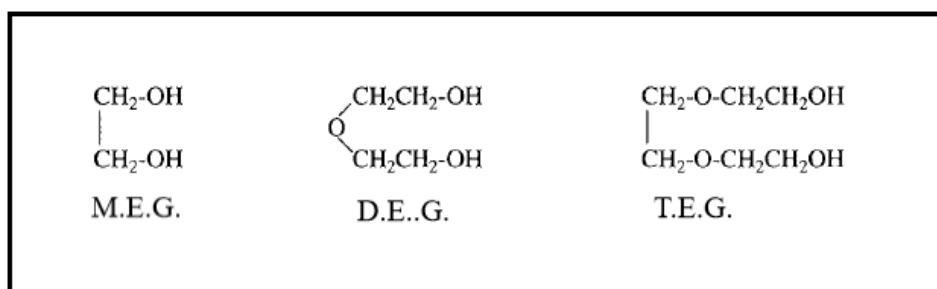


Figure 2.7 Structural information of industrial glycols[6]

M.E.G.: Presents elevated vapor pressures and cannot perform in ambient temperatures due to gas losses. Low decomposition temperature. Difficulty to regenerate. On the other hand it is good for preventing hydrate formation. Suitable for glycol injection purposes.

D.E.G.: Due to high vapor pressures could cause gas losses. Low decomposition temperature.

T.E.G.: Low vapor pressure compared to the pre-described glycols. High decomposition temperature allows adequate regeneration at elevated temperatures producing high purity glycol.

Table 2.5 T.E.G. operating conditions range [1,4,7]

Parameter	Range Value
Dew point depression (°F)	40-100
Gas pressure (psig)	25-2500
Gas temperature (°F)	40-160

Therefore, T.E.G. is the most applicable dehydration agent. Blends of more than one glycol are also used due to their low cost as manufacturing products, but their relatively moderate efficiency is a drawback

Table 2.6 Properties of industrial dehydrating glycols [1, 6, 7]

Properties	M.E.G.	D.E.G.	T.E.G.
M.W.	62,1	106,1	150
Boiling Point (°F)	387.7 F	474.4	288
Decomposition Temperature (°F)	329	328	404
Freezing Point (°F)	9.1	17.6	19
Viscosity (cp) at 77 °F	16.5	28.2	37.3
Density (g/ml) at 77 °F	1.110	1.113	1.119

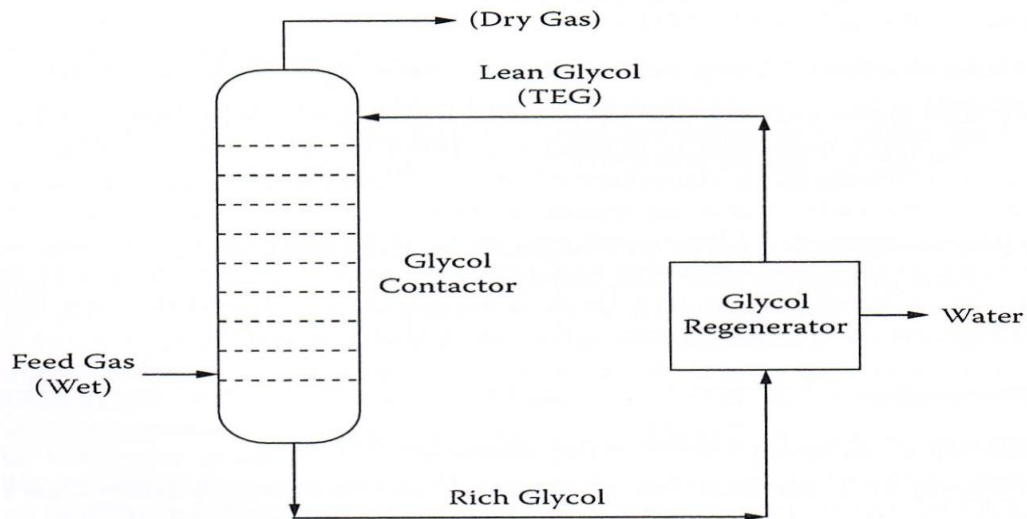


Figure 2.8 Gas dehydration unit with glycol solution [11]

2. Adsorption: Gas flows downwards an adsorption tower filled with solid bed desiccants which can selectively hold on their surface the vapor water content. This can be achieved because of the large effective area of a solid agent, due to its very small pores and capillary openings. Dry gas exits the tower from its bottom. In contrast with absorption process; natural gas flows from the top to the bottom of the tower passing through the bed. The water that is removed from the gas depends on the adsorbent agent; a parameter called adsorption capacity or loading. Adsorbent capacity is typically defined as the mass of water removed per mass of solid desiccant. Typically, a number of towers is needed for the regeneration process to take place when solid bed is fully loaded with impurities. This can be done by introducing a heated gas flowing upwards from the bottom which triggers water evaporation and leaves the tower ready for another dehydration cycle [1,4,6,10,11,12,13]. Typically required adsorption properties are:

- Elevated value of adsorption capacity. High values of this property mean lower volume of desiccant and therefore reduced volume of vessels which subsequently results to lower cost.
- High adsorbent selectivity which promotes the targeted removal of the impurities without affecting the gas stream.
- Low operating temperature which promotes easy regeneration

- Adequate mechanical strength and stability

Bauxite, silica granular gel (SiO_2), activated alumina (Al_2O_3) or molecular sieves are used as dehydration agents. All of them are designed for very high pressures and large gas volumes. Usually a series of towers is needed due desiccant material saturation with water after long operation periods which requires regeneration cycles to take place. Desiccants could last up to 4 years as their effective area is reducing through the usage.

Silica gel exhibits satisfactory loading capacities which reach up to approximately 8% by weight water adsorption. Although it is considered relatively inert as far as H_2S is concerned; feeds containing more than 5% H_2S could create surface depositions and blockages. Stability against liquid water must also be taken into account. Generic silica does not behave adequately against liquid water presence (fractures may be created) ; therefore improved silica grades are preferable. Short lifetime is also a drawback as silica gels are getting saturated in relatively short times [1,4,6,10,11,12,13].

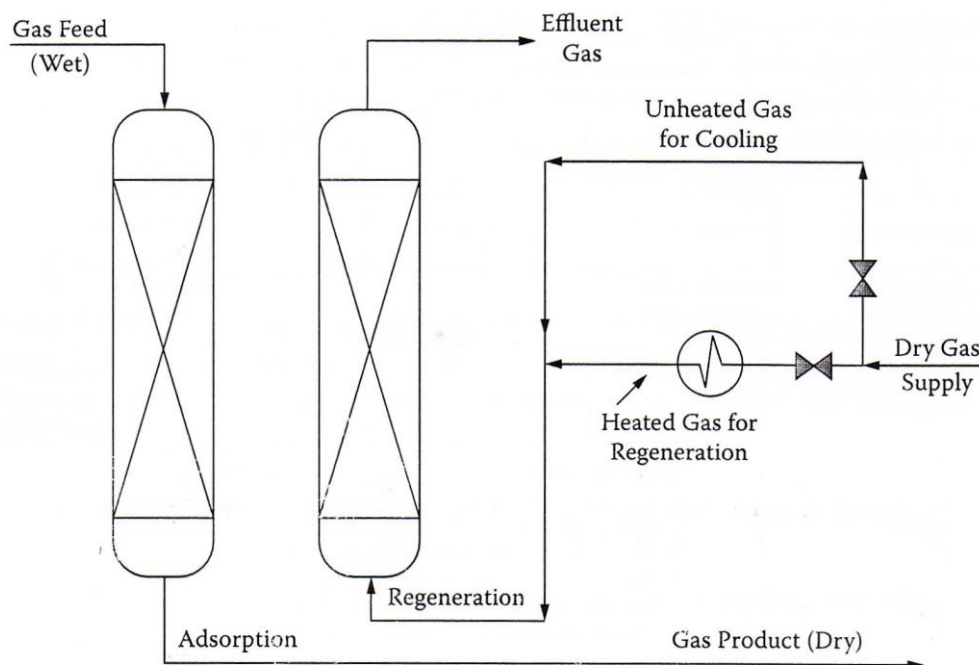


Figure 2.9 Gas dehydration unit with solid adsorption agents [11]

Molecular sieves, also known as zeolites, are aluminosilicates that selectively remove impurities due to their crystalline structure where sites of residual valency existing on their surface adsorb water molecules. Molecular sieves are effective also against other impurities such as acid gases and they are preferably used when extensive dew point depression is required (down to $-150\text{ }^{\circ}\text{F}$). Molecular sieves are considered to be fit for

purpose due to the fact that their lattice pore openings are determined by the cation of the metal existing in the aluminosilicate[1,4,6,10,11,12,13].

Table 2.7 Solid desiccant properties [1,4,7]

Properties	Silica	Activated Alumina	Molecular Sieves
Pore diameter (A°)	10-90	15	3-10
Bulk density (lbf/ft ³)	45	46	43-47
Heat capacity (Btu/lb °F)	0.22	0.24	0.23
Dew Point (o F)	-60 to -90	-60 to -90	-150 to -300
Regeneration temperature (o F)	300-500	300-500	425-550

3. Condensation: Cooling the gas below dew point by using propane refrigeration or ammonia liquid for chilling. A glycol solution is also injected into the gas to prevent solid hydrate solution. The target here is to take advantage of the ability of the gas to host less water content as its temperature is being reduced under constant pressure. Low temperature are operating by enhancing Joule-Thomson effect to take place and water and Hcs are released from the gas stream through condensation. Hydrates are also formed through gas expansion. The resulted mixture enters a coil separation where glycol injection is taking place and gas dehydration occurs.

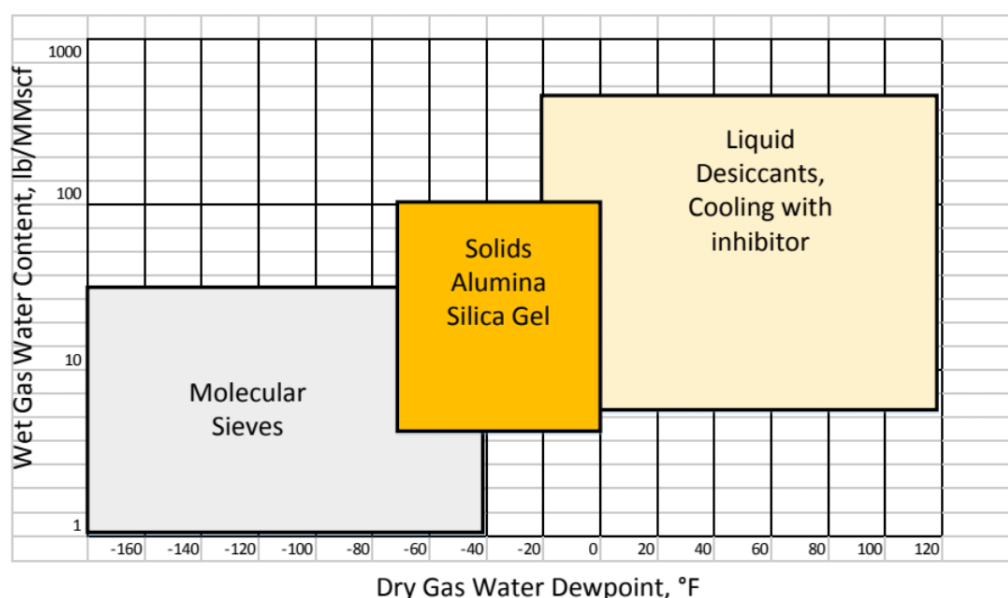


Figure 2.10 Gas dehydration process- Feed water content [1]

Table 2.8 Dehydration Processes Efficiency [1,2,3,4,6,7]

Process	Efficiency			
Absorption	Pipeline Water Content Specification	Deep Dew Point Depression	VOC removal	RSH removal
T.E.G.	✓		✓	
CaCl ₂	✓	✓		
Adsorption				
SiO ₂	✓		✓	
Activated Alumina	✓		✓	
Molecular sieves	✓	✓		✓

The selection of the right process should be based on desirable water dew point depression. Usually all the available processes can provide pipeline quality gas (less than 7-1 lb/MMscf) of water. For deep dehydration molecular sieves seem to be more effective [1,4,10].

2.4.2 Gas dehydration using T.E.G.

Gas containing water vapor enters an inlet filter separation vessel within which free water and any left liquid HCs removal is ensured. This initial separation process is desirable because protects the absorber-contactor from blockages and fouling. Stripped gas then enters the bottom of the absorber where contact with the countercurrent T.E.G. solution takes place. Lean glycol removes vapor water and any volatile organic compounds (VOC) leaving the dry gas stream to exit the absorber from its top section. A demister placed just before the exit of the contactor ensures no glycol losses. Rich T.E.G. is passing through a heat exchanger and flashed into a tank where flashed gas can be regained and introduced as fuel gas for the plant needs. Regeneration of rich T.E.G. takes place inside the reclaimer-regenerator where the reboiler vaporizes only water out of the solution based on the boiling point differences between them. Lean T.E.G. is then cooled down and returns to the top section of the absorber for a new dehydration cycle[1,4,6,10].

2.4.3. Glycol Dehydration Unit Considerations

Key design considerations that must be taken into account when a T.E.G. dehydration unit design are the following:

1) Water removal rate: The amount of vapor water that a dehydration unit can remove; can be calculated by the following equation:

$$Wr = \frac{Qg \times (Wi - Wo)}{24} \quad (2.12)$$

Where

Wr is the water content removed (lb/h)

Wi is the water content of the feed gs (lb/MMscf)

Wo is the water content of the stripped gas (lb/MMscf)

Qg is the gas flow rate MMscfd

2) Glycol circulation rate: It can be defined based on the amount of the desirable removed water. Equilibrium absorption stages do affect the glycol rate. Typical value

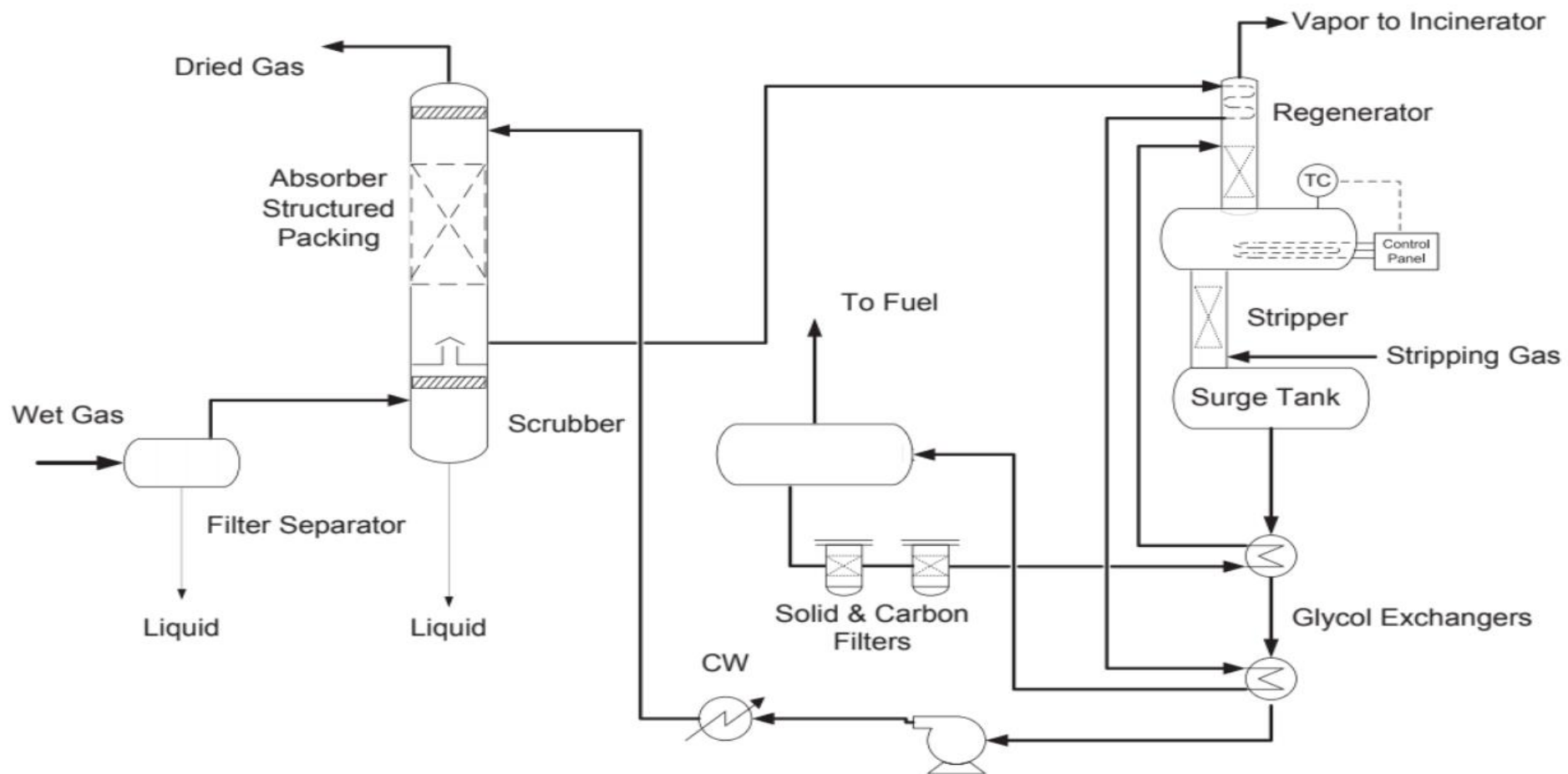


Figure 2.11 General flowchart of a glycol dehydration unit [1]

is 3 gallon of T.E.G. for each pound of water. An introduction of a higher glycol rate would not add any extra dehydration efficiency. The only consequence would be the decrease of the reboiler operating temperature due to heat losses and insufficient water removal from the glycol. Although high circulations could potentially create

operational problems reducing the water stripping efficiency, overdesign is desirable in the range between 10-30% above the minimum glycol rate $Q_{teg,min}$ (gal/h)

$$Q_{teg,min} = G \times W_r \quad (2.13)$$

Where

G is the glycol water ratio and can be expressed by the equation:

$$G = \frac{\text{gallons T.E.G.}}{\text{pounds of removed water}} \quad (2.14)$$

3) Glycol purity: Lean glycol purity control is achieved by the reboiler operating conditions. Optimum pressure and temperature ensures the purity. Of-course decomposition limit should be taken into account. A purity of 99% glycol is achievable. Stripping gas could additionally introduced to ensure glycol purity limits.

4) Feed gas temperature: Elevated inlet gas temperatures increase gas water content and at the same time reduce absorption agent's dehydration capacity. Therefore, a pre-cooling of the feed gas is mandatory especially in hot climate areas. Water can be used as a coolant.

5) Foaming: Low glycol purity and entrained solid contaminants may trigger foaming reducing removal efficiency and disturbing further downstream processes. Inlet separation device and carbon filters ensure the absence of contaminants.

6) VOC: Volatile organic compounds such as benzene, toluene etc. could create emission specification violations as the absorbed gas is released to the atmosphere after exiting the regenerator. Although T.E.G. solution can remove VOC to some extent, more absorption stages in the contactor would ensure the absence of absorbed VOC at the regenerator outlet.

7) Regeneration temperature: Elevated regeneration operating temperatures could lead to T.E.G. losses. On the other hand low temperature values could fire up water condensation and even column flooding.

8) Reboiler salt precipitation: Entrained salts from wellhead gas stream could create corrosive precipitations inside the reboiler and blockages which could affect the heat exchanger efficiency. Even a salt concentration just above 1% in the glycol could create serious problems. Decrease in salt concentration when is mandatory, can be achieved by contaminated glycol disposal which is not preferable[1,4,6,10].

2.4.4 Water Content Determination

Water content data is considered of high importance in natural gas transmission and processing in order to ensure no hydrate formation or corrosion due to condensation. It is usually expressed as pounds of water in MMSCF of gas. At the same time this data contributes to adequate designing in gas dehydration unit (the determination of water content helps to ensure accurate calculation of dehydration agent consumption). The most appropriate method for water content determination is the experimental one and it should be followed for robust and accurate process designing; although it can be described as time consuming and relatively costly procedure[14].

Despite the experimental measurements, different approaches for estimating water content have been also generated over the years. The estimating accuracy of a method is based mainly on two criteria, the divergence and absolute divergence when compared with experimental data.

The majority of them were designed mostly for sweet gas water content calculations. Generally it is believed that acid gases presence increase solubility of water content in gas streams

- 1) Nomographic methods : Charts have been plotted based on limited experimental data; providing good estimation of water saturation concentration as far as sweet natural gas is concerned. The most widely acceptable for sweet gas water content estimation is the one derived by McKetta-Wehe. It can provide good water content estimation for a sweet natural gas containing more than 0.7 of Methane (mole fraction). Campbell and Robinson charts have been generated for sour gas computations but the complexity and the need of interpolations set serious drawbacks. It takes into account only pressure and temperature of a sweet gas stream[14,15,17].
- 2) Thermodynamic models: These models are based on EOS and assume saturation conditions to exist in order to provide a good estimation. Although they can

generate firmly accurate results; they are too complicated and time consuming. Saturated thermodynamic model, Modified ideal thermodynamic model and Simplified thermodynamic Model are the most common [15,17].

- 3) Empirical-Semi empirical Correlations: Correlations that have been generated by matching experimental data to existing equations; seem to be simple, convenient and accurate. That is why they are considered to be popular to engineers for quick estimations. On the other hand, these correlations cannot perform well enough when a gas stream is characterized of increased amount of heavier hydrocarbons. Therefore, they are constrained mostly to high methane content gas streams. Further investigation is needed as far as the low temperature effect on the accuracy of the estimation.
- 4) Analytical methods: These methods are generated from the combination of empirical correlations and thermodynamic models. Analytical methods are preferable due to their computational speed and simplicity and can produce accurate estimations. However, they differ widely in range of applicable conditions and they provide good results mostly for sweet gases. Although they can be also applicable to sour gases when they are supplemented by a correction method, clarification is still required as far as the type of appropriate correction for each analytical method. Recent research has shown that Bahadori Correction and Khaled Correction Method as well as Sloan, Bukacek and Maddox Correction, combined each time with a suitable thermodynamic model, can produce the most accurate estimations of water content in a sour gas[14,15,17].

As far as high pressures and temperatures are concerned where non linearity is observed; Artificial Neural Network method developed by Shirvany, Zahedi and Bashiri, which takes as input H_2S concentration, Pressure and Temperature, has shown extremely high accuracy outperforming traditional methods [16].

2.5 Sulfur Recovery

Hydrogen sulfide contained in produced acid gas streams of the gas processing plants is a major concern due to its high toxicity and flammability[1,3,18].

Table 2.9 H₂S exposure limits for humans [18]

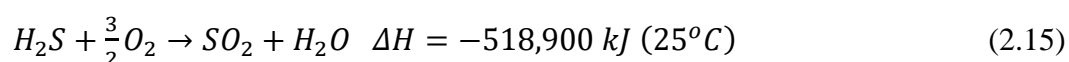
Exposure (mg/m ³)	Effect
0.011	Odour threshold
2.8	Bronchial constriction
5.0	Eye irritation
7 to 14	Increased levels of blood lactate, decreased oxygen uptake
5 to 29	Severe eye irritation
28	Fatigue, headache, poor memory and dizziness
>140	Paralysis
>560	Respiratory distress
>700	Death

Approximately 25% of gas processing plants need a H₂S treatment unit to deal with resulted acid gas streams due to environmental and health regulations. Different approaches may include one of the following:

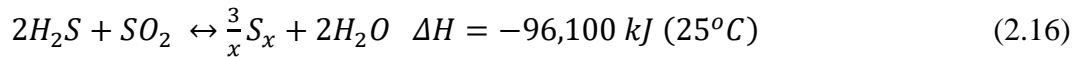
- Sulfur recovery unit, where almost the majority of existing H₂S of an acid gas stream is converted into elemental sulfur, a sticky, fine, yellow-milk inorganic and non-toxic solid which can be further sold as a feedstock to pharmaceutical and fertilizing companies.. It is the most widely acclaimed method for treating H₂S.
- Calcium sulfate (CaSO₄) unit where H₂S is converted into non-toxic CaSO₄ inorganic product which is safe for disposal. This approach lacks of practicality and it is not popular.
- Acid gas injection usages if neighboring injection well are existed. It is favorable in remote regions where enhanced oil recovery is mandatory.

As far as sulfur recovery unit is concerned; the main principle of the process is based on the modification of Claus Process which consists of two main sections[1,4]:

- 1) A high exothermic combustion reaction section where enough air is added along with acid gas stream and oxidization of about 1/3 of H₂S to SO₂ occurs.



- 2) A moderate exothermic catalytic reaction, known as Claus reaction where what is left unburned from H_2S reacts with SO_2 to create elemental sulfur.



Typical sulfur recovery units consist of one free-flame combustion and three catalytic reaction stages.

The feed, which is usually consists of H_2S , CO_2 and remaining HCs traces and it is saturated with water, enters the reaction furnace along with combustion air at approximately 8psig and 100 to 120 °F pressure and temperature respectively. Hot gases produced from the burner (1000-2500°F) are cooled down to 600-800 °F by passing through a heat boiler (Waste heat boiler) where high pressure steam is co-produced. Cooled gas stream leaving the WHB is further cooled down to 350 °F by passing through a condenser where medium pressure steam co-production occurs. Therefore; sulfur is condensed due to temperature reduction, separated from the cooled stream and drained down to a sulfur collection unit. What is left from the cooled gas stream, enters the catalytic section of the process [1,4,10].

Catalytic Claus reaction section typically consists of three catalytic reaction stages. Gas is pre-heated up to 600 °F to avoid condensing existed sulfur vapors inside catalytic reactor and ensure the destruction of COS and CS_2 . Then the heated stream enters the catalytic bed where remaining H_2S reacts with produced SO_2 to form sulfur by achieving 70% conversion. Sulfur vapors are removed by low pressure steam co-production in a sulfur condenser. Remained gas stream is re-heated back and introduced to the next catalytic bed. Final condenser operates at about 300 °F to ensure no sulfur losses. Each one of the pre-described reaction steps transforms a smaller fraction of H_2S into sulfur and contributes to the added result of H_2S reduction. Tail gas of sulfur recovery unit can be either introduced into a tail gas processing unit or a thermal oxidizer which converts all the sulfur compounds into SO_2 before vented into the environment. A tail gas unit, which is the most common practice; recycles sulfur compounds after capturing them and re-introduce them into sulfur recovery unit. This is mandatory because efficiency of a Claus Processing unit can never exceed 97%. This upper limit is imposed from the equilibrium reaction that occurs in the catalytic reaction section [1,4,10].

A further classification of Claus processing methods can be done based on the method used and extent of SO_2 production at the combustion section:

- Straight through Claus: A free flame combustion section and catalyst section exists. More than half of the contained H_2S of the feed gas is converted to sulfur inside the combustion section and therefore catalyst section deals with lower loadings, something that promotes stage efficiency. This type of sulfur units can treat acid gas stream containing from 100% to 40% sulfur as far as the burners flame temperature is highly dependent on the H_2S composition of the stream. Low H_2S concentrations affects flame temperature by making it unstable and a pre-heating stage must be introduced something which is non-profitable.
- Split Flow process: It consists of the same reaction section as Straight through Claus process. The concept here is that small amount of H_2S is burned in the combustion section at such extent capable to provide stoichiometric ratio 2:1 of the two reactants (H_2S and SO_2) for the catalyst section. This approach is used for lean acid gas streams containing relatively low H_2S content (15-50%).
- Direct oxidation: For very lean acid gas streams (H_2S content less than 15%), direct oxidation is more preferable than a Claus unit approach. Acid gas is directly fed in a catalyst section along with combustion air to produce SO_2 . H_2S and produced SO_2 then react together inside a Claus unit for additional sulfur recovery. This option is sensitive as far as catalyst poisoning is concerned and lacks in reliability.
- Acid gas enrichment: Low concentration acid gas streams are undergoing H_2S enrichment by removing CO_2 out of them. This enhances reliability of sulfur recovery unit and prevents overdesigning.
- Oxygen enrichment: Combustion air contains high amount of nitrogen (79%) which considered inert gas. This significant portion of nitrogen affects not only the temperature in the reaction combustion section but also the equipment loadings during sulfur condensation and re-heating. Oxygen enrichment in combustion air feed would improve reaction efficiency and once again prevent overdesigning[1,4,6,10].

Strict environmental regulations impose the removal of any left traces of sulfur compounds from sulfur recovery outlet stream. This can be achieved by a tail gas process. This type of processes can be classified into three main categories:

1. Reduction processes: Include H_2S conversion into sulfur by using a hydrogenation stage (pre-heating, conversion of SO_2 into H_2S by hydrogenation, removal of water) and a scrubbing/recycling stage of H_2S into the main Claus Process (scrubbing can be achieved by using an alkanolamine process e.x MDEA), leaving the outlet stream clean from sulfur species. Shell Claus Off-gas Treating process (SCOT) and Parsons/ Uocal's Beavon Sulfur Removal process are the most common.
2. SO_2 scrubbing processes: During this type of processes; catalytic oxidation of sulfur compounds and transformation into SO_2 is taking place. The produced SO_2 gases can be either vented into the environment if their quantities satisfy the law; or removed by a selective process and then re-introduced into the main Claus Process.
3. Catalytic oxidation Process: The main concept here is to introduce additional stages into catalyst section of conventional Claus process and therefore, increase sulfur recovery up to 99%. Superclaus process and sub-dew point configurations are the most common[1,10].

2.6 Nitrogen Removal

Nitrogen as an inert gas could create several problems in pipeline transmission (increases compression equipment loading, reduces heating value of final product) and further gas processing stages such as LNG (reduces further the liquification temperature with subsequent increase in operational cost) and NGL units (affects recovery efficiency). Therefore; N_2 is not desirable

There are two main methods of nitrogen removal:

1. Cryogenic Process: Takes advantage of the boiling point difference between CH_4 and N_2 . Joule-Thompson cooling effect and simple distillation are used to produce two distinct streams, one containing methane and approximately 1 mol % of nitrogen and vice versa. Inlet gas introduced in a gas-gas exchanger and passes through a valve. Temperature is reduced via a self-refrigeration procedure which contains pressure reduction due to the choke valve. This is the

most widely used method for nitrogen removal and it can be used for any gas feed up to 900 MMSCFD.

2. Non-cryogenic Process: This process involves either the usage of molecular zeolite based sieves or membrane separation technology but it is not preferable as it cannot handle large flows of gas. Furthermore, HCs leakage in nitrogen reject cannot be minimized. Therefore; it is not preferable[1,10].

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3. PROCESS SIMULATION AND DESIGN

3.1 Aspen HYSYS simulation software

Aspen HYSYS is a simulation software; commercial product of Aspen Tech that can facilitate computer-aided modeling and analyses of large-scale chemical processes in a virtual environment. It is considered to have a user-friendly interface based on a step by step simulation strategy that can support process flowchart designing by performing mass and energy balance calculations.

Aspen HYSYS allows the building of a processing plant, the sizing and costing of equipment of a scheduled unit under construction and a potential plant optimization without the dangers, costs and risks of a pilot unit. This can be achieved by substituting a physical system into a mathematical model fully illustrated in the simulation environment by a virtual one. A series of simulation equipment devices and units are available in Aspen HYSYS; each one depicted properly with its representative icon containing the mathematical model defined by the basic operational principles. The transition of a physical process to Aspen Hysys virtual environment can be done through some simple steps. Component selection is the first step, where the user has either the option to choose the suitable components from a HYSYS library or to enter his own specific components by defining molecular weight. Physical properties for the listed components are “known” to HYSYS. Then the proper thermodynamic model for each process has to be defined from an available list. This model is known as the “ Fluid Package”. Material streams that connect devices are introduced. Compositions are required; as well as flow rates and conditions. After defining all the above the simulation environment is opened. By implementing each stream connection with a certain device, a flowchart is successively created; which can be described as a combination of input variables, equipment parameters and equations. Running the designed model or solving the equations results to the output stream data of the simulated process after all the parameters have successfully converged.

Simulation attempts of processing plants can be classified as steady state or dynamic simulations based on whether or not the plant is operating with no variations with respect to time, in a perfectly stable mode. Flexibility and speed of solution is achieved through steady state simulation and it can produce firmly accurate results when main equipment design is considered. On the other hand; dynamic simulation allows the investigation of more realistic approaches but requires specific input data and focuses on transient phenomena that take place in the process; that is why it is used when safety

analysis and advanced process control designing consist the simulation target. Steady state simulation is based on modular sequential approach; which consists of step by step computational solution where every piece of equipment is considered as block which is necessary to be solved before the simulation continues to the next one. This approach can be considered as time consuming but provides robust results [1].

3.2 Selection of implemented simulation data. Iran Pars-Qatar Northfield accumulation

An Iranian North Pars gas composition was selected for simulating the design of an acid gas removal unit using D.E.A. and a dehydration unit using T.E.G. respectively. The selection was made due to availability of composition in bibliography and because it is considered a relatively typical lean and sour gas. North Pars was discovered back in 1967 and it is considered as one of Iran's largest independent gas fields. It has been found [2].



Figure 3.1 Known Gas(red colored) and Oil(green colored) Accumulations in Persian Gulf[3]

120 km southeast of Bushehr province and 10 to 15 km offshore in water depths ranging from 2 to 30 meters; with an estimate of about 1.7 Trillion cubic meters of reserves of which recoverable ones are considered almost 1.3 Trilion cubic meters. Although for many years the exploitation of North Pars gas was pending due to South Pars production competition between Iran and Qatar, a current development plant has begun; consisted of four phases. The recovered sour gas is oriented for LNG production [3].

A typical gas composition expressed in mole percentages and mole fractions at the standards of laboratory representation of Iran Pars gas can be seen in table 3.1 taken by bibliography[4]. It is considered to be a lean gas due to its low concentration in heavier HCs. The given composition has been taken from free of water samples and do not include any water vapor or free water that is co-produced with the related gas. It consists mostly of methane and it is considered of high calorific value due to the low values of non-hydrocarbon content [2,3,4]. Due to the lack of gas well data, it was assumed a pressure of 1000 psia and a temperature of 95°F for the inlet stream of acid gas removal unit.

Table 3.1 Typical Composition of Iran Pars Natural Gas[4]

Component	Mole Percentage	Mole fraction
N ₂	1.70	0.017
CO ₂	3.28	0.0328
H ₂ S	0.66	0.0066
Methane	89.26	0.896
Ethane	2.26	0.0226
Propane	0.51	0.0051
i-Butane	0.12	0.0012
n-Butane	0.13	0.0013
i-Pentane	0.06	0.0006
n-Pentane	0.04	0.0004
Hexanes	0.06	0.0006
Heptane-plus	0.24	0.0024

3.3 Simulation of Amine Sweetening Unit Using D.E.A.

3.3.1 Description of process flow diagram- Material and Energy Balances

A flow chart of the simulated process has been designed in Visio in order to describe more thoroughly the selected amine gas sweetening route. The process has been splitted into section. Each section is defined from a certain device and its inlet and outlet stream. Surge tank which is designed to facilitate the whole DEA inventory; can be considered as the start-up point for the first day of operation. An initial lean amine circulation before the introduction of the raw gas is needed. After the entering of raw gas the system operates according to the following description.

Section 100: A normalized composition of Iran Pars gas with an assumed water content of 0.50 mole percentage was implemented. Inlet gas enters the contactor from its bottom at 90 °F and 1000 psia was assumed with a flow rate of 32 MMSCFD. Lean amine (DEA) aqueous solution is introduced into the contactor from the top inlet section with a typical concentration of 27 wt%. Countercurrent flow, high pressure and relatively low temperature enhance the separation of acid gases from the main gas stream. The contactor is designed with the suggested number of trays (20). 30,54 MMSCFD of Sweet gas containing no H₂S and with 0,9448 mole fraction of methane; exits the top of the absorber and can be processed further to the dehydration unit. Rich amine containing the removed acid gases moves further to a shell-tube exchanger. Rich amine is introduced always on the tube side of the exchanger providing this way optimum conditions for heat exchange. Lean amine flows in the shell side providing the needed amount of heat in order to achieve temperature increase in rich amine stream.

Section 200: Pre-heated rich amine can now enter the amine regenerator, a distillation column, where separation of acid gases takes place from the amine solution based on boiling point differences. A series of perforated trays or plates provide the necessary vapor-liquid contact establishing equilibrium stages. Distillation column operates optimum in low pressures and high temperatures-the opposite of the contactor-absorber. The required regenerator temperature is provided by low a pressure steam stream at the reboiler. A number of 20 trays is selected. Feed operating temperature (206,6 °F) is inside the suggested operating range (200-215 °F). Operational pressure is about 32 psia. Lean amine exits the regenerator bottom while removed acid gases come from the top of the column. Lean amine passes through the lean-rich amine heat exchanger before proceeding to the next section.

Section 300: Lean amine at high temperature and low pressure; moves to the surge tank after exiting the bottom of the column; where constant circulation is applied. The required amine concentration is controlled by introducing water or DEA if the concentration falls out of the specified range. A purge stream ensures

the escape of traces of inert gases like CO₂ and N₂; preventing their build up inside the system.

A pump pressurizes the lean amine solution to inlet pressure for the absorber. A cooler ensures temperature reduction and the lean amine is re-circulated into the system by entering the absorber from its top section for a new cycle.

Table 3.2 Equipment Name and Symbolism (Acid Gas Removal Unit)

Equipment	Equipment Symbolism
Absorber	T-100
Rich-Lean DEA Heat Exchanger	HEAT-EX 100
DEA Regeneration Column	REGEN-201
Surge Tank	S TANK-300
Pressure Pump	P-300
Cooler	Cooler-300

3.3.2 Process Control

Process control is a major sector in process industry. Not only it reduces the variability of the final product, thus ensuring high quality; but it also guaranties safety through the establishment of precise operating conditions. The main principles of process control are the measurement of a variable (measured variable) by a measurement primary element (for example a manometer or thermometer), the comparison with a preset value and finally the adjustment of the variable value based on the calculated difference through the action of a control element on a manipulated variable which affects the measured variable value. Most of the time the control element in process industry is a control valve [5].

A flow controller has been introduced in the top of the absorber in order to adjust the incoming D.E.A. and therefore the liquid level inside the absorber. Pressure and level indicators are also placed.

Flow controller has been introduced also in the heat exchanger section at the rich amine hot stream. Rich amine flow defines the heat exchange that takes place in the device. Therefore; when temperature inside the exchanger is fluctuating over or above the desirable value, a flow controller through a valve reduces or increases the amount of the hot stream that enters the heat exchanger. Protection from potential overflowing is also ensured.

Another point that temperature value is considered very important variable for the optimal operation of the unit, is the distillation column. A temperature controller has

been introduced at the top section and bottom section respectively; controlling a valve which determines the amount of heat entering the absorber through the low-pressure steam current and therefore the temperature at the top section of the column. This ensures good amine condensation and column recirculation and optimal operational temperature. Level controller is introduced at the bottom section in order to control the liquid levels and protect the column from flooding. A flow controller at the inlet section is also needed to adjust the incoming amount of D.E.A.

Temperature indicator can be placed at several streams and devices of the process. A temperature indicator helps to observe the conditions of the stored amine in the amine storage tank. A level controller controls through a valve the level of amine inside the storage tank.

At the surge tank a level controller ensures the flow of the amine solution to the rest of the plant. When a shut-down is needed the valve closes and can facilitate the whole amine inventory. A concentration controller also helps to adjust amine flow from the storage tank in order to reach required solution concentrations

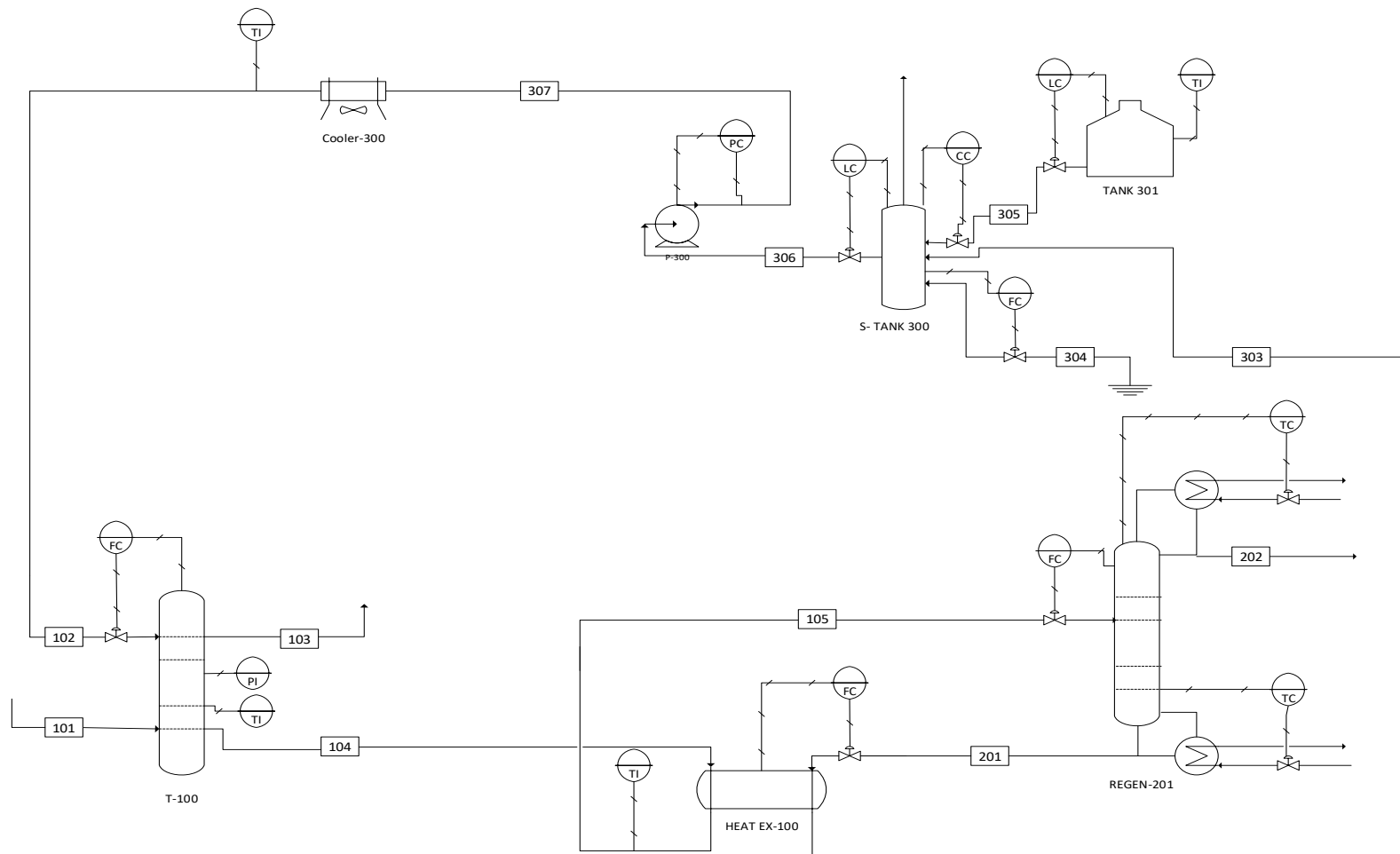


Figure 3.2 Flowchart of Amine Sweetening Unit

M&E	Streams					
Mole Fraction	101	102	103	104	201	202
N2	0,0171	0	0,0179	0	0	0
CO2	0,0331	0,0005	0,0001	0,0229	0,0229	0,0005
H2S	0,0067	0,0001	0	0,0046	0,0046	0,0001
Methane	0,9036	0	0,9448	0,0012	0,0012	0
Ethane	0,0228	0	0,0238	0	0	0
Propane	0,0051	0	0,0054	0	0	0
i-Butane	0,0012	0	0,0013	0	0	0
n-Butane	0,0013	0	0,0014	0	0	0
i-Pentane	0,0006	0	0,0006	0	0	0
n-Pentane	0,0004	0	0,0004	0	0	0
Hexanes	0,0006	0	0,0006	0	0	0
Heptane-Plus	0,0024	0	0,0025	0	0	0
H2O	0,0050	0,9376	0,0011	0,9113	0,9113	0,9347
DEAmine	0	0,0619	0	0,0599	0,0599	0,0648
Mole Flow (Kgmole/hr)	1594	2277	1521	2350	2350	2175
Mass Flow	2,897e+004	5,33e+004	2,613e+004	5,617e+004	5,617e+004	5,149e+005
Volume Flow (S.C.)	33 MMSCFD	53 m3/hr	30,54 MMSCFD	51,59m3/hr	55,73m3/hr	49,75m3/hr
Temperature (F)	90	95	95	131,3	206,6	216,7
Pressure (psia)	1000	1000	970	985	950	34
Enthalpy Flow (kJ/hr)	-7,013e+008	-6,779e+008	-1,443+008	-7,013e+0008	-6,929e+008	-6,307e+008
Vapor Phase Fraction	1	0	1	0	0,0001	0
Liquid Phase Fraction	0	1	0	1	0,9999	1

M&E	STREAMS					
Mole Fraction	203	303	304	305	306	307
N2	0,0001	0	0	0	0	0
CO2	0,3014	0,0005	0	0	0,0005	0,0005
H2S	0,0610	0,0001	0	0	0,0001	0,0001
Methane	0,0167	0	0	0	0	0
Ethane	0,0004	0	0	0	0	0
Propane	0,0001	0	0	0	0	0
i-Butane	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0
i-Pentane	0	0	0	0	0	0
n-Pentane	0	0	0	0	0	0
Hexanes	0	0	0	0	0	0
Heptane-Plus	0	0	0	0	0	0
H2O	0,6202	0,9347	1	0	0,9376	0,9376
DEAmine	0	0,0648	0	1	0,0619	0,0619
Mole Flow (Kgmole/hr)	174,3	2175	101,8	0	2277	2277
Mass Flow	4674	5,149e+004	1834	0	5,33e+004	5,33e+004
Volume Flow (S.C.)	3,5 MMSCFD	49,75 m3/hr	1,835 m3/hr	0	53m3/hr	53m3/hr
Temperature (F)	223,9	189,8	189,8	189,8	189,8	191,9
Pressure (psia)	30	24	24	24	24	1000
Enthalpy Flow (kJ/hr)	-4,678E+007	-6,390e+008	-2,861e+007	0	-6,676e+008	-6,671e+008
Vapor Phase	1	0	0	0	0	0
Liquid Phase	0	1	1	1	1	1

3.3.3 Simulation of a Gas Dehydration Unit Using T.E.G.

3.3.4 Description of process flow diagram- Material and Energy Balances

Section 400: Natural gas stream from the acid gas removal unit, enters a vertical inlet separator where amine solution introduced during gas sweetening is removed.

Section 500: Separator gas then enters the absorber from the bottom inlet. Liquid T.E.G is introduced from the top section. A number of bubble cap trays ensure good liquid distribution. T.E.G. absorbs any water content from the gas. Then rich T.E.G. stream leaves the absorber from the bottom section. Dry gas exits the absorber from the top outlet at ambient temperature (96,84 °F). Rich T.E.G. passes through a valve for pressure reduction. Then the low-pressure stream of rich T.E.G. is introduced in a heat exchanger before entering the regenerator.

Section 600: The rich T.E.G. stream enters the regenerator within which it is stripped of its water content. A wet gas stream leaves the regenerator from its top section while lean T.E.G. is exiting through the bottom section at high temperature. Lean T.E.G. hot stream can be used for temperature increase of the rich T.E.G. feed stream before the latter enters the regenerator.

Section 700: Regenerated lean T.E.G. passes through a mixture at which mixing of the lean stream with additional T.E.G. can be achieved through a connected high purity T.E.G. tank. A pump is used to increase lean T.E.G. pressure. Lean T.E.G. temperature is reduced in a heat exchanger where the produced dried gas is used as a coolant. Dried gas due to its high flow rate and the T.E.G. stream low flow rate is not affected much as far as its temperature is concerned. Dried gas stream is then headed further as sales gas. 30 MMSCFD of high purity gas have been successfully produced.

Table 3.3 Equipment Name and Symbolism (Dehydration Unit)

Equipment	Equipment Symbolism
Inlet Separator	SEP-400
Absorber	T-500
Rich-Lean Heat Exchanger	HEAT EX-500
Regenerator	DI-600
Mixer	M-700
Pressure Pump	P-700
Dried Gas-T.E.G. Heat Exchanger	HEAT EX-700

3.3.5 Process Control

A level controller controls the liquid accumulation inside the separator in order to avoid flooding of the vessel. A flow controller at the top of the absorber defines the T.E.G. entrance flow. Pressure and temperature indicators can also be adjusted at the absorber. A flow controller manipulating hot T.E.G. stream at the rich-lean heat exchanger; ensures no overflowing of the equipment and also helps to control the facilitated heat exchange.

A flow controller at the feed of the regenerator defines the introduced amount of T.E.G.

A concentration controller at the mixer; adjusts the amount of required T.E.G. that flows from the T.E.G. storage tank. A flow controller at the dry gas- lean T.E.G. heat exchanger; adjusts the heat exchange that takes place there.

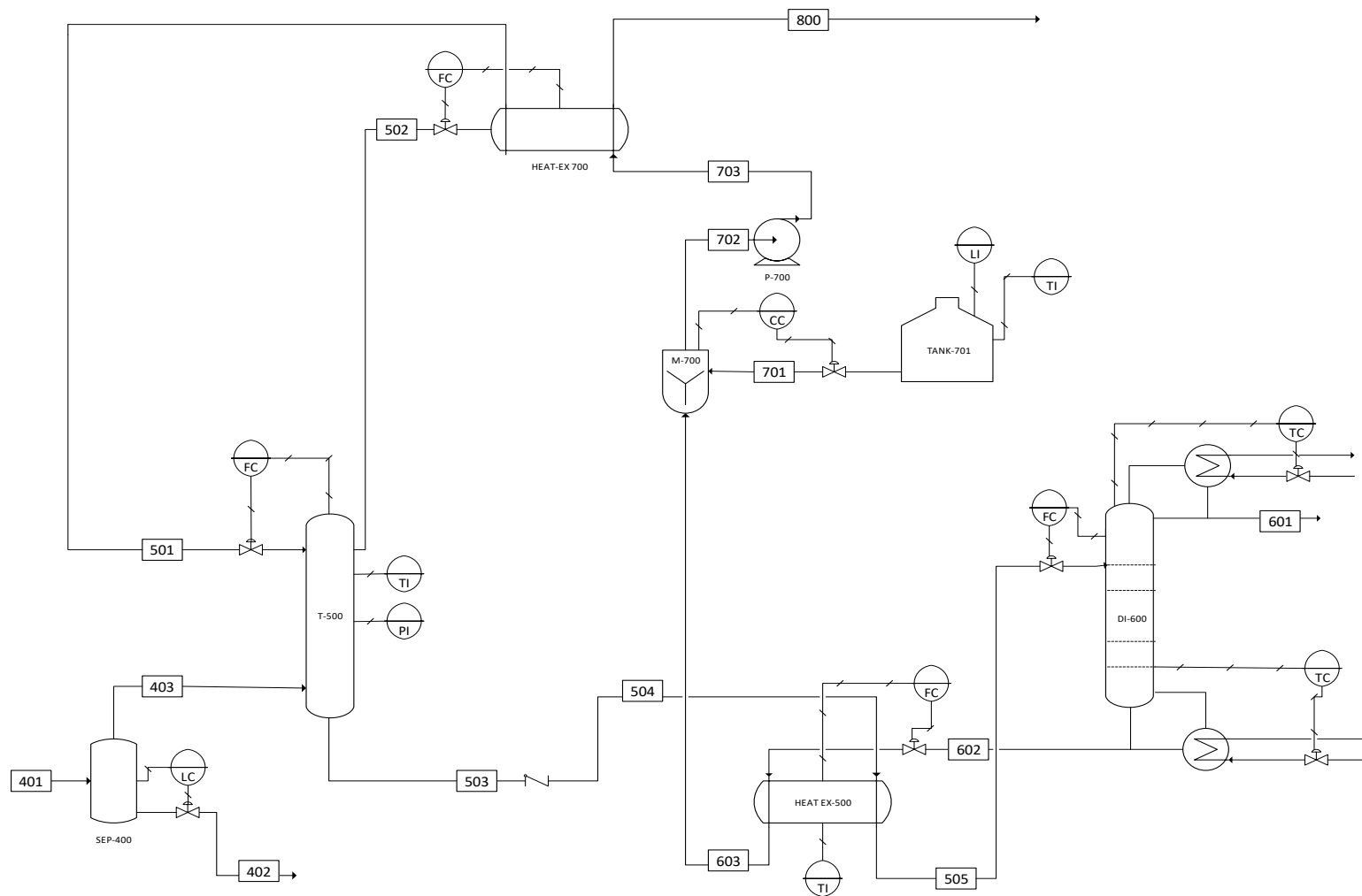


Figure 3.3 Flow Chart of Dehydration Unit

M&E	Streams						
Mole Fraction	401	402	403	501	502	503	504
N2	0,0179	0	0,0179	0	0,0115	0,0001	0,0001
CO2	0,0001	-	0,0001	0	0,0001	0	0
H2S	0	0	0	0	0	0	0
Methane	0,9448	0,0011	0,9472	0	0,9334	0,0061	0,0061
Ethane	0,0238	0	0,0239	0	0,0371	0,0003	0,0003
Propane	0,0054	0	0,0054	0	0,0087	0,0003	0,0003
i-Butane	0,0013	0	0,0013	0	0,0025	0,0007	0,0007
n-Butane	0,0014	0	0,0014	0	0,0026	0,0001	0,0001
i-Pentane	0,0006	0	0,0006	0	0,0013	0,0006	0,0006
n-Pentane	0,0004	0	0,0004	0	0,0008	0,0001	0,0001
Hexanes	0,0006	0	0,0006	0	0,0014	0,0002	0,0002
Heptane-Plus	0,0025	0	0,0003	0	0,0007	0,0003	0,0003
H2O	0,0011	0,9988	0,0009	0,0752	0	0,3127	0,3127
T.E.G.	0	0	0	0,9248	0	0,6786	0,6786
Mole Flow (Kgmole/hr)	1521	0,3293	1521	3,651	1519	4,975	4,975
Mass Flow (Kg/hr)	2,613e+004	5,931	2,582e+004	512	2,580+004	536	536
Volume Flow (S.C.)	30,54 MMSCFD	5,854-e003	30,52 MMSCFD	0,4547m3/hr	30,51 MMSCFD	0,4747m3/hr	0,4803
Temperature (F)	95	95	95	120	93,64	92,73	95,74
Pressure (psia)	970	970	970	900	900	900	26
Enthalpy Flow (kJ/hr)	-1,443+008	-9,375E+004	-1,145+008	-2,808e+0006	-1,141e+008	-3,203e+006	-3,203e+006
Vapor Phase	0,9998	0	1	0	1	0	0,0068
Liquid Phase	0,0002	1	0	1	0	1	0,9932

M&E	Streams							
Mole Fraction	505	601	602	603	701	702	703	800
N2	0,0001	0,0002	0	0	0	0	0	0,0180
CO2	0	0,0001	0	0	0	0	0	0,0001
H2S	0	0	0	0	0	0	0	0
Methane	0,0061	0,90230	0	0	0	0	0	0,948
Ethane	0,0003	0,0011	0	0	0	0	0	0,0239
Propane	0,0003	0,0010	0	0	0	0	0	0,0054
i-Butane	0,0007	0,0027	0	0	0	0	0	0,0013
n-Butane	0,0001	0,0004	0	0	0	0	0	0,0014
i-Pentane	0,0006	0,0022	0	0	0	0	0	0,0006
n-Pentane	0,0001	0,0003	0	0	0	0	0	0
Hexanes	0,0002	0,0009	0	0	0	0	0	0,0006
Heptane-Plus	0,0003	0,0010	0	0	0	0	0	0,0003
H2O	0,3127	0,9672	0,0752	0,0752	0,01	0,0752	0,0752	0
T.E.G.	0,6786	0	0,9248	0,9248	0,99	0,9248	0,9248	0
Mole Flow (Kgmole/hr)	4,975	1,324	3,650	3,650	4,578e-004	3,651	3,651	1519
Mass Flow (Kg/hr)	536,3	24,38	511,9	511,9	6,814e-004	512	512	2,580e+004
Volume Flow (S.C.)	0,4777	2,6542e-002 MMSCFD	0,4547m3/hr	0,454m3/hr	6,047e-005m3/hr	0,454m3/hr	0,454m3/hr	30,45 MMSCFD
Temperature (F)	220	215	399	93,64	60	270,9	272,7	96,84
Pressure (psia)	16	14,7	15	14,9	14,9	14,9	910	895
Enthalpy Flow (kJ/hr)	-3,090e+006	-3,104e+005	-2,569e+0006	-2,683e+006	-375	-2,683e+006	-2,679e+006	-1,140E+008
Vapor Phase	0,0106	1	0	0	0	0	0	1
Liquid Phase	0,9894	0	1	1	1	1	1	0

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4. MAIN EQUIPMENT DESIGN AND ECONOMIC EVALUATION

The size of the equipment and the determination of characteristic dimensions for each device was generated through the combination of empirical equations and Aspen HYSYS output data.

4.1 Acid gas removal unit equipment dimensions

4.1.1 Absorber T-100

The absorption column has 1.5 m diameter and consists of 20 theoretical trays with 0.6 m spacing.

Assuming 50% efficiency, which is typical for absorption towers, the number of the real trays is given by the equation:

$$N_{real} = \frac{N_t - 1}{E_o} \quad [1]$$

Where N_t is the number of theoretical trays

E_o the overall efficiency of the column.

Therefore $N_{real} = 38$ trays

And consequently the height of the column is $H = 26.6$ m from the following equation:

$$H = 1.2(N - 1)TS \quad [2]$$

Where:

H is the column height

TS is the distance between each tray

Table 4.1 T-100 Dimensions

Dimension	Value
Diameter	1.5 m
No. of Trays	38
Height	26.6 m

4.1.2 Rich-Learn Amine Heat Exchanger HEAT EX-100

The rich-lean amine heat exchanger is a tube-shell U type heat exchanger.

The dimensions of the heat exchanger are given by the software:

Heat transfer area: $A = 60 \text{ m}^2$

Overall heat transfer co-efficient: $U = 4381 \text{ kJ/hrxm}^2\text{x}^\circ\text{C}$

Table 4.2 HEAT EX-100 Dimensions

Dimension	Value
Heat Transfer Area	60m ²
Heat Transfer Co-efficient U	4381 kJ/hrxm ² x°C
Duty	7,901 MMBtu/hr

4.1.3 Distillation Column REGEN-201

The distillation column consists of 21 trays with 0.6 m spacing. The number of the real trays is calculated from eq [1] to be 28 trays for a 70% efficiency; which is a typical value for distillation columns.

The height of the column is then computed from eq [2] to 19.4 m. Three additional meters must be considered for the base and the top of the column. The final height of the column will be then 21 m.

Table 4.3 REGEN-201 Dimensions

Dimension	Value
Diameter	1.5 m
No. of Trays	28
Height	21 m
Condenser Heat Flow	3,14 MMBtu/hr
Reboiler Heat Flow	18,09 MMBtu/hr

4.1.4. Pump P-300

All the characteristic dimensions for pump sizing are generated by the software.

Table 4.4 P-300 Dimensions

Dimension	Value
Pump Q	194,1 HP
ΔP	976 psi
Adiabatic Efficiency	69,02%
Pressure Head	687 m

4.1.5. Cooler-300

A typical cooler which uses air as a coolant has a value of A=250 m²

4.2 Dehydration Unit

4.2.1 Separator SEP-400

Characteristic vertical separator dimensions are given by HYSYS.

Table 4.5 SEP-400 Dimensions

Dimension	Value
Height	4,80 m
Diameter	1,37 m
Vessel Volume	7,09

4.2.2 Absorber T-500

A number of 14 theoretical trays were taken into account during simulation. A distance of 0.50 between each tray was generated by Hysys. The diameter was given 1.5 m. Considering a 50% efficiency which is typical for absorption towers, the number of the real trays is (eq [1]) 26. Consequently; the absorber tower height is (eq [2]) 15 m.

Table 4.6 T-500 Dimensions

Dimension	Value
Diameter	1.5 m
No. of Trays	26
Height	15 m

4.2.3 Rich-Lean T.E.G. Heat Exchanger HEAT EX-500

Table 4.7 HEAT EX-500 Dimensions

Dimension	Value
Heat Transfer Area	5 m ²
Heat Transfer Co-efficient U	228 kJ/hrm ² °C
Duty	1,13e+005 kJ/hr

4.2.4 Regenerator DI-600

Table 4.8 DI-600 Dimensions

Dimension	Value
Vessel Length	1,789 m
Vessel Diameter	1,193 m
Duty	2,41e+005 kJ/hr

4.2.5 T.E.G. pump P-700

Table 4.9 P-700 Dimensions

Dimension	Value
Pump Q	1,525 HP
ΔP	870 psi
Adiabatic Efficiency	75 %
Pressure Head	611 m

4.2.6 T.E.G.- Dry Gas Heat Exchanger HEAT EX-700

Table 4.10 HEAT EX-700 Dimensions

Dimension	Value
Heat Transfer Area	5 m ²
Heat Transfer Co-efficient U	588 kJ/hrxm ² x°C
Duty	1,13e+005 kJ/hr

4.3 Main Equipment Cost

The cost of the main equipment can be calculated through diagrams and empirical correlations which relate the basic equipment's dimensions with cost in US dollars; for a given year [1,2]. In order to calculate the cost at different years the Marshall&Swift index is introduced. M&S is a popular cost index among engineers; which expresses the average cost of the equipment based on the cost of a reference year. M&S index is published monthly in *Chemical Engineering Journal* [3]. Therefore; the expression of the diagrams-aided computed costs of the equipment in today values, can be achieved based on the equation:

$$C(2019) = \frac{M\&S(2019)}{M\&S(i)} \times C_i$$

Where C(2019) is the cost of the equipment for 2019

M&S(2019) is the index value for 2019

M&S(i) is the index value for year i

C_i is the cost of the equipment for year i

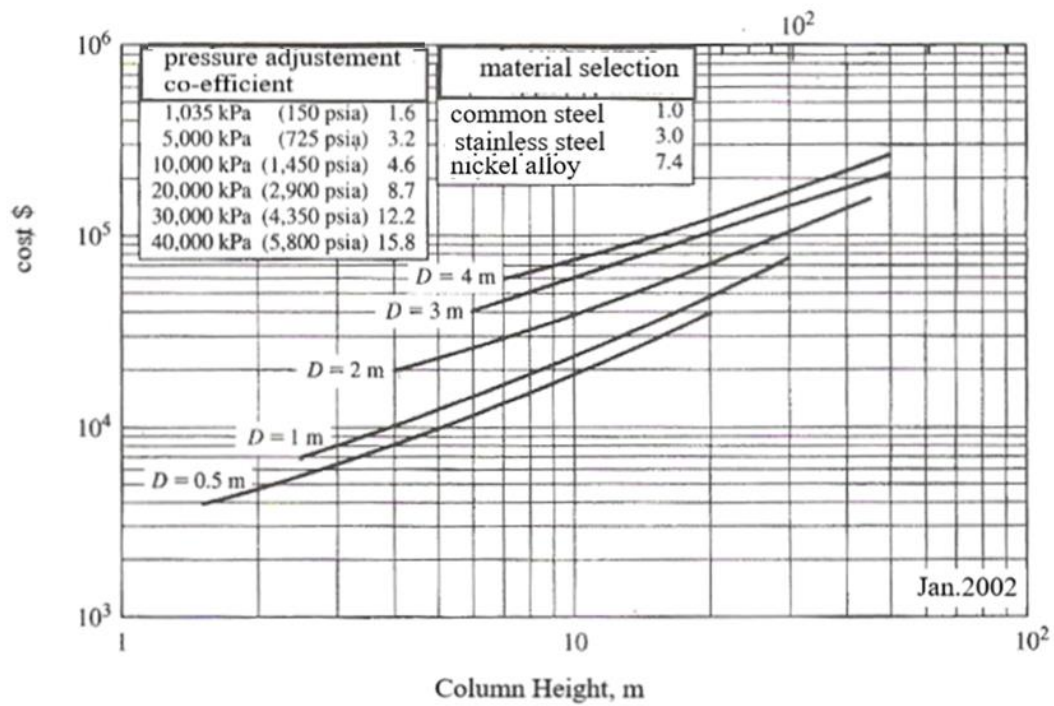


Figure 4.1 Column Cost vs Height [1]

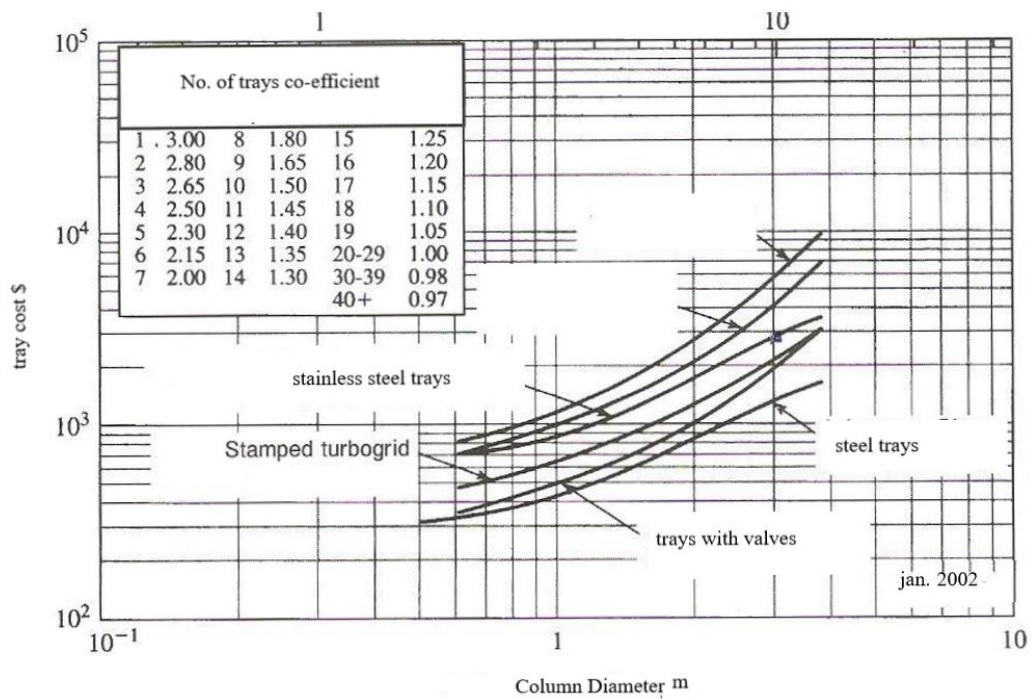


Figure 4.2 Cost of trays vs Column Diameter [1]

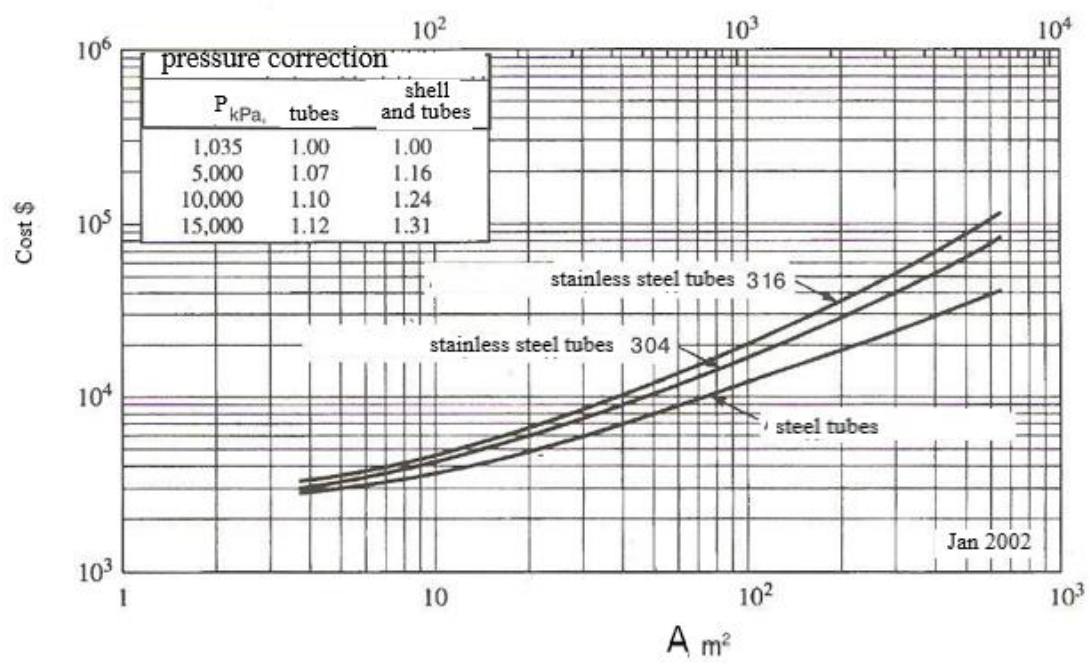


Figure 4.3 Cost of Tube-Shell Heat Exchanger vs $A(m^2)$ [1]

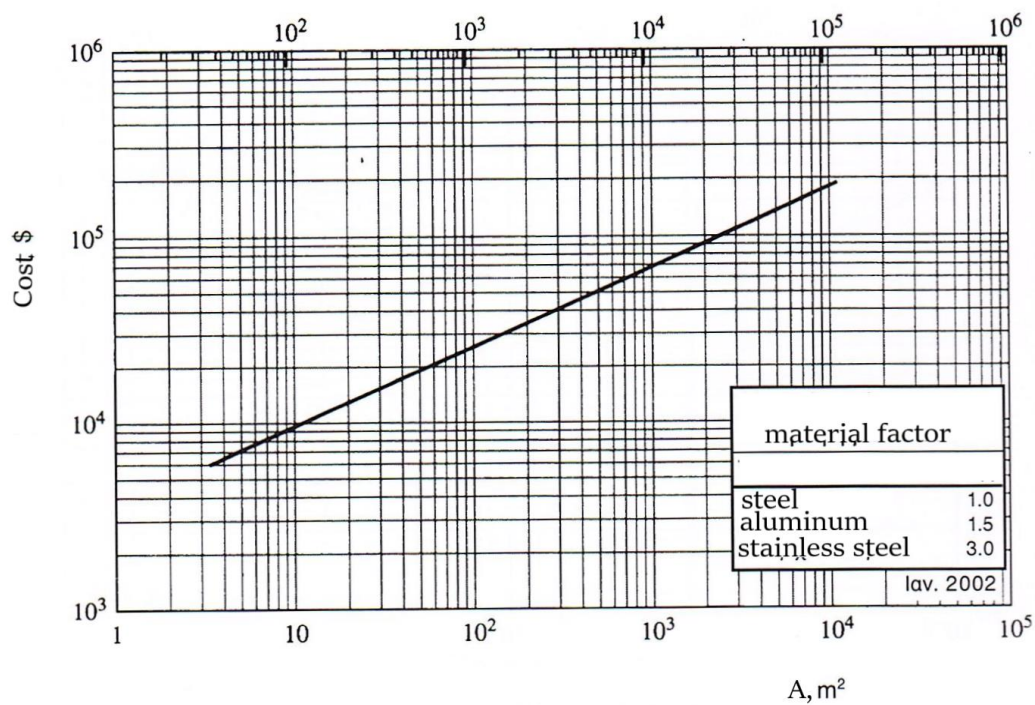


Figure 4.4 Cost of Coolers vs $A (m^2)$ [1]

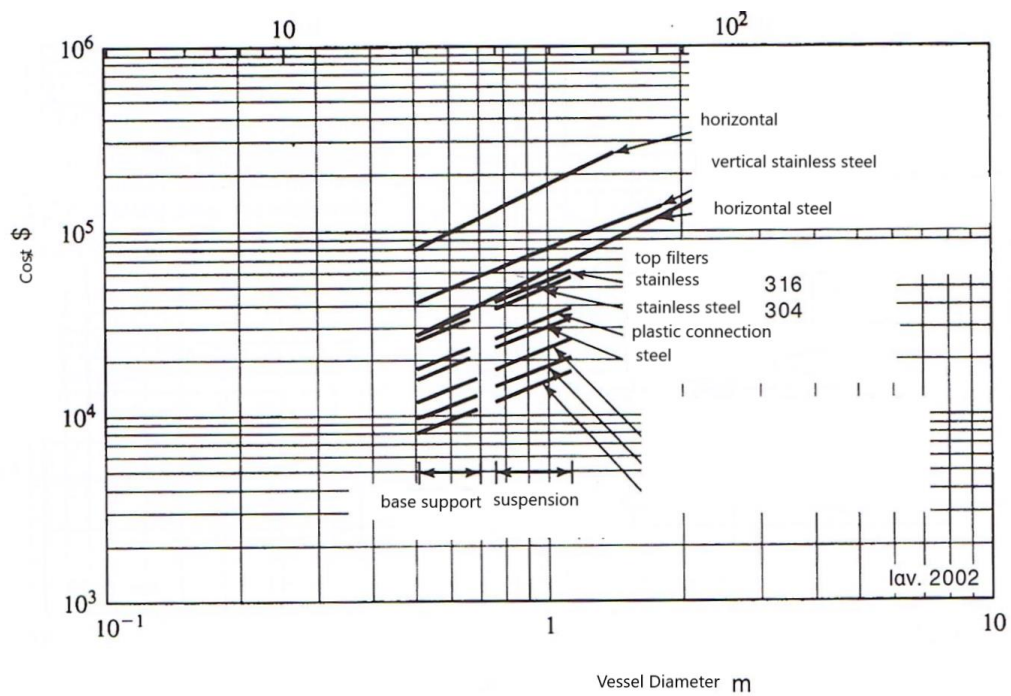


Figure 4.5 Separator cost vs Vessel Diameter [1]

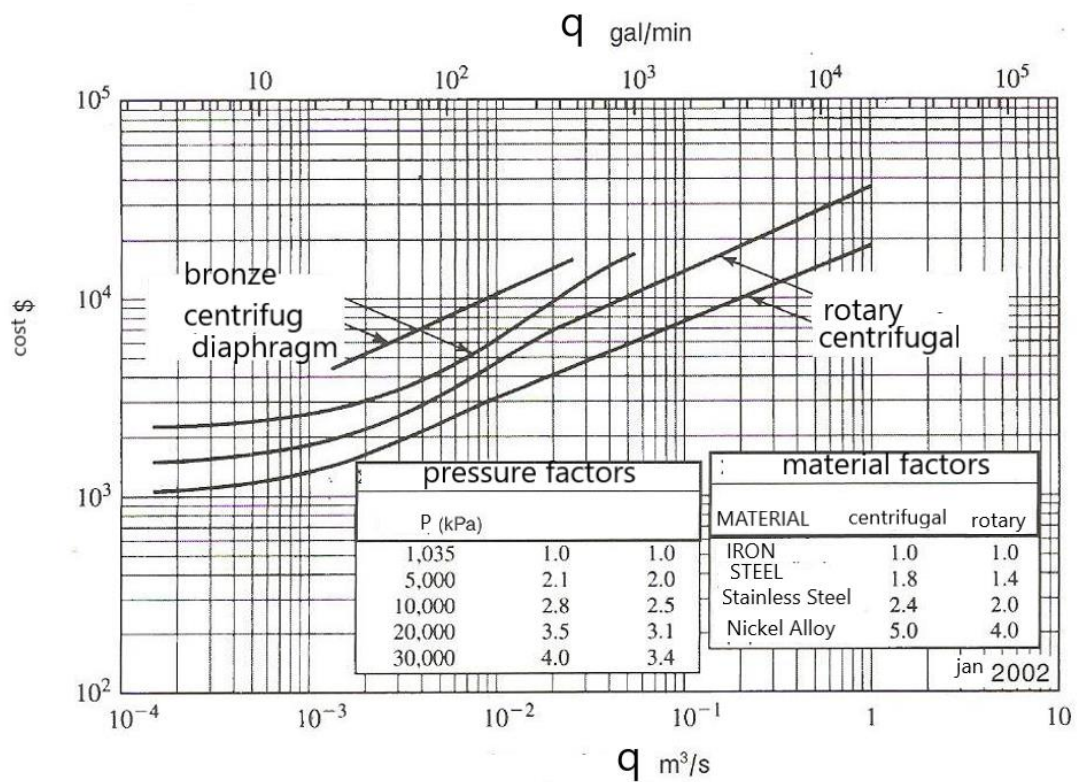


Figure 4.6 Pump costs vs inlet q [1]

The cost of the heat exchangers and reboilers can be computed from the heat exchange surface A - given in ft^2 - by the correlation

$$C = f_d \cdot f_m \cdot f_p \cdot C_b (\$) \quad [3]$$

Where:

$$C_b = \exp [8.821 - 0.30863(\ln A) + 0.0681(\ln A)^2]$$

$f_m = 0.8193 + 0.15984(\ln A)$, factor related to the selected construction material

$f_p = 0.7771 + 0.04981(\ln A)$, factor related to the pressure range.

Table 4.11 f_d value vs Heat Ex Type

Heat Exchanger Type	f_d value
Fixed Tubes	$\exp [-1.1156 + 0.0906(\ln A)]$
Kettle Reboiler	1.35
U-tube	$\exp [-0.9816 + 0.0830(\ln A)]$

The year 2019 is considered as the year where the production starts. Although M&S 2019 index value has not yet been published, it can be easily estimated by following the so far trend line of the M&S values vs the years.

Table 4.12 M&S Index Value vs Year

Year	M&S value	Year	M&S value
2000	1089	2010	1457
2001	1094	2011	1531
2002	1104	2012	1572
2003	1124	2013	1575
2004	1179	2014	1605
2005	1245	2015	1632
2006	1302	2016	1635
2007	1363	2017	1644
2008	1449	2018	1650
2009	1469	2019	To be announced

According to plotted data an estimation of M&S index value can be estimated for 2019, equal to 1670.

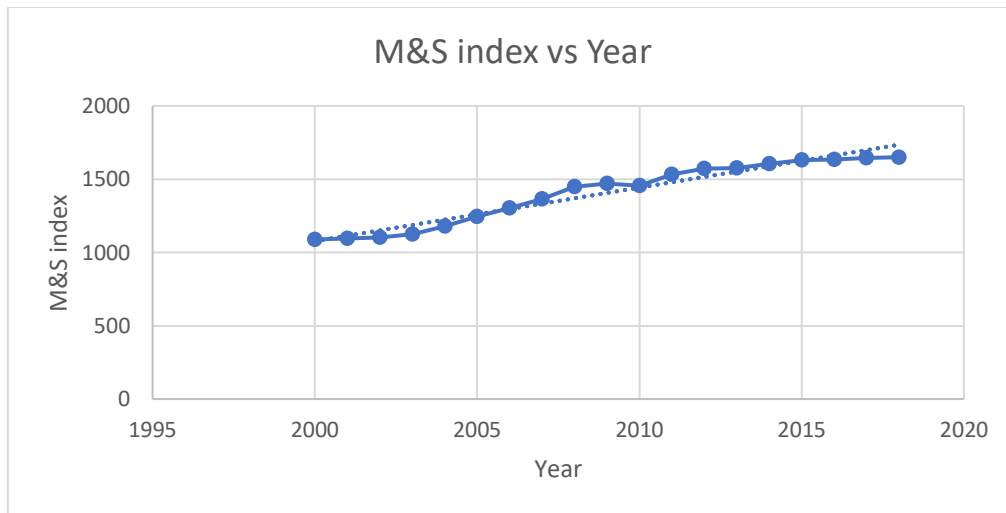


Figure 4.7 M&S index vs Year

The plotted charts of equipment costs in US \$ for year 2002 provide the initial estimations [1]. In order to adjust the costs for 2019 the following equation was used:

$$C(2019) = 1,51 \times C_i$$

All the equipment costs in 2019 US\$ are given in table 4.13

Table 4.13 Units Equipment Cost

Equipment	Cost in US\$
Acid Gas Removal Unit	
T-100	1.362.020
HEAT EX-100	18.724
REGEN-201	767.080
P-300	67.648
Cooler-300	135.900
Total Equipment Cost	2.351.372
Dehydration Unit	
SEP-400	150.000
T-500	471.000
HEAT EX-500	12.573
DI-600	120.800
P-700	30.000
HEAT EX-700	12.752
Total Equipment Cost	797.125
Total Cost	3.148.497

4.4 Capital Investment Cost

The estimation of the fixed capital investment and further economic evaluation is calculated on the basis of the cost of procurement of the plant's mechanical equipment [1]. This method is easy to implement and requires only a good initial estimate of the equipment cost. Table 4.14 presents the cost for a typical fluid processing unit both in US\$ and as a percentage of the total unit cost:

Table 4.14 Fixed Capital Investment

Expenditure	Percentage of Equipment Cost	US \$
I) Direct Costs		
1. Delivered Equipment	100	3.148.497
2. Installation	47	1.479.793
3. Process Control Equipment	36	532.725
4. Electrical networks	11	346.334
5. Buildings	18	566.729
6. Land Improvements	10	314.849
7. Auxillary services	70	2.203.947
<i>Total Direct Cost</i>	360	11.334.589
II) Indirect Costs		
1. Engineering and Supervision services	33	1.039.004
2. Construction Costs	41	1.290.883
3. Legal Services	4	125.939
4. Contractor Fee	22	692.669
5. Unforeseen Expenditure	44	1.385.338
<i>Total Indirect Cost</i>	144	4.533.835
Fixed Capital Investment I_f	504	15.868.424
Working Capital (~15% of total Investment)	89	2.802.162
Total Investment	593	18.670.587

4.5 Basic Auxiliary Supplies Costs & Economic Evaluation

After the calculation of the fixed capital investment, it is necessary to calculate all operating costs and financial rating indicators in order to be able to potentially

investigate further the feasibility of the unit. Initially, the total cost of production is calculated, then the profits from the sales and finally all the economic indices and the profitability criteria of the investment

Table 4.15 Auxiliary Supplies

Auxiliary Supply	Demands	Cost per unit	Cost (\$/yr)
Low Pressure Steam (6 bar – 160 °C)	184982GJ/yr	5\$/GJ	942910
Electricity (220 V)	1287360 kWh/yr	0,05\$/kWh	64368
Total			1.007.278

4.16 Cost of Production Calculation

Expenditure	Estimation	Cost
I) Cost of Production		
A. Direct Costs		
1.Auxiliary Supplies		1.007.278
2.Maintenance	6% of If	952.105
3.Materials	15% of A1	151.091
4.Laboratory costs	15% of A1	151.091
5.Patent rights	0	0
B. Fixed Expenditures		
1.Depreciation	10% If	1.586.842
2.Local taxes	2% If	317.368
3.Insurance	1% If	158.684
C. Additional Expenditure	60% of A2+A3+A5	661.917
II) General Expenditure		
A. Distribution and sales expenses	6% of Total Investment	1.120.235
B. Research and development costs	5% of Total Investment	933.529
C. Interest financing	5% of Total Investment	933.529
III) Total Operational Costs	I+II	7.973.669
Unforeseen	1% of Total Operational	79.736
Total Expenditures		8.053.405

It can be clearly seen after cost calculations that Acid gas removal unit consists of high cost equipment something which is expected due to the process difficulty. Absorption

column and regenerator seem to be significantly costly in both units. A total investment almost 6 time higher than total equipment cost is required. Besides direct costs; indirect costs like legal services etc are considered high enough. Auxiliary supplies do not seem to occupy significant values. On the other hand; operational cost seems to be high enough; due to considerable cost of the general expenditures.

Chapter 4 References

1. Petrs M., Timmerhaus K., West R., Plant Design and Economics, 5th Ed., McGraw & Hill (2003)

5.CONCLUSIONS

The target of the present thesis was to present the basic gas processing technologies of a gas purification plant where the main goal was to achieve production of 30MMSCFD of methane rich gas with high purity. Acid gas removal unit using D.E.A. and dehydration unit using T.E.G. were designed in HYSYS simulation environment for an Iran North Pars gas composition and detailed unit flowcharts were generated describing the mass and energy balance for each stream of the process. It was assumed that sweet gas output of the acid gas removal unit; headed at the inlet section of the dehydration unit. The results showed that a relatively simple process design approach; for both units can achieve excellent contaminants removal and yield a high purity product. A methane rich (0,94 mol fraction) natural gas is produced free of H_2S and water with a HHV=53,1 MJ/kg. As far as process control is concerned flow and pressure controllers are crucial as they ensure good operational conditions and safety. The most expensive part for both units seems to be the absorption column; especially at the acid gas unit where extra attention must be given in order to avoid corrosion. Further economic evaluation based on unit optimization is suggested. Energy recovery options must be taken into account as far as pumps are concerned. To sum up this basic design seems to provide a good approach as far as small gas processing units are concerned for sour and lean feeds.

6.APPENDIX

Source Databank: HYSYS

Component	Type	Group
Nitrogen	Pure Component	
CO2	Pure Component	
H2S	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
n-Hexane	Pure Component	
n-Heptane	Pure Component	
H2O	Pure Component	
DEAmine	Pure Component	

Status: OK

Figure 6.1 Implementation form for Acid Gas Removal Unit-Component List

Property Package Selection

<none>

Acid Gas - Chemical Solvents

Acid Gas - Physical Solvents

Acid Gas - Liquid Treating

Antoine

ASME Steam

Braun K10

BWRS

Chao Seader

Chien Null

CPA

Esso Tabular

Extended NRTL

GCEOS

General NRTL

Glycol Package

Grayson Streed

IAPWS-IF97

Ketner

Required Components

Amine
CO2
H2S
H2O

Supported Amines

DEA
DGA
DIPA
MDEA
MEA
PZ
TEA

Upgrade Parameters

Supported Amine Blends

DEA+DGA DEA+DIPA DEA+MDEA
DEA+MEA DEA+PZ DEA+TEA
DGA+DIPA DGA+MDEA DGA+MEA
DGA+PZ DGA+TEA DIPA+MDEA
DIPA+MEA DIPA+PZ DIPA+TEA
MDEA+MEA MDEA+PZ MDEA+TEA
MEA+PZ MEA+TEA PZ+TEA
SULFOLANE+DIPA
SULFOLANE+MDEA
MDEA+MEA+DEA
SULFOLANE+MDEA+PZ

Figure 6.2 Implementation form for Acid Gas Removal Unit-Fluid Package

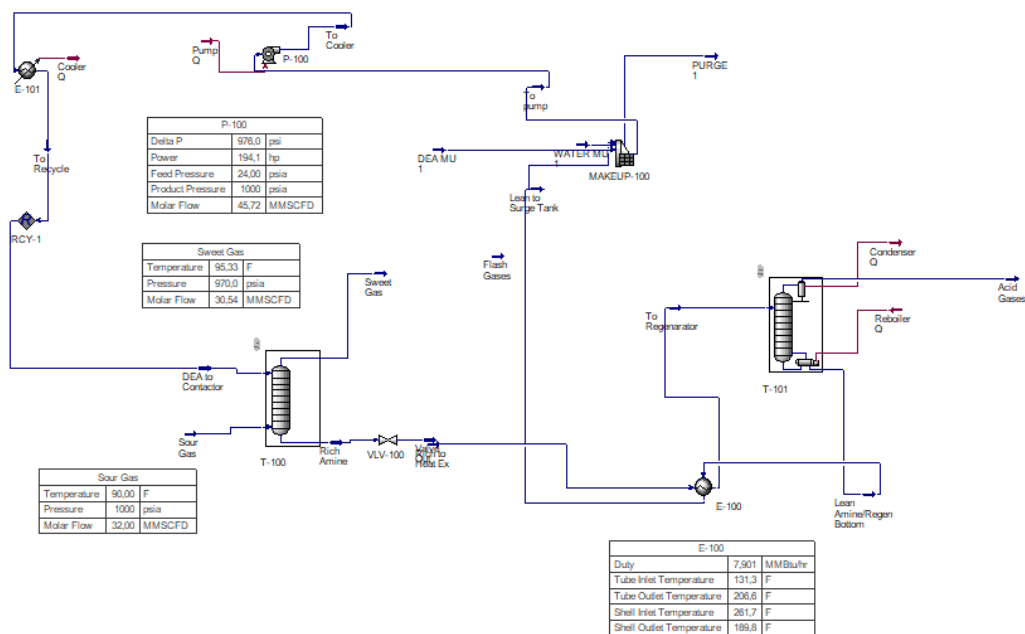


Figure 6.3 Acid Gas Removal Unit

Material Stream: Sour Gas

Worksheet	Attachments	Dynamics
Worksheet		
Conditions	Nitrogen	0,0171
Properties	CO2	0,0331
Composition	H2S	0,0067
Oil & Gas Feed	Methane	0,9036
Petroleum Assay	Ethane	0,0228
K Value	Propane	0,0051
Electrolytes	i-Butane	0,0012
User Variables	n-Butane	0,0013
Notes	i-Pentane	0,0006
Cost Parameters	n-Pentane	0,0004
Normalized Yields	n-Hexane	0,0006
Acid Gas	n-Heptane	0,0024
	H2O	0,0050
	DEAmine	0,0000

Figure 6.4 Sour Gas Composition

Material Stream: Sweet Gas

Worksheet	Attachments	Dynamics
Worksheet		
Conditions	Nitrogen	0,0179
Properties	CO2	0,0001
Composition	H2S	0,0000
Oil & Gas Feed	Methane	0,9448
Petroleum Assay	Ethane	0,0238
K Value	Propane	0,0054
Electrolytes	i-Butane	0,0013
User Variables	n-Butane	0,0014
Notes	i-Pentane	0,0006
Cost Parameters	n-Pentane	0,0004
Normalized Yields	n-Hexane	0,0006
Acid Gas	n-Heptane	0,0025
	H2O	0,0011
	DEAmine	0,0000

Figure 6.5 Sweet Gas Composition

Component List View: Basis-1 Component List [HYSYS Databanks]

Source Databank: HYSYS

Component	Type	Group
Nitrogen	Pure Component	
CO2	Pure Component	
H2S	Pure Component	
Methane	Pure Component	
Ethane	Pure Component	
Propane	Pure Component	
i-Butane	Pure Component	
n-Butane	Pure Component	
i-Pentane	Pure Component	
n-Pentane	Pure Component	
TEGlycol	Pure Component	
H2O	Pure Component	
n-Hexane	Pure Component	
n-Heptane	Pure Component	

< Add

Replace

Remove

Status: OK

Figure 6.6 Implementation form for Dehydration Unit-Component List

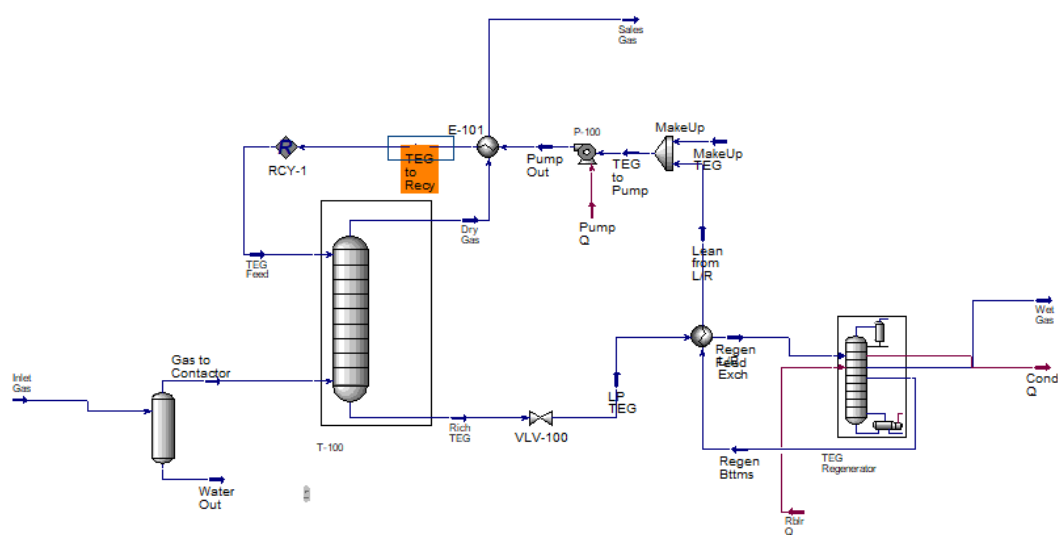


Figure 6.7 Dehydration Unit

Material Stream: Inlet Gas		
Worksheet	Attachments	Dynamics
Worksheet	Stream Name	Inlet Gas
Conditions	Vapour / Phase Fraction	0,9998
Properties	Temperature [C]	35,00
Composition	Pressure [kPa]	6688
Oil & Gas Feed	Molar Flow [kgmole/h]	1521
Petroleum Assay	Mass Flow [kg/h]	2,583e+004
K Value	Std Ideal Liq Vol Flow [m3/h]	82,72
User Variables	Molar Enthalpy [kJ/kgmole]	-7,536e+004
Notes	Molar Entropy [kJ/kgmole-C]	148,8
Cost Parameters	Heat Flow [kJ/h]	-1,146e+008
Normalized Yields	Liq Vol Flow @Std Cond [m3/h]	3,586e+004
	Fluid Package	<i>Basis-1</i>
	Utility Type	

Figure 6.8 Inlet Gas Properties

Material Stream: Inlet Gas		
Worksheet	Attachments	Dynamics
Worksheet	Mole Fractions	
Conditions	Nitrogen	0,0179
Properties	CO2	0,0001
Composition	H2S	0,0000
Oil & Gas Feed	Methane	0,9470
Petroleum Assay	Ethane	0,0239
K Value	Propane	0,0054
User Variables	i-Butane	0,0013
Notes	n-Butane	0,0014
Cost Parameters	i-Pentane	0,0006
Normalized Yields	n-Pentane	0,0004
	TEGlycol	0,0000
	H2O	0,0011
	n-Hexane	0,0006
	n-Heptane	0,0003
Total		1,00000
Edit...		View Properties...
		Basis...

Figure 6.9 Inlet Gas Composition

Material Stream: Sales Gas		
Worksheet	Attachments	Dynamics
Worksheet	Stream Name	Sales Gas
Conditions	Vapour / Phase Fraction	1,0000
Properties	Temperature [C]	36,02
Composition	Pressure [kPa]	6171
Oil & Gas Feed	Molar Flow [kgmole/h]	1519
Petroleum Assay	Mass Flow [kg/h]	2,580e+004
K Value	Std Ideal Liq Vol Flow [m3/h]	82,69
User Variables	Molar Enthalpy [kJ/kgmole]	-7,504e+004
Notes	Molar Entropy [kJ/kgmole-C]	149,8
Cost Parameters	Heat Flow [kJ/h]	-1,140e+008
Normalized Yields	Liq Vol Flow @Std Cond [m3/h]	3,582e+004
	Fluid Package	<i>Basis-1</i>
	Utility Type	

Figure 6.10 Sales Gas Conditions

Material Stream: Sales Gas		
Worksheet	Attachments	Dynamics
Worksheet	Mass Cv [kJ/kg-C]	1,755
Conditions	Cv (Ent. Method) [kJ/kgmole-C]	<empty>
Properties	Mass Cv (Ent. Method) [kJ/kg-C]	<empty>
Composition	Cp/Cv (Ent. Method)	<empty>
Oil & Gas Feed	Reid VP at 37.8 C [kPa]	<empty>
Petroleum Assay	True VP at 37.8 C [kPa]	<empty>
K Value	Liq. Vol. Flow - Sum(Std. Cond) [m3/h]	3,582e+004
User Variables	Viscosity Index	<empty>
Notes	HHV Molar Basis (Std) [kJ/kgmole]	9,028e+005
Cost Parameters	HHV Mass Basis (Std) [kJ/kg]	5,316e+004
Normalized Yields	CO2 Loading	<empty>
	CO2 Apparent Mole Conc. [kgmole/m3]	<empty>
	CO2 Apparent Wt. Conc. [kgmol/kg]	<empty>
	Phase Fraction [Act. Vol. Basis]	1,000
	Mass Exergy [kJ/kg]	578,3
	Ideal Gas Cp/Cv	1,287
	Ideal Gas Cp [kJ/kgmole-C]	37,27
	Mass Ideal Gas Cp [kJ/kg-C]	2,195
	Bubble Point Pressure [kPa]	<empty>

Property Correlation Controls
Preference Option: **Active**

Figure 6.11 Sales Gas Properties

Material Stream: Sales Gas

Worksheet | Attachments | Dynamics

Worksheet

- Conditions
- Properties
- Composition
- Oil & Gas Feed
- Petroleum Assay
- Value
- Derivatives
- Notes
- Cost Parameters
- Normalized Yields

	Mole Fractions
Nitrogen	0.0180
CO2	0.0001
H2S	0.0000
Methane	0.9480
Ethane	0.0239
Propane	0.0054
i-Butane	0.0013
n-Butane	0.0014
i-Pentane	0.0006
n-Pentane	0.0004
TEGlycol	0.0000
H2O	0.0000
n-Hexane	0.0006
n-Heptane	0.0003

Total: 1,00000

Edit... View Properties... Basis...

OK

Figure 6.12 Sales Gas Composition