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Development of Processes for Natural Gas Drying

Further exploring the TEG Injection Concept

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Problem Description

Bakgrunn:
Naturgass som produseres og transporteres fra norske felt for salg i Europa eller USA inneholder en del komponenter som må fjernes fra gassen. I tillegg til CO2 og H2S så må gassen tørkes for å unngå utfelling av vann som igjen kan danne hydrater i prosess- og transportsystem. Hvor mye gassen må tørkes avhenger av bruksområde og hvilke prosesser som er involvert. Det finnes flere tilgjengelige teknologier for tørking av gass som benyttes i dag. På anlegg der gassen sendes videre til sluttbruker via et rørsystem benyttes ofte absorpsjonsbasert tørkeprosess med glykol som kjemikalie. På anlegg der gassen skal prosesseres videre via nedkjøling benyttes ofte adsorpsjonsbasert tørkeprosess med molekular sieve. Både absorpsjon- og adsorpsjonsprosesser er energikrevende og plasskrevende, og det arbeides kontinuerlig med å finne frem til mer kompakte prosessløsninger. Alternative prosesser kan for eksempel innbefatte bruk av membraner eller kompakte absorpsjonskontaktorer.

Mål:
Evaluere eksisterende og nye prosessløsninger for tørking av naturgass med hovedmål om å finne nye konsepter som krever mindre plass (vekt/volum) og mindre energiforbruk.

Oppgaven bearbeides ut fra følgende punkter:

1.Videreutvikling av prosesskonsept (fra prosjektoppgave) for tørking av naturgass med fokus på reduksjon i vekt og volum av anlegget i forhold til konvensjonelle TEG (tri-etylen glykol) anlegg. Anlegget skal kunne tørke gassen både til rørtransportkvalitet og kaldprosesskvalitet

2.Etablering av simuleringsmodell for det nye konseptet i Hysys sammen med CPA (Cubic Pluss Assosiation) tilstandslikning for beregning av glykol-vann-hydrokarbon likevekt. Både tørkedel og regenereringsdel modelleres.

3.Gjennomføre simuleringer av konseptet samt design av hovedkomponenter med bruk av TEG i anlegget for å kunne sammenligne med eksisterende glykol (TEG) tørkeanlegg.

4.Sammenligning av simulerte prosessparametere ved bruk av forskjellige typer kjemikalier i tørkeanlegget som for eksempel glykoler (eksempelvis MEG (mono etylen glykol), DEG (de etylen glykol), TEG) og aminer (eksemplevis MDEA).

Assignment given: 16. January 2008
Supervisor: Arne Olav Fredheim, EPT

Preface

This paper is the result of my master thesis conducted as part of my Master in Science and Technology in Engineering & ICT with specialisation in the field of Industrial Processing at the Norwegian University of Science and Technology (NTNU). The work was done from January to June 2008, counting 30 study points, equivalent to 100% of my work load for this period.

The supervisors for this project have been Professor II Arne Olav Fredheim from NTNU and Lars Henrik Gjertsen from StatoilHydro.

Audun Bråthen
Trondheim, 6.6.2008

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Thanks to Kjersti Omdahl Christensen from StatoilHydro and Bernt Kalgraff from ProPure for giving me an introduction to the ProPure compact mixer technology.

Thanks to Even Solbraa and Svein Puntervold from StatoilHydro for helping out with using the CPA EoS with HYSYS. Also thanks to the IVCSEP research group at the Technical University of Denmark (DTU) for letting me use the CPA EoS for my simulations.

Thanks to Lars Erik Øi from Telemark University College (HiT) for helping me model the TEG regeneration process in HYSYS.

Also thanks to my roommates, friends and family for all the support during the work with this project, and for engaging me in leisure activities from time to time, making me regain my energy and motivation.

Abstrakt

Denne avhandlingen tar for seg videreutvikling av TEG injeksjonskonseptet presentert i Bråthen (2007). En introduksjon til separatorteknologi, konvensjonell glykol regenereringsprosess og kompaktmiksing blir først presentert, ettersom dette er viktige elementer i den alternative tørkeprosessen. Fordeler, ulemper og operasjonelle problemer påpekes, før problemene med TEG injeksjonskonseptet beskrives. Bruk av varm strippegass for regenerering av glykol foreslås som en mulig forbedring, men det viser seg at dette fører til store tap av glykol, stort behov for strippegass samt oksidasjon av glykol, noe som bidrar til at dette ikke inkluderes. De eneste forbedringene som inkluderes er bruk av inline-teknologi for de første ekstraksjonstrinnene, samt bruk av kompaktmiksere for miksing av TEG og naturgass. En simuleringsmodell for prosessen utarbeides i HYSYS, sammen med bruk av CPA-tilstandsligning og operasjonsdata fra Snøhvit LNG anlegget. Både tørkedelen og regenereringsdelen modelleres. Fra simuleringene finnes det at TEG injeksjon krever omtrent 50% større sirkulert mengde TEG for å oppnå samme vanninnhold i den tørka gassen som vanlig absorber tørking. Vekt og volum for absorpsjonsdelen for TEG injeksjon er derimot mye mindre når det sammenlignes med størrelser fra Kristinfeltet i Nordsjøen, hvilket til dels kompenserer for den økte mengden sirkulert TEG. Bruk av MEG og DEG istedenfor TEG i injeksjonskonseptet blir også simulert, men det konkluderes med at TEG er den beste absorbenten på bakgrunn av lavere energiforbruk for regenereringsdelen, mindre tap, samt bedre effektivitet med små til moderate mengder strippegass. På bakgrunn av meget store tap av absorbent konkluderes det også med at MEG er totalt uegnet for tørking.

Abstract

This paper treats further development of the TEG injection process described in Bråthen (2007). An introduction to separation technology, conventional glycol regeneration and compact mixing is presented, as these are important parts of the alternative dehydration concept. Advantages, disadvantages and operational problems are pointed out, before the problems with the TEG injection process is described. Using hot stripping gas for regeneration of the TEG is one of the suggested improvements, but large glycol losses, large flow rates of stripping gas and oxidizing of glycol are found to be the consequences, thus making the alternative unfeasible. The only improvements used, are to use inline separators for the first separation stages and compact mixers for mixing of TEG and natural gas. A simulation model is developed using the simulation software HYSYS with the CPA EoS as fluid package. Both the absorption and the regeneration part of the process are modeled, using operational data from the Snøhvit LNG facility is used as reference. From simulations it is found that TEG injection requires about 50% more circulated TEG than conventional absorber dehydration to obtain the same water content in the dehydrated gas. The weight and volume of the absorption part of the process is however found to be considerably smaller than the operational process at the Kristin field in the Norwegian North Sea, thus partly compensating for the increased TEG circulation rate. Use of MEG and DEG instead of TEG for the injection concept is also simulated, but it is concluded that TEG is the best suited because of lower regeneration energy, lower absorbent loss and best dehydration performance for low to intermediate flow rates of stripping gas. MEG is found to be unsuited for dehydration because of very large losses of absorbent.

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Abbreviations

ASME - American Society of Mechanical Engineers
BTEX - Collective term for benzene, toluene, ethyl benzene and xylene isomers
CFD - Computational Fluid Dynamics
CHIGP - Chemicals in Gas Processing, research project at DTU
CPA - Cubic Plus Association Equation of State
DEG - Di Ethylene Glycol
DTU - Technical University of Denmark
ECR - Elliot Combining Rules
EoS - Equation of State
gal - Gallon
gpm - Gallons per minute
HC - Hydrocarbon
HETP - Height Equivalent to Theoretical Plate
HX - Heat exchanger
LMTD - Logarithmic Mean Temperature Difference [K]
LNG - Liquefied Natural Gas
MAWP - Maximum Allowable Working Pressure
MEG - Mono Ethylene Glycol
micron - Micrometer (μm) [10^{-6}m]
MMscf - Million standard cubic feet
MMscfd - Million standard cubic feet daily
MMscm - Million standard cubic meter
PFD - Process Flow Diagram
ppm - Parts per million
psig - Pound per square inch gauge
SAFT - Statistical Association Fluid Theory
scf - Standard cubic feet
SG - Specific gravity
SRK - Soave-Redlich-Kwong Equation of State
TEG - Tri Ethylene Glycol
TTG - Tetra Ethylene Glycol
VOC - Volatile organic compound
wt% - Percent by weight

Nomenclature

A_G - Area available for gas [m^2]
A_{HT} - Heat transfer area [m^2]
A_L - Area available for liquid [m^2]
b - Component constant for CPA [m^3/kmole]
c_{gl} - Heat capacity of the glycol solution [J/kg*K]
$c_{P,G}$ - Heat capacity of gas at constant pressure [J/kg*K]
C_D - Newtonian friction factor [-]
d_{fil} - Characteristic filament dimension [-]
d_m - Diameter of small droplet [m]
D_i - Inner diameter [m]
g - Acceleration of gravity, approximately 9.81 [m/s^2]
$g(\bar{\rho})$ - Radial distribution function for CPA
h_{fg} - Heat of vaporization [J/kg]
h_L - Height of liquid volume [m]
k_{ij} - Interaction parameter for CPA
K_1 - Semi-empirical constant for Sounders-Brown equation [m/s]
L - Total length [m]
L_{eff} - Effective length usable for separation [m]
m_{gl} - Mass flow glycol [kg/s]
m_{reflux} - Mass flow reflux [kg/s]
M - Mass [kg]
N_A - Number of actual trays
N_{TH} - Number of theoretical trays
P - Pressure [Pa]
q - Heat flow [J/s] (=[W])
Q_G - Volume flow gas [m^3/s]
Q_L - Volume flow liquid [m^3/s]
$\bar{R}$ - Universal gas constant [kJ/kmole*K]
t_G - Retention time gas [s]
t_L - Retention time liquid [s]
t_W - Wall thickness [m]
T_{crit} - Critical temperature [K]
T_{gl} - Temperature of the glycol solution [K]
T_{SAT} - Saturation temperature [K]
U - Overall heat transfer coefficient [W/m^2*K]
U_G - Velocity gas [m/s]
U_L - Velocity liquid [m/s]
U_{rel} - Relative velocity between gas and liquid [m/s]
U_t - Falling velocity of droplet relative to gas [m/s]
$U_{t,L}$ - Falling velocity of droplet relative to liquid [m/s]
$\bar{v}$ - Molar volume [m^3/kmole]
V_L - Volume available for liquid [m^3]
We - Weber number [-]
We_{crit} - Critical Weber number [-]
wt% - Percent by weight
x_i - Molar fraction of substance I [-]
X_{Ai} - Fraction of sites belonging to molecule i that do not bond with other sites

$\alpha(T)$ - Energy parameter for CPA [Pa*m^6/kmole^2]
β_{AiBj} - Association volume for CPA [-]
Δ_{AiBj} - Association strength for CPA [m^3/kmole]
ε_{AiBj} - Association energy for CPA [kJ/kmole]
η_j - Joint efficiency [-]
η_t - Tray efficiency [-]
μ - Dynamic viscosity [cP – centi Poiseuille]
$\bar{\rho}$ - Molar density [kmole/m^3]
ρ_G - Density gas [kg/m^3]
ρ_L - Density liquid [kg/m^3]
ρ_{mix} - Density of mixture [kg/m^3]
ρ_W - Density of water [kg/m^3], approximately 1000 kg/m^3
σ_{surf} - Surface tension [N/m^2]
σ_{max} - Maximum allowable stress [N/m^2]

Introduction

Water is an impurity found in all natural gas reservoirs to various extents all across the world (Guo & Ghalambor (2005) [1]). However, as long as the water is in vapor phase it presents no danger, but as soon as liquid water is formed several problems arise. High pressure hydrocarbons together with liquid water are able to form an ice like structure called hydrates at ambient temperatures, which can plug the entire pipe or channels in process equipment if allowed to develop. Sour gases, another common impurity of natural gas, mixed with liquid water are corrosive, and of course if the temperature gets low enough, regular ice can form (Kohl & Nielsen (1997) [1]). The solution is to reduce the water concentration enough, so that the water present is in vapor form for all conditions encountered throughout the process. For hydrate inhibition use of chemicals is also possible, as is most often employed for pipeline transport from the wellhead to the platform or land based processing facility.

The most commonly used processes for dehydration is absorption into a liquid or adsorption into a solid, but cooling and compression of the gas can also be used (Kohl & Nielsen (1997) [1]). Absorption is used for dehydration when the requirement for water content is moderate, typically for dehydration before sending the gas through long distance pipelines. Adsorption can produce extreme dryness in the gas, and is used before low temperature processes (e.g. liquefaction or heavy hydrocarbon extraction). This paper will focus on absorption.

Today most absorption processes uses glycol, more specifically tri ethylene glycol (TEG), in an absorption column where lean TEG (lean in water) is introduced from the top while wet gas flows in through the bottom (Arnold & Stewart (1999) [1]). As the TEG flows down the column over trays or packing, water is absorbed from the rising gas, letting dehydrated gas flow from the top of the column, while rich TEG (rich in water) exits from the bottom. The rich TEG is regenerated by reducing the pressure and increasing the temperature, before it is re-injected into the absorption column again. This equipment is very large and heavy, two major cons on offshore platforms.

Throughout my project thesis (Bråthen (2007)) I explored different ways of improving upon the conventional absorption process. The result was the proposition for an alternative dehydration concept; injecting TEG directly into the gas stream and utilizing the piping as a contactor. Three stages of injection and extraction was found to be the most promising concept, and simulations were conducted to compare the concept to regular absorption dehydration. It was found that the flow rate of TEG had to be considerably larger for the TEG injection concept to produce the same dryness as the absorber process, thus reducing the potential of the alternative concept (Bråthen (2007)).

This paper will try to further develop the TEG injection concept by introducing inline separators, compact mixer units, and improving upon the regeneration process as well, to potentially make the concept more competitive when compared to conventional dehydration.

First a literature review of separators and the glycol regeneration process is presented, to better be able to suggest improvements and to discuss the suggested improvements. Then the compact mixing unit from ProPure is presented, before further development of the TEG injection process is discussed. Chapter 5 describes the making of the HYSYS simulation model, and lists all the assumptions and decisions made. Simulations comparing the alternative concept to conventional dehydration, as well as simulations using different absorption mediums for TEG injection are presented in chapter 6. Thereafter, based on data from the simulation model, the different components of the TEG injection process are designed using the design models from appendix A and C, and the results compared to conventional absorber dehydration. Finally an overview of the paper is given before a conclusion for the TEG injection process is reached.

Chapter 1: Separators

Separators play a very important part in most processing facilities found today, and is also a fundamental part of the proposed alternative natural gas dehydration concept. Their use is to extract the liquid glycol and water mixture from the gas for the various stages, so the liquid can be recycled to the preceding stage or sent to regeneration (depending on which stage), and the dehydrated gas passed to the next stage or further processing. The type of separator used depends on the tolerances for liquid carry-over, pressure drop and available space, among other factors.

Separators with very high gas to liquid ratio are often referred to as gas scrubbers or simply scrubbers. They are often used upstream of equipment unable to handle liquids, e.g. compressors and contactor towers, or downstream of equipment involving liquids or where liquid might form, e.g. sour gas treating, glycol dehydration units or coolers. As the glycol concentration when injected in the natural gas stream is very small, the separators used in the concept can therefore be referred to as scrubbers.

This chapter gives an introduction to the function of separators and also operation problems experienced during operation. First the function of the separator is explained, discussing the different types of separators and the most common internal equipment. Secondly typical problems experienced with operating separators are described, followed by possible solutions. A short discussion of three-phase separators then follows, describing the differences compared to two-phase separators. Finally an introduction to inline separation is presented. Design procedures for the described separator types can be found in appendix A.

1.1 Function

Generally two types of separators are most commonly used; vertical and horizontal with cylindrical shell (Arnold & Stewart (2008) [1]). Other separator designs include spherical-, centrifugal-, venturi-, filter- and double-barrel horizontal separators. Vertical, horizontal and centrifugal separators will be discussed here, while those interested in other types are referred to Arnold & Stewart (2008) [1]. Regardless of design, each two-phase separator contains four major sections (except centrifugal separators); inlet diverter, liquid collection, gravity settling and mist extractor. See figure 1.1 for an illustration showing the different sections for a vertical separator.

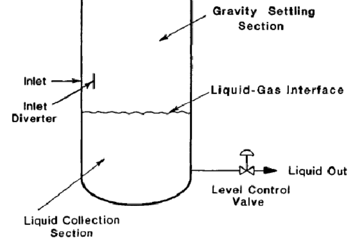

Figure 1.1: The different sections of a separator (adapted from Arnold & Stewart (1998):p.104)

1.1.1 Inlet diverter

The inlet diverter is the first section encountered by the process stream, located just inside the inlet of the separator. Its purpose is to absorb the momentum from the stream by altering the direction of flow, and at the same time working as the first stage of separation. There are many types of inlet diverters, but they can be classified in three main groups; baffle plates, centrifugal diverters and elbows illustrated in figure 1.2 (Arnold & Stewart (2008) [1]).

Baffle plates comes in many shapes and forms, including cones, spherical disks, angle irons and flat plates, about anything able to cause a sudden change in the direction of the flow. The heavier droplets entrained in the gas are not able to adjust for this sudden change in flow direction, and will therefore collide with the inlet diverter, coalesced and fall into the bottom of the separator.

Centrifugal inlet diverters utilize centrifugal forces, created either by a tangential inlet or a cyclonic chimney. Droplets are then forced to the walls by the centrifugal forces, where they coalesced and flow to the bottom of the separator. Centrifugal inlet diverters are the most common in natural gas processing (Guo & Ghalambor (2005) [2]), but baffles are also used. This choice depends on the amount of liquid in the gas, the gas velocity, if slug flow is possible and the regularity of the stream. Centrifugal inlet diverters are very dependent on the flow rate of the gas, as they will not work properly for low flow rates.

Elbows consist merely of a bent pipe section serving the same purpose as a baffle plate.

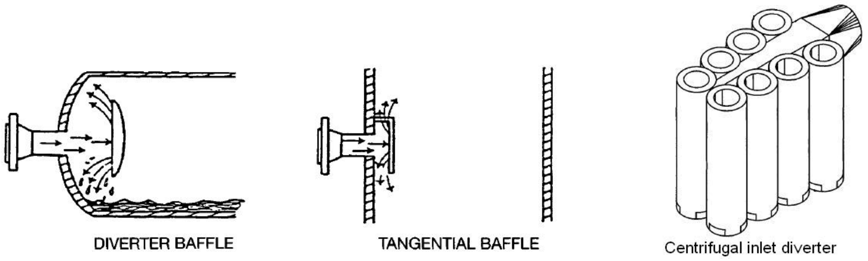

Figure 1.2: Different types of inlet diverters (adapted from Arnold & Stewart (1998):p.109,110)

1.1.2 Liquid collection section

The liquid collection section of the separator, located in the bottom of the vessel, is where the liquid is processed. The liquid has to stay in this section for a given time (retention time) to let the gas bubbles entrained in the liquid escape to the gravity settling section. Retention time is a function of total liquid capacity, feed flow rate, required degree of separation, and the difference in density between the liquid and the gas (Arnold & Stewart (2008) [1]). When designing a separator, the retention time dictates the required size of the liquid collection section. If there is a problem with slug flow, the liquid collection section also acts as a surge volume and has to be sized accordingly.

Depending on the operating conditions for the separator, additional equipment may also be necessary in the liquid collection section. For horizontal separators waves can form on the liquid surface because of intermittent slugs, irregular liquid flow or external movement (e.g. on floating structures), and wave breakers must be installed. A wave breaker consists merely of perforated baffles or plates placed perpendicular to the flow. It is very important to prevent waves from forming, as waves can prevent weirs, liquid level controllers and level safety switches from operating properly (Arnold & Stewart (2008) [1]).

Irregular liquid flow can also result in low liquid levels in the liquid collection section, potentially creating a vortex near the liquid outlet which can suck gas into the liquid outlet line. This can be prevented by installing a vortex breaker, e.g. a cylinder with radially directed flat plates.

1.1.3 Gravity settling section

In the gravity settling section the gas is allowed a greater cross section area, thus reducing the velocity of the gas, thereby also reducing the turbulence. As a result of this, the drag force from the gas on the liquid droplets will decrease (See e.g. White (2003):p.478), and droplets will fall to the liquid surface by the force of gravity. Parallel plates can also be installed to further decrease the turbulence (Asheim & Herfjord (1984)), and also mitigate foaming if that is a problem.

The size of the gravity settling section is dependent on how small droplets that needs to be separated out. A common design criterion for a separator is to remove all liquid droplets greater that 100 – 140 microns, while for a scrubber removal of droplets greater than 500 microns might be sufficient (Arnold & Stewart (2008) [1]). The background for this criterion is to remove enough of the liquid so that the mist extractor will not become flooded, and thereby function properly.

1.1.4 Mist extractor section

The mist extractor section is the final separation stage in a separator, and is designed to coalesce the smallest droplets so that they can be extracted from the gas stream. Intricate flow patterns force the droplets onto the solid surfaces, as the droplets are not able to adjust to the constant changes in direction. There are different types of mist extractors including wire-mesh, corrugated plates and hooked vanes, but the wire-mesh outnumbers all of the others (Arnold & Stewart (2008) [1]). The wire-mesh units are knitted from very thin metal wires (0.1 – 0.28 mm), and features a very large surface area for a given volume (See figure 1.3).

Figure 1.3: A typical wire-mesh mist extractor (adapted from Arnold & Stewart (1998):p.113)

Effectiveness of the wire-mesh unit is strongly dependent on the gas velocity being close to the design velocity. Too high velocity causes the knocked out liquids to be re-entrained, while too low velocity causes the gas to just drift through the wire-mesh without knocking out any liquids (Arnold & Stewart (2008) [1]).

1.2 Configurations

As mentioned earlier, separators appear in a lot of different configurations. The following paragraphs will give a short introduction to horizontal, vertical and centrifugal separators.

1.2.1 Horizontal separator

Horizontal separators are usually the first choice because of their low cost, due to their smaller size for a given gas and liquid flow rate (Guo & Ghalambor (2005) [2]). The feed enters from the left side (See figure. 1.4) and encounters the inlet diverter, forcing the biggest drops out of the gas stream. The feed stream continues horizontally through the gravity settling section with droplets continuously falling toward the liquid/vapor interface. Horizontal separators are usually half filled with liquid, thereby providing the largest possible liquid/vapor interface (Arnold & Stewart (2008) [1]).

As horizontal separators are mounted horizontally, there is always a force acting towards the liquid interface regardless of the gas velocity. There is however still a possibility for liquid carry-over, if the liquid droplet has not reached the liquid surface before the gas outlet (Asheim & Herfjord (1984)). For a proper designed horizontal separator, this requirement is generally met with gas velocities greater than the terminal velocity (See formula A.1 in appendix A) (Asheim & Herfjord (1984)).

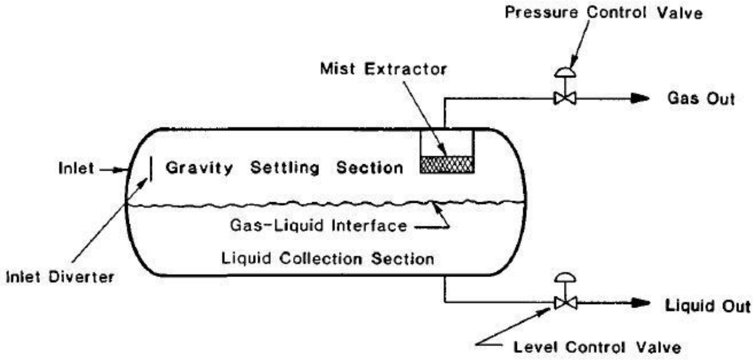

Figure 1.4: Schematics of a typical horizontal separator (adapted from Arnold & Stewart (1998):p.103)

After the gravity settling section the gas enters the mist extractor, where the final separation takes place. The mist extractor is normally mounted horizontally with the gas flowing vertical upwards. This is due to problems with flooding when mounted vertically (Arnold & Stewart (2008) [1]).

Horizontal separators are commonly used in streams with high gas to liquid ratios, foaming well streams or liquid-liquid separation (Guo & Ghalambor (2005) [2]). The large liquid/gas interface enhances both gravity separation and helps gas bubbles escape from the liquid, but at the same time makes the separator vulnerable to liquid slugs and liquid surges. The separator also takes up a larger footprint than a vertical separator, and solids can deposit on the bottom of the vessel (GPSA (2004) [1]). A horizontal separator with an attached second barrel for liquid is known as a double barrel horizontal separator.

Advantages:

- Cheaper to build for a given flow rate compared to other configurations
- Very large liquid/vapor interface
- Can generally handle larger gas velocities than vertical separators
- Better at handling foaming

Disadvantages:

- Large footprint
- Limited surge capacity
- Possible problem with solid deposition

1.2.2 Vertical separators

Vertical separators occupy considerable less floor space than horizontal separators, thereby gaining an advantage for offshore processing facilities where space is limited. As there is a large vertical distance between the feed stream inlet and the gas outlet, chances for liquid to be re-entrained and carried out of the gas inlet is also much smaller. This also makes the vertical separator less vulnerable to liquid slugs (Guo & Ghalambor (2005) [2]).

The feed stream enters from the left side as for the horizontal separator (See figure 1.5), and hits the inlet diverter. Large droplets are flung into the liquid collection section by the abrupt change in direction, while the gas and the smaller droplets flow onward to the gravity settling section. As the force of gravity and drag acts in opposite directions of each other, the velocity has to be limited to reduce the drag force, allowing droplets to fall to the liquid collection section. The drag force exerted by the gas on the liquid droplets is proportional to the square of the gas velocity (See e.g. White (2003):p.478). Gas velocity is proportional to the vessel diameter for a given volume stream. Thus, for the force of gravity to be greater than the drag force, the diameter has to be larger than a given minimum diameter (See appendix A).

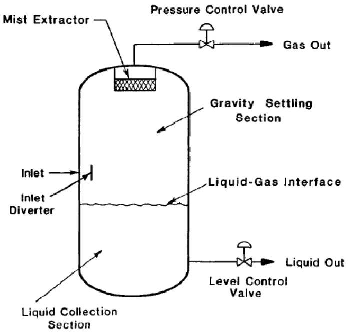

Figure 1.5: The different sections of a separator (adapted from Arnold & Stewart (1998):p.104)

From the gravity settling section the gas flows upwards to the mist extractor for the final separation stage. This section is identical to the horizontal separator, with the gas flowing upwards. The gas then leaves the separator through the top.

Vertical separators are commonly used where the gas to liquid ratio is low to intermediate, or if there is a problem with impurities or solids in the stream. They are also used where the gas to liquid ratio is very high (e.g. gas scrubbers) (Arnold & Stewart (2008) [1]). As they take up less floor space than horizontal separators, they are also frequently employed in facilities with limited space. They are however more costly to produce than horizontal separators (Guo & Ghalambor (2005) [2]).

Advantages:

- Small footprint

- Good at handling solids
- Large surge capacity

Disadvantages:

- Expensive
- Terminal velocity for the liquid droplets dictates a minimum diameter
- Small liquid/vapor interface

1.2.3 Centrifugal separators

Instead of using gravity to separate vapor and liquid, the centrifugal separator utilizes the large centrifugal forces created by rotating a stream at high velocities for separation. This force ranges from 5 to 2500 times the force of gravity, depending on the flow rate of the stream and the diameter of the pipe (Arnold & Stewart (2008) [1]). It is fundamentally different from horizontal and vertical separators, and does not contain the same sections as previously described.

The inlet stream enters through a tilted tangential inlet (See figure 1.6) causing the gas to spiral down towards the liquid outlet. To stratify the inlet stream and enhance separation, the inlet pipe is inclined at an optimum angle determined by CFD analysis (Arnold & Stewart (2008) [1]). As the gas reaches the bottom, most of the droplets have been flung onto the walls by the centrifugal forces, and the liquid free gas can leave upwards through the centre. The liquid drains down from the walls and leaves through the liquid outlet located at the bottom.

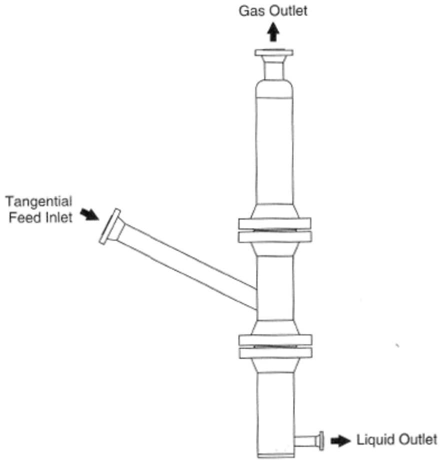

Figure 1.6: Centrifugal separator (adapted from Arnold & Stewart (2008):p.159)

Centrifugal separators are not capable of removing liquid down to the same concentrations as conventional horizontal or vertical separators, and are therefore best suited for bulk separation (Arnold & Stewart (2008) [1]). The high velocity swirl also makes them vulnerable to erosion from solid particles on the pipe walls. Today centrifugal separators are frequently employed downstream of glycol absorption columns for glycol recovery (Arnold & Stewart (2008) [1]), or where space is limited, e.g. offshore platforms.

There are also several special types of centrifugal separators available on the market today, e.g. the Twister supersonic separator (Twister BV (undated)), and the CDS-Gasunie Cyclone (CDS Separation Technology (undated)). See chapter 1.5 for an introduction.

Advantages:

- Small and compact design
- No moving parts
- Little maintenance
- Low cost compared to conventional technology

Disadvantages:

- Sensitive to flow rate; Too low flow rate results in a low centrifugal force hampering separation
- Requires a greater pressure drop
- Each unit must be tailor made for each project

1.3 Three-phase separators

A three-phase separator works in the same way as a two-phase separator, except that in addition to separating the liquid from the gas, the separator must also divide the liquid into separate phases. This requires longer retention times for the liquid and also additional equipment in the liquid collection section. Both vertical and horizontal configurations can be used, subject to the same considerations as for two-phase separation.

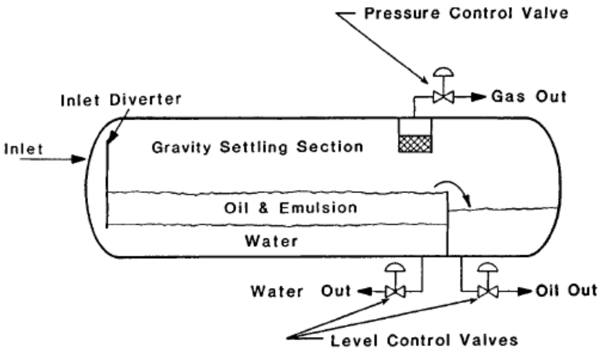

Figure 1.7: Horizontal three phase separator (adapted from Arnold & Stewart (1998):p.137)

To separate the two liquids, the use of weirs is often employed (See figure 1.7) (Arnold & Stewart (2008) [3]). Because of different densities of the two liquids, the lightest liquid will after a certain time float on top of the heavier liquid. By designing a weir to have a height greater than the thickness of the heaviest liquid layer, the only liquid allowed to overflow the weir is the lightest. The use of weir requires however a very strict control of the liquid levels, as too low or too high liquid levels can result in re-mixing of the liquids through the outlets.

In addition to weirs, coalescing plates can also be installed in the liquid collection section to aid liquid separation. The coalescing plates will increase the droplets entrained in both the liquids, enhancing gravity separation and thereby reducing the required retention time, again reducing the required size of the separator (assuming that size is constrained by liquid flow rate) (Arnold & Stewart (2008) [3]). The liquid retention times for three phase separators are generally longer than for two phase separators and retention times as long as 30 to 45 minutes is normal.

1.4 Operation problems

Several problems can arise when operating a separator, most subject to the fluid composition and the separation technique used. Sand and other solids carried with the feed stream can deposit on the vessel bottom, paraffin or other plugging elements may plug the mist extractor, liquid HC may cause foaming, liquid can be carried-over through the gas outlet, or gas can be carried-under through the liquid outlet. Now each of the problems will be addressed more thoroughly, and possible solutions/prevention techniques presented.

1.4.1 Sand deposition

Sand or other solid particles can be carried from the natural gas reservoir, through pipelines and into the inlet separator of the first processing facility (e.g. an offshore platform). If the separator is not properly designed, the solids could plug separator internals or accumulate at the bottom of the separator, thereby decreasing the volume available for the liquid. For vertical separators a cone can readily be placed at the bottom of the separator for sand collection, and the sand can then be removed periodically through an access hole, or continuously through a separate outlet (Arnold & Stewart (2008) [1]). Alternately it can also be transported with the liquid to be handled by downstream equipment. Using a cone is however only done in cases where major problems with sand are anticipated. For minor sand problems, the separator can be periodically cleaned internally with steam or water, washing the sand out with it.

Avoiding sand problems in horizontal separators are harder, as the sand easily deposits along the full length of the separator, and not only around one outlet. One possible solution is to install sand jets and drains at regular intervals (normally less than 1.5 m (Arnold & Stewart (2008) [1])) to flush out the deposited solids (see figure 1.8). Liquid (normally produced water) is then sprayed through the sand jets onto the deposited solids, so the solids break up and drains with the liquid out of the vessel.

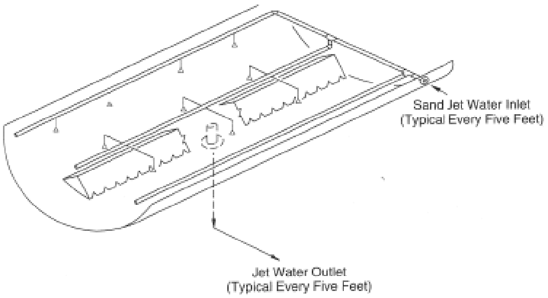

Figure 1.8: Horizontal separator with installed sand jets (adapted from Arnold & Stewart (2008):p.175)

Sand deposition is normally not an issue for centrifugal separators, as the velocity throughout the entire vessel is very high, preventing the sand from settling out. The fast rotating sand particles may however lead to excessive erosion of the vessel walls, requiring units where sand/solids may be present to be manufactured from special materials.

1.4.2 Plugging of mist extractor

Paraffin or other plugging elements can accumulate in tight passages through the separator, and severely hamper separation efficiency. A wire-mesh mist extractor exposed to paraffin over time can for example plug and be rendered useless, resulting in liquid carry-over. Coalescing plates in the gravity settling section can also be plugged, leading to less cross section area available to the gas stream, leading to higher gas velocities and poor separation (Arnold & Stewart (2008) [1]). To counter this problem, alternative mist eliminators should be employed, and the vessel equipped with manways or handholes to allow steam, solvent or other types of cleaning of the separator internals.

1.4.3 Foaming

Foaming is a dispersion of gas in a liquid created by agitation (Toshiba (undated)). In most cases, the foam will rapidly break up as the bubbles expand and the surface tension increases, causing the bubbles to rupture. If this is completed before the outlet of the unit, the foaming may present no problem to the unit`s operation (Arnold & Stewart (2008) [1]). However, some components tend to stabilize the foam, increasing the lifespan of the bubbles and ultimately leading to bubbles being carried over from the separator. For glycol systems, corrosion inhibitors, solids, salt water, thermal decomposition products and degradation products from oxidized glycol have been found to stabilize foam (Toshiba (undated), Kohl & Nielsen (1997) [1]). Typical foaming problems reported include large liquid carry-over, off-spec product, and severely decreased unit efficiency caused by reduced throughput (Toshiba (undated)).

The priority when experiencing foaming problems should be to locate the source of the foam stabilizing components and prevent them from entering the unit. For immediate foam prevention however, anti-foam chemicals can be introduced in the process stream. The use of anti-foam chemicals requires very careful mixing, as excessive dosage of anti-foaming chemicals actually can increase the foaming (Arnold & Stewart (2008) [1]). Coalescing devices (e.g. parallel plates) can also be used to break up the foam.

1.4.4 Carry-over of liquid through the gas outlet

Regardless of how finely tuned a specific separator design is, there will always be traces of liquid left in the gas (The Two-Phase Flow Committee at NTH Lecture Series No.9 (1984)). The designers goal for a separator is however to minimize this amount, as liquid might damage downstream equipment, represent economical losses, or result in off-spec product. Even though a separator is properly designed, carry-over might still occur. Faulty liquid level control resulting in high liquid levels, damage to vessel internals, foaming, plugged liquid outlets or excessive flow rate might all result in liquid carry-over (Arnold & Stewart (2008) [1]).

Carry-over is best prevented by using a properly designed mist extractor, paired with a liquid high level safety sensor that shuts of inlet flow if the liquid level reaches too high (Arnold & Stewart (2008) [1]).

1.4.5 Carry-under of gas through the liquid outlet

Low liquid levels, vortexing and liquid level control failure, are reason that can lead to gas escaping through the liquid outlet. This can result in extensive damage to downstream equipment, e.g. cavitation on pump blades or vessel rupture because of over-pressure. Carry-under can be prevented by using vortex breakers and a liquid low level safety sensor that shuts of liquid outflow if the liquid level becomes too low (Arnold & Stewart (2008) [1]).

1.5 New technology – inline separation

As space on offshore production facilities is very limited and with subsea processing becoming more and more popular, the need for compact and durable process equipment is growing fast. Inline separators is one example of such equipment, replacing big separator tanks with units able to be installed directly in the pipeline (hence the name inline). Inline equipment is also very attractive for de-bottlenecking existing units. This chapter will give a short introduction to the CDS inline separator, as well as the Twister supersonic separator.

1.5.1 The CDS-Gasunie cyclone

The CDS-Gasunie cyclone is an inline separator utilizing centrifugal forces to separate liquid and vapor developed by CDS and StatoilHydro. The cyclone is based on theory developed for centrifugal separators, and can be regarded as a mono-cyclone (spirals in only one direction, contrary to up and down in regular centrifugal separators). It comes in two versions; the Degasser for removal of vapor from a continuous liquid stream, and the Deliquidiser for removal of liquid from a continuous gas stream (Schook & Håland (undated)).

The Degasser

The process stream enters from the left (See figure 1.9) and is passed through a mixing element to prevent a stratified (layered) stream from entering the vessel, as stratified flow will reduce the separation efficiency considerably. A stationary swirling element brings the stream into rotation and because of the difference in density, the liquid is forced to the walls while the gas stays in the center. The gas leaves the vessel through an annular section in the cyclone inner region, and is extracted to the vertical scrubber section on top of the Degasser. In the scrubber section the small amount of liquid entrained in the gas is separated from the gas and re-injected into the liquid stream. Finally the processed liquid is routed through a de-swirling element to recover pressure and reduce the pressure drop over the unit (Schook & Håland (undated)).

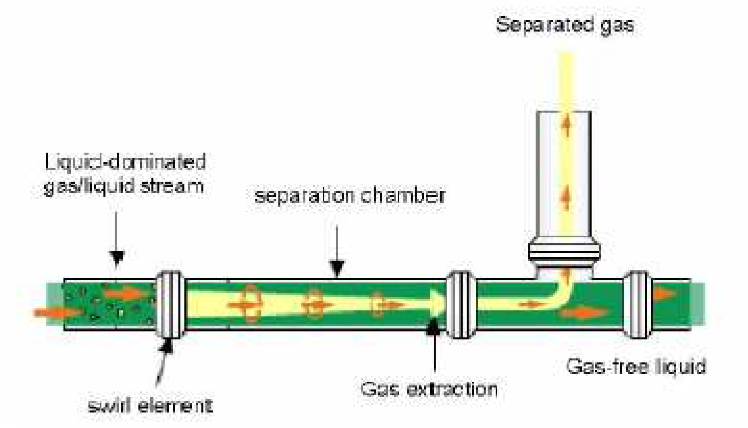

Figure 1.9: Schematics for the Degasser Inline separator (adapted from Schook & Håland (undated):p.4)

Table 1.1: Performance data for the CDS-Gasunie Degasser (adapted from Schook & Håland (undated):p.4)

Maximum diameter (D_i)	Unlimited
Length	8 - 10 * D_i
Height	4 - 6 * D_i
Liquid load	Unlimited
Gas load	< 55% volume fraction
Pressure drop	0.5 - 2.5 bar
Separation efficiency	90 - 99% gas from liquid
Liquid turndown	50 %
Gas turndown	55 - 0 %

All Degasser units are tailor-made for each process, able to optimize for pressure drop, separation efficiency, turndown or a combination of these parameters (Schook & Håland (undated)).

The Deliquidiser

As for the Degasser, the process stream enters from the left (See figure 1.10) and is mixed through a mixing element to prevent stratified flow. A stationary swirl element brings the fluid into rotation, forcing the liquid to form a film on the pipe walls. The gas is removed through a pipe inserted into the center of the unit, as the liquid enters the space between the original pipe wall and the gas pipe and is collected in a vertical boot section. Some gas will inevitably follow the liquid, and is removed from the liquid in the boot and re-injected into the gas stream. A de-swirling element is located close to the exit of the unit to recover pressure and reduce the pressure drop over the unit (Schook & Håland (undated)).

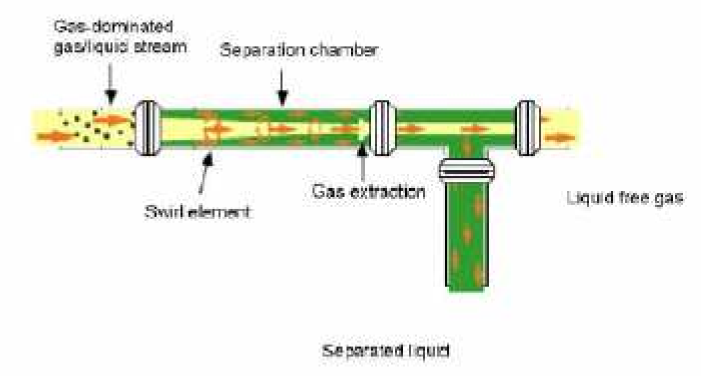

Figure 1.10: Schematics for the Deliquidiser inline separator (adapted from Schook & Håland (undated):p.5)

Table 1.2: Performance data for the CDS-Gasunie Deliquidiser (adapted from Schook & Håland (undated):p.5)

Maximum diameter (D_i)	Unlimited
Length	6 - 8 * D_i
Height	4 - 6 * D_i
Gas load	Unlimited
Liquid load	< 10% volume fraction
Pressure drop	0.4 - 0.7 bar
Separation efficiency	90 - 99% liquid from gas
Gas turndown	50 %

All Deliquidiser units are also tailor-made for each process, able to optimize for pressure drop, separation efficiency, turndown or a combination of these parameters (Schook & Håland (undated)).

Schook & Håland (undated) presents several examples of Degassers and Deliquidisers in operation at various locations in the world today. On the Norwegian continental shelf it is used at the Sleipner-T, Statfjord-B and Veslefrikk fields with great success. However most applications are for de-bottlenecking of existing systems, and no process described employs the Degasser or Deliquidiser as the only unit for separation (Schook & Håland (undated)).

1.5.2 The Twister BV supersonic separator

The Twister BV supersonic separator also employs centrifugal forces to separate gas and liquid, but in addition to rotate the stream, the stream is accelerated to supersonic velocities through a Laval nozzle. Even though the Twister is named “separator”, the main use is however conditioning the gas to meet pipeline specifications.

The feed gas enters from the left (See figure 1.11) and is set into rotation by stationary guide vanes located in the vortex generation section. Passing the stream through a Laval nozzle with a sufficiently high volume stream makes the flow supersonic, greatly intensifying the swirl and lowering the temperature. The swirl flings the liquid droplets onto the vessel walls, and the liquid and some slip gas is extracted through a cyclone separator section. The slip gas is recovered in a hydrate separator, and the liquid sent to treatment (Twister BV (undated)).

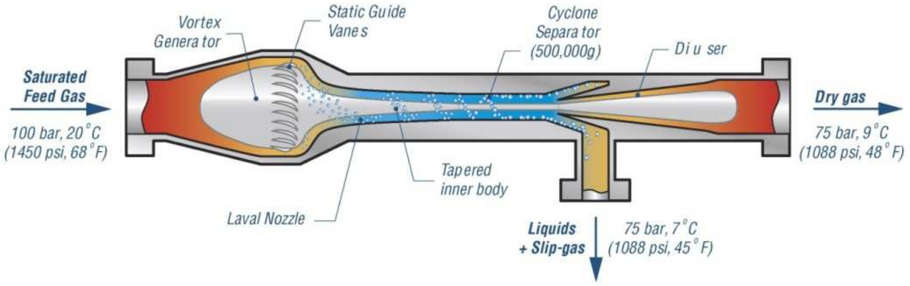

Figure 1.11: Schematics for the Twister BV Supersonic Separator (adapted from Twister BV (undated))

Performance data for the Twister BV is not available, but the unit must be tailor made for all projects. The two main disadvantages of the Twister, is the required pressure loss of 15 – 30%, and the fact that a Laval nozzle is a fixed flow rate device, allowing for very little turndown (Twister BV (undated)). Advantages include no moving parts, compact and lightweight and no need for chemicals when processing for pipeline specifications (**Note:** Not able to produce pipeline specifications for all locations, e.g. the North Sea).

Chapter 2: Glycol regeneration system

Even though the TEG injection concept uses a different method for dehydrating the natural gas, the rich glycol still has to be regenerated using the conventional regeneration process. This chapter will give an introduction to this process, discussing the process and typical operational problems. First the process is presented and the different unit`s operation described. Operational problems are then discussed, shortly describing the problem and giving possible solutions. Finally different processes for enhanced glycol regeneration are presented. Design models are described in appendix C.

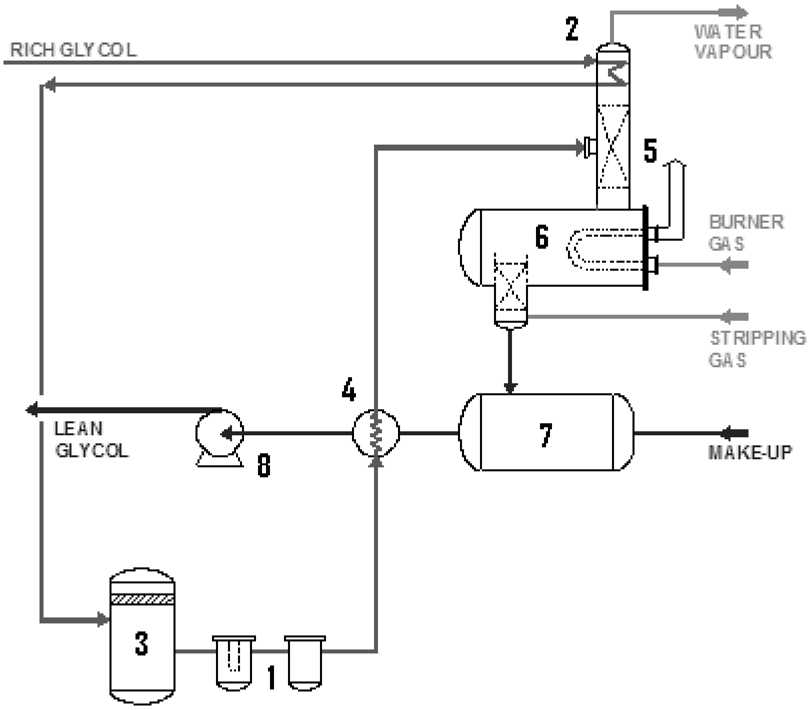

Figure 2.1: Typical glycol regeneration system (adapted from Apex Process Group (2007))

The process

The rich glycol (rich in water) exits the absorber in the top left corner (referring to figure 2.1), is heated through the overhead condenser for the regeneration still, and enters the flash tank. The flash tank operates at a lower pressure than the absorber, and some HC and water are therefore flashed off. From the flash tank, the glycol passes through two filters, one to remove solids entrained in the glycol (called sock filter) and one to remove liquid HC (charcoal filter often employed). It is heated through the glycol/glycol heat exchanger, and finally entered into the regeneration still.

In the regeneration still heat is added through a reboiler (normally by burning natural gas), vaporizing the water together with some glycol. Through the regeneration column, hot rising vapor containing glycol and water is gradually cooled, causing glycol to condense and fall to the bottom, while essentially water vapor leaves from the top. At the top of the regeneration column there is an overhead condenser, providing reflux for the column. From the bottom of the regeneration column lean glycol overflows by gravity to the surge tank, before it is cooled in the glycol/ glycol heat exchanger and pressurized by the glycol pump. Finally the lean glycol is cooled by the dehydrated gas before it is re-entered into the absorber (not shown in the figure) (Carroll (2003), Lyons & Plisga (2005), Guo & Ghalambor (2005) [1]).

2.1 The different units` operation

A typical glycol regeneration system (as shown in figure 2.1) consists of a flash tank for recovering HC trapped in the glycol, several filters for removing impurities, a regeneration column with a reboiler, a heat exchangers for cooling the lean glycol and heating the rich glycol, and a pump for pressurizing the lean glycol before it is sent back to the absorber again (Guo & Ghalambor (2005) [1]). In addition processes for enhanced glycol regeneration (i.e. higher purity in the lean glycol) e.g. introducing stripping gas in the stripping column or using the Coldfinger or the Drizo-processes can be employed requiring additional equipment (See chapter 2.3).

2.1.1 Filters

Even though an inlet separator is used to remove solids and other contaminants before the feed gas enters the absorber, solids will accumulate in the glycol if not removed. A sock filter is therefore employed shortly after the rich glycol leaves the absorber, removing any contaminants. Lyons & Plisga (2005) and Guo & Ghalambor (2005) [1] both claim that the filter should be located before the flash tank filtrating high pressure glycol, while Kidnay & Parrish (2006), Arnold & Stewart (1999) [1] and Carroll (2003) argue it should be placed after the flash tank filtrating low pressure glycol (See point 1 figure 2.1), thus indicating that both placements are used. Filtrating high pressure glycol enhances filtration efficiency and prevent solids from possibly plugging the overhead condenser, but requires pipe joints also in front of the flash tank. Placing the filter after the flash tank enables the sock filter and the charcoal filter to be placed directly after each other, limiting pipe joints, but can result in possible problems with solid deposition in the flash tank.

In addition to a sock filter, a charcoal filter is used to remove small amounts of hydrocarbons that are not flashed off in the flash tank. To serve any purpose, this filter must be placed after the flash tank, filtrating the low pressure glycol. For large glycol regeneration systems, it is common to filtrate only a portion of the glycol through the charcoal filter to extend the operation time of the filter, and to prevent shutdown when replacing the filter (Arnold & Stewart (1999) [1]).

2.1.2 Overhead condenser

To reduce the viscosity and enhance separation, the rich glycol should be heated before it is sent to the flash tank (Point 2 figure 2.1). This is often done by routing a portion (depending on the size of the regeneration unit) through the overhead condenser for the regeneration column. The overhead condenser can be either integrated in the column itself consisting of coils at the top of the column, or be a separate heat exchanger located beside the column (Kohl & Nielsen (1997) [1]). To reduce the thickness of the tubes in the overhead condenser, the pressure is usually lowered either through a nozzle or using a pressure driven pump prior to the rich glycol entering the overhead condenser (Guo & Ghalambor (2005) [1]).

2.1.3 Flash tank

Since the glycol is in intimate contact with the natural gas, some natural gas will become absorbed in the glycol together with the water. The amount of HC absorbed depends on absorber conditions (temperature and pressure), gas composition and contacting time between the glycol and the natural gas, and the type of glycol (Kohl & Nielsen (1997) [1]). High operating pressure and heavy hydrocarbons, especially aromatics, present in the natural gas results in larger quantities of natural gas absorbed by the glycol. Kohl & Nielsen (1997) [1] suggests using 1 scf/gal at 1000 psig and 100^{o}F (0.0075 Sm^{3}/l at 70 bar and 38^{o}C) to estimate the solubility of natural gas in TEG. Hydrocarbons are not allowed to enter the regeneration still as it may result in foaming, thereby reducing the efficiency of the stripping still (Guo & Ghalambor (2005) [1]).

If there are liquid hydrocarbons present in the rich TEG the flash tank should be a three-phase separator, otherwise a two-phase separator is used (as shown in figure 2.1. point 3). For a three-phase separator a retention time of 10 to 45 minutes is required, while 3 to 10 minutes is sufficient for a two-phase separator (Kohl & Nielsen (1997) [1]). Typical operating pressure for the flash tank is 3 to 7 bar (Carroll (2003), Arnold & Stewart (1999) [1]), depending on composition and temperature of the rich glycol. If the pressure is lowered too much glycol may vaporize, resulting in loss of expensive glycol.

2.1.4 Glycol/glycol heat exchanger

To recover some of the heat from the lean glycol exiting the stripping still, it is passed through a glycol/glycol heat exchanger (Point 4 figure 2.1), transferring heat from the lean to the rich glycol. This heat exchanger can be either a simple coil located in the surge drum for the lean glycol (Guo & Ghalambor (2005) [1]), or a dedicated shell and tube or plate/fin heat exchanger located just before the rich glycol inlet to the regeneration still (Kidney & Parrish (2006), Arnold & Stewart (1999) [1]).

2.1.5 Regeneration still

The regeneration still consists of a reboiler, a regeneration column, an overhead condenser and a surge tank. The reboiler heats the rich glycol, vaporizing absorbed water together with some glycol. Operation temperature of the reboiler is governed by the glycol used, e.g. for TEG and DEG the maximum recommended temperature is 204°C and 170°C at sea level pressure, respectively (Kidney & Parrish (2006)). Heat for the reboiler is commonly supplied by burning a portion of the processed gas, flash gas from the flash tank, or fuel gas available from other processes, directly in tubes inside the reboiler (Kohl & Nielsen (1997) [1]). Steam, waste heat and hot oil systems can also be used if available. Regardless of heat source, it is very important to keep the maximum temperature below the decomposition temperature of the glycol, or else large glycol losses due to thermal decomposition may occur.

The regeneration column is a regular distillation column with trays or packing (Figure 2.1. point 5). Hot rising vapor from the reboiler (Point 6 figure 2.1) rises through the column and contacts the falling cold rich glycol, vaporizing the water in the rich glycol and condensing the glycol in the rising vapor. At the top of the column, the vapor consists of nearly only water, and is passed through the reflux condenser and exits the column trough the top (Carroll (2003)).

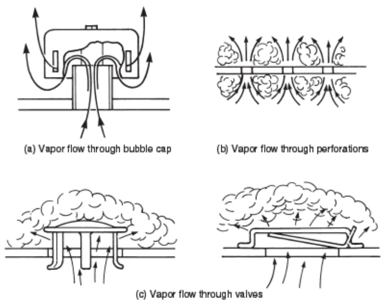

Figure 2.2: Gas flow through the different types of trays (adapted from GPSA (2004):p.19-9)

The mass transfer zone of the regeneration column can use trays or packing. Trays come in many different forms, the simplest being sieve trays which are just holes punched through a plate. A major disadvantage for sieve trays is that for gas flow rates below 70% of design, glycol will start to flow through the sieve holes onto the stage below (called "weeping"), which greatly reduces the efficiency of the tray (Kolmetz & Zygula (2000)). Slightly more efficient and complex are the valve trays, which employ valves over the holes, improving the turn down (the minimum required gas flow relative to design) to 60%. The most efficient, and also most expensive trays, are the bubble cap trays. These are designed to prevent weeping and have unlimited turn down (Arnold & Stewart (1999) [2]). Trays are preferred for large installations; clean, noncorrosive, nonfoaming liquids; and low to medium liquid flow applications (Kohl & Nielsen (1997) [2]) Figure 2.2 shows an illustration of the different types of trays.

Packing is divided into two types; structured- and random packing. Random packing was the first packing developed, just filling a portion of the column with small packing elements in a random order. Structured packing was originally designed for troublesome separation and smaller columns, but is available also for larger units today (Kohl & Nielsen (1997) [2]). Packing is normally used for small installations, corrosive service, foaming liquids, high liquid to gas ratio or applications requiring low pressure drop (e.g. vacuum operation) (Kohl & Nielsen (1997) [2]). Figure 2.3 and 2.4 shows different types of packing elements.

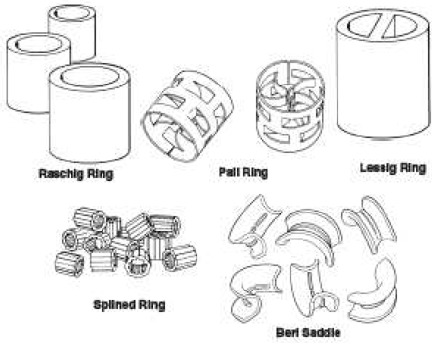

Figure 2.3: Examples of random packing (adapted from GPSA (2004):p.19-16)

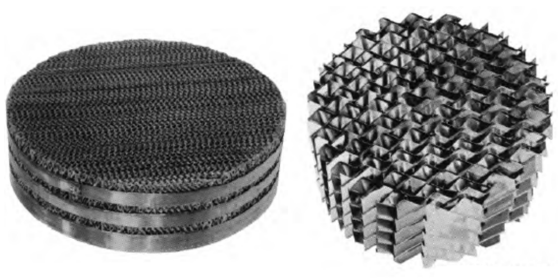

Figure 2.4: Examples of structured packing (adapted from GPSA (2004):p.19-17)

For relative small plants, the regeneration column is packed with ceramic saddles or stainless steel Pall rings with diameters from 1 to 1.5 inch (2.5 – 4 cm) and frequently mounted directly on top of the reboiler (Kohl & Nielsen (1997) [1]). Packing heights range from 4 ft (1.2 m) for small plants, up to 15 ft (4.6 m) for larger units. Columns with larger diameter than 24 inches (60 cm) commonly use bubble cap trays instead of packing, and are connected to the reboiler by piping (Kohl & Nielsen (1997) [1]).

Several reboiler designs are possible to use with a regeneration still, but the most commonly employed is the kettle type reboiler (GPSA (2004) [2]). As seen from figure 2.5, the kettle reboiler (sometimes also referred to as “flooded bundle reboiler”), has a design very similar to a horizontal separator. Natural gas is used as fuel for a burner, and the exhaust gas passed through a pipe bundle emerged in the liquid phase of the reboiler. Water and some glycol are vaporized at the tube surface, rise to the liquid surface, and flow out of the reboiler and into the regeneration column. Using a weir on the left side, the lean glycol is able to overflow into the surge tank, while the rich glycol is kept in the reboiler and heated. The advantages of employing a kettle type reboiler include easy liquid level control, and that the reboiler itself represents one theoretical equilibrium tray (Kohl & Nielsen (1997) [1]). Other types of reboilers include forced circulation, once through natural circulation, and vertical- and horizontal thermosyphon. See GPSA (2004) [2] for an introduction.

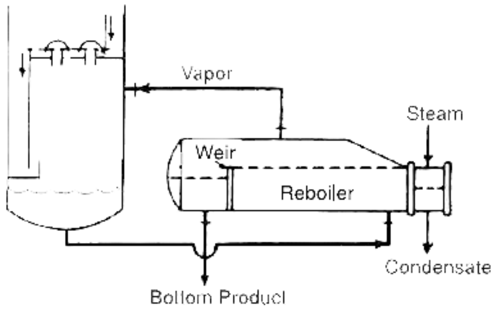

Figure 2.5: A kettle type reboiler employed in a regeneration still (adapted from GPSA (2004):p.19-24)

Because of the large difference in boiling points for glycol (e.g. 285°C for TEG) and water (100°C), the separation of water from the glycol requires very few theoretical equilibrium stages, usually between two and three, where the reboiler is the first (Kohl & Nielsen (1997) [1], Arnold & Stewart (1999) [1]). Depending on the tray efficiency used and the requirement for glycol losses, this results in 10 to 20 actual trays (Kohl & Nielsen (1997) [1]). The reason for using so many trays, is to reduce the loss of glycol through the exiting water vapor to almost zero. The pressure is kept as low as possible, but still at slight overpressure to prevent air leakage into the system.

Using the maximum recommended regeneration temperature for TEG, distillation at sea level results in a maximum theoretical purity for the lean TEG of 98.5 wt% (Kidnay & Parrish (2006)). In practice, purities as high as 99.1 wt% have however been obtained, because of stripping effect from dissolved HC or operation above sea level (Kohl & Nielsen (1997) [1]). If higher purity is required, processes for enhanced glycol regeneration must be employed (See chapter 2.3).

As mentioned in chapter 2.1.2, the overhead condenser can be either integrated in the regeneration column or be a standalone unit located directly downstream of the column. Several systems are available, one of which being the earlier described cooling coil. The simplest solution available, is however to install an uninsulated or finned section at the top of the column, condensing a portion of the vapor and thereby providing reflux. The drawback for this solution is that it is very difficult to control the rate of reflux under adverse weather conditions, and it is therefore only used for small units (Kohl & Nielsen (1997) [1]). Introducing pure water as reflux directly into the top of the distillation column can also be employed, but this requires accurate flow control and pure water to prevent introducing salts and other contaminants into the column. Typical reflux ratios for a regeneration column are 25 to 50% of the rising water vapor at the top of the column (Arnold & Stewart (1999) [1]).

The surge tank is usually located below the reboiler (Point 7 figure 2.1), and stores the lean glycol before it is sent back to the absorber. As mentioned in chapter 2.1.4, the surge tank can also act as the glycol/glycol heat exchanger, transferring heat from the hot lean glycol in the tank to the cold rich glycol flowing through pipes. The surge tank must be large enough to allow for thermal expansion of the lean TEG, in addition to hold sufficient capacity to allow for reasonable time between additions of glycol (Arnold & Stewart (1999) [1]).

2.1.6 Glycol pump

In some glycol regeneration units, a glycol powered pump is employed (See figure 2.6). The high pressure of the rich glycol is used to pressurize the low pressure lean glycol, conserving the pressure energy in the rich glycol which otherwise would have been lost through a nozzle. This solution is however seldom used in larger units or units at remote locations (e.g. offshore), because of the added complexity over a regular pump, reducing the reliability. The alternative is using a regular electric powered pump (Point 8 figure 2.1).

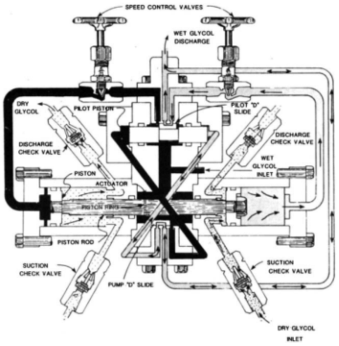

Figure 2.6: Schematics of a glycol driven pump (adapted from Arnold & Stewart (1999):p.221)

2.2 Operational problems

Operating a glycol regeneration unit incur several possible operational problems. The problems described for separators in chapter 1.4 are also applicable for the glycol regeneration process, as the flash tank and the stripping column are both special types of separators. These problems will not be repeated here. In addition problems with corrosion, contamination of the glycol, emission of hazardous compounds and loss of glycol have been reported (Kohl & Nielsen (1997) [1]).

2.2.1 Corrosion

As the glycol is essentially non-corrosive, corrosion caused by the glycol solution is believed to be the result of other components present formed by oxidation or thermal decomposition of the glycol, or by components entering the system from the gas stream (Kohl & Nielsen (1997) [1]). Other factors influencing the extent of the corrosion are the temperature of the glycol and fluid velocity. Corrosion is illustrated in figure 2.7.

Thermal decomposition of the glycol is the result of local temperatures exceeding the thermal decomposition temperature of the glycol. This can happen if the surface of the fire tubes in the reboiler becomes unevenly coated with depositions, resulting in locally low thermal resistances, creating local hotspots. This can be prevented by regular cleaning of the fire tubes` outer surfaces.

Figure 2.7: Example of the result of severe corrosion on a pipe inlet (adapted from TTI Environmental (undated))

Oxidation of the glycol requires that oxygen is present. In some cases the feed gas to be dehydrated contains oxygen, and oxygen can then enter the regeneration system with the rich glycol. The most common source of oxygen is however leakage through the surge tank (Kohl & Nielsen (1997) [1]). Lyons (1957) as cited by Kohl & Nielsen (1997) [1] found that oxidation of DEG resulted in the formation of an organic peroxide as an intermediate product, and formic acid and formaldehyde in considerable quantities. The oxidation rate was found to increase with increased temperature and partial pressure of oxygen, and accelerated by the presence of acid. To prevent oxygen from entering the regeneration system, a gas blanket is often applied in the surge tank (Kohl & Nielsen (1997) [1]).

A third possible source for corrosion is by components entering the system with the rich glycol. CO_2 and H_2S are known to be highly corrosive in combination with free water, and are also soluble in glycol. Dehydration of feed gases with high concentrations of sour gases may therefore require sour gas removal prior to dehydration.

To prevent corrosion in the regeneration system several measures have been mentioned already, but in addition corrosion-resistant alloys can be used for manufacture, corrosion inhibitors added to the glycol and the process designed with minimized temperatures and velocities (Kohl & Nielsen (1997) [1]).

2.2.2 Contamination of the glycol

Contamination of the glycol (See figure 2.8) is a very common problem in operation of glycol regeneration units, and can have many different sources. Corrosion products, together with solid particles brought in by the rich glycol, can lead to fouling of heat exchanger surfaces by deposition and ultimately plugging of heat exchanger pipes. Solids entering the stripping column can plug weep holes in the distillation trays, and also accelerate corrosion (and possibly also erosion) (Kohl & Nielsen (1997) [1]). Glycols also absorb salts at low temperatures, which again are desorbed at high temperatures, leading to extensive salt depositions in heat exchangers and on reboiler fire tubes if not handled.

Figure 2.8: The resulting glycol condition if filtering is not employed can be seen to the left, versus the clean to the right (adapted from Hydrotechnik UK (2008))

As earlier mentioned, sock filters should be employed in the rich glycol line to remove solids (See chapter 2.1.1). Removal of solids larger than 5 microns is regarded as sufficient (Kohl & Nielsen (1997) [1]). If salt is a major problem, a vacuum distillation or ion exchange can be employed to directly remove the salt. For most systems however, it is sufficient to remove deposited salt during scheduled shutdowns (Kohl & Nielsen (1997) [1]).

Absorbed HC are partly flashed off in the flash tank, but some of the heavy HC will remain in the glycol. HC can, as previously noted, lead to foaming in the regeneration still, as well as increased emissions of HC. To prevent heavy HC from accumulating in the system, a portion of the rich glycol is therefore routed through a charcoal filter prior to entering the regeneration still.

2.2.3 Emission of hazardous compounds

Even thought the rich glycol is both flashed and filtrated, some HC will inevitably enter the regeneration still with the glycol. Especially aromatics such as benzene, toluene, ethyl benzene and xylene isomers (BTEX) are absorbed well in glycol, and a significant fraction is carried through the flash tank and into the regeneration still (Kohl & Nielsen (1997) [1]). The amount of BTEX absorbed increases with increasing glycol circulation rate, increasing operating pressure and temperature of the absorber, and increasing concentration in the feed gas (Nassar, Bullin & Lyddon (2000)). If not handled, the BTEX will vaporize in the regeneration still and leave through the top of the column with the water vapor. As BTEX is considered hazardous for the environment, rules and regulations restrict yearly emissions, for the USA maximum 25 tons/year altogether and maximum 10 tons/year of any single compound (Nassar, Bullin & Lyddon (2000)).

Measures for handling BTEX include minimizing the glycol circulation rate, maximizing the amount flashed off in the flash tank, using a glycol with low solubility for aromatics (e.g. DEG/MEG instead of TEG), or processing the overhead vapor from the regeneration still (Kohl & Nielsen (1997) [1]). Processing involves either incineration of the gas (e.g. as fuel gas for the reboiler) or partial condensation with HC recovery. The Drizo process (See chapter 2.3.1) utilizes the condensed liquid HC from the overhead vapor, superheats it, and re-introduces it into the regeneration column as stripping gas.

In addition to BTEX other emissions also occur, including volatile organic components (VOCs), glycol, and breakdown products from the glycol. Depending on the amount, these may also need to be handled, either by distillation or by incineration (Kohl & Nielsen (1997) [1]).

2.2.4 Glycol loss

Sources for glycol loss include glycol carry-over from the absorber, vaporization of glycol in the reboiler, vaporization into the dehydrated gas, and mechanical loss from the system (e.g. leakage through seals, pumps, spillage etc.) (Kohl & Nielsen (1997) [1]). The greatest contribution to glycol loss is however entrainment and resulting carry-over from the absorber (Kidnay & Parrish (2006), Kohl & Nielsen (1997) [1]), and the most effort is therefore made to design the best possible mist extractor for each unit. Typical acceptable glycol losses are in the range from 0.5 – 1.0 lb/MMscf (8 – 16 kg/MMscm), but for some systems even larger losses can be acceptable (e.g. dehydration of sour gases) (Kohl & Nielsen (1997) [1], Kidnay & Parrish (2006)).

2.3 Enhanced glycol regeneration

As mentioned earlier, additional processing is required to regenerate the glycol to purities higher than approximately 98.5 wt%. Several processes are available today, each employing different strategies to enhance glycol regeneration by reducing the effective partial pressure of water in the vapor space (GPSA (2004) [3]). This chapter will give a short introduction to the Drizo, the Coldfinger, vacuum distillation and the stripping gas processes.

2.3.1 The Drizo process

The Drizo process (See appendix E for PFD) utilizes superheated HCs as stripping gas in a column placed after the regeneration still to lower the water vapor pressure. After the stripping column the stripping gas is cooled, and the HCs are separated from the water and off gas in a three-phase separator. The gas is vented to the atmosphere and the water discarded, while the HCs are recovered and recycled to be used for stripping gas again. As the glycol absorbs some HCs in the absorption column, the Drizo process can even produce liquid HC products, enhancing the overall process efficiency (Kohl & Nielsen (1997) [1]).

Using the Drizo process for glycol regeneration has been shown to result in glycol purities higher than 99.99 wt% (GPSA (2004) [3]), which allows for dew point depressions as high as 100^{o}C (Prosernat (undated)). To obtain even higher glycol purities, the superheated HC stripping gas can be dehydrated using a solid desiccant in an adsorption bed before it is injected into the stripping column. This has proven to produce glycol purities as high as 99.999 wt%, allowing dew point depressions of up to 120^{o}C (GPSA (2004) [3]). An added advantage of the Drizo process is that the emission of BTEX to the atmosphere is almost eliminated.

The drawbacks for the Drizo process, is that it requires additional equipment both for the stripping column and the regeneration of the liquid HC, and heat is required to vaporize and superheat the HCs used for stripping.

According to the manufacturer (Prosernat (undated)), the Drizo process has been successfully installed in more than 60 glycol regeneration units worldwide. As there are over 20 000 installed TEG dehydration units in USA alone (Kidnay & Parrish (2006)), this is surprisingly few, indicating that the process is not as advantageous as the manufacturer boasts.

2.3.2 The Coldfinger process

Instead of using gas to reduce the vapor pressure of water, the Coldfinger process (See appendix E for PFD) condenses and extracts water from the vapor phase using a cold tube bundle ("Coldfinger"). The lean glycol is routed from the regeneration still to a separate vessel which contains the Coldfinger process. The vessel is half filled with lean glycol, while the other half is occupied by vapor consisting of water and glycol. A cold bundle is inserted in the vapor space, condensing water from the vapor, which again is extracted from the vessel by the use of troughs. As all systems naturally seek equilibrium conditions, water will evaporate from the liquid glycol to restore equilibrium in the vapor phase, thereby concentrating the glycol. The glycol exiting the vessel therefore has a higher purity of glycol than the glycol entering, reaching as high as 99.9 wt% (Comart undated). In the most common applications, rich glycol from the absorber column is used as coolant for the Coldfinger bundle (GPSA (2004) [3]).

As for the Drizo process, the Coldfinger process requires extra equipment added to the regeneration process. In addition to the Coldfinger vessel itself, equipment for processing the extracted water is also needed.

The manufacturer (Comart (undated)) claims that 61 000 MMscfd is processed using the Coldfinger process for enhanced glycol regeneration. One of the largest gas processing facilities in Norway, Ormen Lange, exports alone approximately 2 500 MMscfd (StatoilHydro undated), indicating that use of the Coldfinger process also is limited.

2.3.3 Vacuum distillation

Partial pressure is the product of the total pressure and the molar fraction of the matter present in the vapor phase (Moran & Shapiro (1998):p.538). The two previously described processes reduce the partial pressure by reducing the molar fraction, while vacuum distillation, as the name implies, reduces the total pressure. Operating at vacuum conditions usually utilize the same equipment as for atmospheric operation, but with the added need to totally condense the overhead vapor as well as provide a vacuum pump for the non-condensable gases (Kohl & Nielsen (1997) [1]).

As the ambient pressure is greater than the process pressure, air will leak into the process if cracks or fractures develop. This would lead to drastic reductions in the process performance, resulting in poor glycol regeneration, and subsequently insufficient dehydration of the natural gas. Exposing the hot glycol solution to oxygen also leads to oxidation of glycol, resulting in large glycol losses.

To reduce the probability of air leaks into the process and limit the amount of vacuum vapor to be handled, the BS&B Engineering Company has developed a process where the first stage in the regeneration process is identical to the original process, operating at near atmospheric pressure.

After the regeneration column the lean glycol is flashed into a vacuum column, resulting in glycol purities reaching 99.9 wt% (Kohl & Nielsen (1997) [1]).

2.3.4 Stripping gas

Stripping gas is the simplest and most employed process for enhanced glycol regeneration used in the natural gas industry today (Kohl & Nielsen (1997) [1]). A non-condensable gas at regeneration conditions is introduced into the regeneration column, lowering the molar fraction of water in the vapor phase, and thereby also lowering the partial pressure of water. Often a small portion of the wet gas is used, because at regeneration temperature it will still be able to absorb large amounts of water (Arnold & Stewart (1999) [1]). The amount of natural gas needed to heat the glycol solution to increase the lean glycol concentration is in most cases lower than if using stripping gas directly. Stripping gas is therefore mainly used to increase the purity above approx 98.5 wt% (Arnold & Stewart (1999) [1]).

Injecting the stripping gas directly into the reboiler has proven to obtain glycol purities between 99.1 and 99.6 wt% at the maximum regeneration temperature and near ambient pressure (Kohl & Nielsen (1997) [1]). To better utilize the stripping gas, a dedicated stripping column can be installed after the regeneration column, counter-currently contacting the stripping gas and lean glycol. Depending on the number of trays in the stripping column and the flow rate of stripping gas, this can produce purities reaching 99.99 wt% (Kohl & Nielsen (1997) [1]).

The greatest advantage when using stripping gas is that it does require very little additional equipment (especially if the stripping gas is injected into the reboiler). This makes stripping gas a very widespread method for extending the operating capacity of existing units (Arnold & Stewart (1999) [1]).

Chapter 3: Compact mixing unit

A fundamental part of the alternative dehydration concept, is the injection and mixing of TEG with the feed gas. To minimize the circulation of TEG and enhance the overall unit efficiency, it is very important that the mixer creates an as near equilibrium mix as possible, at the same time conserving most of the pressure. There are several commercial mixing concepts available, but this chapter will concentrate on the mixing unit from ProPure, as it is one of the newest and most promising compact mixing units for use in the natural gas industry available today.

3.1 The ProPure compact mixer

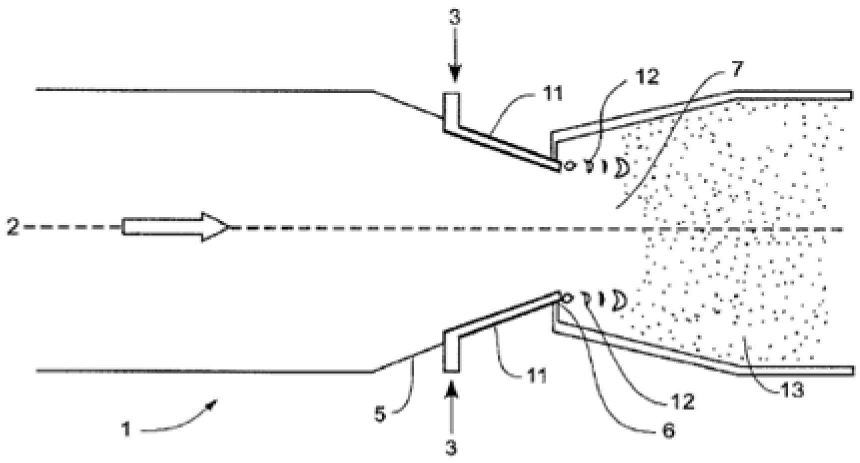

Figure 3.1: Schematics for the ProPure compact mixer (adapted from Nilsen et al. (2006):p.4)

The ProPure compact mixer is an inline unit utilizing turbulence to mix a liquid with a gas. The gas enters from the left (See figure 3.1), and the pipe cross section area is reduced to increase the velocity of the gas. The liquid to be mixed with the gas enters through an annulus, and is evenly distributed around the perimeter. Before the smallest cross section of the pipe (point 3) the liquid is exposed to the gas, causing the gas to drag the liquid with it and form a liquid film on the pipe walls (point 11). At the smallest cross section area (point 6), the diameter is suddenly increased, creating a sharp edge. This causes the liquid to leave the pipe walls and form drops in the gas. The large turbulence created by the sudden cross section expansion and the high velocity aids in breaking up the drops to small droplets, resulting in very effective contact between the liquid and the gas (Nilsen et al. (2006)). To reduce the permanent pressure drop over the unit, the gas is then expanded through a diffuser to the original pipe diameter.

The physics utilized in the mixing unit is surface forces, shear forces and drag forces. The liquid to be mixed with the gas is pumped through a small channel ending in an annulus, and is pushed into the gas stream. As the liquid is pushed into the stream adhered to the wall, the surface force between the liquid and the wall will keep the liquid adhered to the wall, but the shear forces transferred from the fast moving gas will pull the liquid forward, creating a moving film. When the liquid film reaches the sharp edge, the surface forces are reduced enough for the gas to launch the liquid into the gas stream creating liquid filaments. Once in the gas stream the filaments are exposed to drag forces from the faster moving gas, shattering them into smaller droplets. This process is governed by the Weber number (We) defined from formula 3.1 (Nilsen et al. (2006):p.13):

$$We \equiv \frac{\rho_G \cdot U_{rel}^{\ 2} \cdot d_{fil}}{\sigma_{surf}}$$ (Formula 3.1: Weber number)

Where ρ_G is the gas density, U_{rel} the relative velocity between the gas and the liquid, d_{fil} the characteristic filament dimension and σ_{surf} the liquid-gas surface tension.

Figure 3.2: Drop shattering as $We>We_{crit}$ (adapted from ProPure (undated):p.2)

When the Weber number reaches a critical value, We_{crit}, the surface tension is no longer strong enough to hold the filament together, causing it to shatter into smaller droplets (See figure 3.2). From wind tunnel experiments it has been found that We_{crit} has a value between 8 and 10 (Nilsen et al. (2006)). After the filaments have shattered, further mixing is provided by turbulence, governed by the Reynolds number (Nilsen et al. (2006):p.13).

$$\mathrm{Re} \equiv \frac{\rho_{mix} \cdot U_{mix} \cdot D_i}{\mu_{mix}}$$ (Formula 3.2: Reynolds number)

It can be seen from formula 3.1 that the greater the difference in velocity between the gas and the liquid (U_{rel}), the better the mixing, as smaller droplets (smaller values for d_{fil}) will shatter because of reaching We_{crit}. High gas velocities will also push the filaments forward, inertia at the same time causing the filaments to spread out perpendicular to the direction of flow, increasing the d_{fil} and thereby also We (Illustrated in figure 3.2). As velocity is inversely proportional to the square of the pipe diameter, Re (and consequently also the turbulence) will increase when narrowing the cross section area. Accelerating the gas to greater velocities by narrowing the available cross section area will however increase the permanent pressure drop across the unit.

Figure 3.3: Assembled ProPure compact mixer (adapted from ProPure (undated):p.1)

The design of the mixing unit is very strongly dependent on the operating conditions and the fluids involved, and must be reworked for each new project. Several parameters are adjustable including the smallest cross section, the rate of narrowing, the angle of the sharp edge, and the rate of expansion (Nilsen et al. (2006)). Depending on the design criteria, the unit can thereby be optimized for best possible mixing, lowest possible pressure drop, or most compact design, among others. The unit is installed into the same type of pipe used at the operating location, this to ensure compatibility and easily fulfill pressure/temperature requirements (See figure 3.3). Placement and inclination of the unit is arbitrary, but preferentially it is mounted horizontally with 15 to 20 pipe diameters of straight pipe downstream to provide further mixing (Nilsen et al. (2006)).

Although the ProPure compact mixer can be designed to mix any fluid with gas, it is presently marketed for injection of H_2S scavenging chemicals in natural gas. Internationally this concept has been applied to more than 30 applications with up to 30% reduction in scavenger consumption compared to conventional systems (ProPure (2005)).

Advantages using the ProPure compact mixer include (ProPure (2005)):

- Large turndown ratio for both gas and liquid.
- Better mixing performance than conventional mixers.
- Robust towards plugging.
- Great mixing performance also for high viscosity liquids.
- Compact.

Disadvantages using the ProPure compact mixer include:

- Permanent pressure drop (approx 0.3 bar (Kalgraff (2008))).

Chapter 4: Further developing the alternative natural gas drying concept

As earlier mentioned, my semester project work during fall 2007 resulted in a proposition for an alternative concept for dehydrating natural gas. This chapter will give a short introduction to this work and try to further develop the concept, countering some of the major disadvantages when compared to conventional dehydration technology. Alternatives for improvements will be presented, before the different alternatives are discussed and one is eventually chosen for simulation.

4.1 The concept

The proposed dehydration unit consists of multiple stages of injection and extraction of TEG directly into/from the feed gas (See figure 4.1). Following inlet separation, a compact mixer unit injects lean TEG into the feed stream, and the piping is utilized as a contactor. After sufficient contacting time, the rich TEG is separated from the gas stream by using a scrubber, while the dry gas leaves for further processing.

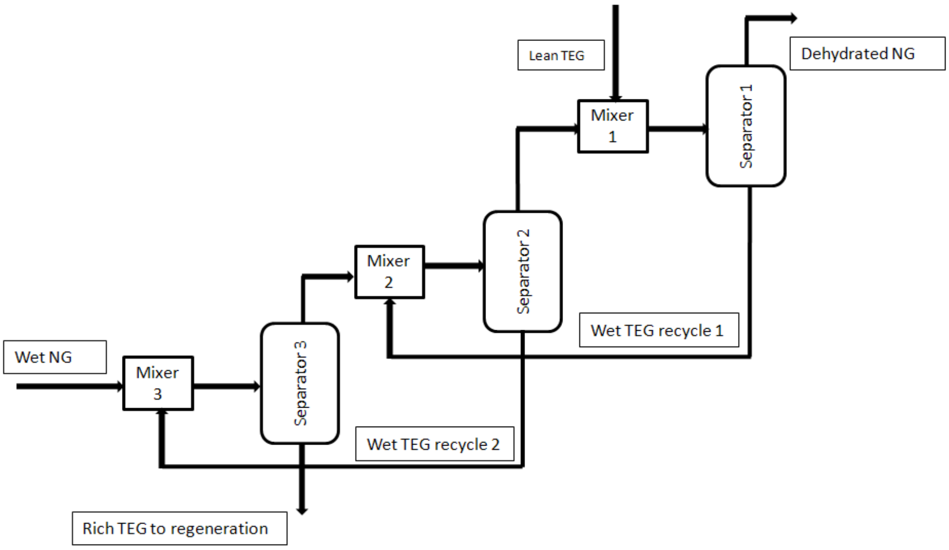

Figure 4.1: The TEG injection dehydration concept

As opposed to counter-current contacting in a conventional absorber column where equilibrium with the lean TEG is obtained, co-current contacting in the gas piping can ultimately only result in the gas obtaining equilibrium with the rich TEG (Kohl & Nielsen (1997) [1]). More than one injection/extraction stage is therefore needed to restrict the TEG flow rate. From simulations it was found that going from one to three stages, the TEG flow rate could be reduced by almost 85% for normal operating conditions (Bråthen (2007)). The leanest TEG is injected in front of the last extraction stage, and the liquid extracted in the last extraction stage injected before the preceding stage and so forth (See figure 4.1). Adding a fourth injection/extraction stage would further reduce the TEG flow rate, but at the expense of adding more weight and volume to the unit. As the reduction was much smaller than going from two to three stages, it was concluded that three stages was the most promising concept.

While the TEG flow rate was found to decrease substantially going from one to three separation stages, the required flow rate required to obtain a given dryness was still roughly four times that of a conventional absorber column (Bråthen (2007)). When comparing weight and volume, the advantages when comparing the absorption units was therefore massively outweighed by the disadvantages when comparing the regeneration units, making TEG injection uncompetitive compared to conventional dehydration. It must however be noted that the absorption column used in the comparison in my project thesis consisted of 10 *theoretical* trays, while it is common to use 6 to 12 *actual* trays (Arnold & Stewart (1999) [1]). 6 to 12 actual trays results in 2 to 8 theoretical trays, depending on the tray efficiency (See chapter 2.1.5). The results for the contactor tower is therefore better than what is found for a typical absorption column in operation, making the comparisons to the TEG injection concept better than they actually are.

Another possible problem with the concept, is the intimate contact between TEG and liquid HC. Heavy hydrocarbons are absorbed by the TEG, and can cause both foaming in the regeneration unit and lead to excessive emission of hazardous components with the water vapor exiting the stripping column. This problem is however also applicable to many conventional absorption units, but still needs to be handled (See chapter 2.2.3).

4.2 Potential improvements

As mentioned above, the TEG circulation rate is the main disadvantage for the TEG injection concept, and focus should therefore be to reduce this even further. One possible solution is to use more injection/extraction stages utilizing inline separators and other compact equipment, but this will at the same time increase both the complexity of the process and also the weight and volume. The separators could be placed on top of each other, minimizing the required footprint, but their weight and volume would still be the same. More stages also require more utility equipment like pumps, gauges, piping etc.

Another possible solution is to use an alternative process for the regeneration of the rich TEG as well. Conventional regeneration technology (See chapter 2) relies on heating the entire glycol/water solution to boil the absorbed water from the glycol. This means that there is a near linear dependence between the flow rate of rich glycol and the energy needed for regeneration (See formula C.6 in appendix C) (The dependence is not directly proportional to the glycol circulation rate, as vaporizing the water requires heat as well). By using alternative processes where energy consumption instead is dependent on the amount of water absorbed, huge energy savings could be achieved.

Intimate contact between HC and TEG is difficult to avoid, as this is a fundamental part of the concept. Using an alternative process for regeneration could however diminish the problem, as foaming in the regeneration column is then mitigated. Anti-foaming chemicals can also be applied, but then at increased operating costs. Other glycols like MEG and DEG with lower solubility for HC can also be used, but then at the expense of increased circulation rate.

Using amines instead of glycols for simultaneous sweetening and dehydration is also possible, and this would improve at least the weight and volume comparisons for the concept. However, to be able to dehydrate the gas, the amine concentration has to be above 95%, requiring very high reboiler temperatures at which severe corrosion will occur (Kohl & Nielsen (1997) [3]). To overcome the corrosion, corrosion-resistant materials must be used, greatly increasing the production costs. Mixtures of glycol and amines have also been tested, but rejected among other factors because of large vaporization losses of amine (Kohl & Nielsen (1997) [3]).

4.3 Alternative concepts

Because most of the improvements pointed out above involve the regeneration process for the glycol, most of the attention in this chapter is devoted to suggesting improvements/alternative processes for the regeneration process.

4.3.1 Alternative 1: Injecting hot stripping gas into the rich glycol stream

Idea: Instead of heating the entire rich glycol solution to boil out the water, hot stripping gas is injected to absorb the water from the glycol.

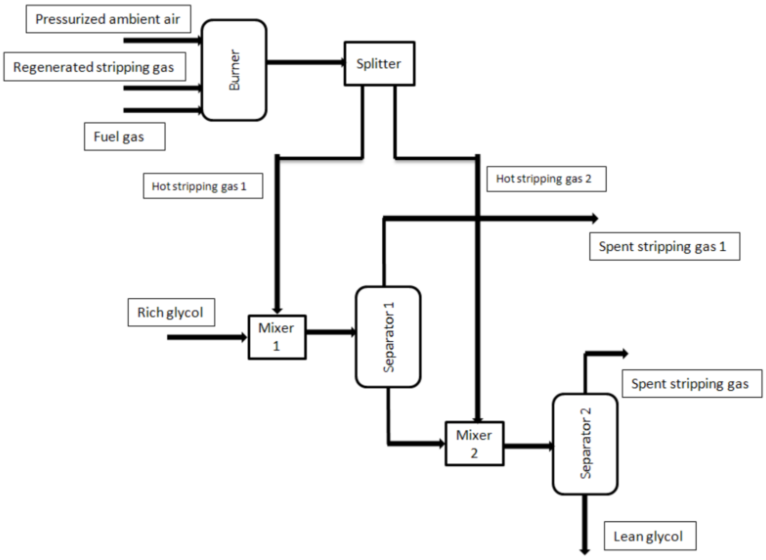

Figure 4.2: Alternative regeneration process using hot stripping gas

Hot stripping gas is provided by burning a portion of the dehydrated gas together with flash gas absorbed by the glycol. Combustion air is supplied from the surroundings, together with recycled stripping gas, and combusted in a gas burner. The stripping gas is then injected into the glycol stream, and separated out again using a separator. To reduce glycol loss, multiple stages are required, but to minimize weight and volume of the unit, two stages is proposed. The off gas from the separators must also be treated to recover some of the entrained glycol, as well as separating the BTEX from the water.

Advantages:

- Energy consumption is much lower than for a conventional regeneration process, as the stripping effect is more important than the heating effect

- Compact; can use separators/scrubbers instead of a distillation column
- Shorter retention time for the glycol in the regeneration process
- Very good heat transmission, since the hot and the cold fluid is mixed
- Less thermal strain on the glycol solution, as lower temperature is sufficient for regeneration

Disadvantages:

- Large glycol entrainment needs to be handled
- Contamination of the glycol; more filtering is needed
- Compressor is needed for pressurizing combustion air; reduces reliability because of introducing additional rotating equipment
- Larger glycol loss than the conventional regeneration process
- Possible degradation of the glycol caused by presence of oxygen

4.3.2 Alternative 2: Using staged injection of hot stripping gas

Idea: Injecting the leanest and hottest stripping gas in front of the last injection stage, recycle the off gas from the last separator and inject it in the preceding stage.

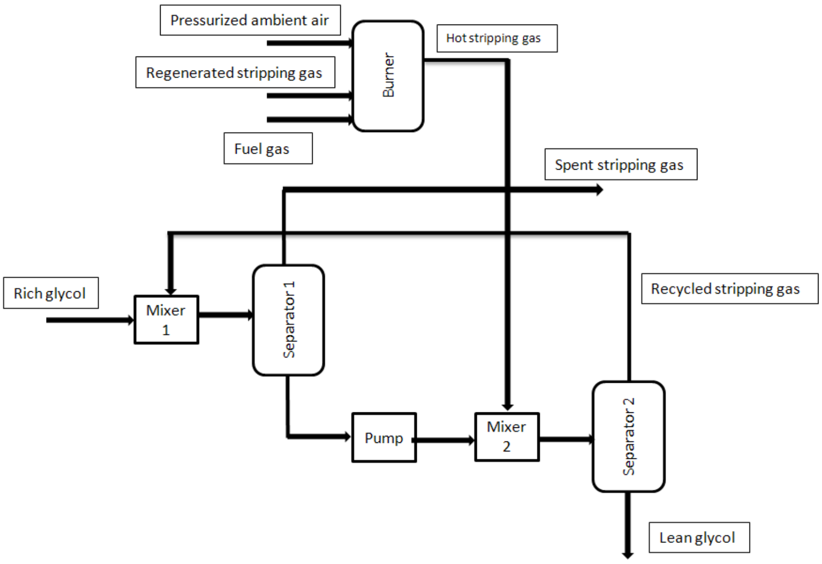

Figure 4.3: Alternative regeneration process using staged injection of hot stripping gas

The hot stripping gas is obtained using the same method as for alternative 1, but instead of dividing the stream into multiple injection stages, all of the stripping gas is injected before the last stage. From the last separator, the off gas is then routed to the preceding stage where it is reused as stripping gas. To avoid condensation of the stripping gas and eliminate the need for recompression, the last stage must operate at a higher pressure than the preceding stage. The liquid entering the last stage must therefore be pressurized using a pump. The off gas from the first stage needs to be processed to recover some of the entrained glycol, as well as separating the volatile HCs from the water.

Advantages:

- Energy consumption is much lower than for a conventional regeneration unit

- Compact; can use separators/scrubbers instead of a distillation column
- Shorter retention time for the glycol in the regeneration system
- Less thermal strain on the glycol
- Very good heat transmission, since the hot and the cold fluid is mixed
- Better fit to equilibrium than alternative 1

Disadvantages:

- Requires pressurizing of the glycol between injection/extraction stages
- Glycol entrainment must be handled
- More filtering is needed because of contamination of the glycol
- Compressor is needed for combustion air
- Possible degradation of the glycol caused by presence of oxygen
- Larger glycol loss than conventional regeneration process

4.3.3 Alternative 3: Using an absorber for regeneration

Idea: Using stripping gas to regenerate the rich glycol in an absorber.

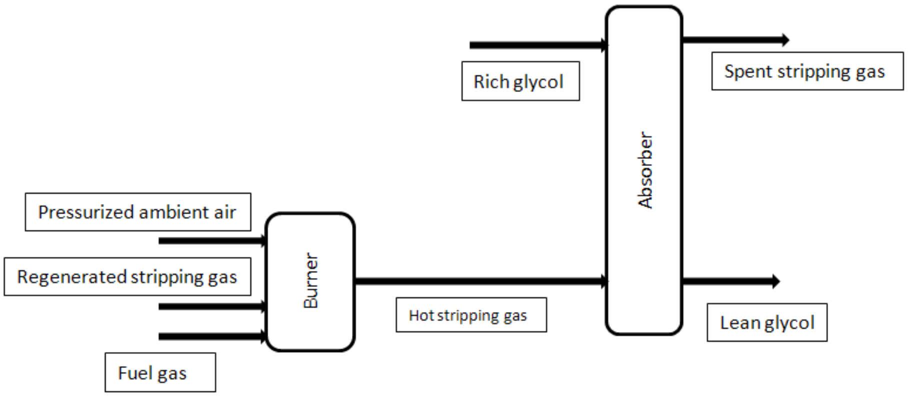

Figure 4.4: Alternative regeneration process using an absorber for regeneration

The rich glycol enters from the top of the column flowing downwards, and lean stripping gas enters from the bottom flowing upwards. As the gas flows up the column it will absorb water from the glycol, and exit from the top saturated with water (ideally). Lean glycol exits from the bottom. Possible stripping gases are atmospheric air, nitrogen or hot combustion gases.

Advantages:

- Less energy consumption than conventional regeneration
- Can re-use known technology
- Less thermal stress on the glycol solution
- Low gas flow rate results in a very compact design of the absorber
- Lower glycol loss than previously described alternatives

Disadvantages:

- Possible problem with oxidation of the glycol if using atmospheric air as stripping gas

- Requires equipment for regenerating stripping gas if other than atmospheric air is used
- Expensive

4.3.4 Alternative 4: Using piping as a heat exchanger for the glycol regeneration unit

Idea: Installing a pipe inside another bigger pipe, and exchanging heat from the hot internal lean glycol to the outer rich cold glycol.

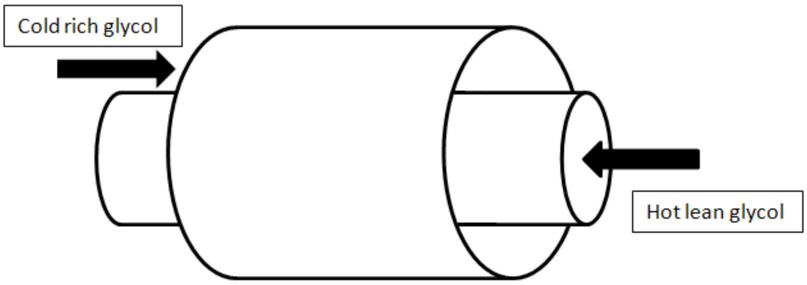

Figure 4.5: Using piping as heat exchanger

A pipe is installed in the center of another pipe and support stands welded between the pipe walls at suitable intervals. The lean hot glycol is passed through the inner pipe, while the rich cold glycol is passed through the outer pipe. As the pressure in the inner and outer pipes is almost the same, the pipe walls can be very thin, enhancing heat transfer and reducing costs.

Advantages:

- Saves weight as the inner pipe walls can be thinner
- Saves floor space as no dedicated heat exchanger is needed
- Easy to manufacture

Disadvantages:

- Increased pressure drop in the fluid flowing in the outer pipe
- May provide insufficient heat transfer, e.g. very long piping is needed
- Difficult to locate leaks in the inner pipe

4.4 Discussion

The concepts involving injection of hot stripping gas for regeneration (Alternative 1 and 2) may initially sound promising, but after some afterthought and calculations they are discarded. One problem discovered by doing coarse simulations, is that very large quantities of stripping gas is needed to regenerate the glycol to even moderate concentrations. E.g. for regeneration of 65 kgmole/h rich TEG to 98 wt% stripping gas flow rates as high as 6500 m^3/h @ 200^oC was found to be needed. This implies that large diameter pipes must be used, in addition to very large TEG losses caused by entrainment in the spent stripping gas. The idea was also that the hot stripping gas should be provided by burning a portion of the dehydrated natural gas, and then mixing the exhaust gases with pressurized atmospheric air to obtain the desired temperature. This would however introduce oxygen into the regeneration system, resulting in oxidation of the glycol, adding further to the glycol losses. Oxidation of glycol would also not be modeled by HYSYS, since it only uses equilibrium calculations for the different units and does not include chemical reactions.

Another drawback with the hot stripping gas concepts is that the equipment for cooling the lean glycol is not incorporated. This does not only result in poor energy efficiency for the unit, but also requires a separate cooling circuit utilizing sea water or air. An alternative is of course to use a heat regenerator as for conventional glycol regeneration, and use stripping gas only for the last heating stage. This would however result in a more complex process than the conventional concept, requiring more equipment and having much larger glycol losses.

The only advantages of using the hot stripping gas concept is very good heat transfer, as the hot and cold fluids are intimately mixed, and that the bulk temperature of the glycol can be lower during regeneration . These advantages are however massively outweighed by the larger glycol losses, faster glycol degradation and huge flow rates of stripping gas.

Using a contactor for regenerating glycol (Alternative 3) is better than staged injection, but still has the problem of obtaining the hot stripping gas. As mentioned above, the mixing of air with the exhaust from burning natural gas introduces oxygen into the system, potentially leading to large glycol loses caused by glycol oxidation. Other stripping gases could have been used, but the stripping gas itself then also needs to be regenerated. This would introduce another regeneration process, both worsening the thermodynamic efficiency of the overall process and requiring additional equipment and energy. If exhaust gas from e.g. a gas turbine was available at the processing location, this could still not be used for stripping gas as most gas turbines operate under lean conditions. That is, there is excess of oxygen compared to the amount of fuel, and consequently there will still be free oxygen present in the exhaust gas. Air is also used for temperature control and to prevent hot spots (Turns (2000)).

Alternative 4 (using piping as heat exchanger) could probably be incorporated in the concept with some success, but its use is still limited. Placing a pipe in the center of another pipe implies that the outer pipe must have a diameter 40% greater than that of the inner pipe to provide the same area for flow. In addition, the flow rate of the rich glycol is greater than the flow rate of the lean glycol (as the water has not been extracted yet), thus requiring an even greater diameter of the pipe. As the perimeter increases as Pi times the diameter difference, the weight saved by using a thin wall for the inner pipe may easily be compensated by the need for a greater diameter of the outer pipe. In addition the pressure in the regeneration process is commonly near atmospheric, further reducing the potential weight savings. Also, to completely replace the heat exchangers a very long pipe is needed, probably requiring more volume and weight than the heat exchangers. A better solution for exchanging heat between the rich and the lean glycol would be to place the pipes beside each other and wrap them in insulation (in addition to using a heat exchanger).

The conclusion of this discussion is therefore that the conventional way of regenerating glycol is difficult to improve upon when considering weight-, volume- and energy requirements. Some minor improvements, like using the piping for heat exchange might be possible to incorporate, but the advantages are too small to make any real difference. None of the proposed alternative concepts for regeneration will therefore be simulated, as they are not considered worth pursuing.

Focus for the rest of this project will therefore be to improve upon the simulation model from my semester project (Bråthen (2007)) by incorporating pressure losses for the different units, carry-over for the different separators and modeling of a conventional glycol regeneration process, as well as utilizing the CPA equation of state (See appendix F) instead of the default glycol fluid package.

Chapter 5: Developing the simulation model

This chapter will describe the modeling of the alternative dehydration concept in the engineering simulation software HYSYS (See appendix G for an introduction), stating the different assumptions made and highlighting the chosen options for the different units. A model for conventional glycol dehydration is also developed. The process flow diagrams (PFDs) are presented in appendix H.

The input data used for the simulations are from the Snøhvit LNG processing facility and are presented in appendix I. As earlier mentioned, a relatively new equation of state (EoS) developed by the Technical University of Denmark (DTU), CPA, is used as fluid package (See appendix F for a description).

General assumptions

- *The entire process is assumed to be adiabatic.*
  Absorption: The flow rate of gas is much greater than the flow rate of TEG, and the gas temperature will therefore govern the process temperature. As the retention time for the gas in the process is very short, the surrounding temperature will have to differ greatly from the gas temperature to influence the process temperature. This assumption is however not so good for the recycling streams of wet TEG, as they are much more vulnerable to the surrounding temperature because of lower flow rates and smaller diameter pipes (e.g. greater perimeter compared to cross section area).
  Regeneration: The temperature difference between the glycol in the regeneration system and the surroundings can often be quite large, indicating that heat transfer will occur. This will result in increased heat load for the reboiler, and ultimately increased fuel gas consumption. As the heat capacity of the glycol is large and the retention time in the piping is small, this contribution is however small, and is assumed to be neglectable.
  Pumps: A pump is a very non-adiabatic unit, but since the flow rate through the pumps are limited the required energy is small, and the difference from adiabatic operation is therefore neglected.
- *Pressure drop through piping is neglected.*
  As the piping used for the processes is very limited, the pressure drop across the different units will constitute most of the pressure drop.

5.1 The TEG injection process

As described in chapter 4.1, the absorption process consists of three stages of injection and extraction of TEG from the natural gas. The mixing unit from ProPure is described in chapter 3, and is used for the mixing of TEG and natural gas. For the two first extraction steps, the Deliquidiser from CDS described in chapter 1.6.1 is used, while for the last extraction step a conventional scrubber (described in chapter 1.2) is used to reduce the liquid carry-over from the system. See appendix H.1 for PFD.

The ProPure compact mixer

The compact mixer is modeled using a mixing unit in HYSYS. This unit assumes perfect mixing of the different streams, which is impossible to achieve in practice. So even though the compact mixer from ProPure has shown to be more efficient than other mixing units available, this still will introduce an error in the simulation model. To model the pressure drop of 0.3 bar across the mixer, a valve is placed after the mixing unit.

The Deliquidiser

The Deliquidiser is modeled using the two phase separator unit in HYSYS using the default options. Separation efficiency is given to be 90-99% in table 1.2, meaning that wet TEG will be carried-over to the following stage. This will marginally reduce the purity of the TEG injected in that stage, but since the amount is very small it will barely affect the results. I have tried to model this using the carry-over calculation option available in HYSYS, but it seems to be something wrong with this function, or that my version of HYSYS does not support it. No matter what I choose and type in, I just get the message “carry-over calculations failed”. Carry-over is therefore not calculated, introducing a minor error to the simulated results.

The pressure loss (assumed to be 0.6 bar, referring to table 1.2) is modeled using the same valve as for the mixing unit, placed in front of the separator. This is based on the assumption that most of the pressure is lost during the acceleration of the stream through the swirling element, in front of the actual separation section.

Two Deliquidiser units are used in the absorption process, making up the two first extraction stages. The liquid from the second Deliquidiser is routed through a pump and injected through the ProPure compact mixer in front of the first Deliquidiser, while the liquid from the first Deliquidiser is sent to the regeneration unit.

Glycol pumps

As some pressure is lost through the mixing and separation units, the glycol needs to be pressurized before it is injected in the preceding stage. This is achieved using regular pumps, one for each of the recycling streams. The pump unit in HYSYS is used to model each pump, and the default values are used for all the available options. Pump efficiency is set to 100%, but this will not affect the simulation results noticeably as the pressure difference across the pump is very small (less than 1 bar) and the flow rate limited.

Gas scrubber

For the last extraction stage a conventional gas scrubber is used instead of a Deliquidiser unit to reduce the liquid carry-over. The scrubber is modeled using the two-phase separator unit in HYSYS, using all the default options. To account for the pressure drop, a valve is placed in front of the scrubber, representing the pressure drop for both the mixing unit and the scrubber. As the scrubber unit has very few internals, the only pressure drop is through the inlet- and exit facilities, assumed to be 0.1 bar.

5.2 Regular absorber process

To be able to compare the simulated results for the TEG injection process with conventional glycol dehydration, an absorption column process is also modeled. The process consists of an inlet separator and an absorption column, but the inlet separator is omitted since the stream used for simulation consists of 100% vapor initially. See appendix H.2 for PFD.

Absorption column

For the absorption column, the absorber unit in HYSYS is utilized. 10 *actual* trays are used for the column, resulting in 4 *theoretical* trays using a tray efficiency of 40% (See appendix C.4). The pressure drop across the column is estimated to be 0.2 bar (Coker (2007)). The rest of the options are left with their default values.

5.3 The regeneration process

The regeneration process is modeled after the standard glycol regeneration process presented in chapter 2. See appendix H for PFD.

Pressure relief valve

To reduce the required thickness of pipe- and vessel walls, the pressure is lowered as soon as the rich glycol enters the system. This is accomplished using the nozzle unit in HYSYS and setting the outlet pressure from the nozzle to 2 bar. Gauge pressure is used because of the pressure drop through the piping and heat exchangers placed before the regeneration unit, and to avoid venting to much glycol in the HC flash tank.

Filters

Filters are most important for removing impurities and liquid HC so it doesn't accumulate in the circulated TEG over time. When doing equilibrium simulations in HYSYS the filters are therefore not employed.

HC flash tank

To regain some of the HC absorbed by the glycol and prevent foaming in the regeneration still, the rich glycol is passed through a flash tank after the pressure relief valve. As there is only one liquid phase in the rich glycol, the two phase separator unit in HYSYS is used for modeling. All the default options are used, with the pressure drop set to 0.1 bar. In chapter 2.1.3 the HC flash tank pressure is recommended to be between 3 and 7 bar, but since the TEG injection concept involves more intimate contact with the natural gas, more HC is probably absorbed by the TEG. To prevent too much HC from entering the regeneration still, the pressure is therefore lowered to 2 bar.

Overhead condenser

The overhead condenser is a heat exchanger where the cold rich glycol is used to provide reflux for the regeneration still. The heater unit in HYSYS is used to model the rich glycol side of the overhead condenser, while the cooling side is integrated in the distillation column unit. For the heater a pressure drop of 0.2 bar (typical pressure drop for heat exchangers recommended by Coker (2007)) is specified, while the pressure drop for the condensing side is neglected. Using a recycle element, the energy stream from the condenser is routed to the heater unit to compute the correct rich glycol outlet temperature.

Glycol/glycol heat regenerator

One heater and one cooler unit in HYSYS are used for modeling the glycol/glycol heat exchanger. The pressure drop is specified to be 0.05 and 0.2 bar for the lean and rich side, respectively (Different pressure drops since a shell-and-tube is commonly used; a smaller pressure drop for the shell side and a larger pressure drop for the tube side). The outlet temperature for the lean TEG is set to 90^{o}C and the energy stream from the lean cooler unit is recycled using a recycle element to the rich heater unit. This forces HYSYS to iterate over the outlet temperature of the rich glycol, as it is dependent on the circulation rate of glycol.

Regeneration still

For modeling the glycol regeneration still, a distillation column unit in HYSYS is used. Four theoretical trays are specified for the column (that is, four theoretical trays including the reboiler and the condenser), with the feed entering on tray 2. Specifications for the column are chosen as recommended by Øi & Bråthen (2000), with the reboiler temperature specified as 204°C and the reflux ratio as 50% for TEG. The rest of the options are used with their default values. The total pressure drop is estimated to 5 kPa including the pressure drop across the reflux condenser (Coker (2007)).

Stripping gas

To be able to simulate dehydration using higher glycol purities, stripping gas is also added to the regeneration column. The stripping gas is taken from the inlet to the dehydration process using a splitting unit. The gas is then expanded to the same pressure as the rich glycol entering the stripping column using a valve unit (110 kPa).

Makeup TEG

As some TEG is lost throughout the process, new TEG needs to be added for the system to converge. This is accomplished by inserting a mixing element before the glycol pump, merging a “Makeup TEG”-stream and the regenerated TEG stream. The flow rate of lean TEG is one of the given parameters for the system, and the “Makeup TEG”-stream is therefore adjusted by HYSYS to obtain this. The composition of the makeup TEG stream is specified to equal the composition of the lean TEG from the regeneration still to avoid influencing the TEG leanness.

Glycol pump

A regular pump unit is used for modeling the glycol pump, using all the default values and setting the efficiency to 100%.

Gas/glycol heat exchanger

As the glycol is injected into the gas stream, the temperature of the lean glycol will very quickly drop to the gas temperature. The gas/glycol heat exchanger is therefore omitted for the TEG injection concept. For conventional absorber dehydration a cooler unit is used, with the outlet temperature of the lean glycol set to 30°C. The pressure drop is specified to 0.2 bar.

5.4 Using different chemicals for dehydration

To decide if TEG is the best suited chemical to use for the compact dehydration process, other chemicals will be simulated as well. This is done by just altering the composition of the lean absorbent stream and adjusting the regeneration temperature for the new compound.

Because of problems with my HYSYS license, I was not able to simulate when using amines for dehydration, so the only alternative components simulated are MEG, DEG and tetra ethylene glycol (TTG). The dehydration performance of TTG was however so poor that the results are omitted.

Also, to be able to make comparisons when using different types of chemicals, the flow rate of absorbent is given in kg/hr instead of kgmole/hr. Preferably the flow rate should have been given in liter/hr or m^3/hr, but this is not supported by HYSYS.

Chapter 6: Simulation results

This chapter presents the results obtained when simulating using the simulation models described in chapter 5 together with input data from the Snøhvit LNG processing plant (See appendix I). First the simulation results for the TEG injection concept is presented and compared to the simulation results for conventional absorber dehydration. Several simulation cases are used, each shortly described before the results are presented. Secondly the injection concept is simulated using different types of absorbent (MEG and DEG), and different parameters of interest compared to TEG. Finally a discussion comparing the TEG injection concept to absorber dehydration, as well as comparing usage of different types of absorbent is given.

Because of problems with the CPA EoS increasing the time required for simulation greatly, the number of data points had to be decreased. This results in that some of the graphs are a bit rough, but the results are still considered reliable.

6.1 Comparing the TEG injection process to absorber dehydration

The three stage TEG injection concept and conventional absorber dehydration are simulated, and the different parameters for the process compared. The absorbent flow rate for these paragraphs is given in kgmole/hr, and one must consider that the molar weight of TEG is approximately 150 kg/kgmole (Kohl & Nielsen (1997) [1]) when evaluating the results.

6.1.1 Dehydration performance

This simulation case is designed to compare the dehydration performance of the three stage TEG injection concept with conventional absorber dehydration. The reboiler pressure has been kept constant at 110 kPa, and the flow rate of lean TEG and the regeneration temperature have been varied. The results are presented below.

Water content (no stripping gas)

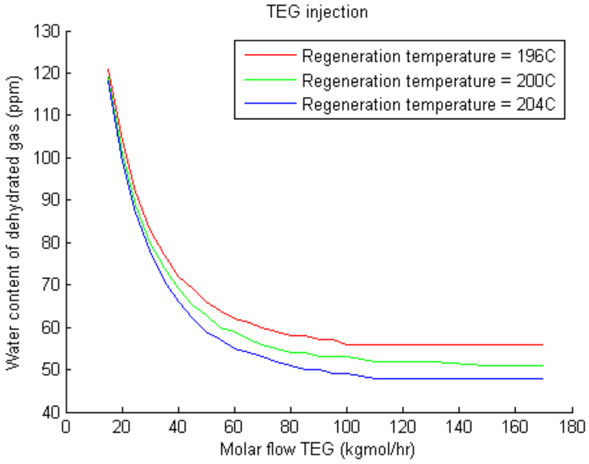

Figure 6.1: Water content of dehydrated gas (TEG injection)

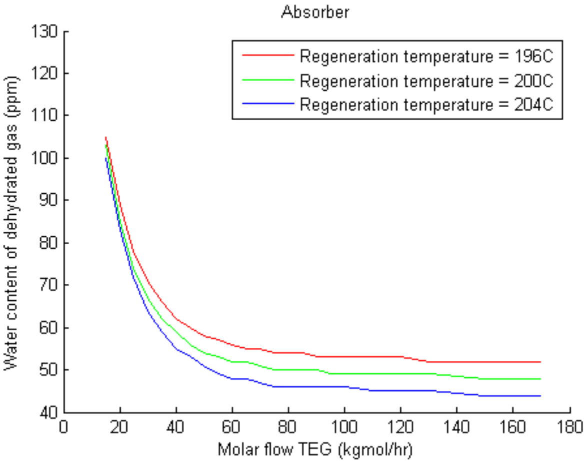

Figure 6.2: Water content of dehydrated gas (Absorber)

The first thing that can be seen from figures 6.1 and 6.2 is that the regeneration temperature has only a small impact on the water concentration of the dehydrated gas. Lowering the temperature from 204°C to 200°C, increases the water content by 4 ppm for TEG flow rates greater than approximately 50 kgmole/hr. For lower flow rates the impact is even smaller, from 2 – 3 ppm. The same increase in water content is found when further lowering the regeneration temperature to 196°C.

Secondly, using an absorber for dehydration is seen to accomplish lower water concentrations in the dehydrated gas using lower flow rates of TEG than the TEG injection concept. E.g. for a regeneration temperature of 204°C and a required water content in the dehydrated gas of 55 ppm, TEG injection requires approximately 60 kgmole/hr, while the absorber requires only 40 kgmole/hr. Alternatively the regeneration process used for the absorber can operate at 196°C and still obtain the same dryness as the TEG injection process regenerated at 204°C using the same TEG flow rate.

The equilibrium water concentration is also affected by the regeneration temperature, and the difference is approximately 4 ppm for each temperature step for the absorber. For the TEG injection process, the difference is 3 and 5 ppm going from 204°C to 200°C and 200°C to 196°C, respectively.

Water content (with stripping gas)

This simulation case is also designed to compare the dehydration performance of the TEG injection concept with conventional absorber dehydration, but now with stripping gas added to the regenerator (i.e. the purity of the lean TEG is higher). The reboiler pressure and temperature has been kept constant for this simulation, while the flow rate of stripping gas and flow rate of TEG is varied. Reboiler pressure has been set to 110 kPa and regeneration temperature to 204°C. Water saturated natural gas at 64 bar is used as stripping gas, taken from the process stream at the inlet to the dehydration unit. The results are presented below.

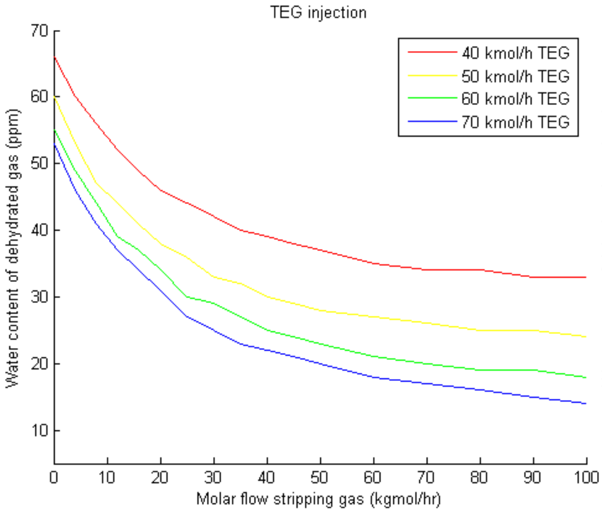

Figure 6.3: Water content of dehydrated gas (TEG injection)

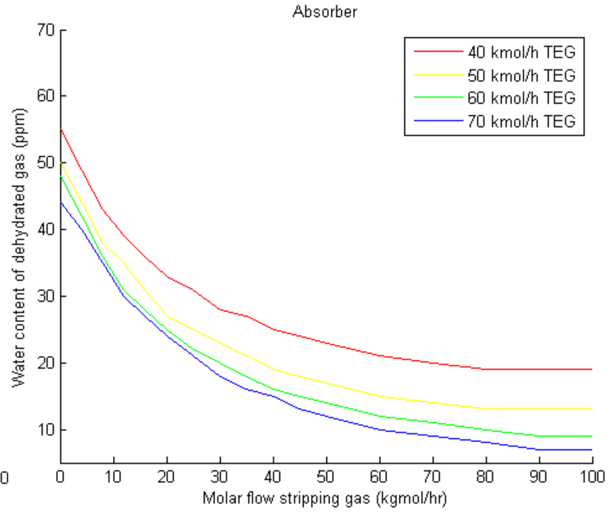

Figure 6.4: Water content of dehydrated gas (Absorber)

While it was required to study the figures when regenerating without stripping gas to observe which was the best dehydration process, it becomes much clearer when using stripping gas (See figures 6.3 and 6.4). E.g. using 50 kgmole TEG/hr and 20 kgmole stripping gas/hr for regeneration results in water content of the dehydrated gas of 27 and 38 ppm for the absorber and TEG injection concept, respectively. To be able to obtain the same dryness, the TEG injection concept requires either an increased TEG flow rate of approximately 20 kgmole/hr or increased stripping gas flow of 40 kgmole/hr (combinations of increasing TEG flow and stripping gas flow is also possible).

It can also be observed from figures 6.3 and 6.4 that the equilibrium concentration for the dehydrated natural gas is very different for the absorber and the injection concept. For a TEG flow rate of 60 kgmole/hr the equilibrium concentration of water is twice as high for the TEG injection process, as for the conventional absorber process.

6.1.2 Energy consumption

This simulation case is designed to determine the connection between flow rate of TEG, regeneration temperature and required heat for the reboiler. Heat is normally supplied by burning a portion of the gas, and is therefore also an important economic parameter. The regeneration pressure is kept constant at 110 kPa for this simulation, and the regeneration temperature and flow rate of lean TEG is varied. As the same regeneration process is employed for both absorption processes, only the data from the absorber process is shown. The results for the injection process is going to be slightly different, as the process is able to remove less water, and consequently decrease the load on the regeneration process. However, the differences are so small that it is not considered prominent enough to show in a separate graph.

Heat required for reboiler (no stripping gas)

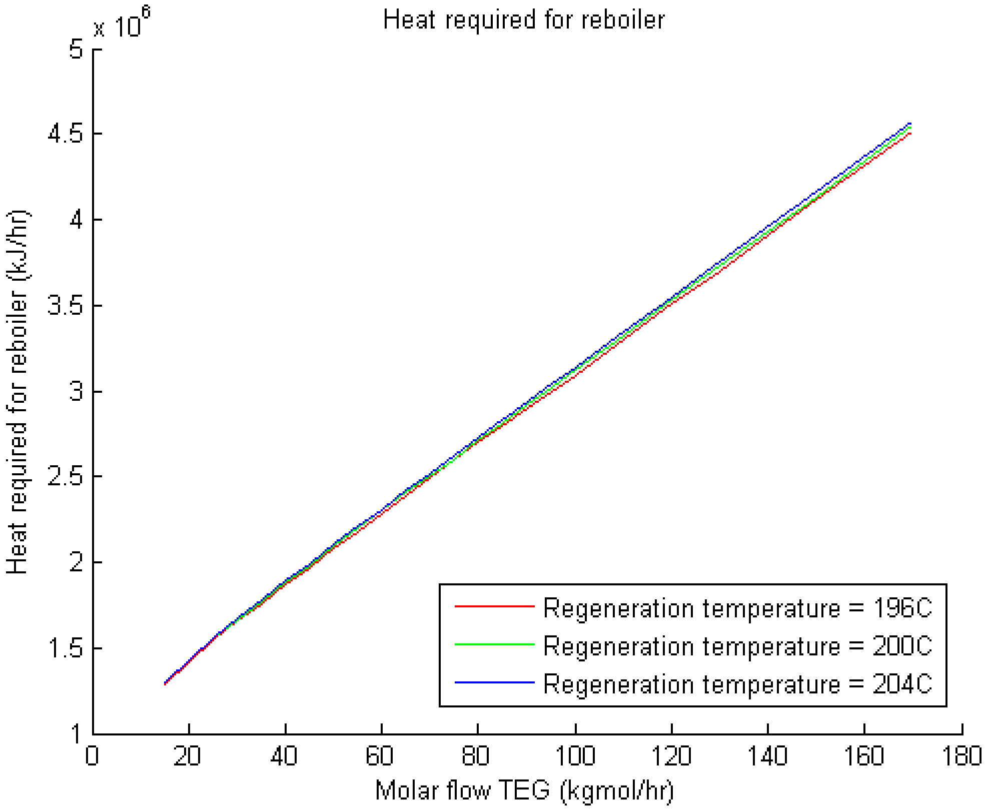

Figure 6.5: Heat requirement for reboiler at different regeneration temperatures and TEG flow rates

As can be seen from figure 6.5, the regeneration temperature has very little impact on the required heat for the reboiler. This is because of the good heat recuperation of the process, exchanging heat between the rich and the lean glycol. So if the glycol is heated another 4°C in the reboiler, this results in the lean glycol being 4°C warmer, and consequently the rich glycol being heated another 4°C in the glycol/glycol heat regenerator. The small difference is caused by additional vaporization in the reboiler when increasing the temperature.

The required heat is also seen to be almost linear dependent on the flow rate of lean TEG. The small deviation from linearity when increasing the flow rate is because the TEG absorbs more water which again must be vaporized in the regeneration column.

Heat required for reboiler (with stripping gas)

This simulation case is designed to investigate how the addition of stripping gas to the regeneration column influences the heat required for the reboiler. The regeneration temperature is kept constant at 204°C and the regeneration pressure at 110 kPa, while the flow rate of lean TEG and stripping gas is varied. Stripping gas is supplied by extracting a portion of the wet natural gas.

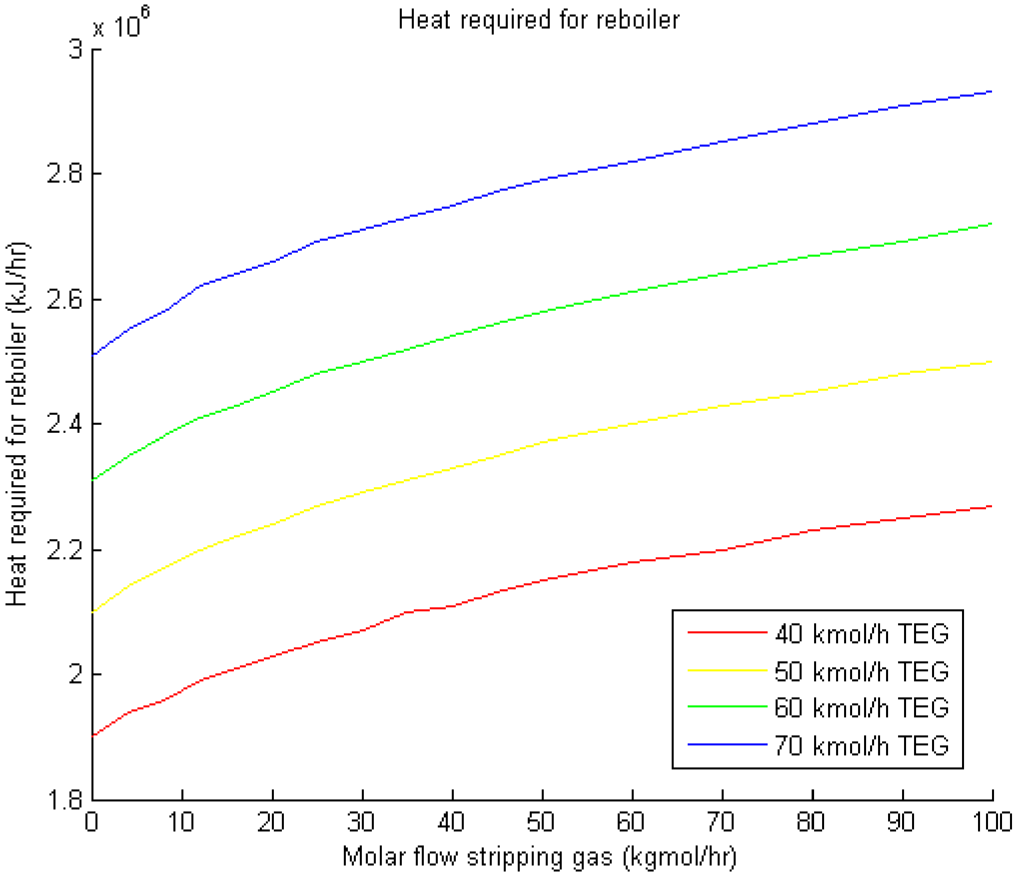

Figure 6.6: Heat requirement for reboiler when introducing stripping gas

From figure 6.6 it can be seen that the addition of stripping gas affects the heat consumption of the reboiler quite a bit. This is both because the stripping gas itself must be heated, but also because more water is absorbed by the glycol and needs to be vaporized. For a TEG flow rate of 40 kgmole/hr, an addition of 30 kgmole/hr of stripping gas increases the heat requirement by almost 10%. The added heat is however observed to be independent of the TEG flow rate, and only a function of the stripping gas flow rate (as all the lines have similar shape).

The gradient for the required heat is observed to decline as the flow rate of stripping gas increases. This is probably because the stripping gas is able to absorb more water from the rich glycol as it enters the reboiler, so that the amount of water that must be vaporized is reduced. The heat required is still rising though, indicating that the amount of heat needed for heating the stripping gas is greater than the heat saved by reduced vaporization.

6.1.3 HC emission

As earlier mentioned, the off gasses from the regeneration column can contain large quantities of HC which are hazardous for the environment. This simulation case is therefore designed to obtain the connection between TEG flow rate and emissions of HC through the overhead vapor from the regeneration column. The regeneration temperature is kept constant at 204°C, the pressure constant at 110 kPa and no stripping gas is added, while the TEG flow rate is varied. The results for both the TEG injection concept and conventional absorber dehydration are presented in figure 6.7 below.

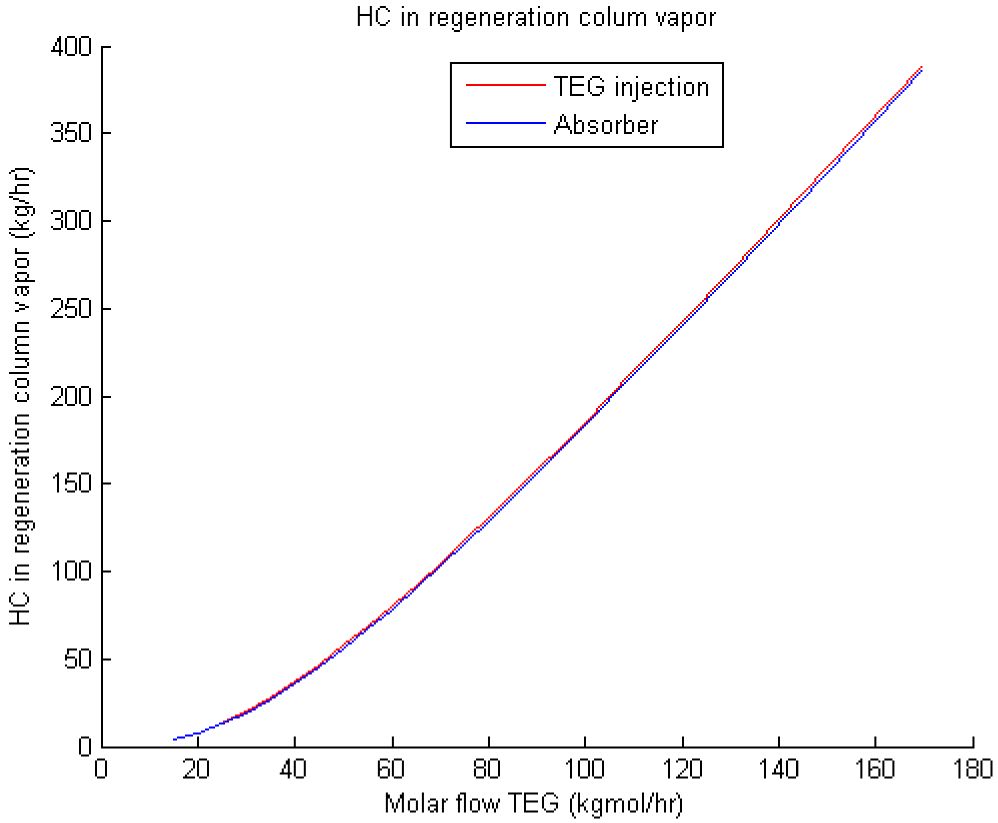

Figure 6.7: HC emissions from regeneration column

From figure 6.7 it is seen that for a given flow rate of TEG, there is very little difference in the amount of HC vented from the regeneration column for the two different dehydration processes. The TEG injection vents slightly more HC to the surroundings, which is expected because of the more intimate contact with the natural gas.

The shapes of the curves are also similar, starting off exponential, but approaches linearity for flow rates greater than approximately 90 kgmole/hr. The reason for this behavior is that when increasing the flow rate of TEG, the water concentration in the wet gas entering the second and third injection/extraction stage decreases, thereby making the glycol able to absorb more HC in these stages. However as the water content decreases and approaches equilibrium, no more water can be absorbed, and the increased amount of HC absorbed becomes linearly dependent of the increase in TEG flow rate.

To be able to estimate the HC emissions for the different processes in relation to water content in the dehydrated gas (which is the parameter of primary interest for the dehydration process), one must however see figure 6.7 in relation to figures 6.1 and 6.2. If the required dryness of the dehydrated natural gas is for example 60 ppm, the required circulated amount of lean TEG is 50 and 35 kgmole/hr for the TEG injection concept and absorber, respectively (provided no stripping gas is used). Looking at figure 6.7 this implies that the TEG injection concept vents about twice the amount of HC compared to absorber dehydration when obtaining a water concentration in the dehydrated gas of 60 ppm.

Summarized

Table 6.1 presents a comparison of the TEG injection process to regular absorber dehydration. A water specification of 50 ppm in the dehydrated gas is chosen, and the rest of the parameters obtained from the simulation results. No stripping gas is used for this comparison.

Table 6.1: Comparing TEG injection to absorber dehydration

	Water content	Flow rate lean TEG	Energy required	HC emissions
TEG injection	50 ppm	12750 kg/hr	2.83 MJ/hr	144.4 kg/hr
Absorber	50 ppm	7875 kg/hr	2.16 MJ/hr	62.9 kg/hr
Difference	--	4875 kg/hr (+60%)	0.67 MJ/hr (+31%)	81.5 kg/hr (+130%)

6.2 Using different absorbents for the TEG injection process

To be able to choose the best suited absorbent for the injection dehydration concept, this paragraph presents the simulation results using different types of absorbent for the injection concept. As earlier mentioned there was a problem with the amine fluid package in HYSYS, so the only compounds simulated are MEG and DEG. To be able to compare the parameters for the different absorbents, the flow rate of absorbent is given as kg/hr in the following paragraphs.

6.2.1 Dehydration performance

This simulation case is designed to compare the dehydration performance when using different types of absorbent for the TEG injection concept. The absorbents used are MEG and DEG. Regeneration temperature is kept constant at the maximum recommended regeneration temperature for the given absorbent at sea level (165°C for MEG , 170°C for DEG and 204°C for TEG (Arnold & Stewart (1999) [1], Kohl & Nielsen (1997) [1])), regeneration pressure kept constant at 110 kPa, while the flow rate of absorbent is varied.

Water content (no stripping gas)

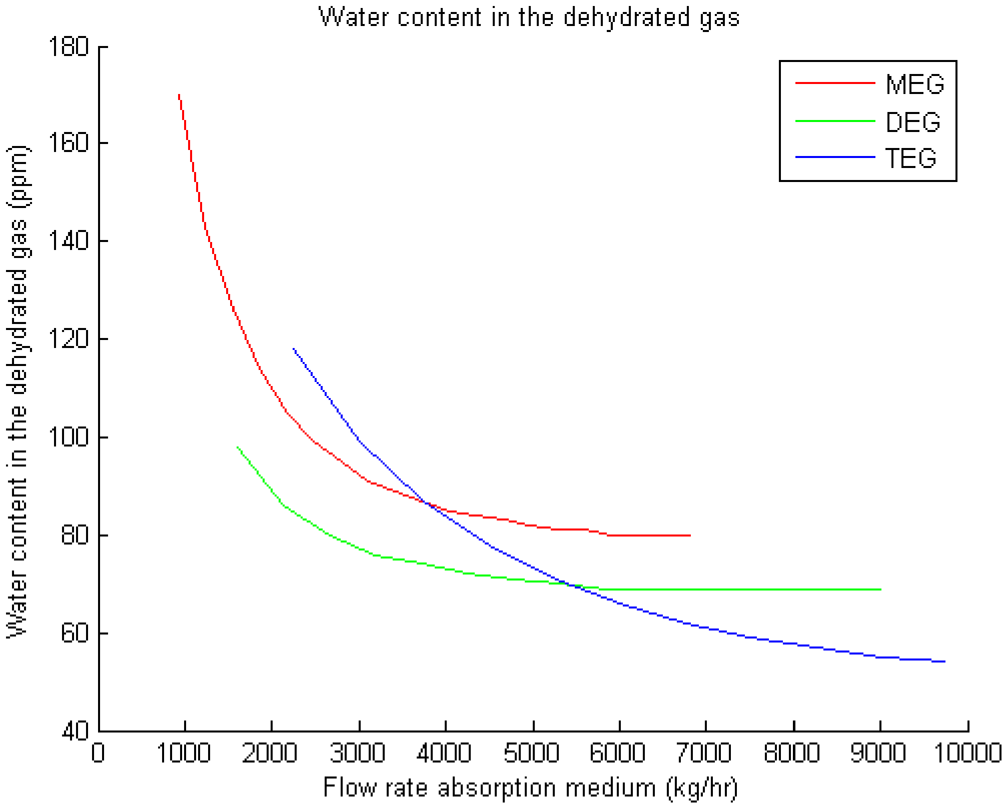

Figure 6.8: Water content in the dehydrated gas for the different absorbents

From figure 6.8 it can be seen that the dehydration performance of the different absorbents vary greatly. Using MEG requires large flow rates of absorbent just to obtain moderate water concentrations in the gas, and the lowest concentration possible is as high as about 80 ppm. DEG actually outperforms TEG with respect to dryness for flow rates lower than approximately 5500 kg/hr, but for higher flow rates TEG is better. It is also noted that DEG and MEG reaches equilibrium conditions with the dehydrated gas already at about 6000 kg/hr, while TEG has yet to reach equilibrium at 10 000 kg/hr.

The explanation for the poor dehydration performance of MEG and DEG when compared to TEG is possibly insufficient leanness of the lean glycol. Using the simulation model it is found that using the maximum recommended regeneration temperature yields 96.40 wt% 96.57 wt% and 99.18 wt% purity for MEG, DEG and TEG respectively for a regeneration pressure of 110 kPa.

Water content (stripping gas)

The same as the above case, but stripping gas is now used to increase the purity of the lean absorbent. Wet natural gas from the inlet of the dehydration process is used as stripping gas, and the flow rate is varied. The flow rate of lean absorbent is kept constant at 6000 kg/hr.

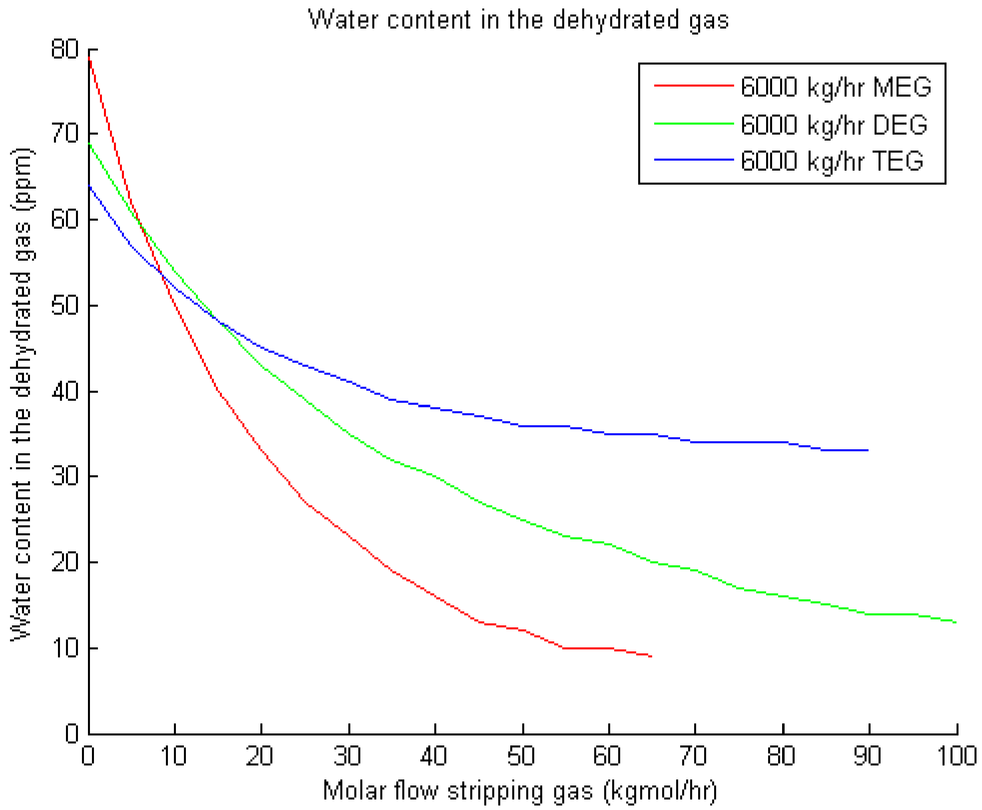

Figure 6.9: Water content for the different glycols when using stripping gas

As expected, the dehydration performance is greatly enhanced by introducing stripping gas in the regeneration column. Especially MEG benefits greatly from enhanced regeneration, and outperforms both DEG and TEG for stripping gas flow rates larger than approximately 10 kgmole/hr. TEG has the best dehydration performance at stripping gas flow rates less than 10 kgmole/hr, but for flow rates larger than 20 kgmole/hr it is outperformed by both MEG and DEG.

Quite surprisingly, MEG has by far the best dehydration performance after introducing stripping gas, and water content as low as 10 ppm is achievable with about 65 kgmole/hr of stripping gas. DEG follows close behind capable of obtaining 15 ppm using 100 kgmole/hr of stripping gas, while TEG is only capable of obtaining 35 ppm. As the purity of the lean TEG without stripping gas already is more than 99 wt%, the reason for the insufficient dehydration performance for TEG is the TEG flow rate.

6.2.2 Energy consumption

This simulation case is designed to compare the heat required for the reboiler when regenerating the different absorbents. No stripping gas is used for this simulation, the regeneration temperature is kept constant at the maximum recommended temperature for each compound, and the regeneration pressure constant at 110 kPa. The flow rate of lean absorbent is varied.

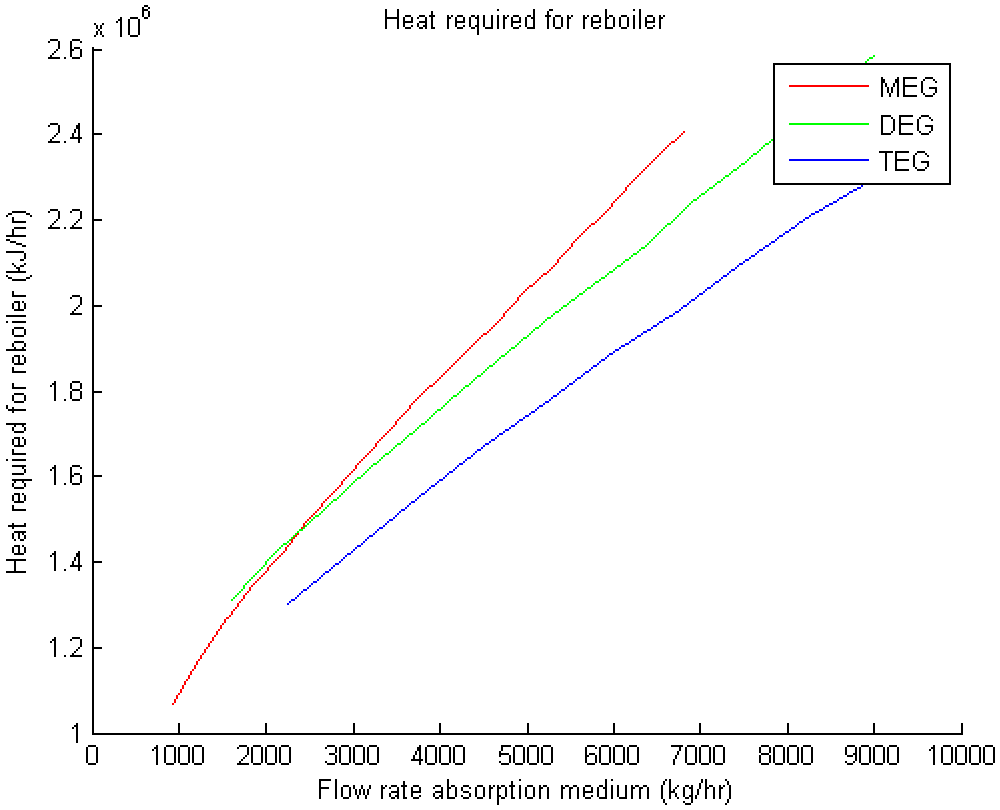

Figure 6.10: Heat required for the reboiler for the different glycols

From figure 6.10 it is observed that MEG is the absorbent which requires the most heat for regeneration, which is quite surprising as it is regenerated at the lowest regeneration temperature of the different compounds. From Kohl & Nielsen (1997) [1] it is however observed that MEG has both the highest specific heat and heat of vaporization, explaining the required added heat. It is also observed that the required heat for regenerating DEG and TEG is almost linear, while MEG increases more than linearly for low flow rates of absorbent. This is probably because the concentration of water in the rich MEG is very large for low flow rates (vaporizing the water requires more heat than increasing the temperature of the glycol). For DEG and TEG these low flow rates are not simulated, but the relation probably also applies to them.

TEG requires the least amount of heat for regeneration, and by closer inspection of figure 6.10, the difference is found to be quite big. E.g. regenerating 6000 kg/hr of MEG requires the same amount of heat as regenerating approximately 8200 kg/hr of TEG. Corresponding, regenerating 6000 kg/hr of DEG requires the same amount of heat as regenerating 7000 kg/hr of TEG.

6.2.3 Loss of absorbent

This simulation case is designed to investigate how the flow rate of absorbent affects the total loss of absorbent for the dehydration and regeneration process. The regeneration temperature is kept constant at the maximum recommended regeneration temperature and the pressure constant at 110 kPa, while the absorbent flow rate is varied.

Total loss of absorbent (no stripping gas)

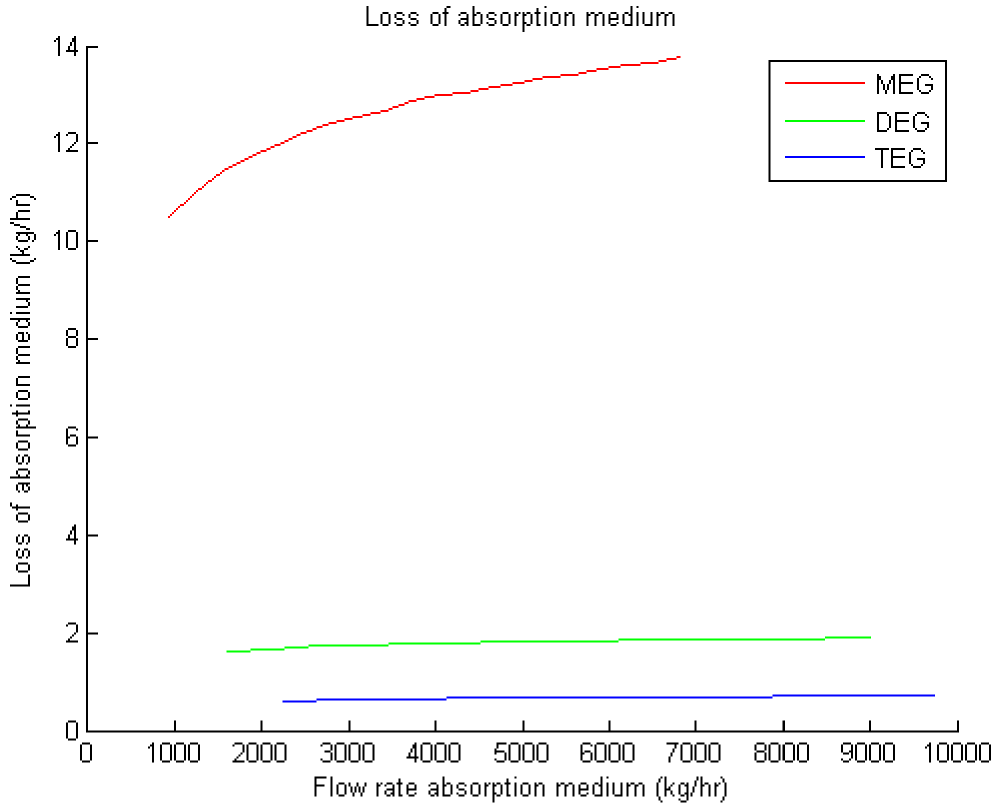

Figure 6.11: Total loss of absorbent for the dehydration process

As can be seen from figure 6.11, the absorbent loss for MEG by far exceeds the losses of DEG and TEG. Already for flow rates as low as 1000 kg/hr the loss is 10 kg/hr, or 1% of the circulated amount. By inspecting the simulation model, it is found that almost 80% of the loss is from vaporization into the processed natural gas. This loss is difficult to reduce, and the only measure is to cool the natural gas to reduce the vapor pressure of the MEG.

The loss when using DEG for dehydration is, although it is much lower than for MEG, observed to be greater than twice the loss as for using TEG. The loss of TEG varies from approximately 0.7 kg/hr to 1 kg/hr depending on the flow rate.

Total loss of absorbent (stripping gas)

This is the same simulation case as above, with the exception of the absorption flow rate being constant and the flow rate of stripping gas varied. The flow rate of lean absorbent is kept constant at 6000 kg/hr.

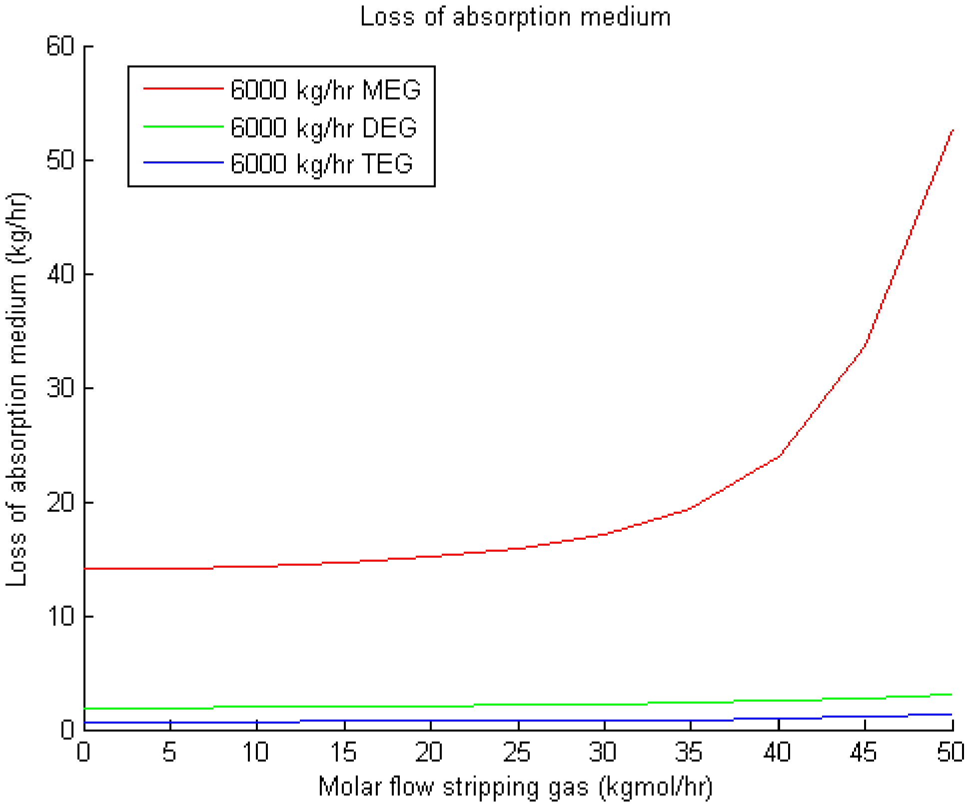

Figure 6.12: Total loss of absorbent when using stripping gas

The results from the previous simulation case are further supported by figure 6.12, worsening the comparison for MEG. For stripping gas flow rates lower than approximately 8 kgmole/hr, the loss does not increase when adding stripping gas, but it is still formidable 14 kg/hr. When the stripping gas flow rate is increased beyond 30 kgmole/hr, the loss starts to increase almost exponentially and totals almost 35 kg/hr when using 45 kgmole/hr of stripping gas. From inspection of the simulation model, it is found that the loss with the dehydrated natural gas has not increased, and that the increased loss is from the regeneration column and the flash tank. The loss from the flash tank could be somewhat reduced by increasing the operation pressure, but then at the expense of more dissolved HC in the glycol.

Because of the huge loss of MEG, the increase in the losses of DEG and TEG is a bit harder to read from the figure. A closer inspection however reveals that the losses increase to approximately 3 kg/hr and 1.5 kg/hr for DEG and TEG respectively for a stripping gas flow rate of 45 kgmole/hr.

6.2.4 HC emissions from the regeneration column

This simulation case is designed to investigate how much HC the different absorbents release in the regeneration column. The regeneration temperature is kept constant at the maximum recommended regeneration temperature for each compound and the pressure constant at 110 kPa. The flow rate of lean absorbent is varied, and no stripping gas is used for this simulation.

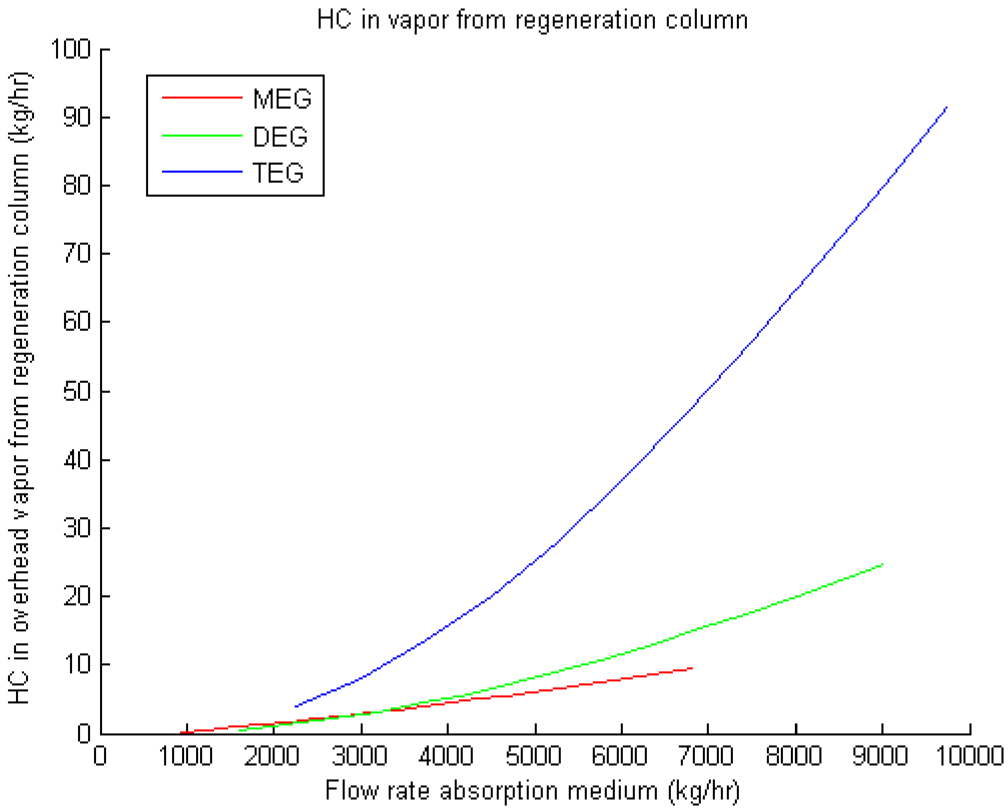

Figure 6.13: HC emissions from the regeneration column for the different glycols

As expected, TEG absorbs a lot more HC than DEG and MEG as can be seen from figure 6.13. For flow rates lower than approximately 3000 kg/hr the difference is not so big, while it increases for larger flow rates. DEG absorbs slightly less HC than MEG for flow rates lower than 3000 kg/hr, but for greater flow rates the situation is reversed.

It can also be observed that the amount of HC vented increases more than linearly for all types of glycol simulated. This is because of the same reason as described in chapter 6.1.3, that the water concentration for the last two stages is lowered when increasing the glycol flow rate, making the glycol able to absorb more HC in these stages. Ultimately a linear relation will be reached, as the water content in the natural gas can not be lowered further by increasing the absorbent flow rate.

Summarized

Table 6.2 presents comparisons data when using MEG, DEG and TEG as absorbent for the injection process.

Table 6.2: Comparison of operation data for different absorbents

	Lean glycol	Stripping gas	Water content	Energy	HC emission	Absorbent loss
MEG	6000 kg/hr	15 kgmole/hr	40 ppm	2.30 MJ/hr	8.0 kg/hr	14.61 kg/hr
DEG	7520 kg/hr	24 kgmole/hr	40 ppm	2.42 MJ/hr	17.5 kg/hr	2.11 kg/hr
TEG	7500 kg/hr	18 kgmole/hr	40 ppm	2.23 MJ/hr	57.3 kg/hr	0.68 kg/hr

6.3 Discussion

The findings from my semester project (Bråthen (2007)) concerning the dehydration performance of the TEG injection concept when compared to conventional absorber dehydration is further supported from the simulation results presented in chapter 6.1.1. The differences are however smaller, caused by using a more realistic number of theoretical trays for the absorber. Also, if the requirements for water concentration are less stringent, TEG injection becomes more competitive and is almost as good as absorber dehydration.

Introducing stripping gas increases the difference in dehydration performance between TEG injection and absorber dehydration, and the effect of an extra theoretical tray becomes prominent. TEG injection is however still able to accomplish water concentrations down to 15 ppm in the dehydrated gas, using 70 kgmole/hr of TEG and 100 kgmole/hr of stripping gas. This is sufficiently low for many low temperature processes, but probably not economically competitive with using an absorber.

As expected, the heat consumption of the reboiler is found to be linearly dependent on the flow rate of TEG and not much affected by the regeneration temperature. Adding stripping gas increases the heat requirement, but the gradient declines as more stripping gas is added and the need for vaporization of water is reduced. However, when calculating the heat requirement for each process related to dryness in the dehydrated gas, the TEG injection concept requires more heat than absorber dehydration, as the TEG flow rate is larger for injection dehydration.

Venting of HC from the regeneration column is a problem both when using an absorber and TEG injection for dehydration. Compared to the emission regulations for USA presented in chapter 2, the flow rate of TEG must be below approximately 10 kgmole/hr to be able to operate the column without processing the vapor. Looking at figure 6.1 and 6.2 this would result in a water concentration of about 130 and 110 ppm in the dehydrated gas for TEG injection and absorber respectively, way above the concentration required for pipeline transport. A vapor handling system is therefore necessary for both dehydration processes to be able to follow the emission regulations.

When simulating using different types of absorbent for the injection concept, DEG is the glycol capable of producing the lowest water concentration for flow rates between 2000 and 5000 kg/hr when no stripping gas is used for regeneration. This is quite surprising, since the lean concentration of the DEG is only about 96.6 wt%. The obtained dryness is however only approximately 70 ppm, which in most cases is too high for even pipeline transport specifications. TEG has the worst dehydration performance for flow rates below 4000 kg/hr, but is the best for flow rates greater than 5500 kg/hr. This is expected, as it has the highest lean purity and therefore will have the lowest equilibrium concentration when the flow rate is increased towards infinity. The dehydration performance of MEG is mediocre for flow rates lower than 4000 kg/hr, and becomes the worst for higher flow rates.

Adding stripping gas to the glycol regeneration column completely changes the dehydration performance characteristics for the different glycols, now making MEG the superior glycol for flow rates higher than 10 kgmole/hr of stripping gas. MEG is even able to obtain water concentrations in the natural gas reaching as low as 10 ppm, which is quite impressive. DEG also performs quite well, and obtains lower water concentrations than TEG for stripping gas flow rates larger than approximately 15 kgmole/hr. Using large quantities of stripping gas it is even able to obtain concentrations below 20 ppm. TEG has the best dehydration performance for stripping gas flow rates lower than 10 kgmole/hr, but is the worst glycol for flow rates larger than approximately 15 kgmole/hr. However, adding 100 kgmole/hr of stripping gas does not improve the dehydration performance much over adding 30 kgmole/hr, indicating that the flow rate of TEG is the limiting factor for the dryness.

The energy consumption for regenerating the different glycols varies quite much for a given flow rate, with TEG being the glycol requiring the least amount. When considering energy consumption, the important parameter is however the amount of energy required to obtain a given dryness in the processed gas, and not the energy consumption for a given flow rate of absorbent. Therefore, figure 6.10 must be viewed together with figure 6.8 and 6.9. Ideally I should have given a 3D-figure where the axis represented glycol flow, stripping gas flow and heat requirement, so it would have been easier to find the most energy efficient glycol and the ideal operating conditions. Unfortunately because of the previously described problems with the CPA EoS, this has not been possible.

The graphs for loss of absorbent clearly illustrate why MEG is not used for dehydration, even though it has proven to have the best dehydration performance of the glycols simulated. Even without stripping gas, the loss of MEG starts off at a whooping 10 kg/hr for a flow rate of 1000 kg/hr. This is about 5 and 10 times the loss for DEG and TEG respectively, and much too high to make MEG an alternative for use in the TEG injection concept. Also as previously noted, most of the loss is with the dehydrated gas, making it almost impossible to recover. When introducing stripping gas, the results become even grimmer, and for stripping gas flow rates larger than 30 kgmole/hr the loss of MEG increases almost exponentially.

Although overshadowed by the huge loss of MEG, the losses of DEG and TEG also increases when introducing stripping gas. The increase is however only linear, resulting in a loss of about 3 and 1.5 kg/hr of DEG and TEG respectively for a stripping gas flow rate of 45 kgmole/hr. This is well within the acceptable limits presented in chapter 2. One must however keep in mind that carry-over is neglected in these simulations, so the real loss for an injection process is going to be higher.

Again, it is of no surprise that TEG absorbs much more HC than MEG and DEG. For all TEG flow rates larger than approximately 2000 kg/hr, the vapor must be processed to meet American emission regulations, as previously noted. DEG absorbs less HC than TEG, but already at a flow rate of 3000 kg/hr the HC emission exceeds the American emission limit, and the vapor must be processed. The vented amount of HC when using MEG has a much gentler slope than TEG and DEG, but also breaches the emission limit already at 3500 kg/hr. The amount of HC vented is therefore not so much of a comparison criterion (at least not for this case, using the reference composition and flow given) when deciding which is the best glycol for the injection dehydration concept. Of course the load on the vapor handling system would be slightly different for the different glycols, but the added flow of HC is very small compared to the total flow of vapor (consisting merely of water).

Chapter 7: Designing a TEG injection system

This chapter uses the simulation model from chapter 5 to obtain operating data for the TEG injection process for a chosen reference case. First the regular requirements for pipeline dehydration are presented, before the reference case is chosen. Using the operating data, the sizing procedures for separators and the different parts of the regeneration process (presented in appendix A and C) is used to design a complete TEG injection system. Operating data from an operational TEG absorber process from the Kristin field in the Norwegian North Sea is then presented, before finally a comparison between the simulated TEG injection system and the operational conventional absorber is made. All calculations are shown in appendix J.

7.1 Dehydrating for pipeline specifications

In short, the dehydration requirement for pipeline transport demands that under no circumstances, is water allowed to condense from the gas forming liquid water (Arnold & Stewart (1999) [1]). As condensation depends on pressure and temperature, the ambient temperature will influence the maximum allowable water content in the gas, thus the pipeline requirements differs for different parts of the world. The water content requirement is commonly substituted by a dew point temperature at a given pressure, representing the temperature at which water from the gas will start to condense at that pressure. In warmer climates where the sea water temperature is high, a water dew point requirement of 10 – 15^{o}C may be sufficient, while for the northern gas fields dew points as low as -20^{o}C can be found (-18^{o}C @ 70 bar is common for the North Sea (Fredheim (undated))).

In this design the water dew point requirement for the Norwegian North Sea of -18^{o}C @ 70 bar will be used. To determine the water dew point of the dehydrated gas, a cooler and a three phase separator is placed after the last TEG injection/extraction stage in the simulation model presented in chapter 5. The dehydrated gas is then cooled until a separate water phase is formed in the three phase separator, and this temperature is used as the water dew point temperature. For -18^{o}C and 70 bar the corresponding water concentration is found to be 29 ppm for the reference gas.

As seen from the simulation results presented in chapter 6, several combinations of TEG and stripping gas flow rates can be used to obtain a given dryness in the dehydrated gas. The main goal for further development of the TEG injection concept is to reduce the weight and volume of the unit, and this is therefore emphasized when determining the operation conditions. From the discussion in chapter 4 and also from my semester project (Bråthen (2007)), it was found that the regeneration unit, and more specifically the regeneration column, contributed most to the overall weight and volume of the process. The size of the column is dependent on the amount of vapor present, and the amount of stripping gas should therefore be limited. By using the simulation model, the following operating conditions are found to be suitable:

Table 7.1: Operating conditions for the TEG injection process

Overhead condenser		**Flash tank**		**Reboiler**	
$T_{gl,i}$	27.4 °C	Q_G	82.67 m^3/hr	Q_L	9,77 m^3/hr
$T_{gl,o}$	77.6 °C	Q_L	9.31 m^3/hr		
m_{gl}	9681 kg/hr	ρ_G	2.07 kg/m^3		
		ρ_L	1021 kg/m^3		
Glycol/glycol HX		**Regeneration column**		**3rd Extraction stage**	
$T_{gl,r,i}$	77.6 °C	$Q_{G,tot}$	1616 m^3/hr	Q_G	12420 m^3/hr
$T_{gl,r,o}$	169.2 °C	Q_L	9.54 m^3/hr	Q_L	9.171 m^3/hr
$T_{gl,l,i}$	204.0 °C	Q_{strip}	553 m^3/hr	ρ_G	54.79 kg/m^3
$T_{gl,l,o}$	90.0 °C	ρ_G	0.804 kg/m^3	ρ_L	1026 kg/m^3
$m_{gl,l}$	8999 kg/hr	ρ_L	977.5 kg/m^3		
		N_{th}	4		

7.2 The TEG injection process

7.2.1 The regeneration process

Overhead condenser

To be able to estimate the size of the overhead condenser, the type of heat exchanger and the overall heat transfer coefficient, U, must be decided. Integrating the heat exchanger in the regeneration column is chosen to save floor space, and U is assumed to be 568 W/m^2K as suggested by Arnold & Stewart (1999) [1]. The procedure described in appendix C.1 is then used to find a value for A_{HT}:

$$A_{HT} = 8\ m^2$$

Assuming the heat exchanger consists of merely pipes placed inside the regeneration column, for 10 pipes with a diameter of 10 cm the resulting length would be 2.5 m for each pipe. Using steel pipes 3 mm thick (the pressure is assumed to be lowered before the overhead condenser, so there is almost no pressure difference), the total weight will be about 190 kg using an average steel density of 7850 kg/m^3 as recommended by Elert (2004).

Because of the high heat transfer coefficient when using a condensing medium, the required heat transfer area is observed to be quite small.

Flash tank

The flash tank is sized using the procedure described in appendix C.2. To minimize the required space, a vertical configuration is chosen and K_1 is assumed to be 0.05 m/s. As there is only one liquid phase, a liquid retention time of 4 minutes and a liquid height of 1.3 m are chosen. This produces the following results:

$$U_t = 1.11\ m/s,\ D_i = 0.78\ m \text{ and } L = 3.08\ m$$

The resulting slenderness ratio is then calculated to 3.95, which is slightly below the maximum recommended for vertical separators, thus the calculated diameter is close to the smallest possible.

To estimate the total weight of the vessel, the wall thickness and material are needed. Using the procedure given in appendix A.6, assuming the material is carbon steel with a maximum stress tolerance of 126.0 MPa (Arnold & Stewart (2008):p.321) and a maximum allowable working pressure (MAWP) for the unit of 230 kPa, I find:

t_W = 6.4 mm and the resulting total weight, M_{tot} = 432 kg without glycol.

The reason for the rather thick wall when compared to the low pressure is to ensure rigidity of the vessel (Coker (2007)).

Glycol/glycol heat exchanger

The glycol/glycol heat exchanger is assumed to be a regular shell and tube heat exchanger with baffles and fins to provide the best possible heat transfer with the smallest heat exchanger possible. From Branan (2002):p.34 it is found that 230 W/m^{2}K is a suitable overall heat transfer coefficient. Assuming the heat exchanger is made from steel, the main dimensions are found to be:

A_{HT} = 125.6 m^2, L = 4 m, D_o = 1.13 m, t_W = 8.1 mm, N_{pipes} = 100, D_{pipes} = 0.05 m, $t_{w,pipes}$ = 3 mm and M_{tot} = 4178 kg

Regeneration column

As the TEG absorbs much more HC than other types of glycol (see chapter 6), structured packing is used for the regeneration column in case of foaming. HETP is assumed to be 1.2 m, in the lower end of the recommended interval from appendix C.4. Using K_1 = 0.1 m/s from appendix C.4 this gives:

U_t = 3.49 m/s, D_i = 0.40 m and L_{tot} = 6.3 m

Using the same procedure as for separators, the thickness of the column wall is calculated and the total weight estimated:

t_w = 6.42 mm and M_{tot} = 708 kg (including the overhead condenser)

Reboiler

The size of the reboiler is calculated using the same procedure as for a horizontal two-phase separator. Assuming the liquid retention time is 5 min, the following is computed:

L = 3.73 m, t_W = 6.4 mm and D_i = 0.745 m

The weight is then calculated to 488 kg and the total volume to 1.68 m^3.

Using the amount of rich glycol to be heated, the required heat is found:

Q_{req} = 1512 kW

For the same operating conditions the simulation model calculates the heat consumption to be only 698 kW, so the formula clearly overestimates the required heat. However, the simulation model does not incorporate heat loss to the surroundings, making the real heat requirement higher than the one simulated.

Glycol pump

The pressure difference across the glycol pump is almost 64 bar, but the flow rate is only 8.71 m^3/hr. Thus the weight of the pump is estimated to be 200 kg and the footprint 0.5 m^2.

Glycol storage tank

As some glycol is lost throughout the process, a storage tank for makeup glycol is required. The TEG loss at the operating conditions is found to be 0.76 kg/hr, thus approximately 550 kg TEG in the tank would be sufficient for one month's operation without refill. This requires a tank with a volume of 0.565 m^3. Choosing a vertical circular storage tank with L=1.5 m and D_i = 0.48 m provides this volume, at the same time reducing the required footprint while maintaining stability. As the storage tank operates at near ambient pressure, the minimum required wall thickness to ensure rigidity is stated to be t_w = 6.4 mm for the chosen diameter (Coker (2007)). This gives a total weight for the storage tank of 134 kg without glycol.

Summarized

The results for the regeneration process are presented in table 7.2:

Table 7.2: Weight, volume and footprint for the regeneration process

Regeneration process				
	Weight [kg]	**Volume [m^3]**	**Footprint [m^2]**	**Glycol volume [m^3]**
Flash tank	432,00	1,52	0,49	0,63
Glycol/glycol HX	4 178,00	4,13	4,52	3,60
Regeneration column	708,00	0,84	0,13	0,00
Reboiler	488,00	1,68	2,78	0,81
Glycol pump	200,00	0,25	0,50	0,00
Glycol storage tank	134,00	0,29	0,73	0,28
Total glycol volume	5 469,00	0,00	0,00	5,32
Total	**11 609,00**	**8,71**	**9,15**	

As can be seen from table 7.2, most of the equipment for the regeneration process is small and lightweight, and actually half the weight is from the glycol contained in the system. This is because of the near ambient pressure, resulting in very thin walls for the different units and the piping. The small size of the regeneration column is quite surprising, but from appendix J.1 it is seen that the low vapor density is the reason for the high terminal velocity, and the resulting small diameter. Compared to operating data presented by Kohl & Nielsen (1997):p.989 it seems to be a bit too small, but this is probably because zero TEG carry-over is the goal for most operational units, thus the dimensions are increased. This is also applicable for the flash tank.

The glycol/glycol heat exchanger is observed to be the biggest part of the regeneration process, constituting almost 50% of the volume and footprint, and about 35% of the weight. This is because of the very large flow rate of glycol through the unit (as both the rich and the lean glycol must pass through this heat exchanger), and the large amount of heat that is transferred.

7.2.2 The absorption process

Pipeline dimensions

To be able to calculate the natural gas pipeline diameter, the gas velocity is assumed to be 20 m/s. Using the definition of volume flow, the diameter is then found to be 0.46 m. As the pressure inside the pipe is much higher than the ambient pressure, it is assumed that higher grade steel is used for manufacture. Using low-alloy steel with a stress allowance of 172.4 MPa (Arnold & Stewart (2008):p.321), the resulting wall thickness is calculated to be 11.9 mm.

Deliquidiser

Assuming the Deliquidiser is produced using the same diameter as the pipe, it is found using table 1.2:

$L = 2.76$ m and $h = 1.84$ m

The height calculated is however the height of the liquid collection section, and not the height of the entire unit (which equals the pipe diameter). The total volume of the liquid collection section is found to be 0.46 m^3 by assuming a liquid retention time of 3 minutes.

Using the data calculated for the pipe, the weight is calculated to 382 kg for the pipe segment required for the unit. Assuming the internals weigh 50 kg and the liquid collection section 100 kg, the total weight for one Deliquidiser unit is 532 kg.

ProPure compact mixer

The ProPure compact mixer is produced using the same diameter pipe as used for the pipeline, as mentioned in chapter 3. Assuming the unit needs to be 5 diameters long to provide sufficient mixing, the weight is found to be 318 kg for the pipe segment. The internal equipment is estimated to weigh 50 kg, making the total weight of the unit 368 kg.

Scrubber

For the last extraction stage a conventional scrubber is employed to reduce the TEG loss for the process. Using a vertical configuration to save space and assuming $K_1 = 0.15$ m/s and $t_L = 15$ min, the following results are obtained:

$U_t = 0.63$ m/s, $h_L = 0.42$ m, $D_i = 2.64$ m and $L = 4.06$ m

The reason for the small U_t and the corresponding large D_i is the high density of the gas, as well as the large gas volume flow. Calculating the slenderness ratio reveals that it is slightly below the minimum required value, but increasing the slenderness ratio further means that the liquid retention time (t_L) must be even greater, increasing the amount of TEG stored in the scrubber.

As the pressure within the scrubber and the diameter is quite large, it is assumed that better quality steel is used for the wall to reduce the required thickness. Using a low-alloy steel with a stress allowance of 172.4 MPa (Arnold & Stewart (2008):p.321) the wall thickness is calculated to 50 mm, while the head thickness is 49 mm. The resulting total weight for the unit is found to be 18 096 kg without TEG.

Piping

Piping is also required between the different units in the absorption process. After the mixing unit it is recommended to have 15 – 20 diameters of straight pipe to enhance the mixing performance. Assuming 17 diameters is sufficient this constitutes 23.46 m of piping for the three mixing units. Gas piping between the other units is also needed, estimated to total 5 m. The combined weight of all the gas piping using a steel density of 7850 kg/hr is then calculated to 3941 kg.

The liquid flow rate is slightly different for the two recycling streams for TEG, but when estimating the piping, the flow rate is assumed to be the same for both (i.e. the same diameter pipes are used). Assuming the liquid velocity is 5 m/s the diameter and wall thickness is calculated to be:

$D_i = 0.026$ m and $t_W = 0.7$ mm

Adding a safety margin to the wall thickness, $t_W = 3$ mm is assumed to be sufficient. Estimating the total length of the pipes to 20 m (a little longer than 2/3 of the total straight pipe length calculated above), the weight is found to be 43 kg.

Pumps

Pumps are needed for pressurizing the recycled glycol in the TEG injection process. However, as the pressure drop is very small (only 90 kPa), the pumps are quite small. Their combined weight is estimated to 50 kg.

Summarized

The results for the TEG injection process is presented in table 7.3.

Table 7.3: Estimated data for the TEG injection process

TEG injection process				
	Weight [kg]	**Volume [m^3]**	**Footprint [m^2]**	**Glycol volume [m^3]**
Deliquidisers	1 064,00	1,01	2,67	0,92
Compact mixers	1 104,00	1,27	3,51	0,00
Scrubber	18 096,00	23,94	5,90	2,30
Gas piping	3 941,00	5,23	13,77	0,00
Recycle piping	43,00	0,02	0,64	0,01
Recycle pumps	50,00	0,003	0,09	0,00
Total glycol volume	3 324,55	0,00	0,00	3,23
Total	**27 622,55**	**31,47**	**26,58**	

For the TEG injection process the last stage scrubber is the by far biggest and heaviest part of the system. It contributes to about 65 % of the weight and 76 % of the volume of the entire unit, indicating that the process would have been much more compact if an inline separator was used for the last extraction stage as well. Of course this would increase the TEG carry-over from the system, probably to a level which would have been unacceptable.

From table 7.3 it can also be observed that the compact mixers are one of the smallest and most lightweight parts of the system. This is somewhat misleading, as most of the gas piping for the system also is considered as a part of the mixing unit. The straight pipe is, as mentioned in chapter 3, used to enhance the mixing performance of the unit, thus it is not required. Removing or reducing the length of the pipe can however lead to incomplete mixing of TEG and natural gas, resulting in insufficient dryness in the dehydrated gas.

7.2 Data from existing TEG dehydration process

To get a better idea of how the TEG injection process performs, the process is compared to a conventional process in operation today. It has been quite difficult to get operational data for a facility, but finally some data from the Kristin field in the Norwegian North Sea was obtained from Bjotveit & Lærum (2005):p.21. The data is quite coarse, and several important facts are missing, but at least some comparisons can be made.

Table 7.4: Operational data for the TEG dehydration process at the Kristin platform (Bjotveit & Lærum (2005):p.21)

	Mass flow [kg/hr]	Size	Pressure [bar]
Absorber	699 300	D = 3.2 m, L = 10.5 m	84,5
Regeneration	14 000	BxLxH = 5.5mx12mx14.075m	unknown

The flow rate of gas is slightly higher than for the reference gas (681 800 kg/hr for reference gas), but the difference is so small (2.5 %) that it is considered to have little impact on the sizing. For the reference gas the pressure is slightly below 65 bar, while the gas for the Kristin field is almost 85 bar. This will greatly affect the required thickness of the wall for the absorber, and thereby also the weight of the unit. Higher pressure also means that the density of the gas processed at the Kristin platform is greater than the reference gas, indicating that the volume flow is similar to or possibly lower than for the reference gas.

The dew point or water concentration for the dehydrated gas is not mentioned, but since the gas is to be transported by the Åsgard Transport pipeline, the dew point must be the same as for the Åsgard platform, which is -18^{o}C at 70 bar (Fredheim (2008)).

Unfortunately the data for the regeneration process is very superficial, but the flow rate is given to be even higher than for the TEG injection system. This indicates that a lower regeneration temperature, a higher regeneration pressure or less stripping gas is used. It is also possible that the simulated TEG amount is underestimated since HYSYS uses equilibrium calculations. The total pressure is higher for the Kristin gas and the initial water concentration is therefore lower than for the reference gas (assuming the gas is saturated), thus also pointing in the direction of underestimation from the simulation model.

Since the intermediate temperatures and flow rates for the regeneration process is unavailable, it is not possible to use the same procedures as for the TEG injection concept to estimate the sizes of the different units. Instead the total volume of the regeneration unit is used for comparison.

Absorber

Using the same procedure as for the TEG injection concept, the required wall thickness of the absorber is calculated. The steel used is assumed to be low-alloy steel with a stress allowance of

172.4 MPa (Arnold & Stewart (2008):p.321) and density of 7850 kg/m^3 (Elert (2004)). The absorber uses structured packing for the mass transfer zone, and assuming 4 theoretical trays and HETP of 1.2 m, the weight of the packing is estimated to 7720 kg (200 kg/m^3).

t_W = 83.2 mm, V_{tot} = 93.5 m^3 and M_{tot} = 90 080 kg

The large height of the absorber indicates that the inlet separator is integrated into the column, thus the weight and volume is already calculated.

7.3 Comparing TEG injection and absorber dehydration

Gathering the estimated values for the TEG injection process and the conventional absorption process gives the following table:

Table 7.5: Comparing TEG injection and conventional absorber dehydration

Conventional absorber process				TEG injection			
	Weight	Volume	Footprint		Weight	Volume	Footprint
Absorption	90 080 kg	94 m^3	8,9 m^2	Absorption	27 623 kg	31 m^3	27 m^2
Regeneration		929 m^3	66 m^2	Regeneration	11 609 kg	8,7 m^3	9,2 m^2
Total	90 080 kg	1 023 m^3	74,9 m^2	Total	39 232 kg	39,7 m^3	36,2 m^2

Starting with the absorption part of the process, it is observed from table 7.5 that the TEG injection concept is much better for both weight and volume of the process. The absorption part of the conventional absorber process is more than twice as heavy as the entire TEG injection system, weighing more than 90 tons. However the process pressure is 20 bar higher for the Kristin process than for the simulated case, thus affecting the required wall thicknesses and thereby also the weight.

The volume for the absorption part of the absorber process is almost three times as big as for the TEG injection process, totaling 94 m^3. Volume is however not a very good comparison criterion, as it is important where the volume is located. For the absorber process very little floor space is required for the volume (i.e. the footprint is small), while for the TEG injection process most of the volume is in contact with the floor. This results in the footprint for the TEG injection process being three times as big as for the absorber process, even though the volume for the absorber process is three times that of the injection process. Most of the footprint for the TEG injection process (about 50%) is required for the piping between the injection/extraction stages. These pipes can potentially be placed on top of each other, thus reducing the footprint to some extent. Also, the total volume of the absorption part of the absorber process is only made up from the absorber column, while the volume for the TEG injection process is distributed over multiple units. Thus it is possible to utilize smaller unoccupied volumes for the TEG injection process, whereas the absorber process requires a large unoccupied volume.

As previously noted the regeneration processes are difficult to compare, as very little data is given for the Kristin process. The size of the regeneration process is given on the form width times length times height, indicating that the entire process is mounted in one module. The height is given to be over 14 meters, but this is probably just for the regeneration column, and not for the entire module. Comparing the footprint for the two processes, it is found that the footprint of the regeneration unit for the absorber process is about 8 times as large as for the TEG injection process. This is because all utility equipment like ladders, pumps, instrumentation and gangways etc are included, while this is neglected for the simulated case. Looking at the flow rate of TEG into the regeneration unit, the flow rate for the absorber process is almost 50% greater than for the TEG injection system, thus implying that the regeneration unit also has to be larger. How much larger is however difficult to say from the obtained operational data.

When comparing the TEG injection process to the operational absorber unit from the Kristin platform, one must however keep in mind that all the values for the TEG injection process are estimated and simulated. Many important factor when designing a real unit is neglected, e.g. maintenance equipment, instruments, safety equipment and so forth. The operating conditions may also change over the lifetime of a facility, which must then be incorporated into the design.

Overview

Based on the literature for separators and the regeneration process for glycol presented in chapter 1 and 2 and an analysis of potential improvements in chapter 4.1, the rest of chapter 4 was used for suggesting and discussing improvements to the TEG injection process. Most of the improvements concerned the regeneration process, using hot stripping gas to extract water from the rich glycol instead of the method of vaporization as is used today. Unfortunately none of the suggested improvements were incorporated into the process, as a couple of major disadvantages were discovered. Firstly the required amount of stripping gas required to regenerate the glycol was much larger than what was first estimated, and secondly the loss of glycol was much too high to be acceptable. Thirdly the stripping gas was to be obtained by burning a portion of the natural gas and mixing the exhaust gases with atmospheric air. This would have introduced oxygen into the system, further increasing the glycol losses. Substituting the regular separators used for the two first extraction stages of the TEG injection process with Deliquidiser units from CDS and using the ProPure compact mixer for TEG injection was therefore the only improvements incorporated.

From simulations using the simulation model from chapter 5, the findings from my semester project (Bråthen (2007)) was again confirmed in chapter 6.1; the TEG injection concept requires more circulated TEG than absorber dehydration to obtain a given dryness. Of course this does not come as a very big surprise, as the absorber has more theoretical trays than the TEG injection process, but the difference was bigger than expected. E.g. to obtain 55 ppm water in the dehydrated gas using the maximum recommended regeneration temperature for TEG, the TEG injection process requires 50% more circulated TEG than the absorber process. Adding stripping gas to the regeneration column further worsens the comparison, and also increases the difference in equilibrium dryness for the two processes; the absorber is able to reach water contents as low as 5 ppm, while the TEG injection process strives to obtain 15 ppm. Comparing the HC emissions through the over head vapor from the regeneration column surprisingly revealed that there is very little difference between TEG injection and absorber dehydration.

Using MEG and DEG as absorbents for the TEG injection concept also revealed some interesting result. When regenerating the absorbents using the maximum recommended temperature, DEG was actually the best absorbents for flow rates lower than 5000 kg/hr, while TEG was the best for higher flow rates. Adding stripping gas however completely changed the situation, with MEG being the by far best absorbent for stripping gas flow rates higher than 10 kgmole/hr. DEG has the second best dehydration performance, while TEG is only able to obtain the highest water concentration. For TEG the flow rate of absorbent seems to be the limiting factor, as there is almost no difference in obtained water concentration when adding 30 kgmole/hr or 100 kgmole/hr of stripping gas. The reason why MEG is not used for dehydration is however clearly revealed when investigating the loss of absorbent. Even when using no stripping gas, the loss of MEG is 10 kg/hr for circulation rates as low as 1000 kg/hr. For comparison, the loss of DEG and TEG are 1.3 kg/hr and 0.6 kg/hr, respectively. Adding stripping gas further increases the loss, and for MEG the loss increases almost exponentially for stripping gas flow rates higher than 30 kgmole/hr.

Comparing the estimated design for the TEG injection concept to the operational TEG absorber process from the Kristin platform showed that the TEG injection concept was considerably smaller and more lightweight. The footprint is however larger, mainly because of the gas piping required between the injection stages and the conventional scrubber used for the last stage. Much weight and volume could therefore be saved by using an inline separator for the final separation stage as well. As noted in chapter 7, it is however difficult to interpret much from these comparisons, as none of the reasons for the design of the operational process is known. The unit could be sized for future increase in the gas load, e.g. if new gas fields were expected to be found in the region, new wells connected to the platform and so forth. Also, the flow rate of TEG is almost 50% higher for the operational absorber compared to TEG injection when handling about the same amount of gas, while the simulated results show that the absorber requires considerably less absorbent than TEG injection. This points in the direction of underestimation of absorbent circulation rate by the simulation model, or that underlying reasons requires the absorber on the Kristin platform to have this high circulation rate.

Conclusion

Based on the presented discussions the following conclusions are reached:

1) As long as the absorbent is regenerated using the conventional regeneration process, the TEG injection concept is not competitive with regular absorber dehydration.
2) When comparing weight and volume for the absorption part of the process, the TEG injection concept is much better than conventional absorber dehydration.
3) TEG injection is able to fulfill both pipeline- and cold process water content specifications.
4) Using injection of hot stripping gas for glycol regeneration is unsuited, as the amount of stripping gas required is too large, the glycol losses to high, and the stripping gas cannot be supplied by mixing hot exhaust gasses with atmospheric air because of oxidizing of glycol.
5) TEG is considered a better absorbent for the injection concept over DEG as TEG requires less energy for regeneration, has lower loss of absorbent, and has the best dehydration performance for low to moderate flow rates of stripping gas (See table 6.2).
6) MEG is unsuited for dehydration at ambient temperatures because of the large glycol loss with the dehydrated gas.

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Appendices

Appendix A: Separator design

Important factors influencing separator design includes (Arnold & Stewart (2008) [1], Guo & Ghalambor (2005) [2]):

- Minimum, average and peak flow rates for both gas and liquid
- Operating and design pressures and temperatures
- Composition and physical properties of the fluids
- Designed degree of separation
- If slugs can develop in the feed stream
- Presence of impurities
- Foaming
- Corrosion from the gas or liquids
- Available space at operation location
- Weight limitations at operation location

After establishing the above factors, a preliminary design can be made based on engineering models, standards and most of all; engineering experience. This appendix will give a short introduction to a set of models enabling preliminary designs for horizontal and vertical separators. Centrifugal separators are not covered, as they require very detailed design including CFD-analysis.

A.1 Falling velocity

The first decision when designing a separator, is choosing either a horizontal or a vertical configuration (or other special configuration, e.g. centrifugal). As earlier mentioned a horizontal configuration is generally cheaper, and it is therefore commonly regarded as the design basis for most projects. Conditions may however dictate the use of vertical separators instead, e.g. space limitations, problems with solid deposition or slug flow.

Regardless of orientation of the separator, the falling velocity of the smallest droplets separated in the gravity settling section must be determined. The falling velocity (also referred to as terminal velocity) (U_t) will dictate the required minimum diameter for the separator, and can be determined from the semi-empirical formula A.1, known as the Sounders-Brown equation (Asheim & Herfjord (1984):p.5).

$$U_t = K_1 \cdot \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{\frac{1}{2}} \qquad \text{(Formula A.1: Falling velocity)}$$

Where K_1 is a semi-empirical constant based on the smallest droplet separated out (in addition to field experience) (normally in the range 0.035 – 0.051 m/s (Asheim & Herfjord (1984))), ρ_G is the gas density and ρ_L is the liquid density [kg/m^3]. Branan (2002):p.132 gives a chart for estimating K_1 based on the liquid to gas ratio.

The formula is developed by manipulating the equation for the different forces acting on a falling drop (drag and gravity), and by assuming that the drag force follows Stokes law. Stokes law assumes laminar flow, which is rarely the situation for the flow around a droplet in an oil/gas separator. This implies that the drag in reality is larger than estimated by Stokes law, making formula A.1 overestimate the falling velocity of the droplet (depending on the chosen value for K_1). Alternatively formula A.2 (Arnold & Stewart (2008):p.199, modified) can be used instead, incorporating the drag coefficient (C_D), the acceleration of gravity (g) and the diameter of the smallest droplet to be separated (d_m [m]). For a complete derivation of formula A.2, see Arnold & Stewart (2008) [1].

$$U_t = \left(\frac{4}{3} \cdot \frac{d_m \cdot g}{C_D} \right)^{\frac{1}{2}} \cdot \left(\frac{\rho_L - \rho_G}{\rho_G} \cdot \right)^{\frac{1}{2}}$$ (Formula A.2: Falling velocity)

A.2 Horizontal separators

Both horizontal and vertical separators must be sized for liquid- and gas capacity. Liquid capacity means that gas bubbles caught in the liquid must have sufficient time to escape to the vapor phase before the liquid leaves the vessel, while gas capacity means that liquid droplets must have sufficient time to fall to the liquid phase before the gas leaves the vessel. For gas capacity the falling velocity is important, dictating the inner diameter (D_i) and effective length (L_{eff}) of the vessel. Effective length is the length of the separator vessel available for separation, i.e. the entire length of the vessel except the inlet and exit sections (See figure A.1). By setting the retention time for the gas equal to the time it takes for a drop to fall from the top of the vessel to the liquid interface (the longest distance possible) and assuming that the vessel is half full of liquid, formula A.3 can be derived (See appendix B):

$$L_{eff} \cdot D_i = \frac{4}{\pi} \cdot \frac{Q_G}{U_t}$$ (Formula A.3: Gas capacity constraint)

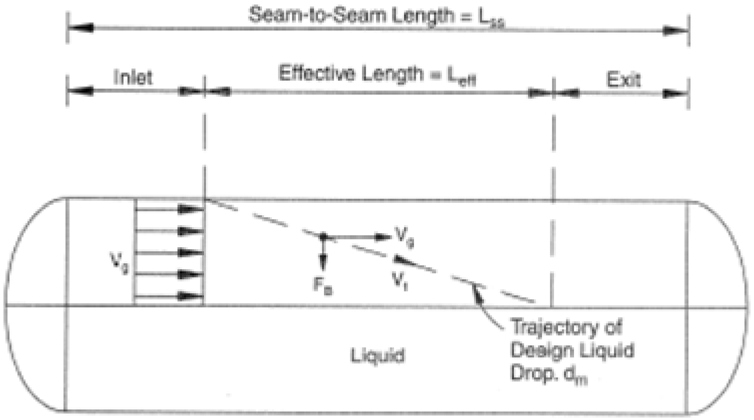

Figure A.1: The different sizes for a horizontal separator (adapted from Arnold & Stewart (2008):p.211)

Where Q_G is the gas volume flow rate under operating conditions [m^3/s] and U_t the falling velocity from formula A.1 or formula A.2. When determining the falling velocity for a horizontal separator, a K_1 value of 0.13 – 0.17 is recommended (depending on the length) (GPSA (2004) [1]). Note that the K_1 value is much higher than recommended in appendix A.1. This is from the previously described reason that there always will be a force acting towards the liquid interface for horizontal separators, regardless of the gas velocity. The falling velocity calculated using the above recommended K_1 value is therefore used to calculate the maximum velocity before carry-over can be anticipated.

By choosing a retention time for the liquid (t_L) (based on engineering experience), a known fluid flow rate (Q_L) and still assuming that the vessel is half full of liquid, formula A.4 can be derived (See appendix B):

$$L_{eff} \cdot D_i^2 = \frac{8}{\pi} \cdot Q_L \cdot t_L \qquad \text{(Formula A.4: Liquid capacity constraint)}$$

As can be seen from formula A.3 and formula A.4, several combinations of L_{eff} and D_i will fulfill the requirements for gas- and liquid capacity. There is however a general design rule, that the ratio of L_{eff} to D_i ("slenderness ratio") should be between 2.5 and 5 for horizontal separators (GPSA (2004) [1]). The smaller the diameter, the lower the cost and the weight of the vessel (Arnold & Stewart (2008) [1]), so it is beneficial to use the biggest slenderness ratio possible. Greater slenderness ratio than 4 are used at the risk of re-entraining liquid in the gas, and should therefore be thoroughly checked.

$$Slenderness\ ratio \equiv \frac{L_{eff}}{D_i} \in \left[2.5, 5.0\right] \qquad \text{(Formula A.5: Slenderness ratio)}$$

After determining suitable values for L_{eff} and D_i based on formula A.3, A.4 and A.5, the total length of the vessel can be determined from the formula A.6 (Arnold & Stewart (2008):p.212):

$$L = \begin{cases} L_{eff} + D_i, \text{ if design is governed by gas capacity} \\ \frac{4}{3} \cdot L_{eff}, \text{ if design is governed by liquid capacity} \end{cases} \qquad \text{(Formula A.6: Total length)}$$

The majority of the two-phase separators operated in the oil and gas industry is operated approximately half filled with liquid (Arnold & Stewart (2008) [1]). Sizing a separator other than half full requires a more complex expression for the gas capacity, in addition to more empirical constants. This is therefore omitted in this report, and those interested are referred to e.g. Arnold & Stewart (2008) [1].

Branan (2002):p.132 presents an alternative method for sizing horizontal separators by multiplying K_1 for formula A.1 with a factor of 1.25, and then use the same procedure as for vertical separators as presented below. This however, may lead to unnecessary large diameters, and consequently higher costs. For a preliminary design it may although still be good enough.

Guo & Ghalambor (2005):p.101 also employs formula A.1 for determining the falling velocity, but gives a table for K_1 based on the configuration of the separator. Based on calculated liquid settling volume and the falling velocity, suitable combinations for total length and inner diameter are given in a set of tables for both vertical and horizontal separators. The maximum available diameter in the table is however only 60 inches (1.5 m), which limits the use to systems with moderate to low gas flow.

A.3 Vertical separators

For vertical separators, the gas flows in the opposite direction of the force of gravity, thereby limiting the gas velocity as previously stated. The maximum allowable gas velocity is therefore equal to the falling velocity of the smallest droplets to be separated out in the gravity settling section derived previously from formula A.1 or A.2. Assuming that the design velocity equals the falling velocity and using the given gas flow rate, formula A.7 can be obtained (See appendix B):

$$D_i = \left(\frac{4}{\pi} \cdot \frac{Q_G}{U_t} \right)^{\frac{1}{2}}$$ (Formula A.7: Gas capacity constraint)

By selecting a settling time for the liquid (t_L) (based on engineering knowledge) and a known liquid flow rate (Q_L), the following formula can be derived (See appendix B):

$$h_L \cdot D_i^2 = \frac{4}{\pi} \cdot Q_L \cdot t_L$$ (Formula A.8: Liquid capacity constraint)

Where h_L is the height of the liquid volume [m].

As for horizontal separators, there are combinations of parameters fulfilling formula A.7 and A.8, but the restriction on the slenderness ratio applies to vertical separators as well. Slenderness ratio is normally no greater than 4 to keep the liquid height to a reasonable level, and ratios between 3 and 4 are normal (Arnold & Stewart (2008) [1]).

After determining the inner diameter and the liquid height, the total height of the vessel can be determined from formula A.9 (Arnold & Stewart (2008):p.224, modified):

$$L = \begin{cases} h_L + 2, \ h_L \leq 0.2 \\ h_L + D_i + 1, \ h_L \geq 0.2 \end{cases}$$ (Formula A.9: Total length)

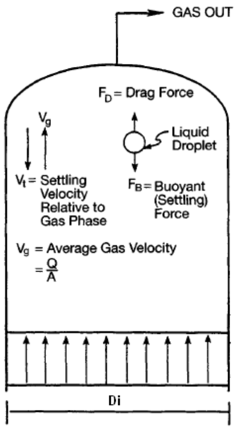

Figure A.2: Forces acting upon a falling droplet in a vertical separator (adapted from Arnold & Stewart (1998):p.127)

A.4 Special considerations for designing scrubbers

As earlier mentioned, scrubbers are separators designed to remove only small quantities of liquid, where the gas to liquid ratio is very high. Because of space limitations vertical configurations are most commonly employed, but horizontal designs can also be used. As the liquid quantities are small, the bulk separation in the gravity settling section becomes less important, and the scrubber can be designed with a higher internal gas velocity than a separator. Austrheim (2006):p.30 presents a discussion for choosing appropriate values for K_1 in the Sounders-Brown equation (formula A.1) for scrubbers, arguing that K_1 should be between 0.1 and 0.3 m/s depending of the scrubber internals. For demisting internals (e.g. mist extractors) values below 0.15 m/s is recommended, while additional equipment e.g. cyclones or vane packs downstream of the demisting section could increase the K_1 value towards 0.3 m/s. It should also be noted that pressure and feed stream composition affects the maximum allowable K_1.

A.5 Three phase separators

Design of a three-phase separator is slightly different than a two-phase separator. The equations for gas/liquid separation are the same (formula A.3 and A.7), but the separation of liquids require additional calculations. It can be shown that the flow around a settling drop of liquid in another liquid is laminar (Arnold & Stewart (2008) [3]), and thus Stokes` law govern the drag forces. The falling velocity of a liquid drop (L1) in another liquid (L2) can then be calculated from (Arnold & Stewart (2008):p.262):

$$U_{t,L1/2} = \frac{5.56 \cdot 10^5 \cdot \Delta SG \cdot d_m{}^2}{\mu_{L2/1}}, \Delta SG = \left(\frac{\rho_{L1}}{\rho_W} - \frac{\rho_{L2}}{\rho_W} \right)$$ (Formula A.10: Falling velocity liquid/liquid)

Where ΔSG is the difference in specific gravity for the two liquids, d_m the diameter of the droplet [m] and $\mu_{L1/2}$ the dynamic viscosity of the continuous liquid (cP). Liquid L1 is assumed to be the heaviest liquid.

It is difficult to predict the droplet size to be separated out without conducting small scale experiments for the same separator design. This is however expensive and not always possible, so Arnold & Stewart (2008) [3] recommends using d_m = 500 micron for oil/water separation and d_m = 200 micron for glycol/HC separation in absence of experimental data.

A.5.1 Horizontal three-phase separators

The design criteria for liquid capacity can be expressed for a horizontal three-phase separator, assuming 50% filled with liquid, as (See appendix B for derivation):

$$L_{eff} \cdot D_i^2 = \frac{8}{\pi} \cdot \left(Q_{L1} \cdot t_{L1} + Q_{L2} \cdot t_{L2} \right)$$ (Formula A.11: Liquid capacity, horizontal 3-phase)

In addition the height of the different liquid layers must be established, to ensure sufficient time for entrained liquids to settle out. The height can be calculated from formula A.12 (See appendix B for derivation).

$$h_{L1/2,\max} = U_{t,L2/1} \cdot t_{L1/2}$$ (Formula A.12: Maximum height of liquid layer)

Where $h_{L,\max}$ is the maximum height of each liquid layer [m] for the entrained droplets to be separated, that is, the heavy liquid droplets must have time to sink to the liquid/liquid interface, and the light liquid droplets must have time to rise, before the liquid leaves the vessel.

Total length and slenderness ratio are calculated as for two-phase horizontal separators, see appendix A.2, but slenderness ratios as high as 5.0 are common (Arnold & Stewart (2008) [3]).

For designing three-phase separators other than half full of liquid, see Arnold & Stewart (2008) [3].

A.5.2 Vertical three-phase separators

The settling constraint for a three-phase separator is based on the requirement that the heavy liquid must have sufficient time to fall to the liquid/liquid interface, and the light liquid must have sufficient time to rise (Formula A.13, see appendix B for derivation).

$$D_i = \left(\frac{4}{\pi} \cdot \frac{Q_{L1/2}}{U_{t,L1/2}} \right)^{\frac{1}{2}}$$ (Formula A.13: Settling constraint)

Formula A.13 gives the minimum diameter required to settle liquid 1 from liquid 2 and vice versa, and must be fulfilled for both liquids. In addition the height of the different liquid layers must be established to later calculate the total length of the separator (See appendix B for derivation):

$$h_L \cdot D_i^2 = \frac{4}{\pi} \cdot \left(Q_{L1} \cdot t_{L1} + Q_{L2} \cdot t_{L2} \right)$$ (Formula A.14: Liquid capacity constraint)

Total length and slenderness ratio are calculated using the same procedure as for vertical two-phase separators. Common slenderness ratios are between 1.5 and 3.0, and for separators governed by the liquid capacity ratios as high as 4.0 can be used (Arnold & Stewart (2008) [3]).

A.6 Shell thickness and choice of material

To determine the shell thickness and the material required, the operating conditions must be established. This includes the maximum and minimum temperatures, the operating pressure and the maximum allowable working pressure (MAWP). The MAWP will govern the settings for the safety pressure relief valve, and should not be set to close to the operating pressure, as this can lead to premature opening of the safety valve (Arnold & Stewart (2008) [2]). It is very important that the MAWP for the separator doesn’t exceed the MAWP for the equipment attached (e.g. valves, nozzles and pipes).

The MAWP and the operating temperature range determine the choice of material. This choice is governed by industrial standards depending on operation location. For USA this standard is the ASME code (American Society of Mechanical Engineers` Boiler and Pressure Vessel Code). Part of the ASME code section VIII for design of pressure vessels is given by Arnold & Stewart (2008) [2]

The ASME code section VII is divided into two divisions; Division 1 governing the design by rules and division 2 governing the design by analysis. Division 1 is less stringent in terms of certain design details and inspection procedures, and therefore incorporates a larger safety factor (a factor of 3.5 for the 2007 edition). Division 2 requires a more thorough analysis and a safety factor of 3 (for the 2007 edition) is therefore considered sufficient (Arnold & Stewart (2008) [2]). See e.g. Arnold & Stewart (2008) [2] for maximum allowable stress for common steels for both division 1 and division 2.

For determining the wall thickness in accordance to the ASME code Section VIII Division 1, the following formula is used for a cylindrical shell (Moss (2004):p.16):

$$t_{W,\min} = \frac{P_{MAWP} \cdot D_i}{2 \cdot \sigma_{\max} \cdot \eta_j - 1.2 \cdot P_{MAWP}}$$ (Formula A.15: Minimum wall thickness)

Where σ_{max} is the maximum allowable stress for the given material [N/m^2], η_j is the joint efficiency ([0,1]) (See Arnold & Stewart (2008):p.325 for table), P_{MAWP} is the maximum allowable working pressure [Pa], D_i the inner diameter [m] and $t_{W,min}$ the minimum wall thickness [m].

In addition the thickness of the heads (top and bottom) of the vessel must be calculated. Ellipsoidal heads are the most common (Arnold & Stewart (2008):p.320) and their thickness is calculated from:

$$t_{W,\min} = \frac{P_{MAWP} \cdot D_i}{2 \cdot \sigma_{\max} \cdot \eta_j - 0.2 \cdot P_{MAWP}}$$ (Formula A.16: Minimum wall thickness for ellipsoidal head)

These equations are valid for cylindrical vessels for $t_{W,min} < 0.25D_i$ and $P_{MAWP} < 0.385 * \sigma_{max} * \eta_j$ (Moss (2004)).

To account for possible corrosion, a safety margin of 0.125 inches (3.2 mm) for non-corrosive service and 0.250 inches (6.4 mm) for corrosive service is added to the wall thickness resulting from formula A.15 and A.16 (Arnold & Stewart (2008) [2]). Also, to ensure rigidity of the vessel, a minimum wall thickness of 6.4 mm for vessels with diameter up to 1.07 m, 8.1 mm for diameters up to 1.52 m and 9.7 mm for greater diameters is recommended (Coker (2007)).

Appendix B: Derivation of functions for sizing separators

The falling velocity (U_t and $U_{t,L}$) used in some of the formulas, can be obtained from one of the below formulas (Asheim & Herfjord (1984):p.5, Arnold & Stewart (2008):p.196,262):

$$U_t = K_1 \cdot \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{\frac{1}{2}}, K_1 \in \left[0.035, 0.051 \right] \qquad U_t = \left(\frac{4}{3} \cdot \frac{d_m \cdot g}{C_D} \right)^{\frac{1}{2}} \cdot \left(\frac{\rho_L - \rho_G \cdot}{\rho_G} \right)^{\frac{1}{2}}$$

$$U_{t,L1/2} = \frac{5.56 \cdot 10^5 \cdot \Delta SG \cdot d_m{}^2}{\mu_{L2/1}}, \Delta SG = \left(\frac{\rho_{L1}}{\rho_W} - \frac{\rho_{L2}}{\rho_W} \right)$$

Formula A.3: Gas capacity constraint for horizontal two-phase separators:

Developing an expression for the effective length (L_{eff}) and inner diameter (D_i), by setting the retention time for the gas equal to the time it takes for a drop to fall from the top of the separator to the liquid interface.

Assumptions:

- Vessel is half full of liquid
- Known Q_G and U_t

Gas retention time: $t_G = \frac{L_{eff}}{U_G}$ Falling time: $t_f = \frac{\frac{D_i}{2}}{U_t}$ Area available for gas: $A_G = \frac{1}{2} \cdot \frac{\pi}{4} \cdot D_i^2$

Gas velocity: $U_G = \frac{Q_G}{A_G} = \frac{Q_G}{\frac{\pi}{8} \cdot D_i^2}$

$$t_G = t_f \Rightarrow \frac{L_{eff}}{U_G} = \frac{D_i}{2 \cdot U_t} \Rightarrow \frac{\pi \cdot D_i^2 \cdot L_{eff}}{8 \cdot Q_G} = \frac{D_i}{2 \cdot U_t} \Rightarrow L_{eff} \cdot D_i = \frac{4}{\pi} \cdot \frac{Q_G}{U_t}$$

Formula A.4: Liquid capacity constraint for horizontal two-phase separators:

Choosing a liquid retention time and using the expression for retention time to develop an expression for the liquid capacity of a horizontal separator.

Assumptions:

- Vessel is half full of liquid
- Known Q_L and t_L

Liquid retention time: $t_L = \frac{V_L}{Q_L}$ Volume available for liquid: $V_L = \frac{1}{2} \cdot \frac{\pi}{4} \cdot D_i^2 \cdot L_{eff}$

$$t_L = \frac{V_L}{Q_L} = \frac{\pi \cdot D_i^2 \cdot L_{eff}}{8 \cdot Q_L} \Rightarrow L_{eff} \cdot D_i^2 = \frac{8}{\pi} \cdot Q_L \cdot t_L$$

Formula A.7: Gas capacity constraint for vertical two-phase separators:

Setting the gas velocity equal to the falling velocity of the smallest droplet to be separated in the gravity settling section, and derive a function for the minimum required diameter of a vertical separator.

Assumptions:

- $U_G = U_t$
- Known Q_G and U_t

Gas velocity: $U_G = \frac{Q_G}{A_G}$ Area available for gas: $A_G = \frac{\pi}{4} \cdot D_i^2$

$$U_G = U_t \Rightarrow U_t = \frac{Q_G}{A_G} = \frac{4}{\pi} \cdot \frac{Q_G}{D_i^2} \Rightarrow D_i = \left(\frac{4}{\pi} \cdot \frac{Q_G}{U_t} \right)^{\frac{1}{2}}$$

Formula A.8: Liquid capacity constraint for vertical two-phase separators:

Choosing a liquid retention time and deriving an expression for the liquid height (h_L) and inner diameter (D_i) by using the expression for retention time.

Assumptions:

- Known Q_L and t_L

Liquid retention time: $t_L = \frac{V_L}{Q_L}$ Volume available for liquid: $V_L = \frac{\pi}{4} \cdot D_i^2 \cdot h_L$

$$t_L = \frac{V_L}{Q_L} = \frac{\pi \cdot D_i^2 \cdot h_L}{4 \cdot Q_L} \Rightarrow h_L \cdot D_i^2 = \frac{4}{\pi} \cdot Q_L \cdot t_L$$

Formula A.11: Liquid capacity constraint for horizontal three-phase separator

Assumptions:

- Known $Q_{L1/2}$ and $t_{L1/2}$
- Vessel 50% full of liquid

Settling time: $t_{L1/2} = \frac{V_{L1/2}}{Q_{L1/2}}$ Volume available for each liquid: $V_{L1/2} = \frac{1}{2} \cdot \frac{\pi}{4} \cdot L_{eff} \cdot D_i^2 \cdot \frac{A_{L1/2}}{A_L}$

Total liquid area: $A_L = A_{L1} + A_{L2}$

$$t_{L1} = \frac{V_{L1}}{Q_{L1}} = \frac{\frac{\pi}{8} \cdot L_{eff} \cdot D_i^2 \cdot \frac{A_{L1}}{A_L}}{Q_{L1}}, \qquad t_{L2} = \frac{V_{L2}}{Q_{L2}} = \frac{\frac{\pi}{8} \cdot L_{eff} \cdot D_i^2 \cdot \frac{A_{L2}}{A_L}}{Q_{L2}}$$

Adding the two above equations gives:

$$t_{L1}\cdot Q_{L1}+t_{L2}\cdot Q_{L2}=\frac{\pi}{8}\cdot L_{eff}\cdot D_i^2\cdot\frac{A_{L1}+A_{L2}}{A_L}\Rightarrow L_{eff}\cdot D_i^2=\frac{8}{\pi}\cdot\left(t_{L1}\cdot Q_{L1}+t_{L2}\cdot Q_{L2}\right)$$

Formula A.12: Maximum height of liquid level in horizontal three-phase separator

Assumptions:

- Known $U_{t,L1/2}$ and $t_{L1/2}$

Retention time: $t_{L1}=\frac{h_{L1,\max}}{U_{t,L2}},\quad t_{L2}=\frac{h_{L2,\max}}{U_{t,L1}}$

Rearranging: $h_{L1,\max}=U_{t,L2}\cdot t_{L1},\quad h_{L2,\max}=U_{t,L1}\cdot t_{L2}$

Formula A.13: Settling constraint for vertical three-phase separator

Assumptions:

- Known $Q_{L1/2}$ and $U_{t,L1/2}$
- $U_{L1}=U_{t,L1}$ and $U_{L2}=U_{t,L2}$

Liquid velocity: $U_{L1/2}=\frac{Q_{L1/2}}{A_L}$ Cross section area: $A_L=\frac{\pi}{4}\cdot D_i^2$

$$U_{L1/2}=U_{t,L1/2}\Rightarrow U_{t,L1/2}=\frac{Q_{L1/2}}{A_L}=\frac{Q_{L1/2}}{\frac{\pi}{4}\cdot D_i^2}\Rightarrow D_i=\left(\frac{4}{\pi}\cdot\frac{Q_{L1/2}}{U_{t,L1/2}}\right)^{\frac{1}{2}}$$

Formula A.14: Liquid capacity constraint for vertical three-phase separator

Assumptions:

- Known $Q_{L1/2}$ and $t_{L1/2}$

Volume of each liquid layer: $V_{L1}=\frac{\pi}{4}\cdot h_{L1}\cdot D_i^2,\quad V_{L2}=\frac{\pi}{4}\cdot h_{L2}\cdot D_i^2$

Retention time for each liquid: $t_{L1}=\frac{V_{L1}}{Q_{L1}},\ t_{L2}=\frac{V_{L2}}{Q_{L2}}\Rightarrow V_{L1}=Q_{L1}\cdot t_{L1},\quad V_{L2}=Q_{L2}\cdot t_{L2}$

$$V_{L1}=V_{L1},\quad V_{L2}=V_{L2}\Rightarrow Q_{L1}\cdot t_{L1}=\frac{\pi}{4}\cdot h_{L1}\cdot D_i^2,\quad Q_{L2}\cdot t_{L2}=\frac{\pi}{4}\cdot h_{L2}\cdot D_i^2$$

$$V_L=V_{L1}+V_{L2}\Rightarrow Q_{L1}\cdot t_{L1}+Q_{L2}\cdot t_{L2}=\frac{\pi}{4}\cdot D_i^2\cdot\left(h_{L1}+h_{L2}\right)\Rightarrow h_L\cdot D_i^2=\frac{4}{\pi}\cdot\left(Q_{L1}\cdot t_{L1}+Q_{L2}\cdot t_{L2}\right)$$

Appendix C: Design of a glycol regeneration unit

As for separator design, a lot of factors influence the design of the glycol regeneration system. In addition to the factors mentioned in appendix A, other important factors are:

- Choice of glycol
- Flow rate of glycol
- Required purity of lean glycol
- Ambient temperature and pressure

C.1 Overhead condenser

The size of the overhead condenser is dependent on the amount of reflux required for the regeneration still, the flow rate of glycol, and the chosen design.

First of all, the outlet temperature of the rich glycol must be calculated (See appendix D for derivation):

$$T_{gl,o} = T_{gl,i} + \frac{m_{reflux}}{m_{gl} \cdot c_{gl}} \cdot \left[c_{p,G} \cdot \left(T_G - T_{SAT} \right) + h_{fg,G} \right] \qquad \text{(Formula C.1)}$$

Where $T_{gl,o}$ is the rich glycol outlet temperature [K], $T_{gl,I}$ the rich glycol inlet temperature [K], m_{reflux} the mass flow of reflux [kg/s], m_{gl} the mass flow of rich glycol [kg/s], c_{gl} the heat capacity of the rich glycol [J/kg*K], $c_{p,G}$ the heat capacity for the gas in the regeneration column at constant pressure [J/kg*K], T_G the temperature of the gas in near the overhead condenser [K], T_{SAT} the saturation temperature for the gas at pressure near the overhead condenser [K], and $h_{fg,G}$ the heat of vaporization/condensation for the gas in the regeneration column near the overhead condenser [J/kg]. Arnold & Stewart (1999) [1] recommends designing for glycol outlet temperatures in the range 353 – 368 K (80 – 95^{o}C). Typical values for required reflux is between 25 and 50% of the water evaporated from the glycol (Kohl & Nielsen (1997) [1], Arnold & Stewart (1999) [1])

After determining the rich glycol outlet temperature, the total heat transfer surface required (A_{HT}) can be estimated from formula C.2 (see appendix D for derivation):

$$A_{HT} = \frac{m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right)}{U \cdot \left[T_{SAT} - \frac{1}{2} \cdot \left(T_{gl,o} - T_{gl,i} \right) \right]} \qquad \text{(Formula C.2: Required area for heat transfer)}$$

Where U is the overall heat transfer coefficient for the overhead condenser [W/m^2*K].

The overall heat transfer coefficient (U) is dependent on both the fluids involved, the geometry of the heat exchanger, the thickness of the walls, the flow regime in the channels/tubes and a lot of other factors. For initial design it can however be estimated from tables, e.g. Branan (2002):p.34. Arnold & Stewart (1999) [1] suggests using U = 568 W/m^{2}K, but this seems conservative compared to Branan (2002):p.34.

Using the calculated heat transfer area (A_{HT}) and choosing the design of the heat exchanger (e.g. plate/fin, shell and tube etc.), the overall size of the heat exchanger can be estimated. For a more thorough design procedure and an introduction to heat exchangers, Incropera & DeWitt (2002) is recommended.

Note: Sizing a heat exchanger is an iterative process, as several of the factors can be varied. For example a portion of the glycol stream can be routed past the heat exchanger, resulting in higher outlet temperature of the glycol and a smaller heat transfer area. At the same time, the pressure of the regeneration still can be changed, altering the saturation temperature and the heat capacity of the gas, giving yet another result.

C.2 Flash tank

The flash tank is sized using the same procedures as described for separators in appendix A. Depending on if there are liquid HC present or not, a three- or two-phase separator design is employed respectively.

C.3 Glycol/glycol heat exchanger

The glycol/glycol heat exchanger must be calculated a little different than the overhead condenser, as there is no condensation, just two liquids exchanging heat. Assuming the outlet temperature of the rich glycol is the unknown, formula C.3 can be derived (see appendix D):

$$T_{gl,r,o} = T_{gl,r,i} + \frac{m_{gl,l} \cdot c_{gl,l}}{m_{gl,r} \cdot c_{gl,r}} \cdot \left(T_{gl,l,i} - T_{gl,l,o}\right) \qquad \text{(Formula C.3: Outlet temperature glycol/glycol HX)}$$

Assuming countercurrent flow, the total required heat transfer area can then be calculated from (Incropera & DeWitt (2002):p.650):

$$A_{HT} = \frac{m_{gl,l} \cdot c_{gl,l} \cdot \left(T_{gl,l,i} - T_{gl,l,o}\right)}{U \cdot LMTD}, \quad LMTD = \frac{\left(T_{gl,l,o} - T_{gl,r,i}\right) - \left(T_{gl,l,i} - T_{gl,r,o}\right)}{\ln\left(\dfrac{T_{gl,l,o} - T_{gl,r,i}}{T_{gl,l,i} - T_{gl,r,o}}\right)} \qquad \text{(Formula C.4)}$$

As mentioned earlier for the overhead condenser, the calculated required surface area for heat transfer together with the chosen design enables the total size of the heat exchanger to be estimated. This design is also considered an iterative process, as the mass flow of glycol (both rich and lean), inlet temperatures and outlet temperature for the lean glycol can be varied. Arnold & Stewart (1999) [1] recommends outlet temperature of the rich glycol to be in the range 408 – 423 K (135 – 150°C). In addition it is important that the outlet temperature of the lean glycol is not too high, as this can damage the glycol pump.

C.4 Stripping column

The first decision to be made in designing a stripping column, is whether to use trays or packing for the mass transfer zone. Trays are typically used for large units handling clean and trouble free fluids (i.e. noncorrosive, nonfoaming etc.), while packing is used for smaller units handling troublesome fluids (See chapter 2.1 and 2.2 for a more thorough discussion). Packing is also preferred for operating locations with limited space (e.g. offshore platforms) or moving operation (e.g. floating operation facilities).

Determining the column diameter can be done by several means, one of which is to use the same procedure as for separator sizing presented in appendix A. For trayed columns with 24 inch tray spacing (0.61 m), Guo & Ghalambor (2005) [2] recommends using K_1= 0.05 m/s, while Kohl & Nielsen (1997) [2] gives different values for K_1 based on the liquid to gas ratio and tray spacing. After determining the falling velocity (U_t) from formula A.1 or A.2, the required diameter can be calculated from formula A.7. When calculating the required diameter, it is important to use the largest gas flow for the column, usually found at the bottom near the reboiler (Kohl & Nielsen (1997) [1]). If stripping gas is used in the stripping column, this must also be taken into account when calculating the required gas capacity.

For packed columns, the Sounders-Brown correlation from appendix A can also be used, but now using K_1= 0.1 m/s (Kohl & Nielsen (1997) [1], GPSA (2004) [2]). Alternatively a more detailed procedure is described by Jones & Pujado (1997), calculating the diameter using the flooding criteria for the tower.

In determining the height of the column, the number of actual trays or feet of packing is required. Actual trays are calculated using the number of theoretical trays (equilibrium is obtained at every theoretical tray) and dividing by the tray efficiency (See formula C.5). Tray efficiency can be estimated using graphical techniques or empirical correlations, see Kohl & Nielsen (1997) [2] or Øi (2003) for an introduction. For conservative design, the efficiency can be estimated as 25% (Arnold & Stewart (1999) [1]). Height required between the trays vary with the tower diameter and flow conditions, and general design guidelines are given by Kohl & Nielsen (1997) [2].Lyons & Plisga (2005) recommends using tray spacing of 55 – 60 cm regardless of column diameter and types of trays.

$$N_A = \frac{N_{TH}}{\eta_t} \qquad \text{(Formula C.5: Actual number of trays)}$$

The number of theoretical trays can be found through analysis of the components of the rich glycol, in addition to operating conditions, usually resulting in 2 – 3 theoretical trays for a stripping column for glycol (Kohl & Nielsen (1997) [1], Arnold & Stewart (1999) [1]). As previously noted many glycol regeneration columns are however designed with an excess of trays to minimize the glycol loss through the overhead vapor (Kohl & Nielsen (1997) [1]).

Feet of packing is found using the HETP ("Height equivalent to theoretical plate") and multiplying by the number of theoretical trays. HETP is difficult to predict accurately, and empirical formulas and engineering experience is often applied (Kohl & Nielsen (1997) [2]). Values between 3.7 and 5.0 ft (1.13 – 1.53 m) are common, but values as high as 8 ft (2.44 m) is also experienced (Kohl & Nielsen (1997) [1], Arnold & Stewart (1999) [1]). Branan (2002):p.87 presents a table for HETP for different types of commercial packing elements.

In addition to allow space for packing or trays, inlet- and exit facilities must also be incorporated. This depends strongly on tower design; if overhead condenser and reboiler are integrated in the column, etc.

C.5 Reboiler

The most common correlation for estimating required reboiler heating duty, is the one proposed by Sivalls (1976) (Cited by Kohl & Nielsen (1997):p.985), see formula C.6 (converted to metric units).

$$q_{req} = 557.43 \cdot Q_{gl}$$ (Formula C.6: Required heat for reboiler)

Where q_{req} is in kilowatts [kW] and Q_{gl} in [l/s].

GPSA (2004) [2] claims that a factor of 418.07 in formula C.6 already overestimates the required heat duty. This is however very dependent on the operating conditions, e.g. water content of rich glycol, rate of reflux, heat loss from the regeneration column etc.

For sizing a kettle type reboiler the same procedure as for sizing two phase horizontal separators presented in appendix A.2 can be used by assuming a liquid retention time.

Appendix D: Derivation of equations for designing a TEG injection process

Formula C.1: Outlet temperature of rich glycol from overhead condenser

Assumptions:

- Known fluid properties
- Known mass flow
- Known vapor temperature near the overhead condenser in the regeneration still

$$q_{reflux} = m_{reflux} \cdot \left[c_{p,G} \cdot \left(T_G - T_{SAT} \right) + h_{fg,G} \right] \text{ , } q_{oc} = m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right)$$

$$q_{reflux} = q_{oc} \Rightarrow m_{reflux} \cdot \left[c_{p,G} \cdot \left(T_G - T_{SAT} \right) + h_{fg,G} \right] = m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right)$$

$$\Rightarrow T_{gl,o} = T_{gl,i} + \frac{m_{reflux} \cdot \left[c_{p,G} \cdot \left(T_G - T_{SAT} \right) + h_{fg,G} \right]}{m_{gl} \cdot c_{gl}}$$

Formula C.2: Required heat transfer area for overhead condenser

Assumptions:

- Known fluid properties, U
- Known temperatures

$$q_{oc} = U \cdot A_{HT} \cdot \left(\overline{T}_G - \overline{T}_{gl} \right), \ q_{oc} = m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right), \ \overline{T}_{gl} = \frac{T_{gl,o} + T_{gl,i}}{2}, \ \overline{T}_G = T_{SAT}$$

$$q_{oc} = q_{oc} \Rightarrow U \cdot A_{HT} \cdot \left(\overline{T}_G - \overline{T}_{gl} \right) = m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right)$$

$$\Rightarrow A_{HT} = \frac{m_{gl} \cdot c_{gl} \cdot \left(T_{gl,o} - T_{gl,i} \right)}{U \cdot \left(T_{SAT} - \frac{1}{2} \cdot \left(T_{gl,o} - T_{gl,i} \right) \right)}$$

Formula C.3: Outlet temperature of rich glycol from glycol/glycol heat exchanger

Assumptions:

$$q_{gl,r} = m_{gl,r} \cdot c_{gl,r} \cdot \left(T_{gl,o,r} - T_{gl,i,r} \right), \ q_{gl,l} = m_{gl,l} \cdot c_{gl,l} \cdot \left(T_{gl,o,l} - T_{gl,i,l} \right)$$

$$q_{gl,r} = q_{gl,l} \Rightarrow m_{gl,r} \cdot c_{gl,r} \cdot \left(T_{gl,o,r} - T_{gl,i,r} \right) = m_{gl,l} \cdot c_{gl,l} \cdot \left(T_{gl,o,l} - T_{gl,i,l} \right)$$

$$\Rightarrow T_{gl,r,o} = T_{gl,r,i} + \frac{m_{gl,l} \cdot c_{gl,l}}{m_{gl,r} \cdot c_{gl,r}} \cdot \left(T_{gl,l,i} - T_{gl,l,o} \right)$$

Appendix E: Process Flow Diagram for Drizo and Coldfinger processes

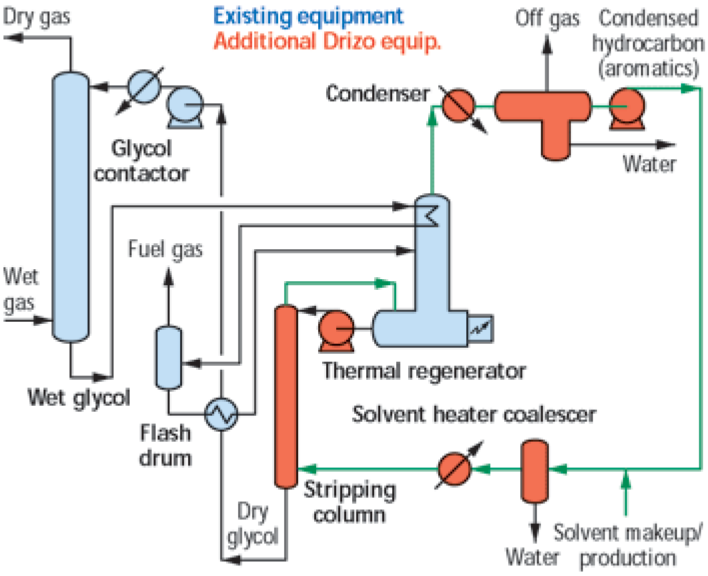

Figure E.1: PFD for glycol dehydration process with Drizo regeneration unit (adapted from World Oil Magazine (2003))

Figure E.1 shows the layout of the glycol dehydration process with a Drizo process added to the regeneration unit. The blue equipment is the basis equipment for a typical glycol dehydration process, while the red equipment is extra equipment needed for the Drizo process.

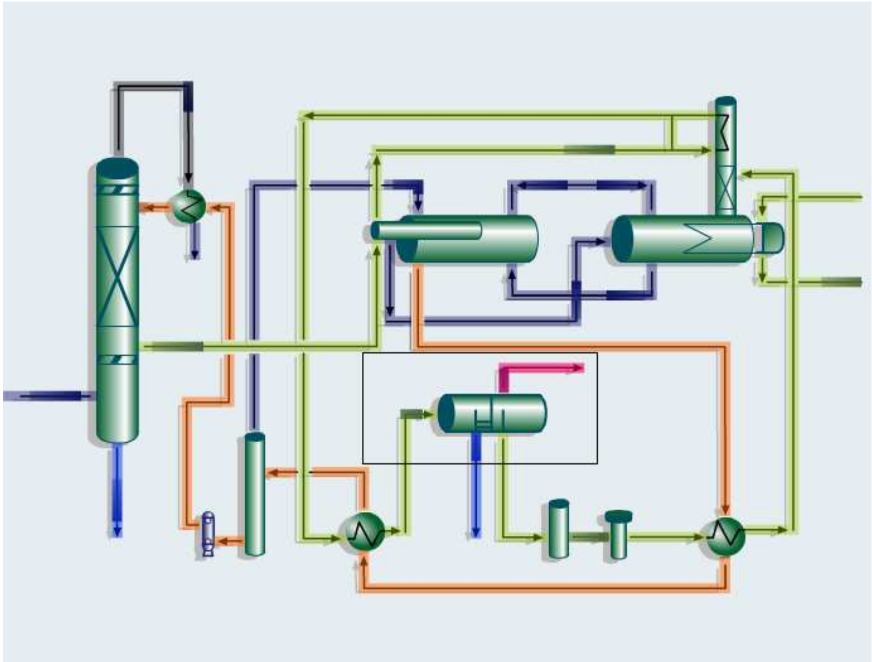

Figure E.2: PFD for glycol dehydration process with Coldfinger regeneration unit (adapted from Comart (undated))

Figure E.2 shows the layout of the glycol dehydration process with a Coldfinger process added to the regeneration unit. The Coldfinger unit is surrounded by a square; the red flow being heat transferred from the vessel to keep the Coldfinger cold, the blue stream being water and the green stream being glycol.

Appendix F: The Cubic-Plus-Association (CPA) Equation of State (EoS)

When simulating a process using simulation software like HYSYS, an equation for the different physical properties (enthalpy, density, entropy etc) is needed to predict the behavior of the components involved. Components interact differently depending on what mixture they are a part of, and the choice of equation is therefore subject to the chosen components. Mixtures of glycol, water and HC are especially hard to model, as both glycol and water are associating components. This appendix will give a short introduction to the CPA EoS, which is able to handle the glycol, water and HC mixture, and is used in this project.

F.1 The equations

The EoS describes the relation between pressure, volume and temperature for a given substance in either vapor or liquid phase, the simplest EoS being the ideal gas law (See e.g. Moran & Shapiro (1998):p.93). The CPA EoS is an extended ideal gas law, which incorporates three component constants and also accounts for associating components (See formula F.1 (Kontogeorgis et al. (2007):p.2)).

$$P = \frac{\bar{R}\cdot T}{\bar{v}-b} - \frac{\alpha(T)}{\bar{v}\cdot(\bar{v}+b)} - \frac{1}{2}\cdot\frac{\bar{R}\cdot T}{\bar{v}}\cdot\left(1+\bar{\rho}\cdot\frac{\partial \ln g(\bar{\rho})}{\partial\bar{\rho}}\right)\cdot\sum_i x_i\cdot\sum_{A_i}(1-X_{A_i}) \qquad \text{(Formula F.1)}$$

Where P is the total pressure [Pa], $\bar{R}$ is the universal gas constant [kJ/kmole*K], $\bar{v}$ the molar volume [m^3/kmole], b a component constant [m^3/kmole], α(T) the energy parameter (See formula F.4), $\bar{\rho}$ the molar density [kmole/m^3], g($\bar{\rho}$) the radial distribution function (See formula F.3), x_i the mole fraction of component i and X_{Ai} the fraction of sites belonging to molecule i that *do not* form bonds with other sites (See formula F.2). The last term in the equation is the association term, and will equal zero for all non-associating compounds (as the fraction of sites not bonded will equal 1.0). The EoS is then reduced to the well known Soave-Redlich-Kwong (SRK) EoS (Kontogeorgis et al. (2007)).

X_{Ai} is found from formula F.2 (Kontogeorgis et al. (2007):p.2):

$$X_{A_i} = \left(1+\bar{\rho}\cdot\sum_j x_j\cdot\sum_{B_j} X_{B_j}\cdot\Delta_{A_iB_j}\right)^{-1} \quad , \Delta_{A_iB_j} = g(\bar{\rho})\cdot\left[\exp\left(\frac{\varepsilon_{A_iB_j}}{\bar{R}\cdot T}\right)-1\right]\cdot\left(\frac{b_i+b_j}{2}\right)\cdot\beta_{A_iB_j}$$

Where Δ_{AiBj} is known as the association strength, ε_{AiBj} as the association energy and β_{AiBj} as the association volume.

The radial distribution function is defined from formula F.3 (Kontogeorgis et al. (2007):p.2):

$$g(\bar{\rho}) \equiv \frac{1}{1-1.9\cdot\eta}, \quad \eta = \frac{1}{4}\cdot b\cdot\bar{\rho} \qquad \text{(Formula F.3)}$$

Finally the energy parameter α(T) is given from formula F.4 (Kontogeorgis et al. (2007):p.2):

$$\alpha(T) = a_0\cdot\left[1+c\cdot\left(1-\sqrt{T_r}\right)\right]^2, \quad T_r = \frac{T}{T_{crit}} \qquad \text{(Formula F.4)}$$

Where a_0 and c are constants subject to the components involved, T_r the reduced temperature and T_{crit} the critical temperature.

In addition to the equation itself it is therefore necessary to know five parameters for each component involved (a_0, b, c, β_{AiBj} and ε_{AiBj}) to be able to solve the equation. These are obtained by fitting vapor pressure and liquid density data to experimental data. For non-associating compounds only the first three parameters are required, which can be obtained via critical properties and acentric factors (Kontogeorgis et al. (2007)).

F.2 Mixing rules

When simulating mixtures, mixing rules must be applied to obtain global values for the model constants. Several mixing rules have been developed, and which one to use is subject to the components involved. Different mixing rules are also employed for the different model constants. For the energy parameter the geometric mean rule is applied (Kontogeorgis et al. (2008):p.2):

$$a_{mix} = \sum_i \sum_j x_i \cdot x_j \cdot \sqrt{a_i \cdot a_j} \cdot (1 - k_{ij}) \qquad \text{(Formula F.5)}$$

Where k_{ij} is an interaction parameter [0,1].

For the b parameter regular mean is used (Formula F.6 (Kontogeorgis et al. (2008):p.2):

$$b_{mix} = \sum_i x_i b_i \qquad \text{(Formula F.6)}$$

If two associating components are involved (as is the case with water/glycol mixtures), mixing rules for the association energy and association volume is also needed. Again several are available, but the ones used for simulation in this project are the Elliot Combining Rule (ECR) rules (Kontogeorgis et al. (2008):p.3):

$$\varepsilon_{A_iB_j} = \frac{\varepsilon_{A_iB_i} + \varepsilon_{A_jB_j}}{2} \qquad \text{(Formula F.7)}$$

$$\beta_{A_iB_j} = \sqrt{\beta_{A_iB_i} \cdot \beta_{A_jB_j}} \cdot \frac{\sqrt{b_i \cdot b_j}}{b_{ij}} \qquad \text{(Formula F.8)}$$

F.3 Experience

The CPA EoS is a relatively new EoS first proposed by Kontogeorgis et al. (1996) from the National Technical University of Athens, Greece in 1996. It combines the physical term from the widely used SRK (Soave-Redlich-Kwong) EoS (See e.g. Soave (1971)) with the association term from the SAFT (Statistical Association Fluid Theory) EoS (See e.g. Wertheim (1986)). For non-associating components CPA reduces to SRK as previously noted, having the advantage that constants developed for the SRK EoS can be used in CPA for non-associating components. CPA has been extensively tested during the last ten years, and proven to be a very good EoS for modeling almost any component mixture (Voutsas et al. (2007)).

Further development of the CPA EoS is conducted by the research project CHIGP (Chemicals in Gas Processing) lead by Georgios Kontogeorgis at the Technical University of Denmark (DTU (2008)).

Appendix G: Introduction to Aspen HYSYS 2006

Aspen HYSYS is a simulation software package provided by the American AspenTech company. The company was founded in 1981 to commercialize technology developed as part of the Advanced System for Process Engineering (ASPEN) project at the Massachusetts Institute of Technology (MIT). Aspen HYSYS has been on the market for more than 20 years, and is one of the leading process modeling tools available today (AspenTech undated [2]).

The program presents a steady state solution for a given process, and can handle both static and dynamic input data. The equation solver runs continuously (as long as it is not stopped by the user), and new output values for changes in input values will therefore appear immediately. As the solution algorithm is designed to calculate both ways, output values can also be given to calculate input values. (AspenTech undated [1])

The process is modeled in a process flow diagram (PFD), using common process components like compressors, separators, heat exchangers, pumps, mixers and valves. Different options are available for each component, including heat and pressure loss, efficiency and working temperature and pressure, and are defined by the user. Detailed options are also available, for example design and layout, reactions within units and carry-over-calculations. The components are connected through material streams, modeling the transportation of mass from one unit to the next. If two process units are going to be very far apart in the physical process, a pipe module can be inserted to also model the pressure loss. If the unit delivers or requires energy, a corresponding energy stream is applied to transfer the energy.

Thermodynamics and fluid properties are contained in “fluid packages”. It is very important to choose a suitable fluid package for the process modeled, as some packages are very accurate for HC calculations, while others are very accurate for processes involving water. Special packages, e.g. Amine package are also available through third-part vendors. (AspenTech undated [1])

When starting to model a project, all the components (the substances) in the process must be defined, resulting in a component list. A suitable fluid package then has to be selected based on the component list, before the user is allowed to enter the simulation environment, where an empty PFD will appear. Different process components can now be selected from the object palette and added to the PFD, where the user gives the required input values (varies for the different components). Material streams can then be defined, and for each input stream (or output stream) two independent variables must be given (e.g. pressure and temperature), in addition to the stream composition and flow rate. When the required input (or output) data is given, the solver will automatically start, calculating the rest of the material streams, unit outputs and other unknowns. The results can be view by adding tables to the PFD, or by simply double-clicking the relevant stream or unit.

Appendix H: PFD for TEG injection dehydration concept

H.1 The TEG injection process

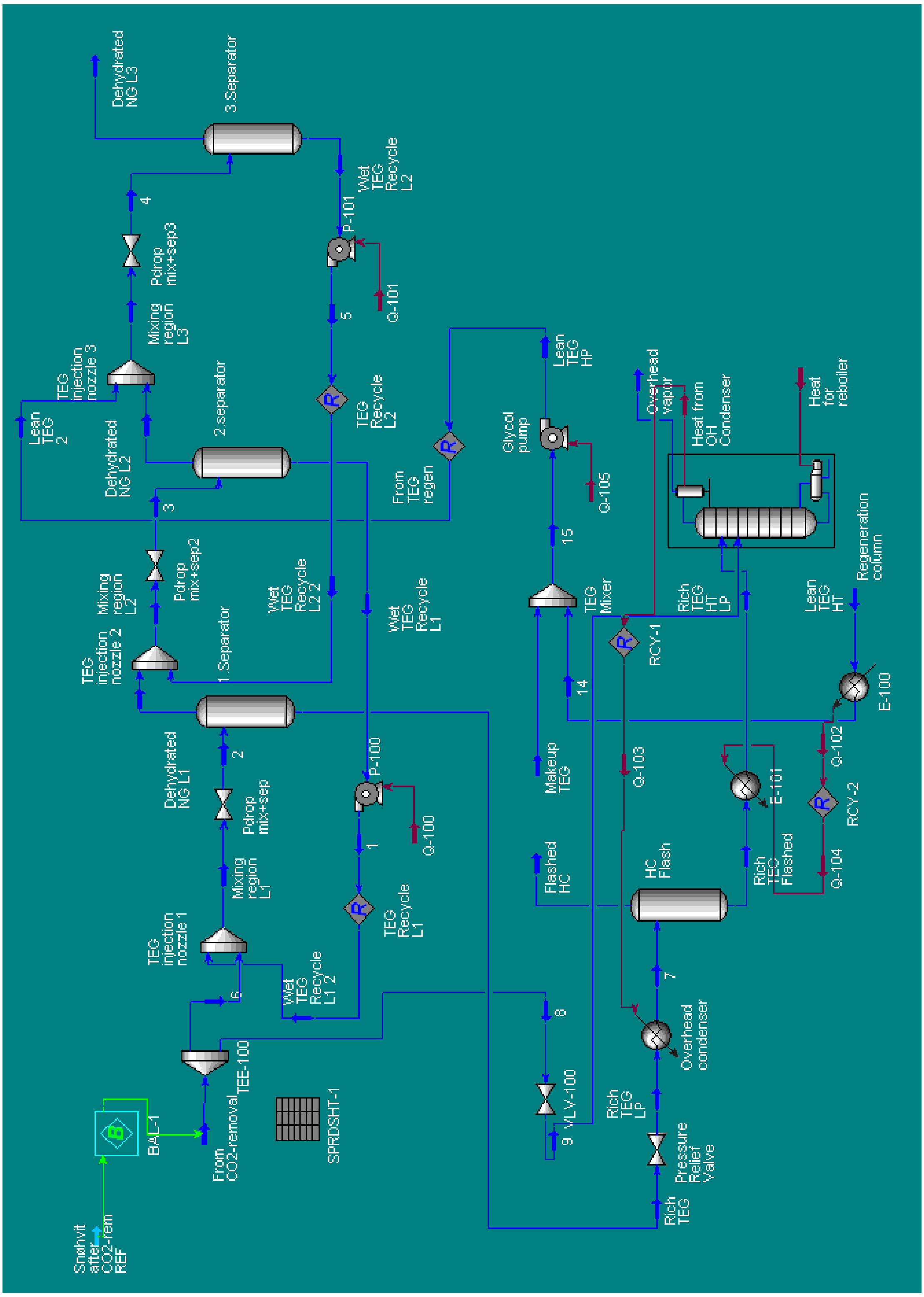

Figure H.1: PFD for the TEG injection concept w/regeneration system

H.2 Conventional absorber dehydration

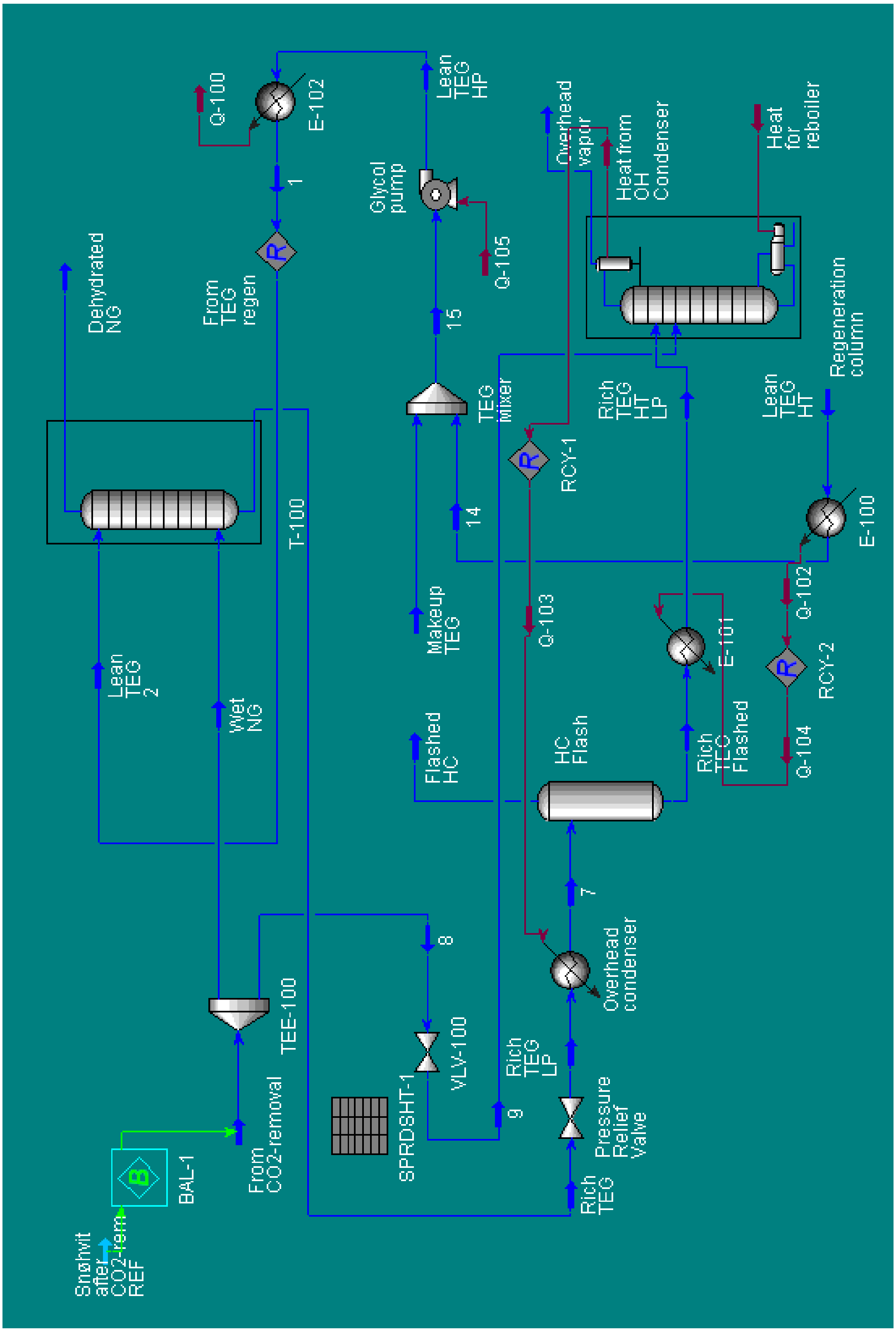

Figure H.2: PFD for conventional absorber dehydration w/regeneration process

Appendix I: Reference gas composition

The reference gas composition used for the simulation is given in the following table:

Table I.1: Reference gas composition

Component	**mol %**	Temperature	26,6	°C
Nitrogen	2,6661	Pressure	64,88	bar
CO2	0,005	Enthalpy	-13274,95	kW
H2O	0,07	Phase	Vapor	Liquid
Methane	86,2807	Fraction	1,0	0,0
Ethane	6,5256	Volume flow	11829,0	m3/h
Propane	2,6825	Molecular Weight	18,88	kg/kgmole
i-Butane	0,4147	Density	57,61	kg/m3
n-Butane	0,8182	Viscosity	0,013	mPa*s
i-Pentane	0,1812	Specific Heat	2,65	kJ/(kg*K)
n-Pentane	0,1701			
Hexanes	0,1003			
Heptanes	0,0449			
Octanes	0,0137			
Nonanes	0,0031			
Benzene	0,0151			
Toluene	0,0058			
Xylenes	0,0015			
C10	0,001			
C11	0,0002			
C12	0,0001			

The reference gas composition is from the Snøhvit processing plant after the sour gas removal unit.

Appendix J: Calculations for chapter 7

All the formulas used in this appendix are found in appendix A and C.

J.1 Regeneration process

Overhead condenser

Using formula C.2:

$$A_{HT} = \frac{9681\frac{kg}{hr}\cdot 2.161\frac{kJ}{kgK}\cdot\left(77.6^{o}C-27.4^{o}C\right)}{3600s\cdot 570\frac{W}{m^{2}K}\cdot\left[89.3^{o}C-\frac{1}{2}\cdot\left(77.6^{o}C-27.4^{o}C\right)\right]} = 8m^{2}$$

Flash tank

Calculating the falling velocity from formula A.1:

$$U_t = 0.05\frac{m}{s}\cdot\left(\frac{1021\frac{kg}{m^3}-2.07\frac{kg}{m^3}}{2.07\frac{kg}{m^3}}\right)^{\frac{1}{2}} = 1.11\frac{m}{s}$$

Calculating the minimum required inner diameter from the gas capacity constraint for a vertical separator (formula A.7):

$$D_i = \left(\frac{4}{\pi}\cdot\frac{82.67\frac{m^3}{hr}}{1.11\frac{m}{s}\cdot 3600s}\right)^{\frac{1}{2}} = 0.16m \quad \text{(Gas capacity)}$$

Calculating the minimum required inner diameter from the liquid capacity constraint (formula A.8):

$$D_i = \left(\frac{\frac{4}{\pi}\cdot 9.31\frac{m^3}{hr}\cdot 4\min}{1.3m\cdot 60s}\right)^{\frac{1}{2}} = 0.78m \quad \text{(Liquid capacity)}$$

The minimum required diameter is the biggest of the calculated diameters, thus $D_i = 0.78m$. The total length of the separator can then be calculated from formula A.9:

$$L = 1.3m + 0.78m + 1m = 3.08m$$

The slenderness ratio is then calculated from formula A.5 to see if the design is suitable:

$$Slenderness\ ratio = \frac{3.08}{0.78} = 3.95$$ -> Within the recommended values

To determine the shell thickness, the maximum allowable stress for the material must be determined. Using the table from Arnold & Stewart (2008):p.321, and assuming the carbon steel is of the cheapest quality (as the pressure difference is very small), σ_{max} is found to be 126.0 MPa. The joint efficiency is assumed to be 1. Using formula A.15:

$$t_{W,\min} = \frac{128kPa \cdot 0.78m}{2 \cdot 1.26E5kPa \cdot 1 - 1.2 \cdot 128kPa} = 0.0004m$$

Adding a safety margin of 3.2 mm, the resulting wall thickness is 3.24 mm. However to ensure rigidity of the vessel, a minimum wall thickness of 6.4 mm for diameters below 1.07 m is recommended by Coker (2007).

Using formula A.16, the corresponding required thickness for the heads (assumed ellipsoidal) is found:

$$t_{W,\min} = \frac{128kPa \cdot 0.78m}{2 \cdot 1.26E5kPa \cdot 1 - 0.2 \cdot 128kPa} = 0.0004m$$

Adding the safety margin, the thickness for the heads is calculated to 3.24mm, the same as for the wall. However to ensure rigidity, a wall thickness of 6.4 mm is used for the heads as well.

Using ρ_{steel} = 7850 kg/m^3 (Elert (2004)) the total weight of the flash tank is calculated to 432 kg (empty).

Glycol/glycol heat exchanger

First the LMTD needs to be calculated:

$$LMTD = \frac{\left(90.0^{\circ}\mathrm{C} - 77.6^{\circ}\mathrm{C}\right) - \left(204.0^{\circ}\mathrm{C} - 169.2^{\circ}\mathrm{C}\right)}{\ln\left(\dfrac{90.0^{\circ}\mathrm{C} - 77.6^{\circ}\mathrm{C}}{204.0^{\circ}\mathrm{C} - 169.2^{\circ}\mathrm{C}}\right)} = 21.71^{\circ}\mathrm{C}$$

Inserting the calculated LMTD into formula C.4 together with the other given properties gives:

$$A_{HT} = \frac{8999\dfrac{kg}{hr} \cdot 2.20\dfrac{kJ}{kgK} \cdot \left(204.0^{\circ}\mathrm{C} - 90.0^{\circ}\mathrm{C}\right)}{3600s \cdot 230\dfrac{W}{m^2K} \cdot 21.71^{\circ}\mathrm{C}} = 125.6m^2$$

Using 100 pipes with a 5 cm diameter and t_w = 3 mm doing two passes for the tube side and assuming the same area is required for shell side flow the volume and weight are calculated as follows:

$$A_{pipes} = 2 \cdot 100 \cdot \frac{\pi}{4} \cdot 5.6^2 cm = 0.5m^2 \Rightarrow A_{tot} = 2 \cdot A_{pipes} = 1.0m^2 \Rightarrow D_{hx} = \left(\frac{4 \cdot A_{tot}}{\pi}\right)^{\frac{1}{2}} = 1.13m$$

The required length of the pipes is then determined:

$$L_{pipes} = \frac{125.6m^2}{\pi \cdot 0.05m \cdot 100 \cdot 2} = 4m$$

Thus, the heat exchanger itself must be 4 m long. Determining the minimum required wall thickness (formula A.15):

$$t_{W,\min} = \frac{10kPa \cdot 1.13m}{2 \cdot 1.26E5kPa \cdot 1 - 1.2 \cdot 10kPa} = 0.000057m$$

Adding a corrosion safety margin the thickness is still below the minimum required for rigidity, and t_w = 8.1 mm is therefore chosen (Coker (2007)).

The total weight of the heat exchanger included the piping is then found to be 4178 kg.

Regeneration column

Calculating the falling velocity from formula A.1:

$$U_t = 0.1\frac{m}{s} \cdot \left(\frac{977.5\frac{kg}{m^3} - 0.804\frac{kg}{m^3}}{0.804\frac{kg}{m^3}} \right)^{\frac{1}{2}} = 3.49\frac{m}{s}$$

Calculating the minimum required diameter from the gas capacity (formula A.7):

$$D_i = \left(\frac{4}{\pi} \cdot \frac{1616\frac{m^3}{hr}}{3.49\frac{m}{s} \cdot 3600s} \right)^{\frac{1}{2}} = 0.40m$$

Using HETP = 1.2 m, the total height is calculated (1.5 m is added for inlet- and exit facilities, as well as overhead condenser):

$$h_{packing} = 4 \cdot 1.2 = 4.8m,\ \mathrm{h}_{\mathrm{tot}} = 4.8m + 1.5m = 6.3m$$

Assuming the same type of steel is used as for the flash tank, the required shell thickness is calculated from formula A.15:

$$t_{W,\min} = \frac{10kPa \cdot 0.40m}{2 \cdot 1.26E5kPa \cdot 1 - 1.2 \cdot 10kPa} = 0.00002m$$

Because of the near ambient pressure, the required wall thickness is very small, but by adding a corrosion safety margin, the resulting thickness is 6.42 mm (6.4 mm is added, since corrosion can occur).

This gives the total weight and volume of the shell to be 0.82 m^3 and 418 kg respectively. The packing also contributes to the total weight, estimated to 100 kg, thus the total weight for the column is 708 kg (including the overhead condenser).

Reboiler

Calculating the size using the same procedure as for a two-phase horizontal separator (formula A.4):

$$L \cdot D_i^2 = \frac{8}{\pi} \cdot 9.773 \frac{m^3}{hr} \cdot 5\min \cdot \frac{1}{60s} = 2.07m^3$$

Assuming a slenderness ratio of 5 (maximum recommended), L and D_i is found:

L = 3.73 m and D_i = 0.75 m

Shell thickness from formula A.15:

$$t_{W,\min} = \frac{10kPa \cdot 0.75m}{2 \cdot 1.26E5kPa \cdot 1 - 1.2 \cdot 10kPa} = 0.000038m$$

Using the minimum wall thickness to ensure rigidity recommended by Coker (2007), t_w = 6.4 mm.

The required heat for the reboiler is calculated from formula C.5:

$$q_{req} = 557.43 \frac{kJ}{l} \cdot 9770 \frac{l}{hr} \cdot \frac{1hr}{3600s} = 1512kW$$

J.2 TEG injection process

Calculating the pipe diameter from the definition of volume flow assuming the velocity is 20 m/s:

$$D_i = \left(\frac{4 \cdot 11940 \frac{m^3}{hr}}{\pi \cdot 20 \frac{m}{s} \cdot 3600s} \right)^{\frac{1}{2}} = 0.46m$$

Calculating the wall thickness for the pipe from formula A.15:

$$t_{W,\min} = \frac{6387kPa \cdot 0.46m}{2 \cdot 1.724E5kPa \cdot 1 - 1.2 \cdot 6387kPa} = 0.0087m$$

Adding the safety margin, the resulting pipe wall thickness is 11.9 mm.

Deliquidiser

Calculating the length and height using the values given in table 1.2:

$$L = 6 \cdot 0.46m = 2.76m,\ h = 4 \cdot 0.46 = 1.84m$$

Scrubber

Calculating the falling velocity from formula A.1:

$$U_t = 0.15\frac{m}{s}\cdot\left(\frac{1026\frac{kg}{m^3} - 54.79\frac{kg}{m^3}}{54.79\frac{kg}{m^3}}\right)^{\frac{1}{2}} = 0.63\frac{m}{s}$$

Calculating the minimum required diameter from the gas capacity constraint (formula A.7):

$$D_i = \left(\frac{4}{\pi}\cdot\frac{12420\frac{m^3}{hr}}{0.63\frac{m}{s}\cdot 3600s}\right)^{\frac{1}{2}} = 2.64m$$

Calculating the required liquid height assuming the retention time for the liquid is 3 minutes (formula A.8):

$$h_L = \frac{4}{\pi}\cdot\frac{9.171\frac{m^3}{hr}\cdot 15\min}{2.64^2 m^2\cdot 60s} = 0.42m$$

The length of the entire vessel is then calculated using formula A.9:

$$L = 0.42m + 2.64m + 1m = 4.06m$$

Checking the slenderness ratio (formula A.5):

$$Slenderness\ ratio = \frac{4.06}{2.64} = 1.54$$

The resulting minimum wall thickness is calculated from formula A.15:

$$t_{W,\min} = \frac{6387kPa\cdot 2.64m}{2\cdot 1.724E5kPa\cdot 1 - 1.2\cdot 6387kPa} = 0.050m$$

Calculating the thickness for the heads (assumed ellipsoidal) from formula A.16:

$$t_{W,\min} = \frac{6387kPa\cdot 2.64m}{2\cdot 1.724E5kPa\cdot 1 - 0.2\cdot 6387kPa} = 0.049m$$

Piping

Calculating the TEG recycle pipe diameter from the definition of volume flow:

$$D_i = \left(\frac{4\cdot 9.218\frac{m^3}{hr}}{\pi\cdot 5\frac{m}{s}\cdot 3600s}\right)^{\frac{1}{2}} = 0.026m$$

Calculating the wall thickness for the TEG recycle streams piping using formula A.15:

$$t_{W,\min} = \frac{6387kPa \cdot 0.026m}{2 \cdot 1.26E5kPa \cdot 1 - 1.2 \cdot 6387kPa} = 0.0007m$$

Adding the safety margin, t_W = 3 mm is assumed to be sufficient.

J.3 Comparing with existing TEG dehydration process

Calculating the wall thickness using formula A.15:

$$t_{W,\min} = \frac{8350kPa \cdot 3.2m}{2 \cdot 1.724E5kPa \cdot 1 - 1.2 \cdot 8350kPa} = 0.080m$$

Adding a corrosion allowance of 3.2 mm, the resulting thickness is 83.2 mm.