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Novel Processes for Power Plant with CO2 Capture

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MASTER THESIS

for

Student Kjetil Vinjerui Ekre

Spring 2012

Novel processes for power plant with CO_2

Nye prosesser for kraftverk med CO_2-fangst

Background and objective

One of the main challenges with CO_2 capture in power generation systems is the high energy penalty related to the capture of CO_2. The current status with respect to energy penalty is that post-combustion systems seem to imply a slightly less penalty compared to pre-combustion and oxy-combustion methods. A main disadvantage with the post-combustion method is that capturing CO_2 from an atmospheric flue gas requires large equipment because of the very large volumetric flow rates, as well as a significant energy penalty because of the very low partial pressure of CO_2 in the flue gas.

The main objective of this work is to suggest and analyse power cycles where the CO_2 capture is enabled with an elevated partial pressure of CO_2. Various separation methods for separation of CO_2 from the flue gas are to be investigated, such as absorption with amines or other solvents, and membranes. The analysis should include issues like plant efficiency, component sizing and technological maturity.

The following tasks are to be considered:

1. The candidate shall carry out a literature review related to novel power cycles, with emphasis on processes adapted for CO_2 capture at elevated pressure.
2. Based on the findings in the literature review, the candidate will select a number of plant configurations for further evaluation.
3. The candidate shall make simulation models for a selected number of cases of power cycle configurations. For each of the power cycle configurations, various separation methods for CO_2 shall be evaluated. A reference plant should be modelled in order to perform parametric evaluations and comparisons of the processes.
4. Based on the simulation models, the selected cycles shall be compared with respect to plant efficiency, component sizing, and technological maturity.

Within 14 days of receiving the written text on the master thesis, the candidate shall submit a research plan for his project to the department.

When the thesis is evaluated, emphasis is put on processing of the results, and that they are presented in tabular and/or graphic form in a clear manner, and that they are analyzed carefully.

The thesis should be formulated as a research report with summary both in English and Norwegian, conclusion, literature references, table of contents etc. During the preparation of the text, the candidate should make an effort to produce a well-structured and easily readable report. In order to ease the evaluation of the thesis, it is important that the cross-references are correct. In the making of the report, strong emphasis should be placed on both a thorough discussion of the results and an orderly presentation.

The candidate is requested to initiate and keep close contact with his/her academic supervisor(s) throughout the working period. The candidate must follow the rules and regulations of NTNU as well as passive directions given by the Department of Energy and Process Engineering.

Risk assessment of the candidate's work shall be carried out according to the department's procedures. The risk assessment must be documented and included as part of the final report. Events related to the candidate's work adversely affecting the health, safety or security, must be documented and included as part of the final report.

Pursuant to "Regulations concerning the supplementary provisions to the technology study program/Master of Science" at NTNU §20, the Department reserves the permission to utilize all the results and data for teaching and research purposes as well as in future publications.

The final report is to be submitted digitally in DAIM. An executive summary of the thesis including title, student's name, supervisor's name, year, department name, and NTNU's logo and name, shall be submitted to the department as a separate pdf file. Based on an agreement with the supervisor, the final report and other material and documents may be given to the supervisor in digital format.

Department of Energy and Process Engineering, 23. February 2012

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Olav Bolland
Supervisor

Preface

This thesis, Novel Processes for Power Plant with CO_2 Capture, is written as the Master's Thesis of stud. techn. Kjetil Vinjerui Ekre. The thesis was produced at the Department of Energy and Process Engineering at the Norwegian University of Science and Technology in Trondheim, Norway. The study was written as the final thesis for the 5-year Master of Science Degree in the field of Energy and Environmental Engineering.

The author of the study is hopeful that the information and results provided in this work can contribute as an element in the campaign of mitigating CO_2 emissions from fossil fuel based power production.

Trondheim, August 26th 2012

Kjetil Vinjerui Ekre

Kjetil Vinjerui Ekre

Acknowledgements

I would like to express my gratitude to my supervisor, Olav Bolland for his help and guidance through this year. He has always been available for me and shown interest and enthusiasm all the way.

I would also thank my fellow students at B340 for their endless coffee breaks and heated discussion we have had this year. Especially thanks to Bjørn Jordheim Halvorsen who contributed on the work for our joint project thesis, which laid much of the ground work for this Master's Thesis.

Lastly, I would like to thank my mom for pushing me in the right direction of Energy and Environmental studies and my dad for always showing me enthusiasm for mathematics.

Abstract

The purpose of this thesis was to examine different technologies, which enhances the CO_2 partial pressure in the flue gas from the natural gas combined cycle. A base case has been created as a reference for comparison of the other cycles. The base case includes a MEA capture plant with a reboiler duty of 3,6 MJ/kg CO_2. To simulate the process in this thesis HYSYS and GT PRO have been used as simulation tools. The thesis has also looked into ways of extracting steam from the steam cycle to be used in the reboiler. The chosen extraction point was the crossover between the intermediate-pressure turbine and the low-pressure turbine, the steam was saturated with water from the low-pressure boiler and have a pressure and temperature of 3,6 bar and 140 °C into the reboiler.

Four different technologies have been evaluated in this thesis; a natural gas combined cycle with the use of exhaust gas recycle and, three elevated pressure cycles; post-compression CO_2 capture, post-expansion CO_2 capture, and tail-end CO_2 capture. These processes have been compared against each other with regards to the net plant efficiency, absorber size at the capture plant, and the technological maturity. The most promising of these technologies is the natural gas combined cycle with exhaust gas recycle and the tail-end CO_2 capture processes, with respectively 52 % and 51,7 % net plant efficiency. The smallest absorber size is achieved by the use of post-compression CO_2 capture, with a diameter of 2,9 m and a height of 10,5 m. The elevated pressure cycles have also been tested with the use of MDEA as solvent in the capture plant. By use of elevated pressure and MDEA the reboiler duty was reduced to 2 MJ/ kg CO_2.

Sammendrag

Formålet med denne rapporten var å undersøke ulike teknologier som forbedrer CO_2 partialtrykket i røykgassen fra et kombinert gasskraftverk. Et referanseanlegg har blitt satt opp som sammenligning for de andre prosessene. Referanseanlegget omfatter også et MEA fangstanlegg med en fordampnings varme på 3,6 MJ/kg CO_2. For å simulere prosessene i denne rapporten har HYSYS og GT PRO blitt brukt som simuleringsverktøy. Rapporten har også sett på forskjellige måter å utvinne damp fra dampsyklusen til bruk i fordamperen. Det valgte uttakspunket var ved overgangen mellom middelstrykkturbinen og lavtrykksturbinen, dampen ble mettet med vann fra lavtrykkskjelen og hadde et trykk og temperatur på 3,6 bar og 140 ° C ved inngangen til fordamperen.

Fire ulike teknologier har blitt evaluert i denne rapporten, et kombinert gasskraftverk med bruk av resirkulert eksosgass, tre høytrykksprosesser; post-compression CO_2-fangst, post-expansion CO_2-fangst og tail-end CO_2-fangst. Disse prosessene har blitt sammenlignet mot hverandre med hensyn på netto anleggs effektivitet, absorber størrelse ved fangstanlegget og teknologisk modenhet. Den mest lovende av disse teknologiene er et kombinert gasskraftverk med bruk av resirkulert eksosgass og tail-end CO_2-fangst, med henholdsvis 52 og 51,7 % netto anleggs effektivitet. Post-compression CO_2-fangst prosessen reduserer absorberenstørrelsen mest, en diameter på 2,9 m og høyde på 10,5 m. Høyttrykksprosessene har også blitt testet med bruk av MDEA som solvens i fangstanlegget. Ved bruk av høyere trykk og MDEA reduseres fordampnings varme til 2 MJ/kg CO_2.

Acronyms and Abbreviations

CW	Cooling water
EFC	Externally fired gas turbine cycles
EGR	Exhaust gas recycle
GT	Gas turbine
HP	High-pressure
HPT	High-pressure turbine
HRSG	Heat Recovery Steam Generator
HX	Heat exchanger
IP	Intermediate-pressure
IPB	Intermediate-pressure boiler
IPT	High-pressure turbine
LP	Low-pressure
LPB	Low-pressure boiler
LHV	Lower heating value
LPS	Low-pressure super heater
LPT	Low-pressure turbine
MEA	Monoethanolamine
MDEA	Methyl diethanolamine
NGCC	Natural gas combined cycle
PFBC	Pressurized fluidized-bed combustion
RHE	Recuperator heat exchanger
SFC	Supplementary fired cycle
ST	Steam turbine
Wt. %	Weight percentage

Nomenclature

Parameters

α	Ratio between heat potential and power reduction	[-]
C	Ratio of formed CO2 and fuel	[kg/kg]
f	CO2 capture rate	[-]
c_p	Heat capacity	[kJ/kg K]
ΔH	Enthalpy of formation	[kJ/kg]
$\dot{m}_{steam}$	Mass flow rate of steam	[kg/s]
$\dot{m}_{cw}$	Mass flow rate of cooling water	[kg/s]
$\dot{m}_{water}$	Mass flow rate of water	[kg/s]
$\dot{m}$	Mass flow rate	[kg/s]
$\dot{m}_{fuel}$	Mass flow rate of fuel	[kg/s]
$\dot{m}_{cross}$	Mass flow rate of steam at crossover	[kg/s]
$\dot{m}_{ext}$	Mass flow rate of steam at extraction point	[kg/s]
$\dot{m}_{inject}$	Mass flow rate of water injected from deaerator	[kg/s]
$\dot{Q}_{cw}$	Heat removed in cooling water	[MJ]
$\dot{Q}_{reb}$	Heat requirement in reboiler	[MW]
$\Delta\dot{Q}$	Heat difference	[MJ]
α_{cross}	Steam quality by extraction from IPT/LPT crossover	[-]
$h_{sub\ cooled}$	Enthalpy after reboiler, subcooled condition	[kJ/kg]
$h_{reb\ exit}$	Enthalpy at reboiler exit	[kJ/kg]
h_{cross}	Enthalpy at crossover	[kJ/kg]
$h_{exit\ turb}$	Enthalpy at turbine exit	[kJ/kg]
$h_{cold,in}$	Enthalpy of cooling water inlet	[kJ/kg]
$h_{cold,out}$	Enthalpy of cooling water outlet	[kJ/kg]
$h_{ST\ inlet}$	Enthalpy at steam turbine inlet	[kJ/kg]
$h_{ST\ outlet}$	Enthalpy at steam turbine outlet	[kJ/kg]
h_{exit}	Enthalpy at HX exit	[kJ/kg]
h_{inlet}	Enthalpy at HX inlet	[kJ/kg]
$h_{inlet\ turb}$	Enthalpy at turbine inlet	[kJ/kg]
h_{ext}	Enthalpy in the extraction point, prior to the reboiler	[kJ/kg]
h_{cond}	Enthalpy of saturated liquid at reboiler exit	[kJ/kg]
$E^{CO_2}_{rem,mech}$	Mechanical work consumption in capture process	[MJ/kg CO2]
$E^{CO_2}_{rem,heat}$	Heat consumption in stripper process	[MJ/kg CO2]
$E^{CO_2}_{comp}$	Work requirement for compression of CO2	[MJ/kg CO2]
ΔE_{aux}	Work difference with no capture and reference NGCC	[MJ/kg CO2]
$\dot{W}_{ext}$	Steam turbine work in NGCC with steam extraction	[MW]
$\Delta\dot{W}_{lost}$	Lost steam turbine work due to steam extraction	[MW]
$\dot{W}_{turb}$	Turbine work	[MW]
$W_{net\ output}$	Net power output in power plant	[MW]
$\dot{W}_{no\ ext}$	Steam turbine work in NGCC without steam extraction	[MW]

T_{exit}	Temperature at HX exit	°C
T_{inlet}	Temperature at HX inlet	°C
ΔT	Temperature difference	°C
P_{ext}	Pressure at the extraction point, prior to the reboiler	bar
P_{cross}	Pressure at the IPT/LPT crossover	bar
P_{LPB}	Pressure in low-pressure boiler	bar
ΔP_{loss}	Pressure loss	bar
η_{NGCC}	Efficiency of natural gas combined cycle	%
$\eta_{no\ cap}$	Efficiency of power plant with no CO2 capture	%
η_{is}	Isentropic efficiency	%
η_{shaft}	Shaft efficiency	%
η_{gen}	Generator efficiency	%

Contents

1 Introduction

The climate is changing and the temperature in the atmosphere is increasing. The Intergovernmental Panel for Climate Change have indicated that during the 21st century the global surface temperature is likely to rise from 1.1-2.9 °C for their lowest emissions scenario and from 2.4-6.4 °C for their highest[1]. These temperature changes are somewhat caused by human emission and CO_2 have the greatest impact on the greenhouse effect and represents approximately 5 % of the global warming[1]. The amount of CO_2 in the atmosphere has lately been measured at the highest level ever. From Figure 1.1 we see that the CO_2 amount gradually has increased over the last years, and is still increasing. While the population keeps on increasing, the energy demand increases proportionally. Fossil fuel represents over 60 % of the total electricity production, and in waiting for a better primary energy source, the usage of fossil fuel will increase dramatically in the next decades, thereby releasing more and more CO_2 into the atmosphere. There are many ways to curb this growth; one of them is to introduce more renewable energy sources. Another approach is to capture and store the CO_2 away from the atmosphere, this study will go into depth on some of the capture methods.

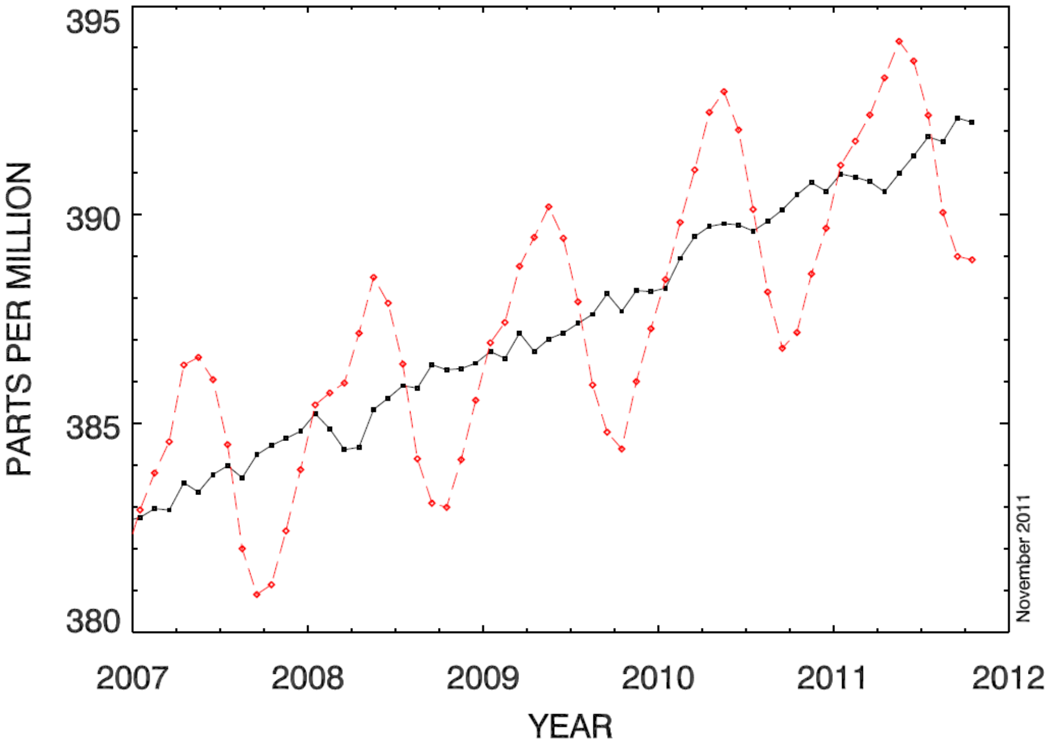

Figure 1.1 - Trends in Atmospheric Carbon Dioxide[2].

Prior to the finance crisis in 2008 a lot of money was given for research on Carbon Capture and Storage (CCS) and the building of pilot plants for CO_2 capture from power generation. After the crisis most of the pilot plants projects have been postponed or canceled due to the lack of further funding. Capturing and storaging of CO_2 requires a lot of energy and is very expensive, with a result that many investors became too concerned about their own finance and not on how prevent the growing climate change as best they can. The energy demand of the capture process is too high at the current state of development. The capture technology needs improving to get the attention of investors. Money will be invested as soon as the capturing of CO_2 becomes more economical thereby enhancing the production rate of new capture plants and improving the capture efficiency.

1.1 Carbon Capture

There are three main principles for capturing carbon from a power plant. As seen in Figure 1.2 the different ways are pre-, post- and oxy-fuel combustion capture. Within each of these capture stages membranes can be used in order to separate CO_2. This study will focus on post-combustion based on chemical absorption with amines.

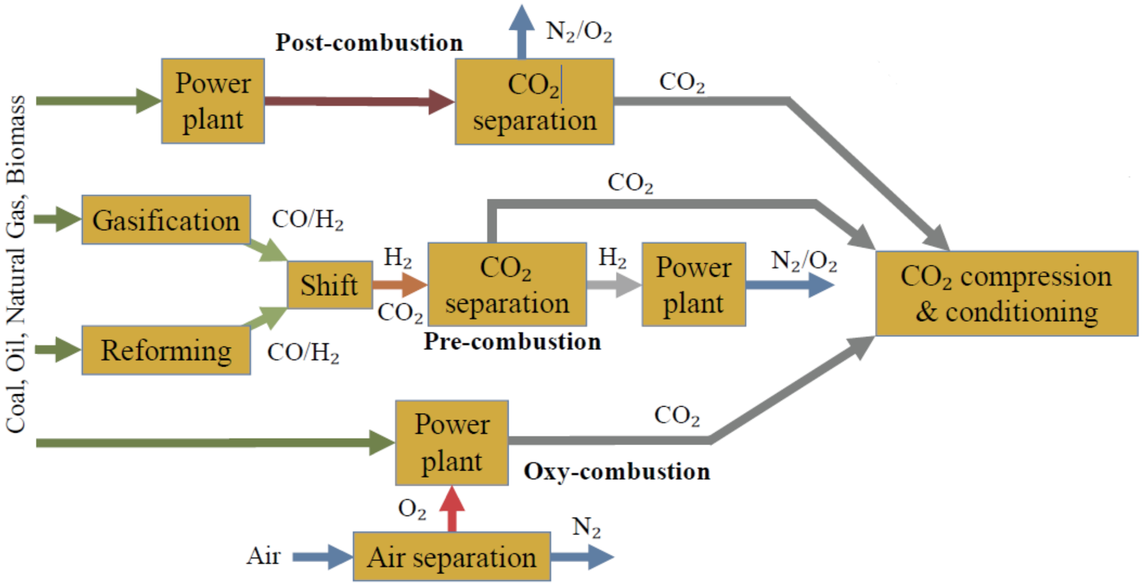

Figure 1.2 - Principle methods for CO_2 capture[3].

1.1.1. Pre-combustion

In pre-combustion the process is divided into two main reactions. In the first reaction the fuel is mixed with either pure oxygen equation 1.1 often called partially oxidation or steam equation 1.2 called steam reforming, these reactions transforms the fuel into carbon monoxide and hydrogen.

$$C_xH_y + \frac{x}{2}H_2O \leftrightarrow xCO + \left(\frac{y}{2}\right)H_2 \qquad \Delta H - ve \qquad 1.1$$

$$C_xH_y + xH_2O \leftrightarrow xCO + \left(\frac{x+y}{2}\right)H_2 \qquad \Delta H + ve \qquad 1.2$$

After the steam reforming the temperature is lowered so that a shift reaction can transform the CO into CO_2 by adding more steam (Water Gas Shift Reaction, equation 1.3)

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \Delta H = -41\ kJ/mol \qquad 1.3$$

The CO_2 is then removed from the process, and the concentration of CO_2 before the separation can be in the range of 15-60%(dry basis)[1].

1.1.2. Oxy-fuel Combustion

Oxy-fuel combustion is a type of combustion where fuel is burned with pure oxygen or a blend of oxygen and carbon dioxide instead of air. By eliminating the nitrogen from the combustion the products is mostly a mixture of CO_2 and water vapor equation 1.4.

$$C_xH_y + (x + \frac{y}{4})O_2 \leftrightarrow xCO_2 + \frac{y}{2}H_2O \qquad 1.4$$

After the combustion the flue gas is cooled down and water is condensed out, leaving a product of almost pure CO_2, depending on how pure the O_2 are and the fuel. The flue gas often requires further cleaning. When burning natural gas with pure oxygen the temperature in the combustion can get very high, up to 3500 °C. The temperature in an oxy-fuel coal-fired boiler using current technology is around 1900 °C. To decrease the temperature in the combustion flue gas is recycled back into the combustor; by regulating the amount of flue gas that is recycled it's possible to get the desired temperature in the combustion. The problem with oxy-fuel combustion is that the process of separating oxygen form the air is very energy demanding[1].

1.1.3. Post-combustion

There are many methods for capturing carbon after the combustion. Since CO_2 is a rather weak acid alkaline solvent can be used to absorb it, membrane can be used to separate the CO_2 and micro porous structures can be used to adsorb it. The most common process is absorption and that will be explained in more depth in chapter 3.1. With adsorption the process take place on the surface of the adsorbent, the most often used adsorbent is a solid, but liquids are also used. The micro porous structure that is used is often activated carbon or molecular sieves[1]. In this process the CO_2 is accumulated in the pores of the adsorbent, as can be seen in Figure 1.3 [4].

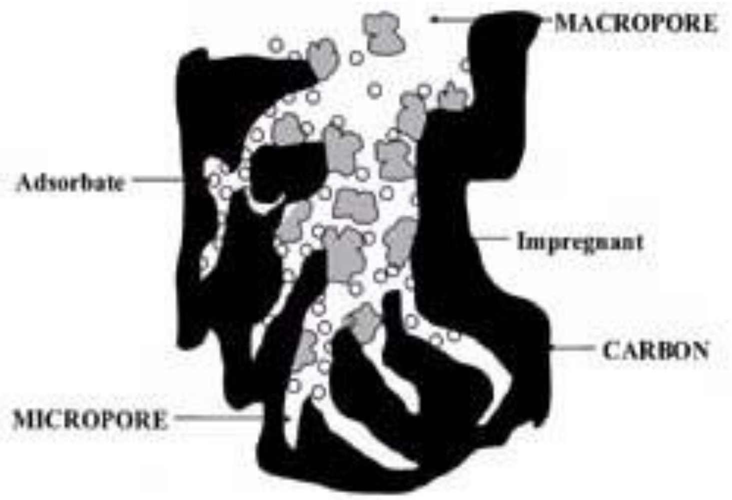

Figure 1.3 - Adsorption with activated carbon[4].

In adsorption pressure swings and/or temperature swings are used to desorb the CO_2 from the structure. An adsorption process is one of the most researched CO_2 capturing technologies at the moment [3].

1.1.4. Membrane

Membrane is a technology that selectively transfers certain chemical components over other components through a wall. A schematic of a membrane is presented in Figure 1.4. The feed gas containing e.g. CO_2 or O_2 is transported past the membrane wall and some of the CO_2 or O_2 is transported through, the amount of CO_2 depends on the membranes' permeability and selectivity. On the other side of the wall sweep gas is used to transport the CO_2 away from the wall. This is done in order to keep the partial pressure low on the permeate side. By having low partial pressure on the permeate side more CO_2 is diffused through. This technology can be used in pre-, post- and oxy-fuel combustion. It can be used to separate O_2 from air or separate CO_2 from flue gas. The problem with membrane today is its low tolerance to high and varied temperatures. With post-combustion membranes can be used in combination with absorption where the absorbent is flowing on the permeate side as the sweep gas to absorb the CO_2[3].

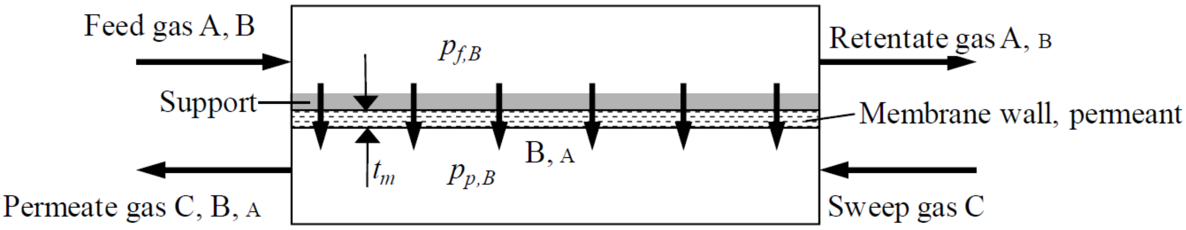

Figure 1.4 – A Typical membrane [3]

1.2 Risk Assessment

In the writing of this master thesis there have not been performed any laboratory work or excursions, therefore the risk assessment required for most master thesis have not been done.

1.3 Scope of Thesis

The main goal of thesis is to suggest and analyze different power plant cycles where the CO_2 capture is enabled with an elevated partial pressure of CO_2. A literature study will be carried out with focus on different methods for CO_2 capture for power plant with elevated CO_2 partial pressure. Some of the CO_2 capture methods will be further evaluated, and the focus shall be on the net plant efficiency with capture, the components size and the maturity of the technology. There will not be completed any economical evaluations.

The thesis is divided into two parts: In the first part a detailed theoretical introduction is given of the fundamentals of CO_2 capture and power generation. Thereafter a literature study has been carried out, where different power plant configurations with capture at elevated pressure have been compared. In the next part process integration have been taken into consideration, the focus have been on the extraction point of steam and the condition of the steam turbine and how to include exhaust gas recycle(EGR) in the natural gas combined cycle. Conclusions are made to be further used in the simulation that is presented in the last part of the thesis.

The last past main part of the thesis is the simulations of the different power plant configuration with CO_2 capture. In this part the simulation results is evaluated and presented. The tools to complete the simulation are: HYSYS and GT PRO each of these programs have been linked together in Microsoft Excel via ELINK and Aspen workbook. Conclusions are made based on the efficiency of each of these technologies.

2 Power Generation

In order to evaluate the possibilities for integration of a power plant and the accompanying CO_2 capture process, it is important to look into and analyze the power generation process. In particular the steam cycle in a combined cycle is essential. In this chapter, a brief overview of the most important components in a natural gas combined cycle (NGCC) is given.

Combined Cycle

A combined cycle could be defined as a combination of two thermal cycles in one plant[5]. This combination makes it possible to extract more energy from the fuel, and contributes to a higher total efficiency in the power plant. In a NGCC these two cycles consists of a gas and a steam cycle. The gas cycle is operating on a higher temperature level than the steam cycle, and is called the topping cycle[5]. Because of the high gas turbine exit temperature it is possible to utilize the energy remained in the exhaust stream. This heat is used to generate steam in a second process, and the steam will be expanded in a steam turbine. The net electric power efficiency of a NGCC could be close to 60%[3].

2.1 Gas Turbine

A gas turbine consists of a compressor, a burner and a turbine. The gas turbine cycle is called a Brayton cycle, see Figure 2.1. In the compressor, ambient air is compressed to a pressure of 14 to 30 bar, depending on the gas turbine[5]. The air is used to burn the fuel, producing a hot gas in the combustion chamber. In this work natural gas is the only fuel type considered. The fuel could be pre-heated before the combustion. An important property of the fuel is the fuels lower heating value (LHV). It defines the mass flow of the fuel, which must be supplied to the gas turbine. After the burning, the temperature of the hot gas may be up to 1500 °C, and the gas is expanded in a turbine, producing power[6]. The power is produced by conversion of pressure energy to kinetic energy with stators, and then the kinetic energy is converted to power because of rotation of the shaft at the rotors that is connected to a generator.

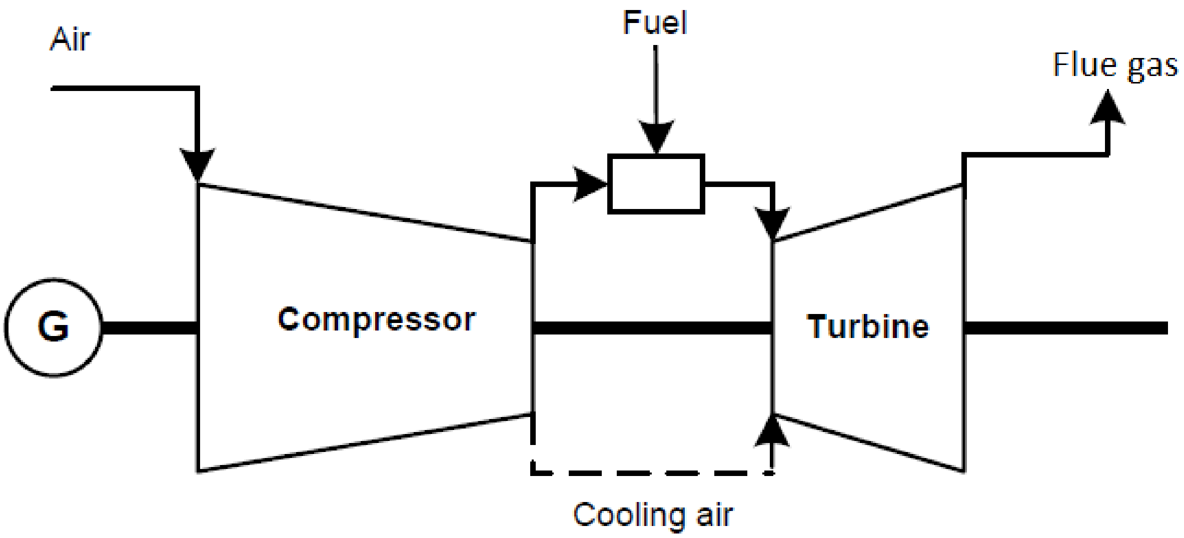

Figure 2.1 - Brayton cycle

When increasing the inlet temperature of the gas turbine, a higher useful enthalpy drop is produced thus generating more power and the efficiency of the gas turbine increases. The inlet temperature of the gas turbine is limited by the materials of the blades in the gas turbine. Turbine cooling is necessary for high temperatures. The most common cooling system is to inject excess air from the air compressor into small bleeds in the turbine blades.

The efficiency of the gas turbine is defined as the conversion of the fuels LHV to electric power. For large gas turbines in power plants the efficiency is in the range 35-40%[6]. The flue gas out of the turbine is slightly above atmospheric pressure, and has a temperature from 450-650 °C. The exhaust stream contains almost all the not converted energy from the fuels lower heating value. As seen from the efficiency value, this energy is significant. By connecting the gas turbine to a steam cycle some of this energy could be recovered. The connection takes place in a Heat Recovery Steam Generator (HRSG).

2.2 Heat Recovery Steam Generator

The second cycle is the steam cycle, where steam is produced in a HRSG before it is expanded through a steam turbine for power generation. The HRSG consists of three heat exchanger sections that transport the heat from the hot exhaust gas to a water cycle.

First, the water in the steam cycle is pre-heated in an economizer from a subcooled condition to a close to saturated condition. The saturation temperature depends on the pressure of the water. It is possible for the HRSG to operate at more than one pressure level. This will be explained more in depth in chapter 2.2.1. The reason why the water is not at saturated conditions at the economizer exit is to avoid evaporation in the economizer at off-design conditions[5]. After the economizer the water is fed into a boiler. In the boiler, the water is heat exchanged with the exhaust gas and evaporated at constant pressure and temperature. At the end, the steam is superheated in a superheater. Superheating of the steam increases the enthalpy in the stream. The work generated in the steam turbine, except mechanical and generation losses, is given by the following equation:

$$\dot{W}_{turb} = \dot{m}_{steam}(h_{ST\ inlet} - h_{ST\ outlet})\eta_{gen}\eta_{shaft} \qquad 2.1$$

As seen from equation 2.1, an increase of the inlet enthalpy contributes to an increase of the steam turbine power output. Another reason for superheating of the steam is to avoid liquid in the turbine. Liquid in the turbine could reduce the power output by slowing down the turbine blades. High moisture content at the end of the steam turbine also increases the risk of erosion. It is recommended a moisture content limit of 16% at the steam turbine exit[5].

2.2.1 Pressure Levels

One of the most important parameters in the HRSG design is the minimum temperature difference between the exhaust gas and the water within a given pressure level. This temperature difference is called the pinch point. The location of the pinch point depends on the HRSG inlet temperature of the flue gas and the pressure in the water cycle. For low flue gas temperature, as is the case for natural gas as fuel, the pinch point is located between the economizer and the evaporator.

The heat transfer on the water side in a heat exchanger is given by the following equation:

$$\dot{Q} = \dot{m}_{water}(h_{inlet} - h_{exit}) \tag{2.2}$$

The equation could be rewritten by assuming a constant specific heat capacity for water:

$$\Delta\dot{Q} = \dot{m}_{water}c_p(T_{inlet} - T_{exit}) \tag{2.3}$$

The slope of the water lines in the TQ diagram is then given by equation 2.4.

$$\frac{\Delta T}{\Delta\dot{Q}} = \frac{1}{\dot{m}_{water}c_p} \tag{2.4}$$

In order to achieve a high power generation in the steam cycle, in relation to equation 2.1, the amount of steam produced is of importance. A moderate slope of the water in the TQ diagram is preferred, giving a high steam mass flow. A typical TQ diagram is presented in Figure 2.2.

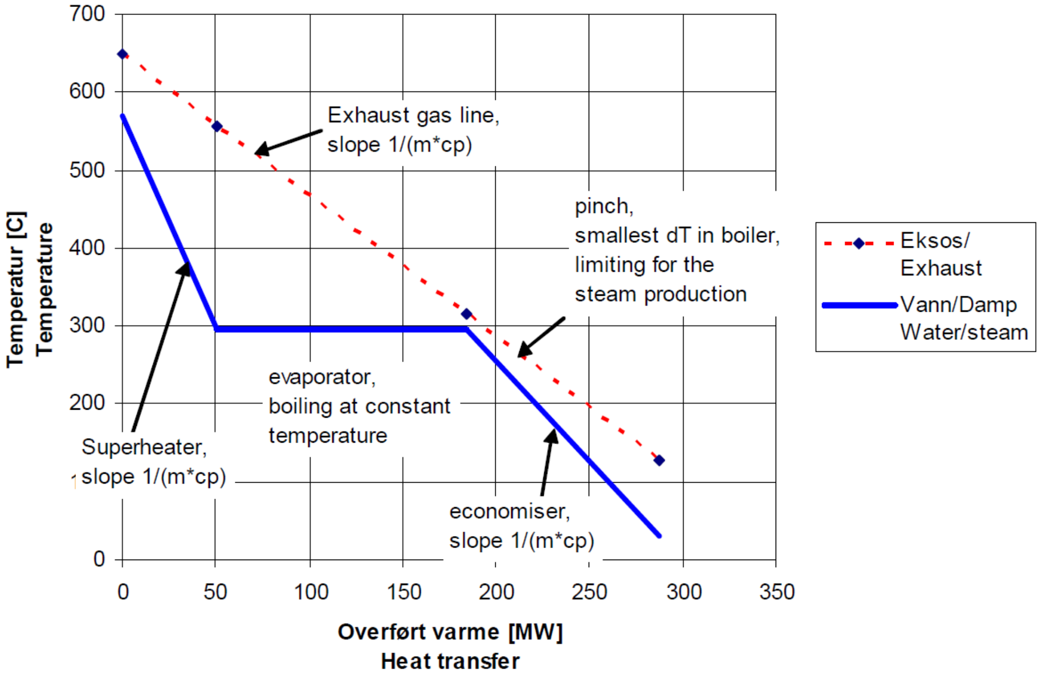

Figure 2.2 – TQ-diagram for heat recovery process[6].

The high heat of evaporation of water is a disadvantage for the steam generation. The effect of a high heat of evaporation on the TQ diagram is a large temperature difference between the water and the flue gas at the steam inlet of the superheater. As seen in Figure 2.2, the low flue gas temperature contributes to a steep slope of the superheater line, in order to reach the live steam temperature. That is a drawback for the steam generation. An option in order to increase the steam generation is to introduce multiple pressure levels.

By introducing multiple pressure levels, it is possible to utilize the lower temperatures in the flue gas because of reduced evaporation temperature at the lowest pressure levels. The mean temperature difference will be reduced through the HRSG, and more steam is produced. In a combined cycle with natural gas as fuel, the inlet temperature of the HRSG is about 450-650 °C, and for this low temperature range multiple pressure levels will increase the steam cycle efficiency[3].

The pinch point refers to the smallest temperature difference through a HRSG. With higher temperature difference less heat exchanger area is needed to evaporate the water, compared to a lower pinch point. This is equal to a cheaper heat exchanger. With lower pinch point, the exhaust heat is utilized better and more steam is generated. How to design the pinch point is an optimization problem, but if the goal is to maximize the efficiency of the plant the lowest pinch point is optimal. There has to be a temperature difference for heat exchanging between the mediums to occur. Pinch points are often in the range of 8-15K[5].

2.2.2 Pressure Drop

There will be a pressure drop through the HRSG. The size of the pressure drop of the flue gas is a design problem of the heat exchangers. The biggest disadvantage related to the flue gas pressure drop, is the backpressure of the gas turbine. To overcome the pressure drop, the exit pressure out of the gas turbine has to increase. This increase leads to less work generated in the gas turbine. The lost work is proportional to the pressure drop and the temperature close to the turbine exit among other factors [3]. Therefore, the power generation in the gas turbine is very sensitive to a change in the pressure drop.

There are also advantages related to an increased pressure drop. A higher pressure and enthalpy at the exit of the gas turbine is related to a higher temperature out of the turbine. The flue gas temperature at the inlet of the HRSG increases and more steam could be generated in the steam cycle; increasing the steam turbine power output. However, the decrease in the gas turbine power output is dominant and the net combined cycle output is also decreased with increased pressure drop[5].

At the same time, there is a relation between pressure loss, gas velocity and the heat transfer inside a heat exchanger. A lower gas velocity contributes to less turbulence and thereby low heat transfer and pressure loss[7]. For most of the heat exchangers the relations are:

$$h \sim \Delta P^{0,33} \qquad 2.5$$

h is in this case the heat transfer coefficient. As seen from the expression, the heat transfer is reduced when the pressure loss decreases. In order to obtain the total heat transfer through the HRSG, the heat exchanger area must increase. Increased heat exchanger surface is equivalent to higher costs. The designing of the HRSG is an optimization problem, with power output and costs as parameters. Usually the pressure loss on the flue gas side of the HRSG is about 30 mbar[3].

2.3 Steam Turbine

After the superheater the steam enters the steam turbine. The steam expands through the turbine and there will be generated power. Depending upon the number of pressure levels in the HRSG, there will be multiple inlets at the turbine. In the case of three pressure levels with reheat, there will be three inlets, two outlets and one crossover for the steam turbines. The moisture content at the steam turbines exit is a limiting factor for the highest pressure levels. By introducing reheat of the steam expanded in the high-pressure turbine (HPT), the pressures could be higher in relation to the moisture content. The high-pressure steam, normally in the range 100-130 bar and a temperature of 450-560 °C, enters the HPT[3]. At the outlet of the HPT, the steam exits and mixes with the intermediate pressure steam in the HRSG. The mixed steam is superheated in a re-heater and enters the intermediate-pressure turbine (IPT) with conditions about 30 bar and temperatures of 450-560 °C. There is a crossover between the exit of the IPT and the inlet of the low-pressure turbine (LPT), where the steam flows. In addition, steam at low pressure conditions is transferred to the LPT from the HRSG. The steam conditions at the LPT inlet are about 3-5 bar and 200-300 °C. Finally the expanded steam exits the steam turbine at a low pressure, around 0,04-0,05 bar. The isentropic efficiency is different for each of the three turbines in a three pressure level cycle.

In chapter 5.1, the possibilities for steam extraction from the steam turbine for heating purposes are examined. Steam extraction contributes to lower mass flow through the steam turbine, thus resulting in lower power output.

2.4 Cooling System

When the steam exits the steam turbine the pressure is about 0,04-0,05 bar. Low pressure out of the turbine gives the highest power output, related to equation 2.1. At the same time, low pressure causes larger dimension of the condenser and steam turbine[6]. Larger dimensions are linked to higher costs. There is a limit of the exit pressure, and it varies with type of cooling system for the exit steam. The low-pressure steam out of the steam turbine is condensed before it is pumped up to a higher pressure. To reject the heat of condensation from the steam, a cooling system is needed. In the cooling system, a heat exchanger is used for condensing the steam by heat transferring from the steam to a coolant. The coolant is heated, and before it is returned to the heat exchanger it has to be cooled.

There are basically three different types of cooling systems[5].

- Direct/once-through water cooling
- Wet cooling tower with evaporative cooling
- Direct/indirect air cooling in an air cooled condenser

In the direct water cooling system water is taken from a water source like the ocean, a river or a cooling pond, and enters the condenser. The water is used as a heat sink for condensation of the steam, before it’s returned to the water source. This cooling system is normally the first choice for a power plant, because it allows the lowest condenser pressure[6]. It is also often the most economic system. The drawback for this method is that the power plant needs to be located near one of the water sources described.

If there is no water source available, a system with a wet cooling tower is a good option. The principle of the condenser heat exchanger is the same as for direct water cooling, but the difference is that the heated water needs to be cooled in a second process. This cooling takes place in a wet cooling tower. Often, a counter flow cooling tower is used, as presented in Figure 2.3. Unsaturated air is entering in the bottom of the tower and blows upwards. The hot water from the condenser heat exchanger is sprayed as droplets downwards in a counter-current flow with the air stream. Some of the water droplets are evaporated until the air is saturated with water vapor, and the saturated air is flowing out of the top of the tower. The not evaporated water droplets are cooled because the evaporation of the droplets' surface is taking place at the wet bulb temperature of the air[6]. By this reason, the wet bulb temperature of the air sets the limit of the water cooling and the condenser temperature. The cooled water droplets are collected in the bottom of the tower and sent back to the condenser heat exchanger. Because of the evaporation of some of the water droplets, make-up water has to be added continuously to the process. In the simulations of the steam cycle in this thesis, the wet cooling tower is used as the cooling system.

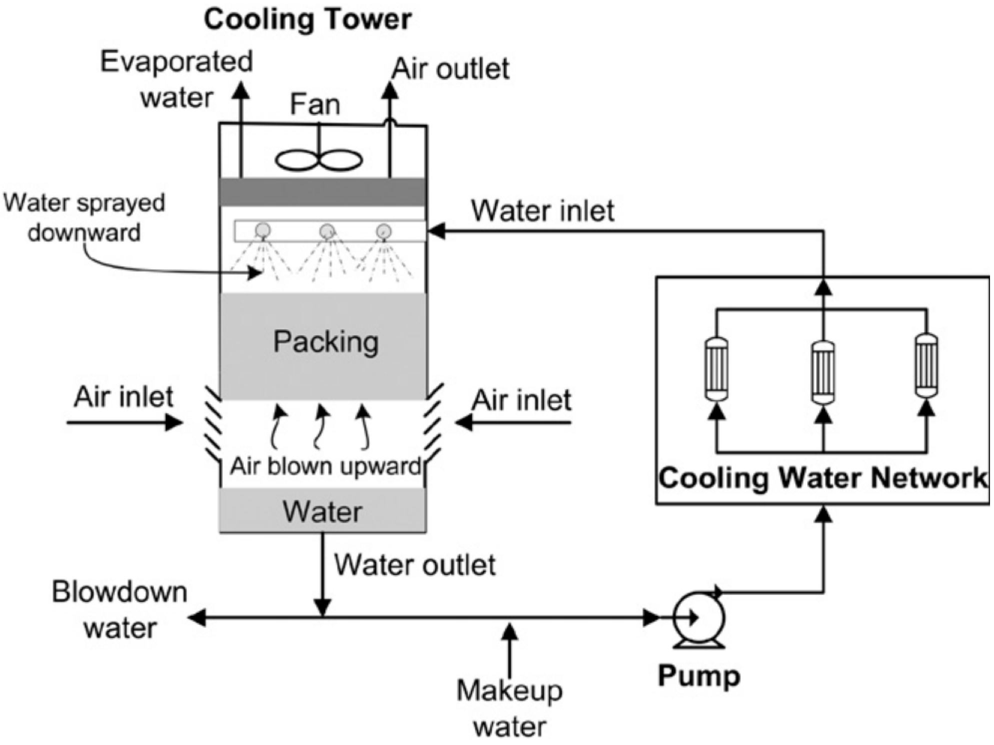

Figure 2.3 - Wet cooling tower[8].

An air cooled condenser has the advantage that there is no water needed for the process. This system is used when water is a not available source. The temperature of the steam/water in the condenser is limited by the dry bulb temperature of the air.

2.5 Feedwater Tank/Deaerator

The condensed water is pumped to a higher pressure and enters the feedwater tank. In the feedwater tank, the condensate is mixed with makeup-water. Makeup-water is added because of small leakages through the steam cycle. In the case of process steam extraction, returned process condensate is also fed into the feedwater tank. The water mix is often heat exchanged with the flue gas in the last part of the HRSG, and enters the deaerator. In the feedwater, dissolved oxygen and CO_2 are presented and causes corrosion and acid attack in the boilers. By introducing a deaerator, the level of O_2 and CO_2 in the feedwater can be controlled. The deaerator utilizes the principle of how gas solubility in a solution decreases when the temperature increases.

2.6 Steam Quality

Some of the energy in the steam expanded in the steam turbine is converted to power, but not all of the energy is converted. The work produced in the steam turbine depends on the enthalpy difference between the inlet and exit stream of the turbine, see equation 2.1. At the exit, the steam is at saturated conditions at the given condensation pressure. The steam/water is slightly condensed, with a vapor fraction of about 0,84-1[5]. Because of the high heat of evaporation for water, the enthalpy at the turbine exit is high compared to the condensed liquid at the same pressure. This is illustrated in Figure 2.4. A typical condensation pressure of 0,05 bar is indicated.

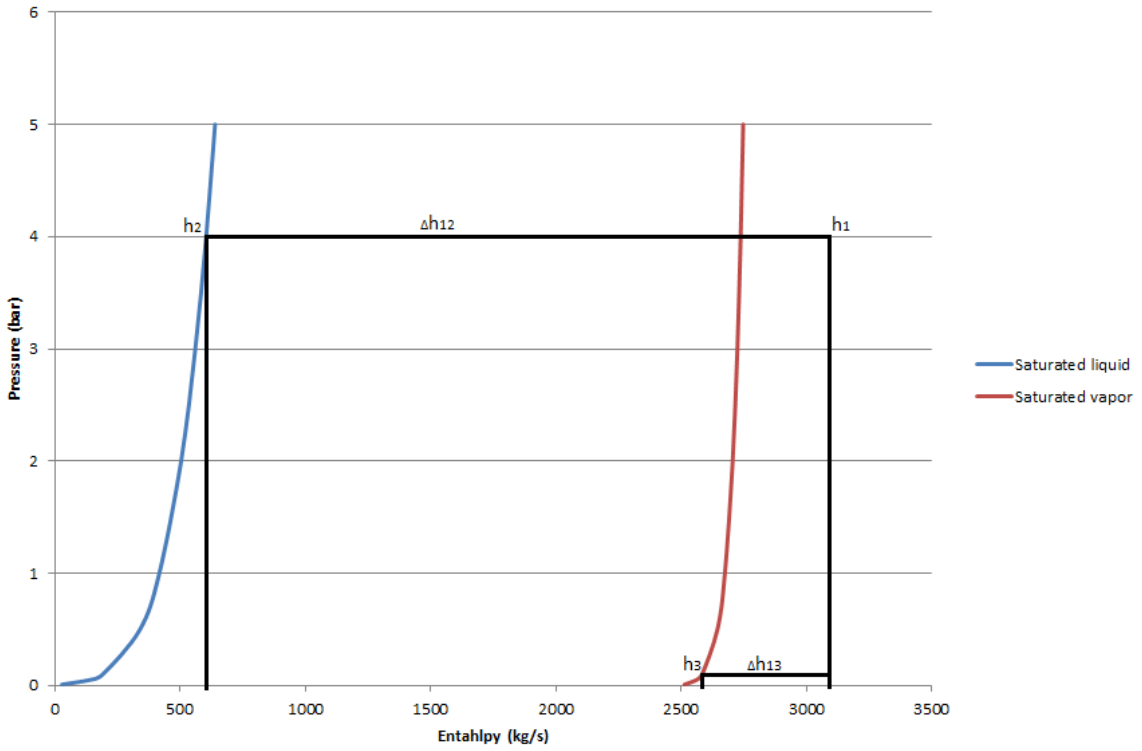

Figure 2.4 – Enthalpy curves for saturated water liquid (blue line) and saturated water vapor (red line) against pressure[9].

The relative high enthalpy at the turbine exit leads to a low enthalpy difference through the turbine, and therefore a low power production.

By extracting steam from the steam turbine, the steam could be used for heating purposes instead. The high heat of evaporation could be utilized in a heat exchanger process. For a given stream, the heat potential is given by equation 2.6 if the water is returned as saturated liquid:

$$\dot{Q} = \dot{m}(h_1 - h_2) \qquad 2.6$$

The enthalpies are indicated in the figure above for a constant inlet/extraction pressure of 3,9 bar, together with the enthalpy difference $\Delta h_{1,2}$. If the stream is not extracted, it could have been further expanded in the steam turbine. The power generated in the steam turbine, excluded mechanical losses, is:

$$\dot{W} = \dot{m}(h_1 - h_3) \qquad 2.7$$

The enthalpies are illustrated in Figure 2.4, and from the figure it is clearly that the heat potential is higher than the turbine work. This tendency is also the case for higher inlet/extraction pressure with a fixed condenser pressure. However, the ratio between the heat potential and the power generation is decreasing as the pressure increases. In Figure 2.5, it is illustrated how the saturated liquid enthalpy increase with pressure. If the condenser pressure is fixed, the enthalpy difference in equation 2.7 is growing more rapidly with higher pressures than the enthalpy difference in equation 2.6, indicating a decrease in the ratio.

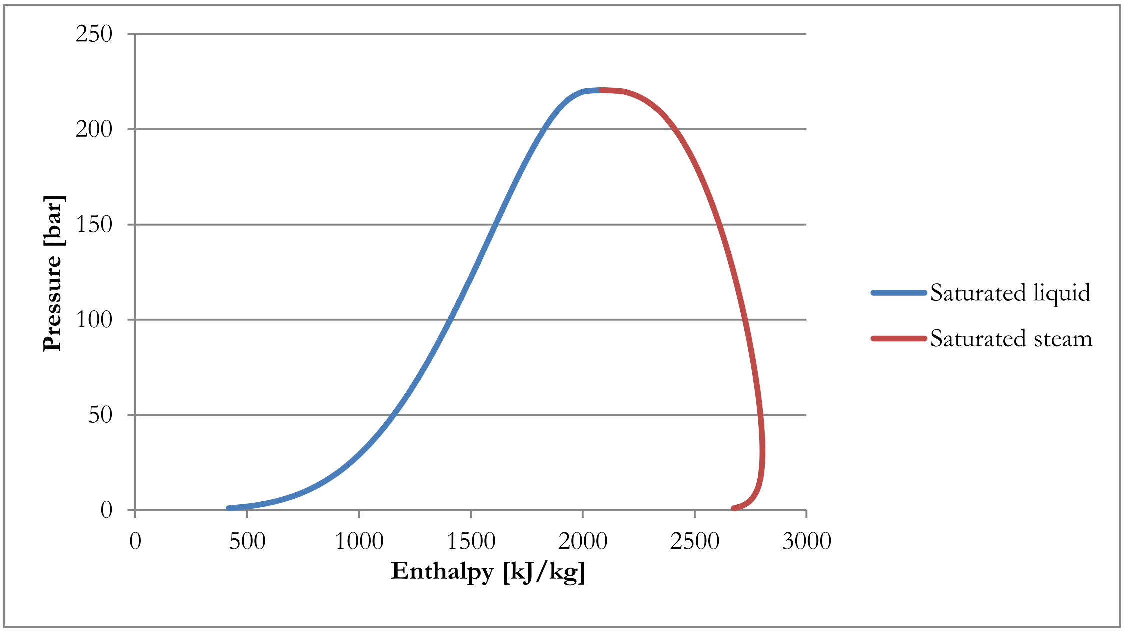

Figure 2.5 - Enthalpy curves for saturated water liquid (blue line) and saturated water vapor (red line) against pressure[9].

The lost turbine work because of steam extraction is defined as the change in generator terminal power output, with and without steam extraction[3]. This lost expansion work is the difference in power output between two steam cycles with the exact same design, apart from the extracted steam. The change in auxiliaries and losses is neglected. The lost power output in the steam cycle is:

$$\Delta\dot{W}_{lost} = \dot{W}_{no\ ext} - \dot{W}_{ext} \qquad 2.8$$

The ratio between the extracted process heat and the lost power production is defined as the steam quality, α.

$$\alpha = \frac{\dot{Q}}{\Delta\dot{W}_{lost}} \qquad 2.9$$

The α-value is preferred as high as possible, indicating high heat potential in the extracted steam compared to the lost power generation in the steam turbine.

2.7 Efficiency

The efficiency of a power plant can be defined in many ways. In this thesis, the efficiency of a NGCC is defined as net power output divided by the LHV of the fuel multiplied with the consumption of fuel.

$$\eta_{NGCC} = \frac{W_{net\ output}}{LHV * \dot{m}_{fuel}} \qquad 2.10$$

3 Capture Technologies

3.1 Absorption

The absorption process is a well-established process in the gas processing industry; it is used in the natural gas sweetening process for removal of CO_2 and other acidic gases. The main difference from the gas production to the flue gas cleaning is the pressure level. Natural gas sweetening takes place at a pressure level of around 60 bar, and in post-combustion capture the pressure is atmospheric 1.013 bar[10]. Because of the high pressure in the gas from the well the CO_2 has a high partial pressure which enables use of both pressure and temperature swing for desorption of the CO_2.

In all absorption processes gas is absorbed by a reactive liquid solvent, often called absorbent. There are many different types of absorbent that are being used for removing CO_2; amines, carbonates, calcination reactions or amino alternatives as amino acid salts and ionic liquids[11]. The most common solvent is an aqueous mixture that includes an amine. An amine consists of ammonia, hydrogen and one or more organyl group. There are three different groups of amine solvents, primary amine which has one of the hydrogen replaced, a secondary amine have two of the hydrogen replaced, when all of the hydrogen's have been replaced it forms a tertiary amine. Primary and secondary amines are characterized by fast reaction kinetics and form carbamates of varying stability, they also have a maximum loading capacity of 0,5 mol CO_2/ mol amine. Tertiary amine have slower reaction kinetics and can form bicarbonates in the presence of water, they can have a loading of up to 1 mol CO_2/mol amine[3].

Of the amines that are used the primary amine monoethanolamine (MEA) are most often preferred for post-combustion capture. This is because have a low heat of absorption[1]. MEA is also preferred because it is suited for gas streams containing a low concentration of CO_2, typical in the area of 0.03-0.15 wt. %[3]. MEA reacts with the CO_2 and forms carbamate (reactions 4.1). This reaction takes place with a rather low temperature 40-60 °C. To release the CO_2 from the MEA the rich solution goes through a temperature swing, which means heating it up to around 100-150 °C[12]. This takes place in the stripper column.

$$C_2H_4OHNH_2 + H_2O + CO_2 \rightarrow C_2H_4OHNH_3 + HCO_3 \qquad 3.1$$

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[1] Heat of absorption: The heat necessary to break chemical bonds between the solvent and the CO_2 (heat of reaction) and to drive out the CO_2 from the liquid (heat of dissolution).

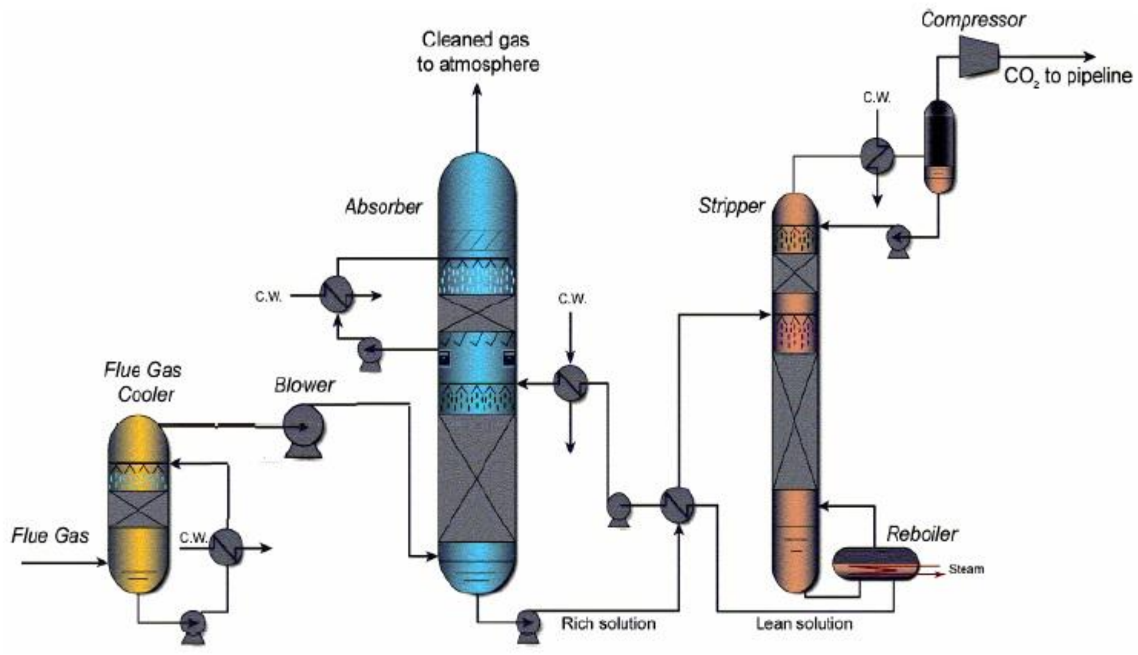

Figure 3.1 - Flow sheet of post-combustion CO2 capture plant[11].

The amount of MEA being used may vary from 15-30 wt. %[13]. Due to the high corrosiveness of MEA only low wt. % can be used, but with effective corrosion inhibitors in the system the wt. % of MEA may come up to 30. If the flue gas contains SO_2, NO_2 or high levels of O_2 degradation of the MEA may occur. Degradation of the solvent reduces its loading ability, and may even destroy it completely.

When capturing CO_2 at elevated pressure, a different amine than MEA is more suited methyl diethanolamine (MDEA). This is a tertiary amine so it has a much lower heat of reaction and low vapor pressure. There are many other advantages by using MDEA, e.g. aren't as corrosive as MEA which gives the possibility of using amine solutions with wt. % up till 55. It also has a much lower degradation than MEA. As mentioned before temperature swing are used to strip the CO_2 from the amine, but with the use of MDEA a combination with pressure swing are often utilized as well. The drawback by use of MDEA is the slow reaction rate for capture of CO_2. This may result in a need for a bigger absorber because the amine requires more time to react with the CO_2 in the column. [14]

3.2 Absorption Process

The flue gas coming from the power plant has a temperature in the range of 80-100 °C[3], the pressure my vary for different power plant configurations. The temperature should not be too low because of condensation of water in the pipes prior to the capture plant. This gas normally contains 3-4 mol. % of CO_2[15]. The rest of the gas is a variation e.g. O_2, N_2 and H_2O. In some flue gases traces of NO_X and SO_2 may be found, these components needs to be removed by pretreating the gas before it enters the absorption column. That's because they may cause high degradation of the solvent. The flue gas needs to be cooled down to the desired temperature of 40-60 °C. This is done by a flue gas cooler, and in the cooler liquid is condensed out of the gas.

The liquid is mostly water, but it is possible to find traces of CO_2 in it as well. Pressure drop will occur at the inlet of the absorption column, to prevent this from affecting the turbine in the power plant a blower is installed to maintain an even pressure level. In the absorption column (see Figure 3.2 for a detailed schematic) the flue gas is mixed with the water and the amine mixture. The absorption column is divided into different trays. On the top tray the solvent is sprayed over structured packing or random packing[16], which is designed to have a very high area per volume (typical 150 m2/m3[17]). Because of the high area/volume all the gas that flows through the column comes in contact with the solvent. The gas flows into the bottom of the absorber and the solvent at the top. The goal of an absorber is to have as much of the CO_2 absorbed by the solution, therefore a rich loading as high as possible. The better rich loading and absorption that is achieved, the lower circulation rate of amine is needed and thus lower reboiler duty[2][18]. After the solvent has gone through a number of trays, liquid collectors are installed to collect and redistribute the solvent over the next tray. Liquid collectors are only used in high columns[3]. At the bottom of the column the CO_2-rich solution leaves with a temperature of 40-60 °C. The clean gas leaves the absorber in a vent at the top of the column. This type of absorption column can get an efficiency of over 90%[19] with the right operation condition and design.

The rich solution is pumped through a heat exchanger before it goes into the stripper column. In the heat exchanger the rich solution is heated up with the warm lean solution (110-120 °C) leaving the stripper column. By using this rich/lean heat exchanger the rich solution can be heated up to around 100-110 °C before entering the stripper. That energy is saved reboiler duty. Another benefit from having this rich/lean heat exchanger is that the necessity of cooling water for cooling of the lean solution is reduced.

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[2] Reboiler duty is presented as energy required for capturing per kilogram of CO_2.

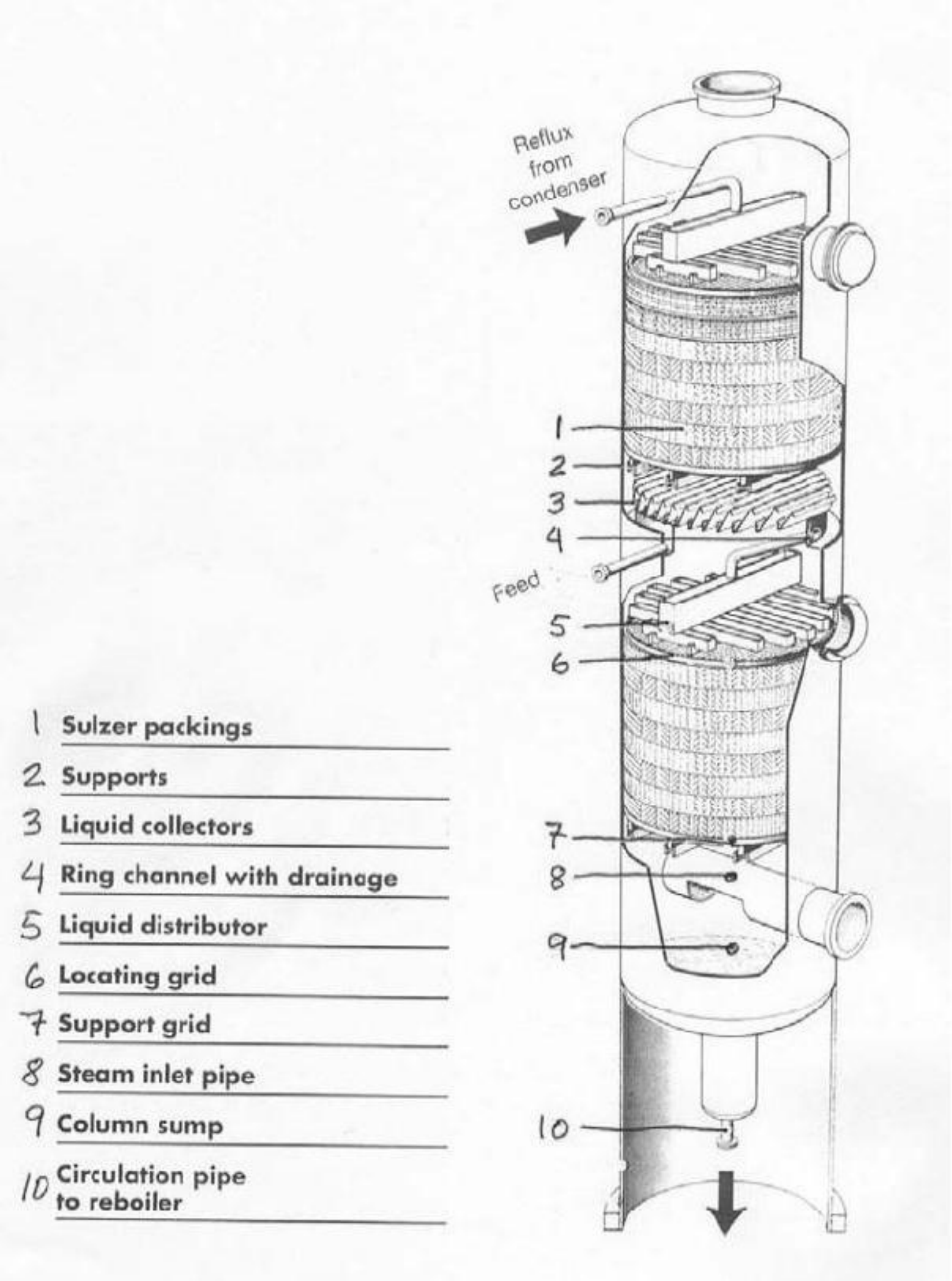

Figure 3.2 - Absorption column schematic[20].

3.3 Stripper

Before entering the separator the rich solution is throttled down to around 2 bar. The desorption process works the opposite way than the absorption process. Here heat is added to release the CO_2 from the amine. The separator works as a distillation column and it is also here important to have a high gas/liquid ration, because this gives a good contact between the rich solution and the warm gas. The main concept of the separator is the heating of the rich solution. The rich solution flows downward in the column and the steam rises up. The heat is produced in the reboiler by heat exchanging with the superheated steam from the HRSG (see chapter 2.2). The stripping of the CO_2 from the amine happens in the reboiler. The warm rich solution comes into the boiler as liquid, and there it is boiled and the CO_2 is therefore released from the amine. The boilup from the reboiler is sent back into the separator to heat up the rich solution. The liquid from the reboiler is called the lean solution and is transported back to the absorption column (see Figure 3.1). The temperature of the lean solution leaving the desorber is typically around 110-120 °C, but it should not exceed 122 °C because then the degradation and corrosion may become intolerable[21].

3.4 Energy Demand

The process of capturing CO_2 from flue gas is very energy consuming. The most demanding part is the reboiler. That is because the amine require a temperature in the range of 110-120 °C to boil and to release the CO_2 the amine needs to be brought to the boiling point. The energy use in the reboiler can be divided into three parts, see equation 3.2[3].

$$Q_{reb} = Q_{solv} + Q_{vap,H_2O} + Q_{des,CO_2} \qquad 3.2$$

Q_{reb} = Total heat demand in the reboiler to regenerate the solvent.

Q_{solv} = Heat to raise the solvent from inlet stripper temperature to reboiler temperature.

$Q_{vap, H2O}$ = The heat required to evaporate the water in the stripper.

$Q_{des, CO2}$ = The heat of absorption of the solvent with CO_2. In the stripper heat equal to the heat released in the absorber needs to be supplied back to the solvent in the reverse process.

Many other parts of the process require energy. There are pumps to pump the rich/lean solution and cooling water around, and the process also includes a blower before the absorber. This blower is the second most energy consuming unit.

3.5 Loading Capacity

The loading of the solvent are defined by how much CO_2 is captured in the solvent per amount of solvent in the stream (mol CO_2/mol amine). A typical loading curve can be seen in Figure 3.3, this figure shows that for a partial pressure of 0,03-0,04 bar the loading of the solvent should be around 0,5-0,6 for the rich stream and 0,1-0,2 for the lean stream.

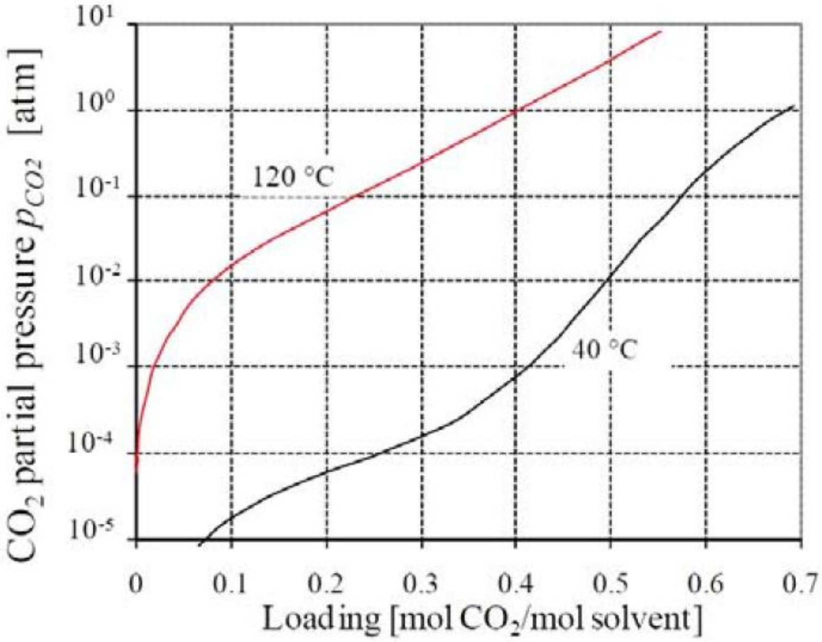

Figure 3.3 - Solubility of CO_2 in aqueous alkanolamine[3].

3.6 Cooling Water

The CO_2 capture process requires a lot of cooling water. There is need for cooling of the flue gas in the flash tank prior to the absorption column and in the lean solvent stream before it re-enters the absorption column. There is also a large cooling demand in the overhead condenser in the stripper column. All these cooling demands are being met by the usage of water or air. A typical cooling circuit is presented in Figure 3.4. Each circuit has a pump and the pump require mechanical work to be driven.

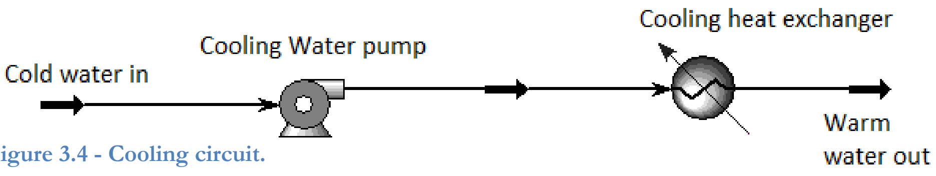

Figure 3.4 - Cooling circuit.

To calculate the amount of cooling water needed, the following equation can be used:

$$\dot{m}_{cw} = \frac{\dot{Q}_{cw}}{h_{cold,out} - h_{cold,in}} \qquad 3.3$$

Equation 3.3 requires that the pressure and temperature before and after the cooler is known. With these the enthalpy can be found and the amount of cooling water needed can be calculated. In this thesis the cooling water temperature is set to an inlet temperature of 15 °C and a maximum an outlet temperature of 25 °C.

3.7 Absorber Size

The components in a capture plant tends to be very large, especially the absorber. When the main focus of a NGCC with CO_2 capture is the total efficiency of the plant, the size of each component tends to get very big, but when economics comes into consideration the size becomes important as well because bigger e.g. absorbers costs more to produce and maintain. To calculate the height and diameter of the absorber the following equations can be used:

$$d = \sqrt{\frac{4\dot{m}_{fg}}{\pi \rho_{fg} \mu_0}} \qquad 3.4$$

Where:

$\dot{m}_{fg}$ = Flue gas mass flow rate into absorber

ρ_{fg} = Density of the flue gas

μ_0 = Superficial gas velocity (in these thesis assumed to be 4 m/s[3])

For the height:

$$H = \int_{y_1}^{y_2} \frac{-G}{K_G a P (y - y_{eq})} dy = \frac{G}{K_G a P} \ln\left[\frac{y_1 - y_{eq}}{y_2 - y_{eq}}\right] \qquad 3.5$$

Where:
G = the molar flux of the flue gas, and is defined in equation 3.5.
K_G = Total mass transfer coefficient
A_c = Cross-sectional area of absorber
P = Pressure
a = Gas-liquid transfer area divided by the column volume
y = Mole fraction of CO_2 in the inlet (1) gas and outlet (2) gas
y_{eq} = Equilibrium mole fraction for CO_2 between gas and liquid

To calculate the height of an absorber are a complicated process, to make it easier certain assumptions have been made. According to Aroonwilas et al [22] a value of 0,02 for $K_g a$ can be assumed and from Bolland [3] the y_{eq} can be used as a constant of 0,001.

$$G = \frac{\dot{m}_{fg}}{MW_{fg}A_c} = \frac{4\dot{m}_{fg}}{MW_{fg}d^2\pi} \qquad 3.6$$

Where:
MW_{fg} = Molecular weight flue gas
A_c = Cross-sectional area

These equations will only give theoretical values of the absorber size, but they can be used to compare the different processes with each other. The calculations of the real height of the absorber require a lot more and advanced equations.

3.8 CO_2 Compression

After the CO_2 has been released from the amine in the desorber it needs to be compressed before it can be transported to storage. The best way to transport and store CO_2 is when it is in liquid form. The CO_2 leaving the stripper has a pressure of 2 bar and a temperature of 30 °C. For the CO_2 to be in liquid form at ambient temperature the pressure needs to be above 80 bar (see Figure 3.5). When transporting CO_2 in pipes there are always pressure losses and to prevent two phase flow in the pipes the CO_2 is normally compressed to 100-150 bar[3]. The number of compression stages that is used in this thesis is a 4 step process with 3 compressors and a pump. The pressure ratio is 2-4 per compression. After each compressor a water cooler is used to cool down the CO_2 to 25 °C. Two flash tanks are used to flash out water from the CO_2 stream. See appendix 12.1, Figure 12.1 for a detailed schematic of a compression process.

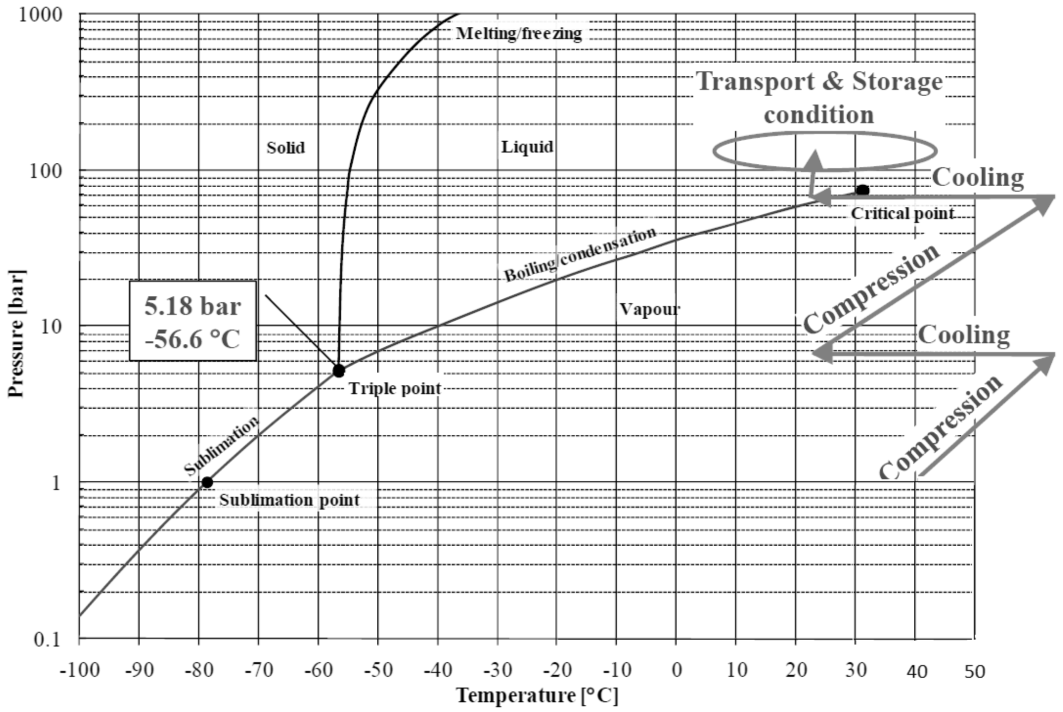

Figure 3.5 - Gas-phase separation[3].

4 Literature Study

This chapter presents different process done in the field of CO_2 capture from power plants with flue gas at elevated CO_2 partial pressure. Here different power plant models for capture are evaluated and described. The focus here will be on process that has the goal of increasing the partial pressure of CO_2 in the flue gas and decreasing volumetric flow rate into the absorber. The cases evaluated presents different ways of accomplishing these problems. The processes that have been evaluated are: postcompression capture of CO_2 in gas turbine, two different cases called Combicap and Sargas. There will also be looked into a way of increasing the CO_2 concentration by burning extra fuel at the turbine outlet, and a process which uses separate gas turbine and combustion chamber.

4.1 Sargas

The Sargas process has been developed by Sargas AS, the flow sheet of the process is presented in Figure 4.1. The process uses pressurized fluidized-bed combustion (PFBC), and is designed to use coal, but by changing the PFBC to a pressurized combustion chamber with steam tubes natural gas can also be used as fuel. The fuel used in the initial design is a coal-water slurry, the mixture is mixed with air in the bed that has a pressure around 12-16 bar[23]. The pressurized air is mixed with the fuel in the PFB combustor, the flue gas is then take into the CO_2 separation unit. Prior to the separation the flue gas goes through a recuperator heat exchanger (RHE) with a temperature of 850 °C and is cooled down to around 200 °C by use of the clean gas from the capture unit. Before capture can take place the gas needs to be further cooled down to 90 °C. After the flue gas have been cleaned it needs to be heated up again before entering the gas turbine, by use of the RHE the clean gas is heated up to around 814 °C. The RHE is the most critical units of this technology because it has a very high thermal loading and it is also the only link between the gas turbine and the combustion chamber. Should it shut down the capture unit will need to be by passed. Heat exchangers operating at high temperature are very expensive to produce and they require robust constructions, they also need to be replaced more often than lower temperature heat exchangers. The capture unit uses a hot potassium carbonate process to capture the CO_2. The CO_2 concentration in the flue gas is 16,4 % by volume. The manufacture claim the energy demand of the CO_2 capture is as low as 1,6-2,1 MJ/kg CO_2. Technical data from the Sargas process is presented in Table 4.1.

Feature (100 MW unit block)	Value	Comments
Gross power generated [MW]	105	
Fuel input [MW]	267	
Overall plant efficiency (HHV) – [%]	39,3	Not accounting for CO_2 pressurization
Overall net plant efficiency	36,3	CO2 pressurization taken into consideration
Fuel supply [kg/s]	7,5	Dry coal. HHV equals 33,9 MJ/kg
Unit power [MWe]	100	Full plant 400MWe requiring 4 unit blocks
CO_2 generated [kg/s]	26,5	
CO_2 captured [kg/s]	23,9	Thus emitting 2,6 kg/s CO_2
Power for CO2 pre-treatment [MW]	8,2	Based on 94,5 kWh per tonne CO_2
CO_2 capture rate [%]	90,4	

Table 4.1 – Technical data from the Sargas process [24].

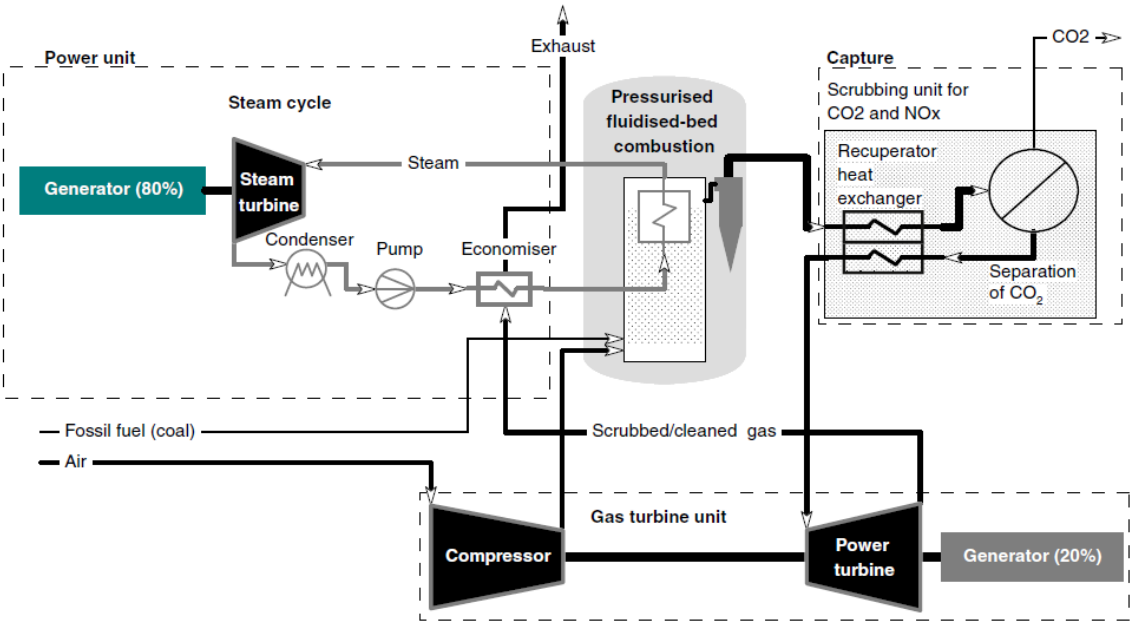

Figure 4.1 - Simplified outline of the Sargas technology[24].

4.2 Combicap

This is a process developed by Statoil in 2004, and it utilizes CO_2 capture at elevated pressure. The process consists of two gas turbines, two HRSG, a steam turbine and a combustion chamber that also works as a heat exchanger. In the combustion chamber compressed air is burned with natural gas and heat is transferred from the combustion to heat up the clean flue gas from the CO_2 capture plant. The gas from the combustor is then expanded in a turbine and then the remaining heat is used to produce steam in the HRSG. The exhaust gas is then compressed and some of it is recycled back into the combustion chamber. The remaining flue gas is cleaned in the capture plant; thereafter it is heated up by use of heat from the combustion chamber. The warm flue gas is taken through a turbine and then used in the last HRSG to produce steam. The process schematic can be seen in Figure 4.2.

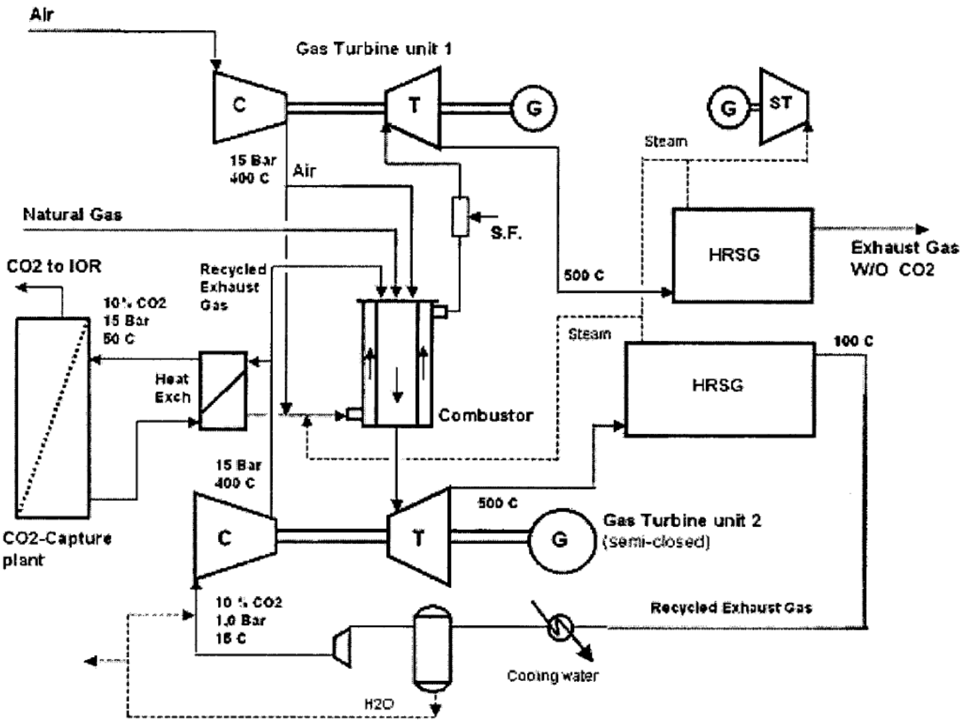

Figure 4.2 – Schematic of the Combicap cycle[25].

Finkenrath et al[26] have evaluated the Combicap process, the results are presented in Table 4.2. For the evaluation these assumptions are made:

- Membrane capture does not require extra energy.
- For absorption with amine, 90 % of the CO_2 is separated, reboiler duty 2,4 MJ/kg CO_2 and a total energy demand of 2,9 MW.

CO_2-capture	Total efficiency [%]	Capture rate of CO_2 [%]	Power [MW]	CO_2 emission [g/kWh]
Membrane	52,9	68,3	986	118,7
Amine	50	68,3	935	125,5

Table 4.2 – Results from the Combiecap process.

4.3 Postcompression Capture of CO_2 in Gas Turbines

The idea behind this process is to use the air compressor to increase the pressure of the gas, thereby increasing the CO_2 partial pressure, so the penalty of separating the CO_2 is decreased. Flue gas from the HRSG will be recirculated back to the air compressor as in the prior processes. A flow sheet of the process can be seen in Figure 4.3. The capture plant is located after the air compressor; this will give the gas a much higher pressure than the normal location of the plant would give, which is around atmospheric pressure. Before the gas enters the plant a heat exchanger is used to cool the gas using the returning clean gas. This is a disadvantage with this process, because the high temperature after the compressor, results in a need of very large heat exchangers. This drawback can be disregarded if capture processes with high temperature are utilized.

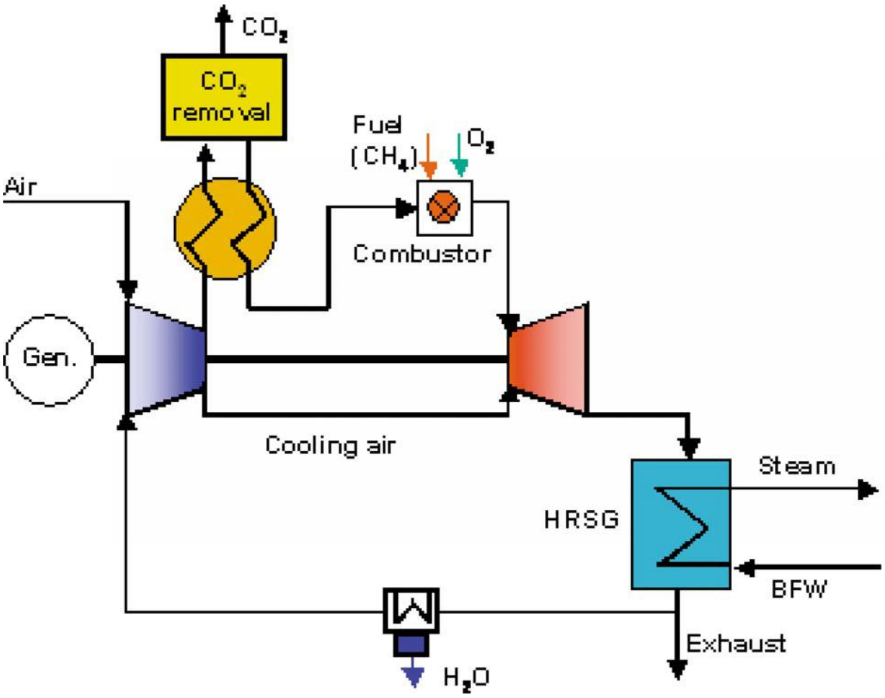

Figure 4.3 – Postcompression CO_2 capture[27].

This study has 4 cases, where the amount of CO_2 is varied as well as the supply of O_2 in the burner. The results are presented in Table 4.3. The benefit of recycling fuel gas is lost when the removal rate of CO_2 in the separation unit is high. Since case 3 and 4 uses enriched fuel the EGR ratio can become rather high compared to other processes, the cost for increasing the O_2 concentration in the air will decrease the overall efficiency of the plant.

Case	O_2 supply	CO_2 removed in separation unit (%)	Total CO_2 removed (%)	CO_2 molar fraction entering separation unit
1	Air	80	61	0,04
2	Air	20	26	0,08
3	Enriched air	80	93	0,05
4	Enriched air	20	53	0,14

Table 4.3 - Postcompression CO_2 capture: Effect of CO_2 removal rate[27].

4.4 Supplementary Fired Cycles

The difference between a NGCC and supplementary fired cycles(SFC) is that the SFC uses an extra combustion at the turbine outlet. This is done in order to improve the CO_2 concentration of the flue gas. The CO_2 concentration increases but so does the temperature leaving more heat available for the steam cycle to utilize. A schematic of a SFC is presented in Figure 4.4. Since burning extra fuel an extra turbine is also installed, and the flue gas from the first turbine is reheated by use of the combustion at the turbine outlet before it enters the second turbine. The usage of extra fuel will decrease the net plant efficiency. Simulations done by Li et al[28] shows that by use of this technology the overall plant efficiency will be at 48,1 %. The CO_2 concentration in the flue gas will increase from 3,8 to 6,7 mol-%, but the use of supplementary firing will increase the mass flow into the capture plant instead of lowering it as EGR. This will result in large size of the capture plant as well as more mechanical work from pumps etc.

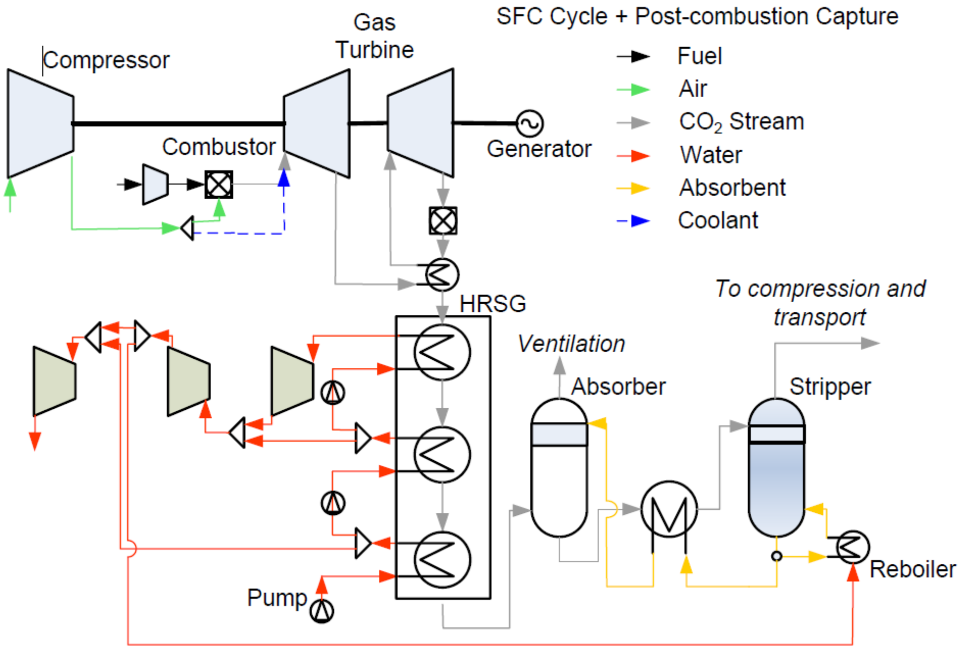

Figure 4.4 - Schematic of a supplementary fired gas turbine cycle[28].

4.5 Externally Fired Gas Turbine Cycles

An externally fired gas turbine cycle (EFC) uses technology where the combustion chamber is located outside the GT. The burned gas from the combustion is heat exchanged with the compressed air in a heat exchanger, so there is no direct contact between the GT air flow and the flue gas from the combustion. There are many benefits from this type of combustion, firstly the combustion chamber does not need to be pressurized as in normal GTs and the problem of unstable combustion from low O_2 concentration will not affect the GT. Another benefit is the possibility of using of so-call "dirty" fuel, which is cheap but cannot be used in normal GTs. Since the power generation and the combustion chamber are separated, the combustion does not need to have as high excess air ratio as normal. Therefore the flue gas will be less diluted and there will be lower flue gas to be treated in the absorber. With stoichiometric combustion the CO_2 concentration will be 11,8 mol-% and the mass flow as low as 17,1 kg/s. The electric efficiency at these conditions will be 44,3 % for this process.

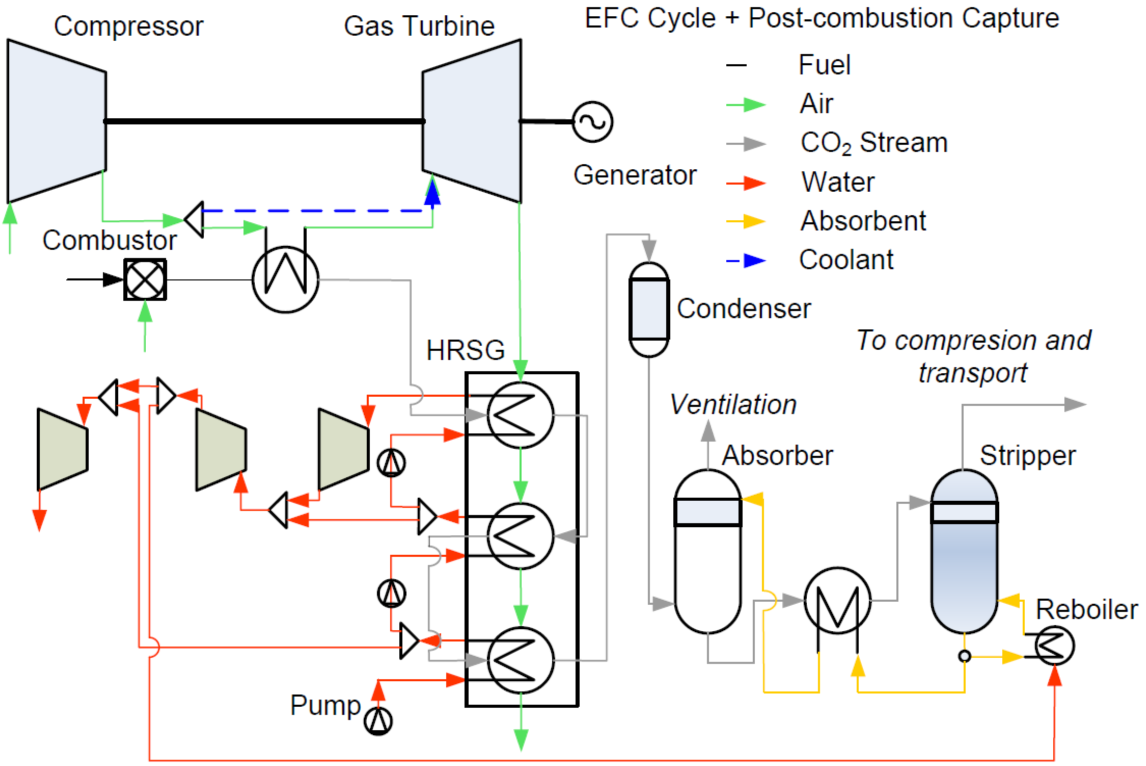

Figure 4.5 - Externally fired gas turbine cycles[28].

4.6 Conclusion

In Table 4.4 the results from the literature study are presented. Here the benefit from each of the processes are graded on how well they improved the CO_2 concentration in the flue gas as well as reducing the flue gas flow into the absorber. The challenges with usage of the technology are presented and the ability to produce a high net plant efficiency.

	Sargas	Combicap	Pcc	EFC	SFC
Increased concentration	-	-	+	+++	++
Increased total pressure	+++	+++	+++	-	-
Reduced volumetric flow	+++	+++	-	+++	-
Technological challenges	+++	+++	++	++	+
Possibility of achieving high net plant efficiency	++	++	+	-	-

Table 4.4 – Summary of the different technologies.

From Table 4.4 the Sargas and Combiecap technology scores high on the possibility of achieving high net plant efficiency, this is closely linked with the fact that these processes use equipment's and technologies that is immature, and by this scoring very high on the technological challenges part. The externally fired cycles and supplementary fired cycles both have low net plant efficiency but both are less complicated to use and have great benefit on the partial pressure of CO_2. The process which reduces the flow rate to the largest extent are the externally fire gas turbine.

5 Process Integration

In this chapter there have been looked into ways of integrating the power plant and the capture plant. Since the stripping process requires a lot of steam, way of extracting steam from the steam cycle will be evaluated. The energy required to separate CO_2 from the flue gas are heavily linked with the partial pressure of CO_2 in the gas. Modifications to the power plant to include exhaust gas recycle (EGR) will be evaluated in this chapter.

5.1 Steam Extraction

The reboiler duty related to the capture process is high. By integrating the heat demand into the power plant, a lot of energy could be saved. The steam in the steam cycle could be used for other purposes than generating power in the steam turbine. Because of the high heat of evaporation for water, steam is superb as a heating source. For that reason it is preferable to use the steam in the steam cycle to deliver the heat demand to the reboiler. There are several options for integration. In this thesis it is focused on steam extraction from the crossover between the IP/LP turbines. The affection of steam extracting on the power plant's performance is analyzed in terms of varied parameters.

5.1.1 Extraction from Crossover

In a three pressure level steam cycle, there is a crossover pipe that connects the IPT and the LPT. The steam expanded in the IPT flows through the crossover, is mixed with steam produced in the low-pressure cycle and expands further in the LPT. At this crossover it is possible to extract steam. The steam is extracted upstream of the low-pressure steam mixing, see Figure 5.1.

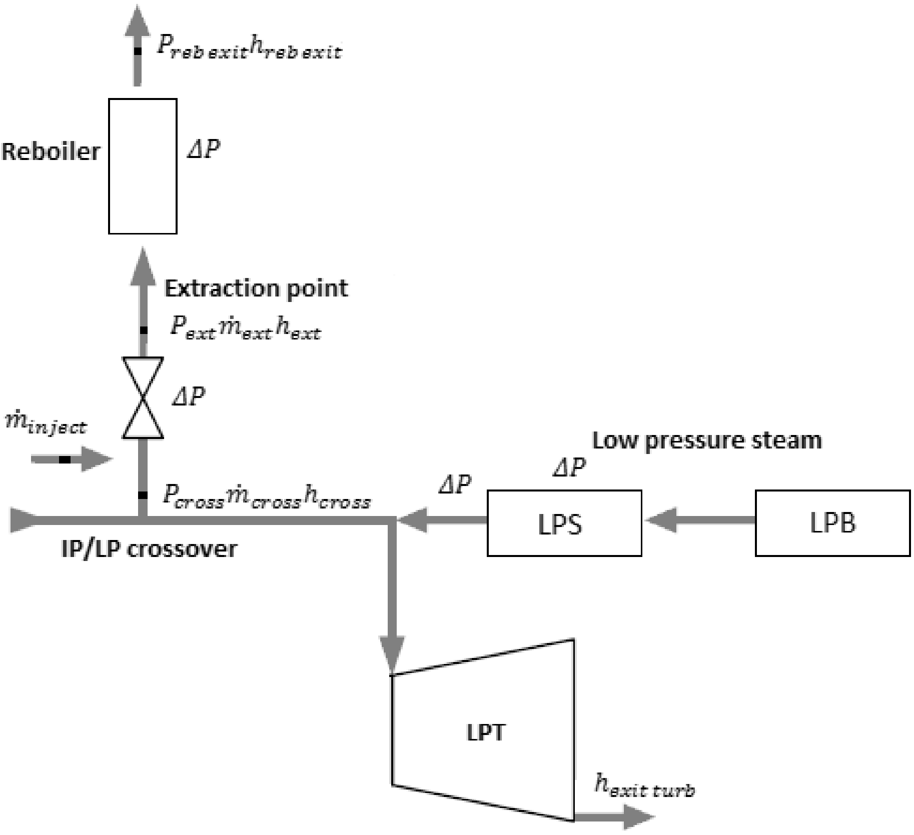

Figure 5.1 – Illustration of steam extraction from IPT/LPT crossover.

The steam at the crossover is superheated, and in this work additional water from the deaerator is injected in order to saturate the steam. This will be explained in more details later. In the piping between the crossover and the reboiler, pressure loss occurs. The extraction point defined in this thesis is in front of the reboiler, with saturated steam conditions, see Figure 5.1.

When extracting steam from a crossover, it is possible to design the crossover to exactly match the preferred extraction pressure. How to determine the preferred extraction pressure is explained in chapter 5.1.1.2.

A disadvantage by extracting steam from the crossover is the change in the low-pressure boiler (LPB) pressure. The pressure in the LPB is designed in order to match the pressure in the crossover, and is given by equation 5.1.

$$P_{LPB} = P_{crossover} + \Delta P_{loss} \qquad 5.1$$

ΔP_{loss} is the pressure losses in the piping and the heat exchangers between the LPB and the crossover. If the extraction pressure is high, the crossover and the LPB pressures are high. From the theory about the HRSG in chapter 2.2, less steam is produced in the low-pressure cycle when the LPB pressure is high and the other conditions are held constant. Less steam produced could contribute to a lower power generation in the steam turbine, but also other factors as enthalpy affects the power generation.

5.1.1.1 Water Injection

Steam extracted from the crossover is superheated and at a high enthalpy level. The steam that is extracted would otherwise generate power in the last expansion process in the steam turbine. The power generated in a steam turbine, apart from mechanical and generator losses, is given by the conservation of energy principle. By assuming no heat loss, the expression becomes:

$$\dot{W} = \dot{m}(h_{inlet\ turb} - h_{exit\ turb}) \tag{5.2}$$

The lost work related to steam extraction at the IPT/LPT crossover is expressed as:

$$\Delta\dot{W}_{lost} = \dot{m}_{cross}(h_{cross} - h_{exit\ turb}) \tag{5.3}$$

One option in order to reduce the power loss is to reduce the extracted mass flow from the turbine. The steam is superheated, and water at a lower enthalpy level could be injected to the steam in order to cool the steam to saturated conditions. As stated in equation 5.4, the required mass flow at saturated conditions at the extraction point is the sum of the extracted steam from the crossover and the water injected.

$$\dot{m}_{ext} = \dot{m}_{cross} + \dot{m}_{inject} \tag{5.4}$$

5.1.1.2 Preferred Extraction Pressure

The heat demand is significant in the reboiler in a CO_2 capture process. If steam from the IPT/LPT crossover is extracted and saturated with water, the necessary steam mass flow in the extraction point is given by equation 5.5.

$$\dot{m}_{ext} = \frac{\dot{Q}_{reb}}{(h_{ext} - h_{reb\ exit})} \tag{5.5}$$

The reboiler duty is fixed for a given capture process. From the equation 5.5 it is seen that only the enthalpy difference affects the mass flow at the extraction point. For the reboiler, the most common exit condition is saturated condensate. Accordingly; the enthalpy difference is the heat of evaporation for water at the extraction pressure, included a small modification. The pressure loss of 0,5 bar through the reboiler results in a bit lower condensate enthalpy at the outlet of the reboiler. The mass flow expression becomes:

$$\dot{m}_{ext} = \frac{\dot{Q}_{reb}}{(h_{ext} - h_{cond})} \tag{5.6}$$

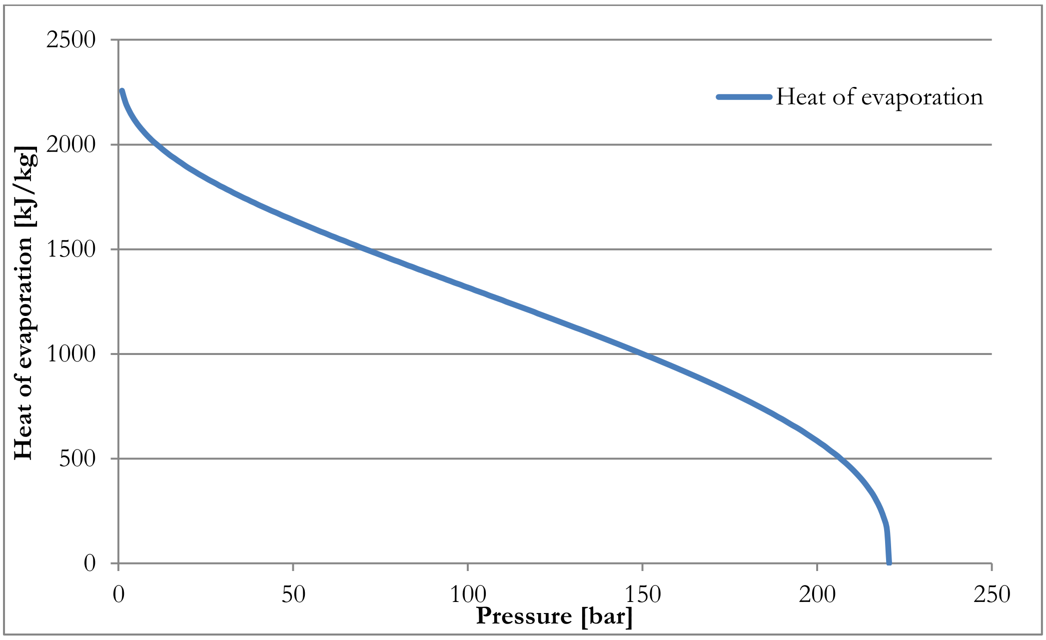

Figure 5.2 - Heat of evaporation for water[9].

As the figure above illustrate, the heat of evaporation for water decreases with higher pressure. With a decreasing enthalpy difference, the required mass flow at the extraction point increases for a given reboiler duty. However, an increase of the total mass flow is not the same as an increase of the extracted steam from the crossover. Simulations done by GT PRO shows that an increase of extraction pressure gives a small decrease of the extracted crossover steam, while the mass flow of the injected water increases. The simulations are presented in Figure 5.3. This is related to the steams enthalpy upstream in the steam turbine. If the extraction pressure increases, the steam extracted from the steam turbine contains a higher enthalpy and is more superheated. More water is needed in order to saturate the steam.

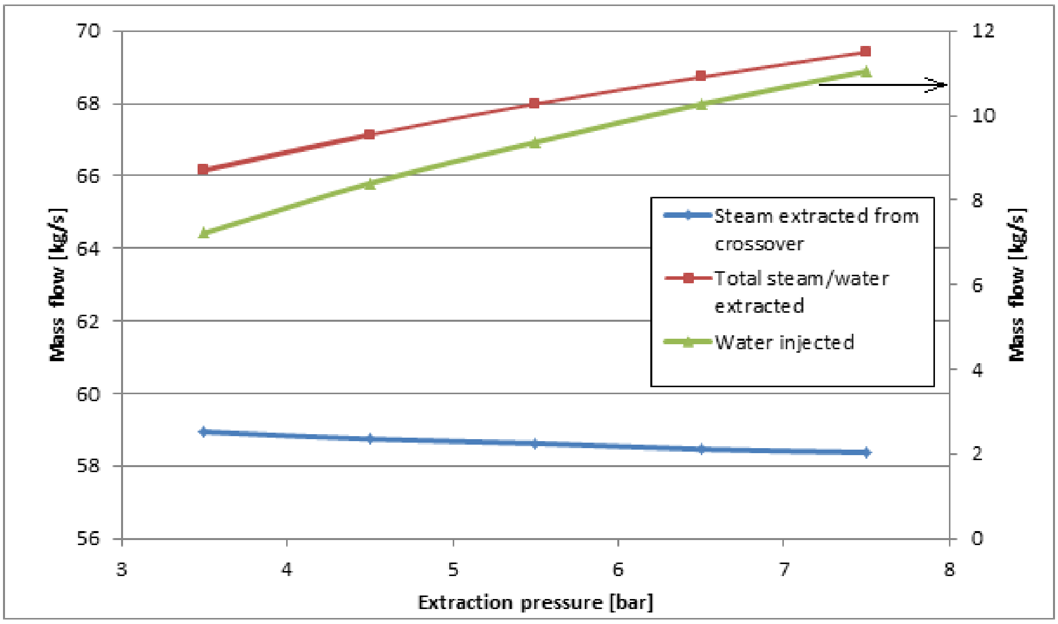

Figure 5.3 – Extracted mass flows from the steam cycle for different extraction pressures. The reboiler heat demand is constant. The diagram is developed from GT PRO simulations.

In equation 5.3, the lost turbine work due to steam extraction from the crossover is expressed. As the simulations in Figure 5.3 describes, the extracted mass flow from the crossover decreases slightly with increased extraction pressure. However, when the extraction pressure increases also the pressure at the crossover increases, and with that the enthalpy in the steam from the IPT. Figure 5.4 illustrates how the lost expansion work depends on the extraction pressure.

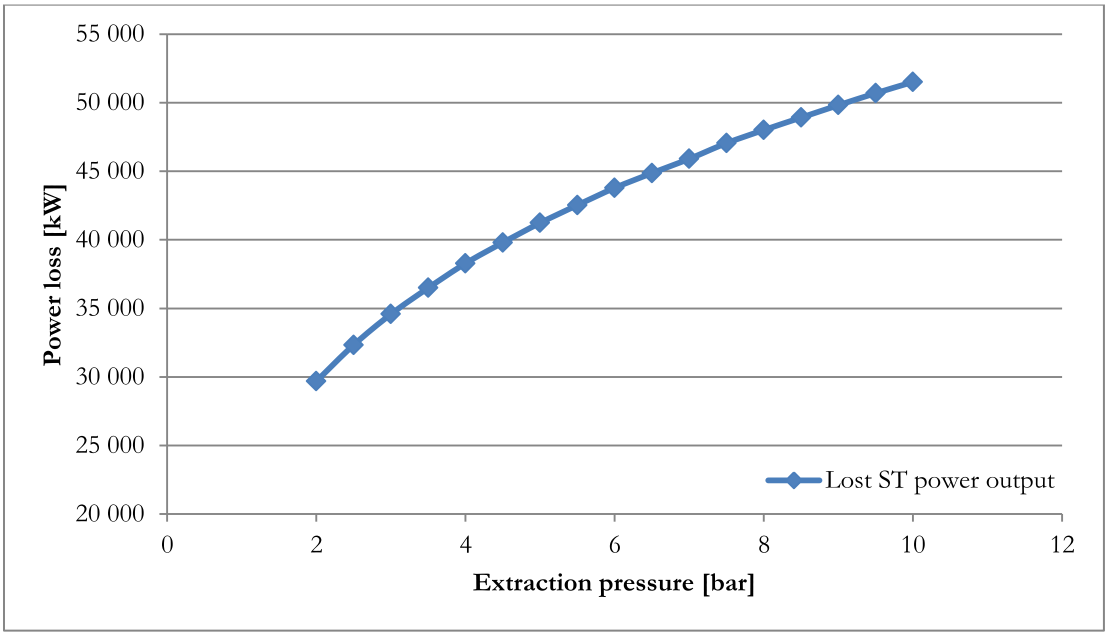

Figure 5.4 - Lost power in steam turbine for different extraction pressures. The reboiler heat demand is fixed. The diagram is developed from GT PRO simulations.

The lost work increases with pressure. In other words; the effect of the increased enthalpy is more dominating than the reduction in the extracted steam mass flow. The definition of the α-value is stated in equation 2.9. If steam is extracted from the IPT/LPT crossover, the α-value is:

$$\alpha_{cross} = \frac{\dot{Q}_{reb}}{\Delta \dot{W}_{lost}} = \frac{\dot{Q}_{reb}}{\dot{m}_{cross}(h_{cross} - h_{exit\ turb})} \qquad 5.7$$

In Figure 5.5, the α-value is plotted against the extraction pressure for a constant reboiler duty.

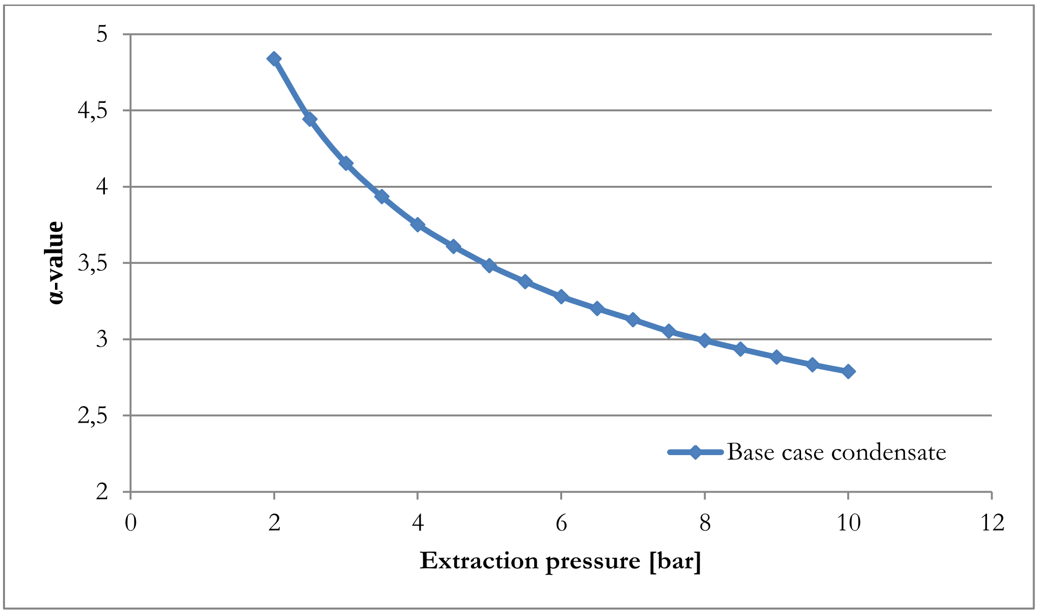

Figure 5.5 - Variation of α-value against extraction pressure, with a fixed heat demand in the reboiler. The graph is calculated from simulations in GT PRO.

As seen from the graph; in case of low extraction pressure the α-value is high and with that the power generation penalty of extracting steam from the turbine is small. The steam is therefore preferably extracted at lowest pressures possible within the conditions for the simulation. Another observation is how the curve is increasing more rapidly the lower the extraction pressures become. A small change in extraction pressure has a greater impact on the lost power generation for low pressures compared to higher pressures.

It is important to distinguish between α-value and net plant efficiency. The α-value is only taken into account the lost turbine power output between a plant with steam extraction and a plant without steam extraction, at the same design conditions. It does not say anything about how a different extraction pressure, and thereby different crossover pressure, affects the steam generation in the HRSG. Therefore, it is not possible to compare the total efficiency of a power plant between two different steam extraction pressure designs with only regard to the α-value. However, the α-value gives a significant guiding principle for steam extraction designs.

In accordance with the theory from chapter 2.2, an increase in the crossover pressure and the accompanying LPB pressure could result in a lower LP steam generation in the no extraction case. This is because of the increased evaporation temperature. If that is the case, less power might be produced as long as the effect of the reduced steam mass flow exceeds the effect of a higher enthalpy drop in the LPT due to increased crossover pressure.

5.1.2 Extraction at Turbine Casing

It is also possible to extract steam from the casing of one of the steam turbines. In that case, the steam is extracted through bleed ports between two turbine stages. By applying this method, the steam could be extracted at the preferred pressure. The same problem due to superheating of the steam occurs, as was the case for the extraction from crossover. Water could be injected to the stream in order to saturate the steam.

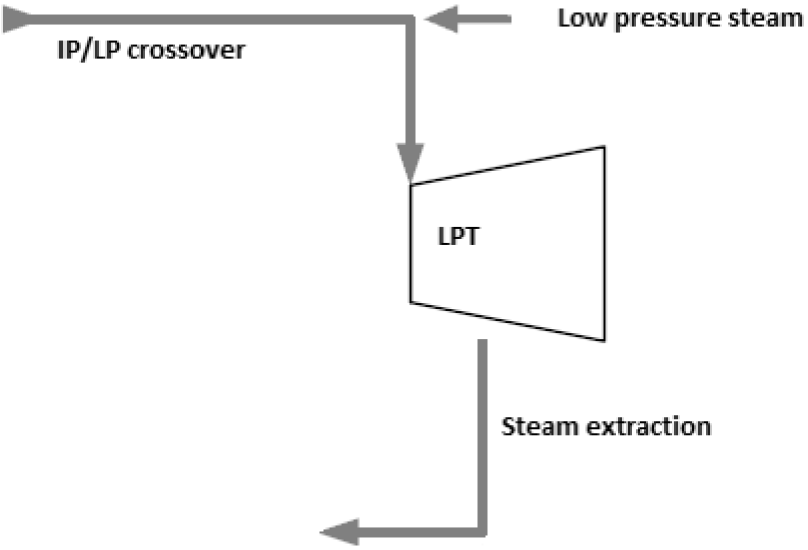

Figure 5.6 – Illustration of steam extraction from the casing of the LPT.

A problem related to extraction from the turbine casing is the large pressure reduction between two stages. The pressure between two stages is reduced with a factor of about 2[29]. If the steam is extracted from the casing of the LPT, a fairly high crossover pressure is required in order to satisfy the limitations for the amine regeneration temperature. By increasing the crossover pressure, the LPB pressure is increased and less steam is generated in the low-pressure cycle.

For a high amine regeneration temperature, with a corresponding high extraction pressure, it could be preferable to extract steam from the casing of the IPT. In that case, the crossover and LPB pressure could be lower than in the case of steam extraction from the crossover.

In terms of retrofitting of an existing power plant, there may be advantages related to steam extraction from casing compared to extraction from crossover. The extraction pressure could be determined based on the amine regeneration temperature, as long as the necessary pressure reduction between the turbine stages is feasible. If the existing power plant has a high design pressure at the crossover, and the required reboiler pressure is low, steam extracting from the casing of the LPT could give a greater α-value than by extraction from the IPT/LPT crossover.

5.1.3 Extraction from LPB/Crossover

Another opportunity is to extract steam directly from the LPB in addition to the steam from the crossover between the IPT and the LPT. The steam extracted from the LPB is saturated, and exergy losses due to superheating and enthalpy rise of the steam are avoided.

If the capture plant is not in operation, the steam generated in the LPB has to be utilized in the steam turbine, in order to maintain high plant efficiency. For that reason, the superheater between the LPB and the crossover cannot be excluded. Also, the pressure in the LPB has to be designed in order to match the crossover pressure.

Because of the crossover pressure constraint, the same amount of steam is to be generated in the LPB as for the steam extraction case from the crossover. However, when the steam is transported directly to the extraction point without superheating, the heat from the flue gas could be used to generate more steam in the IP cycle. From Figure 5.7, the low-pressure super heater (LPS) is placed before the intermediate-pressure boiler (IPB) in the HRSG, and the temperature of the flue gas is high enough for steam generation. Accordingly, more steam could be generated in the case of extraction from the LPB directly together with IPT exit steam than from the crossover alone. This additional steam generation will in theory improve the net efficiency.

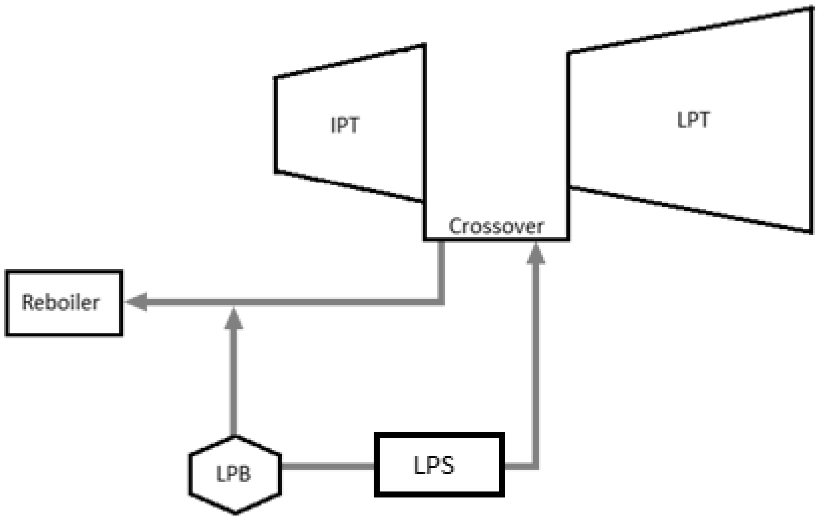

Figure 5.7 – Illustration of steam extraction from the LPB and the crossover between the IP and LP turbine.

The steam flow from the LPB is not enough to cover the entire heat duty in a typical amine process. Additional steam from the crossover is mixed with the steam from the LPB, and the total stream is superheated. Also in this case, water is necessary in order to saturate the steam.

5.1.4 Efficiency Penalty

The efficiency of the power plant including the capture process can be calculated using the following equation 5.8[3]. This equation includes all the losses that consist with capturing CO_2, all the way from fuel gas to transport ready CO_2. The result from this equation is the net plant efficiency.

$$\eta_{Post-Comb} = \eta_{no\ capture} - \frac{E^{CO_2}_{rem,mech} C}{LHV} - \frac{E^{CO_2}_{rem,heat} \alpha_{str,atm} C f}{LHV} - \frac{E^{CO_2}_{comp} C f}{LHV} - \frac{\Delta E_{aux}}{LHV} \qquad 5.8$$

1) Efficiency of a standard plant with no capture, calculated from equation 2.10.
2) Efficiency penalty by use of mechanical work or electricity in the CO_2 absorption process. The work includes fans and pumps. The C gives the ratio between formed CO_2 and consumed fuel.
3) It is the largest penalty and it consists of the loss by extracting steam from the steam cycle. The f is the percentage of CO_2 captured in the absorption process. The α express the loss of steam turbine power output compared to the potential heat in the steam, see chapter 2.6.
4) This penalty is the energy required to compress and condense the CO_2 after it has been captured. Make the CO_2 transport and storage ready.
5) The α value is only taking into account the loss of the steam turbines power output and neglects the change in auxiliaries and losses in the cycle. This term represents the difference

in auxiliary work in case of steam extraction. The pump work related to the condenser cooling water will be reduced when steam is extracted from the steam turbine.

5.2 Exhaust Gas Recycling

This is a quite simple way of both increasing the partial pressure of CO_2 and as well decreasing mass flow of flue gas into the absorption column. With this technology parts of the exhaust gas will be recycled back to either the air compressor or a separate compressor before it enters the combustion chamber. Figure 5.8 gives a simplified schematic of a gas turbine cycle with EGR. The ratio of exhaust gas being recycled can vary form 10-55% and is defined as:

$$EGR = \frac{\text{volume flow of recirculated exhaust gas}}{\text{volume flow of exhaust gas}} \tag{5.9}$$

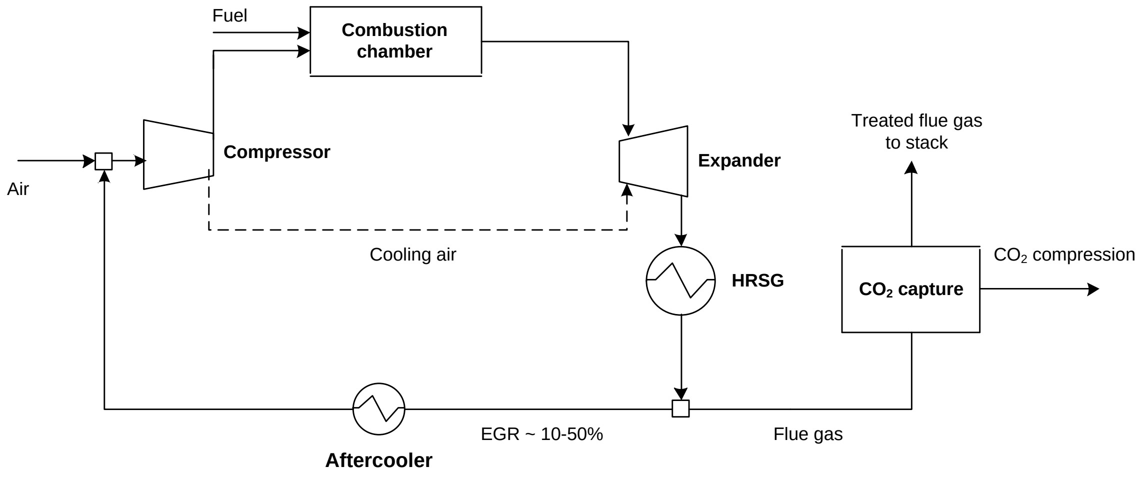

Figure 5.8 - Simplified flow sheet of a NGCC with EGR.

Figure 5.9 represents the variation of the EGR ratio compared against the mass flow of flue gas going into the absorber and CO_2 mole composition of the flue gas after the HRSG. Her it can be seen that by increasing the EGR ratio from 0 to 55 % the mole composition of CO_2 will increase from 4,22 mol-% to 9,7 mol-%. At the same time the mass flow of flue gas into the absorber column decreases from around 660 to 292 kg/s. This is a percentage decrease of around 56 %, which may result in smaller absorber and stripper column as well as lower heating duty in the reboiler.

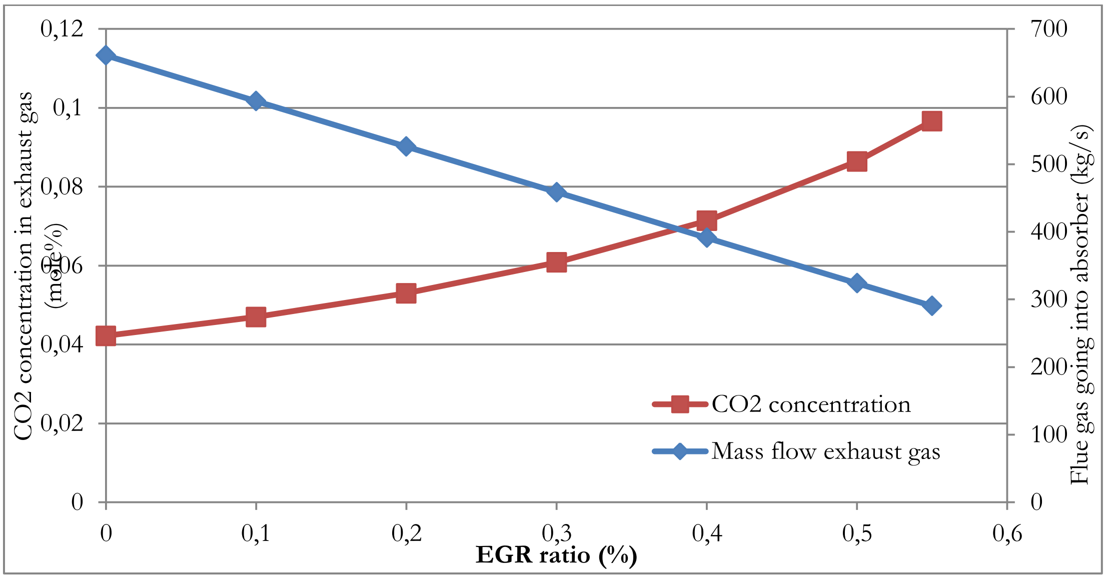

Figure 5.9 - CO_2 concentration and mass flow of exhaust gas at the absorber inlet as function of EGR ratio.

The drawback by use of the technology is the change in the composition of the flow going into the gas turbine. By mixing in flue gas into the burner there can be problems getting a complete combustion because of lack of oxygen. This can cause a reduction in the efficiency as well as create CO instead the desired CO_2. Figure 5.10 gives the concentration of O_2 before and after the combustion chamber, here the ratio of EGR should not exceed 55 % because then the concentration of O_2 will be below what is required for a stoichiometric combustion. Normal burners are designed to operate with around 21 mol-% O_2, when using EGR with a ratio of 55 % the concentration of O_2 can be as low as 11 mol-%. This value is theoretically possible but according to Ditaranto et al[30] the flame requires a minimum of 14 mol-% to sustain the combustion. The same study indicate that when burning with O_2 mol-% below 16 there will be extensive amounts of unburned hydrocarbons and CO in the combustion outlet.

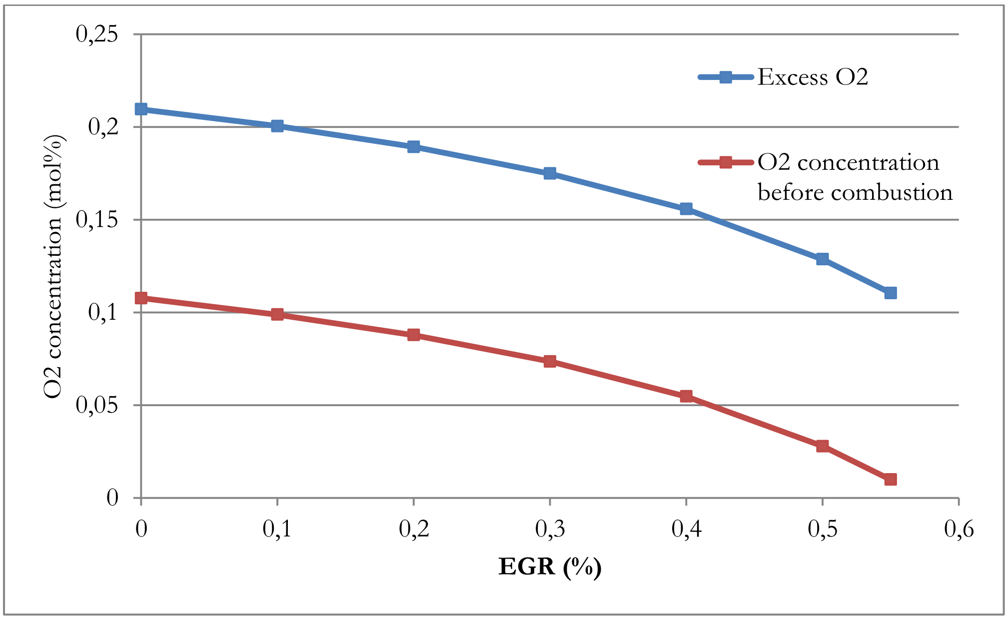

Figure 5.10 – O_2 concentrations in the recirculated exhaust gas at different EGR ratios.

When EGR is used on a NGCC, the gas turbine is taken outside its designed operational area and it becomes an off-design case. When the gas turbine operates in off-design, the mass flow into the compressor and the compression ratio will change. The change in compression ratio can be evaluated by the use of chocked nozzle equation:

$$\frac{P_{TIT}}{P_{r,TIT}} = \frac{\dot{m}_{TIT}}{\dot{m}_{r,TIT}} * \sqrt{\left(\frac{T_{TIT} * MW_{r,TIT}}{T_{r,TIT} * MW_{TIT}}\right)} \tag{5.10}$$

Where:
TIT = Turbine inlet temperature
MW = Molecular weight
P = Pressure
r = Design case value
T = Temperature

From this equation 5.10 it can be seen that the pressure ratio over the compressor will decrease when the mass flow in the off-design case are reduced.

By implementing this equation into the gas turbine calculations the compressor outlet pressure will vary when the EGR ratio changes.

The compressor flow rate will also be changed when the gas turbine is operating in off-design, this change can be seen in equation 5.11.

$$\frac{\dot{m}_{comp\ inlet}}{\dot{m}_{r,comp\ inlet}} = \frac{T_{r,comp\ inlet}}{T_{Comp\ inlet}} \qquad 5.11$$

Where:

$T_{\text{comp inlet}}$ = Temperature at compressor inlet

$\dot{m}_{comp\ inlet}$ = Mass flow at compressor inlet

Equation 5.11 shows that when the temperature into the compressor increases the mass flow will decrease compared to the design case. When using EGR, the recycled flue gas have a temperature of around 80 °C, this will decrease the mass flow. To prevent some of this decrease a flue gas cooler can be installed prior to the compressor.

5.3 Conclusion

- For steam extraction from the IPT/LPT crossover, with water injection from the deaerator and constant reboiler duty, the α-value is reduced when the extraction pressure increases. Lower α-value is related to an increased power loss in the steam turbine, and it is preferable to extract steam at the lowest crossover pressure possible.
- Even though the extracted mass flow from the IPT/LPT crossover decreases with increased extraction pressure, the α-value goes down for a given reboiler duty. The effect of a higher enthalpy drop in the expression for lost expansion work exceeds the effect of a lower mass flow extracted from the crossover.
- The α-value is an indication of how much power is lost in the steam cycle because of the steam extracted to the capture process.
- The reboiler duty has great impact on the net plant efficiency.

5.3.1 Pressure of Extracted Steam

The steam needed in the reboiler in capture plant will be extracted from the crossover between the IPT/LPT. From the simulations and theory in the chapter 5.1.1.2, the steam is preferable extracted at its lowest possible pressure due to the temperature limit of the solvent. The regeneration temperature of the absorbent is in the area of 110-120 °C. A minimum temperature difference of 10 K is required in the reboiler. It is assumed saturated steam/vapor at the inlet and saturated liquid at the outlet of the reboiler. According to the equation 5.1, the minimum extraction pressure is:

$$P_{ext} = P_{sat}(130°C) + 0{,}5 = 2{,}7 + 0{,}5 = 3{,}2bar \qquad 5.1$$

2,7 bar is an absolute minimum pressure at the reboiler exit. Included a safety margin, 3,2 bar is chosen as the reboiler exit pressure. The extraction pressure is thus 3,6 bar. This extraction pressure gives a temperature of 139,9 °C for the steam in the reboiler.

5.3.2 EGR

Because of the minimum required O_2 mol-% concentration in the combustion the maximum EGR used hence forth in this thesis have been set to 45 %, this results in O_2 concentration of below 16 mol-% into the combustion chamber. In a state of the art gas turbine this maybe too low, but higher EGR ratio have been evaluated to find the benefit of having burners that can operate with lower O_2 concentration.

6 Simulation Processes

The most promising processes have been chosen based on the literature study; they have been further evaluated and will be presented in this chapter. The processes were chosen because they use a combination of both EGR and higher flue gas pressure. These processes have gone through thorough investigation and computational simulations have been performed, the results are presented in chapter 8. Each of these processes is based on the GT in the NGCC and they use identical steam cycles.

6.1 Post-expansion CO_2 Capture

The post-expansion is similar to a normal NGCC the major difference is the utilization of two turbines; one high pressure and one low pressure. After the HPT the flue gas is cooled down and the CO_2 is removed from the gas in the capture plant. The remaining pressure of the flue gas can be utilized in the LPT. This process also uses EGR, but here the gas that is being recycled is only the clean gas so the partial pressure into the capture plant will not be affected as much as in the NGCC with EGR.

The heat exchanger after the HPT is a critical factor for this process. The temperature into the heat exchanger can vary form 700-1100 °C depending on pressure for the HPT. It requires a robust heat exchanger to withstand these kinds of temperatures.

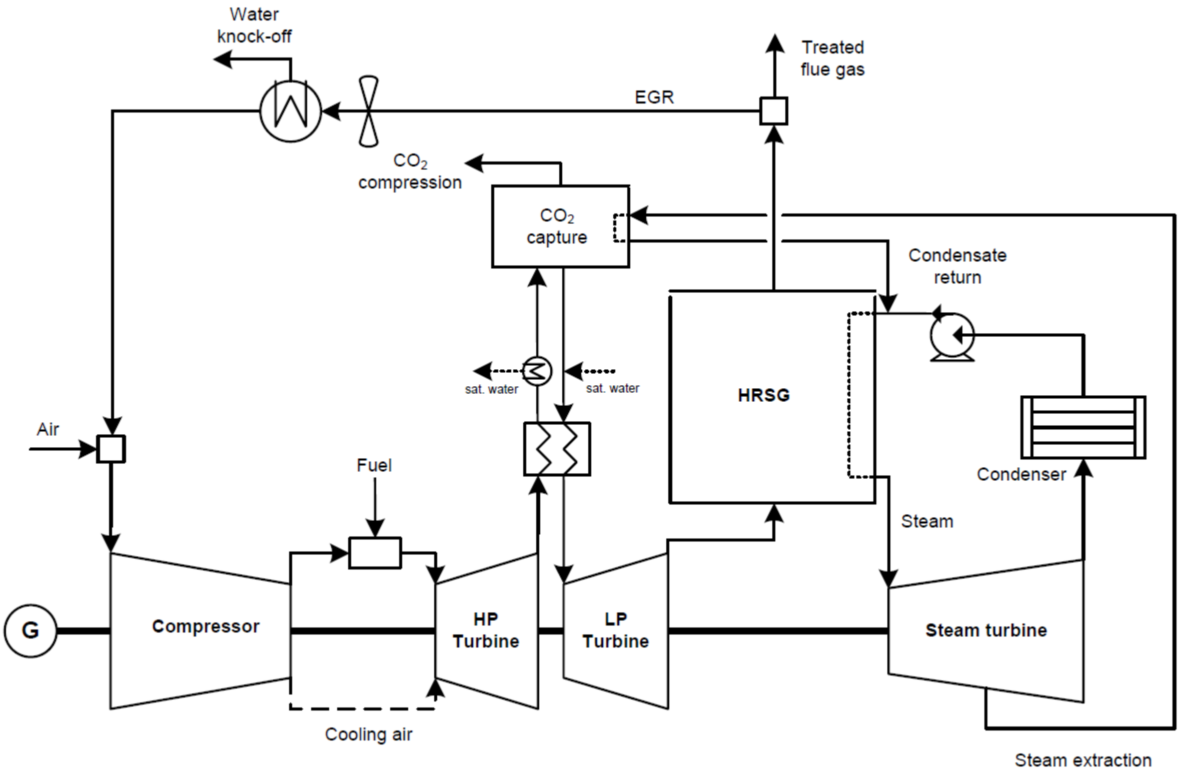

Figure 6.1 – Post-expansion CO_2 capture.

6.2 Tail-end CO_2 Capture

The tail-end capture process is very similar to a normal NGCC plant with CO_2 capture; the one of the main differences is the EGR which recycles a certain amount of CO_2 back to the air compressor. The other is the compressor which compresses the remaining flue gas before it enters the capture plant. This is done in order to increase the partial pressure of the CO_2 in the flue gas as well as the volumetric flow rate of the gas. By increasing the pressure the CO_2 can be more easily captured in the absorber. After the capture plant the remaining the flue gas is taken through an expander before being release into the stack.

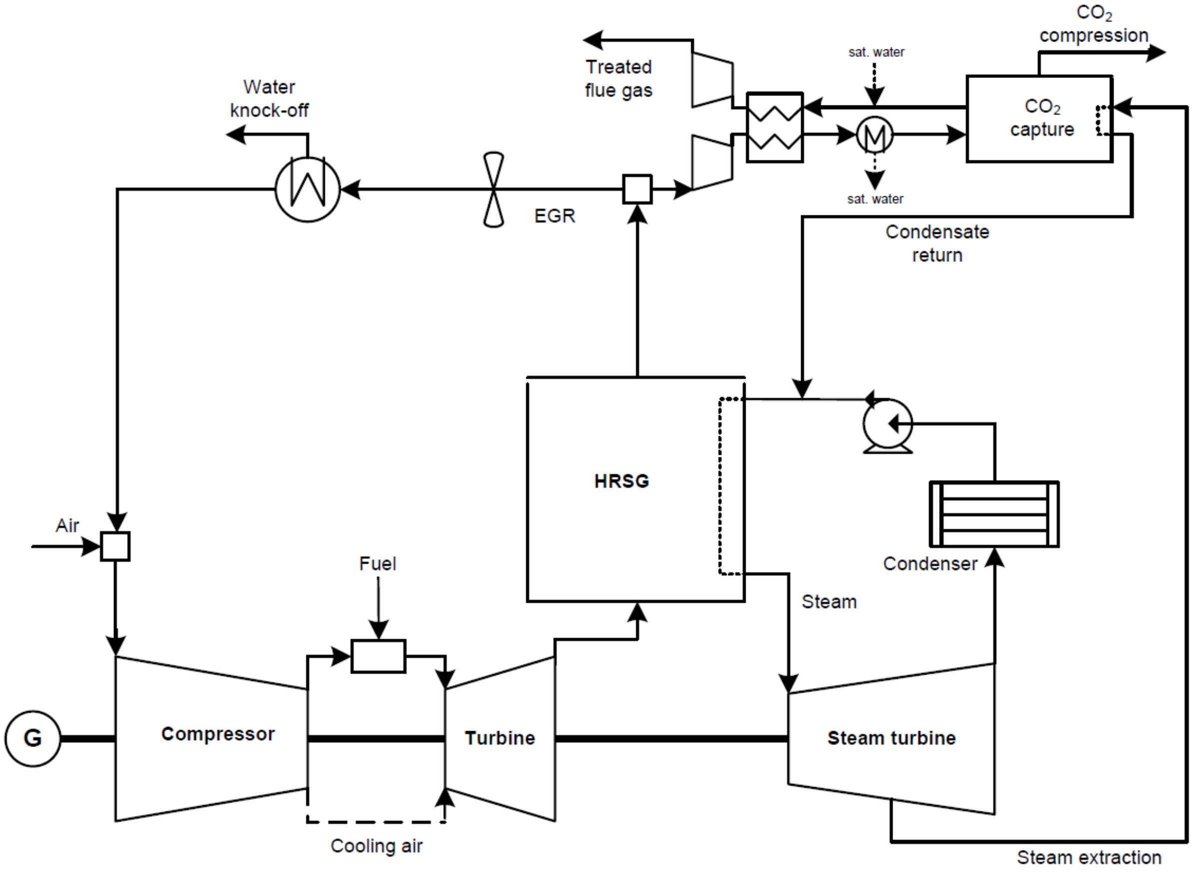

Figure 6.2 - Tail-end CO_2 capture.

6.3 Post-compression CO_2 Capture

The post-compression are a variant of tail-end capture, her all the flue gas is compressed after the HRSG. Then a percentage of the flue gas are recycled directly back into the combustion chamber. The rest of the flue gas is now at an elevated pressure and is taken through the CO_2 capture plant to be cleaned. Before entering the plant the gas is cooled down using the clean gas leaving the absorber in capture plant. After the capture plant the flue gas is expanded in an expander.

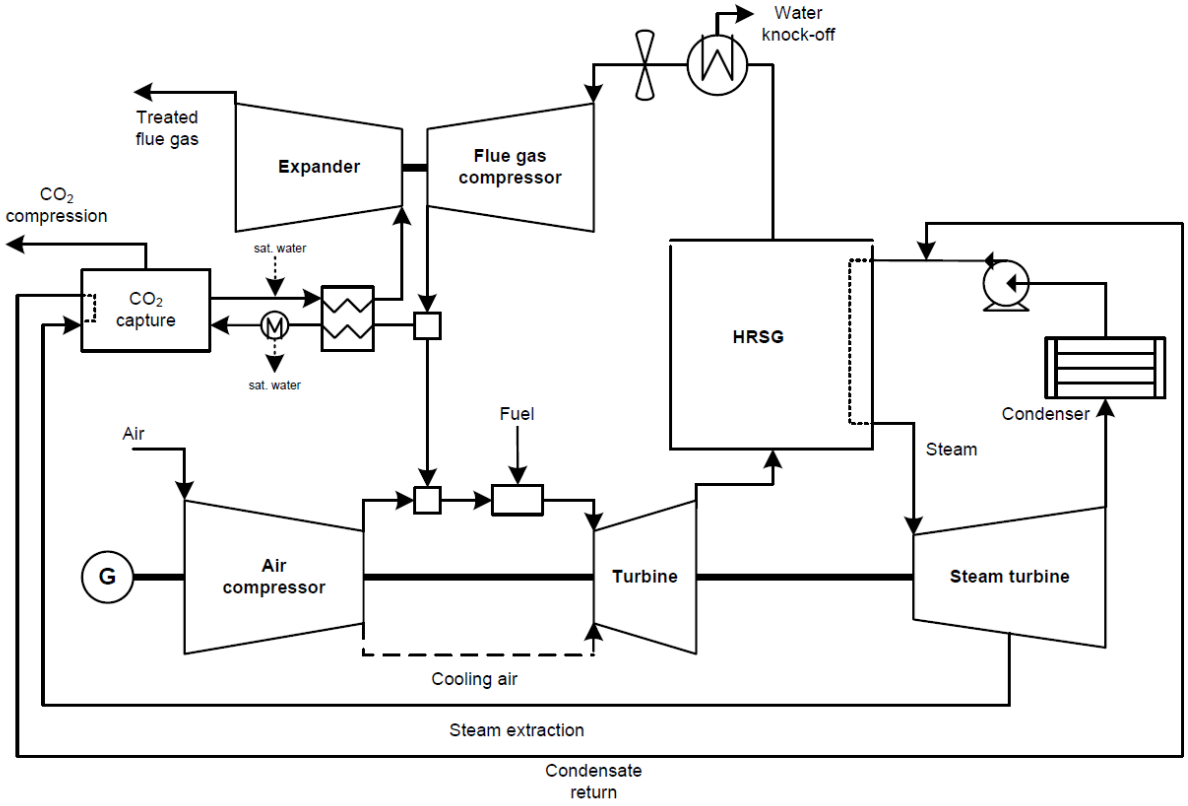

Figure 6.3 - Post-compression CO_2 capture.

7 Simulations and Methodology

In this chapter a short description will be given on the different modeling cases and programs that have been used in this thesis. All the assumptions that are made will be presented and explained for each of the processes models.

7.1 Simulation Software

For designing and simulation of the CO_2 capture process and the novel power cycles the software Aspen HYSYS 7.3 from AspenTech[31] has been used. This is a straight forward program with no programming needed. The simulation environment in HYSYS is graphical. Where compressors, turbines and columns are drag-and-drop as process blocks and connected by stream lines. The parameters are set inside each process block. Color codes are used to systematically provide information on whether or not components are sufficiently defined or within the boundaries of this thermodynamic fluid package. Because each of the processes gives a lot of output data, HYSYS have been linked to Microsoft Excel with Apsen Woorkbook 7.3 to more easily extract and handle the large amount of data. For the capture simulation the fluid package Amine Pkg is used, this fluid package is not as stable as Peng-Robinson and SRK which is used for the NGCC. Therefore some of the results for the capture plant can vary much from case to case. The cooling circuits uses ASME steam as fluid package.

In order to design the steam cycles, the simulation program GT PRO is used[32]. GT PRO is a simulation software from Thermoflow. The program computes all necessary heat balances and equipment designs for the given plant. The user types the input criteria and other assumptions, while GT PRO computes the heat and mass balances, system performance and component sizing[33]. In addition, GT PRO is automatically suggesting several input values. The user has the choice of utilizing these values or to make adjustments. It is possible to either optimize the economy or the net efficiency of the plant. In this thesis, GT PRO is used in order to optimize the efficiency. GT PRO have been linked up to Microsoft Excel by the use of an add-in called Thermoflow ELINK.

7.2 Process Design and Specifications, NGCC

A reference NGCC plant is designed without CO_2 capture in order to create a starting point for the CO_2 capture study. The output values of the flue gas conditions are used as input values in the capture process. In addition, the plant performance is analyzed and creates a standard that can be used for comparison of the loss by capturing CO_2. The basic designs of the NGCC plant is taken from the European guidelines for assessment of CO_2 capture technologies, published by DECARBit[34]. One of the challenges in compering different simulation results between published reports is the different input values of the plants. With the guidelines from DECARBit, it could be easier to compare the simulations with other results.

A summary of the most important input values used in the reference plant without CO_2 capture is given in Table 7.1

Parameter	Unit	Value
Fuel composition		
Methane	89	Mol-%
Ethane	7	Mol-%
Propane	1	Mol-%
I-Butane	0,05	Mol-%
N-Butane	0,05	Mol-%
N-Pentane	0,004	Mol-%
Pentene	0,005	Mol-%
Hexane	0,001	Mol-%
Carbon dioxide	2	Mol-%
Nitrogen	0,89	Mol-%
Plant design		
Polytropic turbine efficiency	85[3]	%
Mass flow air	642,1	kg/s
Polytropic compressor efficiency	90[3]	%
Turbine cooling air (percentage of compressor inlet)	13,17	%
Generator and mechanical efficiency	98,6	%
Combustion outlet temperature	1427	°C

Table 7.1 – Input values for NGCC reference plant.

7.3 Process Design and Specifications, CO_2 Capture Plant

7.3.1 MEA Capture Plant

The MEA process is defined with the parameters given in appendix 12.3, Table 12.1. The capture rate in the absorber is set to 90 %, and this value is achieved by using an adjuster in HYSYS. The adjuster regulates the mass flow rate of MEA into the absorber so that 90 % capture is attained. In the desorber two parameters are selected for the iteration process, those parameters are; the condensation temperature out of the top condenser and the amount of CO_2 leaving the reboiler with the lean solution. The condensation temperature is set to 30 °C and the maximum mol-% CO_2 leaving the reboiler is 2,5. The results from HYSYS may vary; this is because the reboiler duty is very sensitive to the mass flow, fraction of MEA and amount of CO_2 leaving the stripper. HYSYS varies both the mass flow and fraction of MEA by use of the adjuster. In this study the start values for solvent mass flow and MEA wt. % have been set to 650 kg/s and 29,85. This is done so that all the different simulations have the same starting point, and can therefore more easily be compared.

For both the MEA and MDEA processes the height of the absorber and the stripper are not taken into consideration when completing the simulations in HYSYS. The diameter for both absorber and stripper has been set constant at 10 m and 8 m in both of the processes. This is done in order to compare the processes more easily. In chapter 8.6 the theoretical height and diameter of the absorber have been calculated by use of the flue gas data from the simulated cases in HYSYS

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[3] The polytropic efficiency is taken from the course material in TEP4185: Industrial Process and Energy Technology.

7.3.2 MDEA Capture Plant

The MDEA capture process will only be used on the high pressure flue gas processes, tail-end, post-compression, and post-expansion. That is because MDEA are mostly suited for higher flue gases with higher pressure. The processes are defined in appendix 12.3, Table 12.2. The MDEA have the same setup as the MEA simulation. The big change is the wt. % of amine in the solvent and the gas/liquid separator prior to the stripper. By use of the valve in front of the separator, the solvent will be given a pressure swing and some of the CO_2 will be released from the solvent and vented out through the top of the separator. The gas entering the absorber needs to be saturated with water in order for the absorption process to operate properly. The starting parameters for the MDEA process have been set at 650 kg/s and a wt. % of 49,9.

7.4 Process Design and Specifications, Steam Cycle

When simulating the NGCC with EGR, tail-end capture, post-compression and post-expansion, HYSYS are used as the simulation tool all of the process except the steam cycle. The flue gas from the gas turbine is used in GT PRO to simulate the steam production in the HRSG and the steam turbine power production.

A three pressure level steam cycle with reheat of the HPT exit flow is used. As cooling system in the condenser, a natural draft wet cooling tower is used, with the ambient conditions described in the DECARBit report[34]. The temperature increase of the cooling water is assumed to be 11 K through the condenser. A temperature difference of 3 K is used between the cooling water and the condensed steam/water. It is assumed no bleeds and leakages in the steam turbine. The most important input data for the steam cycle is presented in Table 7.2.

Parameter	**Value**	**Unit**
HP pressure and temperature	125/566	bar/°C
IP pressure and temperature	30/566	bar/°C
Crossover IPT/LPT pressure	3,92	bar
LPB pressure	4,36	bar
LPS exit temperature	290	°C
Dearator pressure and temperature	1/99,63	bar/°C
Condenser pressure and temperature	0,048/32,17	bar/°C
Pinch LP/IP/HP	5/10/10	°C
Isentropic efficiency HPT/IPT/LPT	92/94/90	%

Table 7.2 – Input data for steam cycle in GT PRO.

7.5 NGCC with EGR

GT PRO does not have the possibility of including EGR; therefore HYSYS has been used for simulating the NGCC with EGR. The HYSYS model has been based on the base case gas turbine from GT PRO. The main change is the EGR, to calculate the change for the gas turbine in off-design the equation explained in chapter 5.2 will be implemented into the model. In Table 7.3 the assumptions and input values used in this model are presented.

Air compressor		
Polytropic efficiency	90	%
Turbine cooling air (percentage of compressor inlet)	13,17	%
Mass flow	642,1	kg/s
Inlet temperature	15	°C
Turbine		
Polytropic efficiency	85	%
Mass flow	642,1	kg/s
Combustion outlet temperature	1427	°C
Molecular weight	28,49	

Table 7.3 - Assumptions and input values for NGCC with EGR.

It has been assumed that there is no pressure drop through the combustion chamber, which means that the pressure at the compressor outlet is the same as the turbine inlet.

7.6 Tail-end-, Post-compression- and Post-expansion Capture

These processes will be so-called Greenfield power plants. Because of this the gas turbine used will be designed to operate with an EGR ratio that will increase the overall efficiency of both NGCC and capture plant, thus eliminating the need for off-design calculations with regards to EGR.

The different cases simulated for these processes are EGR ratio from 0-45 % and the pressure of the flue gas into the absorber 2,4,6,8 and 10 bar. The input data for the case are the same as in Table 7.3.

Each of these requires a heat exchanger that is connected to the streams going into and leaving the capture plant. The pressure drop through that heat exchanger is assumed to be 3 % for these cases. The temperature drop over the heat exchanger in the post-compression is set to 20 °C and in the tail-end the drop is 10 °C. The temperature difference over the heat exchanger in the post-expansion process is set to 60 °C, there have to be a larger drop over this heat exchanger because of the high temperature of the inlet gas. To have a pressure drop of 20 °C with this kind of temperature requires enormous size on the heat exchanger.

8 Results and Discussion

In this chapter the results from the different power generation and CO_2 capture processes will be presented and discussed. They will be compared with each other with regards on the total efficiency, reboiler duty, component size, and the maturity of the different technologies. The different cases to be evaluated are NGCC with uses of EGR, post-expansion, post-compression, and tail-end capture. All of the cases will be evaluated with use of both MEA and/or MDEA as amine in the capture plant. The variables that will vary from each of the cases will be the pressure of the flue gas and the amount of exhaust gas that is being recycled back to the gas turbine.

8.1 NGCC Base Case

8.1.1 Without Capture

A NGCC without CO_2 capture is used as comparison for the other cases. The plant is presented in chapter 7.2 and a summary of the most important results are presented in Table 8.1.

NGCC	Value	Unit
Mass flow fuel	15,67	kg/s
LHV fuel	46,5	MJ/kg
CO_2 emitted	41,8	kg/s
Gross gas turbine output	289,03	MW
Gross steam turbine output	148,37	MW
Auxiliaries & losses	4,87	MW
Plant net output	432,53	MW
Net plant efficiency	58,80	%

Table 8.1 - NGCC base case without CO_2 capture.

8.1.2 With Capture

An important factor when capturing CO_2 is the reboiler duty and to reduce it when optimizing the capture process. To reduce the heat needed in the reboiler one can look at the amount of released CO_2. As mentioned in chapter 7.3 the fraction of CO_2 leaving the desorber is set to 2,5 mol-%. This number can be optimized to find the value that reduces the reboiler duty. In Figure 8.1 the reboiler duty and the fraction of CO_2 in the lean solvent flow is plotted. With a higher mass flow the reboiler needs more energy to boil the rich solvent, as can be seen in Figure 8.2. The reboiler duty increases a lot when the flow rate becomes high. When looking at CO_2 fraction in Figure 8.1 below 2,6 % the reboiler duty starts to increase. This is because the amount of CO_2 leaving with the solvent starts to become so small that it requires more energy to liberate the CO_2 from the solvent than is saved by having a lower mass flow. The optimal amount of CO_2 leaving the desorber is the low point of the curve 2,6 %. This gives a reboiler duty of 3,64 MJ/kg CO_2. For this case the mole fraction of 2,6 gives a reboiler duty of 3,6 MJ/kg CO_2. Whit that duty the steam demand becomes 66 kg/s.

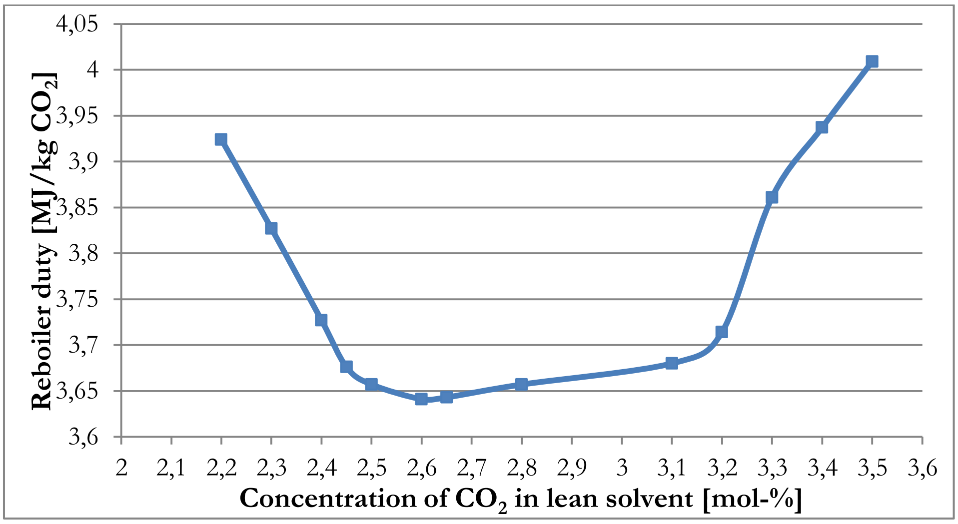

Figure 8.1 - Variation of reboiler duty as a function of concentration of CO_2 in lean solvent flow.

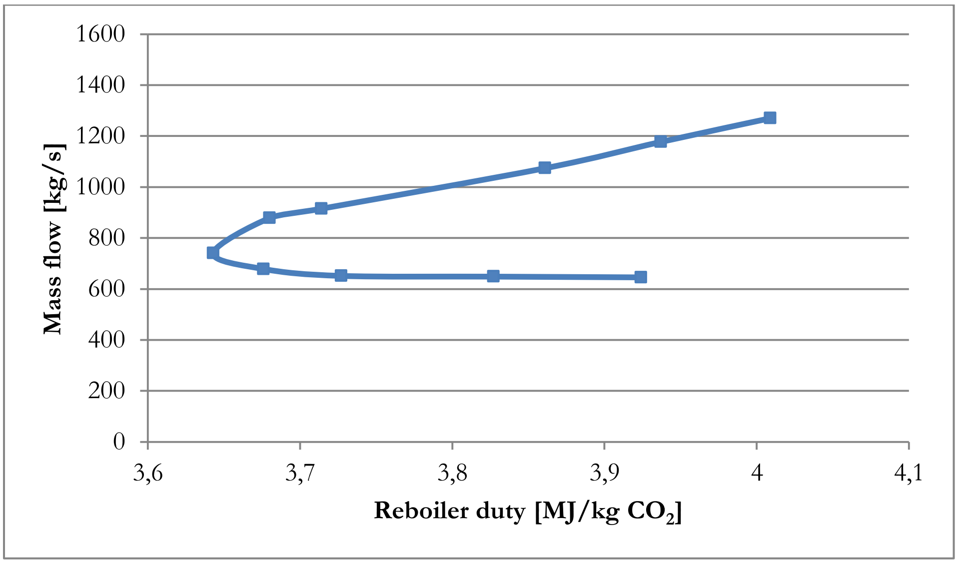

Figure 8.2 - Mass flow of regenerated MEA solution plotted vs. reboiler duty.

In Table 8.2 the results from the NGCC base case with CO_2 capture is presented. The capture plant is described in chapter 7.3.1. The GT in the NGCC are unchanged, the ST output have been lower because of the 66 kg/s steam extracted to the reboiler. From the results it can be seen that by capturing the CO_2 the net plant efficiency have taken a penalty of 7,4 %-points.

NGCC	**Vaule**	**Unit**
LHV input	728,73	MW
Auxiliaries & losses	3,71	MW
Gross gas turbine output	289,03	MW
Gross steam turbine output	112,69	MW
Plant efficiency	54,06	%
MEA capture		
Reboiler duty	3,64	MJ/kg CO_2
Steam demand in reboiler	66,1	kg/s
Mechanical work	7,84	MW
CO_2 compression	11,43	MW
CO_2 emitted	4,14	kg/s
Net plant efficiency	51,42	%
Efficiency penalty of CO_2 capture	**7,38**	**%-points**

Table 8.2 - Results from NGCC base case.

8.2 NGCC with EGR

The benefit this process has on the mass flow rate into the absorber and the CO_2 partial pressure are presented in chapter 5.2. The impact on the capture process is evaluated here. Since the exhaust gas is at atmospheric pressure only the MEA case has been used in these evaluations.

8.2.1 Power Plant

When EGR is used there will be a slight fall in GT efficiency, as observed in Figure 8.3, and this is closely related to the temperature increase at the turbine outlet, see Figure 8.6. Both these changes are because of the change in composition in the gas entering the turbine. By use of EGR the CO_2 concentration increases and this changes the total specific heat in the gas. CO_2 have a higher specific heat capacity and molecular weight than O_2 and N_2. As mentioned in chapter 5.2 the temperature at the compressor inlet will rise when the hot flue gas is recycled and this will again increase the power demand in the compressor. This is true, but from equation 5.10 the pressure ratio on the compressor will change for a GT in off-design, and from Figure 8.5 the pressure will decrease with higher EGR ratio. Along with the decrease in mass flow into the compressor, these factors will play a more important roll on the total compressor power demand than the temperature, resulting in a slight fall in the compressor power output, Figure 8.4. This will play a positive part of the change in GT efficiency.

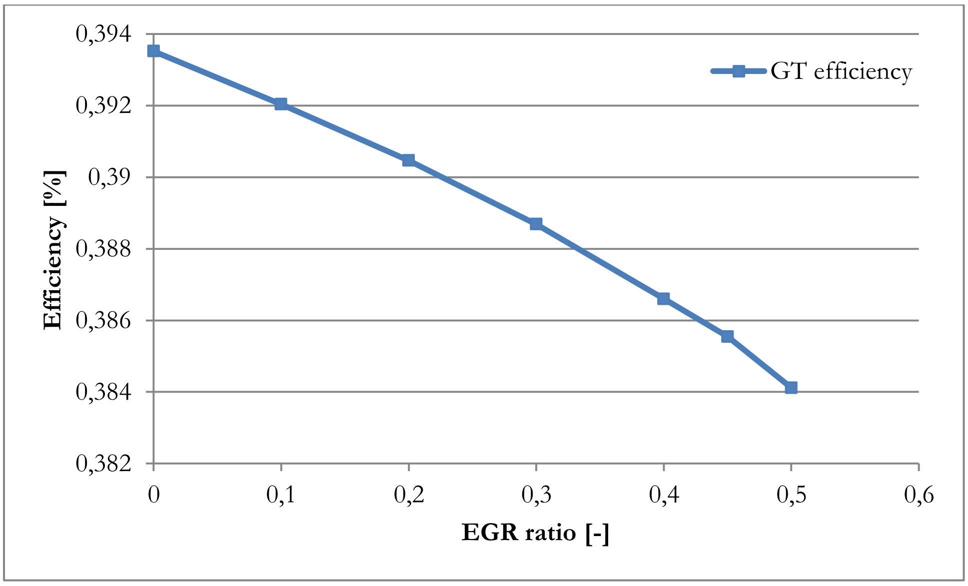

Figure 8.3 - GT efficiency with changing EGR ratio.

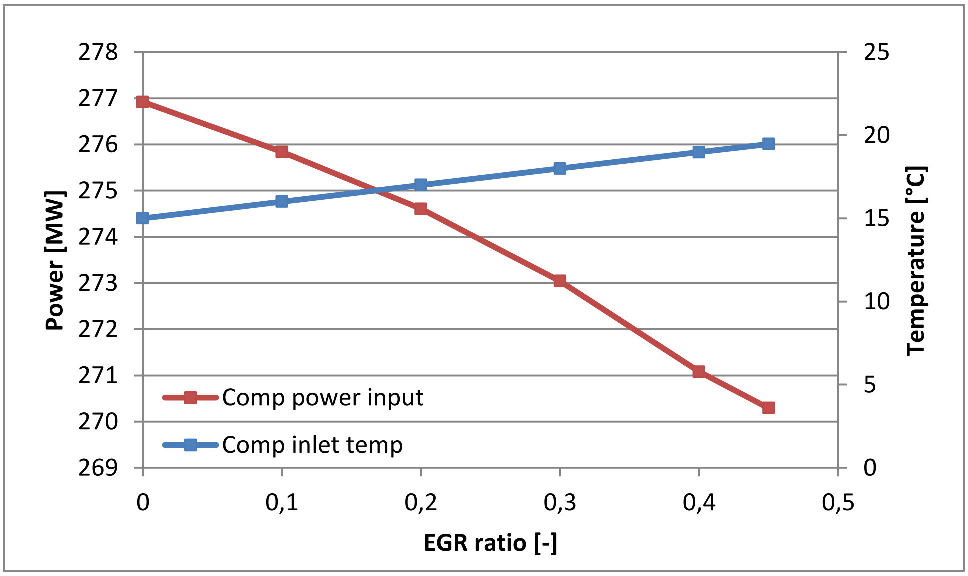

Figure 8.4 - Power demand in the compressor and temperature at the inlet vs. change in EGR ratio.

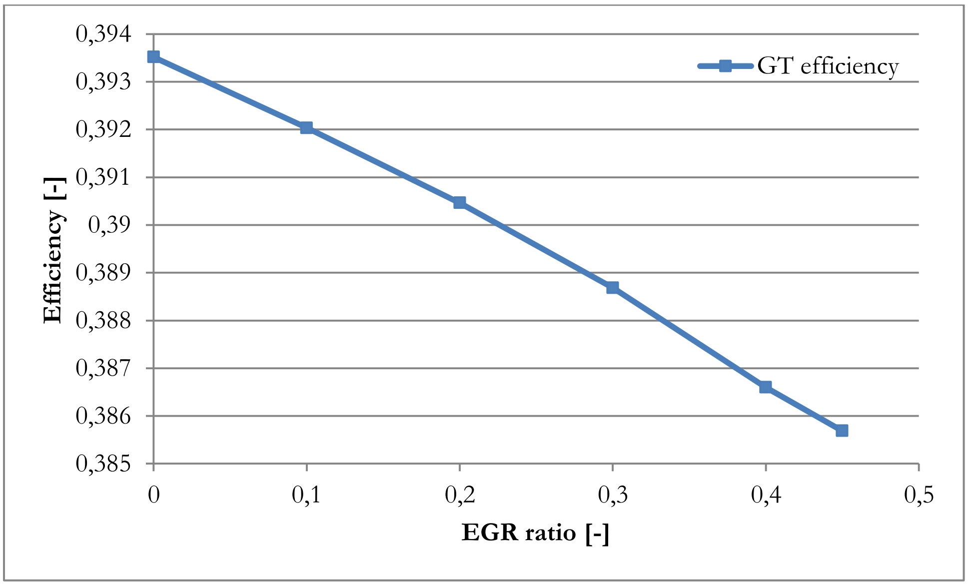

Figure 8.5 - Compressor pressure vs. EGR ratio.

Higher temperature at the turbine outlet play as negative factor on the GT but it plays an important positive factor on the ST. From Figure 8.6 the increase in ST power output can be observed along with the higher temperature at GT outlet.

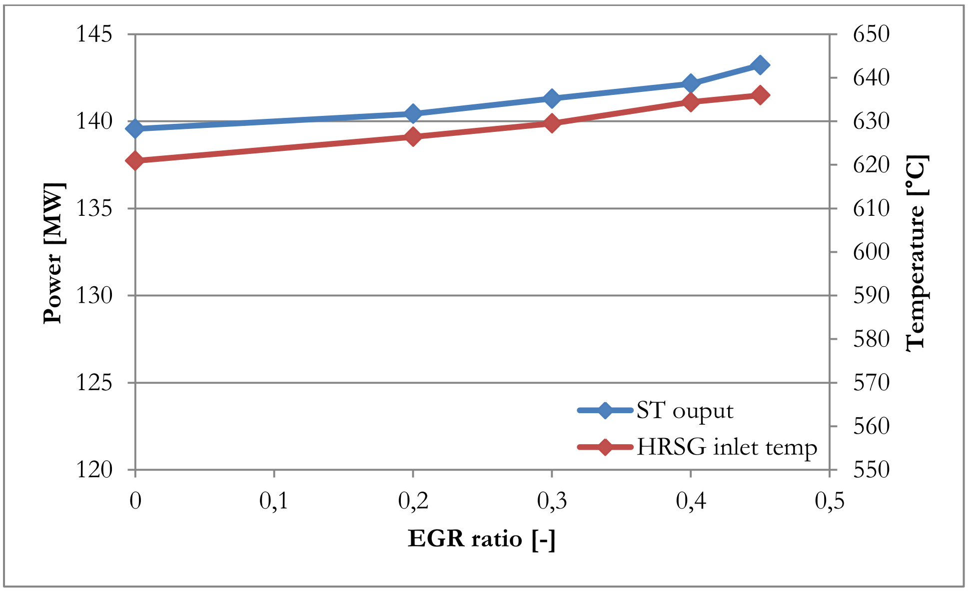

Figure 8.6 – ST power output and HRSG inlet temperature for changing EGR ratio.

8.2.2 Capture Plant

In the base case the best fraction of mole CO_2 leaving the absorber was 2,6 %. The same study has been made with different EGR ratios. From the results in Figure 8.7 a trend can be observed that indicates that the best mole fraction CO_2 is around 2,5 %. That will give the lowest reboiler duty for each of the EGR ratios. For simplifications that mole fraction will be used for the rest of the processes in this thesis. The lowest reboiler duty are for a EGR ratio of 45 %, this fits well with the theory because for higher partial pressure the energy demand of the capture process should be lower. The reboiler duty is here 3,4 MJ/kg CO_2.

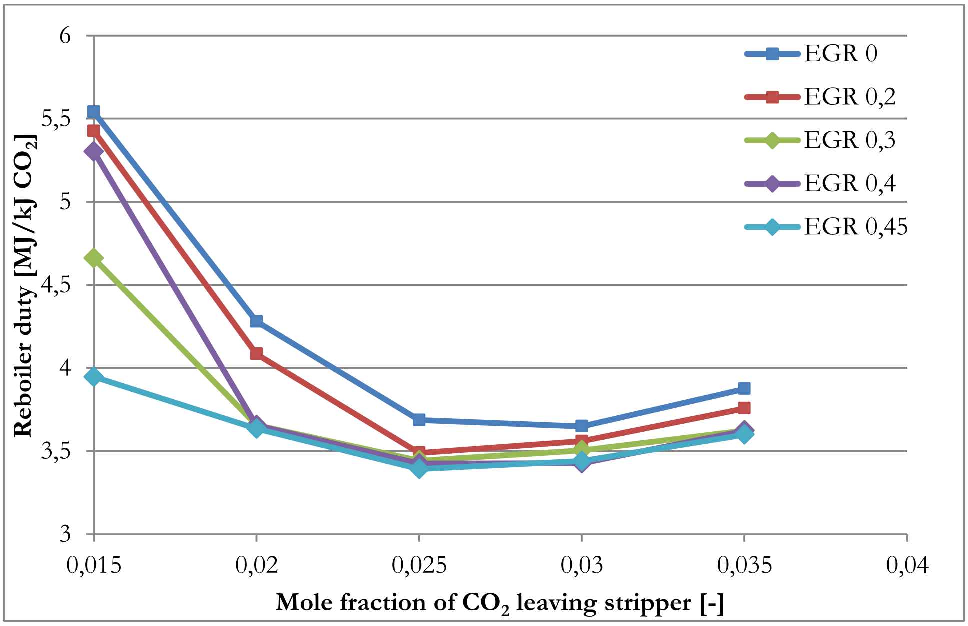

Figure 8.7 - Reboiler duty plotted against the mole fraction of CO_2 leaving the stripper for flue gas with different EGR ratio.

8.2.3 Conclusion

From Figure 8.8 it can be observed that the best net plant efficiency is with an EGR ratio of 45 %, the efficiency is the 52 %. All the losses have been included, from CO_2 capture and CO_2 compression. At EGR ratio of 0-20 % the change is quite pronounced compared with 20-45 %, this tell that the use of EGR are more dominating at lower ratios. All the data from the best EGR ratio are presented in Table 8.3. If the EGR ratio could have been increased further the net plant efficiency would have been increased as well.

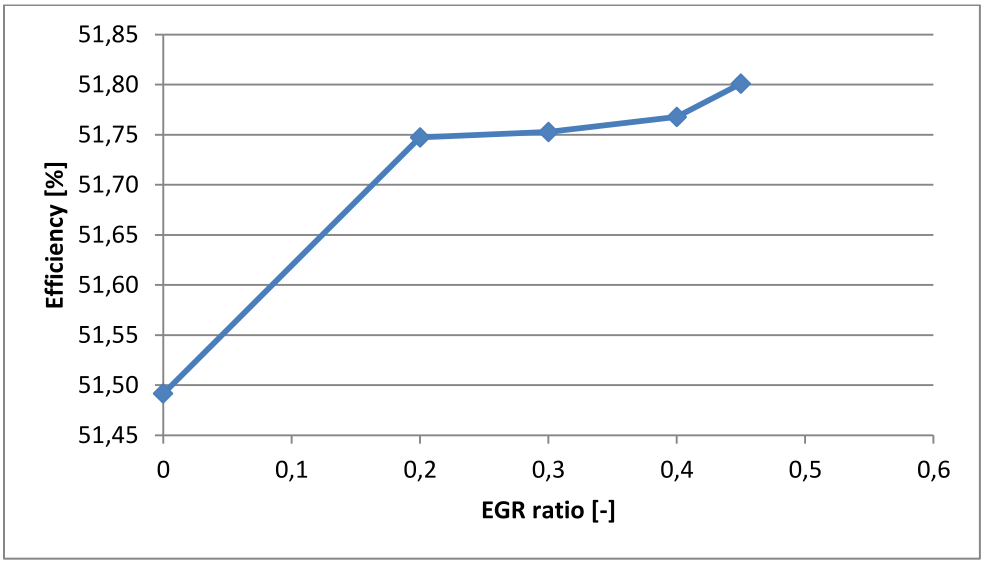

Figure 8.8 - Net plant efficiency for changing EGR ratio.

NGCC with EGR	Value	Unit
LHV input	733,08	MW
Auxiliaries & losses	6,46	MW
Gross gas turbine output	284,76	MW
Gross steam turbine output	119,37	MW
Plant efficiency	53,70	%
MEA capture		
Reboiler duty	3,39	MJ/kg CO_2
Steam demand in reboiler	58,99	kg/s
Mechanical work	0,97	MW
CO_2 compression	12,1	MW
CO_2 emitted	4,32	kg/s
Net plant efficiency	51,99	%
Efficiency penalty of CO_2 capture	**6,81**	**%-points**

Table 8.3 – Results from NGCC with EGR and CO2 capture.

- When EGR is used the GT efficiency is decreased, but the power output of the ST increases because of the higher turbine outlet temperature.
- The reboiler duty have been lower by 5,4 % compared to the base case.
- By using EGR with a ratio of 45 % the total efficiency of the plant with capture increases with 1,11 % compared to the base case.
- Simulation results show that the net plant efficiency is 52 % which is a drop in efficiency of 6,8 % compared to a NGCC without capture.

8.3 Tail-end Capture

The tail-end capture process is described in chapter 6.2. Here the results for the simulation of the process along with CO_2 capture with the use of both MEA and MDEA as amine. A complete schematic of the tail-end capture process from HYSYS can be found in Appendix 12.1, Figure 12.4.

8.3.1 Power Plant

As with the NGCC with EGR the maximum amount of recycled flue gas is 45 %, this is due to the amount of O_2 going into the combustor. With 45 % the O_2 concentration is 14,5 % this also below the recommended value, but for the same reasons as with NGCC with EGR it will be tested.

Since the GT in the tail-end process are design to operate with a chosen EGR ratio, the pressure ratio over to compressor will be the same as for the base case, 18,28 bar. The change to the GT with use of EGR for this process comes from the change in composition and temperature for the gas flowing into the compressor. The inlet temperature increases as with the compressor in NGCC with EGR, but here the pressure is constant, therefore the power input to the compressor will behave differently, this can be seen in Figure 8.9. Here the compressor work decreases; this has some what to do with the CO_2 concentration at the compressor inlet. CO_2 has a lower ratio of specific heat than the other components so when it increases less work is required on the compressor.

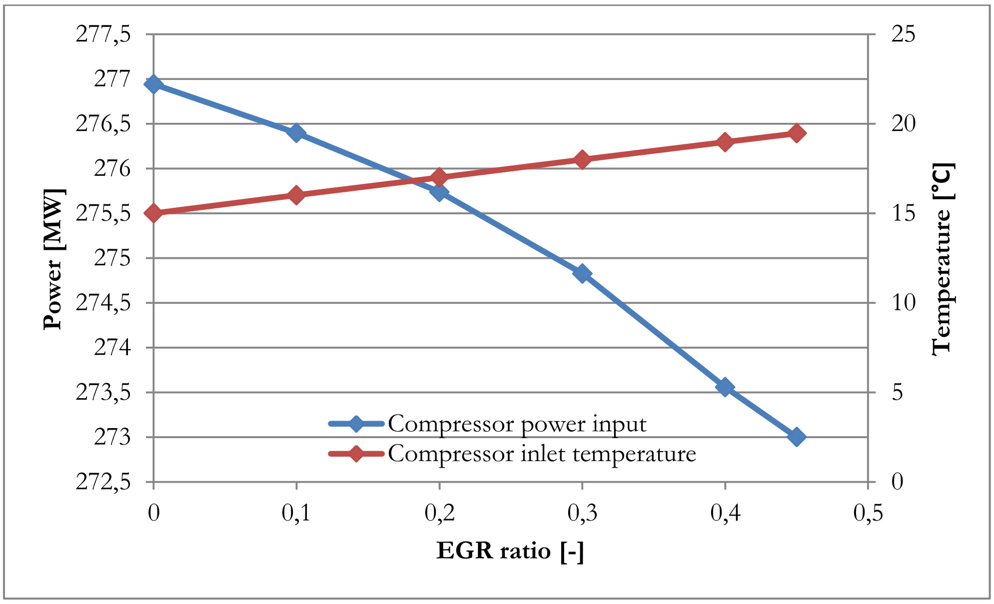

Figure 8.9 - Compressor inlet temperature and power consumption with different EGR ratio.

In Figure 8.10 the efficiency of the GT including the flue gas compressor and the expander after the capture plant is plotted against the EGR ratio. Here it can be seen that the efficiency increases for higher EGR ratio. That is because the mass flow into the flue gas compressor is reduced with higher EGR. The flue gas compressor and the expander will always have a negative output because there is no external firing between the two components as with a normal GT.

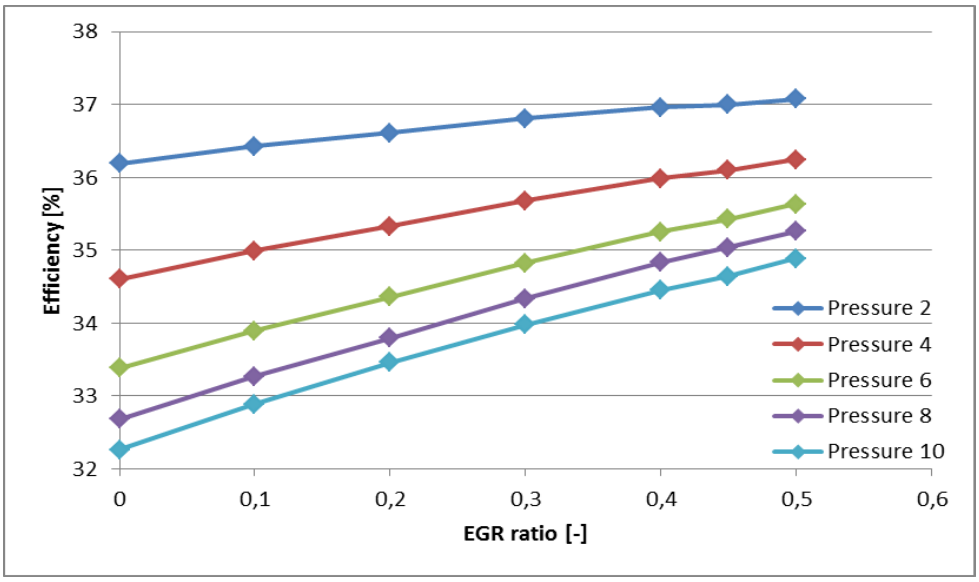

Figure 8.10 - GT efficiency with varying flue gas compression for different EGR ratio.

Through the capture plant there is a pressure loss and the mass flow is reduced because CO_2 is capture from the flue gas, these factors will also increase the loss in power output for GT#2. See Figure 8.11. Since GT#2 have a negative output the lowest pressure from the flue gas compressor gives the best efficiency for both GTs combined.

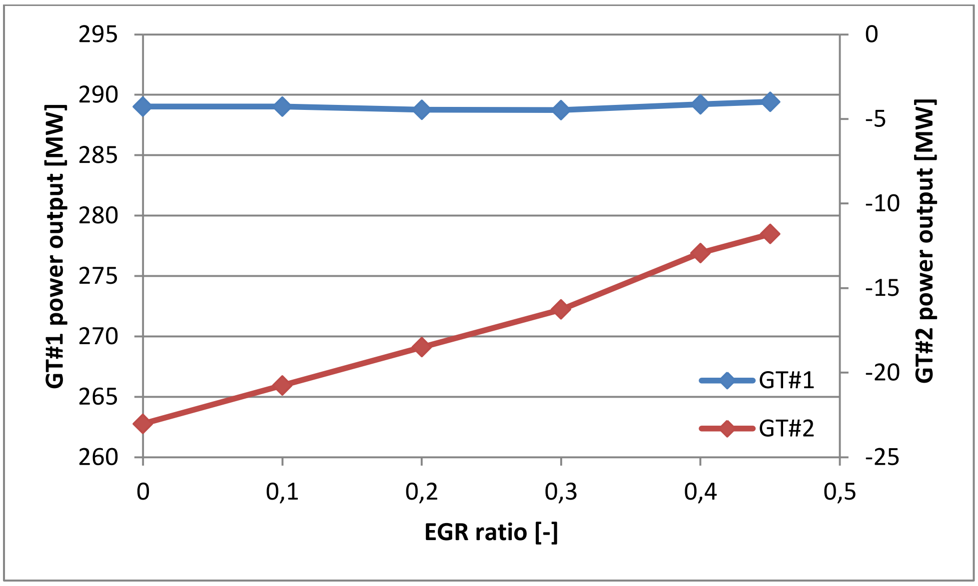

Figure 8.11 - GT#1 and GT#2 power output for different EGR ratio, the pressure ratio over flue gas compressor is constant on 2 bar.

The CO_2 partial pressure increases a lot with when the pressure and EGR ratio increases. This can be seen in Figure 8.12. Without the use of the flue gas compressor the change in partial pressure is very small, but when compressing the gas to 10 bar the partial pressure becomes 0,9 bar. This will improve the performance of the capture plant, because it is easier to capture CO_2 with higher partial pressure.

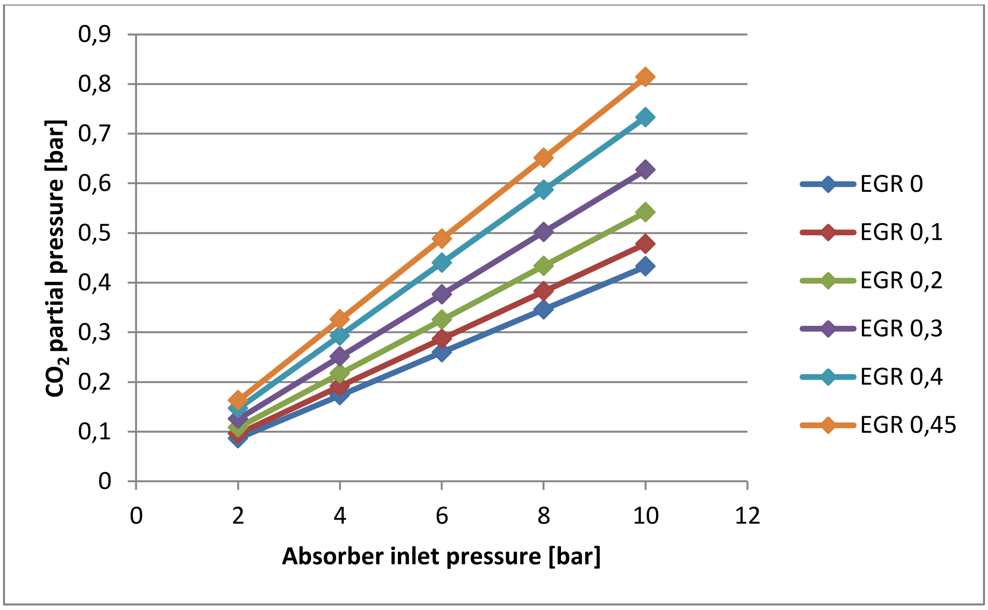

Figure 8.12 - CO_2 partial pressure against absorber inlet pressure for changing EGR ratio.

Another positive factor by using EGR and a flue gas compressor is that the volumetric flow rate will decrease drastically, from Figure 8.13 it can be seen that the flow rate is 285 m^3/s for EGR 0 % and pressure of 2 bar, when increasing the pressure to 10 bar and a EGR ratio of 45 % the flow is down to 30 m^3/s. The effect of EGR is much more noticeable for the gases with low pressure. When the pressure is 10 bar the change from an EGR ratio of 0 % to an EGR ratio of

45 % is 55-30 m^3/s and 2 bar the change is 285-139 m^3/s. This has a major impact on the size of the capture plant; this has been further evaluated in chapter 8.6.

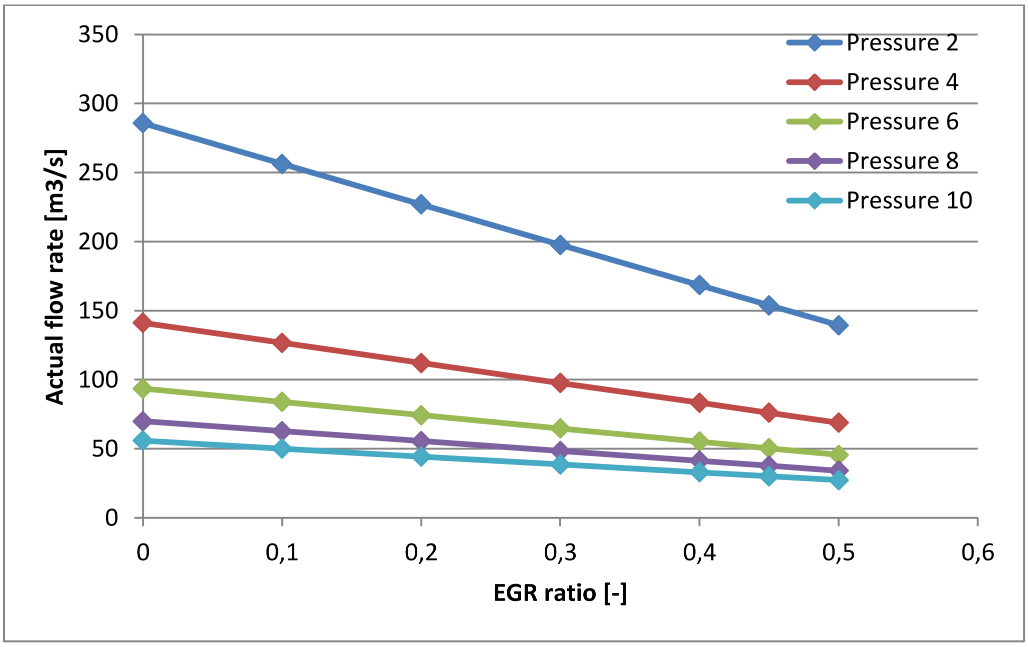

Figure 8.13 – Actual volume flow rate of the flue gas into the absorber for changing EGR ratio and flue gas pressure.

8.3.2 Capture Plant

8.3.2.1 MEA

For this case the variable parameters are the pressure of the flue gas and the amount of recycled gas. The results are presented in Figure 8.14. Here it can be observed that the change in reboiler duty are quiet small, it varies from 3,26-3,41 MJ/kg CO_2. All the results are within this area. When doing this sensitive analysis the simulation tool HYSYS can vary too much to get the desired results, but from Figure 8.14 a trend can be observed. The trend here is that for higher pressure the reboiler duty becomes lower and this fit according to the theory described in chapter 3. The lowest reboiler duty occurs at an inlet pressure of 10 bar and a EGR ratio of 45 %, the duty then becomes 3,26 MJ/kg CO_2.

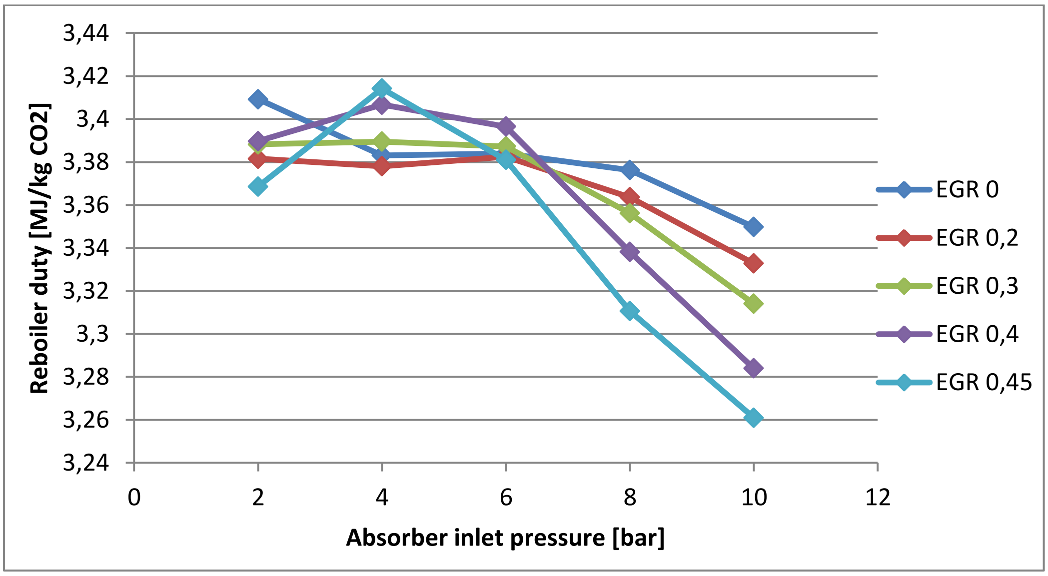

Figure 8.14 – Reboiler duty for different pressure into absorber for changing EGR ratio.

8.3.2.2 MDEA

In Figure 8.15 the reboiler duty needed when using MDEA is plotted against the absorber inlet pressure. Since MDEA are best suited for capture at higher pressure, capture was tested with absorber inlet pressures of up to 30 bar. This was done in order to better see the trends, for low-pressure the results for the MDEA is not so consistent as preferred, but for the higher pressure it can be clearly seen that the reboiler duty decreases. Absorber pressure from 14-30 bar have not been simulated in the NGCC because GT#2 have a negative output, and for high pressure that output will be more negative on the net plant efficiency than the gain from the lower reboiler duty.

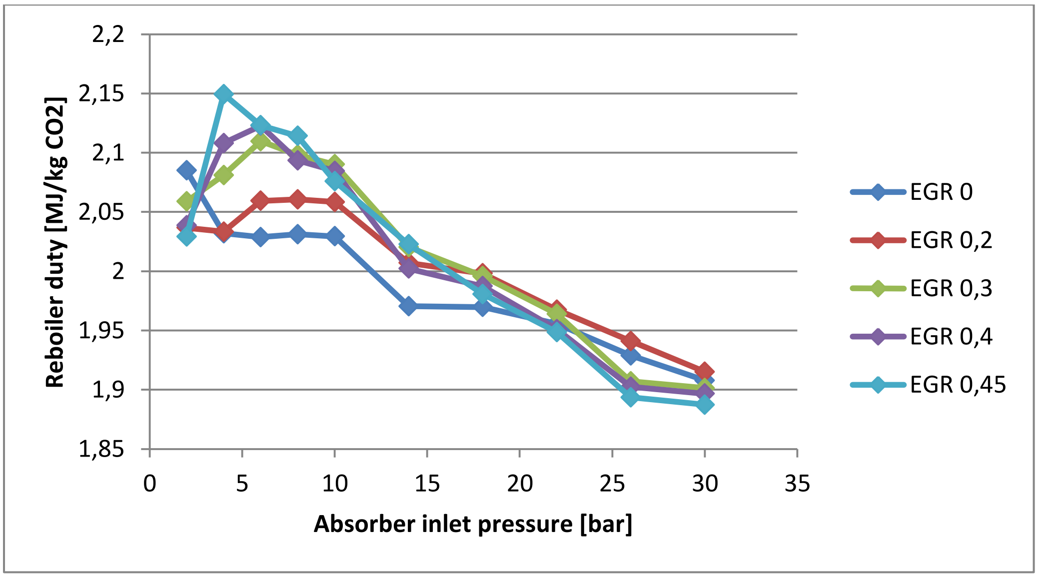

Figure 8.15 - Reboiler duty for different pressure into absorber for changing EGR ratio.

For a pressure of 10 bar the lowest reboiler duty with an EGR ratio of 0 %, there is quite small variation, but this does not fit according to the theory which states that the reboiler duty should decrease for higher partial pressure. The reason for this may be that MDEA do not react properly at lower pressures.

8.3.3 Conclusion

The total efficiency of tail-end with MEA can be seen in Figure 8.16, the best results from the use of MEA takes place with an EGR ratio of 45 % and an absorber inlet temperature of 2 bar. The net plant efficiency then becomes 50,9 %.

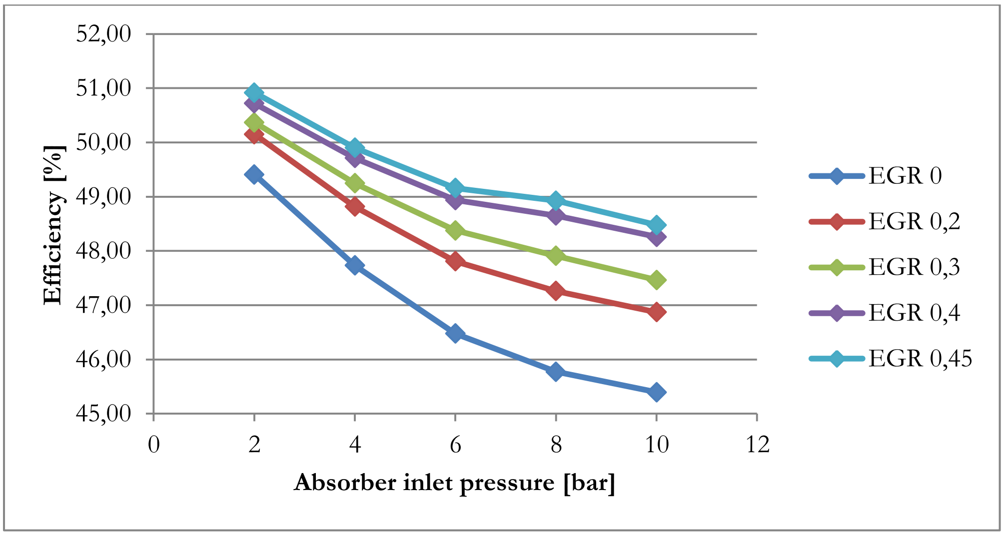

Figure 8.16 - Total efficiency for Tail-end with MEA.

The same results is presented for the MDEA case in Figure 8.17 as with the case with MEA the best results is with an EGR ratio of 45 % and that gives a net plant efficiency of 51,7 % which is a great deal higher than with MEA. So when capturing CO_2 at pressures above atmospheric pressure MDEA is more suitable than MEA, even as low as 2 bar. The loss in GT#2 is bigger than the gain of having a higher partial pressure of CO_2 in the flue gas. The net plant efficiency is higher than for the base case, this is mostly because of the gain from using MDEA instead of MEA.

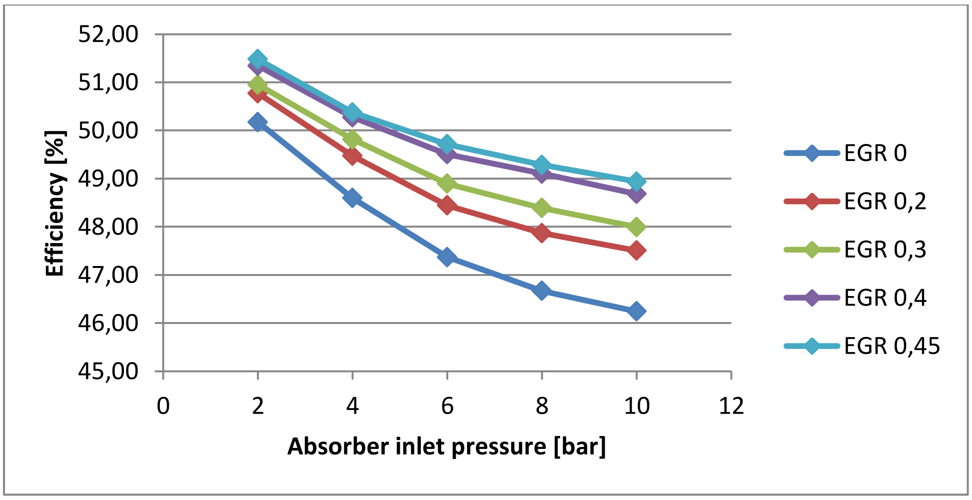

Figure 8.17 - Total efficiency for Tail-end with MDEA.

Tail-end MEA	Value	Unit
LHV input	740,62	MW
Auxiliaries & losses	9,54	MW
Gross gas turbine output	276,28	MW
Gross steam turbine output	123,43	MW
Plant efficiency without steam extracted	52,16	%
MEA capture		
Reboiler duty	3,37	MJ/kg CO_2
Steam demand in reboiler	59,20	kg/s
Mechanical work	0,82	MW
CO_2 compression	12,24	MW
CO_2 emitted	4,66	kg/s
Net plant efficiency	50,41	%
Efficiency penalty of CO_2 capture	**8,39**	**%-points**
Tail-end MDEA		
LHV input	740,62	MW
Auxiliaries & losses	9,94	MW
Gross gas turbine output	276,28	MW
Gross steam turbine output	131,77	MW
Plant efficiency without steam extracted	53,60	%
MDEA capture		
Reboiler duty	2,03	MJ/kg CO_2
Steam demand in reboiler	35,83	kg/s
Mechanical work	0,36	MW
CO_2 compression	15,5	MW
CO_2 emitted	4,46	kg/s
Net plant efficiency	51,69	%
Efficiency penalty of CO_2 capture	**7,11**	**%-points**

Table 8.4 - Results from Tail-end capture.

- Tail-end increases the partial pressure of CO_2 in the flue gas by utilizing EGR and a flue gas compressor.
- The reboiler duty is decreased by 44,3 % from the base case, this is mainly due to the use of MDEA. By using MEA the reboiler duty is reduced by 7,4 %.
- The best efficiency from the tail-end capture process comes from using MDEA as amine and the total efficiency is 51,7 %.
- The result gives a 7,1 %-points loss compared with the base case without capture, but an increase of 0,3 %-points compared to a base case with CO_2 capture.

8.4 Post-expansion

The post-expansion capture process is described in chapter 6.1. Here the results for the simulation of the process along with CO_2 capture with the use of both MEA and MDEA as amine. A complete schematic of the post-expansion process from HYSYS can be found in Appendix 12.1, Figure 12.2.

8.4.1 Power Plant

Since 90 % of the CO_2 from the recycled gas have been removed the O_2 concentration in the mixture of air and recycled gas have been changed compared with the other cases. From Figure 8.18 it can be seen that the O_2-concentration at the combustion inlet is higher than for the NGCC EGR case, 15,3 mol-%, but when increasing the EGR ratio to 50 % the O_2-concentration still becomes too lower to have satisfying combustion parameters in the combustion chamber. Therefore the maximum EGR used in this case is 45 %.

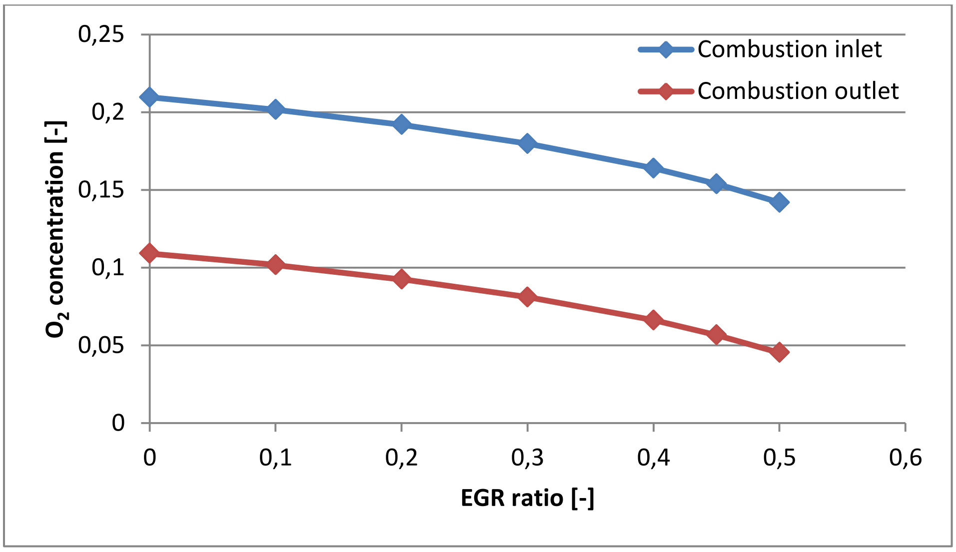

Figure 8.18 - O_2 concentration at the combustion inlet and outlet for different EGR ratio.

As with the other cases the use of EGR increases the temperature at the compressor inlet, for this case it makes the compressor power consumption increase, see Figure 8.19. The reason for the increase instead of the decrease as with the NGCC with EGR is the mole fraction of CO_2 is almost constant. CO_2 has a lower ratio of specific heat than the other components, therefore with less than in the NGCC with EGR case the result is higher temperature at the compressor outlet and therefore a higher power demand. The temperature increases at the compressor inlet will give an increase in compressor outlet temperature; form around 414-422 °C. This will result in a lesser fuel demand to reach the desired temperature of 1427 °C at the combustion outlet.

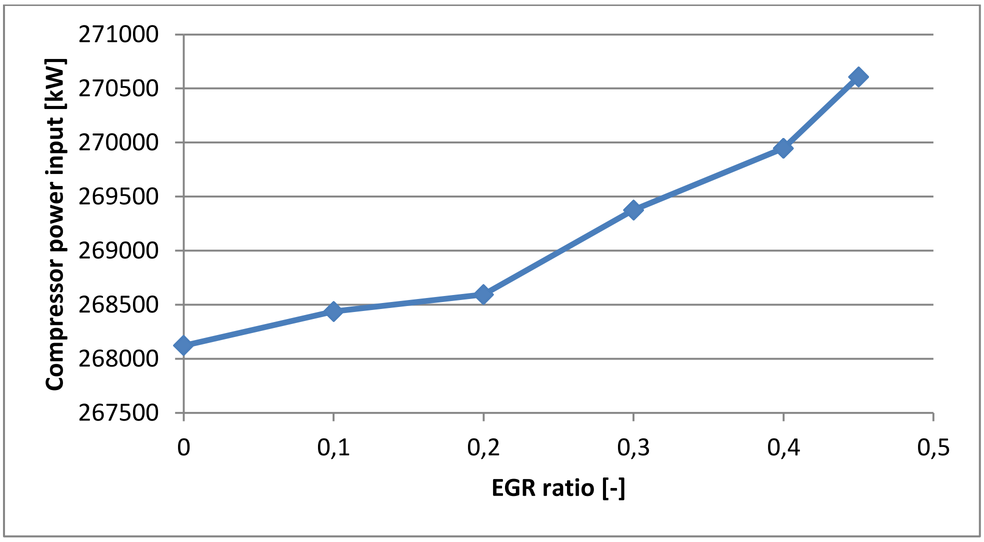

Figure 8.19 – Compressor work with changing EGR ratio.

When looking at the GT efficiency, see Figure 8.20 it can be seen that the efficiency is fairly constant for changing EGR ratio, but when the gas is extracted at a higher pressure from the HPT the efficiency decrease a lot. From 2-4 bar the efficiency decrease is 1,1 %-points, but the decrease is more moderate for higher pressure. The reason for the low efficiency at higher pressure is partly because of the 3 % pressure and 60 °C temperature drop over the heat exchanger prior to the capture plant; there are also pressure losses in the capture plant itself. It can be observed that the best GT efficiency is with EGR ratio of 0 %. This has to do with the increase in compressor work, due to the temperature increase at the inlet.

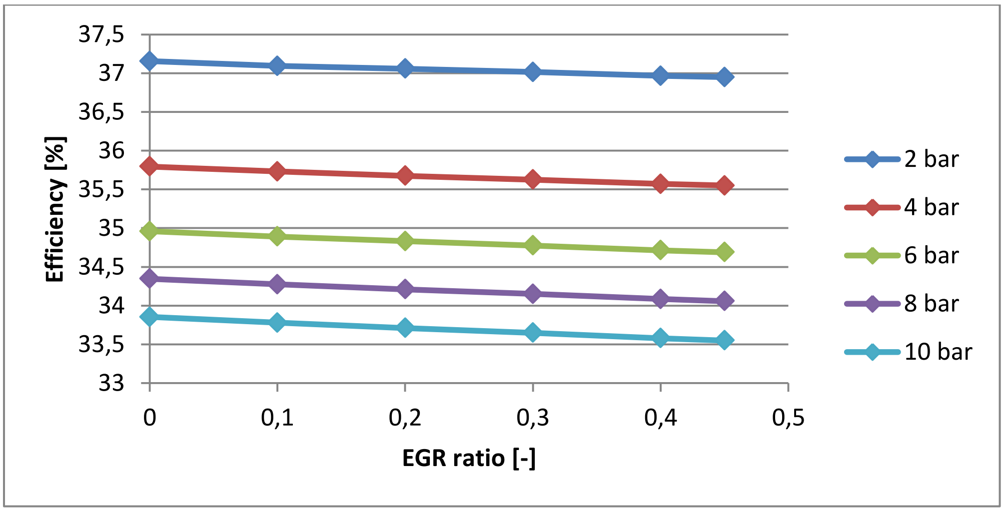

Figure 8.20 - GT efficiency with for different EGR ratio.

The steam cycles produces less power for this case than any other, this has to do with the low temperature at the LPT outlet. The temperature becomes as low as 570 °C, this has major consequence for the ST, see Figure 8.21. The low temperature is because of the high temperature drop over the heat exchanger prior to the LPT. When comparing the power output from the ST with the ST power output from the NGCC with EGR case, the difference is quite severe.

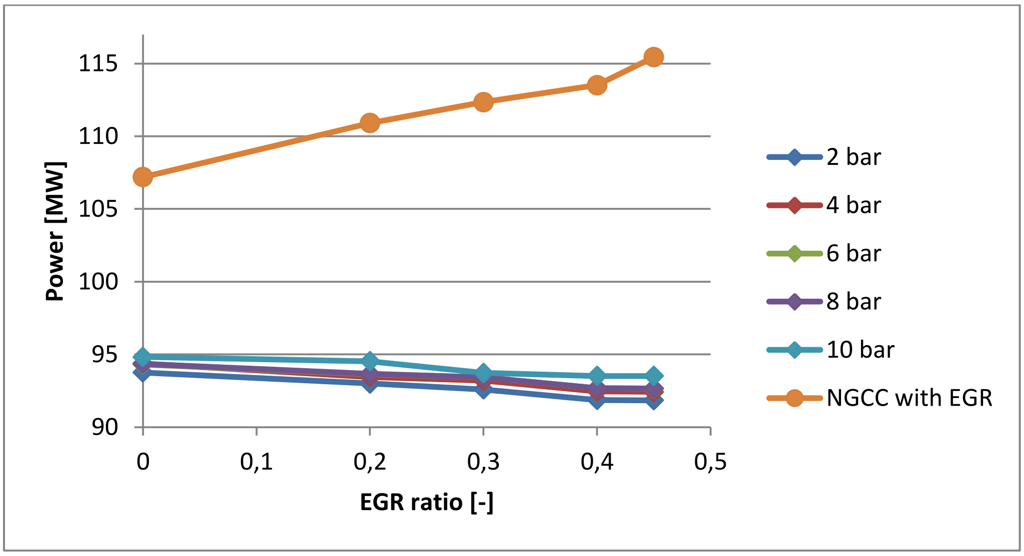

Figure 8.21 – ST power output with post-expansion MEA compared against NGCC with EGR.

8.4.2 Capture Plant

In this case the EGR does not help with increasing the CO_2 mol-% into the capture plant. When the EGR ratio is varied from 0-45 %, the mole fraction of CO_2 varies from 0,045-0,047. The same can be seen with the mass flow of flue gas. The mass flow decreases from 642-622 kg/s with 0 and 45 % EGR. This is because the capture plant is placed in between the HPT and the LPT. The gas leaving the LPT is already cleaned so the gas that is being recycled is gas containing only 10 % of the CO_2. That is the reason EGR does not increase the CO_2 mole fraction in the gas to the capture plant. The same goes for the mass flow, since capture takes place in between the turbines all of the mass flow has to be cleaned. Therefore is post-expansion a process where the use of EGR does not benefit for improving the CO_2 capture or the net plant efficiency of the plant in the same regards as the other processes.

8.4.2.1 MEA

The benefit from this process on the capture plant can be seen in Figure 8.22, the reboiler duty decreases slightly when the absorber inlet pressure increases. For the different EGR ratios the reboiler duty is almost constant for each of the pressures. This is because the mole composition and the mass flow into the absorber have not been changed to any notable degree by the use of EGR.

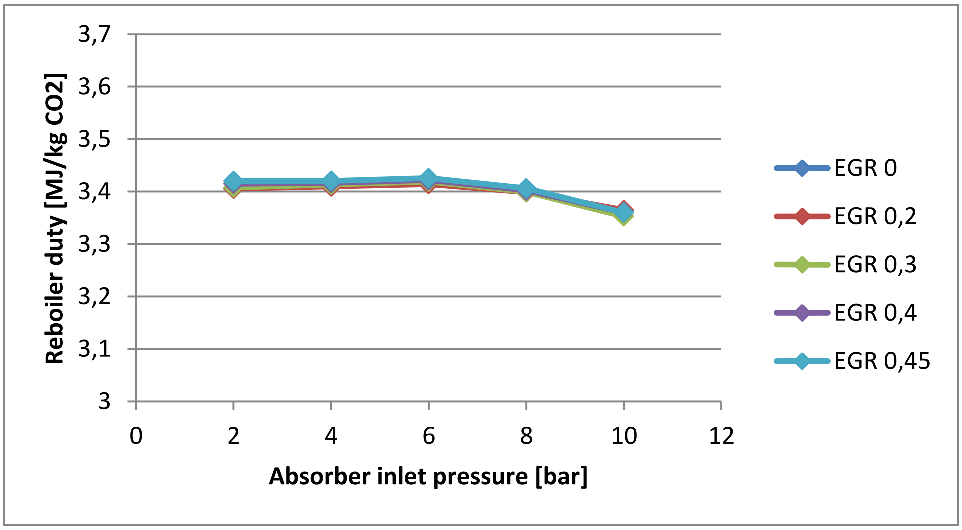

Figure 8.22 - Reboiler duty for different pressure into absorber for changing EGR ratio.

8.4.2.2 MDEA

As well as with the MEA the MDEA capture case have not been improved by the use of EGR, as seen in Figure 8.23, the reboiler duty are almost identical for the different EGR ratios, the only change in the duty from 0-45 % is 0,016. The reboiler duty decrease with higher pressure, but in this case the trend is a bit more marked than in the MEA case. This has to do with that MDEA are designed for higher pressure than MEA.

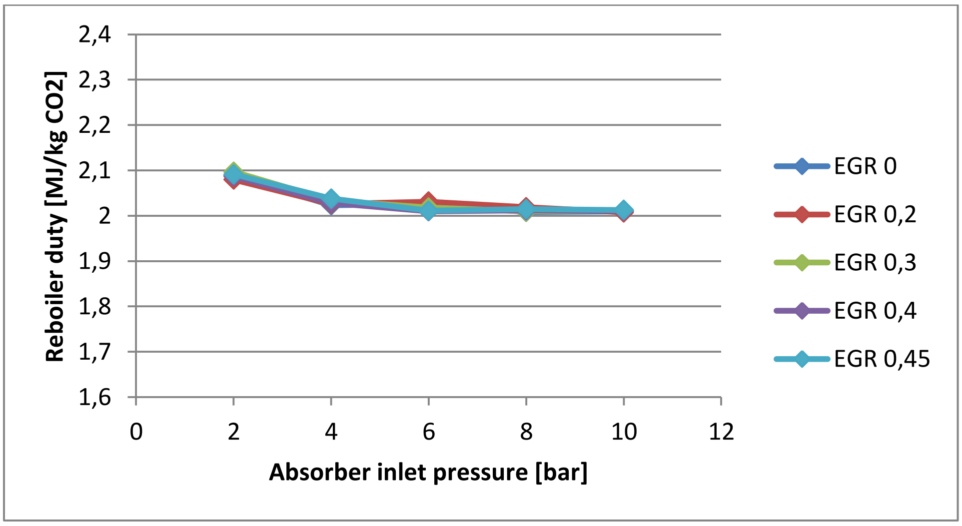

Figure 8.23 - Reboiler duty for different absorber inlet pressure with changing EGR ratio.

8.4.3 Conclusion

The net plant efficiency curves, Figure 8.24 and Figure 8.25 have the same development, the big difference is that MEA have a lower efficiency than the MDEA, this can be explained by the lower reboiler duty in the stripper. For both cases the best net plant efficiency is with the use of an EGR ratio of 0 %, this is closely linked with the compressor work and power output from the ST. With the EGR ratio of 0 the power from the fan and cooler pump for the recycled flue gas is removed.

For the absorber inlet pressure, a pressure of 2 bar gives the best benefit on the total plant. This is because there are higher pressure drop over the components between the two turbines at a higher pressure.

This process also have been tested with an 0 temperature loss heat exchanger and the result on the net plant efficiency was 52,15 % which is very good results. By developing better heat exchanger this type NGCC with CO_2 capture can be a great improvement.

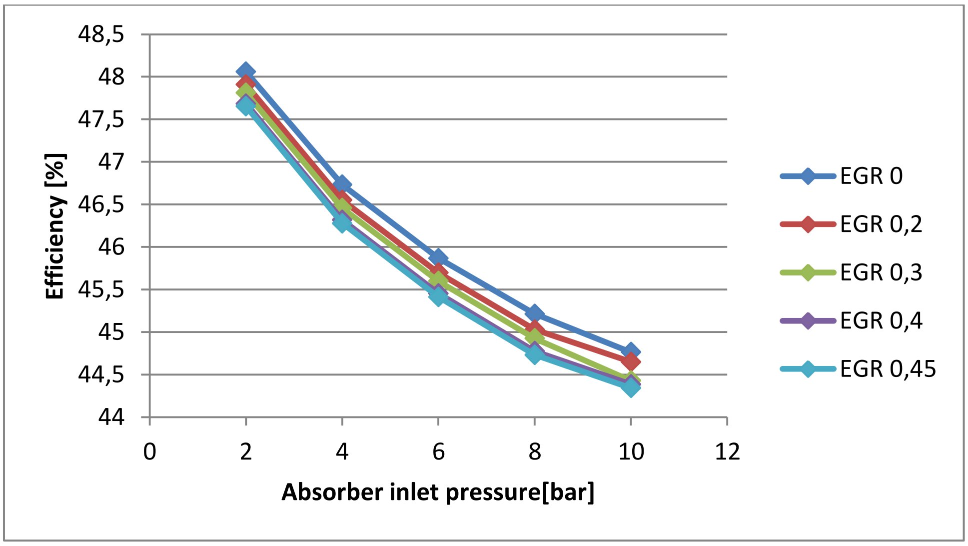

Figure 8.24 - Net plant efficiency with the use of MEA.

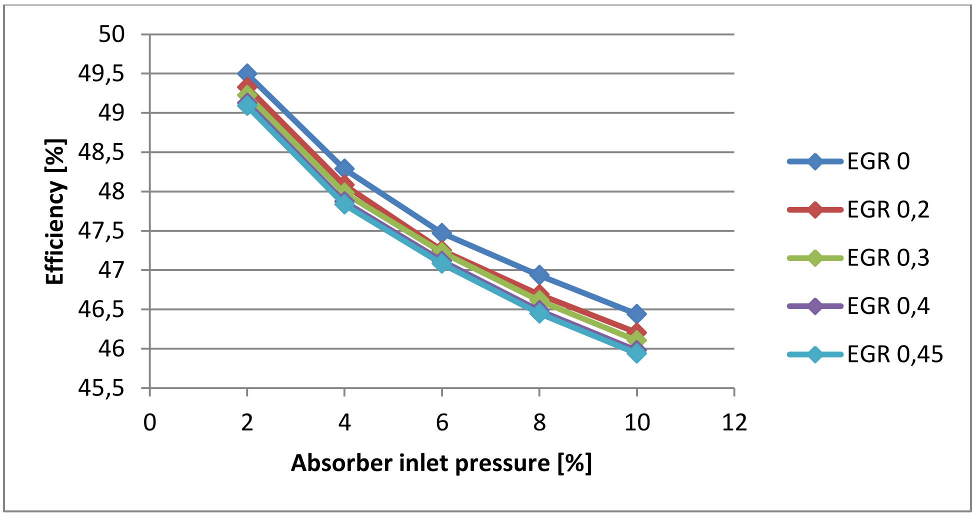

Figure 8.25 - Net plant efficiency with the use of MDEA.

Post-expansion MEA	Value	Unit
LHV input	737,09	MW
Auxiliaries & losses	3,23	MW
Gross gas turbine output	277,75	MW
Gross steam turbine output	96,98	MW
Plant efficiency without steam extracted	49,87	%
MEA capture		
Reboiler duty	3,41	MJ/kg CO_2
Steam demand in reboiler	60,96	kg/s
Mechanical work	0,6	MW
CO_2 compression	12,78	MW
CO_2 emitted	4,24	kg/s
Net plant efficiency	48,06	%
Efficiency penalty of CO_2 capture	**10,74**	**%-points**
Post-expansion MDEA		
LHV input	737,09	MW
Auxiliaries & losses	3,59	MW
Gross gas turbine output	277,75	MW
Gross steam turbine output	110,34	MW
Plant efficiency without steam extracted	51,64	%
MDEA capture		
Reboiler duty	2,09	MJ/kg CO_2
Steam demand in reboiler	36,74	kg/s
Mechanical work	0,3	MW
CO_2 compression	15,52	MW
CO_2 emitted	4,24	kg/s
Net plant efficiency	49,49	%
Efficiency penalty of CO_2 capture	**9,31**	**%-points**

Table 8.5 - Results from Post-expansion and CO_2 capture.

- Since capture plant is placed prior to the EGR spilt all the mass flow is taken through the capture plant.
- The use of EGR does not improved the partial pressure of the flue gas, extracting the gas at higher pressure does, but the loss in power output are too great for this improvement to help the net plant efficiency.
- The temperature drop over the heat exchanger between the two turbines reduces the temperature at the turbine outlet, resulting in low ST output.
- The best capture technology is the MDEA, giving a reboiler duty of 2,1 MJ/kg CO_2.
- The net plant efficiency with the use of MDEA is 49,5 %. Which is a 9,3 %-points loss compared with the base case without capture.
- Compared with the base case with capture this case have a 1,9 %-points loss.

8.5 Post-compression

With this process the benefit of compressing all the flue gas have been tested out. The process did not have the option of varying the absorber pressure due to the fact that the GT modification then would have involved too large capital costs associated with extra turbomachinery. The EGR compressor exit pressure was set to be the same as for the air compressor, 18,28 bar; however, due to the pressure losses through the heat exchangers the absorber pressure was only 17,73 bar. The HYSYS schematic of this process can be found in appendix 12.1, Figure 12.3.

8.5.1 Power Plant

In this process the GT efficiency increases quite a lot with higher EGR ratio, see Figure 8.26. This due to the fact that the all the flue gas is being compressed, therefore with high EGR ratio less extra air is needed from the air compressor. This can be seen in Figure 8.27, the air flow rate is decreasing with increasing EGR. The flue gas expander power output also decreases but the power demand from the compressor is higher. The reason for this is the loss in pressure through the heat exchanger and the absorber. The loss in mass flow by capturing the CO_2 also diminishes the output.

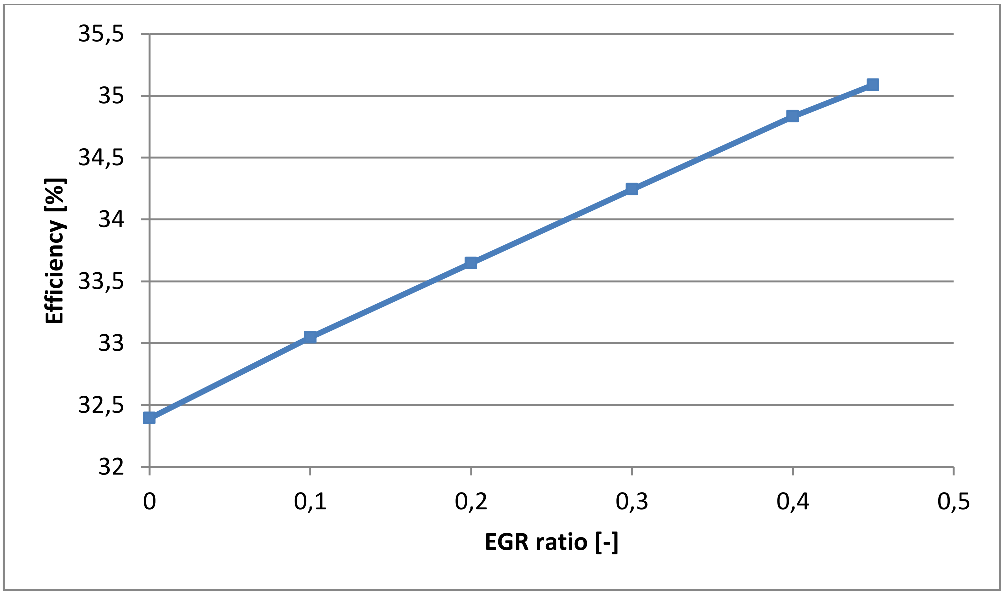

Figure 8.26 - GT efficiency for different EGR ratio.

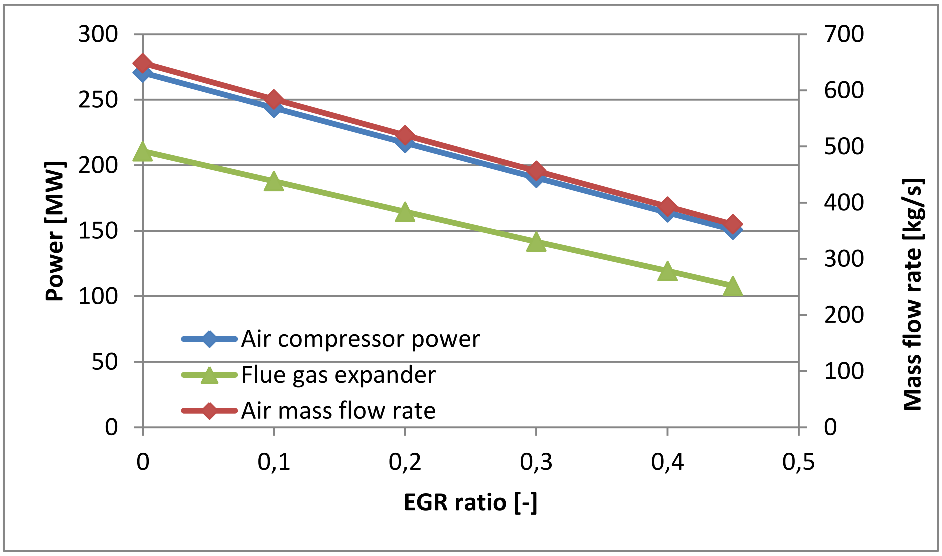

Figure 8.27 - Air compressor power input and flue gas expander on the left vertical axis and air mass flow rate on the right vertical axis, each plotted for different EGR ratio.

The temperature at the turbine outlet increases for higher EGR ratio, from 620 to 632 this will result in more steam production and higher power output from the ST. The increase is 144MW for EGR 0 % and 151 MW for EGR 45 %. This improves the overall plant efficiency.

8.5.2 Capture Plant

By having a pressure of 17,73 bar into the absorber the partial pressure of CO_2 increases from 0,8-1,5 bar with EGR ratio of 45 %. The benefit of this increase can clearly be seen by the decrease in reboiler duty, Figure 8.28 and Figure 8.29. The mass flow rate into the absorber have also been reduced from 633-345 kg/s, this change is the same as with the other cases, but the volumetric flow rate is much smaller compared to the other cases. For EGR ratio 0-45 % the flow is reduced from 31,6-17 m^3/s.

8.5.2.1 MEA

The reboiler duty decreases from 3,3 to 3,1 MJ/kg CO_2. Which is a 3,5 % decrease, resulting in a steam demand reduction of 2 kg/s from the ST cycle.

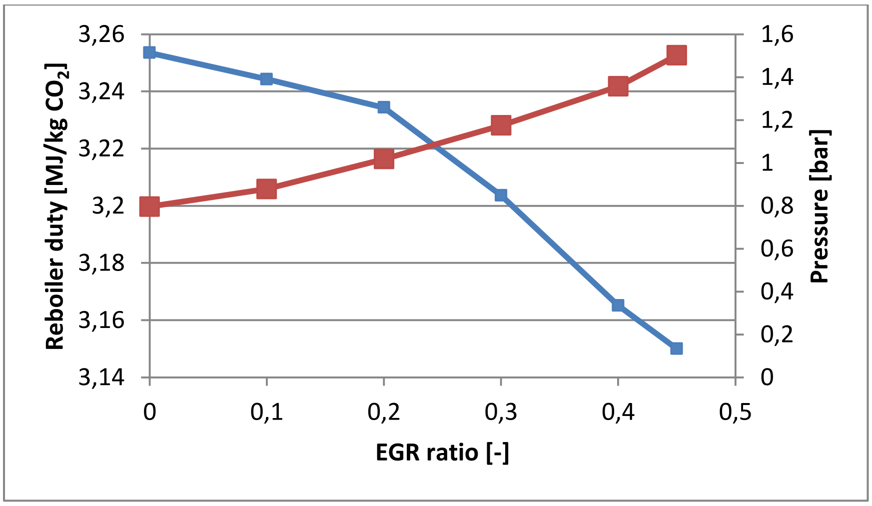

Figure 8.28 - Reboiler duty and CO_2 partial pressure in flue gas vs. EGR ratio.

8.5.2.2 MDEA

As for the MEA case the reboiler duty decreases for higher EGR ratio. The benefit with higher EGR ratio is not so large for MDEA in this case, the percentage increase is 1,8 %, form 1,98-1,95 MJ/kg CO_2. The high pressure has more impact in to reboiler duty than the concentration.

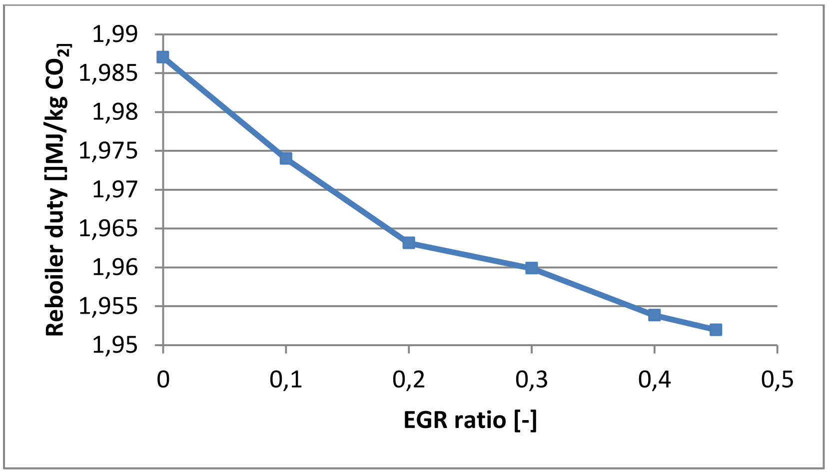

Figure 8.29 - Reboiler duty for different EGR ratio.

8.5.3 Conclusion

The total power plant efficiency with the use of both MEA and MDEA can be seen in Figure 8.30. The use of MEA is below MDEA for all the different EGR ratios. This can be expected since MEA is designed for lower partial pressures. The best overall efficiency is 49,5 %. This efficiency is much lower than the base case. Therefore it can be concluded that by compressing all the flue gas, too much energy is lost from pressure, temperature and mass loss in the capture plant.

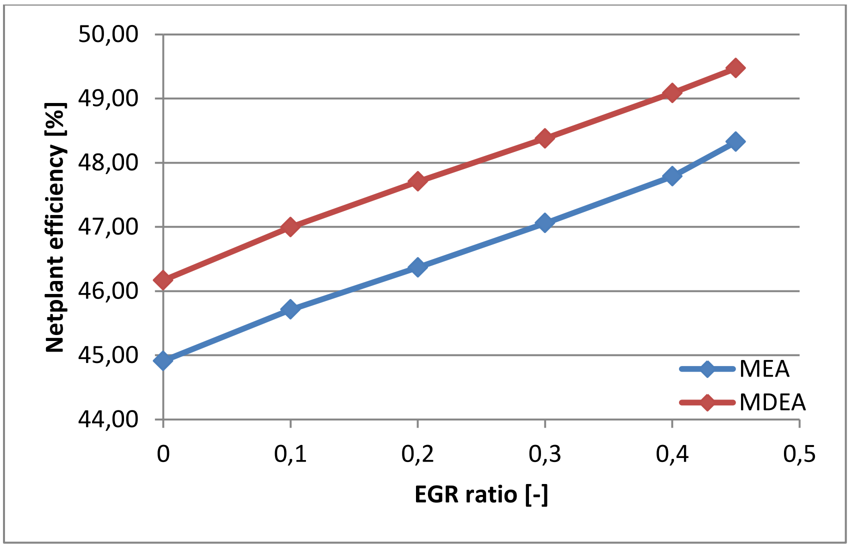

Figure 8.30 - Net plant efficiency for both MEA and MDEA for different EGR ratio.

Post-compression MEA	Value	Unit
LHV input	758,4	MW
Auxiliaries & losses	4,15	MW
Gross gas turbine output	679,63	MW
Compressor input	413,53	MW
Gross steam turbine output	126,27	MW
Plant efficiency without steam extracted	51,19	%
MEA capture		
Reboiler duty	3,15	MJ/kg CO_2
Steam demand in reboiler	56,71	kg/s
Mechanical work	5,52	MW
CO_2 compression	12,58	MW
CO_2 emitted	4,79	kg/s
Net plant efficiency	48,31	%
Efficiency penalty of CO_2 capture	**10,49**	**%-points**
Post-compression MDEA		
LHV input	758,4	MW
Auxiliaries & losses	4,51	MW
Gross gas turbine output	679,63	MW
Compressor input	413,53	MW
Gross steam turbine output	137,93	MW
Plant efficiency without steam extracted	52,68	%
MDEA capture		
Reboiler duty	1,96	MW
Steam demand in reboiler	35,4	kg/s
Mechanical work	4,85	MW
CO_2 compression	15,87	MW
CO_2 emitted	4,52	kg/s
Net plant efficiency	49,46	%
Efficiency penalty of CO_2 capture	**9,34**	**%-points**

Table 8.6 - Results from Post-compression and CO_2 capture.

- The GT efficiency increased for higher EGR ratio, from 32-35 %. This was mainly due to decreased mass flow into the air compressor.
- ST power output increased by 4,7 % for EGR ratio 0-45 %.
- The partial pressure of CO_2 in the flue gas became 1,5 bar and the volumetric flow rate decreased from 31,6-17 m^3/s for 45 % EGR.
- The reboiler duty decreased by 44,2 % compared with the base case.
- The net plant efficiency with the use of MDEA have a 2 %-points drop compared with the base case.

8.6 Absorber Sizing

The size of the absorber has been calculated for each of the different cases, both height the diameter. From the equations in chapter 3.7 it can be seen that the partial pressure of the CO_2 in the flue gas as well as the mass flow plays an important role in the calculations of the absorber size. A summary of the absorber size from the different processes can be found in Table 8.7. In the Figure 8.31 the diameter of the absorber for different EGR ratios for each of the processes are plotted. The tail-end process and the post-expansion process is represented by a pressure of 2 and 10 bar into the absorber, this is to see the benefit higher pressure have on the diameter of the absorber. The lowest diameter achieved from all of the processes is with the post-compression process. This case has a diameter of only 3 m, with 45 % EGR ratio. The flue gas is almost the same in each of the processes, therefore the main reason for the different diameters are the pressure at the inlet. When comparing the diameter of the absorber in the NGCC with EGR and tail-end with 2 bar the pressure difference is only around 1 bar and the decrease in diameter is from 10 to 7 m. This means that the change in diameter is higher for lower pressure. The benefit on the diameter is best at low pressures, 2-4 bar. When the pressure becomes high as around 10-17 bar the gain is almost neglectable compare to lower pressure. From the tail-end with 10 bar to post-compression the only difference is 0,8 m, and the pressure difference is 7,73 bar. The post-expansion have close to constant diameter for each of the different absorber inlet pressures. This has to do with the before mentioned mass flow into the absorber. Since it does not change much for different EGR ratio the diameter becomes constant.

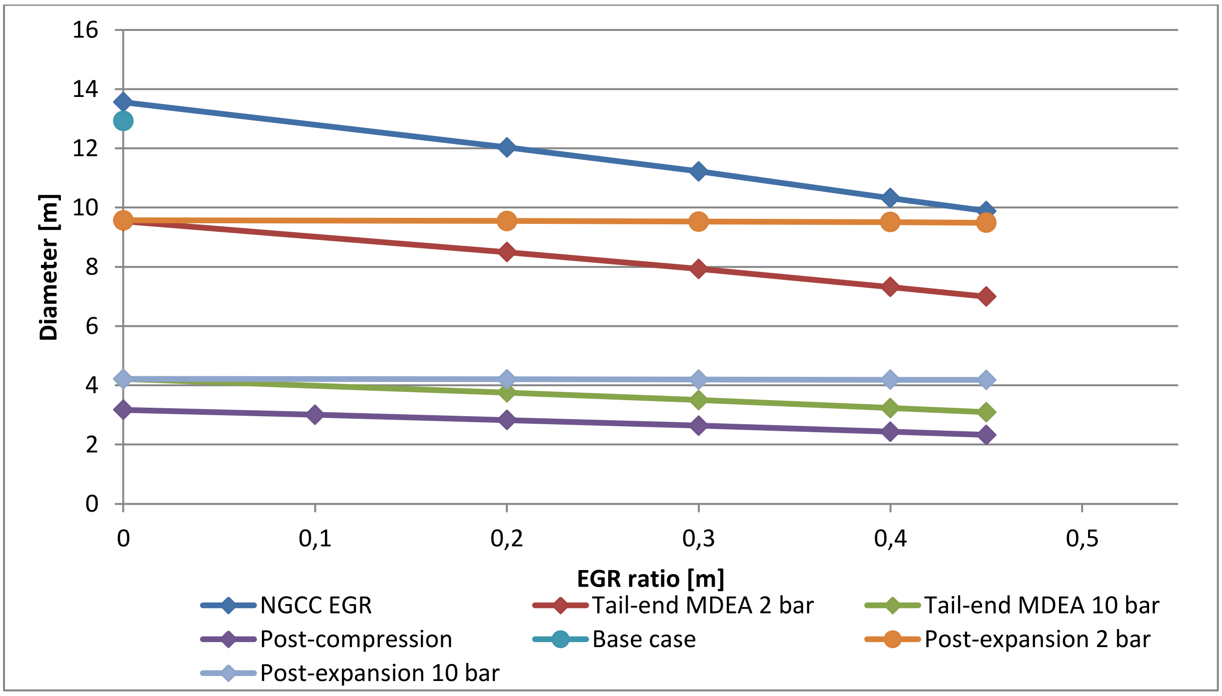

Figure 8.31 – Absorber diameter for different EGR ratio for each of the technology.

The height of the absorber for each of the processes is presented in Figure 8.32. It can be observed with the tail-end case the height is not so dependent on the absorber inlet pressure, the pressure of 2 bar and 10 bar is almost at equal standing when it comes to the height. The height is more dependent on the amount of CO_2 captured in the absorber. The reason for the small jumps along the tail-end height line is because the capture percentage can vary between 89,5 and

90,5 %. The amount of CO_2 into the absorber increases, but the percentage captured varies for each case. Since the mass flow occurs in the diameter equation and on the opposite side in the height equation the height is not depended on the mass flow into the absorber. The lowest absorber comes with the use of post-compression, this is because the pressure in this case is as high as 17,73 bar. The height of the absorber in the post-expansion case is almost constant for each of the pressure levels. This because the molar composition of the flue gas into the absorber is fairly constant, which results in an even capture ratio for each of the EGR ratios.

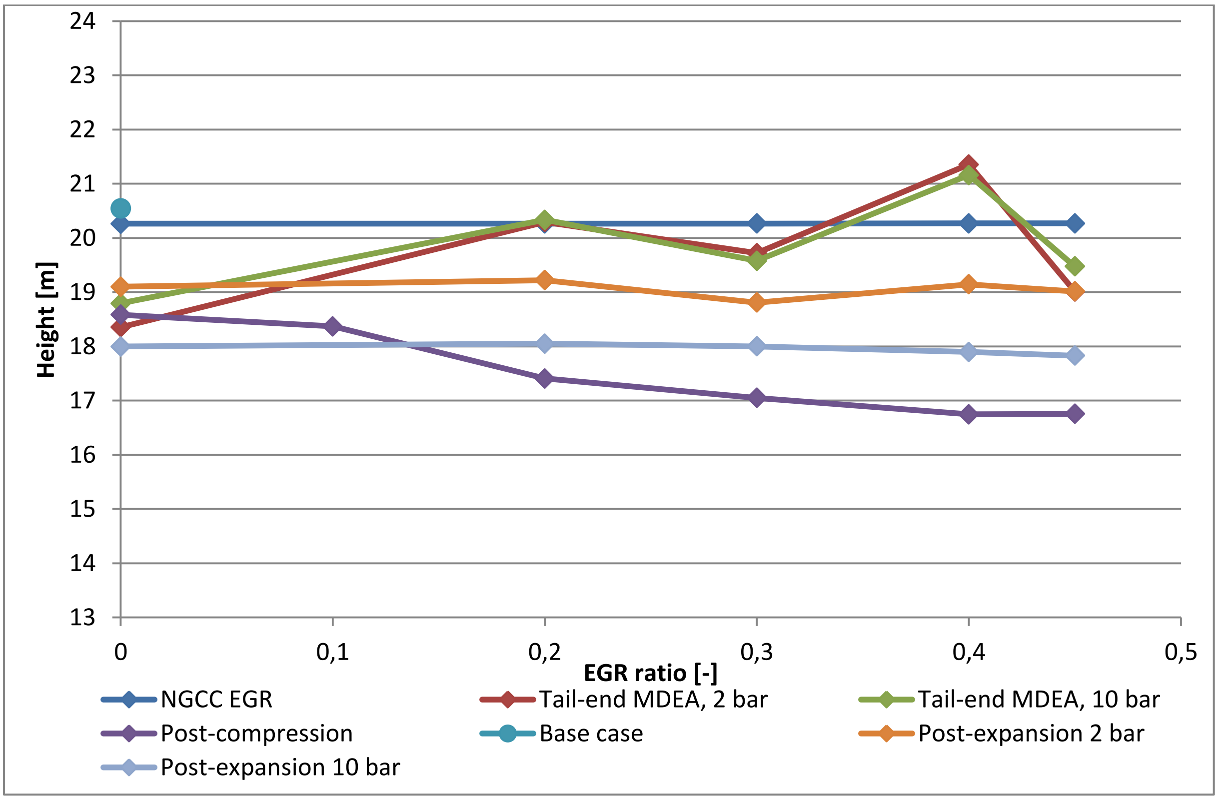

Figure 8.32 - Absorber height for different EGR ratio for each of the technology.

Process	Inlet pressure	EGR	Mass flow [kg/s]	Diameter[m]	Height [m]
NGCC base case	1,044	0	658	13	20,5
NGCC with EGR	1,044	45	352	9,4	20,3
Tail-end MDEA	2	45	348	6,6	18,3
Post-compression MDEA	17,73	45	345	3	10,5
Post-expansion MDEA	2	0	640	9,6	19,1

Table 8.7 – Absorber size for each of the process at the best net plant efficiency.

8.7 Summary

Here all the different processes will be compared with each other based on the net plant efficiency and the losses from each of the processes, the technological maturity will be considered for each of the processes.

In Table 8.8 the summary of some of the most important data for each of the processes are presented. Here it can be observed that the best technology to use is either a NGCC with EGR or a tail-end capture process with MDEA as solvent. Both of these have almost the same net plant efficiency 52 and 51,7 %. The two processes post-expansion and post-compression have both low net plant efficiency.

When comparing the results from the NGCC with EGR to other studies of this technology[28], which have gotten the result of 50,5 %. The results from this thesis are to some degree better. That is because the NGCC without capture has a high net plant efficiency of 58,8 %, and since the NGCC with EGR and the elevated pressure cycles uses the same design, the net plant efficiency of these cases will also be higher than in other studies.

Parameter	Unit	Base case	NGCC with EGR	Tail-end	Post-compression	Post-expansion
EGR	%	0	45	45	45	0
Absorber pressure	Bar	1,044	1,044	2	17,73	2
Reboiler duty	MJ/kg CO_2	3,64	3,39	2,03	1,96	2,09
Mechanical work[4]	MW	19,3	15,6	15,9	20,7	15,8
CO_2-emission	kg/s	4,14	4,32	4,46	4,52	4,24
GT efficiency	%	39,7	38,5	36,6	35,1	37,1
Net power output	MW	394	381	383	375	385
Net plant efficiency	**%**	**51,4**	**52**	**51,7**	**49,5**	**49,5**
Efficiency penalty	**%-points**	**7,4**	**6,8**	**7,1**	**9,3**	**9,3**

Table 8.8 - Summary of the different power plant configurations with CO_2 capture.

The elevated pressure cycles does not need a fan prior to the absorber to prevent backpressure drop, this should give a lower mechanical work compared with the other cases. The rich solution in the MDEA cases is throttled down to a lower pressure than in the MEA cases, therefore the CO_2 compression process requires more power than in the MEA capture cases, therefore high overall mechanical work. The post-compression process requires a fan installed prior to the flue gas cooler to prevent backpressure drop on the turbine, resulting in the highest mechanical work.

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[4] The mechanical work includes the compression work for compressing the CO_2 to transports ready condition as well as capture plant mechanical work.

All of the processes have 90 % CO_2-capture so the CO_2 emission is fairly constant for each case. The lowest is for the base case with 4,1 kg/s and the highest for the post-compression with 4,5 kg/s.

All the losses in percentage of the net plant efficiency of the base case without capture are presented in Figure 8.33. Here the impact each loss have on the total plant can be seen. The reboiler duty is clearly the dominating factor for the base case and EGR. For the elevated pressure cycles the CO_2 compression work is almost as demanding, this has to do with lower reboiler duty for MDEA and the low pressure for the CO_2 out of the stripper.

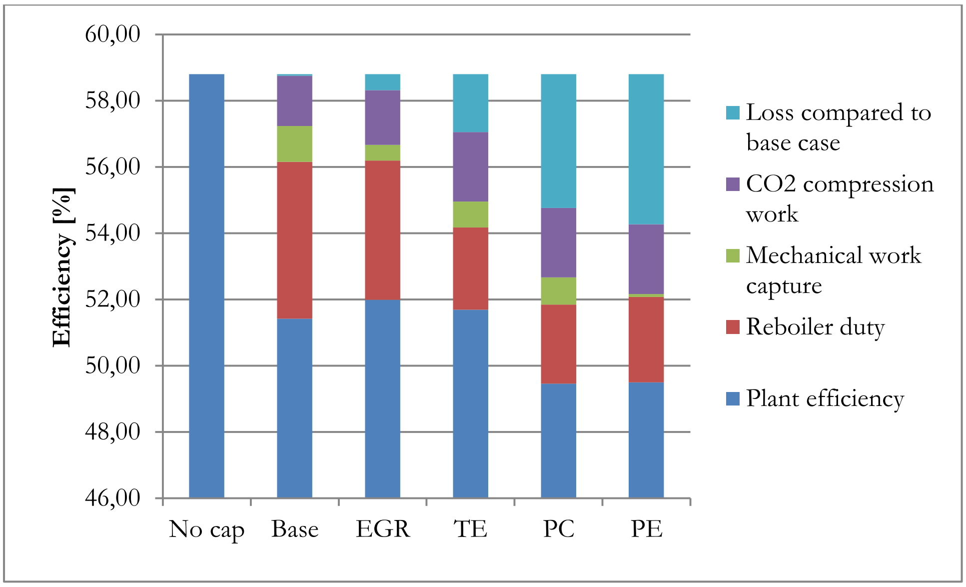

Figure 8.33 - Diagram for the efficiency penalty in each of the processes.

In the Figure 8.34-Figure 8.38 the percentage loss from each part of the processes is compared against the total loss from the processes. Here it can be observed that the mechanical work in the post-expansion process is almost neglect able compared to the other losses. The mechanical work for the rest of the cases is more dominating.

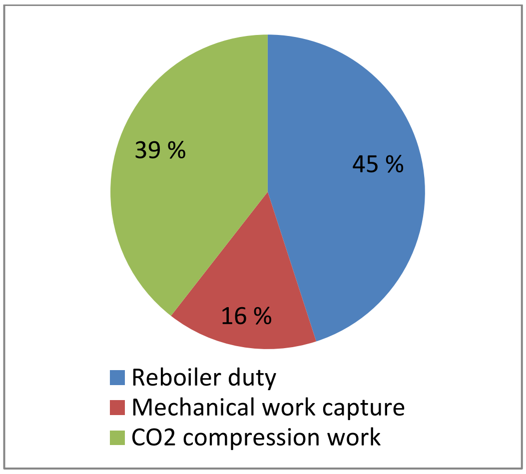

Figure 8.34- Post-compression CO2 capture.

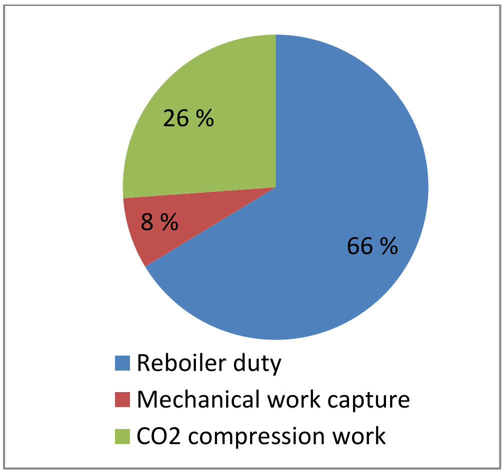

Figure 8.35 - NGCC with EGR.

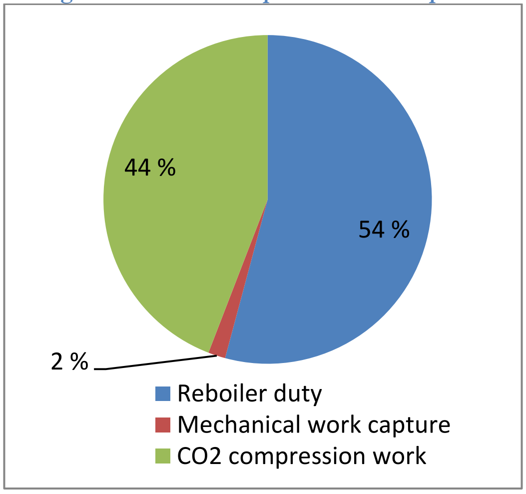

Figure 8.36 - Post expansion CO_2 capture.

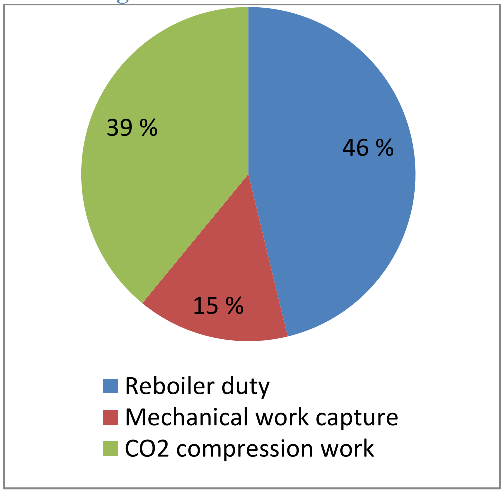

Figure 8.37 - Tail-end co2 capture.

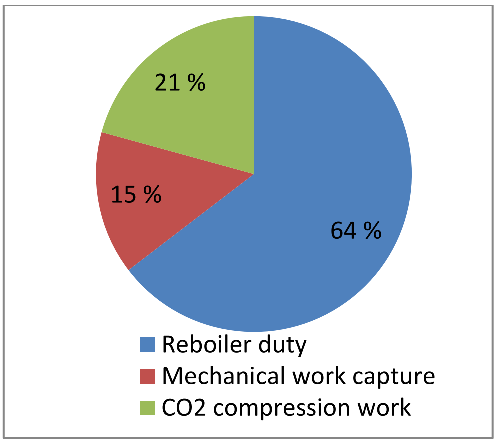

Figure 8.38 - Base case NGCC with CO_2 capture.

Technological Maturity

The maximum EGR ratio used in today's gas turbine is 35 %, to look into the maturity of the technology the results from this thesis have been compared with the maximum used with today's GT technology. The results are presented in Table 8.9.

Parameter	Unit	NGCC with EGR	Tail-end	Post-compression	Post-expansion
Efficiency EGR 45 %	%	52	51,7	49,5	49,1
Efficiency EGR 35 %	%	51,8	51,1	48,7	49,2
Difference 35%-45%	**%-points**	**0,2**	**0,6**	**0,8**	**-0,1**

Table 8.9 - Comparison between EGR 35 % and EGR 45 %.

The post-compression is sensitive to the use of EGR, when the it have been reduced to 35 % the net plant efficiency is reduced by 0,8 %-points. The same goes for tail-end with a reduction of 0,6 %-points. The post-expansion process have the best efficiency with 0 % so when reducing the EGR the efficiency increases with 0,1 %-points.

The elevated pressure cycles requires more equipment's than the base case and the EGR case. This will increase the capital cost of building plant with this design as well as maintains costs. The size of the NGCC will also be increased. In the table below the different number of equipment's each of the processes uses are presented.

Parameter	Base case	NGCC with EGR	Tail-end	Post-compression	Post-expansion
Compressors	1	1	2	2	1
Turbines	1	1	2	2	2
Heat exchangers	0	0	1	1	1
Flue gas coolers	1	2	2	2	2

Table 8.10 - Number of process equipment for the different NGCC cases.

The post-compression and the post-expansion process require heat exchangers that can handle high temperature. Improvements can be made both for increasing net plant efficiency but also reduction on the heat exchanger size. The post-expansion process with a heat exchanger with 0 temperature difference on the hot streams results in a net plant efficiency of 52,2 %. This is clearly not a mature technology, but something that can be used in the future.

All of the processes use EGR and all of them except post-expansion can benefit from having a higher EGR ratio. Therefore by developing better combustion chambers all of these processes will improve, because they can recycle more of the flue gas.

9 Conclusion

This thesis have looked into different ways modifying the natural gas combined cycle to be able to capture CO_2 either at elevated pressure and/or with higher CO_2 concentration. There has been established a base case which is used as a comparison, this case uses MEA as solvent in the capture plant. The net plant efficiency of each of the process modification has been compared. The reduction in absorber size for the different processes has been looked into, as well as maturity of the technology. Each of the elevated processes has been compared with both MEA and MDEA as solvent.

The steam used in the reboiler for stripping the CO_2 from the amine is extracted for the steam cycle. This thesis has looked into different extraction point and pressures. The conclusion is to extract the steam for the crossover between the intermediate pressure turbine and the low-pressure turbine. The extracted steam will be saturated with water to minimize the amount of required steam. The desired extraction pressure and temperature was found to be 3,6 bar and 140 °C.

The most promising power cycle in this thesis was the NGCC with EGR, this process uses MEA as solvent in the capture plant, and there are small changes done to the power plant. The net efficiency of this process is 52 %, which is a 6,8 %-point loss compared to a NGCC without CO_2 capture.

For the elevated pressure cycles the best option was the tail-end CO_2 capture process with the use of MDEA as solvent. This process requires one extra set of turbine and compressor, also an extra flue gas cooler. With this process part of the flue gas is recycled back to the gas turbine, the rest is compressed and sent to the capture plant at elevated pressure. After the CO_2 has been removed the remaining flue gas is taken through an expander before being released to the atmosphere. This process was best used with as high an exhaust gas recycle ratio as possible, in this thesis the maximum recycled gas was 45 %. The result show that the net plant efficiency was 51,7 % which is 7,1 %-point drop from NGCC without capture.

Post-expansion is a very promising technology for the future. The process extracts the flue gas from the turbine, cools it down and removes the CO_2 in a capture plant. The clean gas is heated up again and expanded in a second turbine. The main problem with this process is the pressure and temperature difference in the heat exchanger, with improvements on that part the net plant efficiency can improve quite a lot. This thesis have a net plant efficiency of 49,5 %, this is a 9,3 %-points drop from the NGCC without capture.

Post-expansion is the only process where all the flue gas is compressed to 18,3 bar. This is the least promising technology, there can be made improvements on the heat exchanger but the temperature difference was set to 20 °C which is quite good. The net plant efficiency became 49,5 %. Almost the same as post-expansion, but the room for improvements on this process are not as clear as for post-expansion.

When using MDEA as solvent for flue gas with high pressure the reboiler duty can be reduced up 50 % compared with a normal MEA capture plant, this is very encouraging results. The problem is that the loss from the gas turbine is greater than the gain from the capture plant.

The combination of EGR and high-pressure absorption with the use of MDEA is a potentially effective method in order to reduce the high capital and operating costs of today's state-of-the-art amine plant. The main problem with this type of processes is the combustion chamber and the high temperature heat exchangers. In the future combustion chambers that can burn with very low concentration of O_2 needs to be developed and better heat exchangers.

10 Future Work

In this thesis Aspen HYSYS have been the main simulation program, but for future work a more accurate program should be used with regards to the capture processes, either AspenTech's Aspen Plus or Sintef's CO_2-sim could be used. The models used in HYSYS can also be improved to get better and more consistent results.

To improve the capture process further, the used of lean vapor recompression could be used. By compressing parts of the lean solvent and sending it back into the stripper the reboiler duty could be reduced to some degree.

The amount of CO_2 leaving the stripper with the lean solution is fixed in this thesis. To optimize each process better, this amount should be varied for each of the flue gas composition from the power plant. Thereby the reboiler duty could be reduced more for the specific cases.

For the post-expansion and post-compression both uses high temperature heat exchangers, the temperature difference over the heat exchanger is a big loss on the processes. Ways of separating the CO_2 from flue gas at high temperature could be looked into, thereby removing the use of the heat exchanger.

The tail-end process is quite promising and should be further tested with different capture systems as well as new kind of solvents. The use of membranes technology as a part of the capture plant could be evaluated.

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12 Appendix

12.1 HYSYS Simulation Schematics

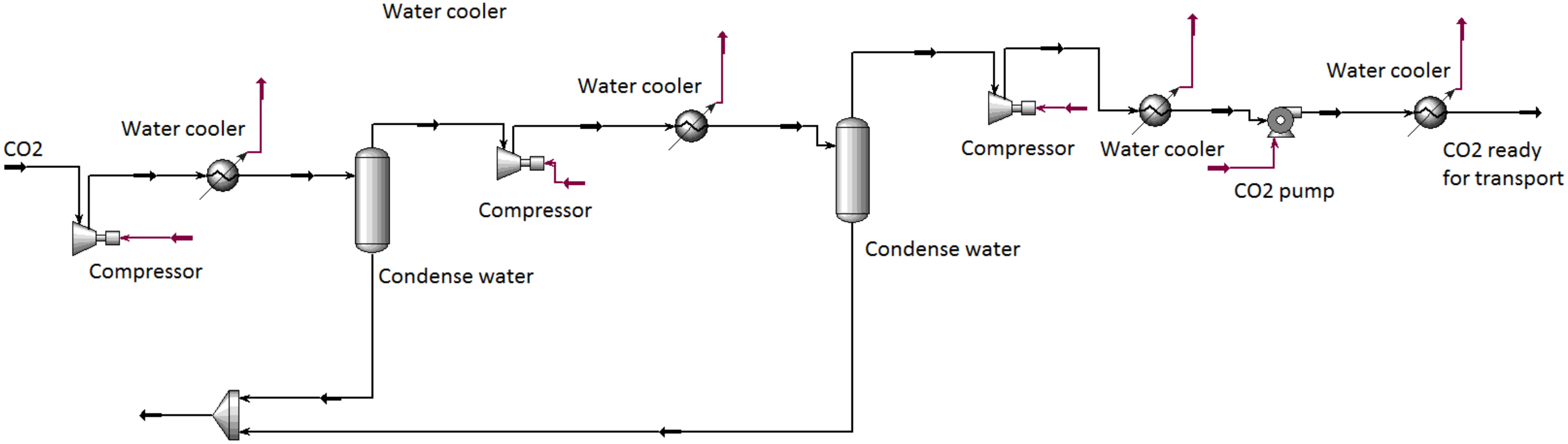


Figure 12.1 - Schematic of CO_2 compression and condensing process.

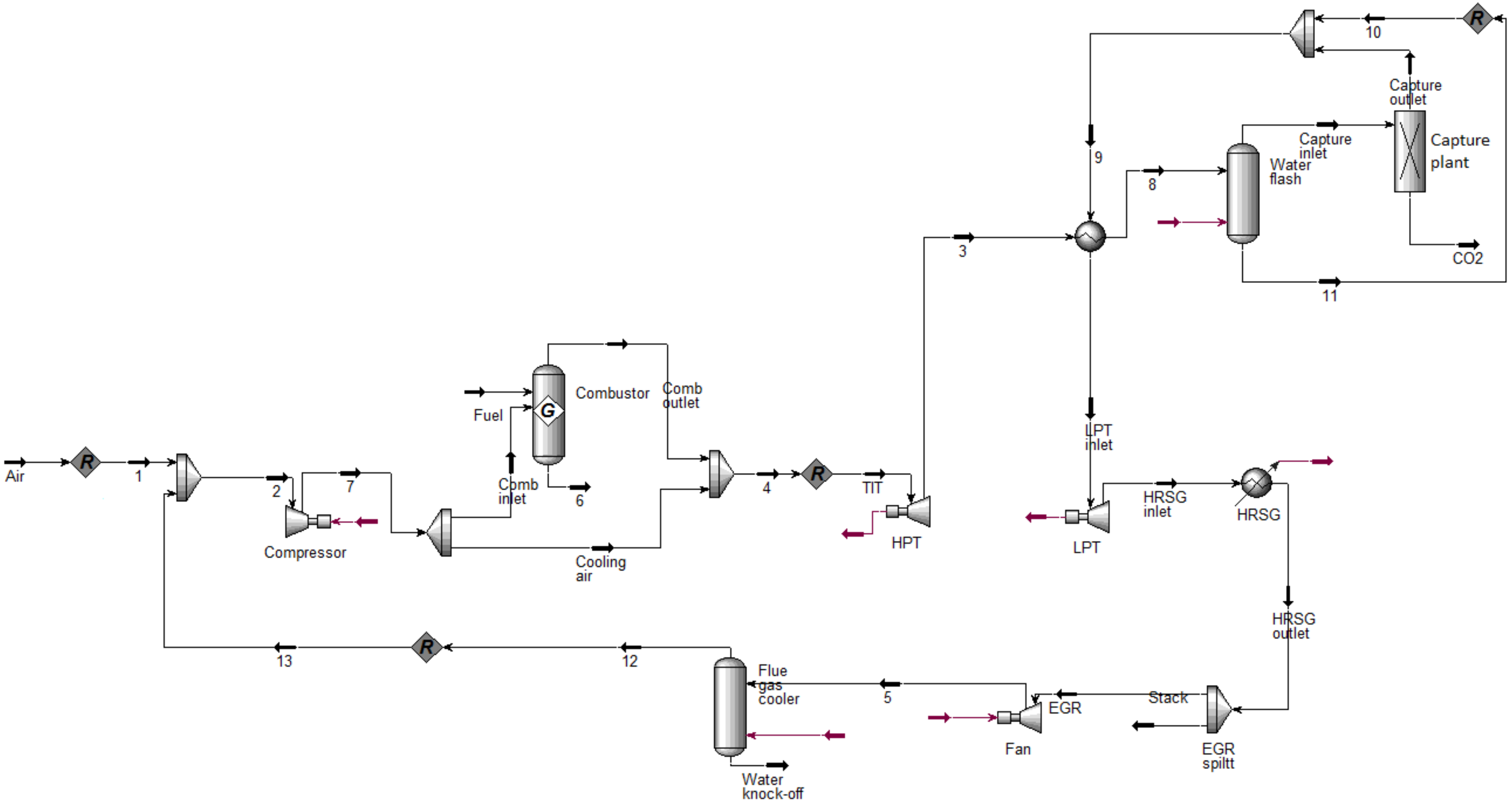


Figure 12.2 - Post-expansion CO_2 capture, Hysys schematic.

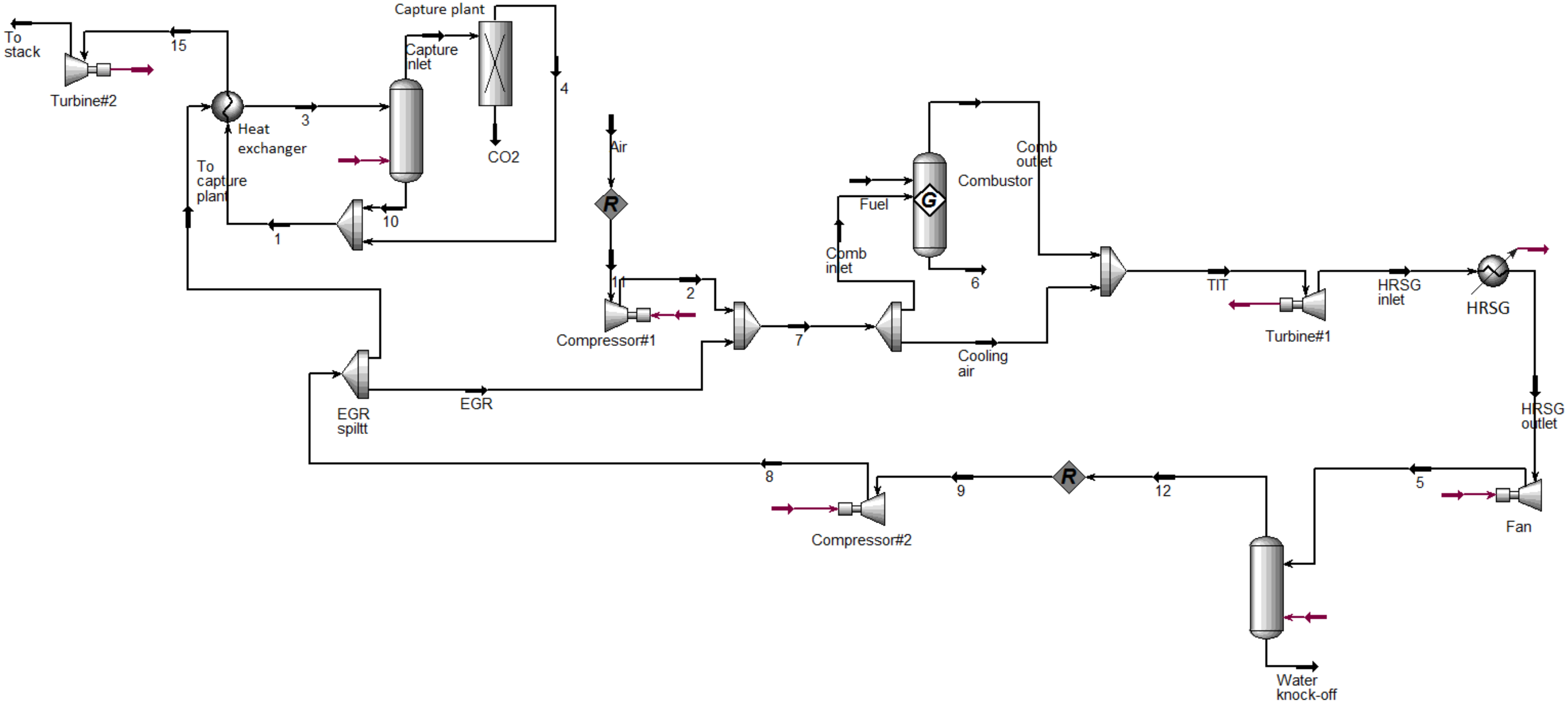


Figure 12.3 - Post-compression CO_2 capture, Hysys schematic.

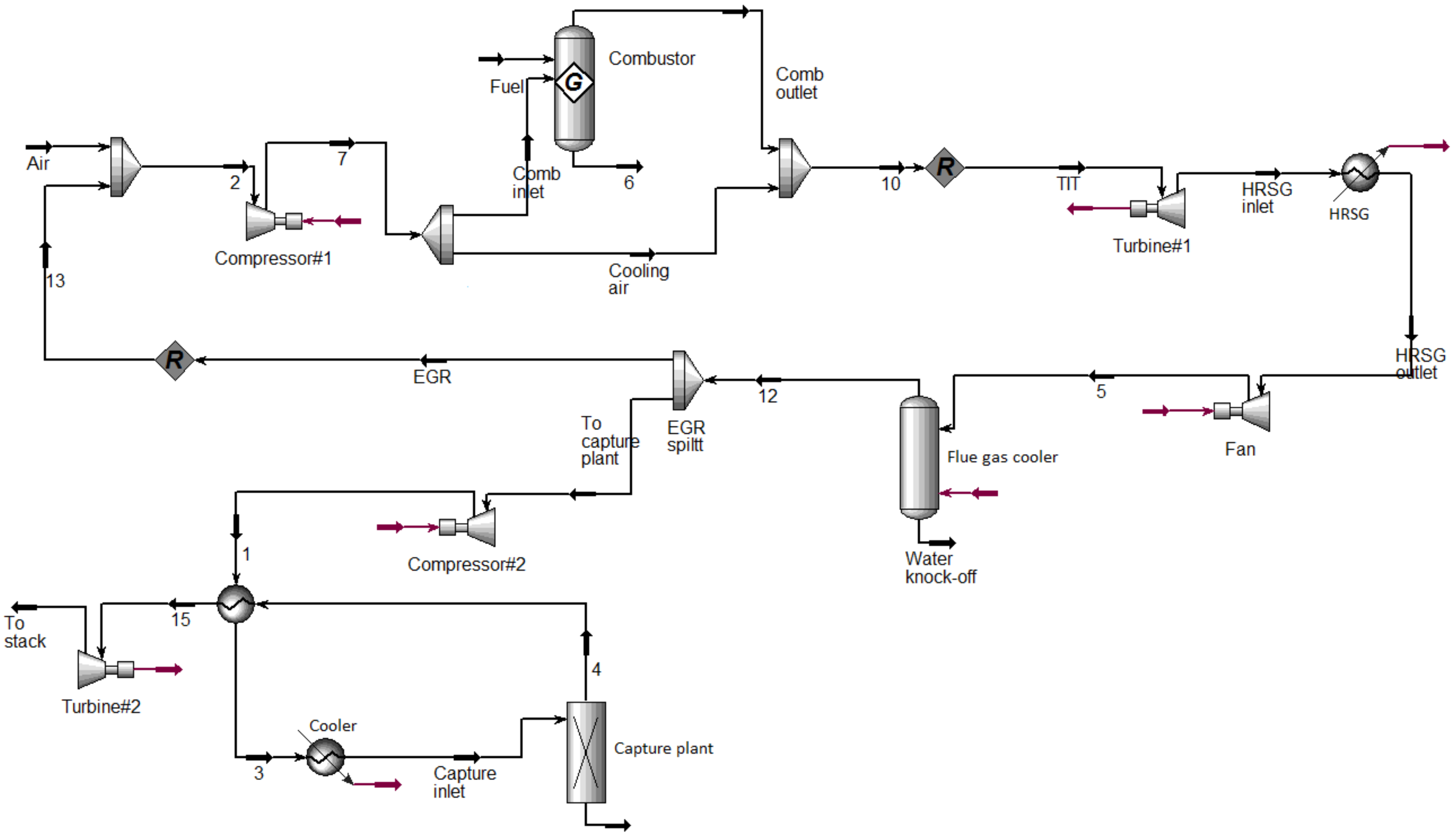


Figure 12.4 - Tail-end CO_2 capture, Hysys schematic.

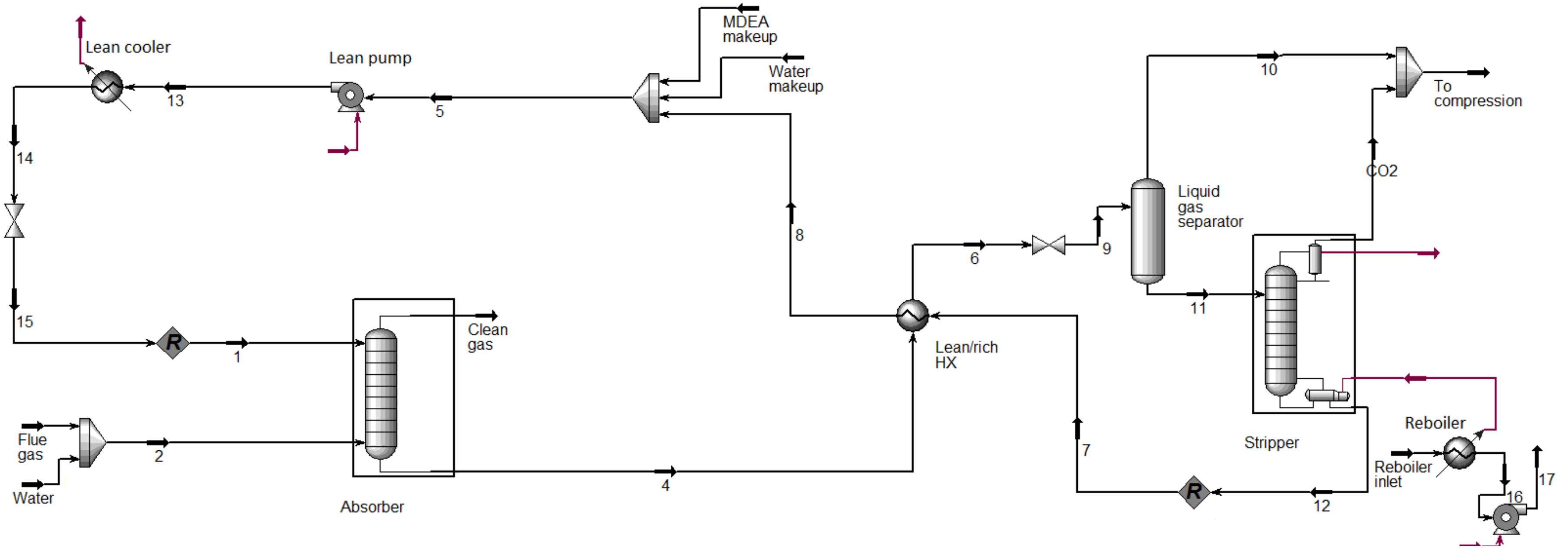


Figure 12.5 - MDEA capture process, Hysys schematic

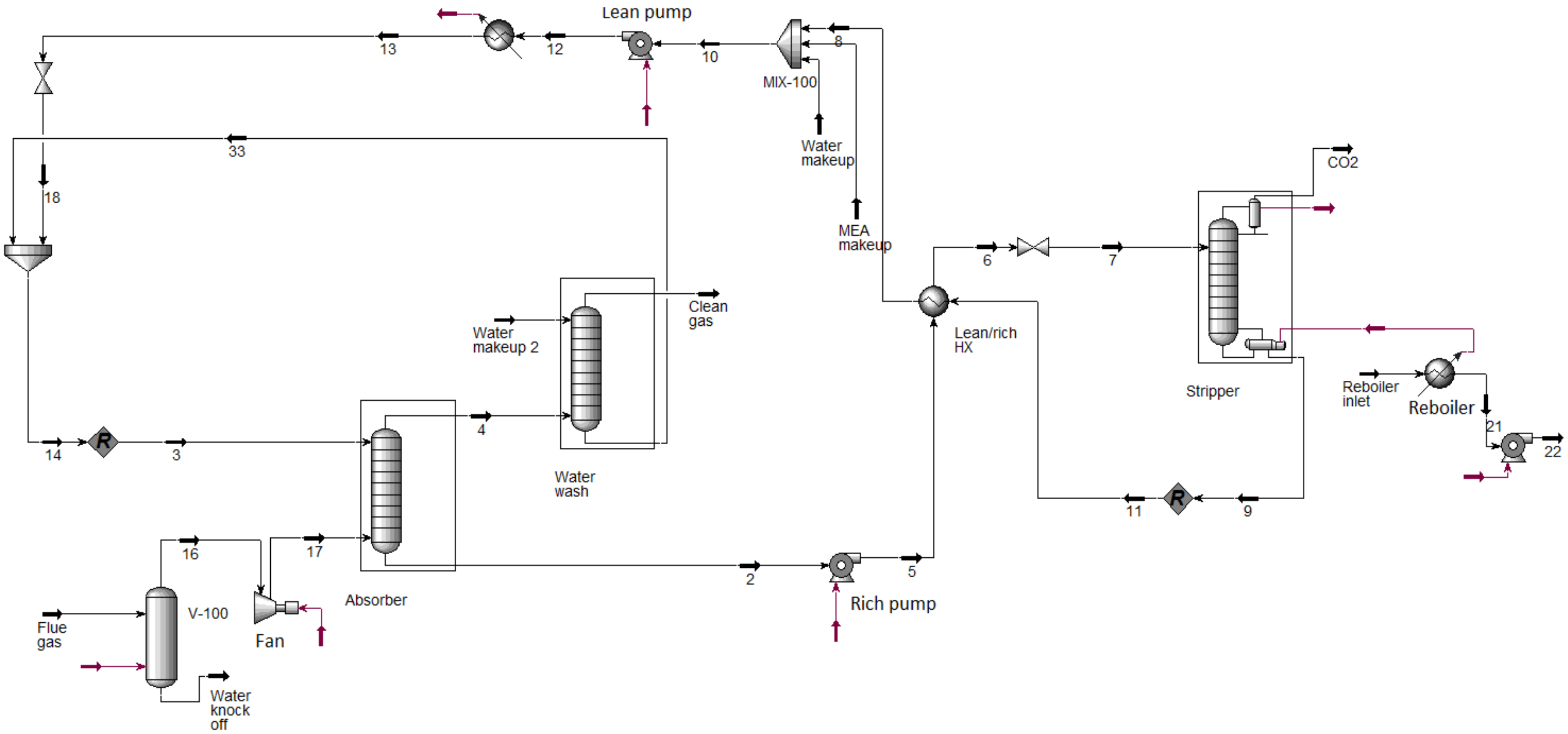
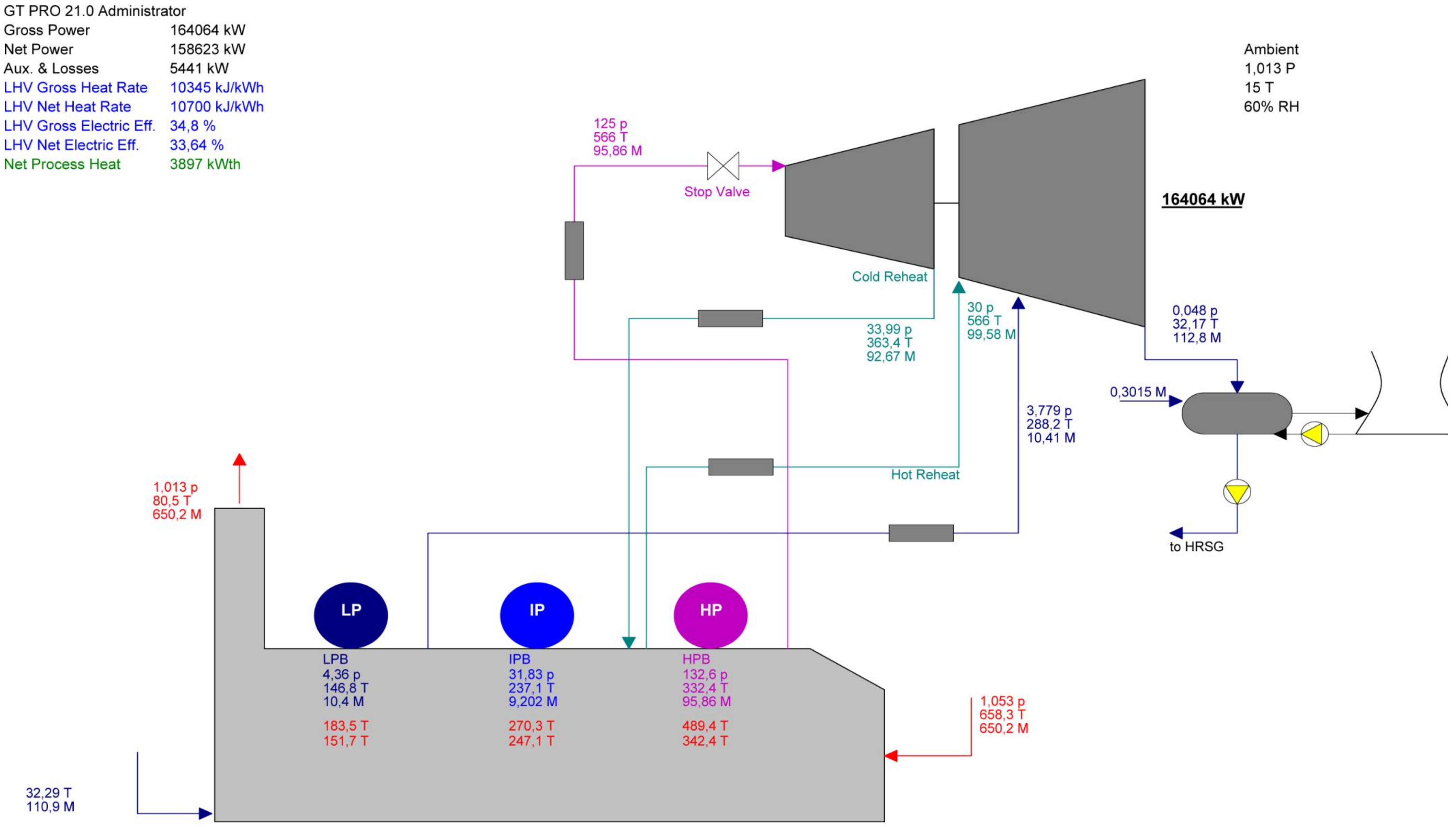


Figure 12.6 MEA capture process, Hysys schematic.

12.2 GT PRO



GT PRO 21.0 Administrator

Gross Power	164064 kW
Net Power	158623 kW
Aux. & Losses	5441 kW
LHV Gross Heat Rate	10345 kJ/kWh
LHV Net Heat Rate	10700 kJ/kWh
LHV Gross Electric Eff.	34,8 %
LHV Net Electric Eff.	33,64 %
Net Process Heat	3897 kWth

Figure 12.7 - Schematic of simplified steam cycle.

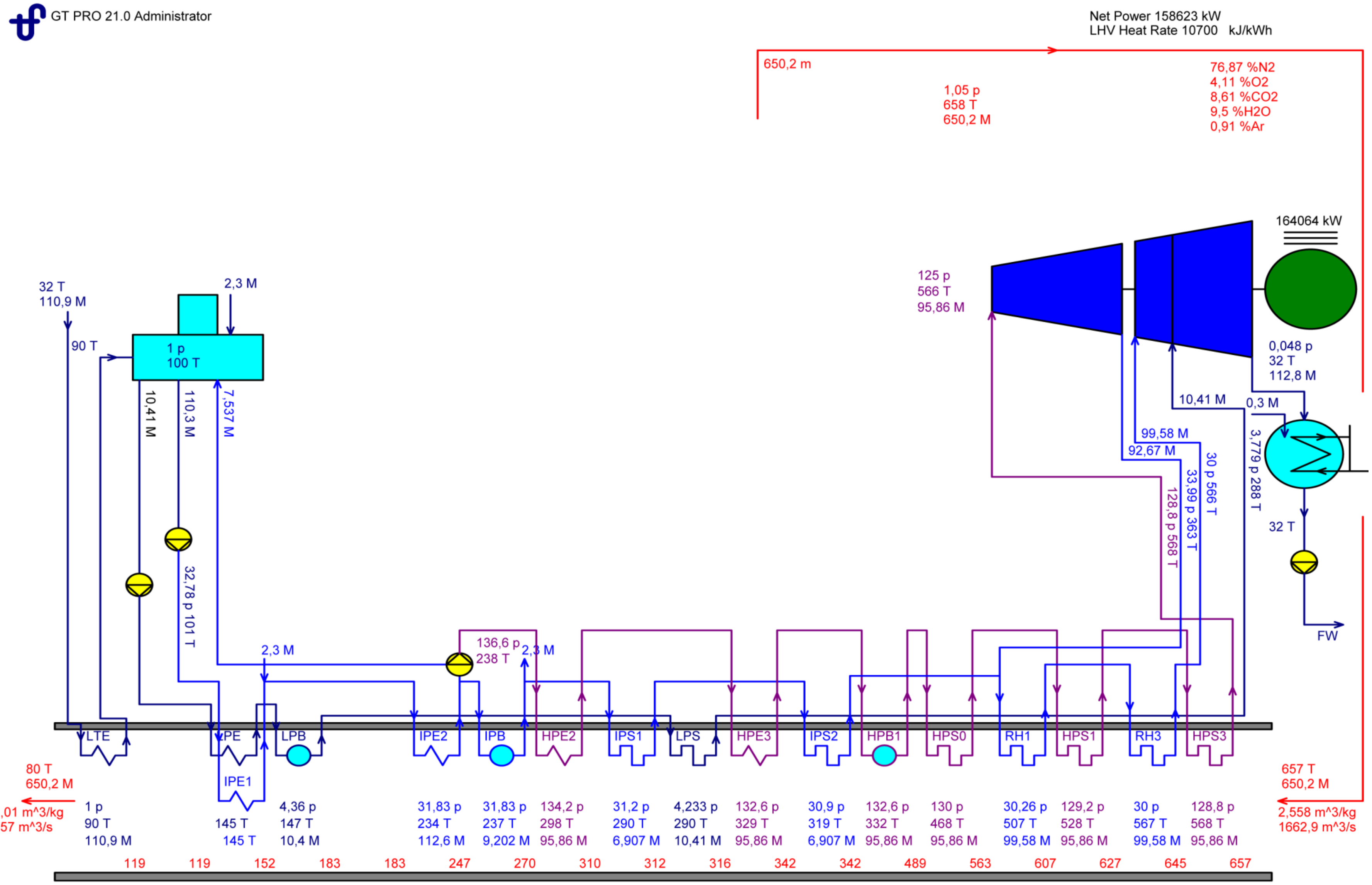


Figure 12.8 - Advanced schematic of the steam cycle.

12.3 Capture Processes

Solvent start values	**Value**	**Unit**
Unloaded composition		
MEA	29,85	wt. %
H2O	65,47	wt. %
Lean loading (mol CO_2/mol MEA)	0,205	
Lean solvent temperature out of desorber	121,5	°C
Lean solvent temperature in to absorber	40	°C
CO_2 removal	90	%
Rich loading (mol CO_2/mol MEA)	0,4649	
Rich loading temperature	38,41	°C
Flow rate	733,8	kg/s
Absorber		
Diameter	10	m
Pressure drop	10	mbar
Number of stages	11	
Lean/rich heat exchanger		
Pressure drop	10	kPa
Temperature approach, hot end	10	K
Stripper		
Number of stages	25	
Diameter	8	m
Bottom pressure	2	bar
Pressure drop	20	mbar
Overhead condenser temperature	30	°C
Various		
Pressure drop over water coolers	10	kPa
Pumps polytrophic efficiency	84	%

Table 12.1 – Basic design data for the MEA simulation.

Solvent start values	**Value**	**Unit**
Unloaded composition		
MDEA	49,90	wt. %
H2O	50,07	wt. %
Lean loading (mol CO_2/mol MEA)	0,004377	
Lean solvent temperature out of desorber	116,1	°C
Lean solvent temperature in to absorber	55[35]	°C
CO_2 removal	90	%
Rich loading (mol CO_2/mol MEA)	0,5276	
Rich loading temperature	54,02	°C
Flow rate into absorber	414,6	kg/s
Absorber		
Diameter	10	M
Pressure drop	10	mbar
Number of stages	11	
Lean/rich heat exchanger		
Pressure drop	10	kPa
Temperature approach, hot end	10	K
Stripper		
Number of stages	15	
Diameter	8	m
Bottom pressure	1,31	Bar
Pressure drop	20	mbar
Overhead condenser temperature	30	°C
Various		
Pressure drop over water coolers	10	kPa
Pumps polytrophic efficiency	84	%

Table 12.2 – Basic design data for the MDEA simulation.