CHALMERS

Lignite gasification in gas turbine applications with CO$_2$ capture

Master's Thesis within the Sustainable Energy Systems programme

DANIEL BONDÉUS

MASTER'S THESIS

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Abstract

This thesis work has evaluated the TopCycle technology for power production from lignite, integrated with drying, gasification and CO_2-capture. The TopCycle concept is a high pressure, steam injected gas turbine with up to 50% steam in the flue gases. Within the frames of the thesis work the TopCycle concept has been evaluated for oxygen- and air-blown gasification, equipped with pre- and post-combustion CO_2-capture, respectively. The results have been benchmarked against a conventional IGCC concept. This thesis has also specifically studied the utilisation of low temperature heat to drive a lignite dryer and the post-combustion CO_2 capture unit. The concepts were evaluated by simulations in the heat balance software Ebsilon Professional.

The simulations have shown that when using a 100 MW TopCycle unit as a building block, the configuration with an oxygen-blown gasifier and pre-combustion capture has an efficiency of 37% while the configuration with an air blown gasifier and post-combustion capture has an efficiency of 39-49%.

The results from the air-blown configurations show a very high potential without CO_2 capture, with net electrical efficiencies up to 56% (LHV). With CO_2-capture the concepts are able to achieve net electrical efficiencies of up to 49% (LHV), with the assumption that a low pressure CO_2-capture unit stripper is able to utilise the large amounts (2.5 MJ/kg CO_2) of low temperature heat (<69°C) available in the TopCycle flue gas. If a steam dryer is utilised, instead of a low temperature dryer, the efficiency is reduced to 47 %, but the CO_2-capture unit is able to utilize higher grade flue gas (<71°C).

The TopCycle works with a high fraction of water as working media. This can be inserted as superheated, pressurised steam prior to combustion, or by humidifying the uncompressed air before entry. The latter case means that lower temperature heat can be used for evaporating water in the uncompressed air, leaving more high temperature heat for the CO_2-separation process. It does, however, mean less heat recovery to the power generation process and more compression work, leading to lower efficiencies for the power generation. Depending on the characteristics of the CO_2-capture unit, it can be required to humidify the incoming air to achieve an operational system.

There are a series of crucial technologies that need to be further developed and validated before the TopCycle concept with post-combustion CO_2-capture can be realised:

- Operation of a 70 bar air blown gasifier
- Cost effectiveness and technical feasibility of a CO_2 capture unit driven with low temperature heat and partly operated below atmospheric pressure
- An efficient hot gas filter at 500°C
- Insensitivity of the gas turbine to SO_2 compounds or hot sulphur cleanup in the syngas
- A TopCycle gas turbine able to operate with a syngas of low heating value (≈3.8 MJ/kg)

Despite many technical uncertainties the conclusion is that the TopCycle unit should be able to reach extraordinary efficiencies together with CO_2-capture, showing promise as an important future power production technology. Further work should be focused on optimizing the post-combustion CO_2-capture unit to match the heat characteristics of the TopCycle unit, utilizing low temperature heat from the flue gases.

Contents

Nomenclature

AGR – Acid Gas Removal

ASU – Air Separation Unit

CC – Combined Cycle

CCGT - Combined Cycle Gas Turbine

CCS – Carbon Capture and Storage

HRSG – Heat Recovery Steam Generator

HTW – High Temperature Winkler

IGCC – Integrated Gasification Combined Cycle

IGTC – Integrated Gasification TopCycle

MHI – Mitsubishi Heavy Industries

1. Introduction

In early 2007 the leaders of the European Union proposed an energy policy that would aim to slow climate change while simultaneously strengthening the EU's energy security and strengthening its competitiveness.

The EU Heads of State and Government set a series of targets to be met by the year 2020. These are:

- A reduction in EU greenhouse gas emissions of at least 20% below the levels of 1990
- 20% of the EU's energy consumption to come from renewable resources
- A 20% reduction in primary energy use compared with projected levels, to be achieved by improving energy efficiency

These 3 statements are in the public debate often collectively known as the 20-20-20 targets. (European Commission, 2009)

All targets are directly applicable to the power producing sector. A large portion of the reduction in greenhouse gas emissions are theoretically possible to achieve in the power sector since the emissions from fossil power plants are large and each power plant is point source of emissions. The total share of fossil fuels (coal, oil and gas) used for power generation in EU27 is 55.3%. This corresponds to a GHG emission share of 36% within the EU27. (European Commission, 2010)

Almost half of the electricity in Germany is produced by hard coal and lignite. Germany is self sufficient in terms of lignite while about two thirds of the hard coal is imported. Lignite is an indispensable source for Germany because it is abundantly available for long-term use and employs tens of thousands of people. Apart from being a secure source of supply the German lignite has a steady extraction cost providing high economic safety. Lignite is therefore a cost efficient fuel to use in conjunction with carbon capture and storage (CCS). (Bundesministerium für Wirtschaft und Technologie, 2010)

Vattenfall is the fifth largest producer of electricity in Europe and the largest producer of heat. The major portion of the electricity produced by Vattenfall is produced in Germany and Poland (Vattenfall AB, 2010). A newly published press release states that Vattenfall is the 3rd largest company in Europe in regards of CO_2 emissions (Carbon Market Data Ltd., 2010), strongly connected with the coal power plants operated by Vattenfall in mainly Germany (Vattenfall AB, 2010). To enable the continued use of fossil fuels, new technologies are needed to decrease emissions and increase plant efficiencies. At Vattenfall Research & Development novel technologies are continuously evaluated to achieve Vattenfall's internal goal of a being carbon emission neutral by the year 2050 (Vattenfall AB, 2010).

A new novel gas turbine concept known as TopCycle has been introduced in recent years. Vattenfall, in conjunction with the technology owner Euroturbine AB, has been researching and evaluating the potential of this new technology for a number of different applications since 2001. The TopCycle is essentially a steam injected gas turbine operated with maximum steam injection and at high pressure ratios. To achieve this, an existing gas turbine is modified and complemented with a topping gas turbine, a so-called TopSpool unit. A high dew point in the flue gas is achieved which means that there is heat available to operate additional processes such as: a fuel dryer in a gasification

configuration, an amine-based CO_2-capture process or to operate the plant in combined heat and power (CHP) configuration, supplying useful heat to external costumers. (Bartlett M. , Vattenfall Applications with the TopCycle concept, 2010)

A simplified schematic of the basic TopCycle configuration is shown in Figure 1.
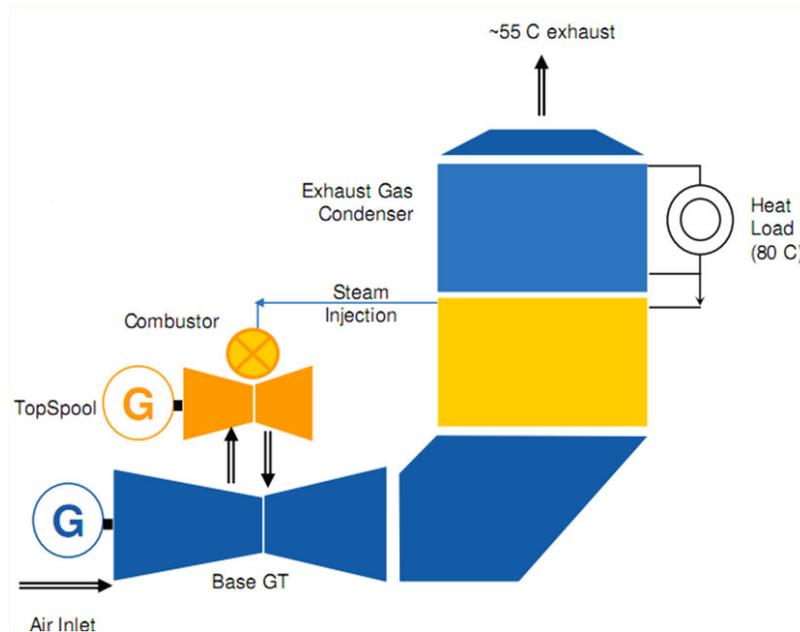
Figure 1. Overview of the generalized TopCycle unit (Bartlett M. , Vattenfall Applications with the TopCycle concept, 2010)

There are numerous expected advantages with the TopCycle concept compared to a classical combined cycle (CC) concept, which incorporates both a gas and a steam turbine:

- Similar or better electrical efficiency when integrated with gasification processes
- Lower specific investment – due to the absence of a steam turbine
- Heat extraction without penalising the power cycle

1.1. Thesis formulation and objectives

The thesis focuses on simulating the TopCycle concept together with lignite gasification and CO_2-capture. The combination of TopCycle and gasification has previously been named IGTC (Integrated Gasification TopCycle) within Vattenfall. The concept has proved promising in previous studies and the integration with CO_2 separation and drying is believed to give unique synergy effects, resulting in a competitive future power solution.

The thesis is outlined to study the TopCycle concept and compare it to a plant of the size corresponding to the reference case of an F-class gas turbine lignite pre-combustion IGCC concept with CO_2-capture.

The detailed objectives of the thesis are to:

- Evaluate the process layout and performance of different IGTC concepts together with integrated lignite drying and CO_2 capture. The main IGTC concepts will be with either:
  - Pre-combustion capture (oxygen-blown gasifier) or
  - Post-combustion capture (air-blown gasifier)

- Relate the performance of the IGTC concepts to competing technologies.

- In more detail, the parameters of most interest for the technical evaluation are:
  – Net electrical efficiency (first law efficiency)
  – Emission levels like CO_2 and SO_2
  – CO_2 capture unit global characteristics and the integration with the power cycle
  – Dryer heat requirements and integration with the power cycle
  – Ability of the TopCycle to provide low temperature heat to both the lignite drying and CO_2 capture process
  – Gasifier performance (oxygen consumption, inert gas requirements for fuel feeding, outlet syngas composition, etc.)
  – TopSpool performance and main assumptions
  – Highlight key sensitivities to assumptions

In order to pinpoint the limitations of the Topcycle unit, in post-combustion CO_2-capture configuration, a generalized study of the heat available for a generic post-combustion CO_2-capture process is performed. This is done after the initial process specific simulations which simulate a specific CO_2-capture unit with predefined heat and temperature requirements.

1.2. System boundaries

The system boundaries for the thesis are illustrated in Figure 2 with the red dashed rectangle. The input to the simulated system will be moist raw lignite and ambient air (and water when the system is not self sufficient with water). Inside the system the lignite will be dried, gasified and used to generate power. The CO_2 will be removed through pre- or post-combustion capture (see chapter 2) and the outgoing result from the system will be electricity, flue gases, compressed CO_2, a heat sink duty and waste water.
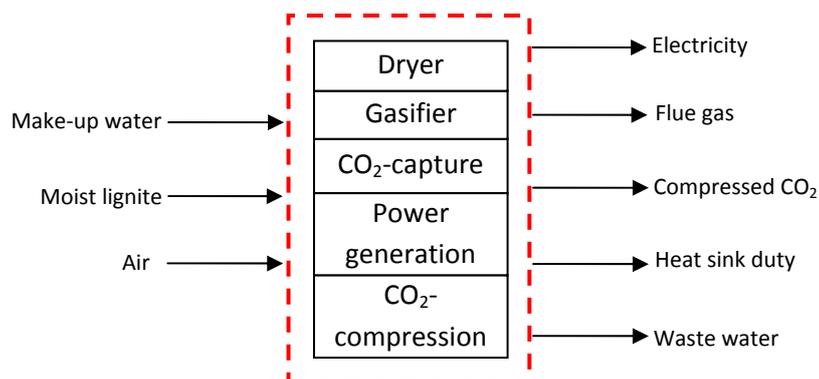
Figure 2. System boundaries for the simulation preformed. Note that the content within the system boundary (dashed red rectangle) is merely indicative

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1.3. Methodology

The heat and mass balance software Ebsilon Professional 9.0 is used to perform all simulations. Some process components are simulated as black box models with global characteristics to be able to be included in the simulations, due to the fact that the simulation software, Ebsilon Professional 9.0, is not designed for these kinds of simulations. The process components which are simplified are mainly chemical processes stages like the gasifier, CO_2-capture process (AGR stages in oxygen blown pre-combustion case and the post-combustion CO_2-capture unit in the air blown cases) and shift reactor.

An extensive literature study has provided the basic knowledge about the process steps involved in the overall system. A more thorough investigation involving both available literature and communication with co-workers within Vattenfall Research and Development concerning each process step and component has, after initial evaluation, lead to a system concept selection for each of the pre- and post-combustion configurations evaluated.

After the specific cases are simulated and evaluated an approach to generalize the post-combustion process is preformed to highlight the impact of the characteristic of the post-combustion CO_2-capture unit to the overall plant efficiency.

2. Case studies and concept selection

The TopCycle concept is simulated together with CO_2-capture, lignite gasification and drying. The principles of CO_2-capture options to be included in the heat and mass balance simulations are twofold; a pre-combustion CO_2-capture configuration, and a post-combustion CO_2-capture configuration. For benchmarking purposes a third case is also considered, an integrated gasification combined cycle (IGCC) with integrated drying and pre-combustion CO_2-capture.

The three different cases, pre-combustion TopCycle, pre-combustion reference IGCC and post-combustion TopCycle, are split into four simulated plant options. This is because the post-combustion CO_2-capture case with air blown gasification is simulated with two different gasifiers, a high temperature Winkler gasifier (HTW) and a Mitsubishi Heavy Industries (MHI) gasifier. For a thorough investigation of the impact of the post-combustion CO_2-capture unit, three different CO_2-capture units are simulated together with the HTW gasifier.

In summary the main simulated cases are:

- **Reference IGCC:** an oxygen blown Integrated Gasification Combined Cycle (IGCC) with pre-combustion CO_2-capture and integrated lignite pre-drying based on the WTA process
- **IGTC – Oxygen blown:** TopCycle process with an oxygen blown gasifier with pre-combustion CO_2-capture and integrated lignite pre-drying based on the WTA process
- **IGTC – Air blown HTW:** TopCycle process with an air blown HTW gasifier with a low temperature lignite pre-drier and three different post-combustion CO_2-capture alternatives:
    - *MEA*
    - *Atmospheric ECO_2*
    - *A combination of sub-atmospheric and atmospheric ECO_2*
- **IGTC – Air blown MHI:** TopCycle process with an air blown MHI gasifier and a low temperature lignite pre-drier

A summary of the technology selections used for each simulated case is shown in Table 1 and illustrated in the form of block diagrams in Figure 3 to Figure 6. The selections are treated in more detail in the following subchapters. For more detailed information regarding each process the reader is referred to chapter 4, Theory: Process Units. To get basic knowledge about the theory behind each process the readers is referred to chapter 3, Theory: Basics.

Table 1. Technology summary of the simulated cases

Process	IGCC	IGTC oxygen	IGTC HTW	IGTC MHI
CO_2-capture configuration	Pre-combustion CO_2-capture	Pre-combustion CO_2-capture	Post-combustion CO_2-capture	No CO_2-capture
Dryer	Fluidised bed (WTA)	Fluidised bed (WTA)	Low temperature	Low temperature
Gasifier	Oxygen blown Siemens	Oxygen blown Siemens	Air blown HTW	Oxygen enriched air blown MHI
Sulphur removal	Physical absorbent (Selexol)	Physical absorbent (Selexol)	SO_x-scrubber	SO_x-scrubber
CO_2-capture unit	Physical absorbent (Selexol)	Physical absorbent (Selexol)	Chemical absorbents (MEA & ECO_2)	No CO_2-capture
Power island	Combined Cycle	TopCycle	TopCycle	TopCycle
Other processes	Stand-alone Air separation unit, Sour shift reactor	Stand-alone Air separation unit, Sour shift reactor	Hot gas cleanup	Hot gas cleanup, Stand-alone Air separation unit

A simplified layout of the reference IGCC plant is shown in Figure 3. The configuration resembles existing IGCC plants. The main difference is the existence of a pre-combustion CO_2-capture unit, a sour CO-shift unit and a gas turbine able to handle firing almost pure hydrogen. (Hannemann, Schiffers, Karg, & Kanaar, 2002)
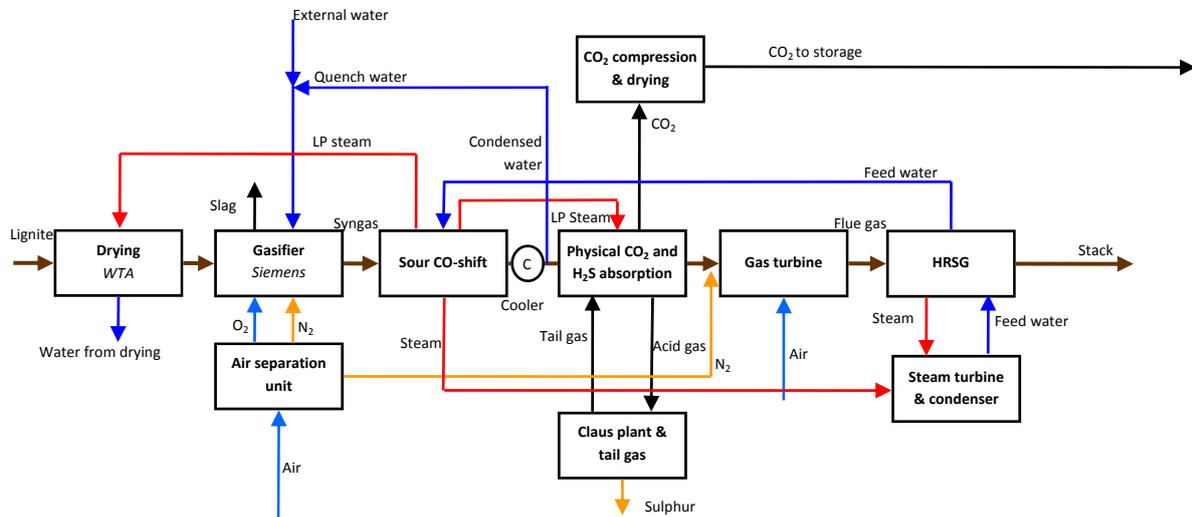
Figure 3. Simplified process flow diagram of the reference IGCC case

In the oxygen blown IGTC case (IGTC Oxygen), compared to the IGCC reference case, the steam turbine is eliminated and no nitrogen is supplied from the air separation unit (ASU) to the gas turbine (or TopSpool) combustor. This is due to the fact that steam is used for cooling the combustion chamber instead of nitrogen. In the oxygen blown IGTC case the exothermic sour shift unit produces enough steam to operate the WTA fluidized bed dryer, the Selexol process and to supply some high pressure steam for the TopSpool turbine. Figure 4 shows the simplified process layout.
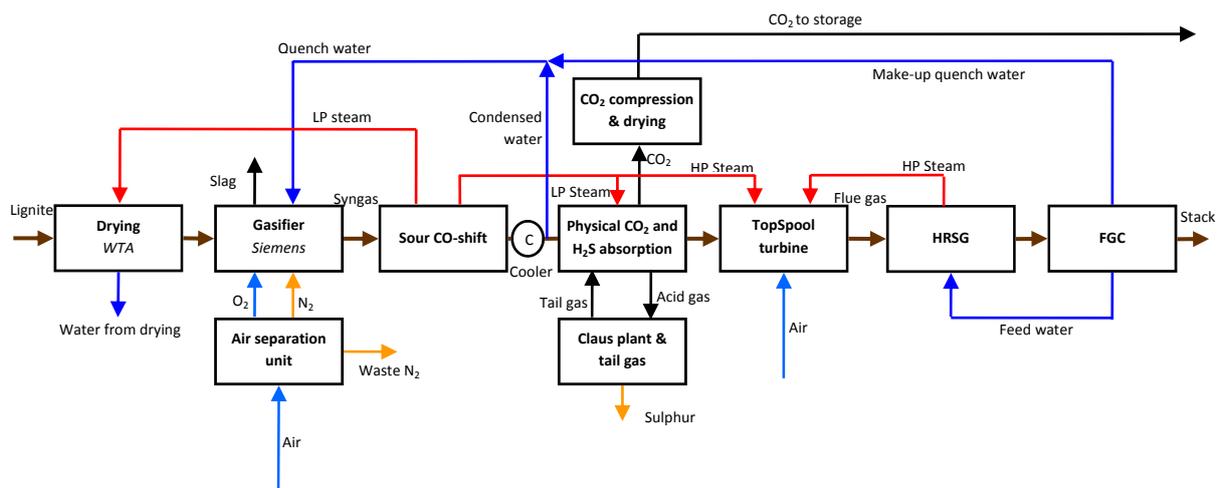
Figure 4. Simplified process flow diagram of the IGTC oxygen blown concept

In the air blown, post-combustion CO_2-capture options, a hot gas filter captures particles from the gasification process. This particle filter is designed for a maximum temperature of 500°C. To cool the

hot raw syngas leaving the gasifier, water will be injected up-streams of the hot gas filter in order to reach sufficiently low temperatures, as shown in Figure 5 and Figure 6.

The condenser in the air blown concept is designed for a very low pinch temperature of 2 K - to enable extensive use of the condensing water in the flue gas. The overall air blown concept is strongly dependant on the effectiveness of the flue gas condenser to be an overall efficient solution together with both drying and CO_2-capture.

The main differences between the air blown options lie in the gasifiers, their operating temperatures, feeding systems and carbon conversions. The HTW gasifier has lower carbon conversion but uses a screw feeding system, minimizing inert gas dilution, but requires an ash boiler in order to reach sufficient efficiency, as shown in the simplified process layout in Figure 5. The MHI concept (shown in Figure 6) on the other hand does not need an ash boiler due to the high carbon conversion but has considerably higher inert gas dilution requirement due to the dry feeding system. The MHI is also a slagging gasifier and uses oxygen-enriched air.

Due to limited information about the MHI gasifier the IGTC air blown MHI option is only studied without CO_2-capture, to show the potential of the gasifier.
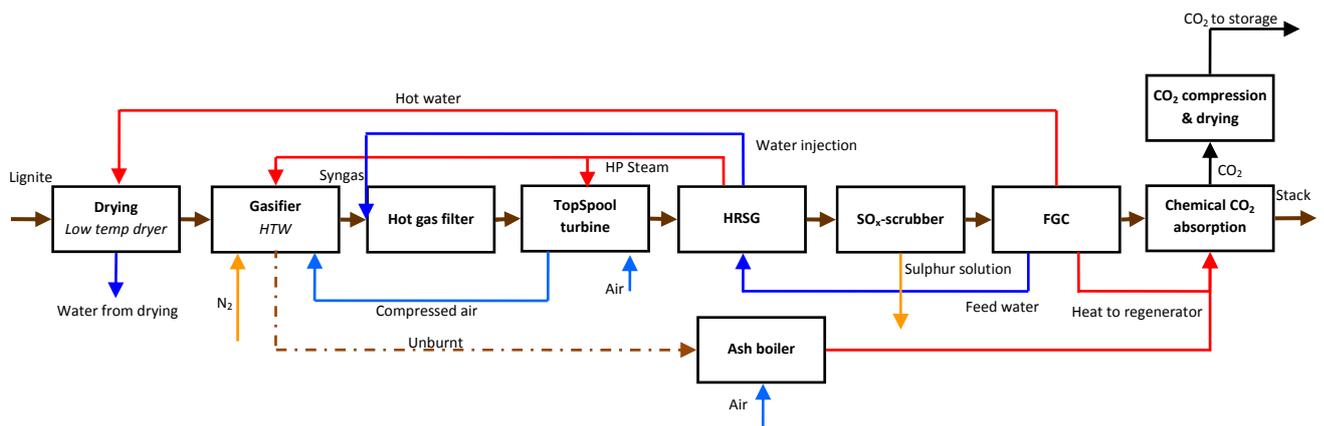
Figure 5. Simplified process flow diagram of the IGTC with air blown HTW gasifier
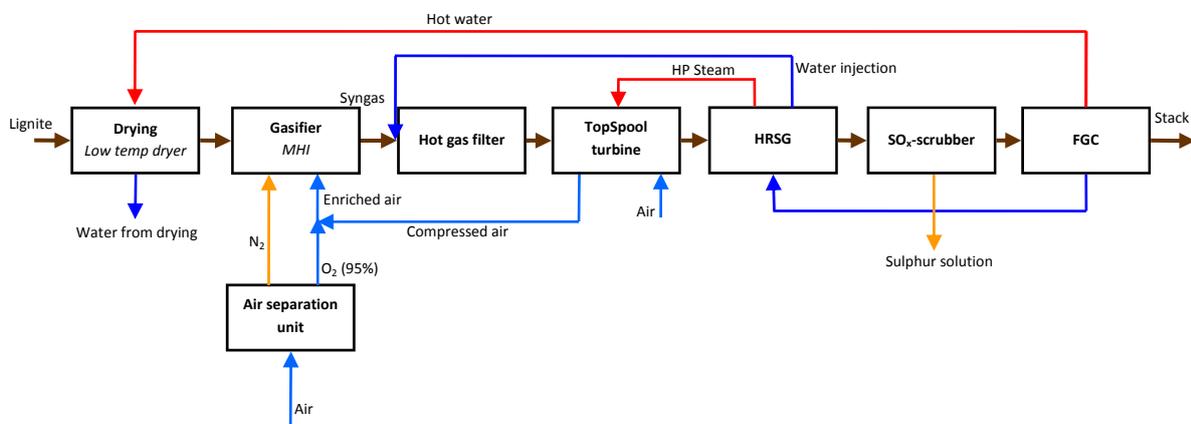
Figure 6. Simplified process flow diagram of the IGTC with oxygen enriched air blown MHI

2.1. Dryer

The moist lignite fuel is dried to increase the overall efficiency of the process. This can be done by mill drying, fluidised bed drying utilizing steam or using a low temperature source such as the latent heat from the flue gas condenser.(Allardice et al, 2004)

The steam fluidised bed drier utilizes superheated steam to dry the fuel. The steam is introduced in the bottom of the fluidising bed vessel, drying the fuel and the steam is then recycled and recompressed. The fluidizing bed drier accomplishes very high heat transfer and therefore enables a compact design(Allardice, Chaffee, Jackson, & Marshall, 2004). A semi-commercial concept of the fluidised bed technology is the WTA drier from RWE (Hoehne et al, 2006).A low temperature drier uses low temperature heat which is supplied to a working medium such as air, which blown over or through the fuel to be dried. The low heat transfer of the low temperature based drying means the installations would be large and bulky (Johansson, Larsson, & Wennberg, 2004). The Swiss Combi belt dryer is a low temperature dryer utilizing heat at around 80°C to dry the fuel (W, Kunz dryTec AG, 2010). Fluidized beds with internal heat exchangers are an alternative to reduce the unit size and air consumption.

The IGCC and IGTC oxygen-blown case can be quite easily integrated with a WTA dryer whereas the IGTC air-blown cases are severely penalized when having to extract the steam required. As base cases the IGCC reference and the IGTC oxygen use the WTA dryer, and the air blown IGTC cases are coupled with a low temperature dryer.

2.2. Gasifier

Through gasification solid fuels such as lignite can be converted into a gaseous fuel that can be used in a gas turbine process. The principle is to heat the solid fuel to drive off volatile matter and to partly oxidize the carbon in the fuel in the presence of an oxidation medium, commonly oxygen. (Bolhàr-Nordenkampf, 2004)

Details regarding different gasifiers and their operation principles can be found in chapter 3.4.4.

The commercial gasifiers relevant for gasification of lignite are:

- Siemens, oxygen-blown entrained flow gasifier
- HTW, air- or oxygen-blown fluidised bed gasifier
- BGL, air- or oxygen-blown moving bed gasifier

The Siemens gasifier with water quench is chosen as gasifier for the TopCycle oxygen-blown case (IGTC oxygen), as it is specifically developed for lignite and well-suited to the high ash fuel considered (Siemens, 2009). The entrained flow design allows high throughput and eliminates the need for parallel series of gasifiers. The high operating temperature of the oxygen blown entrained means that there is no methane in the outgoing syngas (Higman & van der Burgt, Gasification Processes, 2008) that would otherwise slip through the pre-combustion CO_2 removal unit and decrease the carbon capture rate of the plant. The design is also simple and hopefully able to be modified for higher pressures. This latter point is important as the TopCycle operates at 60 bar instead of the usual 15-25 bar for combined cycles.

The oxygen blown configuration keeps the nitrogen dilution of the syngas to a minimum and means that the pre-combustion CO_2-unit has a high effectiveness. Keeping the nitrogen dilution to a minimum is also important as the pre-combustion CO_2-capture system requires cool syngas (37°C), which is reheated before being fed to the combustion chamber. It is important to remember that the gasifier has a dry feeding system, meaning inert nitrogen is needed as a transport and feed-in gas, resulting in some dilution of the syngas. All gasifiers simulated need nitrogen for feeding fuel into the gasifier. Appendix B explains the calculation behind the specific nitrogen needs for the gasifiers, but in short it is strongly related to the operating pressure of the gasifier. The IGCC case is based on conventional technology and gasification is operated at a lower pressure than for the cases including the TopCycle unit. This means the IGCC has less nitrogen dilution of the syngas through the acid gas removal systems.

A cryogenic based air separation unit (ASU) is used for supplying the oxygen to the oxygen blown Siemens gasifier, as this is the only commercially available technology (Allam, Castle-Smith, Smith, Sorensen, & Stein, 2000). Due to expected operational difficulties, the air separation unit is not integrated with the gas turbine compressor, despite the potential to gain a higher efficiency from a higher isentropic efficiency of the gas turbine compressor compared to the smaller ASU-compressor. The reason is purely based on lowering the plant integration and complexity to hopefully make plant start-up and load change less problematic.

The reference IGCC plant utilizes the same entrained flow gasifier with quench, mainly for purpose of comparison. An ideal IGCC plant would perhaps rather use a syngas cooled method to take advantage of the sensible heat in the syngas after gasification. However, syngas coolers are generally designed for hard coal and not lignite, and usually need much maintenance.

The main air blown IGTC options are based on a fluidised bed gasifier, the High Temperature Winkler gasifier (HTW). This gasifier is capable of using air as the oxidizing medium and it has been proven on a commercial scale. Additionally, know-how in fluidised bed gasifiers exists within Vattenfall. Air blown gasification does not suffer the efficiency penalty of having an ASU and the nitrogen demand in the air blown configuration is low as the gasifier has a screw feeding system, reducing inert gas consumption. Due to the low carbon conversion achieved by the HTW gasifier, the combustibles recovered from particle filters and dry ash will be fed to a waste boiler to provide extra heat needed for auxiliary systems such as dryer and CO_2-capture unit. This will reduce the losses imposed by the lower carbon conversion of the gasifier. The carbon capture rating of the entire plant will be maintained as the flue gases from the waste boiler will be fed to the main flue gas stream prior to CO_2-capture.

The additional air blown case using the entrained flow MHI gasifier is simulated to show the potential of a concept with higher carbon conversion and therefore excluding the need for an ash boiler to reach sufficient fuel utilisation.

2.3. Shift reactor

In a pre-combustion CO_2-capture configuration the syngas after oxygen blown gasification mainly consists of water, hydrogen, carbon dioxide and carbon monoxide. In order to successfully capture the carbon dioxide in the pre-combustion CO_2-capture unit the carbon present as carbon monoxide must be converted into carbon dioxide, this is done in a shift reactor. The purpose of the reactor is to

enhance the reaction speed of the reaction shown in Equation 1, promoting the conversion from carbon monoxide to carbon dioxide.

Equation 1. The homogeneous water-gas shift reaction

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

The reaction is exothermic and there is therefore an opportunity to raise steam from the heated syngas.

Since the lignite contains considerable amounts of sulphur, using a clean shift reactor would impose a risk. Moreover the additional cooling and heating of the syngas which would be required when using a clean shift reactor would increase plant complexity and mean considerable water condensation after gasification and water quench. This water would have to be added after H_2S removal to achieve an appropriate H_2O/CO ratio in the clean shift reactor. The sour shift reactor is chosen for all applicable concepts since the Siemens water quench gasifier suits the sour shift perfectly with the moist outgoing syngas. Further, the reaction is split into 2 separate shift reactors in the simulations, working at different temperatures, to achieve a high degree of total conversion.

2.4. Acid gas removal

After the shift reactor, in the pre-combustion CO_2-capture configuration, the carbon dioxide and sulphur compounds are to be removed. The CO_2 is removed to achieve specified carbon capture rate and the sulphur compounds are required to be removed in order to not risk corrosion in the turbine.

Rectisol and Selexol are two of the physical solvents offered commercially. The Rectisol process is a methanol based solvent which requires refrigeration of the flue gas while Selexol is able to operate at a higher flue gas temperature, hence not requiring any syngas refrigeration. (Higman, Gasification, 2008)

In the reference IGCC and the IGTC oxygen-blown cases, the H_2S and CO_2 will be removed through a Selexol process taking advantage of the high partial pressure and requiring no refrigeration of the syngas.

2.5. Hot syngas cleanup

The air-blown post-combustion concepts will feature hot syngas clean up. It is desirable to feed high temperature syngas from the gasification process directly into the TopCycle unit. This is thermodynamically favourable as gasifier heat is utilised directly in the gas turbine rather than recovered or rejected to a cooling sink. To enable this, a hot gas cleaning system is required. The options available for hot gas cleaning are limited and cannot yet be considered fully commercial.

To remove particles, metallic and ceramic filters can be used and are able to operate at temperatures of about 500°C. These have challenges with lifetime due to physical stresses or indirect failures due to corrosion related issues. In the process of cooling the syngas to 500°C, both heavier and lighter metal compounds are condensed onto fly ash. Therefore, the hot gas filters act as both particle filters and metallic compound separators, preventing particles from reaching the combustion chamber and turbine and hot alkali sulphate attack on the turbine blades.

There is no commercial process to remove sulphur from the syngas at high temperature. In this thesis, no sulphur removal is assumed in the syngas for the air-blown cases. The sulphur hence

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passes through the gas turbine and HRSG as SO_x. This may cause sulphidation problems with the blades and requires acid-resistant material in the back end of the HRSG.

2.6. Flue gas clean-up

In a post-combustion CO_2-capture configuration the CO_2 is removed after the TopCycle unit, from the flue gas. The flue gases are at a close to atmospheric pressure and does therefore not benefit from an especially high carbon dioxide partial pressure. Chemical absorption unit to separate the carbon dioxide is therefore a more natural choice.

Chemical absorbents are among others offered from BASF (Badische Anilin- und Sodafabrik), Flour and Mitsubishi Heavy Industries. The most common solvents are MEA, DEA and MDEA and MEA has been used commercially in the natural gas industry for over 60 years. MEA is normally regenerated in a stripper column with steam at a temperature of 100-200°C. Steam is generated in a reboiler at the column base, in which external steam is used to boil the lean solvent from the exit of the column.

A chemical absorption process called ECO_2 from Powerspan Corp. is an ammonia based solution designed for post-combustion capture for traditional coal-fired plants. The ECO_2 process is indicated to use less energy than traditional amines and does therefore show promise for CO_2-capture applications. The stripper of the ECO_2 process can also operate with lower temperature heat, as the boiling point of the solvent is lower than for amines.

For the HTW air blown options with post-combustion CO_2-capture, the CO_2-capture will be a chemical post-combustion solution. Both a MEA and ECO_2 based chemical absorption system are simulated due to their different characteristics in terms of heat and temperature demand, resulting in different heat integration opportunities.

Sulphur removal in the flue gas is done in a limestone/lime scrubber. A water based slurry of inexpensive limestone/lime is sprayed into the flue gas flow in a scrubbing tower. The sulphur dioxide in the flue gas reacts with the aqueous solvent to from calcium sulphate, gypsum (Kohl & Nielsen, Sulphur Dioxide Removal, 1997).

To remove the sulphur compounds to an acceptable level a conventional scrubber will be installed in the flue gas stream of the air blown cases in order to remove sulphuric compounds before reaching the stack.

2.7. Power island configurations

A brief overview of the different power island configurations is presented.

2.7.1. Combined Cycle

The IGCC reference case features a combined cycle. The combined cycle consists of a gas turbine with a HRSG and steam turbine. Heat is mainly recovered from the flue gases from the gas turbine and additional heat is also recovered from the exothermic sour shift reactors.

For a more detailed description of a combined cycle the reader is referred to chapter 3.9.1.

2.7.2. TopCycle Process

The TopCycle process is based on an existing gas turbine, which is modified and complemented with a topping gas turbine, a so-called TopSpool unit. The main features of the TopCycle concept are to

utilize a large amount of steam injection, recovered from a heat recovery steam generator (HRSG) and to run the gas turbine at high pressures. A standard gas turbine is used as a bottoming cycle and the combustion chamber is replaced with a very high pressure topping gas turbine cycle (TopSpool) increasing the pressure to >50 bar. In the high pressure combustion chamber the fuel is burnt at close to stochiometric conditions.

The TopCycle unit is based for applications in the ≈100 MW_e size. This is due to the first generation design which is adapted to a commercial bottoming cycle.

The heat recovery system consists of the intercooler and the single pressure HRSG system in the flue gas train. The feed water is evaporated in the intercooler and evaporator in the HRSG. The steam is used for blade and vane cooling in the TopSpool turbine and auxiliary process, the rest of the steam is superheated in the HRSG prior to injection into the combustion chamber. After the HRSG the flue gas is cooled down in a flue gas condenser. The condensed water is recovered, treated to sufficient quality and led back to the HRSG. The TopCycle process can be self sufficient with water, or even produce water, as the water formed during combustion can be condensed.

2.7.3. TopCycle in pre-combustion CO_2-capture

The pre-combustion CO_2-capture configuration means that the gas producing process, which involves gasification, gas clean-up and CO_2-seperation can almost be considered as a separate block of process units, similarly to the IGCC reference case with combined cycle. There is some heat exchange between the TopCycle unit and the exothermic shift reactor, but to a large extent this configuration is very similar to the one shown in Figure 7.
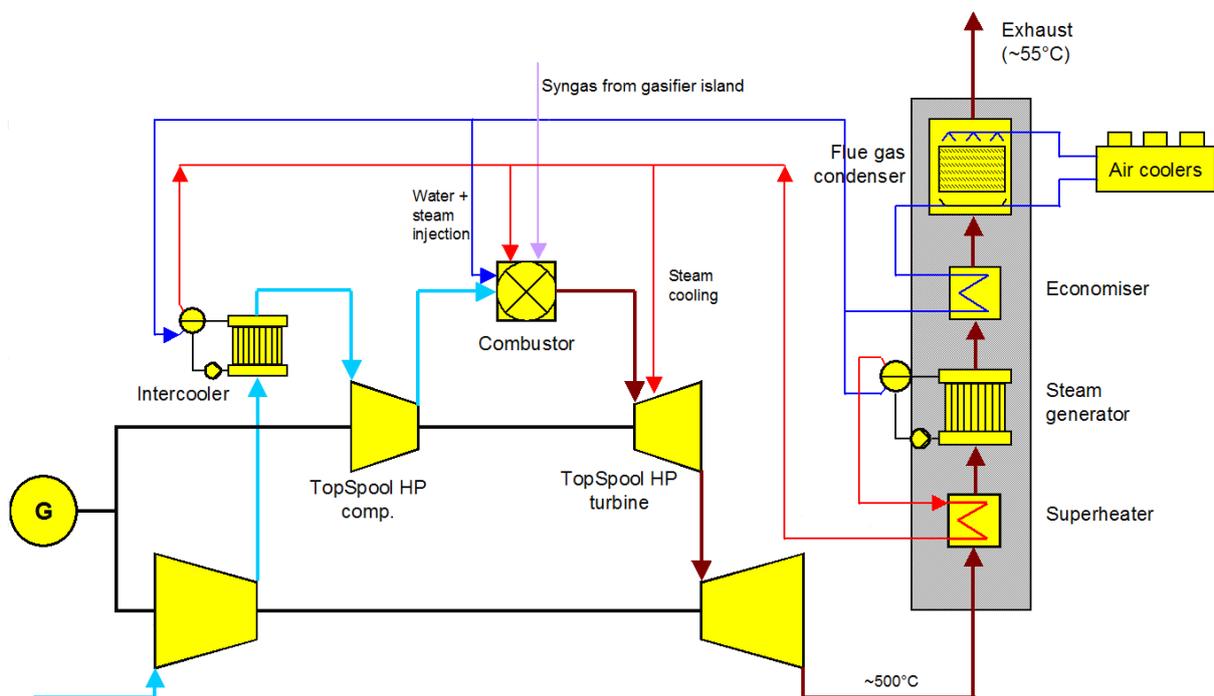
Figure 7. General TopCycle layout when firing natural gas (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

2.7.4. TopCycle in post-combustion CO_2-capture

When applying the TopCycle concept together with drying and gasification of lignite and CO_2-capture there is a large difference in working principle when comparing pre- and post-combustion CO_2-capture.

When employing post-combustion CO_2-capture with air blown gasification, there is a large need of low temperature heat to operate the post-combustion chemical CO_2-capture unit. When additional heat is needed from the TopCycle unit there is a way to increase the available heat in the flue gases, by introducing a two-stage flue gas condenser as illustrated in Figure 8. In the first stage, water is sprayed counter current to the flue gas, condensing a portion of the flue gas water content, while being heated to near the flue gas dew point. In Figure 8 the return water flow from a district heating network is used as a heat sink, this heat sink is replaced by the CO_2-capture unit in a post-combustion configuration. In the second stage, heat is extracted to drive an atmospheric humidifier for increasing the humidity of the compressor intake air. This inlet humidification uses low temperature heat in the flue gas to increase the water content in the inlet air and hence the flue gas dew point somewhat. Simplified it can be said that the gas turbine acts as a heat pump, allowing low temperature heat from the second FGC stage to become available at higher temperatures in the first stage of the flue gas condenser. (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

Figure 8. Principal layout and function of the inlet humidifier and two-stage flue gas condenser in a TopCycle process. Note that temperatures in the figure are only indicative (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

Although the total efficiency of the cycle will increase, the electrical efficiency of the cycle will be penalized by inlet humidification since the compressor work will increase. Therefore inlet humidification is only of interest in applications where low grade heat is needed for the overall process.
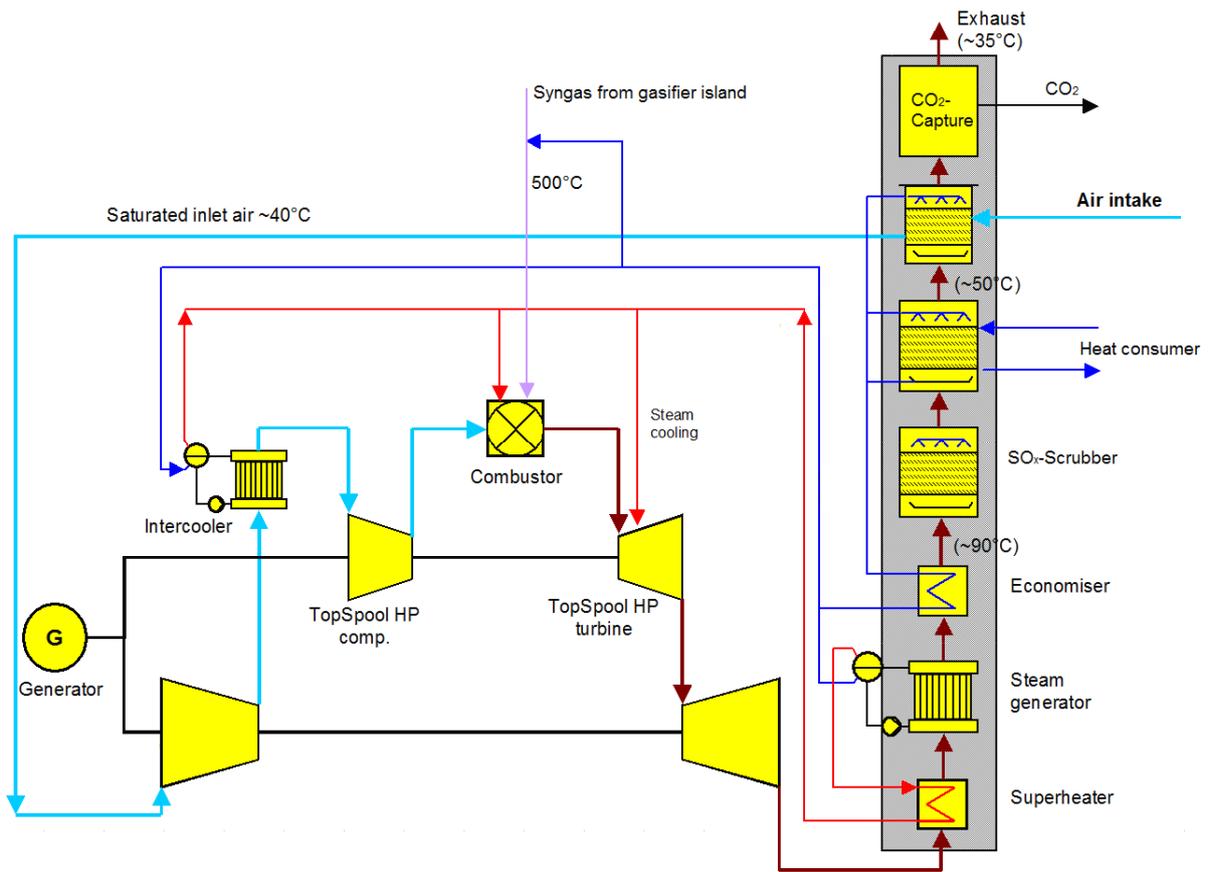
Figure 9. TopCycle with inlet humidification and CO_2-capture (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

The inlet humidification increases the amount of moist present in the air supplied to the combustor unit. This will mean that less steam and/or water injection is needed to achieve the design condition of low excess O_2 in the combustion chamber. The incoming syngas, shown in Figure 9, is sprayed with water in an air blown case, to achieve a sufficient low temperate of the syngas for hot gas clean-up before entry into the combustion chamber. This means that there is a limited amount of steam which can be supplied to the combustion chamber, in order not to go below the specified value of excess O_2 for a certain firing temperature. The result is that the more humidification applied to the system the less steam is needed for the TopCycle unit. By superheating the steam fed into the combustor the energy recovered in the HRSG can still be made available for the TopCycle unit. The superheating is however restricted by the temperature in the exiting flue gases and the required pinch temperature in the super heater.

An overview of the heat integration of flue gas condensation for the HTW case to supply heat to the CO_2-capture unit is shown in Figure 10.
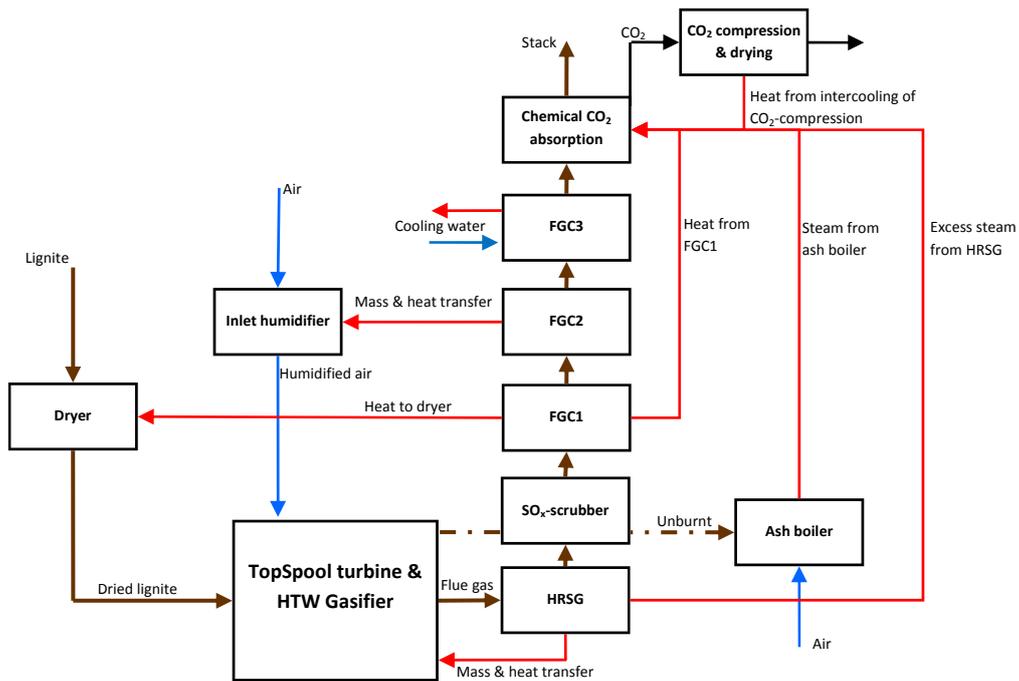
Figure 10. Overview of the heat integration principle for IGTC HTW with CO_2-capture

Heat from the first flue gas condenser (FGC1) will supply heat to the low temperature dryer. Any excess heat will be used together with heat from the ash boiler and any excess steam from the HRSG to operate the chemical CO_2-absorption unit. Additionally, some heat from the intercooled CO_2-compressor can be used. The operating principle for the chemical CO_2-absorption unit is illustrated in Figure 11. The role of the second (FGC2) and third (FGC3) flue gas condenser is to humidify the incoming air, if needed, and to bring down the flue gas to appropriate temperature before the CO_2-absorption unit with cooling water.
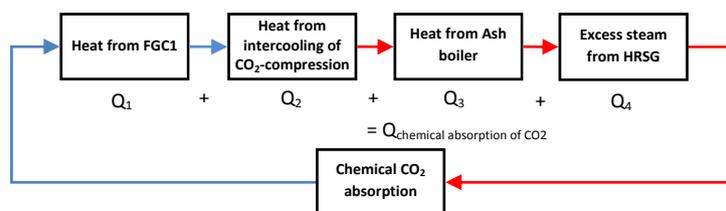
Figure 11. Heat sources for operation of the chemical CO_2-absoption unit

3. Theory: Basics

In order to appropriately choose and simulate the process units, which are presented in Theory: Process Units, it is important to understand the underlying working principles, which will be presented in this chapter.

3.1. Concepts of pre- and post-combustion CO_2 capture

The TopCycle concept is simulated together with pre-combustion and post-combustion CO_2-capture. For a short introduction the principals of pre- and post-combustion CO_2-capture with gasification and drying are shown in Figure 12 and Figure 13 respectively.
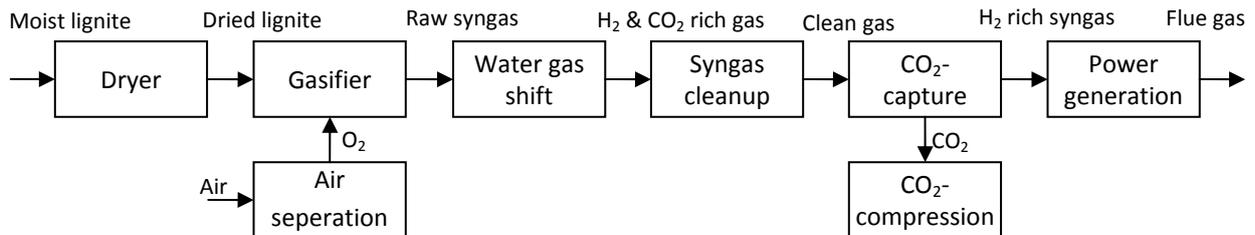
Figure 12. Basic schematic of pre-combustion CO_2-capture

In the pre-combustion CO_2-capture configuration (Figure 12) the carbon in the fuel is separated as CO_2 from the process before combustion. In the water gas shift unit any carbon monoxide (CO) in the syngas is oxidized into CO_2, which the down-stream CO_2-capture unit can capture. The water gas shift unit is therefore critical in order to reach sufficient carbon capture rates in a pre-combustion CO_2-capture configuration. The water gas shift unit, or shift conversion, will be discussed more thoroughly in the following chapter.
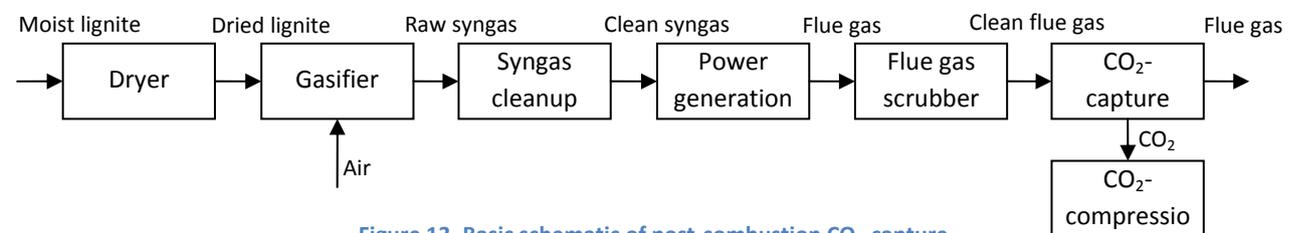
Figure 13. Basic schematic of post-combustion CO_2-capture

In the post-combustion CO_2-capture configuration (Figure 13) the CO_2 is separated as CO_2 after combustion. There is therefore no requirement on the syngas to have a certain composition to be able to separate CO_2 as in the pre-combustion case.

3.2. Fuel characteristics

The combustible substances in the solid fuel are composed of volatile matter and solid coke. Lignite contains of about 50% of volatile matter and the rest is composed of coke. As reference hard coal contains around 30% volatile matter (Energy Research Centre of the Netherlands, 2009). Lignite has, compared to hard coal, a couple of disadvantages due to its very high water and sulphur content. The water content can be as high as 60 % on mass basis while the content of sulphur can be about 3 % (Hoehne, Lechner, Schreiber, & Krautz). Corresponding values for hard coal can be found in the region of 10 % water content and 2% sulphur, on mass basis (Higman & van der Burgt, Feedstocks and Feedstock Characteristics, 2008). The sulphur can result in severe corrosion problems and the

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need for an efficient desulphurization system. The high moisture content means that the lignite should be pre-dried to avoid high efficiency losses in a normal pulverized fired power plant. This is achieved by using high temperature flue gases to dry the lignite prior to milling, resulting in efficiency losses. (Hoehne, Lechner, Schreiber, & Krautz)

3.3. Drying theory

Drying of moist lignite fuel is essential to achieve high overall plant efficiencies, and is in certain cases a requirement to be able to crush and feed the lignite into a gasifier because of the high moisture content of the fuel (Hoehne, Lechner, Schreiber, & Krautz). Previous studies have shown that pre-drying of lignite can increase the thermal efficiency by at least 4 % (Decon Deutsche Energie-Consult Ingenieurgesellschaft mbH).

In order to understand the drying characteristics of lignite it is necessary to consider how the water is bound in the lignite. Lignite is described as a capillary-porous colloid. The walls of its capillaries are elastic and swell up to its elasticity limit when taking up a liquid. These capillaries greatly influence the drying process.

The water on the surface of the lignite particles and inside the larger capillaries is bound through weak physical-chemical adhesion forces. The water film is bound by stronger physical-mechanical adsorption forces and the intermediate capillary water is bound by weak capillary forces. Within the smaller capillaries (<120 nm) the forces binding the water to the lignite particle are of the capillary and dipole nature, forming an efficient bond between particle and liquid. In the smallest capillaries, diameter of a few nm, the water is bound by the strong hydrogen bonds. (Hoehne, Lechner, Schreiber, & Krautz)

When drying lignite the energy supplied to the drying process will need to overcome the bonding between particle and water in order to remove the water from the fuel. The variation of bonding forces, hence energy needed to break the bonds, results in distinct drying phases during the drying of a lignite fuel as illustrated in Figure 14.

During the first drying phase the intermediate capillary water and the water bound by adhesion forces is drawn off at a constant speed until water content w_1. The lower drying speed of the second phase is due to the capillaries in the fuel. The water has to diffuse out of the capillaries before being evaporated. In the third drying phase the drying speed is reduced even further due to the practically strong hydrogen bonds, eventually coming to a halt at the equilibrium point, w_E, in respect to the level of supplied drying energy. (Hoehne, Lechner, Schreiber, & Krautz)
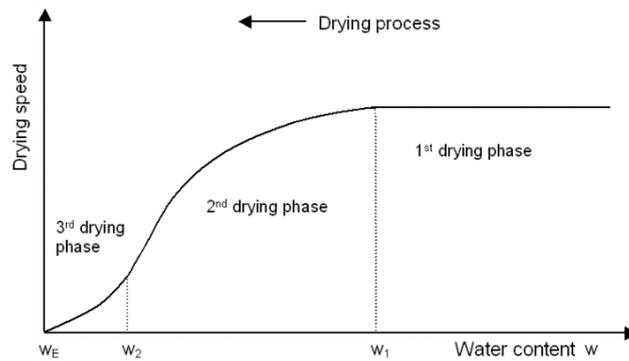
Figure 14. Different drying phases of lignite (Hoehne, Lechner, Schreiber, & Krautz)

The total amount of energy needed to dry a certain amount of lignite to a specific water content must therefore not only heat up the lignite particle, evaporate the water and compensate for heat losses in the process but also overcome the physical and chemical binding forces that bind the water to the lignite.

It has been observed that the binding forces of the lignite/water system become especially important when it is desired to reach moisture contents of less than 23 %. This is clearly illustrated in Figure 15, where the differential binding energy is plotted against moisture content for Niederlausitz lignite. (Confidential, Source 1)
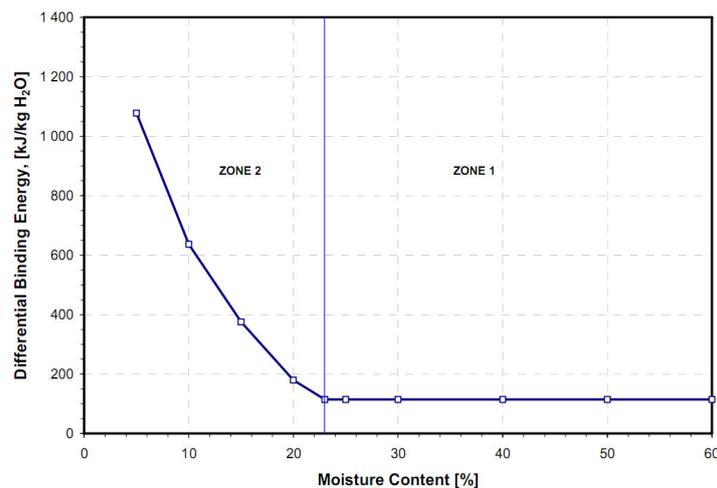
Figure 15. Binding energy against moisture content for lignite (Confidential, Source 1)

To calculate the total binding energy the differential binding energy in Figure 15 has to be integrated from the moisture content wanted to the initial water content. However this above graph is a kind of worst case scenario since the binding energy should go towards zero with increasing moisture content. (Confidential, Source 1)

3.4. Gasification theory

Gasification is a process where solid fuel is converted to a product gas (syngas) by heating. The main steps of the gasification process itself are drying, pyrolysis, gasification and combustion. The first step is - as the name implies - drying of the fuel, evaporating the moisture content of the fuel. The next step, pyrolysis, involves the release of the volatile components in a reducing atmosphere producing pyrolysis gas. The remaining coke is thereafter gasified together with the produced gas. That is, oxidised in the presence of a gasification agent and finally again reduced, to

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produce the raw product gas (Bolhàr-Nordenkampf, 2004). The general path of reactions is shown in Figure 16.
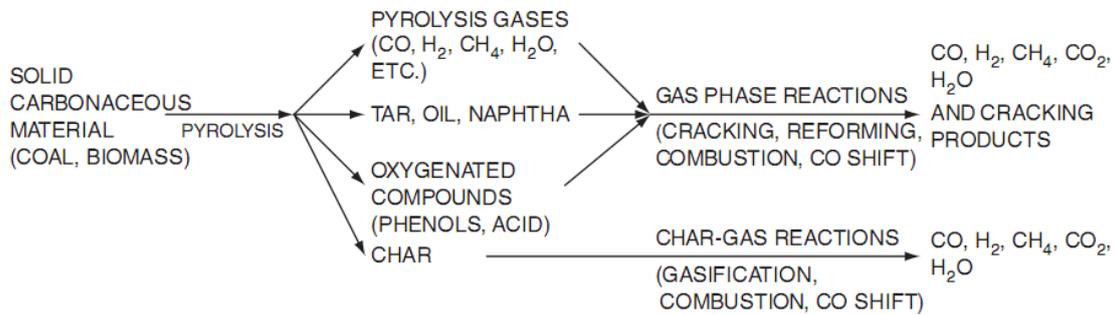
Figure 16. Reaction sequence for gasification of coal or biomass (Higman & van der Burgt, The Thermodynamics of Gasification, 2008)

Depending on the gasification technology used different compositions of the outgoing gas, syngas, is obtained, this in turn decides which subsequent steps in the process are needed to achieve the desired end product.

In the vessel where gasification takes place, the gasifier, a mixture of oxygen and steam reacts with the coal in the lignite to produce a mixture which mainly consists of hydrogen, carbon monoxide, carbon dioxide and steam. The most important reactions to roughly evaluate the gasification of coal is based on reactions according to Equation 2, Equation 3, Equation 4 and Equation 5. (Hiller, et al., 2009)

Equation 2. Combustion reactions

$$C + 1/2O_2 \iff CO \qquad \Delta H = -111 \quad kJ/mol$$

$$CO + 1/2O_2 \iff CO_2 \qquad \Delta H = -283 \quad kJ/mol$$

$$H_2 + 1/2O_2 \iff H_2O \qquad \Delta H = -242 \quad kJ/mol$$
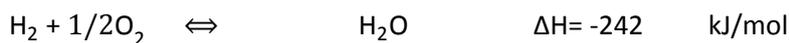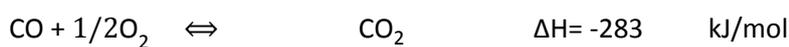
Equation 3. The homogenous water-gas shift reaction

$$CO + H_2O \iff CO_2 + H_2 \qquad \Delta H = -41 \quad kJ/mol$$
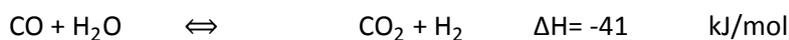
Equation 4. Steam-methane reforming

$$CH_4 + H_2O \iff CO + 3H_2 \qquad \Delta H = +206 \quad kJ/mol$$
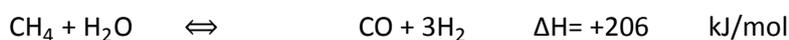
Equation 5. The Boudouard reaction

$$C + CO_2 \iff 2CO \qquad \Delta H = +173 \quad kJ/mol$$

The sign after *ΔH*, implying difference in enthalpy, indicate if the reaction is endothermic, needs heat (positive sign) or exothermic, produces heat (negative sign). The double arrow signs indicate that these are reversible, that is to say that they may proceed from right to left as well as from left to right. The reactions usually occur at the same time within a reactor, only at different rates. So for given conditions the reactions occur in both directions and the final equilibrium concentration, after a period of time is therefore dependant on the temperature and pressure.

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Due to the principal that a system in equilibrium tries to evade a change forced upon it, formulated by Le Châtelier – Braun, the steam-methane reforming reaction (Equation 4) tends to favour the formation of methane at higher pressures due to the higher molecular count on the right hand side of the reaction equation. Similarly the Bouduard reaction (Equation 5) and the homogenous water-gas shift reaction (Equation 5) equilibriums are practically unaffected by changes in gasification pressure. (Hiller, et al., 2009)

The temperature in the gasifier is controlled by adjusting the steam/oxygen ratio. High levels of steam will lower the temperature in the gasifier and increase the amount of hydrogen in the gas mix. Oxygen on the other hand increases the temperature but lowers the hydrogen content in the outgoing syngas. A high temperature is however needed to reach sufficient reaction rates in the gasification process.

There are a number of indicators to describe the performance of gasifiers. The carbon conversion represents how much of the carbon in the fuel that is transferred to gaseous phase, hence an indication of how much of the carbon is left in the ash. A high carbon conversion is accomplished at high temperature and sufficient residence time.

The hot gas efficiency is calculated by taking the sum of the calorific value and the heat energy of the raw syngas exiting the gasifier and dividing by the calorific value of the fuel input. Typically entrained flow gasifiers achieve hot gas efficiencies of 90%.(Göttlicher, 2004)

In a similar manner, the cold gas efficiency is calculated by taking the calorific value of the raw syngas and dividing with the calorific energy of the fuel input. Hence the heat of the raw syngas is neglected. (Higman & van der Burgt, The Thermodynamics of Gasification, 2008) Entrained flow gasifiers typically achieve close to 80% cold gas efficiencies. Moving bed gasifiers produce syngas at lower temperatures and can therefore reach higher efficiencies but as tars, methane and other hydrocarbons - which are produced at lower temperatures - necessarily are not beneficial or wanted it can still result in a lower total efficiency of the plant. (Higman & van der Burgt, Gasification Processes, 2008)

3.4.1. Effects of pressure

When increasing the operating pressure of gasification the reactants to be used in the gasifier have to be pressurised. When the final products after gasification are to be used in a gas turbine, as studied in this report, the syngas has to be fed into the gas turbine at elevated pressure. There are two ways to achieve this, either by compressing the reactants and operating the gasifier at a high pressure; or by operating the gasifier at lower pressure and compressing the final syngas before the gas turbine. As shown by Higman and Burgt (Higman & van der Burgt, The Thermodynamics of Gasification, 2008) in Table 2, where the final product is wanted at 50 bar, it is intuitively advantageous to pressure the reactants instead of the final product.

Table 2. Example of energy requirements for pressurization of reactants or products (Higman & van der Burgt, The Thermodynamics of Gasification, 2008)

		5 bar gasification	50 bar gasification
Feed pumping energy	35 450 kg/h	0.03 MW	0.09 MW
Oxygen compression	21 120 Nm^3/h	2.85 MW	4.97 MW

Syngas compression	100 000 Nm^3/h	19.70 MW	0.00 MW
Total		22.58 MW	5.05 MW

However there are practical issues that make the increased pressure have some inherent disadvantages. At very high pressures the reactant feeding might become severely problematic and result in large inert gas leakage, as the feeding system require a pressurizing gas, counteracting the theoretical advantages.

In addition the equipment used downstream of the gasifier but before the gas turbine are affected by the operating pressure. If the pressure is high and a physical absorbent is used to capture CO_2 prior to combustion in the gas turbine the partial pressure is of importance determining the effectiveness of the CO_2 capture unit or even deciding the technology of the capture unit.

Regarding syngas composition a higher pressure favours the formation of methane as previously stated. However at high temperature the methane formation is not as apparent as at lower temperature. (Higman, Gasification, 2008)

In summary the advantages of high pressure gasification are typically: (Higman & van der Burgt, Practical Issues, 2008)

- the gasification vessel can be made smaller - lowering capital cost and space requirements
- the absorption of acid gases from a physical solvent is more efficient the higher the pressure
- the shift catalyst vessels can be made smaller - decreasing capital costs

The disadvantages of pressurised gasification can generally can be summarized as a matter of higher plant complexity, resulting in lower plant availability.

3.4.2. Effects of temperature

The operating temperature of a gasifier is generally determined by the ash properties in relation to the gasifier design (Higman & van der Burgt, The Thermodynamics of Gasification, 2008). This will be more thoroughly treated in chapter 4.3 (Gasification technologies) but in short is a combination of the ash removal system and ash melting temperature. If the ash is to be removed dry, the operating temperature is to be kept below the ash melting point and if the ash is to be removed as a slag, the operation temperature has to be above the ash melting temperature.

The syngas composition is quite strongly affected by the operating temperature as shown in Figure 17. A high operating temperature of the gasifier lowers the methane content in the syngas but also increases the CO_2 content, and lowers the hydrogen yield.
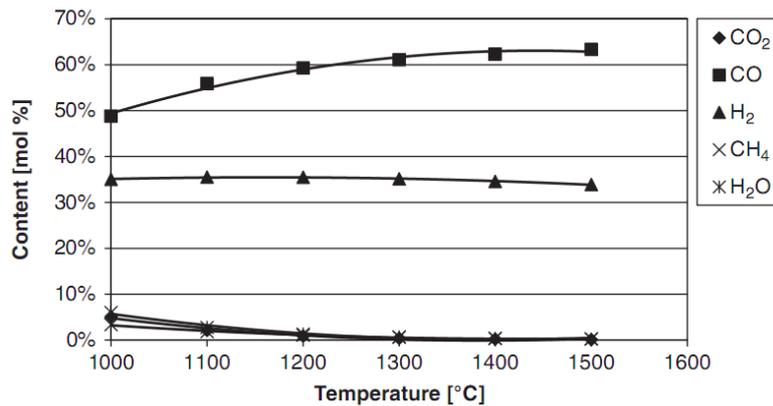
Figure 17. Example of syngas composition depending on temperature at 30 bar (Higman, Gasification, 2008)

3.4.3. Effects of oxidizing medium

When using oxygen or air as a gasifying agent the process will be self sustaining or autothermal, meaning the heat is added directly from the partial combustion in the gasifying zone. No external heat source is needed. When pure oxygen is used, the product gas will be very high quality (12-18 $MJ/_nm^3$) since there is no nitrogen dilution. However, when using air, the quality of the product gas out of the gasifier will be substantially lower (4-6 $MJ/_nm^3$), which is directly linked to the nitrogen dilution. The drawback of the oxygen is that it must be generated at the plant site or bought, introducing efficiency losses or direct economical drawbacks. (Bolhàr-Nordenkampf, 2004)

The third kind of gasification agent is steam. When using steam an external source of heat must be supplied, this is called allothermal gasification. The outgoing product gas is however free from nitrogen and the gas will be of high quality (12-18 $MJ/_nm^3$). This kind of principle is often used in the twin fluidised bed configuration, where the combustion chamber is fuelled by the excess products from gasification, which in return supplies the gasification chamber with external heat. This configuration is often seen in atmospheric biomass gasification plants when the purpose is to produce synthetic natural gas (SNG). (Bolhàr-Nordenkampf, 2004)

3.4.4. Gasifier types

There are three main types of concepts available for coal gasification: entrained flow, fluidised bed and moving bed gasifiers. The aspects of the different techniques are briefly explained in the following sub-chapters.

3.4.4.1. Entrained flow

In an entrained flow gasifier the fuel is fed to a burner where part of the fuel is oxidised in sub-stochiometric conditions in a cylindrical pressure vessel. The residence time in the vessel is counted in seconds and this is possible because of the fine fuel particles used (or sprayed liquid fuel) and the high operating temperature in the gasifier. (Higman & van der Burgt, Gasification Processes, 2008)

The walls of the gasifier vessel are commonly lined with refractory material or cooled by water tubes. During operation the walls are covered with a layer of slag which is molten on the inner side of the gasifier, the side exposed to the gasification conditions. The additional slag produced within the vessel is therefore molten and runs down the sides of the vessel with the help of gravity. This phenomenon makes the viscosity of the slag to a crucial design parameter when constructing this kind of gasifier vessel. A too low viscosity makes the protective layer of slag too thin and exposes the

refractory lining or water tubes to the full force of the gasifier environment. On the contrary, if the viscosity is too high, the flow of slag is impaired, making slag removal problematic. The slag viscosity can be controlled by means of adding fluxing minerals or varying the operation temperature. Dolomite and limestone are additives that can lower the slag viscosity. (Higman & van der Burgt, Gasification Processes, 2008)

The entrained flow gasifier is designed for high throughput and operation at high pressures and temperatures. These operation characteristics results in low capital costs and high carbon conversion efficiencies (95-99%). Additionally the high temperatures results in low amounts of troublesome tar and hydrocarbon formations in the outgoing syngas.(Higman & van der Burgt, Gasification Processes, 2008)

There are two main configurations to the entrained flow gasifier - where the outgoing syngas exits either at the top, called updraft, or at the bottom, downdraft.

Entrained flow gasifier characteristics:

- **High pressure operation and low residence time leading to a compact design**
- **High operating temperature resulting in high carbon conversion, low tar and hydrocarbon formation**
- **Important to take ash characteristics into account during design phase**
- **Requires pre-drying of the fuel**

3.4.4.2. Fluidised bed

In a fluidised bed gasifier, the fuel is fed into a bed of heated inert bed material, often sand or limestone. The fluidisation effect is achieved by a fluidising medium, being air or oxygen, depending on configuration, that is flown through the bed. The stream of fluidisation medium enters the bed from below through pipes at a velocity that enables complete fluidisation. Further increasing the stream velocity of the fluidisation medium increases the amount of feedstock material that is thrown into the volume above the bed, the freeboard.

The residence time of the feedstock is in-between that of the entrained flow (couple of seconds) and the moving bed type (30-60 minutes). The operating temperature is limited to about 800 to 900°C because of the risk of sintering in the bed material, effectively interrupting fluidisation. This relatively low temperature operating temperature results in high levels of hydrocarbons in the exiting syngas. As the ashes do not melt, they are extracted through a cooled screw conveyor.

Fluidised bed gasifier characteristics:

- **Moderate operating temperature resulting in high hydrocarbon content in the raw syngas**
- **Medium residence time**
- **Requires pre-drying of the fuel**

3.4.4.3. Moving bed

In a moving, or sometimes called fixed, bed gasifier, the gasifying medium passes through a bed of granular fuel. The fuel is fed from above and the moving bed is fed with the gasifying agent in the middle of the bed or from the bottom of the bed. In order to sustain gasification the feedstock must be of suitable size and of appropriate size distribution. If the feedstock is large but the size

distribution is not controlled the smaller components of the fuel will block the crucial passages of gas flow through the bed.

The residence time for the fuel in the bed is around 30 to 60 minutes. A unique feature of the moving bed gasifier is that the fuel does not have to be dried before being fed into the gasification vessel as the raw outgoing product gases pre-heats and evaporates the moisture in the feedstock as it exits the vessel in a counter current matter.

Moving bed gasifier characteristics:

- **Requires briquetting**
- **Integrated drying**
- **High hydrocarbon content in raw syngas**

3.5. Shift reaction

After gasification the syngas mainly consists of water, hydrogen, carbon dioxide and carbon monoxide. In order to transform the carbon monoxide into carbon dioxide, which can be separated in a CO_2 separation unit further downstream - while also producing more hydrogen - a shift conversion reactor is needed. This application is needed when using pre-combustion CO_2 separation, separating the CO_2 from the gas stream prior to combustion.

In the shift reactor a catalyst is used to increase the speed of reaction to come closer to the equilibrium of Equation 6, towards the formation of hydrogen and carbon dioxide.

Equation 6. The homogeneous water-gas shift reaction

$$CO + H_2O \quad \Longleftrightarrow \quad H_2 + CO_2 \quad\quad\quad \Delta H= -41kJ/mol$$

The reaction is exothermic and there is therefore an opportunity to raise steam from the heated syngas. (Kohl & Nielsen, Thermal and Catalytic Conversion of Gas, 1997)

3.6. Acid gas removal

The purpose of an acid gas removal unit (AGR unit) is to remove sulphur and carbon dioxide from the syngas stream. This can be done in several different configurations. The sulphur present in the fuel will during gasification react with hydrogen or carbon monoxide, producing H_2S or COS respectively (Higman & van der Burgt, The Thermodynamics of Gasification, 2008). These compounds will produce SO_2 during combustion and small amounts of SO_3 (Kohl & Nielsen, Sulphur Dioxide Removal, 1997).

There is a principle difference in the acid gas removal in the two main CO_2-capture configurations which are simulated. In the pre-combustion capture concept, the CO_2 and sulphur compounds, COS and H_2S, are captured prior to combustion. In the post-combustion case the sulphur compounds can be partly removed prior to combustion, while the CO_2-capture and main sulphur removal is done in the flue gas stream, which will be treated in the next chapter.

Because the delivered CO_2 should practically only contain CO_2 it is assumed that co-capture of CO_2 and H_2S from the syngas is not a viable option for on-shore aquifer storage (for Enhanced Oil Recovery applications co-capture of CO_2 and H_2S could be an option).

The removal of CO_2 and H_2S can be done by applying two consecutive stages of acid gas removal, due to the fact that the sulphur removal unit (SRU) needs a H_2S-rich gas feed, hence requiring the AGR unit to have a selectivity of H_2S over CO_2. The operation of the AGR plant to be able to remove both H_2S and CO_2 and delivering these substances to two separate streams is shown in principal in Figure 18. (Karlsson, Köpcke, Simonsson, & Wolf, 2007)
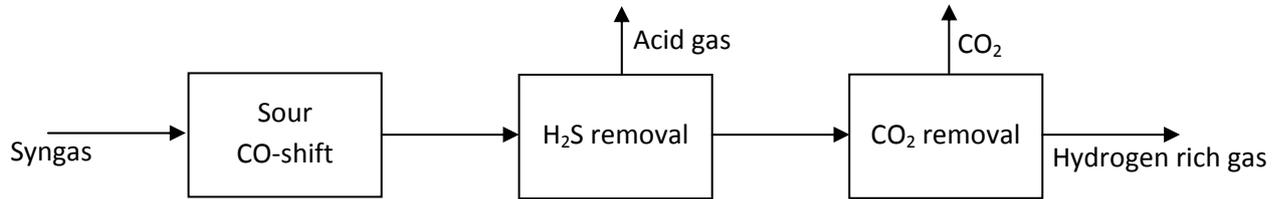
Figure 18. Sour shift with consecutive stages of acid gas removal(Karlsson, Köpcke, Simonsson, & Wolf, 2007)

This configuration includes a sour shift reactor stage prior to the removal of H_2S and CO_2. This is so that there will be no CO in the syngas that would otherwise pass through, reducing the total carbon capture rate of the plant.

If a clean shift reactor is used instead then the syngas entering the shift reactor must be cleaned from sulphur. Therefore the sulphur removal system must come before the clean shift reactor and the CO_2-capture unit is then placed after the clean shift reactor as shown in Figure 19. (Karlsson, Köpcke, Simonsson, & Wolf, 2007)
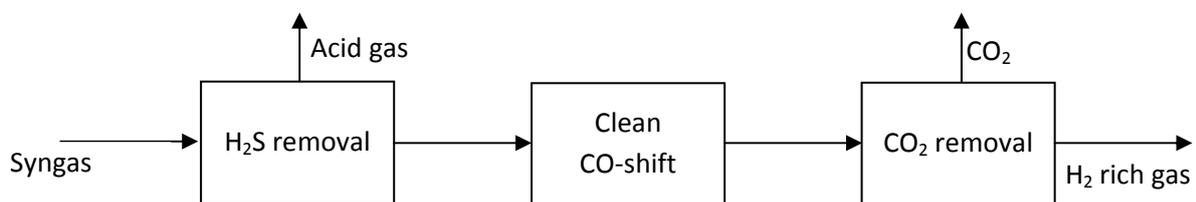
Figure 19. Clean shift initial sulphur removal and downstream CO_2 removal (Karlsson, Köpcke, Simonsson, & Wolf, 2007)

3.7. Flue gas cleanup

Flue gas cleanup concerns cleanup of a flue gas, while the before mentioned acid gas removal refers to cleaning of a syngas stream. The most common working principles for removing sulphur and carbon dioxide from a flue is presented in the following chapters.

3.7.1. CO_2 removal systems

Absorption systems remove acid compounds from the gas by washing with a liquid solvent. The type of solvent can be chemical or physical or a mix of the two and relates to how the solvent binds the acid gas to the solvent. Chemical solvents bind the acid gas through chemical bonds and therefore an increased partial pressure of the acid gas only increases the loading capacity of the solvent to a certain extent. For physical solvents, however, the partial pressure of the acid gas is directly proportional to the loading capacity of the solvent, as shown in Figure 20. (Higman & van der Burgt, Auxiliary Technologies, 2008)
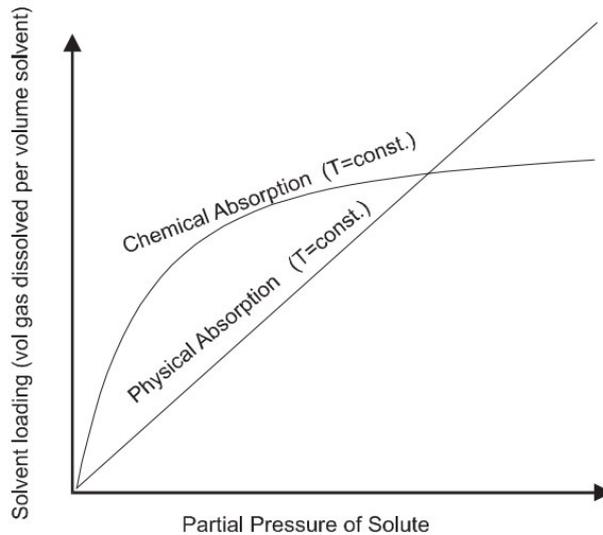
Figure 20. Solvent loading as a function of solute partial pressure (Göttlicher, 2004)

The chemical solvents are usually based on amines like mono- and diethanolamine (MEA and DEA), methyldiethanolamine (MDEA). The chemical solvents are regenerated by boiling the solvent. The increased temperature of the loaded chemical solvent breaks the chemical bonds, releasing the acid gas in a regenerator as shown in principal in Figure 21. The chemical solvents therefore require large amounts of heat. (Higman & van der Burgt, Auxiliary Technologies, 2008)
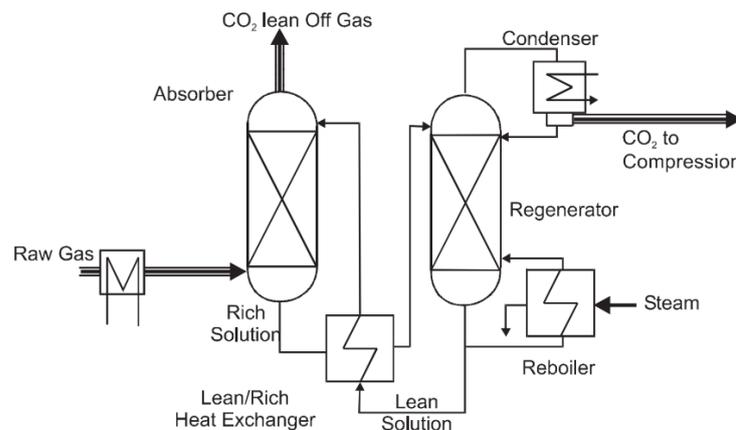
Figure 21. Generic flow diagram of a chemical absorption system (Göttlicher, 2004)

In physical solvents the loaded acid gas is released in a so called stripper where the partial pressure of the absorbed gas is reduced. Instead of a high heat demand physical solvents require more electricity to drive the solvent pumps to bring the solvent back to the operating pressure and to operate the refrigeration system. Figure 22 presents a simplified schematic of a physical absorption process.

3.7.2. SO$_2$ removal systems

The dominating technology used for removing sulphur dioxide from flue gases is through water slurry scrubbing of the flue gas. The active substance dissolved in the water reacts with the sulphur dioxide and the flue gas is cleaned. The resulting slurry can commonly be regenerated and a usable sulphur-based by-product can be produced. (Kohl & Nielsen, Sulphur Dioxide Removal, 1997)

Alternatives to the wet scrubbing processes are the dry injection and spray drying desulphurization systems. In the dry injection system there is no water present when injecting the absorbent into the flue gas. The spray drying or semi-dry desulphurization system commonly uses lime as an absorbent and the water in the slurry is evaporated in the flue gas stream. The reaction product, calcium salts, must be removed in a particle control device downstream of the spray drying desulphurization system. (Bruce, 2005) These simple systems lead to less complex installation but render no usable by-product and lower SO$_2$ removal efficiency. (Kohl & Nielsen, Sulphur Dioxide Removal, 1997)

3.8. Hot gas particle removal

The removal of particles prior to syngas feed into the gas turbine is essential to protect the gas turbine from being permanently damaged. (Wright & Gibbons, 2007)

To remove particles from a hot gas there are two options available; cyclones and filters. Cyclones use the gravitational force to separate out coarse particle from the gas. This principle is mostly applied right after a fluidized bed combustor or gasifier, to recycle entrained bed particles (Higman & van der Burgt, Gasification Processes, 2008).

In order to remove fine particles from a gas at elevated temperature, filtering is the only viable option. At temperatures of 250-300°C the filters may be blinded by deposits of alkali compounds. At temperatures lower than 600°C the alkali compounds condense and stick to the fly ash which then is filtered out. The filters may be either ceramic or metallic and are discussed more in Theory: Process Units under chapter 4.6, Hot gas filters. (Higman & van der Burgt, Practical Issues, 2008)

3.9.Gas turbine power cycles

A schematic diagram of the simple-cycle gas turbine is shown in Figure 23.
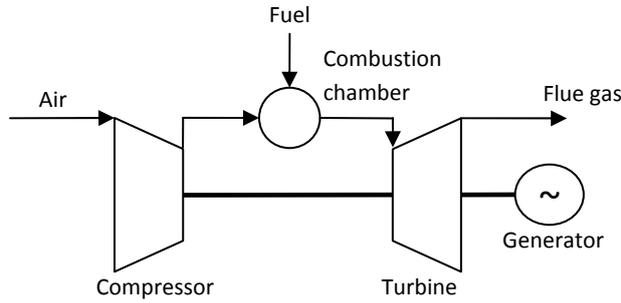
Figure 23. Simple gas turbine system

Air enters the axial compressor and is compressed to some higher pressure. There is no heat added, however compression raises the air temperature so that the air at discharge is at a higher pressure and temperature. In the combustion chamber fuel is injected into the air stream and continuous combustion occurs. In the turbine section, after combustion, the energy of the hot gases is converted into useful work. The conversion takes place in two steps. In the nozzle section of the turbine, the hot gasses are expanded and a portion of the thermal energy is converted into kinetic energy. In the following blade section the turbine, a portion of the kinetic energy is transferred to the rotating blades and converted to work. Some of the work is used to drive the compressor, and the remainder is used to drive the generator producing electricity (Saravanamuttoo, Rogers, Cohen, & Straznicky, 2009). The flue gases exiting the turbine have a temperature of 400-600°C, but despite this large energy loss modern gas turbines achieve efficiencies up to 42% (Poullikkas, 2005).

3.9.1. Combined cycle

Even though the simple gas turbine cycle today is able to reach efficiencies of over 40%, the system still looses much efficiency by ejecting high temperature heat to the surroundings as flue gas exhaust. The exhaust heat can instead be utilised in a variety of ways. One is to produce steam in a heat recovery steam generator (HRSG) for a steam turbine, to increase the total work output the system. This is referred to a combined cycle plant, shown in its simplest form in Figure 24.
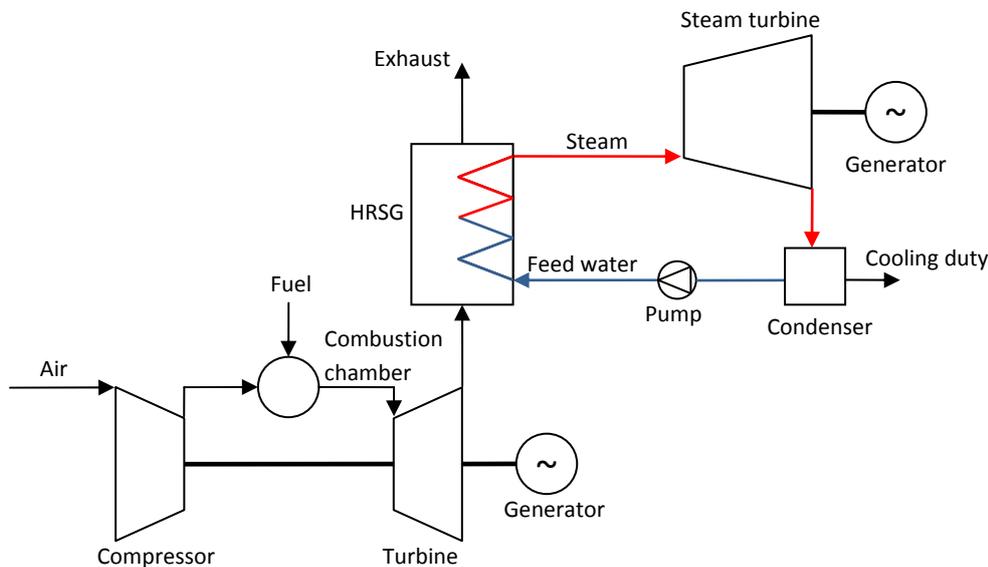
Figure 24. The Combined cycle

Heat utilization is maximised by applying multiple-pressure boilers to match the characteristics of the flue gas stream. A combined cycle is able to achieve electrical efficiencies of 58-60 % when employing triple-pressure HRSG with a reheat cycle and using natural gas as fuel (Khartchenko, 1998).

3.9.2. Humidified gas turbine

The principle of the humidified gas turbine compared to the simple cycle gas turbine is that steam is generated using process heat to increase the flow of working fluid passing through the turbine without increasing compression work. The increased mass flow through the turbine consequently increases the work output of the turbine (Khartchenko, 1998). By raising steam in a heat recovery steam generator (HRSG) and injecting it into the combustor the efficiency increases. The compression work of the compressor is the same and a relatively small portion of work is required to increase the pressure of the feed water to the HRSG. As schematic of this, so called, steam-injected gas turbine configuration is shown in Figure 25. (Jonsson & Yan, 2005)
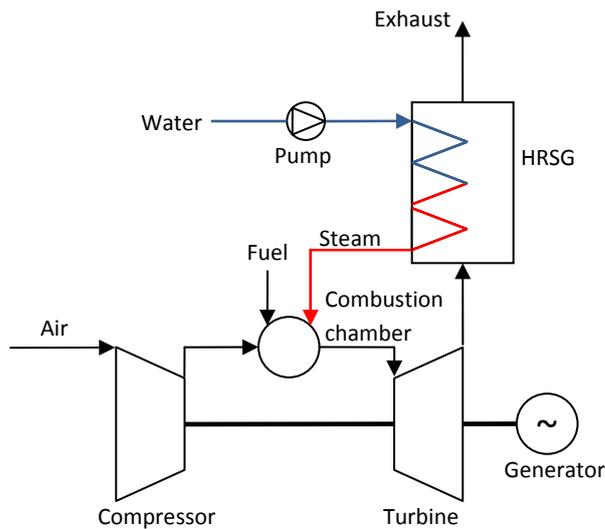
Figure 25. Schematic of a steam-injected turbine

3.9.3. Syngas purity requirements

Hot corrosion in gas turbines is caused by molten salts which deposit on surfaces at high temperature. The salts are in vapour phase harmless, but once they condense they accelerate the oxidation of the surface alloy in the temperature range of 700°C and 925°C. It is customary to call corrosion that occurs above the melting point of the salt "type I" corrosion. The "type II" corrosion occurs at the lower temperature range. In both types of corrosion the salts corrode the protective oxide scale which covers the alloy surface. Once the protective layer is breached the alloy becomes vulnerable to oxidation. (Bose, 2007)

The salts responsible for the hot corrosion are typically alkali and alkaline sulphates (Bose, 2007). . It is therefore important to limit the amount of sulphur and alkali metals which could lead to corrosion in the turbine. Experience shows that sulphur levels up to 1% (by volume or weight depending on source) (GE Power Systems, 2002; Boyce, 2006) does not significantly affect oxidation/corrosion rates in the turbine.

More specific recommendations regarding fuel requirements for gas turbines can be found in Specification for Fuel Gases for Combustion in Heavy-Duty Gas Turbines (GE Power Systems, 2002). It should however be noted that the limits differ between turbine models, and that figures regarding the alkali metals sodium and potassium limits are commented by GE to be reviewed "on a case-by-case basis"(GE Power Systems, 2002).

3.9.4. Hydrogen combustion

The syngas after gasification mainly consists of H_2, CO, CO_2 and H_2O. Upon entering the gas turbine the syngas composition will depend heavily on the system layout, being pre- or post-combustion CO_2-capture. In pre-combustion CO_2-capture case the CO_2 will be removed and the syngas will on a volume basis mainly consist of hydrogen and dilution gas (N_2), needed for fuel feeding into the gasifier, resulting in a heating value of about 22 MJ/kg. In a post-combustion CO_2-capture case the syngas mainly consist of N2, H2O (due to water injection and air blown gasification) and a medium amount of the combustible gases CO and H_2, as well as a small portion of CH_4 and CO_2. Naturally this kind of syngas has a very low heating value, in the region of 4 MJ/kg. In the Värnamo IGCC demonstration plant, Sweden, it has been shown that low heating value syngas (4 MJ/kg) can be fired

in an almost standard gas turbine. The gas is fired without a pilot flame in a Typhoon gas turbine with a modified combustor, from ABB Alstom Gas Turbines in Lincoln, England (Ståhl, Waldheim, Morris, Johansson, & Gårdmark, 2004).

The low heating value of syngas, about 20% of the heating value for natural gas, from air or oxygen (without CO_2-capture) blown gasification affects the layout of the burner nozzles and fuel supply system for the gas turbine. A larger fuel flow leads to a large fuel feeding system and higher costs.

The combustion of hydrogen is very different from combustion of natural gas, which is the ordinary fuel is in gas turbines. Hydrogen has compared to natural gas, which mainly consists of methane, several distinct different characteristics, as shown in table Table 3.

Table 3. Comparison between methane and hydroge (Paschereit & Göke, 2009)

	Methane	Hydrogen
Density	0.7 kg/m^3	0.09 kg/m^3
Lower heating value	33 MJ/m^3	10 MJ/m^3
Lower heating value	50 MJ/kg	120 MJ/kg
Quenching distance	2 mm	0.7 mm
Limits of flammability (vol% in air)	5 – 15	4 – 76
Limits of flammability (mass% in air)	2.9 – 9	0.3 – 18
Maximum laminar flame velocity	0.4 m/s	2.9 m/s
Adiabatic flame temperature	2236 K	2391 K

The low heating value per unit volume for hydrogen together with the low quenching distance means that a combustor for hydrogen has to have different burner design than for natural gas. There are also flashback issues and the hydrogen flame has a higher temperature and generally different kinetics and this is the reason to why there is no technique today to burn hydrogen at a low level of NO_x emissions. (Paschereit & Göke, 2009)

Simulations have however shown that by injecting steam, hydrogen can be burnt at low NO_x emission levels similar to that of methane combustion. Furthermore the hydrogen flame has shown to be stable at high humidity. (Paschereit & Göke, 2009)

The increased water vapor presence in the gas turbine is however shown to lower the corrosion resistance of alumina-forming alloys in the gas turbine, by favouring the formation of less protective alumina scales. (Wright & Gibbons, 2007)

4. Theory: Process Units

This chapter describes the different technologies and process units which are evaluated and used as basis for the concept selections.

4.1. Drying technologies

There are numerous ways to dewater lignite (Allardice, Chaffee, Jackson, & Marshall, 2004) but only mill drying, steam fluidized bed drying and hot or warm gas drying will be discussed.

4.1.1. Mill drying

Mill drying is the conventional process where the lignite is milled and dried simultaneously. The required heat to dry the fuel is provided by furnace off-take gas (Allardice, Chaffee, Jackson, & Marshall, 2004). Since a furnace in an ordinary fashion will not be available in this context, the most similar solution would be to use flue gases after the gas turbine.

4.1.2. Steam fluidized bed drying

The steam fluidised bed drying technology utilizes superheated steam to fluidize a bed of fuel. Through immersion heaters in the bed, steam is condensed to supply a majority of the heat for the drying process. In this way an efficient high heat transfer drying process is achieved, enabling a very compact design compared to warm gas drying. The almost pure steam atmosphere also effectively eliminates the risk of fires and explosions. (Allardice, Chaffee, Jackson, & Marshall, 2004)

There are principally two different configurations of this drying technology, with different steam supply strategies. In the full heat recovery configuration the steam supplied is taken from product gas exiting the fluidizing vessel, which is de-dusted, then recompressed and fed to the immersion heaters. A more straightforward approach is simply to take external steam. In both configurations however, the steam enabling fluidization is taken from the saturated product gas exiting the fluidizing vessel which is recompressed and fed to the nozzles in the bottom of the bed. The choice of having full heat recovery or using external steam is in the end a question of cost for the extra complexity versus efficiency. A presentation of the two concepts can be seen in Figure 26. (Allardice, Chaffee, Jackson, & Marshall, 2004)
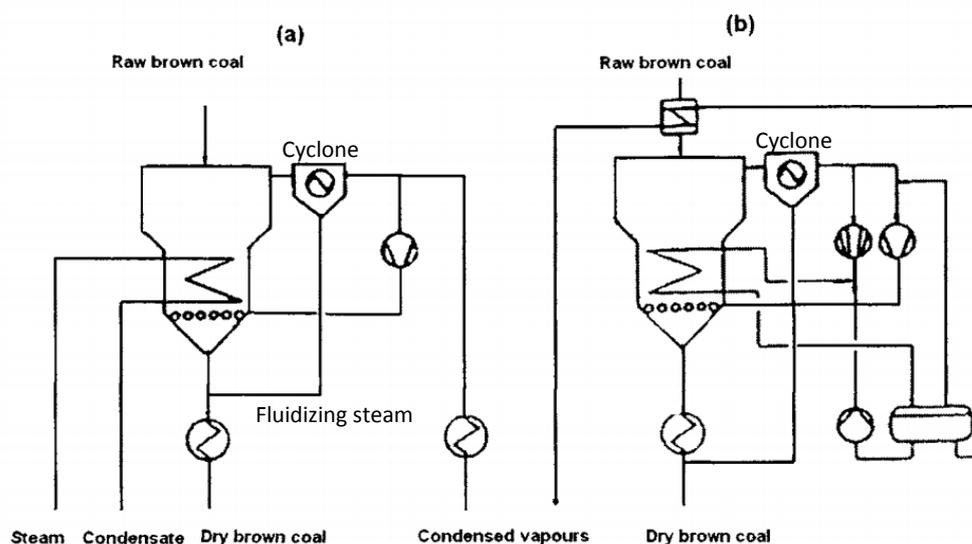
Figure 26. Schematic picture of (a) external steam supply and (b) vapour recompression (Allardice, Chaffee, Jackson, & Marshall, 2004)

RWE and Vattenfall have two different configurations mainly differing in operational pressure. The WTA technology, applied by RWE operates at atmospheric pressure while the Vattenfall counterpart operates under pressure. The pressure results in a higher heat transfer coefficient, enabling a smaller drying unit and lower investment costs (Hoehne, Lechner, Schreiber, & Krautz). The WTA technology is currently demonstrated in Niederaußem, Germany (210 t/h of raw coal) where the coal is milled prior to drying to increase heat transfer.

A low-temperature lignite dryer, has been developed by the Great River Energy (GRE) and Lehigh University's Energy Research Center (ERC) in the U.S. This drying technology is a combination of the steam fluidized bed dryer and a warm gas dryer, described in principal in section 4.1.3. The GRE-dryer uses air as fluidization medium instead of steam but include immersion heaters fed with steam just as in the steam fluidized bed concept. (Clean Coal Power Initiative, 2008)

4.1.3. Hot or warm gas drying

An alternative to steam drying is to use another drying medium, such as air or off-gas nitrogen from an air separation unit. Gas drying is a better option when utilizing low grade heat, since steam needs to be overheated in order to dry fuel, which requires higher temperatures or lower pressure. The drying medium in a gas drying process is supplied with a fan to a heat exchanger, indirectly heated by the low grade heat available and then supplied to the dryer. A simplified overview of the process is illustrated in Figure 27.
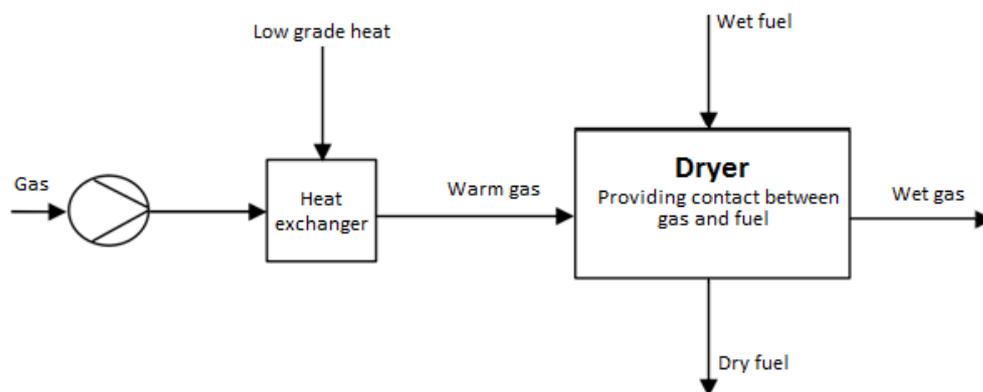
Figure 27. Simplified flow scheme of a warm gas dryer

It is important to be aware of the flammability risks when drying the fuel with air and this limits the temperature of the air that can be supplied to the dryer. However, low grade heat implies low temperatures so flammability is not necessarily a problem in all cases. (Johansson, Larsson, & Wennberg, 2004)

4.1.3.1. Swiss Combi

Swiss Combi is a belt dryer which utilizes low grade heat (~80°C) to heat air which is blown trough the layer of moist fuel situated on the belt. The dryer is built in modules and designed for easy transport in standard trucks and sea-containers. An overview of the dryer is shown in Figure 28. (W, Kunz dryTec AG, 2010)

1. Wet material
2. Feeding screw 1
3. Discharing screw 1
4. Feeding screw 2
5. Discharge of dried material
6. Heat exchanger
7. Extractor fan
8. Cleaning brush
9. Belt washing device
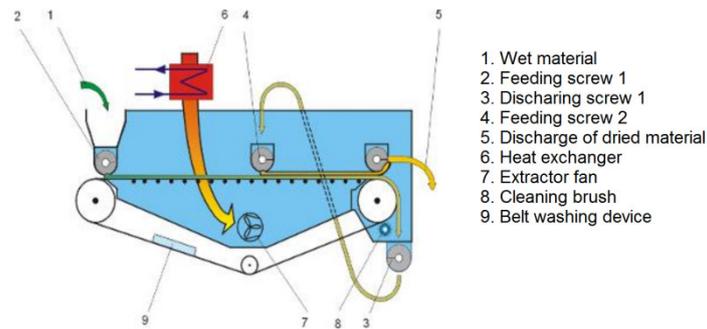
Figure 28. Simplified process flow diagram of the Swiss Combi belt dryer. (W, Kunz dryTec AG, 2010)

4.2. Air separation technologies

4.2.1. Cryogenic

In order to provide a sufficiently pure stream of oxygen to the gasifier an air separation unit (ASU) of considerable size is required. Today the only commercial technology to supply oxygen for this purpose is a cryogenic based air separation unit.

The working principle behind the cryogenic air separation unit is logically enough cryogenic separation of oxygen from air. This means oxygen is separated from the other components - mainly nitrogen - at a very low temperature after the initial air compression.

The basic process is shown in Figure 29. The process demands a large amount of electricity and also some heat for the regeneration of molecular sieves used to remove contaminants. (Allam, Castle-Smith, Smith, Sorensen, & Stein, 2000)
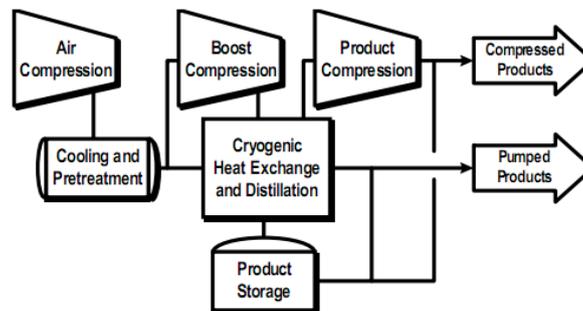
Figure 29. Simplified flow diagram of a cryogenic ASU. (Allam, Castle-Smith, Smith, Sorensen, & Stein, 2000)

In Table 4 below some numbers regarding the electricity consumption from various reports are presented to give an indication of the power consumption for a cryogenic ASU.

Table 4. Examples of electrical consumption for a cryogenic ASUs from various literature sources

Source	Power demand [kWh/kg O_2]	Oxygen purity [vol %]	Oxy deliv. P. [bar]	Nitrogen deliv. P. [bar]	Integration level
Göttlichers 2004	0.250 kWh/kg O_2	0.95 %	1 bar	1 bar	No GT integration
Chiesa et al. 2005	0.261 kWh/kg O2	0.95 %	1.05 bar	1.05 bar	No GT integration
Confidential, Source 2	0.225 kWh/kg O_2	0.95 %	2.379 bar	2.379 bar	No GT integration

As previously mentioned – a small amount of heat is needed to regenerate the molecular sieves.

4.2.2. High temperature membrane

High temperature membrane air separation unit (HTM-ASU) is a concept under development. The concept is based on a dense ceramic membrane which under high temperature begins to conduct oxygen ions, shown in Figure 30. The driving force for the process is the oxygen partial pressure - requiring highly compressed air – which can be further enhanced by increasing membrane temperature and decreasing membrane thickness. (Kather & Pfaff, 2009)
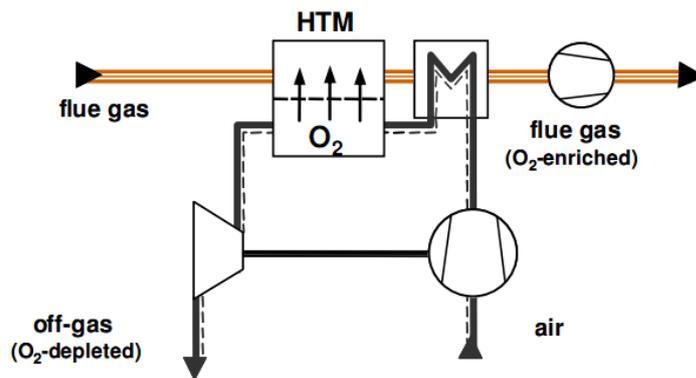
Figure 30. Schematic picture of a HTM-ASU (Kather & Pfaff, 2009)

4.3. Gasification technologies

The presented gasifiers are divided into oxygen blown and air blown gasifiers although most gasifiers can be operated with more or less success on either gas composition. A summary of the gasifiers can be seen in Appendix D together with an overview of the ongoing IGCC projects in the world.

4.3.1. Oxygen blown gasifiers

The following gasifiers are in literature treated as primarily designed for oxygen blown operation.

4.3.1.1. British Gas Lurgi gasifier

The gasifier from British Gas Lurgi (BGL) is of the moving bed type, shown in Figure 31. It is an updraft gasifier, meaning that the produced syngas exits at the top of the vessel, where it is also washed with quench water upon exit. The fuel is fed with a dry feed lockhopper system and the slag is lead to a quench bath at the bottom of the gasifier through a water cooled slag tap for periodic discharge.

The commercial version of the BGL gasifier is 3.6 metres in diameter which corresponds to about 90 MW_e in coal IGCC configuration, which means that several gasifiers would be required to achieve an

IGCC concept of true commercial scale(Higman & van der Burgt, Gasification Processes, 2008). The BGL gasifier can be expected to achieve cold gas efficiencies of 85-88%. (Hiller, et al., 2009)

A characteristic derived from the large hearth is that the gasifier is able to be turned down to zero output and then quickly started from a hot state due to the retained heat in the hearth. However start-up from cold state would take considerably more time, about 2 days. A further characteristic originating from the moving bed design is the high amount of methane that would be present in the exiting syngas due to the low operating temperatures.(Higman & van der Burgt, Gasification Processes, 2008)
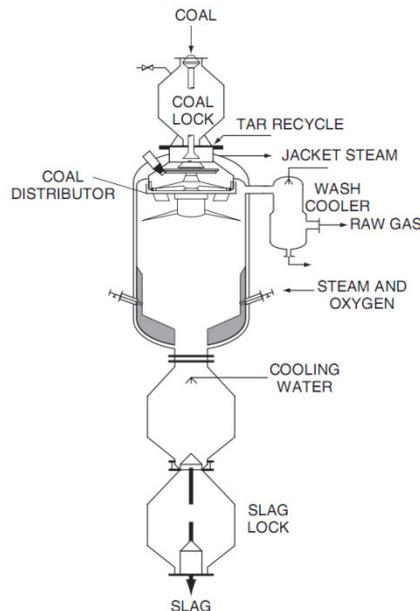
Figure 31. BGL gasifier (Higman & van der Burgt, Gasification Processes, 2008)

4.3.1.2.   Conoco Phillips E-Gas gasifier

The Conoco Phillips E-Gas (CoP E-Gas) gasifier is an updraft entrained flow gasifier with 2-stage slurry fuel feed system. The gasifier also employs a water quench and a full syngas heat recovery system, raising steam.

About 80% of the fuel is fed into the bottom part of the gasifier together with oxygen. In the second upper stage the remaining 20% of the fuel stock is fed but in absence of oxidizing medium, efficiently reducing the temperature of the syngas to just over 1000°C.

The shape of the vessel results in a limited flexibility in regards of operating pressure. This is due to the stress concentrations and welding problems imposed by the joining of two cylindrical pressure vessels (Figure 32).
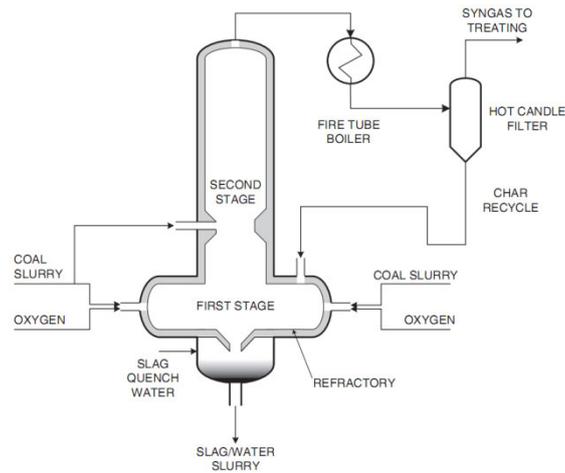
Figure 32. Conoco Phillips E-gas gasifier (Higman & van der Burgt, Gasification Processes, 2008)

Estimated cold gas efficiency from the Conoco Phillips gasifier is in the region of 70% with about 2% methane present in the syngas. The carbon conversion efficiency is estimated to be around 98% which implies that there might be some tars present in the syngas. (Higman & van der Burgt, Gasification Processes, 2008)

4.3.1.3. Siemens gasifier

The gasifier from Siemens is a downdraft entrained flow gasifier with a water cooled slag wall and water quench as seen in Figure 33.
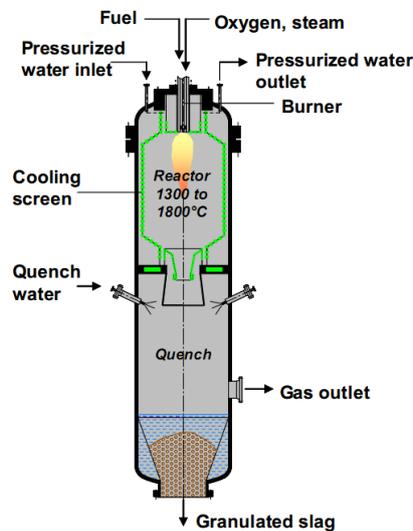
Figure 33. Siemens gasifier (Siemens, 2009)

The gasifier wall is protected by a layer of slag. The cooled wall solidifies the deposits in contact with the wall while further deposits form a melting slag which drips down the inner wall. This protects the wall of the gasifier and makes the slag drip down into the quench bath to instantly solidify. The design does however impose a requirement on the fuel which has to contain at least 2% of ash. (Siemens, 2009)

The cooling screens low thermal mass allows for rapid start-up and shut-down compared to refractory lined design solutions and has a life-time of at least 10 years. The burner produces a swirling flame to increase residence time and reduce vessel size.

When the hot syngas exits the throat of the gasifier into the quench water vessel underneath the syngas is immediately sprayed with water. The water sprays effectively washes any particulate matter into the water quench bath and the syngas exits from the gasifier saturated.

A carbon conversion above 99% and a cold gas efficiency of about 76-79% is expected from the Siemens gasifier.(Confidential, Source 2)

4.3.1.4.   General Electric gasifier

All gasifiers provided from General Electric (GE), originally developed by Texaco, are entrained down-draft gasifiers which employ slurry feed systems. There are three different configurations of the GE gasifier: GE Water Quench, GE Full Heat Recovery and GE Radiant Cooler. All configurations utilise the same gasification concept with a carbon conversion of >95% and a cold gas efficiency of 60-63%. (Confidential, Source 2)

The GE Water Quench has refractory-lined gasifier walls with a water quench at the bottom. The GE Full Heat Recovery gasifier also uses a refractory-lined wall but the syngas is cooled after the main gasifier in a radiant cooler section down streams. At the bottom of the gasifier vessel there is a slag quench and a lock hopper system is used to remove the solidified slag. After the radiant cooler the syngas is led into a separate fire-tube convective heat exchanger, from which the syngas exits at approximately 400°C.(Confidential, Source 2)

The last configuration is a compromise between the first two. The gasifier vessel with the radiant cooler remains the same but the syngas is sprayed with water before exiting. The fire-tube heat exchanger is also removed, hence resulting in a solution which combines the water quench system with a radiant heat exchanger generating steam. (Confidential, Source 2)

4.3.1.5.   Lurgi Dry Ash gasifier

The Lurgi dry ash gasifier is a moving bed gasifier with dry feed and a water cooled jacket wall. The ash is removed dry, as the name implies, so there is no melting of the ash in the fuel. The syngas exits above the bed in an updraft manner and is quenched upon exit. (Higman & van der Burgt, Gasification Processes, 2008)

The Lurgi gasifier was developed in early 20th century and has been used for producing town gas, as well as but also chemical feedstock. The gasifiers are 3 to 5 meters in diameter (Higman & van der Burgt, Gasification Processes, 2008), depending on model, and operate at around 30 bar. This corresponds to about 100 MW electrical output and means that several units would have to be used in order to run a full scale IGCC plant.

The fuel is fed from the lockhopper system at the top of the vessel as illustrated in Figure 34. As the coal enters it is distributed over the bed by a rotating stirrer. Steam and oxygen is fed to the vessel from under the bed. In the hearth of the bed the reaction temperature is about 1100°C – sufficient to allow gasification reactions to take place but low enough to not melt the ash. The ash is removed in a lockhopper system below the rotating grate which is cooled by the flow of steam and oxygen from underneath. (Higman & van der Burgt, Gasification Processes, 2008)

38
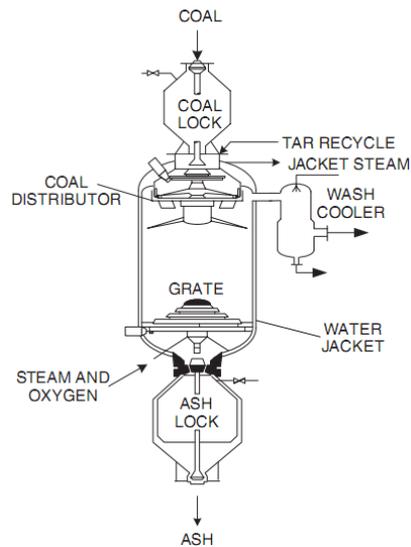
Figure 34. Lurgi dry ash gasifier (Higman & van der Burgt, Gasification Processes, 2008)

As with the BGL gasifier the Lurgi gasifier is a moving bed gasifier and has similar characteristics. The exit syngas is high in methane content and the tar and phenols washed out in the water quench cannot be to recycle back into the gasifier, resulting in an efficiency loss for the overall system. Depending on fuel the Lurgi dry ash moving bed gasifier has shown cold gas efficiencies of 76% to 87% (Hiller, et al., 2009).

4.3.1.6. Oxygen blown High Temperature Winkler gasifier

The High Temperature Winkler (HTW) is a fluidised bed gasifier. The dry feedstock is pressurised in a lockhopper and fed to a fuel bin from which the fuel is screwed to the middle of the gasifier. The fluidizing medium is oxygen and steam and the vessel diameter is 2.75 m. The bed material and fuel flows up the freeboard where additional steam and oxygen is added to complete the gasification reactions. A cyclone separates the particles from the hot syngas before it exits the gasifier and the particles are recycled back to the fluidized bed. The temperature of the syngas leaving the gasifier is in the region of 900°C. (Higman & van der Burgt, Gasification Processes, 2008)

The bed temperature is kept low enough to avoid bed sintering and avoid total operational halt. The second oxidizing injection in the freeboard increases the temperature in order to enhance reaction speed and increase the overall carbon conversion – but is kept below ash melting temperature to avoid ash sticking to the upstream cyclone. (Higman & van der Burgt, Gasification Processes, 2008)

At least three commercial size HTW gasifier have been built; one in Oulu, Finland, and two in Germany in Wesseling and Berrenrath. The plant in Berrenrath supplied syngas based on lignite (Collot, 2006) to a methanol plant for 12 years with an availability of over 84%. Today none of the plants are in operation. (Higman & van der Burgt, Gasification Processes, 2008)

On brown coal the HTW has achieved cold gas efficiencies of 80% with a carbon conversion of 94%. (Hiller, et al., 2009)

4.3.1.7. Prenfo gasifier

The Prenflo gasifier, uses four burners mounted tangentially in the cylindrical vessel. The fuel is blown into the gasifier using inert gas, nitrogen, together with steam and oxygen. The four burners

produce a swirling flame, increasing residence time of the flame and causes the molten slag to stick on the walls of the gasifier, which eventually starts to drip down into the quench bath. Figure 35 shows then Prenflo gasifier with steam generation, the so called PSG. (Udhe)
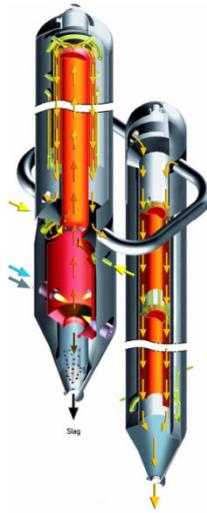
Figure 35. PSG gasifier (Udhe)

The cooling screen is water cooled and has the same advantages as with other water cooled designs such as Siemens, long lifetime and quick start-up/shut-down. The carbon conversion from the PSG is expected to be close to complete (>99%) and the cold gas efficiency in the region of 74-77%. Recently a direct quench version of the Prenflo gasifier has been presented, the Prenflo direct quench (PDQ), shown in Figure 36. (Udhe)
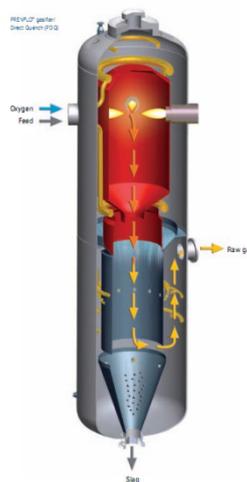
Figure 36. PDQ gasifier (Udhe)

The quench version of the Prenflo gasifier means lower investment costs making the system similar to the Siemens gasifier with water quench.

4.3.2. Air blown gasifiers

The presented gasifiers under this chapter are primarily designed for air blown operation (Higman & van der Burgt, Gasification Processes, 2008).

4.3.2.1. HRL gasifier

The HRL gasifier is a fluidised bed gasifier specifically designed for the very moist Australian lignite. The special feature of this gasifier is the co-current recycling of syngas which dries the fuel, prior to entry into the gasifier, as shown in Figure 37. This enables the gasifier to handle high moisture lignite without additional pre-processing. The incoming fuel is dried to about 5-10% moisture before being fed into the gasifier and while the syngas post-drying has a temperature of approximately 200-250°C. (Higman & van der Burgt, Gasification Processes, 2008)
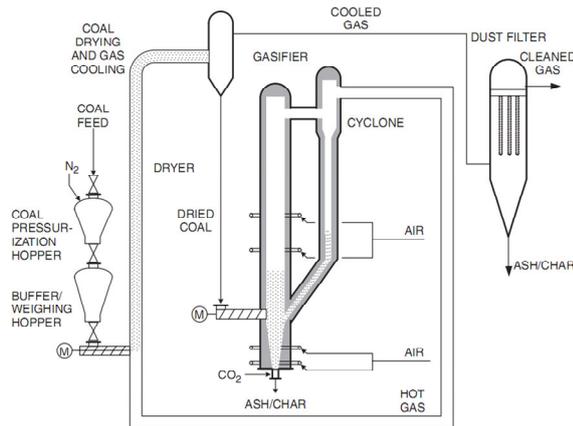
Figure 37. HRL gasifier (Higman & van der Burgt, Gasification Processes, 2008)

Due to the fact that the system is air blown the final heating value of the outgoing gas is very low, in the region of 3-3.5 MJ/kg. The system is reported to be able to achieve energy efficiencies of 38-41% on HHV-basis which is high considering the high moisture fuel used. (Higman & van der Burgt, Gasification Processes, 2008)

The process is yet to be demonstrated on a larger scale and the pilot rig has been demonstrated on the scale of 240 t/d of lignite, which supplied syngas to a 5 MW Typhoon gas turbine. (Higman & van der Burgt, Gasification Processes, 2008)

4.3.2.2. Kellogg Brown and Root transport gasifier

The Kellogg Brown and Root (KBR) transport gasifier (Figure 38) is a circulating fluidized bed gasifier operating at a high gas velocity. The higher than usual gas velocity promotes mixing and allows higher fuel throughput. The KBR gasifiers has been operated both air and oxygen blown and carbon conversion ratios of 95% to 98% have been obtained. (Higman & van der Burgt, Gasification Processes, 2008)
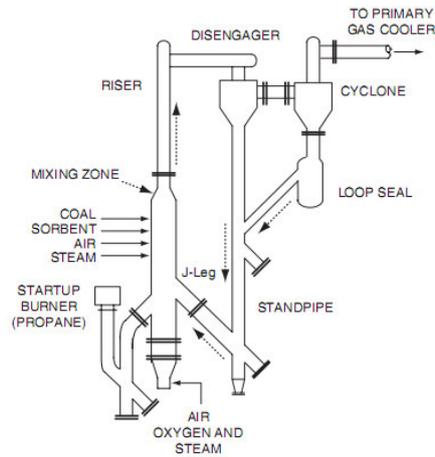
Figure 38. KBR gasifier (Higman & van der Burgt, Gasification Processes, 2008)

4.3.2.3. Mitsubishi Heavy Industries gasifier

The gasifier from Mitsubishi Heavy Industries (MHI) is an updraft entrained flow gasifier with dry feed and two-stage feeding (Figure 39). Although the gasifier is said to be air blown it does in fact use oxygen enriched air, which means surplus oxygen from the nitrogen plant (ASU) - needed to supply inert gas for feed transport - is mixed with the air fed supplied to the gasifier.

The two-stage feeding system is similar to that of Conoco Phillips E-gas. The first fuel inlet operates as a combustor and promotes very high temperatures, enabling slagging ash removal. The second stage fuel inlet supplies additional fuel to the reactor in the absence of further oxidising medium, which effectively lowers the temperature of the syngas.
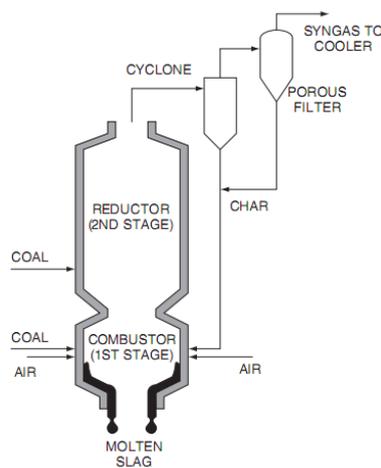
Figure 39. MHI gasifier (Higman & van der Burgt, Gasification Processes, 2008)

The syngas exits the reactor at a temperature of about 1000°C (Higman & van der Burgt, Gasification Processes, 2008) and the carbon conversion rate and cold gas efficiency is reported to be >99.9% and 77% respectively (Mitsubishi Heavy Industries, 2009).

4.3.2.4. Kellogg Rust Westinghouse gasifier

The Kellogg Rust Westinghouse (KRW) gasifier is an agglomerating fluid bed gasifier, operating at a slightly higher temperature than conventional circulating fluidised bed gasifiers. The idea is to let the ash soften and fuse together into larger ash particles which become too heavy to stay in the fluidized

bed and will therefore fall to the bottom of the bed to be removed. The hope is to achieve a higher carbon conversion than in a circulating bed. (Higman & van der Burgt, Gasification Processes, 2008)

This kind of gasifier was operated in Piñon Pine, Nevada, but the plant had severe problems starting up correctly, which was mainly related to the hot gas clean-up. (DOE/NETL, 2002)

4.4. Shift reactors

A short explanation of the technical impacts of the two different shift reactor principles are presented in the following subchapter.

4.4.1. Sour shift

In a so called sour or raw gas shift reactor the catalysts are tolerant to sulphur in the incoming syngas, therefore requires no prior desulphurization of the syngas. The sour shift catalyst is based on cobalt-molybdenum and the catalyst also hydrolyses COS to H_2S. The catalyst actually requires sulphur to be operational and it is most often fed a gas saturated with water at around 250°C. (Higman, Gasification, 2008)

4.4.2. Clean shift

The clean shift reactor is extremely sensitive to sulphur and therefore requires a COS hydrolysis reactor and subsequent solvent scrubber to remove the sulphur prior to shift conversion. Additionally this means multiple stages of heating and cooling the gas in order to fulfil the required temperatures to enable an operational shift and sulphur removal train. (Higman, Gasification, 2008)

In a clean shift system the syngas first has to enter a COS hydrolysis reactor to transform all COS to the more easily removed compound H_2S. The syngas then has to be cooled prior to entering a H_2S removal system, after which the syngas enters the clean shift reactor, oxidizing CO to CO_2 and releasing heat, increasing the temperature of the syngas, which again has to be cooled down to enter the CO_2 capture unit.

4.5. Acid gas removal technologies

Acid gas removal systems consider the removal of sulphuric compounds and carbon dioxide from a syngas stream. Conventional sulphur cleaning of cold gas streams is treated in the following chapter called Flue gas cleanup.

4.5.1. Hot gas sulphur cleaning

For removal of sulphur from a hot syngas after gasification there are many metal oxides which each have of their own advantages and disadvantages. It should be noted that it is not a commercial technology as of yet.

One of the most popular metal oxides is the Zink oxide which exhibits the most favourable thermodynamic sorbent property of H_2S. Due to problems with vaporization and agglomeration at high temperatures the ZnO is limited to about 600°C as sulphur sorbent. Zinc titanate (ZT) and zinc ferrite ($ZnFe_2O_4$) are two zinc oxide based materials which has undergone extensive research. The oxides are further doped with other metals as copper (Cu) in order to enhance their sulphur capturing characteristics. It has been shown that zinc ferrite can remove H_2S from 4000 ppmv to less than 1 ppmv at 500°C in a simulated coal gasification gas. Recently, it has been shown that a zinc and copper oxide mix supported with Al_2O_3 has achieved a high degree of desulphurization by reducing 10 000 ppmv to less than 5 ppmv. (Meng, de Jong, Pal, & H.M. Verkooijen, 2010)

Even though hot gas desulphurization shows promising potential the regeneration of the sorbents are not yet fully understood. In a test based on zinc ferrite, 100 regeneration cycles were preformed and the capture capacity of the sorbent was reduced from 16 g to 3 g of sulphur adsorption per 100 grams of sorbent. (Meng, de Jong, Pal, & H.M. Verkooijen, 2010)

For an extensive summary of the literature regarding high temperature desulphurization a newly published article in the journal of Fuel Processing Technology is recommended. (Meng, de Jong, Pal, & H.M. Verkooijen, 2010)

4.5.2. Physical absorption systems

Rectisol is a methanol based physical solvent and has a typical operating range of -28 to -60°C resulting in high refrigeration duty. Figure 40 shows the Rectisol process in methanol production, removing both H_2S and COS and then CO_2. The raw incoming gas is cooled to -28°C by refrigeration and the exiting gas is free from H_2S and COS and let into a CO shift stage to transform all CO into CO_2. The CO_2 removal unit is split into two separate sections. First a CO_2 removal unit using flash regenerated methanol and then a fine CO_2 removal unit using hot regenerated methanol. The CO_2 removal unit operates at considerably lower temperatures, around 60°C. The Rectisol process can achieve extreme gas purities and is limited by cost and practical issues. (Higman, Gasification, 2008)
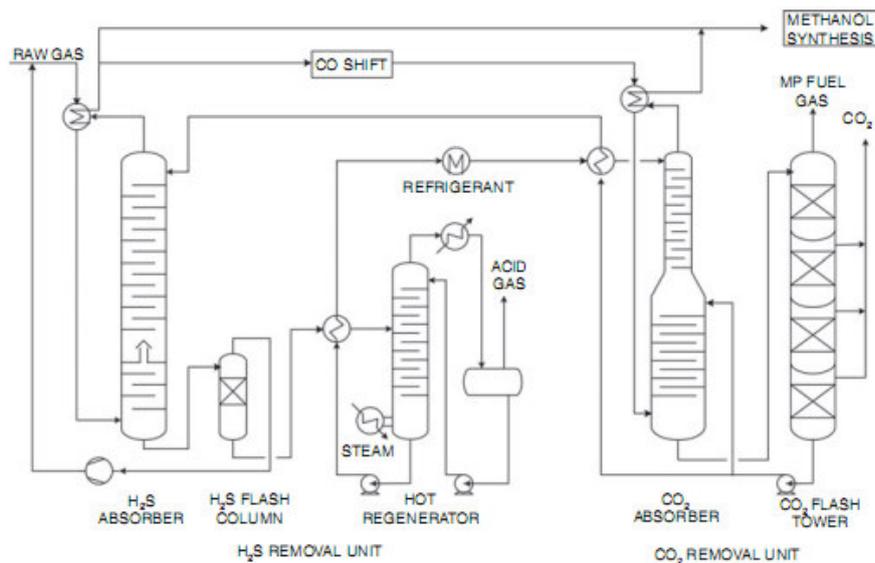
Figure 40. The Rectisol process (Higman, Gasification, 2008)

The physical absorption Selexol process operates at higher temperature than the Rectisol process, typically around 0 to 40°C resulting in substantially lower refrigeration load. Due to its lower selectivity to COS compared to H_2S it is desirable to convert COS to H_2S prior to the Selexol unit if sulphur removal is of importance. Otherwise the circulation rate of the solvent would have to be greatly increased to be able to reach low COS levels. In IGCC applications with CO_2 capture however a sour shift CO reactor also transforms COS to H_2S, which makes a Selexol process down streams very suitable. (Higman, Gasification, 2008)

When the Selexol or Rectisol process is used to capture H_2S, the H_2S stream is sent to an OxyClaus and tail gas treatment (TGT) unit, which together make up the sulphur recovery unit, SRU. In the

OxyClaus unit the H_2S is subjected to a partial oxidation process leading to the production of pure sulphur. A further description of the operation of a OxyClaus unit can be found in Gas Purification by Kohl and Nielsen (Kohl & Nielsen, Sulphur Recovery Processes, 1997).

4.6. Hot gas filters

To be able to withstand the high temperature (≈500°C) the candle filters are either made of ceramic material or metal. The ceramic filters are able to withstand the temperature without sintering while the metallic filters are more robust and less prone to rupture. (Higman & van der Burgt, Practical Issues, 2008)

4.6.1. Metallic filters

The higher thermal conductivity of metals opposed to ceramics means that the transient thermal shock effects introduced by back pulsing are limited. Additionally, the tensile strength of the metal means the filters can be made very thin, reducing pressure drop over the filters. Metal filters are however more prone to corrosion and this limits their operating temperature. (Sharma, o.a., 2008)

Metals filters based on iron aluminide (Fe_3Al) has been extensively tested at the Power Systems Development Facility (PSDF) in the United States. Tests were performed with coal derived syngas at a temperature of 400°C, and exposure times of 5780 hours were achieved for the most successful filters. Even though corrosion of the elements was indentified, no decrease in mechanical strength was reported. Overall the collection efficiencies are reported to be acceptable for gas turbine applications. (Guan, Gardner, Martin, & Spain, 2008)

4.6.2. Ceramic filters

Ceramic filters are able to withstand very high temperatures and efficiently clean a hot gas from particles. The problem is that they are very brittle and break. During back-pulse cleaning the fly ash, which is separated by the filters, is knocked off in a process where nitrogen is blast towards the filter. The sudden change of temperature leads to thermal chock and can crack the brittle ceramic material. Contrary to the metal filters, the ceramic filters need to be quite thick for robustness, increasing the pressure drop over the filter. A further limitation for the ceramic filters is that their metallic support elements cannot withstand the high temperatures, even though the filters themselves are able to. (Sharma, o.a., 2008)

In the IGCC demonstration plant in Värnamo ceramic filter failure was experienced due to mechanical failure in hot gas filtration at a temperature of about 350-400°C (Ståhl, Waldheim, Morris, Johansson, & Gårdmark, 2004). In Japan a ceramic candle filter produced and developed by Ashahi Glass Company and Mitsubishi Heavy Industries has been used in a 71 MW_e pressurised fluidized bed combustor: The flue gas had a temperature of 400°C and the longest lifetime of a filter was 8000 hours (Sasatsu, Misawa, Kobori, & Iritani, 2002).

4.7. Flue gas cleanup

Included in the flue gas cleanup topic in this context is aside the traditional sulphur cleaning also the removal of carbon dioxide in a post-combustion process.

4.7.1. Chemical absorption systems

The three main types of amine-based chemical absorbents are:

- Primary amine, usually MEA (monoethanolamine)
- Secondary, usually DEA (diethanol amine)
- Tertiary amine, usually MDEA (methyl diethanol amine)

MEA-based absorption has been commercialized used in the natural gas industry for over 60 years. The sorbent is normally regenerated in a stripper column with steam at a temperature of 100-200°C. There are some disadvantages with the MEA technology: the MEA solution is very corrosive and the solvent is degraded by SO_2, NO_2, HFC, HF and O_2. The MEA technology therefore requires the use of expensive corrosions resistant alloys and makeup of degradation of MEA (Olajire, 2010). In terms of corrosion, MDEA is less reactive than DEA, and DEA is less reactive than MEA (Higman & van der Burgt, Auxiliary Technologies, 2008).

Today the most widely used amine for H_2S absorption is MDEA, due to its selectivity over H_2S compared to CO_2. BASF (Badische Anilin- und Sodafabrik), the world's largest chemical company, has variation called activated MDEA (aMDEA) with accelerated CO_2 absorption, when selectivity for H_2S over CO_2 is not the primary concern. (Higman & van der Burgt, Auxiliary Technologies, 2008) Major suppliers within the field of amine-based sorbents are also Mitsubishi Heavy Industries (MHI) and Flour. (Olajire, 2010)

The chemical absorption ECO_2 process from Powerspan Corp. uses an ammonia-based solution designed as a post-combustion capture technology for traditional coal-fired power plants. The ECO_2 process can also be combined with Powerspan's ECO-SO_2 sulphur removal system. In Figure 41 the process flow of the ECO_2 process is shown in detail. (PowerSpan)
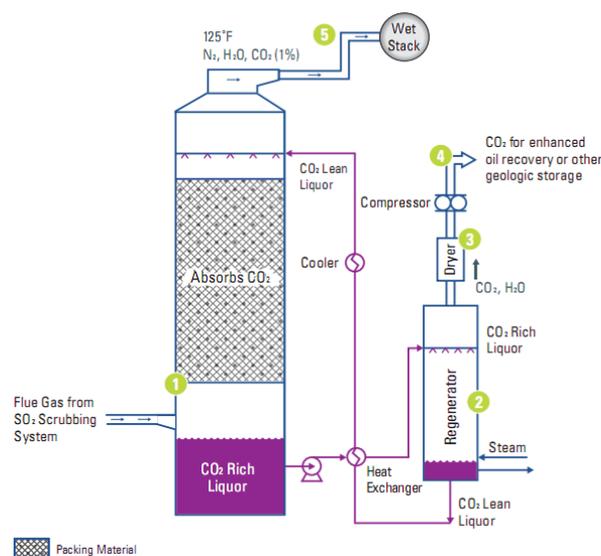
Figure 41. The ECO_2 process (PowerSpan)

The ECO$_2$ solution uses less energy than traditional amines and does therefore show promise for future applications in CO$_2$ capture. Table 5 shows some estimations of the energy requirement of the ECO$_2$ process.

Table 5. Heat demand estimations for the ECO$_2$ process

Source	Heat demand for regeneration	Steam pressure	CO$_2$ partial pressure
WorleyParsons Group Inc.(PowerSpan)	2326 kJ/kg CO$_2$	-	≈0.15
Vattenfall R&D (Bartlett & Simonsson, 2010)	2900 kJ/kg CO$_2$	1.2 bar	≈0.15

4.7.2. Sulphur cleaning

The most method to clean sulphur dioxide out of a flue gas stream is through the limestone/lime scrubbing process. A water based limestone/lime slurry is sprayed into the flow of flue gas in a scrubbing tower, the sulphur dioxide reacts with the aqueous solvent to from calcium sulphate and calcium sulphite. (Kohl & Nielsen, Sulphur Dioxide Removal, 1997)

More specifically, the limestone process with forced oxidation is the most common process. Limestone is and an inexpensive and efficient solvent and the system produce a sellable by-product, gypsum. Removal efficiencies in excess of 95% are possible with the limestone scrubbing process. A flow diagram of the process is shown in Figure 42. (Kohl & Nielsen, Sulphur Dioxide Removal, 1997)
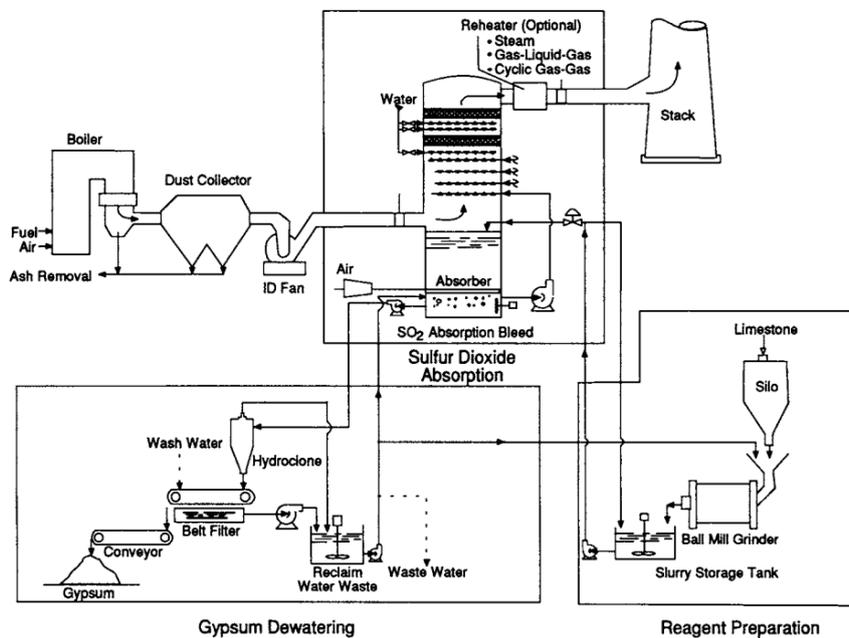
Figure 42. SO$_2$ removal by limestone process with forced oxidation (Kohl & Nielsen, Sulphur Dioxide Removal, 1997)

4.8. TopCycle concept

The main features of the TopCycle concept are utilization of a large amount of steam injection, to run the gas turbine at high pressures and to efficiently recover heat from the flue gas condensation. A standard gas turbine is used as a bottoming cycle and the combustion chamber is replaced with a very high pressure topping gas turbine cycle (TopSpool) increasing the pressure to >50 bar. (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

4.8.1. Gas turbines

The TopCycle unit is based for applications in the ≈100 MW_e size. This is due to the first generation design which is adapted to a commercial bottoming cycle of a certain size (Simonsson, Starfelt, Petersen, & Bartlett, 2010). The basic TopCycle layout is shown in Figure 43.
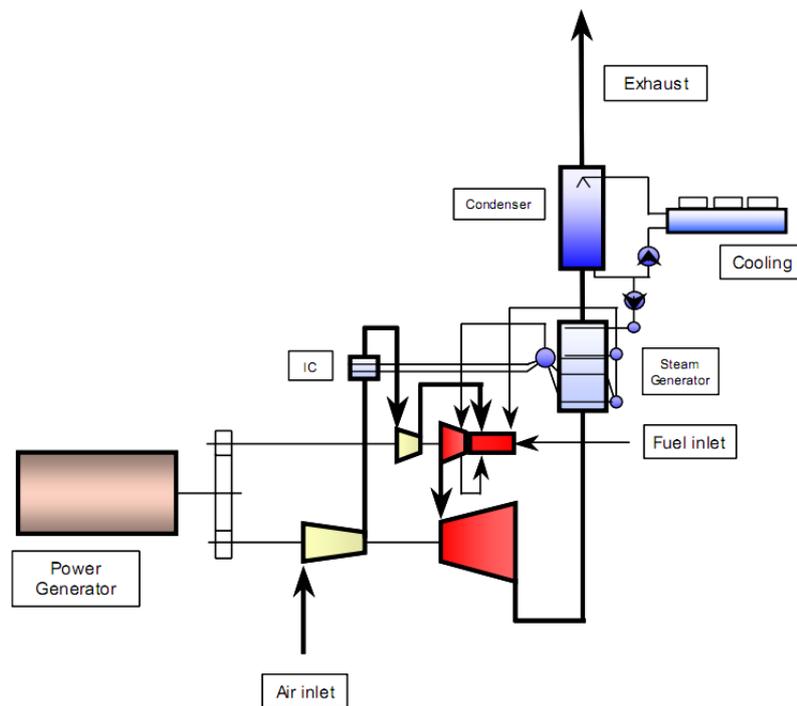
Figure 43. Schematic illustration of the TopCycle concept (Bruhn, Kretschman, & Simonsson, 2007)

Air enters the low pressure compressor, in which the air is compressed to approximately the same pressure as the selected base gas turbine originally was designed for. After the first compressor the hot air stream is intercooled with a heat exchanger, effectively lowering the air temperature, before entering the high pressure compressor. The heat recovered in the intercooler and from the gas turbine flue gas is used to raise high pressure steam which is injected into the combustion chamber.

In the high pressure combustion chamber the fuel is burned with a very low excess air ratio resulting in almost stochiometric conditions with the injected steam acting as coolant. After combustion the flue gases are expanded in the high and low pressure turbine stages. The low pressure turbine is air cooled with bleed-off air from the low pressure compressor, while the TopSpool turbine is cooled with steam. Steam cooling has two main advantages; the steam possesses good cooling properties and the use of steam also contributes to an increased specific work leading to increased efficiency. (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

4.8.2. Heat recovery

The heat recovery system consists of the intercooler and the single pressure HRSG system in the flue gas train. The feed water is evaporated in the intercooler and the evaporator of the HRSG. Some steam is used for blade and vane cooling in the TopSpool turbine, the rest is superheated in the HRSG and injected into the combustion chamber. Downstream of the HRSG the flue gas is cooled down in a flue gas condenser. The condensed water is recovered, treated to sufficient quality and led back to the HRSG. The TopCycle process can be self sufficient with water, or even produce water, as the water formed during combustion can be condensed.

4.8.3. Inlet humidifier

When additional heat is needed from the TopCycle unit there is a way to increase the available heat in the flue gases, by introducing a two-stage flue gas condenser as illustrated in Figure 44. In the first stage, water is sprayed counter current to the flue gas, condensing a portion of the flue gas water content, while being heated to near the flue gas dew point. In Figure 44 the return water flow from a district heating network is used as a heat sink, this heat sink is replaced by the CO_2-capture unit in a post-combustion configuration. In the second stage, heat is extracted to drive an atmospheric humidifier for increasing the humidity of the compressor intake air. This inlet humidification uses low temperature heat in the flue gas. To increase the water content in the inlet air and hence the flue gas dew point somewhat. Somewhat simplified it can be said that the gas turbine acts as a heat pump, allowing low temperature heat from the second FGC stage to become available at higher temperatures in the first stage of the flue gas condenser. (Simonsson, Starfelt, Petersen, & Bartlett, 2010)
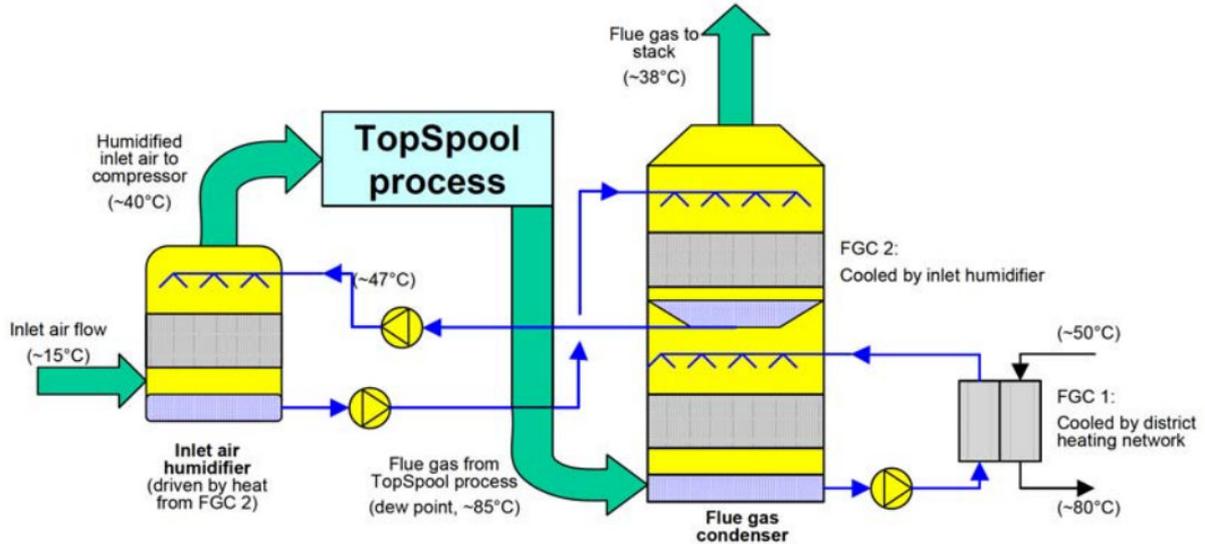
Figure 44. Principal layout and function of the inlet humidifier and two-stage flue gas condenser in a TopCycle process. Note that temperatures in the figure are only indicative (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

Although the total efficiency of the cycle will increase, the electrical efficiency of the cycle will be penalized by inlet humidification since the compressor work will increase. Therefore inlet humidification is only of interest in applications where low grade heat is needed for the overall process.

4.8.4. TopCycle concept when integrated with gasification

When applying the TopCycle concept together with drying and gasification of lignite and CO_2-capture there is a large difference in working principle when comparing pre- and post-combustion CO_2-capture.

The pre-combustion configuration means that the gas producing process - which involves gasification, gas clean-up and CO_2-separation - can be considered as a separate process. There will be some need to exchange heat between the TopCycle unit and the gas producing process but to a large extent this configuration is very similar to the one showed in Figure 45.
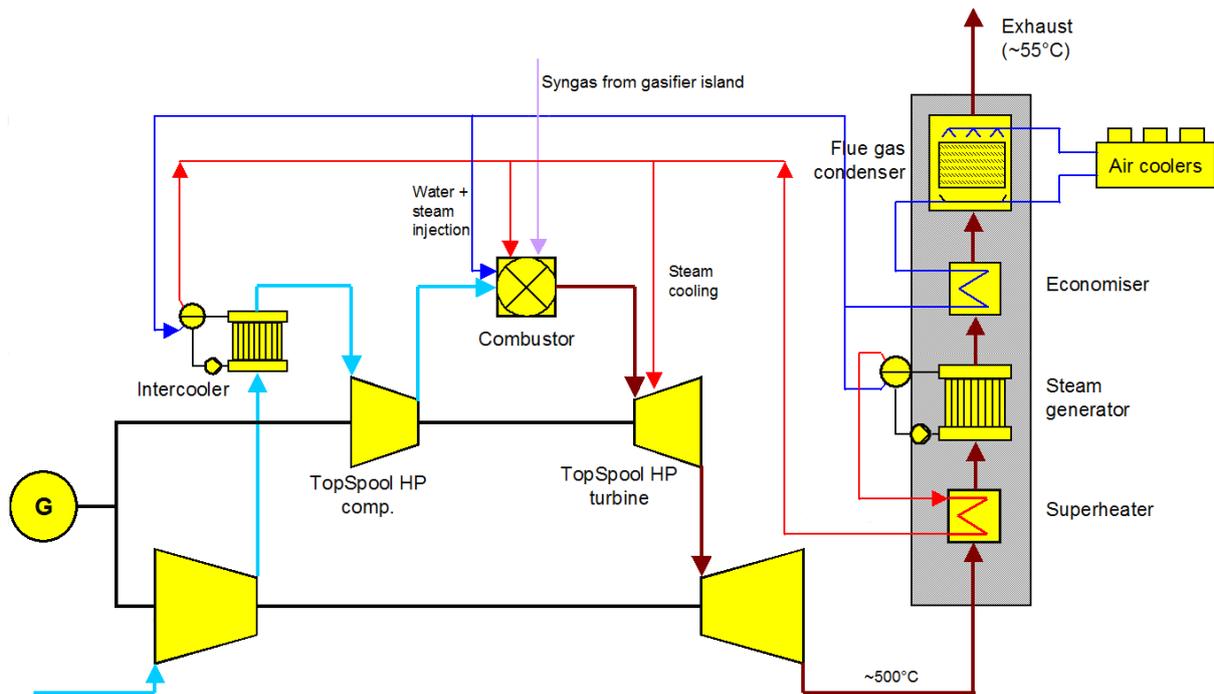
Figure 45. General TopCycle layout (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

When employing post-combustion CO_2-capture with air blown gasification however, there is will be a large need of low temperature heat to operate a post-combustion chemical CO_2-capture unit. The layout will change to require the previously discussed inlet humidifier and the TopCycle unit will be similar to that of Figure 46.
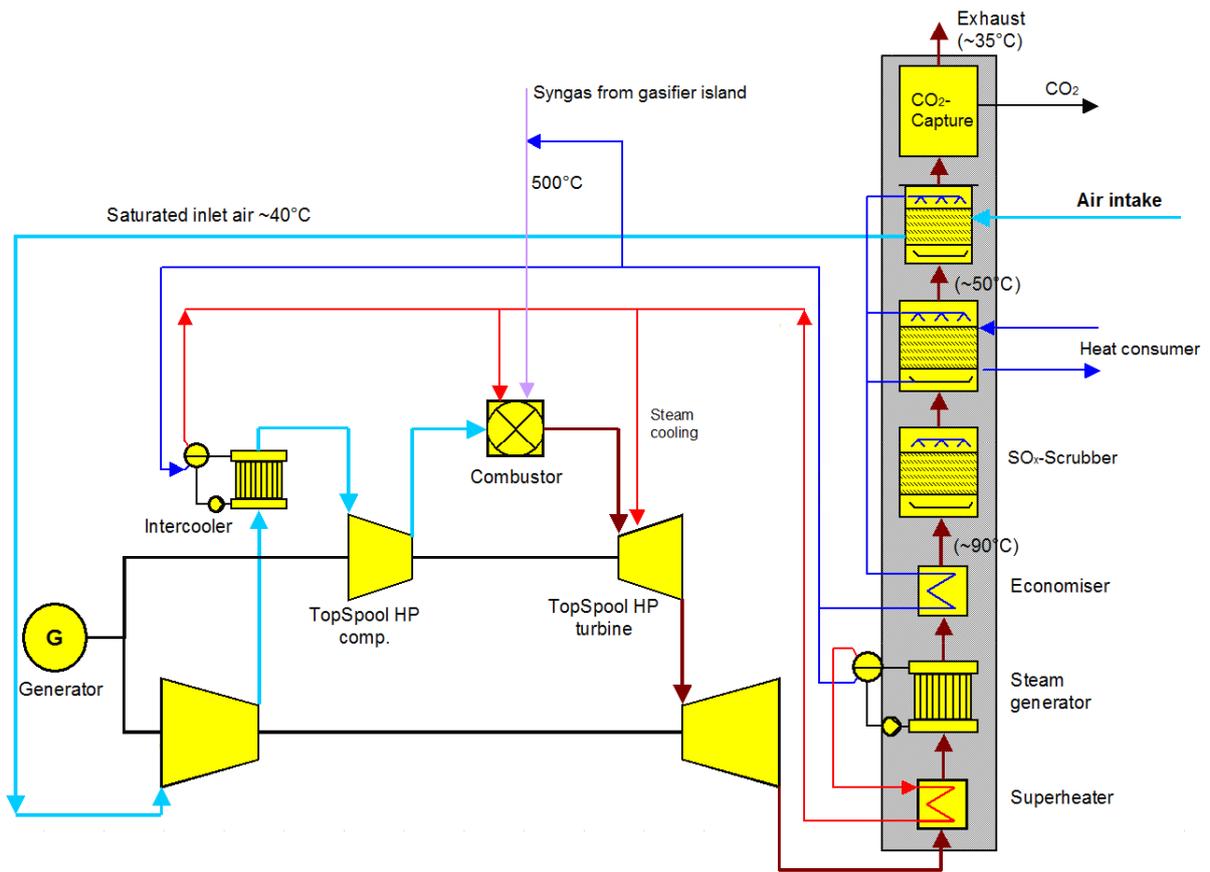
Figure 46. TopCycle with inlet humidification and CO_2-capture (Simonsson, Starfelt, Petersen, & Bartlett, 2010)

In the air blown case the incoming syngas, shown in Figure 46, is sprayed with to achieve a sufficiently low temperate of the syngas, to be able to apply hot gas clean-up before entry into the combustion chamber.

The inlet humidification increases the amount of moisture present in the air supplied to the combustor unit. This will mean that less water injection is needed to achieve the design condition, unique to the TopCycle unit, of very low excess O_2 in the combustion chamber. The result is that the more humidification applied to the system, the less steam injection to the combustor is needed for the TopCycle unit to reach the specific value of excess O_2. By superheating the steam fed into the combustor the energy recovered in the HRSG can still be made available for the TopCycle unit, when the amount of steam injection is limited by the excess O_2 design value. The superheating is however restricted by the temperature in the exiting flue gases and the required pinch temperature in the super heater.

5. Results

5.1.Main results

The main results of the simulations are shown in Table 6. The results from cases with CO_2 capture have a 90% carbon capture rate as default. A breakdown of the power consumers is also visualised in Figure 47.

The air blown IGTC with HTW without CO_2-capture is also included to show the difference of not having a CO_2-capture unit, and to highlight the potential of the configuration.

Table 6. Main results of simulations

		IGCC	IGTC Oxygen	IGTC HTW air				IGTC MHI air
		Siemens with Selexol	Siemens with Selexol	HTW with MEA	HTW with ECO2	HTW with ECO2 Combi	HTW without capture	MHI without capture
Key results	Fuel input, as received [MW]	1038.2	289.1	185.8	185.2	184.5	184.3	192.6
	Net power [MW]	386.5	107.2	72.8	80.2	86.5	99.5	107.2
	Net efficiency	**37.2%**	**37.1%**	**39.2%**	**43.3%**	**46.9%**	**54.0%**	**55.6%**
Power balance	Power produced [MW]	527.4	140.4	88.7	94.3	101.0	103.8	116.1
	Power consumed [MW]	140.9	33.2	4.3	15.9	14.1	14.5	8.9
	Net power [MW]	386.5	107.2	72.8	80.2	86.5	99.5	107.2
Main Power Consumers	ASU [MW]	28.8	8.1	N/A	N/A	N/A	N/A	1.4
	Dryer [MW]	22.4	6.2	1.3	1.3	1.3	1.3	1.4
	Gasifier [MW]	8.2	2.3	1.0	1.0	1.0	1.0	1.5
	CO_2-capture [MW]	21.1	5.1	1.4	2.2	2.2	N/A	N/A
	CO_2-compression [MW]	25.8	6	9.6	6.9	7.5	N/A	N/A
	O_2/N_2 compression [MW]	29.2	3.7	N/A	N/A	N/A	N/A	2.3
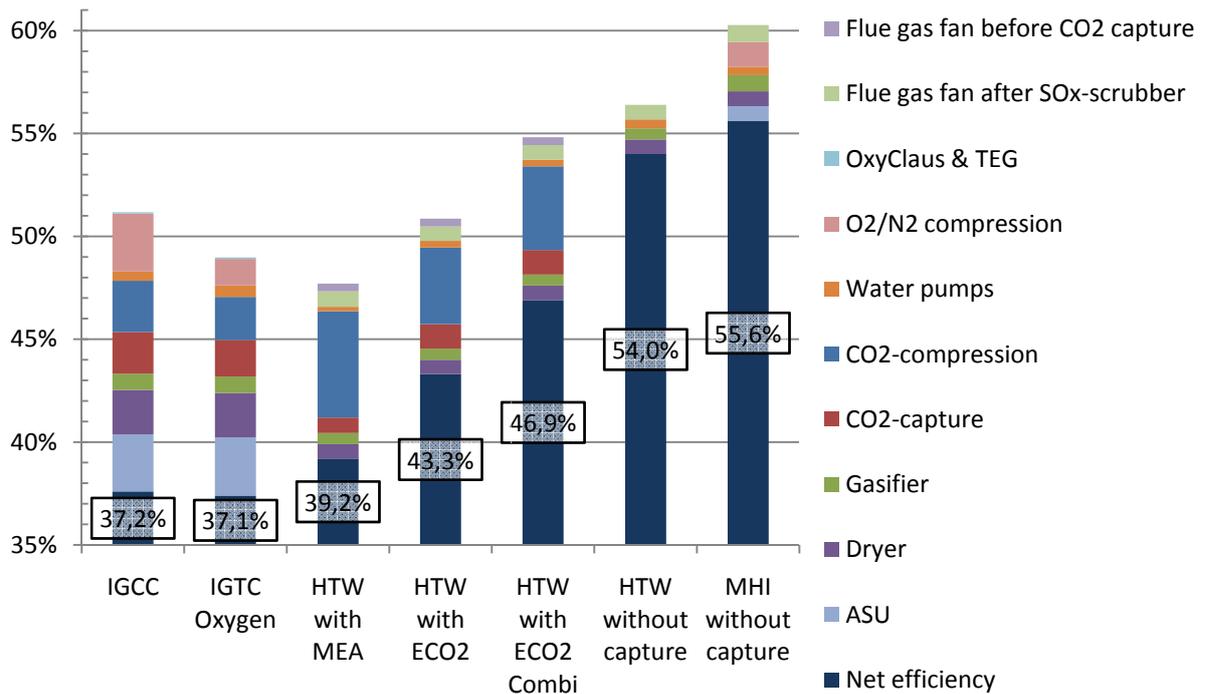
Figure 47. Efficiency loss breakdown. The total height of the bars indicate gross electrical efficiency and the numbers indicate net electrical efficiency.

IGTC oxygen blown and IGCC reference

The IGCC reference case achieves an efficiency of 37.2% in its base layout with CO_2-capture and lignite drying with the WTA technology. The IGTC oxygen blown case with CO_2-capture and WTA drying reaches an efficiency of 37.1%, which is slightly lower than for the IGCC reference case. Although most unit processes are identical, there are some differences.

The efficiency of power systems, combined cycle and the TopCycle unit, are slighty different. The syngas to power efficiency is higher with the combined cycle at 63%, while the IGTC configuration achieves 60%.

The higher nitrogen dilution of the oxygen IGTC case, due to the higher operating pressure of the gasifier, results in the need for more oxygen to reach the given temperature out of the gasifier, resulting in a larger ASU, related to the total fuel input. There is a major difference in the usage of the outgoing 2.38 bar nitrogen from the ASU. The IGCC reference case utilizes a very large part (95% mass) of the excess nitrogen, the nitrogen not needed in the gasification process, to dilute the syngas to achieve a LHV of 7.5 MJ/kg of the diluted syngas, as specified for the turbine. This nitrogen is however pressurized from 2.38 bar to 20.1 bar, resulting in very large compression duty "O_2/N_2-compression" for the IGCC reference case.

The Selexol process differs slightly because of the increased pressure of the syngas. Upon entering the Selexol process the syngas pressure is about 37 bar in the IGCC case whereas it is above 66 bar in the TopCycle case. Hence, the CO_2 partial pressure is higher in the latter case, making the Selexol process of the TopCycle case more efficient and therefore giving a slightly lower specific electrical and heat load. The increase in syngas pressure is, however, partly countered by the fact that the syngas in the TopCycle case is a little more diluted with nitrogen.

Furthermore the pressure of the CO_2 stream exiting the Selexol process and fed into the CO_2 compression process is increased by the ratio of the IGTC to IGCC operating pressure. This is a rough approximation to take into account the higher operating pressure of the Selexol process. In reality there are several streams exiting the Selexol process, each stream corresponding to the different pressures at which the Selexol solvent is regenerated in flash tanks.

IGTC air blown cases

The air blown concepts show great potential. The TopCycle unit with only gasification, drying and no CO_2-capture or compression exhibits efficiencies of 54.0% or 55.6% depending on configuration with HTW or MHI gasifier respectively. The higher efficiency for the MHI concept is derived from the higher carbon conversion and full utilization of raised steam. The HTW case produces excess steam (3.3 kg/s) which is left unused in the case with no CCS due to the oxygen fraction constraint in the combustor on volume basis. Any addition in steam injection to the combustor would lower the excess oxygen ratio below design level, resulting in an infeasible combustion process.

When introducing the CO_2-compression and CO_2-capture units the efficiency drops first by about 5-6%, due to the electrical consumption of these processes (at nominal 90% carbon capture rate) and an additional fan used to compensate for the additional pressure drop over the CO_2-capture unit. This is shown by the introduction of electric power for the CO_2-capture, CO_2-compression and flue gas fan before CO_2-capture, shown in Table 6. Furthermore, the TopCycle options with CO_2-capture have inlet humidification in order to supply the required heat demand to the CO_2-capture unit. Due to differentiating heat demands in the flue gas condenser (FGC) the cases are humidified to differing temperatures. The different humidification demands strongly influence the electrical output from the TopCycle unit, as is clearly seen in Table 6, ranging from 101 MW down to 89 MW. A more detailed review of the heat recovery, leading to the differentiating needs of inlet humidification, is presented in Appendix F.

5.2. Heat recovery characteristics

Due to the exceptional potential of the IGTC air blown cases and complex heat integration they are treated in more detail. A summary of the heat duties in the IGTC air blown cases is shown in Table 7.

Table 7. Heat recovery information for the IGTC air blown cases

		IGTC HTW air				IGTC MHI air
		MEA	ECO$_2$	ECO$_2$ Combi	No CO$_2$-capture	No CO$_2$-capture
	Fuel input	185.8	185.2	184.5	184.3	192.6
TopSpool heat exchangers	TopSpool IC [MW]	27.2	22.4	15.6	12.3	12.3
	TopSpool Evaporator [MW]	29.6	29.2	28.7	28.4	33.0
	TopSpool Super heater [MW]	2.8	4.5	6.4	7.1	6.2
	TopSpool Economizer [MW]	22.2	25.3	28.7	32.7	34.1
	Temp. of flue gas after economizer [°C]	179	157	131	105	111
	TopCycle electrical output [MW]	88.7	94.2	101	103.8	116.1
CO$_2$ stripper reboiler breakdown	Total duty [MW]	88.1	55	59.3	N/A	N/A
	From CO$_2$-compressor train [MW]	N/A	1.5	1.8	N/A	N/A
	From flue gas condenser [MW]	43.5	20.4	38.3	N/A	N/A
	From ash boiler [MW]	8.4	8.4	8.4	N/A	N/A
	From HP steam [MW]	36.2	25.0	10.8	N/A	N/A
Dryer breakdown	Total duty [MW]	34.9	34.8	34.6	34.6	36.2
	From flue gas condenser [MW]	28.5	34.8	34.6	34.6	36.2
	From CO2-compressor train [MW]	6.3	N/A	N/A	N/A	0
Humidification	Total duty [MW]	31.3	18.9	5.0	N/A	N/A
	Humidified air temperature [°C]	58.6	50.0	29.0	N/A	N/A
	Flue gas temperature, before FGC2 [°C]	68.5	71.4	62.2	N/A	N/A
	Global temperature driving force [°C]	9.9	21.4	33.2	N/A	N/A

The differences between the two cases without inlet humidification are not very large. Due to the higher carbon conversion and cold gas efficiency of the MHI gasifier, it supplies more fuel to the turbine and achieves a larger TopCycle output and overall a higher net efficiency as shown in Table 6. The HTW option as previously stated, is unable to utilise all steam generated, with a theoretical excess steam flow of 3.3 kg/s. The degree of superheat in the HRSG is limited by the temperature difference set at 25 K, i.e. there is not sufficient high temperature heat to decrease latent heat recovery and make the energy available to the TopSpool combustor in terms of sensible heat. The degree of superheat in the MHI case, however, is not maximized and the super heater has an upper terminal temperature difference of 136 K. But due to the higher steam flow in the MHI case the

absolute amount of heat exchange in the MHI option (6.2 MW) is quite similar to the HTW option (7.1 MW)

5.3.Gasifier, syngas and turbine characteristics

The resulting performance of the gasifiers can see in the table below.

Table 8. Gasifier characteristics

			IGTC HTW air				IGTC MHI air
	IGCC	IGTC Oxygen	MEA	ECO$_2$	ECO$_2$ Combi	No CO$_2$-capture	No CO$_2$-capture
Fuel input [MW]	1038.2	286.9	185.8	185.2	184.5	184.3	192.6
Dried fuel [MW]	1187.0	327.8	212.4	211.8	211.0	210.7	220.2
Cold gas efficiency (after particle removal)	80.8%	80.3%	81.2%	80.9%	80.6%	80.5%	86.0%
Gasifier Steam demand [kg/s]	N/A	N/A	1.0	2.1	3.3	3.6	N/A
Gasifier Quench water demand [kg/s]	81.1	22.1	N/A	N/A	N/A	N/A	N/A

The cold gas efficiency of the MHI gasifier is much higher than figures reported in literature of ≈77% (chapter 4.3.2.3, page 42). The simulation of the gasifier is however limited by the Ebsilon calculations and is a direct consequence of the assumed high carbon conversion, the air temperature from the compressor and the set outlet temperature of a moderate 950°C. Thermodynamically the cold gas efficiency is correct.

The steam demand for the fluidized bed HTW gasifiers changes as the inlet humidification is changed. With more inlet humidification the water content of the air supplied to the gasifier is increased, and to achieve the same amount of water content in the gasifier, which the simulation is based on, the steam demand of the gasifier has to decrease.

Information about the syngas composition prior to the gas turbine combustor is shown in Table 9.

Table 9. Syngas characteristics prior to gas turbine feeding

			IGTC HTW air				IGTC MHI air
Syngas into Combustor	IGCC	IGTC Oxygen	MEA	ECO$_2$	ECO$_2$ Combi	No CO$_2$-capture	No CO$_2$-capture
Temperature [°C]	70	160	500	500	500	500	500
Pressure [bar]	20.1	65.1	71.8	69.0	65.9	64.9	65.0
LHV [MJ/kg]	7.5	22.5	3.8	3.8	3.8	3.8	4.6
Cold gas efficiency to turbine	70.4%	70.5%	81.1%	80.8%	80.6%	80.5%	86.0%

The 10% cold gas efficiency drop for the oxygen blown IGCC and IGTC from the gasifier to the turbine is due to the exothermic sour shift reactors where steam is generated. This is a very important difference between the concepts of post- and pre-combustion CO$_2$ capture, which leads main difference in electrical efficiencies between the two concepts. The steam raised in the sour shift reactors is fed to the steam turbine and TopCycle combustor respectively, but does not yield the same power output as it would have as chemically bound energy in the syngas. It is easiest to realize

in the case of the combined cycle, where the raised steam from the sour shift reactor enters the steam turbine instead of the alternative, entering the gas turbine, generating work and then heating the fuel gas which generates steam in the HRSG which then enters the steam turbine.

The fuel to the gas turbine is very high in hydrogen for the oxygen blown IGTC concept (over 70%) although this can be moderated freely with steam injection to prevent flash back issues and lower flame speeds. Conversely, the air blown IGTC concepts have large quantities of steam from the water quench function and the low LHV may pose flammability issues.

The characteristics of the gas turbine is that for the HTW cases, a higher degree of humidification, leads to a decrease in steam injection to the combustor, while the flow from the compressor stages increase. The more the air is humidified, the more water enters the combustion chamber through the incoming air, and the less steam can be added to achieve the specific amount of excess oxygen after combustion. Because of the additional water in the inlet air the increased humidification leads to a lower net turbine output, because of the increase in compressor work. The gross turbine output however, does not vary as much.

More information about the syngas compositions is shown in Appendix F, together with flue gas emissions and turbine performance data.

5.4.Key sensitivities

The sensitivity analysis is divided into two separate parts for the IGCC oxygen and IGTC air blown options

IGTC oxygen blown

The sensitivity analysis for the IGTC oxygen case is set in relation to the IGCC reference case to highlight the characteristics of the TopCycle concept.

Table 10. Sensitivity analysis of the IGTC oxygen blown case, net electrical efficiency

	IGCC	IGTC Oxygen
Base case (with WTA dryer)	37.2%	37.1%
With low temperature dryer	37.6%	38.6%
Increased inert gas consumption to 150%	36.6%	36.1%
Decreased inert gas consumption to 50%	37.8%	38.0%

As shown in Table 10, when switching to the low temperature dryer for the IGTC oxygen blown and IGCC concepts the result is that the TopCycle benefits greatly. The difference is that the IGTC can supply waste heat from a flue gas condenser while the IGCC supplies low pressure steam to the low temperature dryer.

The changes in inert gas consumption influences the IGTC configuration the most because of the higher base consumption of inert gas.

IGTC air blown options

The HTW gasifier with atmospheric ECO_2 CO_2-capture at 90% capture rate is used as a base case for the initial sensitivity analysis.

The duty of the CO_2-capture and CO_2 compression units vary linearly with the amount of CO_2 captured. Therefore it is of little surprise that the efficiency of the IGTC HTW case varies linearly from 50% to 95% capture rate as shown in Table 11, and illustrated in Figure 48. The lower the capture rate the lower the heat demand is in the flue gas stream. A direct result is then that the humidification can be decreased and therefore the efficiency increases.

Table 11. Effects of varying CO_2-capture rate for the IGTC HTW, net electrical efficiency

Capture rate	Efficiency	Inlet humidification to
ECO_2 - 95%	42.5%	52°C
ECO_2 - 90%	43.3%	50°C
ECO_2 - 80%	44.9%	46°C
ECO_2 - 70%	46.6%	41°C
ECO_2 - 60%	48.3%	33°C
ECO_2 - 50%	50.0%	23°C
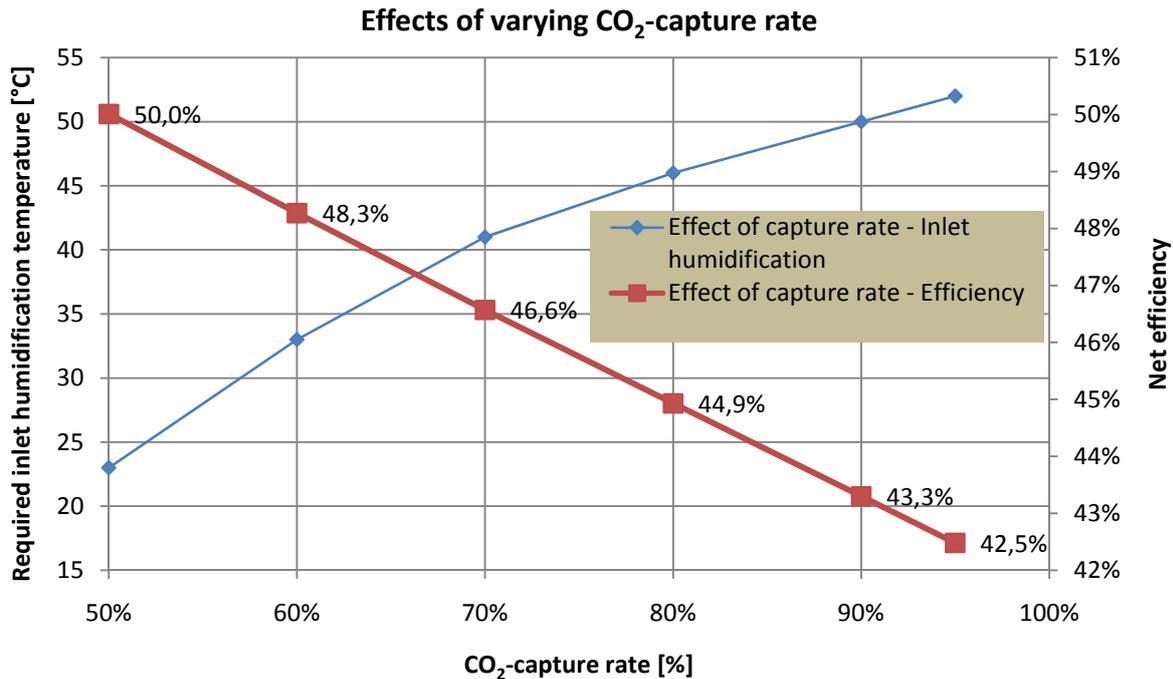
Figure 48. Effects of varying CO2-capture rate for the HTW gasifier with atmospheric ECO$_2$ as the base case

With the 90% capture rate as a base case some of most important parameters are varied. Table 12, summarizes the results of changes in temperature over the hot gas filter, increased moisture in the fuel fed to the gasifier, as well as changes in the inert gas dilution.

Table 12. Effects of key parameter changes for the IGTC HTW

Sensativity case	Efficiency	Inlet humidification temperature
Base case	43.3%	50°C
Hot gas filter -50 K (450°C)	42.3%	48°C
Hot gas filter +50 K (550°C)	44.2%	52°C
Fuel moisture 15%	42.4%	48°C
Inert gas need +50%	43.0%	50°C
Inert gas need -50%	43.6%	50°C

As can be seen in Table 12 the operating temperature of the hot gas filter is decisive for the overall plant efficiency. An increase in 50 K leads to an efficiency increase of 0.9%. However, the increase temperature allowance of the syngas entering the filter means that less water has to be injected prior to the hot gas filter to cool the syngas. This means that more steam can be added in the combustor to achieve the desired concentration of O$_2$ after the combustor unit. While this means more efficient heat recovery, the larger steam demand means there is less excess steam to use for the CO$_2$-capture unit, resulting in a larger need for humidification.

5.5. Generalisation of the air blown IGTC with CO_2-capture

A number of specific scenarios of the air blown IGTC post-combustion CO_2-capture configuration has been investigated. But to get an overview of the technology and its characteristics in a wider sense, a generalisation of the results for the post-combustion CO_2-capture configuration is performed.

The results are shown in the following three figures. The solid black lines in Figure 49 and Figure 50 represent the heat available in the flue gas at the specified temperature, varying with the inlet humidification temperature. With more inlet humidification, the heat available at a certain temperature increases, which results in the positive gradient of the parallel black lines.

The intersection of a red dashed line and a solid black line represents at which inlet humidification a feasible solution is achieved. For example if the CO_2-stripper requires 2.5 MJ/kg CO_2 at a temperature of 75°C with a 2 K pinch in the condenser, the heat available to the stripper is all the heat which follows the "77°C"-line. The lines intersect at around 45°C for the configuration with a low temperature dryer (Figure 49), which means the inlet air has to be humidified to 45°C in order to successfully operate the CO_2-stripper. With the inlet humidification known, the efficiency can be estimated from the last figure, Figure 51 , to about 44.5%.

When the temperature of the flue gas in Figure 49 drops below 77°C the heat required for the low temperature dryer comes into play. The low temperature dryer is taken into account with identical characteristics as in previous simulations, requiring a temperature of 75°C with a 2 K pinch and supplying a water temperature of 54°C to the flue gas condenser. Therefore the distance between the black temperature lines decrease when the flue gas temperature drops below 77°C in Figure 49, there is simply less heat available beyond this point, due to the low temperature dryer.
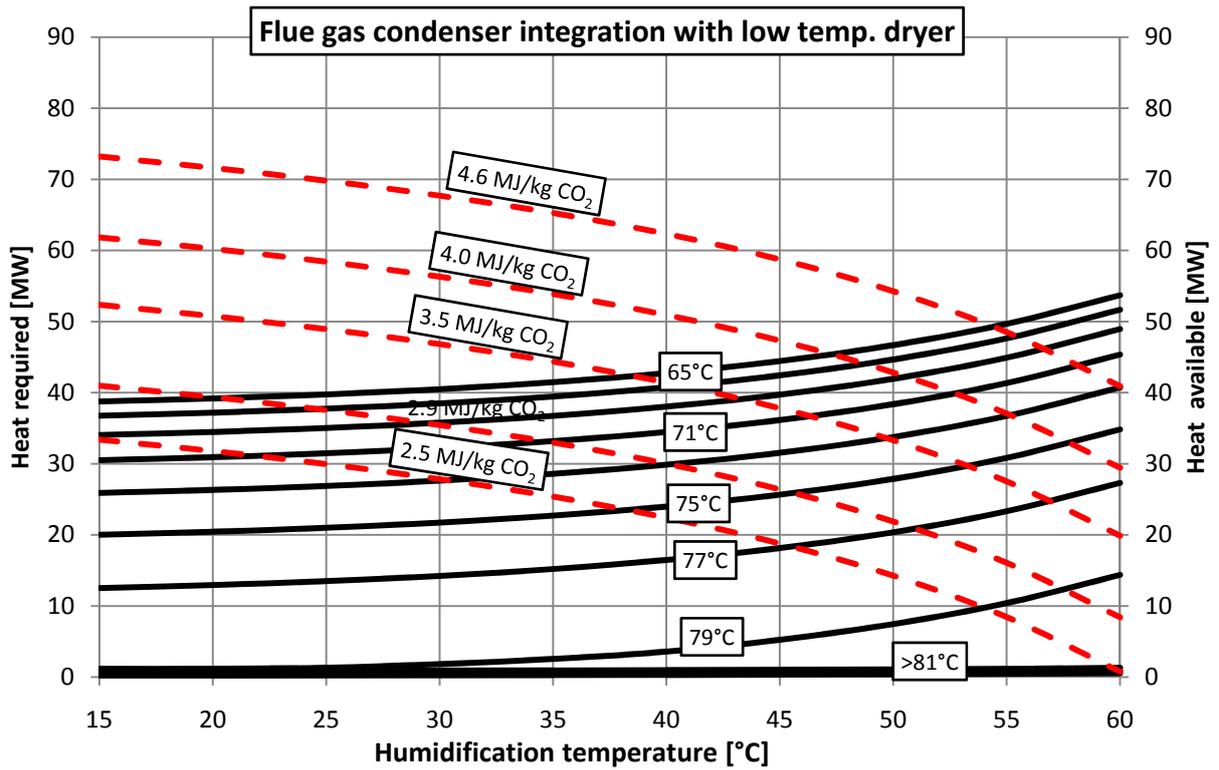
Figure 49. Flue gas condenser characteristic of air blown IGTC post-combustion CO_2-capture with low temperature dryer
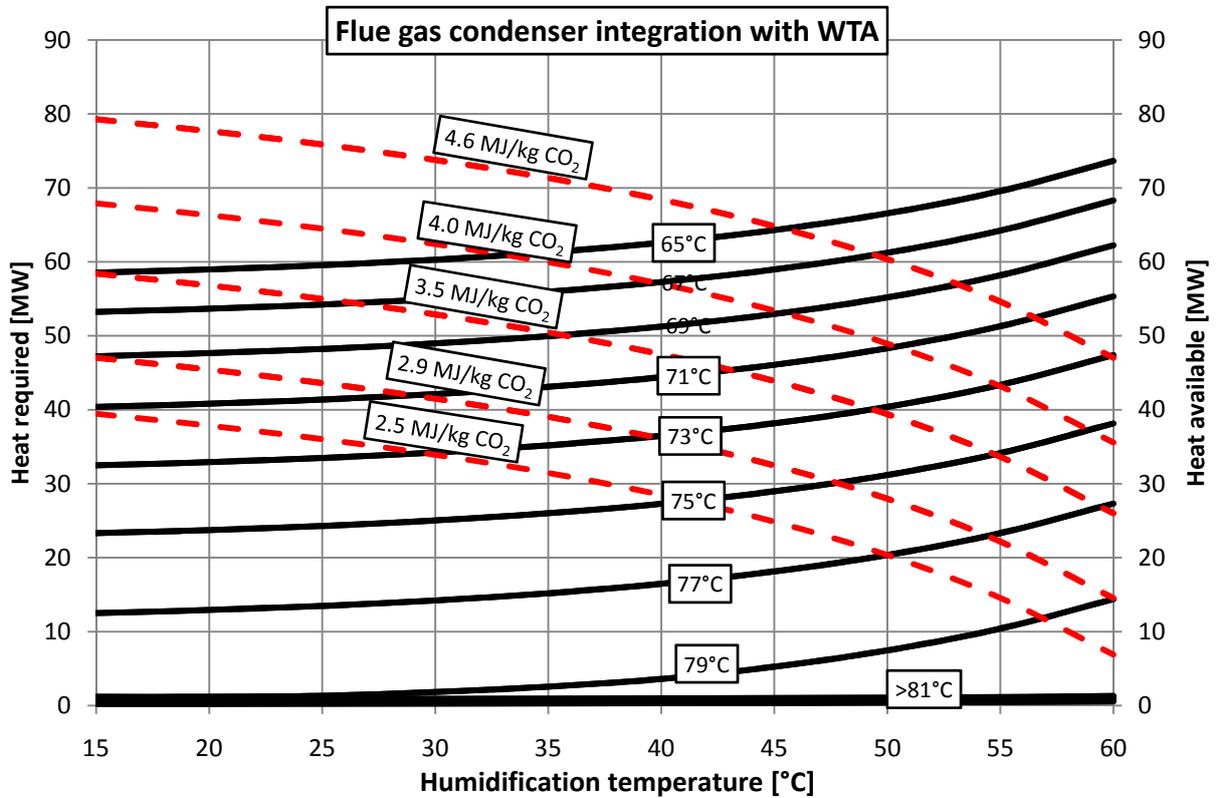
Figure 50. Flue gas condenser characteristic of air blown IGTC post-combustion CO_2-capture with WTA dryer
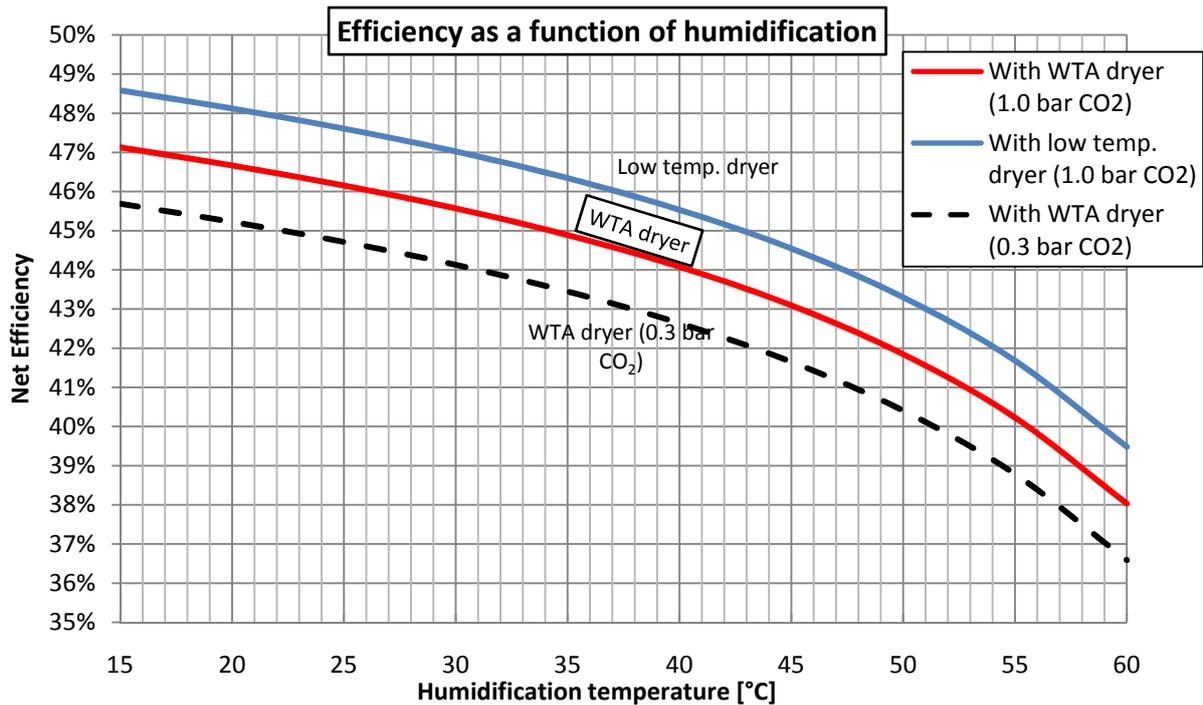
Figure 51. Net efficiency as a function of inlet humidification temperature

In a similar fashion, Figure 50 represents the characteristics of the post-combustion CO_2-capture IGTC system with the WTA dryer. The difference is that no flue gas heat is required to operate a low temperature dryer and that steam is needed to operate the WTA dryer. The result is that the red dashed lines, representing the required heat, are pushed to a higher starting position - while at the same time - more heat is available in the flue gas.

Figure 51 shows the net electrical efficiency as a function of the inlet humidification for the post-combustion CO_2 capture with either the low temperature dryer or the WTA dryer. Integrated into Figure 51 is also the efficiency of the IGTC system with a special case, showing the impact of a lower pressure discharge of CO_2 from the CO_2-separation unit, leading to higher compression work, a kind of "worst case" scenario.

The influence of the condenser pinch temperature and required temperature of post-combustion CO_2-capture process becomes clear when looking through Figure 49, Figure 50 and Figure 51. Small changes in specific heat requirements and temperature has significant impact on the overall efficiency.

Larger copies of Figure 49, Figure 50 and Figure 51 can be found in Appendix C, and a more detailed explanation of the generalization can be found in Appendix F.

6. Discussion

6.1. IGTC - Post vs. pre-combustion CO_2-capture

There simulations have shown that there is large difference between the two concepts with IGTC. The pre-combustion CO_2-capture concept achieves efficiencies comparable to more conventional IGCC plants with pre-combustion CO_2-capture, in the region of 37%. The post-combustion concepts reach efficiencies up to 49% with CO_2-capture, and 56% without CO_2-capture, which is impressive.

The conversion efficiency from moist fuel to combustible syngas into the turbine have been shown to differ by about 10% pt between the two concepts and is due to the exothermic shift reactor needed in the pre-combustion configuration. The syngas is able to enter the combustion chamber at 500°C in the post-combustion case while as much heat as possible is recovered in the pre-combustion case. The reason to the lower temperature of the syngas is due to the pre-combustion CO_2-capture unit which requires cooling of the syngas before CO_2-capture, which results in a loss of sensible heat.

The results for the air blown post-combustion CO_2-capture configuration however depend on a number of processes which are not yet commercial. The nitrogen needed for feeding the fuel into the gasifier is also not accounted for in the air blown IGTC configurations with CO_2-capture, it is considered as supplied from an external source.

6.2. IGTC Oxygen blown

It has been shown that the TopCycle unit is an alternative to the integrated gasification combined cycle when it comes to CO_2-capture applications. The pre-combustion configuration with IGTC achieves efficiencies similar to that of an IGCC (37%). Additionally the TopCycle unit is based on a relatively small gas turbine, being disfavoured by the smaller scale means that the TopCycle unit has lower efficiencies for gas turbine components and generator. Considering the unit on a large scale would probably increase the efficiency of TopCycle specific components.

The IGTC system utilizes a single pressure heat recovery steam generator and does not need a steam turbine, which should mean lower costs. Further, the power density of the gas turbine is strongly increased. However, the high pressure gasification is the large uncertainty in the IGTC case. It needs to be proved on a commercial scale that high pressure gasification is possible. An alternative approach could be to compress the syngas, to still benefit from the potential cost saving due to single pressure HRSG and absence of steam turbine. This scenario seems beneficial in a development stage of the TopCycle concept.

The oxygen blown IGTC is more proven than the air blown cases. The TopCycle unit in itself is still on the drawing board but other than that the oxygen blown case has its major technological drawback in unproven high pressure gasification.

6.3. IGTC Air blown

Simulations have shown that the IGTC air blown plant concept has great potential, reaching efficiencies of 54-56% without CO_2-capture. After extensive concept evaluation it is clear that the humidification of the inlet air is a powerful tool for the IGTC concept to be able to adapt the core process, the TopCycle unit, to fit the requirements of ancillary services. With CO_2-capture, the air blown IGTC is able to reach efficiencies up to 49 %. The study should be seen as a study of future potential as no techno-economic optimisations were performed and many technologies were

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included that are not commercial. The efficiencies found, therefore, represent the driving force for a series of development & commercialisation steps, in particular:

- High pressure air gasification

- Hot syngas cleanup (particles, metals and potentially sulphur)

- Flue gas CO_2 capture units driven with low temperature heat

- Large scale lignite dryers driven with low temperature heat

The air blown IGTC cases with hot gas cleaning can seem to be quite simple configurations. However, the air integration with the TopCycle compressor, supplying compressed air to the gasifier, means the power and gasifier island are more integrated than in the oxygen blown concept.

The uncertainties regarding the post-combustion CO_2-capture unit are large. This is critical due to the strong dependency on the temperature and heat requirements of the CO_2-capture unit for the overall performance of air blown IGTC. The simulations preformed have showed that for a beneficial CO_2-capture unit, able to utilize low temperature heat (ECO_2 Combination), the IGTC can reach efficiencies of 49% while simulations show efficiencies of 39% if the heat demand is at higher temperatures or of larger quantities. Strongly connected to the plant performance and cost-effectiveness is the performance of the low temperature heat exchangers. In these simulations, a low pinch of 2 K is assumed to maximise the temperatures available. This will drive up both the flue gas condenser and stripper reboiler surfaces and does most likely not represent an economic optimum.

Due to the large amount of nitrogen present in the air blown cases the syngas is of very low heating value (≈3.8 MJ/kg). This might be a problem during start-up or even to get a stable flame in the TopCycle combustor. An alternative to increase the heating value of the gas slightly would be to use a syngas cooler after the gasifier to raise steam instead of injecting water to cool the syngas prior to the hot gas filter. This would however mean that the syngas cooler would be subjected to a very corrosive environment probably leading to operability issues and high costs.

Operating a low temperature dryer would require large amounts of space and may have operability issues. The WTA dryer on the other hand has been proved to be possible to integrate with the air blown TopCycle concept, providing the CO_2-capture unit has a low heat demand. This seems like a very promising option and should be investigated further.

Due to relatively low carbon conversion of the HTW gasifier the ash may contain enough unburned coal to not be classified as normal disposable ash. Therefore it might be needed to completely burn the ashes in a boiler, as simulated, or to be forced to pay to dispose of the unburned material and ashes in an acceptable way.

The simulated TopCycle concepts are all based on a small scale model. If the TopCycle unit would be scaled up the efficiencies of the individual components such as the compressor and turbine stages as well as the generator would increase, increasing the competitiveness of the concept, this should not be forgotten when comparing the efficiencies reached with the TopCycle concept with other CO_2-capture concepts.

6.4. Error analysis

The simulations performed depend on a large number of assumptions and simplifications. As shown in the sensitivity analysis small changes in the pinch temperature in the flue gas condenser or changes in the amount of nitrogen dilution during gasification can have substantial impact on the final result.

The software, Ebsilon professional, is not developed for evaluating chemical units as shift reactors or gasifiers. The simulations of these process units have therefore been extra uncertain as it has not been possible to simulate the units to more than black-box models, even though the heat and mass balance is maintained.

Some components have been especially problematic. The MHI gasifier, for example, has been modelled on the basis of a very limited amount of data and is therefore only considered without CO_2-capture. The inlet humidifier for the post-combustion cases is another component which has been greatly simplified. The performance of the humidifier and the limits of its operation has however been double-checked with an excel-model available at Vattenfall Research & Development.

The compressor stage, compressing the captured CO_2 from the CO_2-capture unit up to 110 bars is a rough model and needs refining.

7. Conclusions

This thesis has shown that TopCycle integrated with gasification (IGTC) of lignite and CO_2 capture is a promising concept with the potential for very high efficiencies. When using a 100 MW TopCycle unit as a building block, the configuration with an oxygen-blown gasifier and pre-combustion capture has an efficiency of 37% while the configuration with an air blown gasifier and post-combustion capture is 39-49%. The latter figure is highly dependent on the use of low temperature heat from the power island. The most efficient IGTC systems include a low temp dryer, a post-combustion CO_2-capture unit able to utilize low temperature heat (<69°C) with a low specific heat demand (2.5 MJ/kg CO_2) and a hot syngas cleanup system. As a reference, a 500 MW combined cycle system with pre-combustion capture and oxygen blown gasifier reaches efficiencies of 37% according to the simulations preformed in this thesis.

The main difference between the IGTC cases with post- and pre-combustion CO_2 capture lies in the treatment of the syngas after the gasifer. Due to losses in heating value of the gas, related to the use of a exothermic shift reactor, and loss of sensible heat due to restrictions regarding syngas cleanup, the pre-combustion configuration delivers a syngas with lower heating value to the power generating TopCycle unit.

The difference between the simulated pre-combustion configurations of the reference IGCC and IGTC are marginal in terms of net electrical efficiency. The main reason is that a combined cycle configuration and TopCycle unit achieves similar efficiencies when supplied with the same fuel into the gas turbine unit. This fact, combined with almost identical layouts of ancillary equipment, leads - unsurprisingly - to two system with very similar net electrical efficiencies.

The efficiency of the promising IGTC concept with post-combustion CO_2-capture depends to a large degree on the efficiency of the post-combustion CO_2-capture unit. For a good effectiveness, the capture process has to be able to utilise low temperature heat from the flue gas condenser (<79°C). Inlet humidification is a useful tool to modify the amount of heat available and needed from the flue gas condenser. Increased air humidification leads to lower electrical efficiencies but at the same time more process steam and heat in the flue gas condenser is made available to drive a CO_2 capture unit.

The post-combustion CO_2-capture IGTC configuration with air blown gasification achieves unmatched efficiencies with CO_2-capture. The deciding factor for the potential of the TopCycle concept is concluded to be the heat requirements of matching CO_2-capture units. Further research in regards of heat requirements of post-combustion CO_2-capture technologies are crucial if the TopCycle concept is to be considered as a future alternative power production technology.

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Water treatment

Scour shift 2

Selexol CO2 +H2S separation unit

Claus unit
TGT
Elec. cons.=0.20 MW
Sulphur produced=13.25 t/d

CO2 compression

Humidifier

Nitrogen

TopCycle

Syngas

Steam injection

CMB

DEAL

FGC 1

ECO

HPC

HPT

G

EVAP

IC

LPC

Air

LPT

G
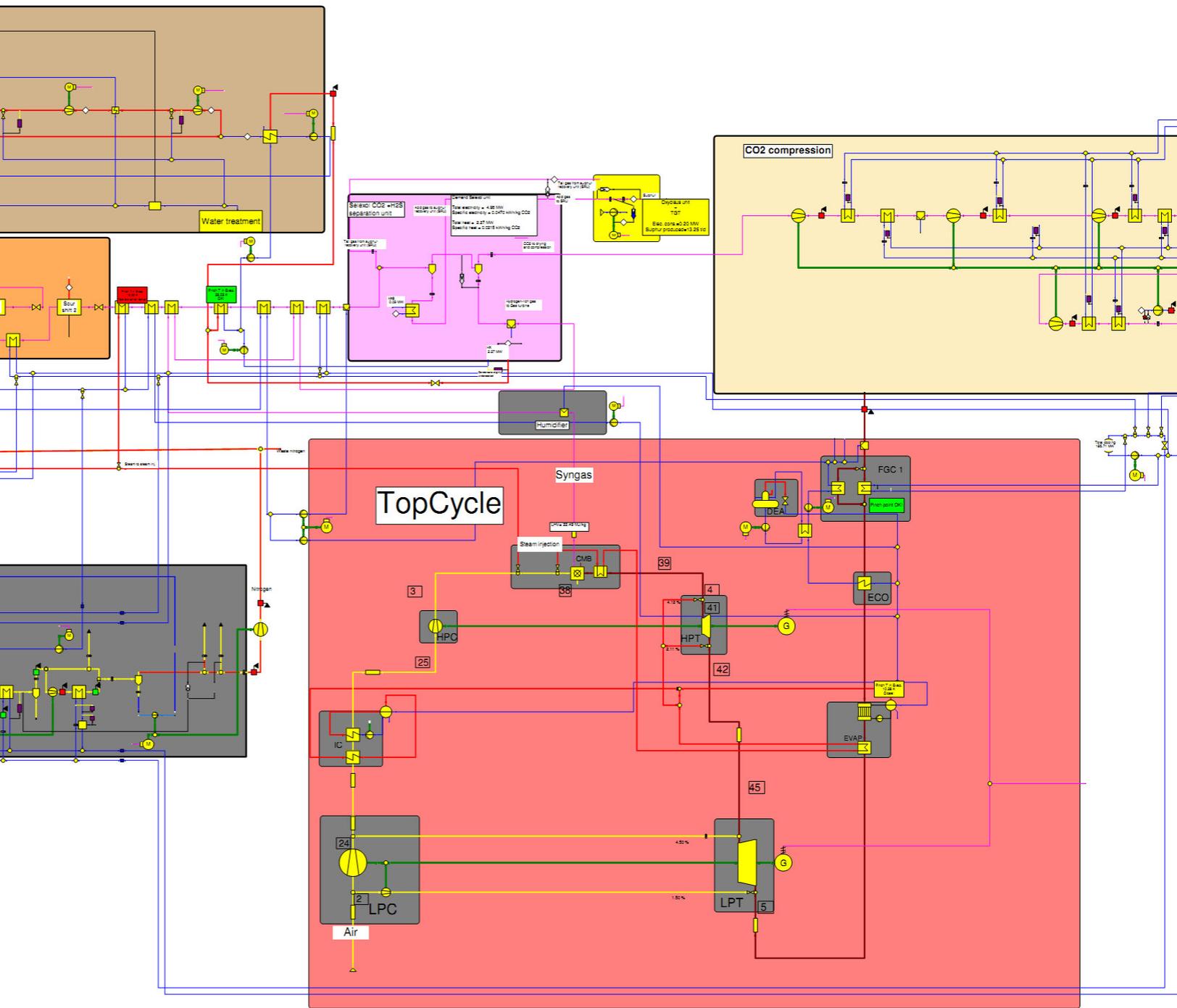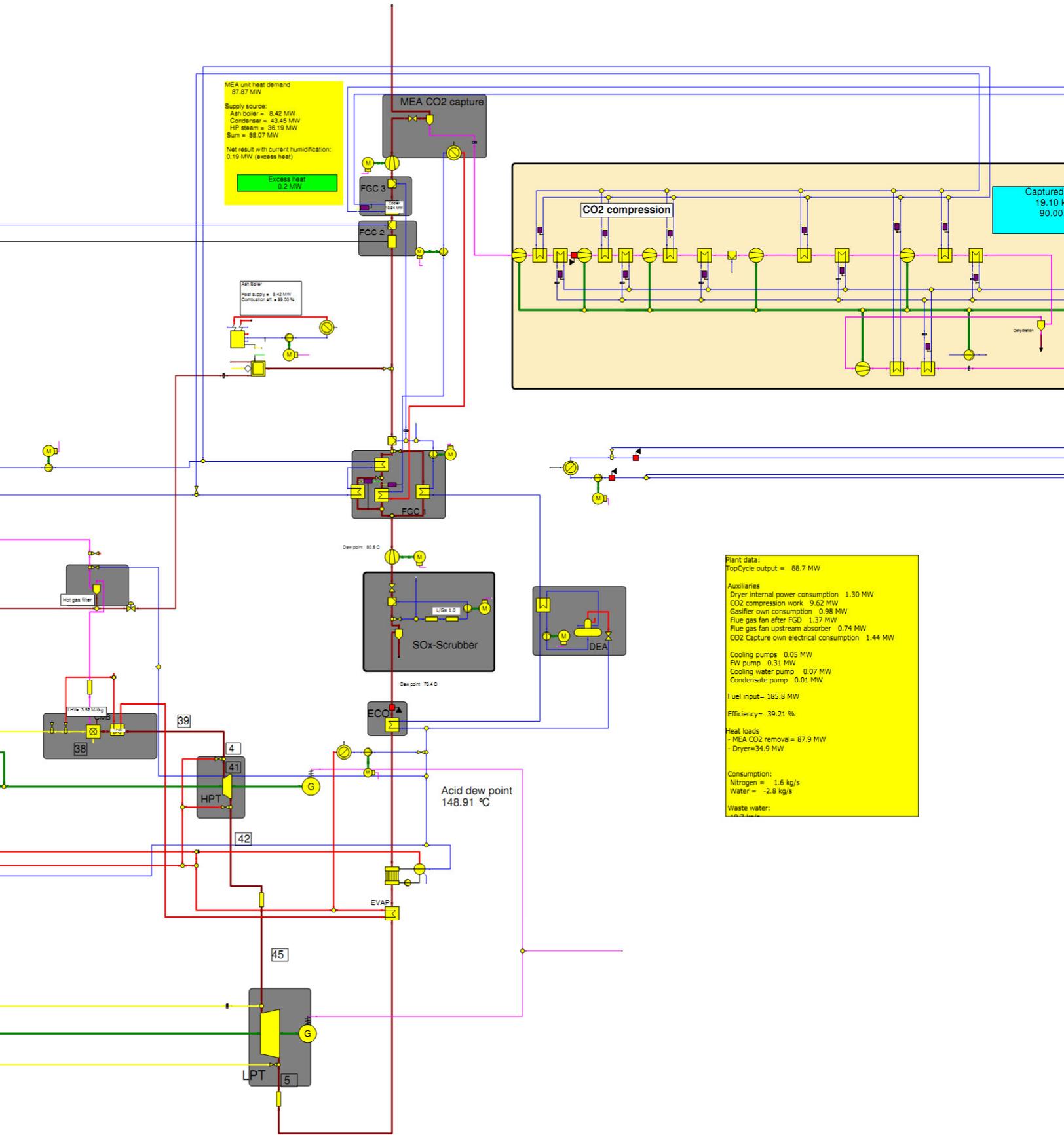
MEA unit heat demand
87.87 MW

Supply source:
Ash boiler = 8.42 MW
Condenser = 43.45 MW
HP steam = 36.19 MW
Sum = 88.07 MW

Net result with current humidification:
0.19 MW (excess heat)

Excess heat
0.2 MW

MEA CO2 capture

FGC 3

Cooler

FCC 2

CO2 compression

Captured
19.10 k
90.00

Dehydration

Ash Boiler

Heat supply = 8.42 MW
Combustion eff = 99.00 %

FGC 1

Dew point 80.5 C

SOx-Scrubber

L/G= 1.0

DEA

Hot gas filter

LHV= 3.82 MJ/kg

39

38

4

41

HPT

42

Dew point 78.4 C

ECO

Acid dew point
148.91 ℃

Plant data:
TopCycle output =   88.7 MW

Auxiliaries
Dryer internal power consumption   1.30 MW
CO2 compression work   9.62 MW
Gasifier own consumption   0.98 MW
Flue gas fan after FGD   1.37 MW
Flue gas fan upstream absorber   0.74 MW
CO2 Capture own electrical consumption   1.44 MW

Cooling pumps   0.05 MW
FW pump   0.31 MW
Cooling water pump   0.07 MW
Condensate pump   0.01 MW

Fuel input= 185.8 MW

Efficiency= 39.21 %

Heat loads
- MEA CO2 removal= 87.9 MW
- Dryer=34.9 MW

Consumption:
Nitrogen =   1.6 kg/s
Water = -2.8 kg/s

Waste water:

EVAP

45

LPT

5

G

Atmospheric ECO2

CO2 compression

Captured
19.05
90.00

FGC 3

FGC 2

Ash Boiler
Heat supply = 8.39 MW
Combustion eff. = 99.00 %

Heat from Ash boiler
8.40 MW

Dehydration

Total cooling
55.32 MW

FGC 1

Dew point 79.8 C

Hot gas filter

Syngas

SOx-Scrubber

L/G= 1.0

DEA

LHWe 3.60 MJ/kg

CMB

39

38

Dew point 78.1 C

ECO

ECO
25.32 MW

4

41

4.12

HPT

G

42

Acid dew point
148.75 °C

EVAP

45

LPT

G

5
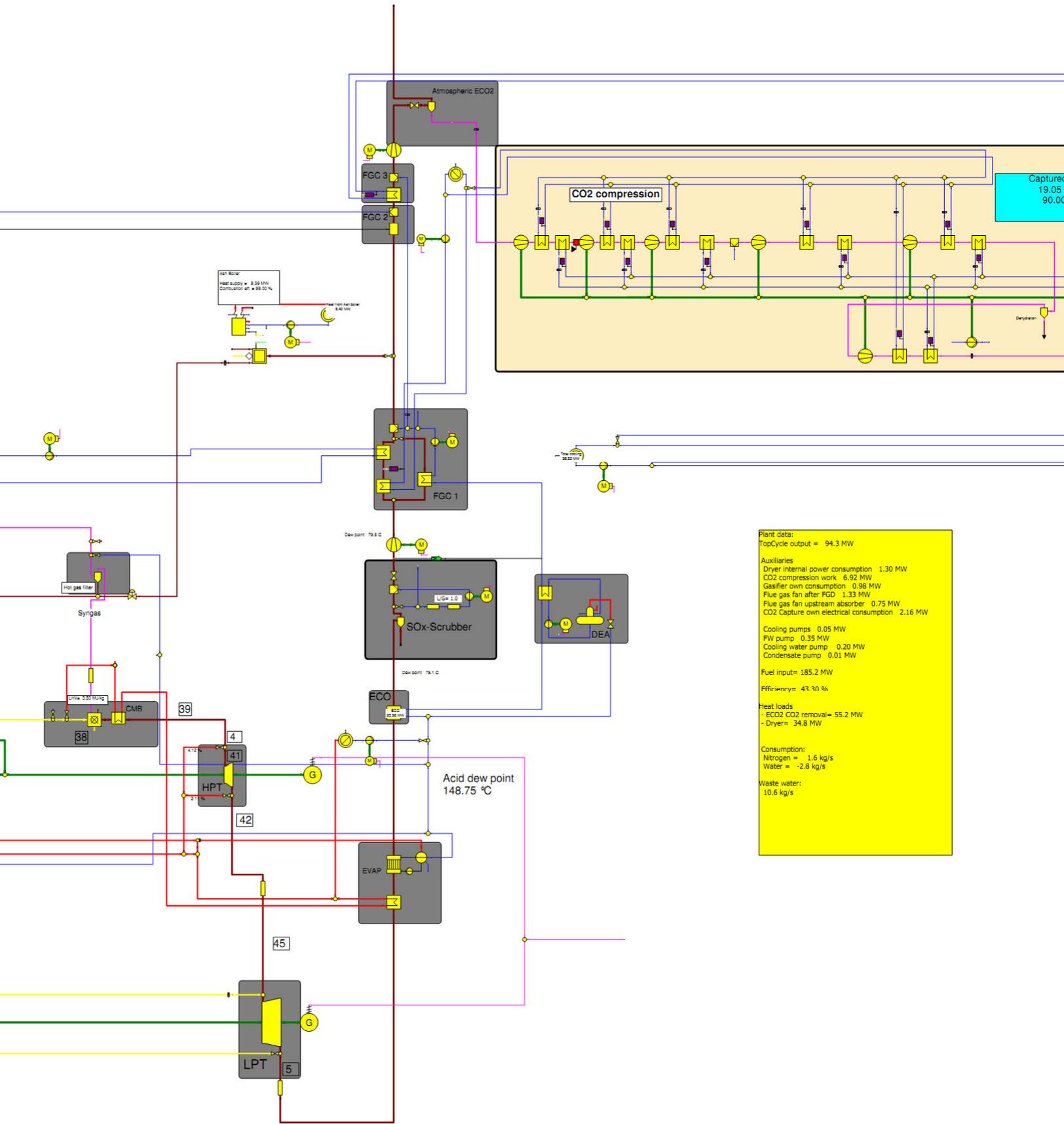
Plant data:
TopCycle output =   94.3 MW

Auxiliaries
 Dryer internal power consumption   1.30 MW
 CO2 compression work   6.92 MW
 Gasifier own consumption   0.98 MW
 Flue gas fan after FGD   1.33 MW
 Flue gas fan upstream absorber   0.75 MW
 CO2 Capture own electrical consumption   2.16 MW

 Cooling pumps   0.05 MW
 FW pump   0.35 MW
 Cooling water pump   0.20 MW
 Condensate pump   0.01 MW

Fuel input= 185.2 MW

Efficiency= 43.30 %

Heat loads
- ECO2 CO2 removal= 55.2 MW
- Dryer= 34.8 MW

Consumption:
 Nitrogen =   1.6 kg/s
 Water =   -2.8 kg/s

Waste water:
 10.6 kg/s

ECO2 combination of atmospheric
and sub-atmospheric stripper

Captured CO
18.97 kg/s
90.00 %

FGC 3

FGC 2

CO2 compression

Dehydration

Ash Boiler
Heat supply = 3.35 MW
Combustion eff. = 99.00 %

el from Ash bo
5.37 MW

Hot gas filter

Syngas

FGC 1

Dew point 79.0 C

SOx-Scrubber

L/G= 1.0

DEA

Dew point 77.7 C

LHVs 3.78 MJ/kg
DR B

38

39

ECO
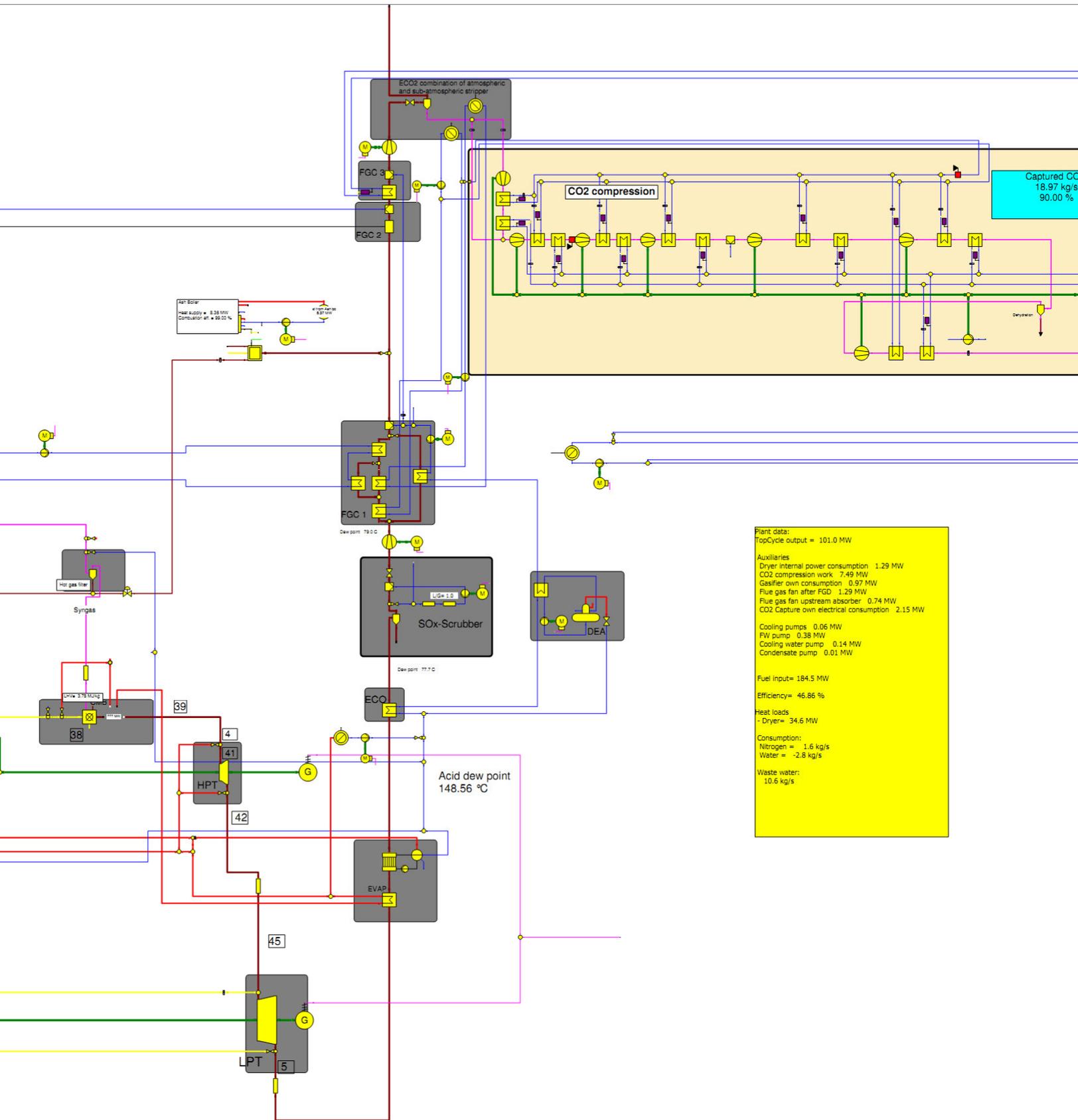
Plant data:
TopCycle output = 101.0 MW

Auxiliaries
 Dryer internal power consumption 1.29 MW
 CO2 compression work 7.49 MW
 Gasifier own consumption 0.97 MW
 Flue gas fan after FGD 1.29 MW
 Flue gas fan upstream absorber 0.74 MW
 CO2 Capture own electrical consumption 2.15 MW

 Cooling pumps 0.06 MW
 FW pump 0.38 MW
 Cooling water pump 0.14 MW
 Condensate pump 0.01 MW

Fuel input= 184.5 MW

Efficiency= 46.86 %

Heat loads
- Dryer= 34.6 MW
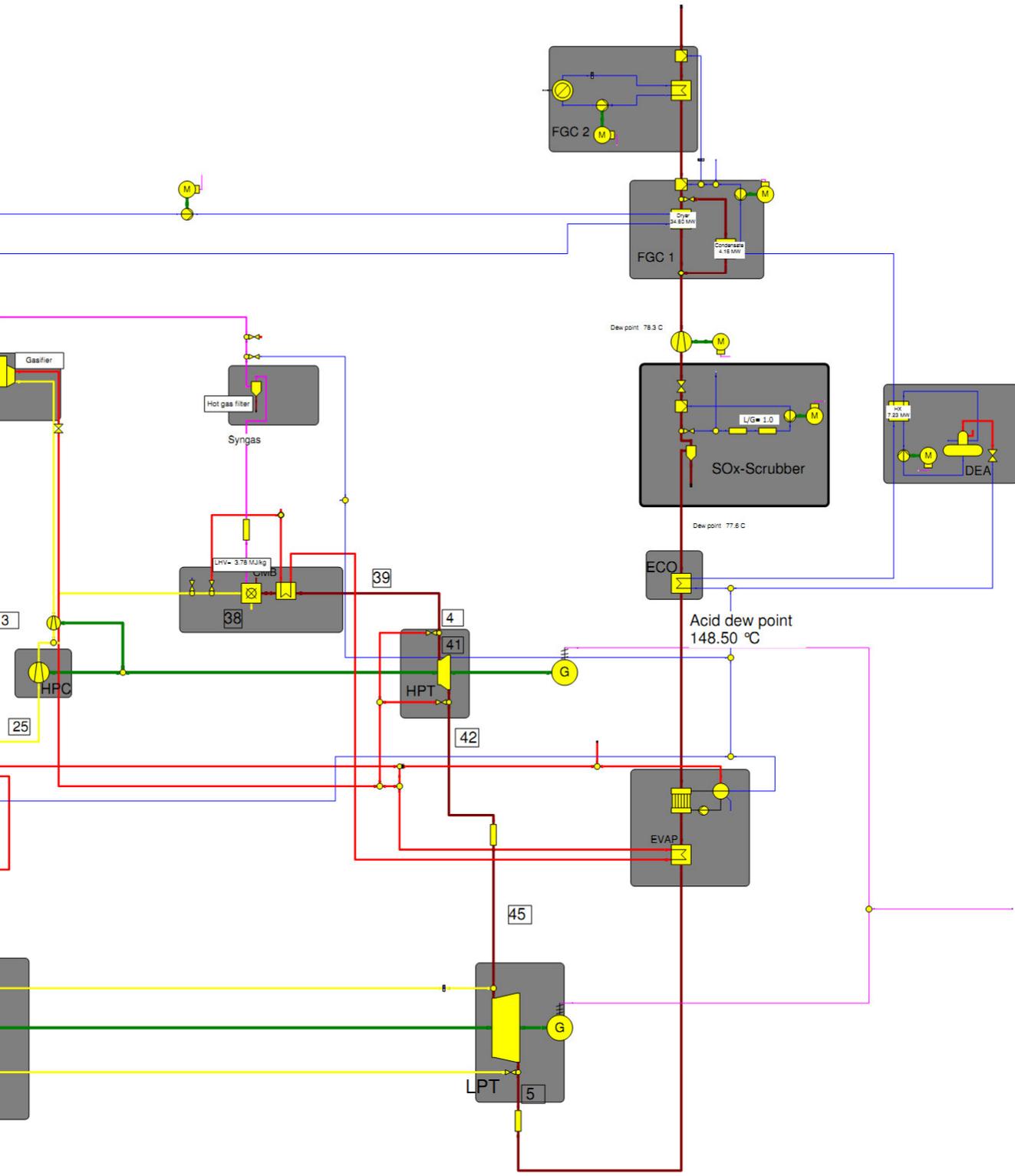
Consumption:
 Nitrogen =   1.6 kg/s
 Water =   -2.8 kg/s

Waste water:
 10.6 kg/s

4

41

HPT

42

Acid dew point
148.56 °C

EVAP

45

LPT

5

G

G

G

FGC 2

FGC 1

Dryer
34.60 MW

Condensate
4.16 MW

Gasifier

Hot gas filter

Syngas

Dew point 76.3 C

HX
7.23 MW

SOx-Scrubber

L/G= 1.0

DEA

Dew point 77.6 C

LHV= 3.76 MJ/kg

COMB

38

39

4

41

ECO

25

HPC

HPT

42

G

Acid dew point
148.50 ℃

EVAP

45

LPT

5

G

Plant data:
  TopCycle output = 103.8 MW

Auxiliaries:
  Dryer internal power consumption   1.29 MW
  CO2 compression work 0 MW
  Gasifier own consumption   0.97 MW
  Flue gas fan after FGD   1.26 MW
  Flue gas fan upstream absorber 0 MW
  CO2 Capture own electrical consumption 0 MW

  Cooling pumps   0.04 MW
  FW pump   0.43 MW
  Cooling water pump   0.29 MW
  Condensate pump   0.01 MW

Fuel input= 184.3 MW

Efficiency= 53.98 %

Heat loads:
  Dryer= 34.6 MW

Consumption:
  Nitrogen =   1.6 kg/s
  Water =   1.6 kg/s

Excess steam =   3.3 kg/s

Waste water:
  10.6 kg/s

temperature dryer

Low temperature dryer
Electricity consumption = 1.95 MW
Heat demand = 36.16 MW
Supply temp. = 75 °C
Return temp. = 25 °C

Gasifier

Hot gas filter

Syngas

Nitrogen

3

25

IC

LPC
24
2

FGC 2

FGC 1

Dew point 80.1 C

SOx-Scrubber

L/G= 1.0

DEA

Dew point 79.1 C

ECO

Acid dew point
149.27 °C

39

4
41

HPT
G

42

45

EVAP

LPT
5

G
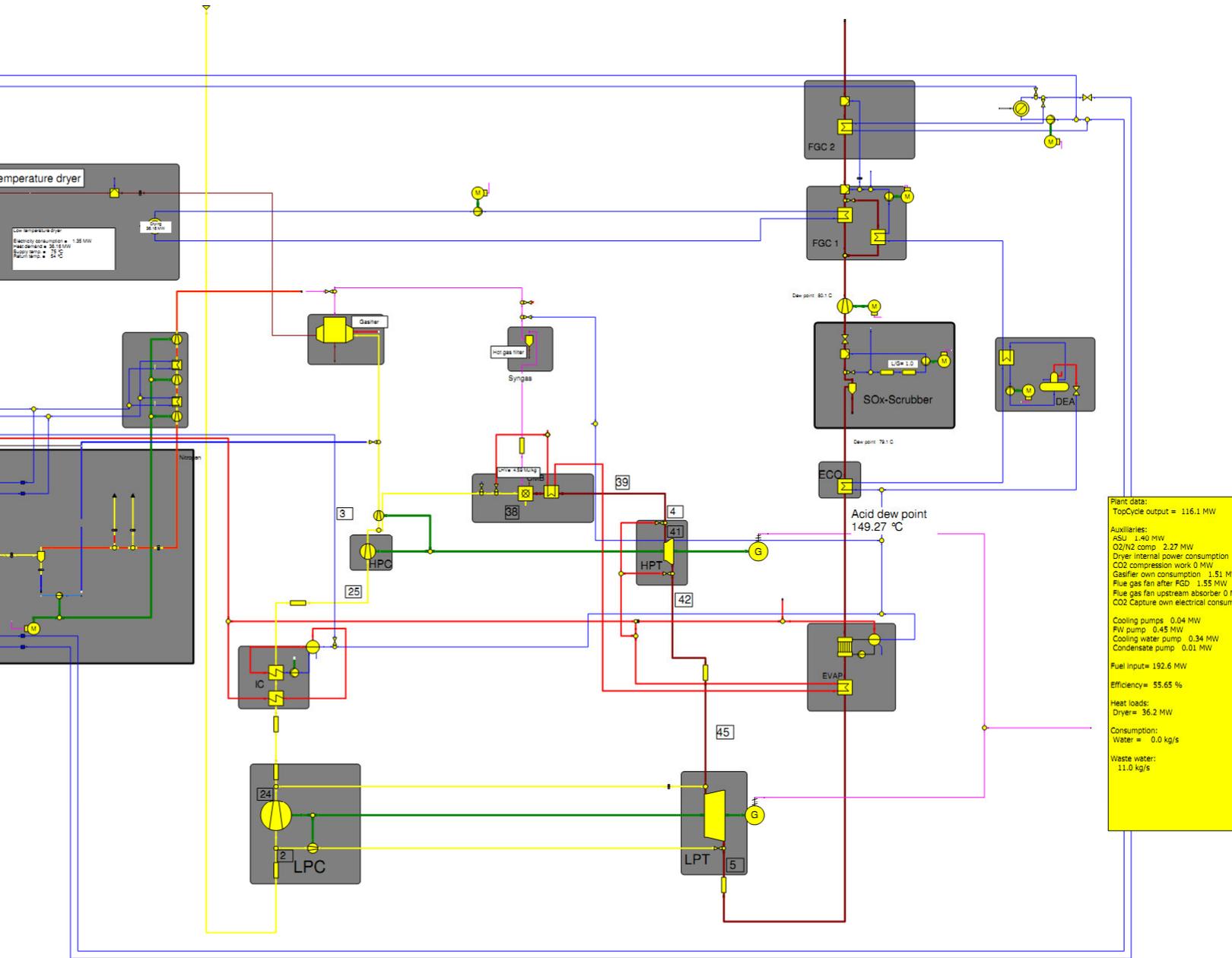
Plant data:
TopCycle output = 116.1 MW

Auxiliaries:
ASU 1.40 MW
O2/N2 comp 2.27 MW
CO2 compression work 0 MW
Dryer internal power consumption
Gasifier own consumption 1.51 MW
Flue gas fan after FGD 1.55 MW
Flue gas fan upstream absorber 0 M
CO2 Capture own electrical consum

Cooling pumps 0.04 MW
FW pump 0.45 MW
Cooling water pump 0.34 MW
Condensate pump 0.01 MW

Fuel input= 192.6 MW

Efficiency= 55.65 %

Heat loads:
Dryer= 36.2 MW

Consumption:
Water = 0.0 kg/s

Waste water:
11.0 kg/s

Captured CO2
18.95 kg/s
90.00 %

CO2 compression

FGC 3

FGC 2

Ash Boiler
Heat supply = 8.35 MW
Combustion eff = 99.00 %

FGC 1

Dew point 78.7 C

Dehydration

LiG= 1.0

DEA

Hot gas filter

Syngas

CMB

ECO

38

39

4

41

HPT

42

Dew point   77.6 C

Acid dew point
148.50 ℃

EVAP

45

LPT

5

G

Plant data:
TopCycle output = 103.8 MW

Auxiliaries
 Dryer internal power consumption 3.96 MW
 CO2 compression work 7.87 MW
 Gasifier own consumption 0.97 MW
 Flue gas fan after FGD 1.28 MW
 Flue gas fan upstream absorber 0.74 MW
 CO2 Capture own electrical consumption 2.15 MW

 Cooling pumps 0.05 MW
 FW pump 0.39 MW
 Cooling water pump 0.26 MW
 Condensate pump 0.01 MW

Fuel input= 184.3 MW

Efficiency= 46.71 %

Heat loads
- Dryer= 0.0 MW

Consumption:
 Nitrogen =   1.6 kg/s
 Water = -2.8 kg/s

Waste water:
 10.6 kg/s

Appendix B

Details of the inert gas system consumption and simulation aspects of the flue gas condenser and water injection for the air blown gasifier cases.

Inert gas system calculations

The true density of the dried lignite is derived from the equation proposed by Parkash to calculate the true density for dry, ash free (daf) subbituminous coals:

Equation 7. Equation for calculating true density of subbituminous coals on daf basis

$$True\ density\ (daf)\ =\ 3.5742\ -\ (0.0197 \cdot C\ + 0.0192\ \cdot O) -\ 0.0691 \cdot H\ [^g/_{cc}]$$

The true density on daf-basis is calculated and corrected with the moisture and ash components assuming an ash density of 2.33 g/cc, also proposed by Parkash (Parkash, 1985). The calculated true density ($\approx$1500kg/m^3) is then used in further calculations to estimate the amount of inert gas dilution.

In more detail, the calculation of the nitrogen consumption in an entrained flow gasifier is based on the sum of three components:

1. Sluicing of pulverized fuel
2. Dense phase transport of pulverized fuel
3. Residual nitrogen consumption

The first factor (1) is calculated by assuming a vessel volumetric efficiency (η_v=0.67) and a true and bulk density of the lignite entering the gasifier, of ρ_t=1500 kg/m^3 and ρ_b=500 kg/m^3 respectively. The fuel takes up a bulk volume of:

Equation 8.

$$\frac{1}{\rho_b} = 0.002\ ^{m^3}/_{kg\ fuel}$$

The occupied volume gives a vessel volume together with the vessel volumetric efficiency:

Equation 9.

$$\frac{0.002}{\eta_v} = 0.002985\ ^{m^3}/_{kg\ fuel}$$

By subtracting the true volume of the fuel from the vessel volume the total void which needs to be filled up with nitrogen is obtained:

Equation 10.

$$0.002985 - \frac{1}{\rho_t} = 0.002985 - 0.000667 = 0.002318\ ^{m^3 N_2}/_{kg\ fuel}$$

The second factor (2) is calculated in a similar manner for the dense phase transport by assuming a dense phase porosity (η_p=0.7) and assuming no slip between the fuel particles and the transport nitrogen.

IX

Equation 11.

$$\frac{1/\rho_b}{\eta_v} - \frac{1}{\rho_t} - \left(\frac{1}{\rho_b} - \frac{1}{\rho_t}\right) \approx 0.00085 \; m^3 N_2 \Big/ kg \; fuel$$

The true volume of the fuel and the nitrogen in the pores is subtracted from total volume of the imagined transport vessel and a dense phase transport nitrogen can be calculated.

The last factor is a residual nitrogen factor (3) which is equal to the sum of the sluicing of pulverized fuel nitrogen (1) and the nitrogen needed for dense phase transport (2).

The sum, the total nitrogen demand of the entrained flow gasifier, is then given in the unit of m^3 N_2/kg fuel. The ideal gas law (Equation 10) is then used to calculate the nitrogen demand given on the from kg N_2/kg fuel.

Equation 12.

$$p \cdot V = \left(\frac{m}{M}\right) \cdot R \cdot T \rightarrow m = \frac{p \cdot V \cdot M}{R \cdot T}$$

Where p is pressure, V is volume, M is the molar mass of the substance, R is the gas constant and T is the temperature in Kelvin. The nitrogen demand is calculated at 80°C and at a pressure 10% above the operating pressure of the gasifier.

The dilution of inert gas into the syngas stream however is not the same as the nitrogen demand for the gasifier. The above procedure to calculate the nitrogen demand for the gasifier and fuel feeding system is used to determine how much nitrogen is needed to pressurize, since this is a costly procedure, in means of efficiency penalty.

The dilution of nitrogen for an entrained flow gasifier is calculated by Equation 13.

Equation 13.

$$\frac{1/\rho_b}{\eta_p} - \frac{1}{\rho_t} = 0.00219 \; m^3 N_2 \Big/ kg \; fuel$$

The equation is very similar to Equation 11, but instead of subtracting the nitrogen which is in the pores, this nitrogen also follows the dense phase transport into the gasifier, diluting the syngas. The amount of nitrogen, in terms of mass flow is calculated by using the ideal gas law and Equation 12.

The HTW gasifier uses a screw feeding system and has its nitrogen dilution of the syngas rescaled with the ideal gas law to appropriate pressure based originally on a standard HTW case at 35 bar operating pressure(Confidential, Source 3).

The resulting consumptions of each gasifier in regards of nitrogen dilution and consumption is given in Table 13.

X

	Siemens, 49,5 bar, IGCC ref.	Siemens 77 bar, TopCycle oxygen blown	MHI, 77 bar, TopCycle air blown	HTW, 77 bar, TopCycle air blown
Nitrogen dilution	10.34 %	16.08 %	16.08 %	14.74 %
Nitrogen consumption	29.98 %	46.63 %	46.63 %	14.74 %

Simulation of flue gas condenser

An important parameter when dealing with the condensation of water in the flue gases is the actual dew point of the flue gas. Ebsilon has its limitations in this case because of the way the heat balance is calculated.

In the air blown case where the condensation heat of water is utilized to a great extent every degree is crucial for the overall efficiency. After the fan situated after the SO_X-cleaning in the flue gas train the temperature of the flue gas is approximately 86°C. The flue gas dew point is however in the region of 79°C. This means that considerable heat can only be extracted below the dew gas temperature, since everything above 79°C is purely sensible heat introduced from the fan, since the flue gas is saturated after the SO_X-scrubber.

It is however possible to extract heat from 84°C all the way down to 54°C as long as the end and starting points of enthalpy of the flue gas is correct in relation to the heated water. In other words, Ebsilon simplifies a condenser by just checking that the starting and ending temperatures match, even though the gradual condensation of water makes it impossible to actually extract heat at a temperature of 85°C – as indicated by the thick dashed line in Figure 52, where the blue line represents the characteristics of the flue gas. The additional red line represents the actual heat possible to extract as an example when utilizing a 2 Kelvin pinch temperature at the inlet, relative to the dew point.
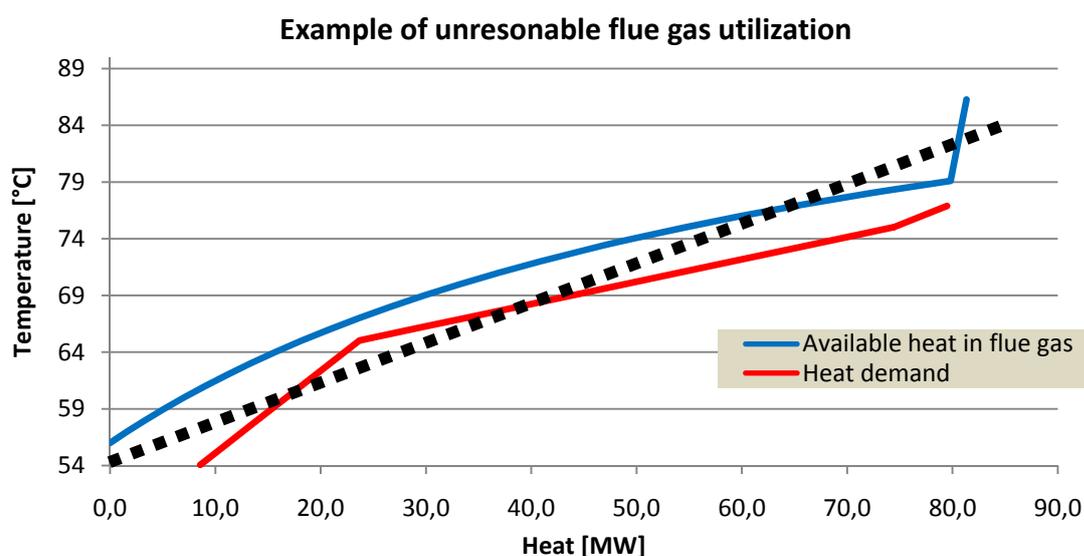
Figure 52. The dashed line represents the possible approximation error in Ebsilon when dealing with flue gas condensation

The issue is resolved by keeping an eye on the actual dew point of the flue gas and comparing it to the heat extracted in the first condenser. This can be done manually or by introducing a controlling mechanism in Ebsilon. Throughout simulations the temperature of the extracted heat has been set to a temperature of 2 K below the dew point of the flue gas to make sure the condition of the minimum pinch temperature is met.

Choice of heat recovery system for TopCycle air-blown gasifier case

The hot syngas from the HTW air blown gasifier has to be cleaned from particles. As the design of the hot gas filter cannot withstand a higher temperature than 500°C. There are three principle ways to lower the temperature of the syngas exiting the gasifier by water injection, syngas cooling or steam injection, shown in Figure 53, Figure 54 and Figure 55.
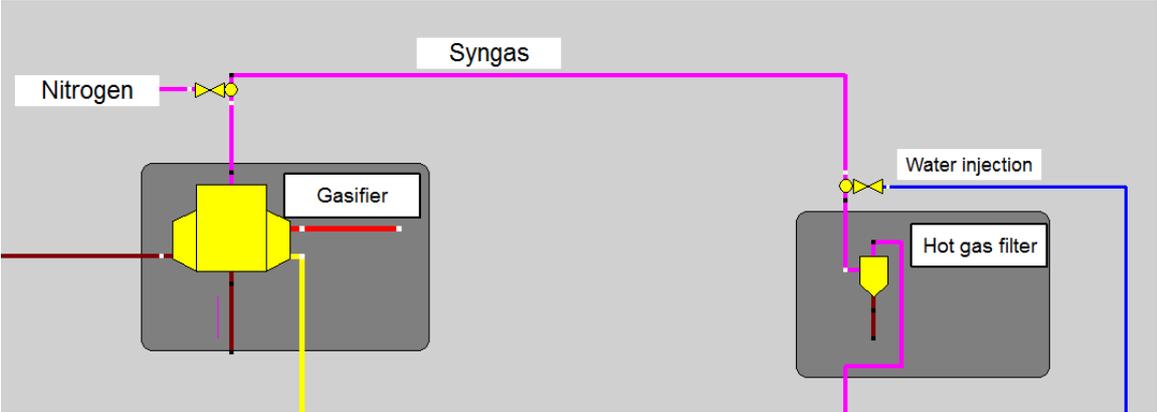
Figure 53. Water injection
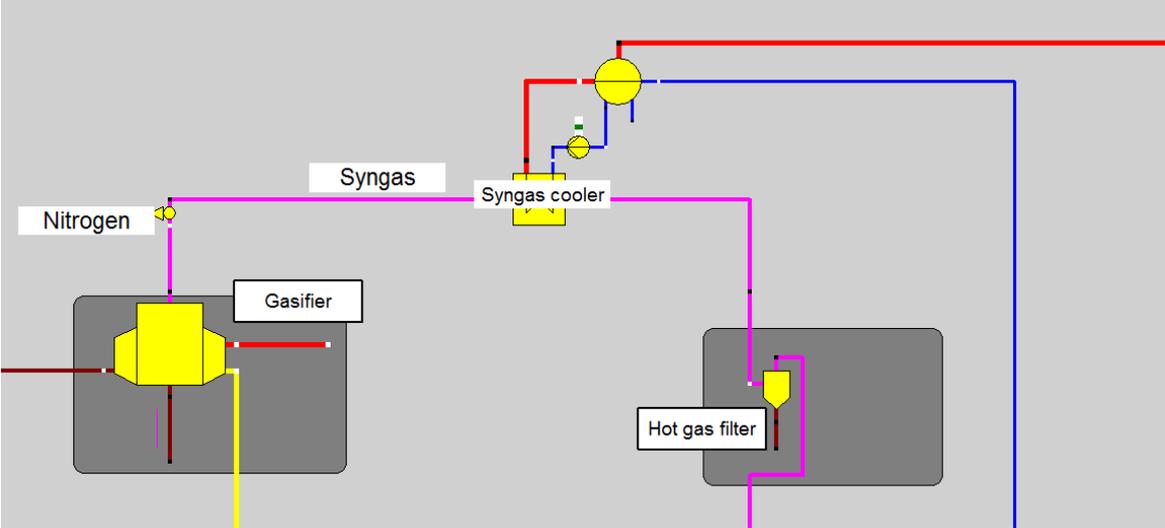
Figure 54. Syngas cooling.
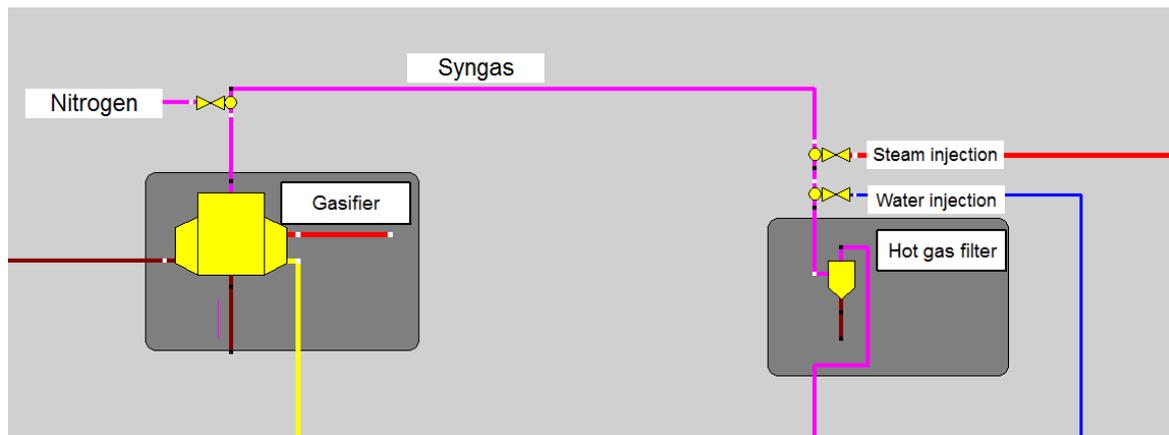
Figure 55. Steam injection

Thermodynamically in Ebsilon, there is no difference from injecting water into the syngas, cooling the syngas down to 500°C or to use an evaporator to raise steam and then superheat it to 500°C somewhere else in the system. However, for these two principles to be exactly the same, then the syngas cooler also must have a super heater in the syngas stream or utilize a high temperature source from somewhere else, like in the HRSG. The syngas cooler however does not "dilute" the syngas with any water vapour, unlike the other two options.

Injecting steam into the syngas is a good way of pressing in as much moisture as possible into syngas, at the cost of syngas heating value. The working principle is that as much steam as possible is injected into the syngas and water is then used to come down to sufficiently low temperature. When no superheating is used within the whole system (not in the HRSG either) then steam injection is superior. This is simply due to the fact that any water which enters the syngas is super heated to the temperature of 500°C. In initial simulations there were no superheaters in the system, and steam injection seemed completely superior, because of this reason.

Water injection is the middle way - moderate dilution of the syngas, since less water is put into the syngas than in the steam injection case, and no expensive heat exchanger areas need to be installed to handle the corrosive, high temperature syngas. The water injection principle is used in all air blown simulations. However, the syngas cooler seems as an attractive alternative due to the low heating value of the syngas.

Appendix C

Flue gas temperature [°C] vs Available heat [MW]

Legend:
- Humidification to 15°C
- Humidification to 20°C
- Humidification to 25°C
- Humidification to 30°C
- Humidification to 35°C
- Humidification to 40°C
- Humidification to 45°C
- Humidification to 50°C
- Humidification to 55°C
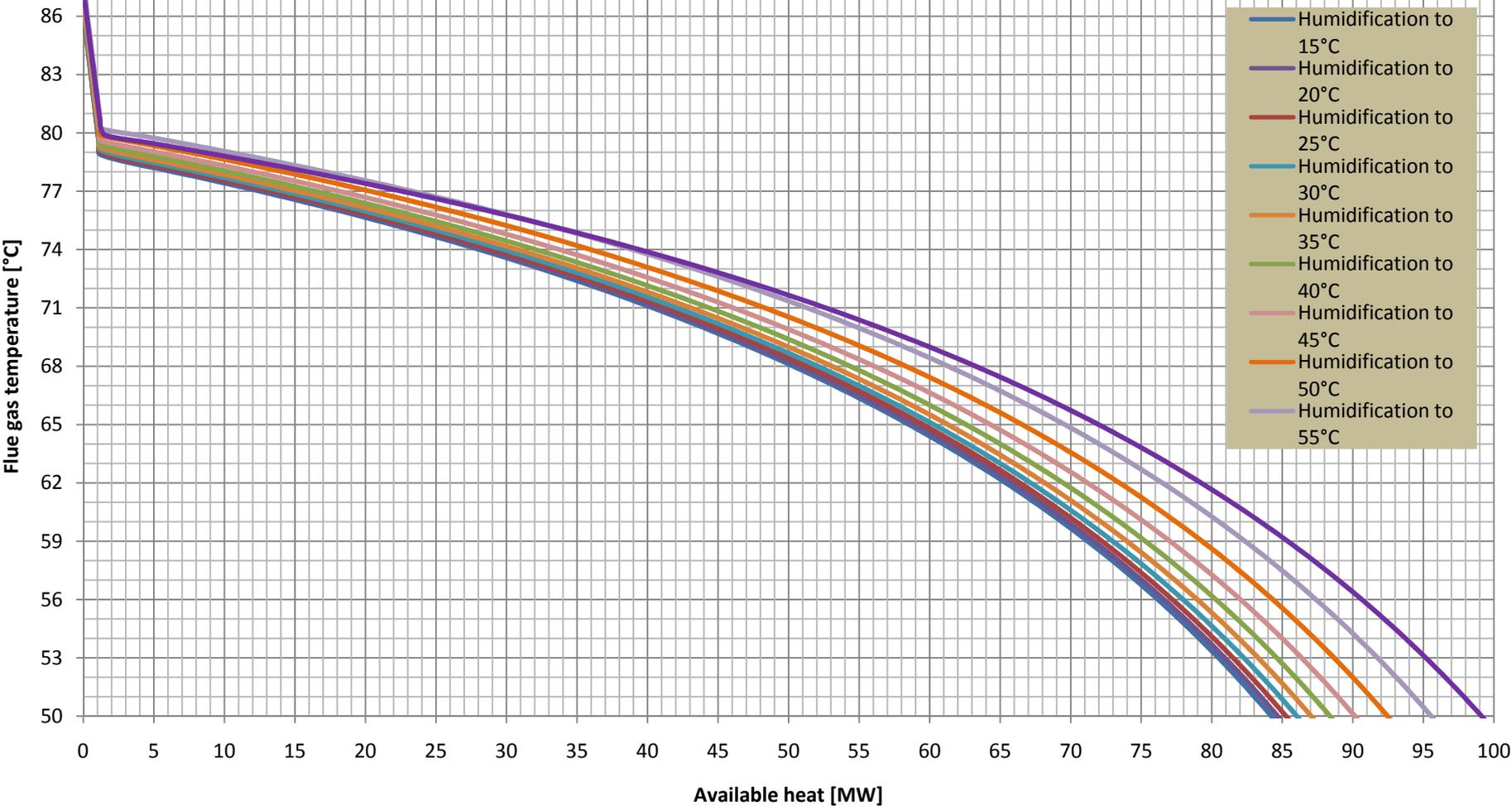
Figure 56. Available heat in the flue gas condenser depending on inlet humidification
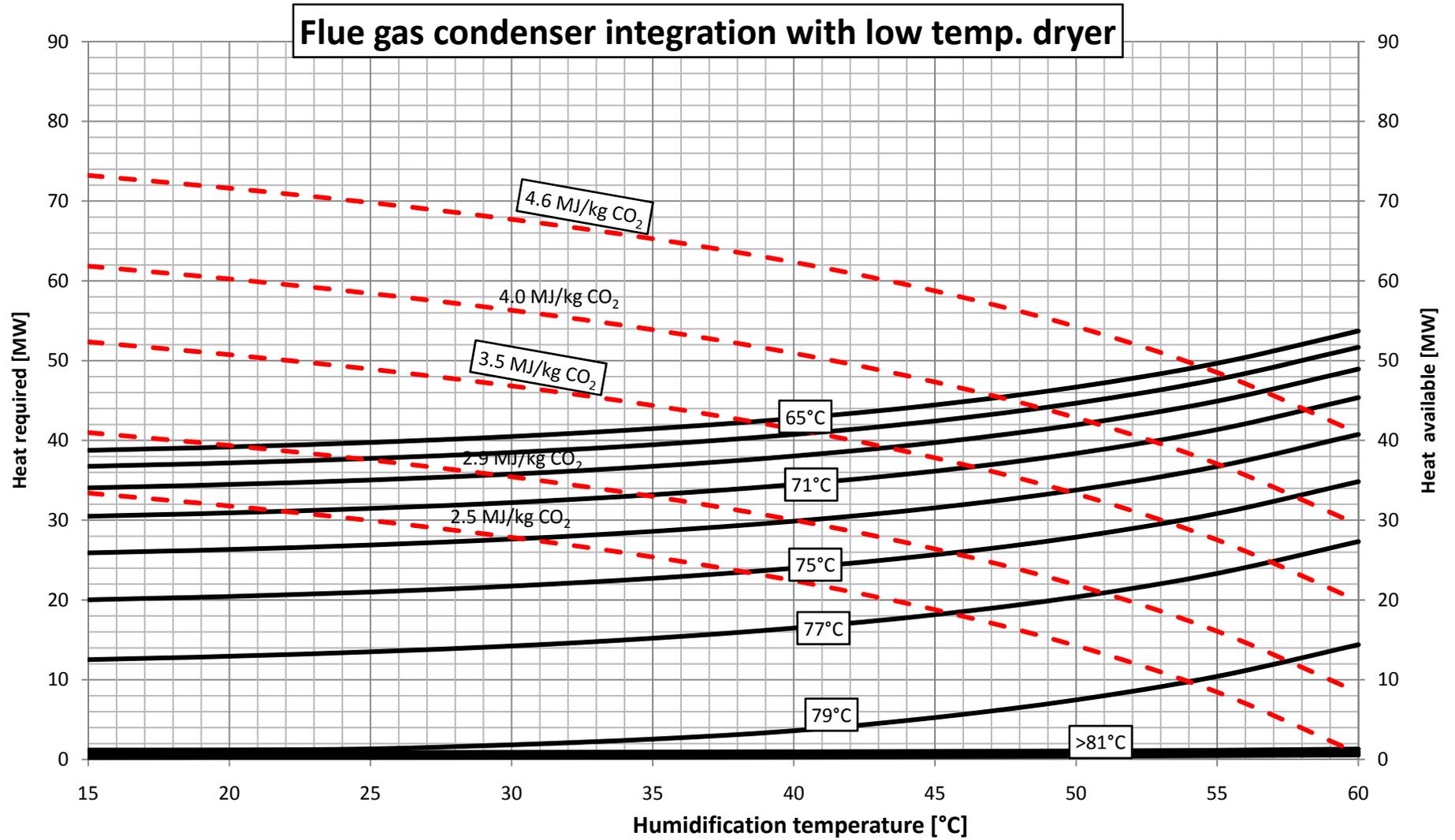
Flue gas condenser integration with WTA

XVI
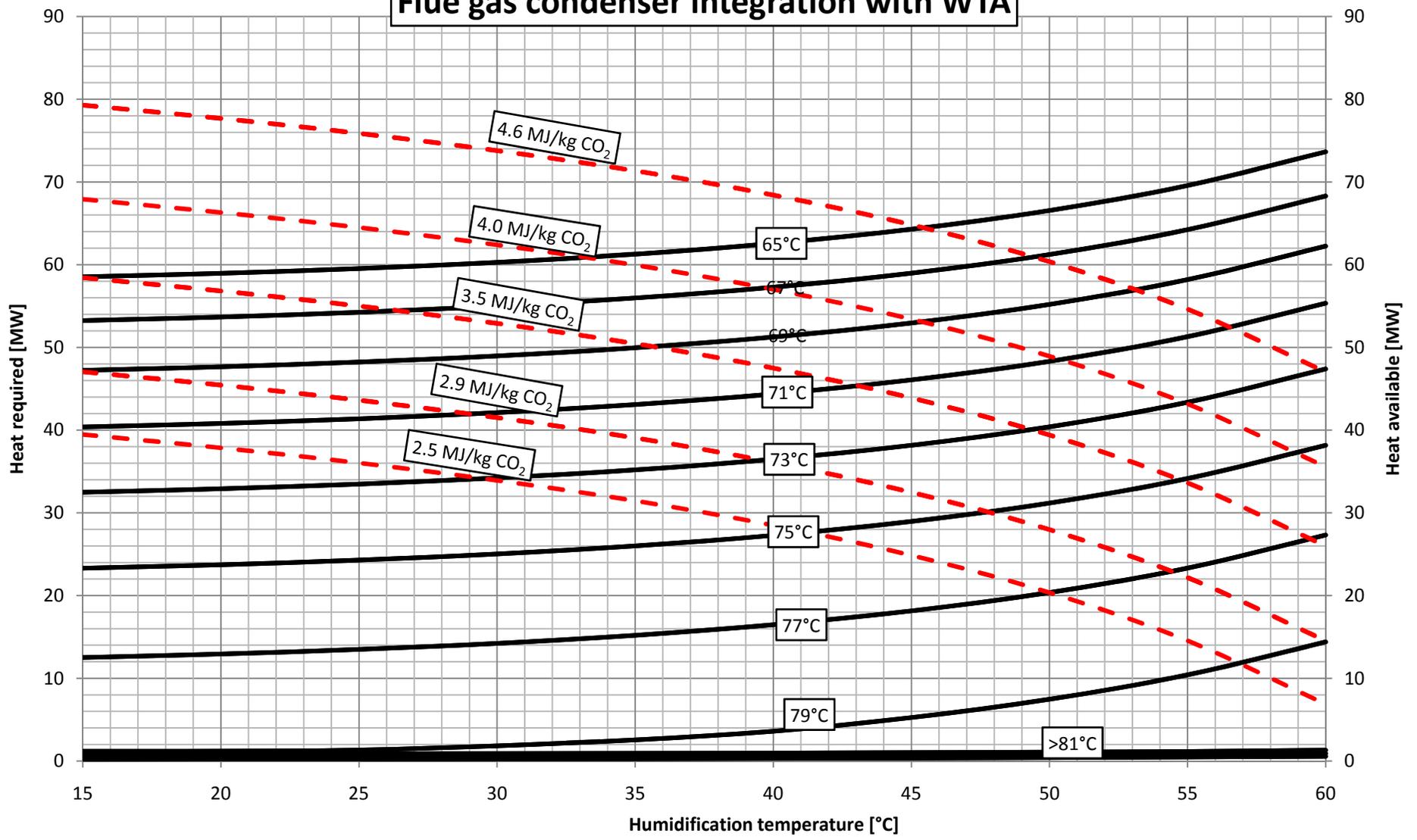
Efficiency as a function of humidification

Legend:
- With WTA dryer (1.0 bar CO2)
- With low temp. dryer (1.0 bar CO2)
- With WTA dryer (0.3 bar CO2)

In-chart labels: Low temp. dryer; WTA dryer; WTA dryer (0.3 bar CO_2)

Y-axis: Net Efficiency (35% to 50%)

X-axis: Humidification temperature [°C] (15 to 60)
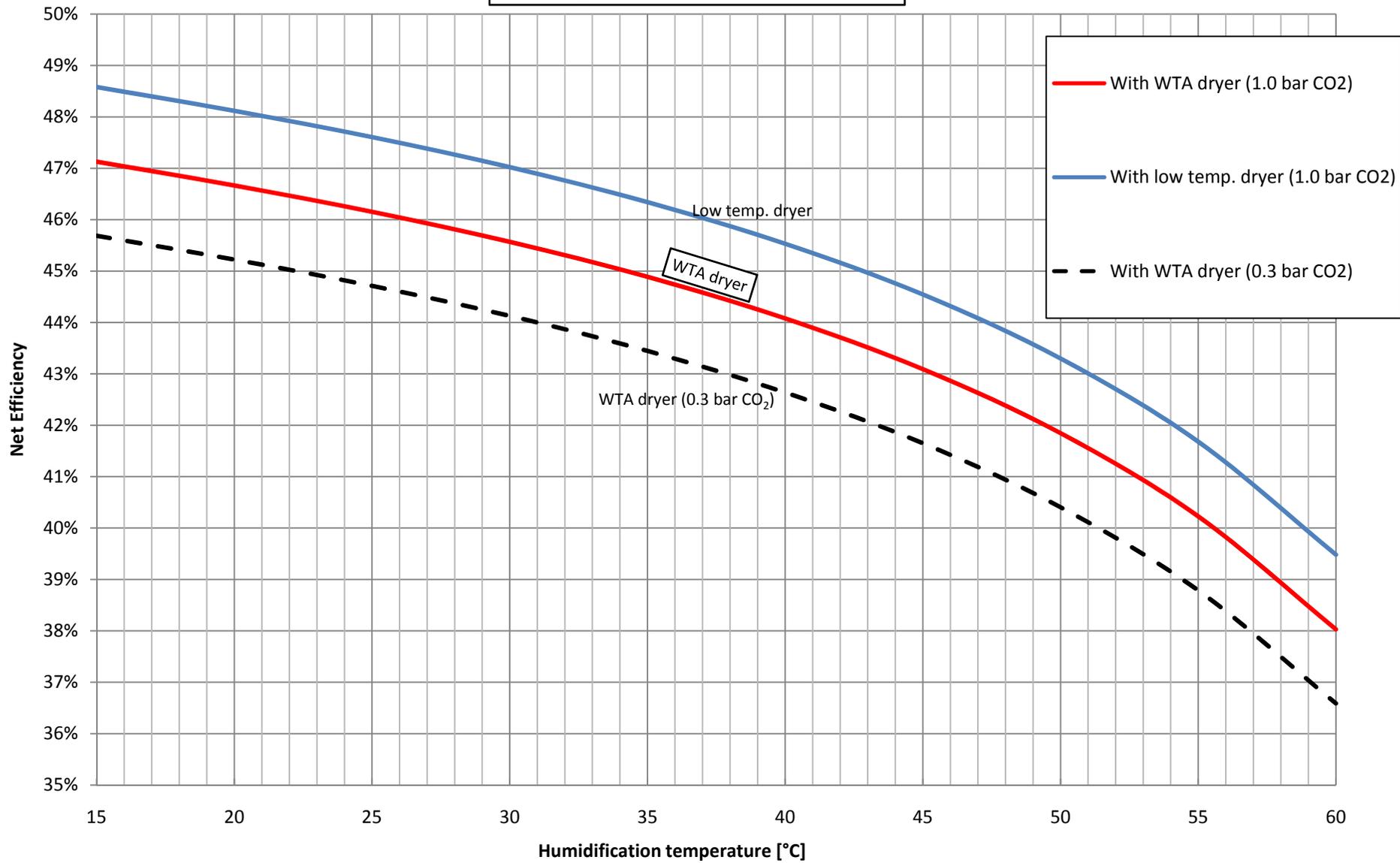
Appendix D

A summary of the gasifiers considered in the concept selection is presented in Table 14.

Table 14. Short summary of gasifiers considered

Category	Moving bed		Fluidized bed				Entrained flow				
	BGL	Lurgi	HTW	HRL	KBR	KRW	C/P E-Gas	Siemens	GE	Prenflo (Shell)	MHI
Ash conditions	Slagging	Dry ash	Dry ash	Dry ash	Dry ash	Agglomerating	Slagging	Slagging	Slagging	Slagging	Slagging
Fuel feeding	Dry feed	Dry feed	Dry feed	Dry feed	Dry feed	Dry feed	Slurry, 2-stage feed	Dry feed	Slurry feed	Dry feed	Dry feed, 2-stage feed
Oxidant	St / O_2	St / O_2 or St/Air	St / O_2 or St / Air	Air	St/O_2 or St/Air	St/Air or St/O2	O_2	O_2	O_2	O_2	Air
Other	Hydrocarbons in syngas	Hydrocarbons in syngas	-	Integrated fuel drying	-	Gasifier used in Piñon Pine	Gasifier used in Wabash River		Used in Polk/Tampa	Used in Puertollano and Buggenum	
Size offered	3.6 m diameter	3-5 m diameter	2.8 m diameter	-	-	-	-	500 to 1200 MW$_{th}$	-	1200 MW$_{th}$	≈800 MW$_{th}$
#Commercial	2	>165	2	-	-	-	1	>9	>63	>2	>1
Carbon conversion	-	-	94%	-	-	-	98%	>99%	95%	>99%	>99%
Cold gas efficiency	85-88%	76-87%	80%	-	-	-	70%	76-79%	60-63%	74-77%	77%

A summary of ongoing IGCC projects throughout the world in 2010 by Gas Turbine World is shown in Table 15 (Jaeger, 2010).

Table 15. IGCC project under development in 2010 (Jaeger, 2010)

Project	Fuel	Year	Gasifier	GT	Net output	Comments
Alter NRG, Canada	Petcoke / Oil field waste	N/A	Alter NRG	N/A	120 MW	90% CO_2 capture
BHEL/Andhra Pradesh Power, India	N/A	N/A	N/A	N/A	125 MW	Memorandum of understanding
Cash Creek, Kentucky	Kentucky #9 coal	2013	GE Energy	2 x Fr 7FB	720 MW	Coal-to-SNG plus NGCC, permit in progress
Dongguan IGCC, China	Low grade coal	2011	KRB-TRIG	N/A	120 MW	Chinese demo of TRIG and KBR
Duke Energy, Indiana	Coal	2012	GE Energy	2 x Fr 7FB	630 MW	Construction over 30% complete
FutureGen, Illinois	N/A	2013-2014	N/A	N/A	275 MW	Pending DOE funding
Good Spring IGCC, Pennsylvania	N/A	2012	(China)	N/A	270 MW	Seeking DOE funds
Great Bend, Ohio	Coal	2010	GE Energy	2 x Fr 7FB	630 MW	Project delay
GreenGen, China	Coal	2010	Thermal Power	N/A	250 MW	CO_2 capture
Hatfield/Powerfuel, UK	Coal	2014	Shell	GE hydrogen	900+ MW	Initial operation on NG. EEC funds for CCS
Hydrogen Energy California, California	Petcoke / Western bituminous coal	2015-2016	GE Energy	1 x Fr 7FB	390 MW + NG 100 MW peaking	90% CO_2 capture for EOR injection
Lima Energy, Ohio	Coal	N/A	CoP E-Gas	2 x Fr 7FA	600 MW	Project on hold. Sierra Club opposed
Magnum / Nuon, Netherlands	Coal / Biomass	2011	Shell license	3 x M701F4	1200 MW	Initial operation on NG fuel 2011
Mesaba Energy, Minnesota (Phase 1)	PRB / Petcoke	2012	CoP E-Gas	2 x 5000F	600 MW	Recived PUC permits. Needs PPA for financing
Mississippi Power, Mississippi	Lignite	2013-2015	KBR-TRIG	N/A	582 MW	PUC permitting process in progress
Mountaineer, West Virginia	Coal	2012	GE Energy	2 x Fr 7FB	630 MW	Project delay
RWE Rhenish, Germany	Lignite or Hard coal	2014	N/A	N/A	450 MW polygeneration	360 MW net and 300 tph hydrogen
Taylorville/Tenaska, Illinois	Coal	2014	Siemens	TBD	630+ MW	Coal-to-SNG plus NGCC. DOE loan guarantee
Taean No. 1, Korea	Coal	2014	Shell license	N/A	N/A	In progress
Texas Clean Energy IGCC, Texas	Coal	N/A	Siemens	1 x 5000F	400 MW	To sell CO_2 for EOR
Wandoan, Australia	N/A	2016	GE Energy	N/A	400 MW	90% CO_2 capture. Seeking Queensland funds
ZeroGen, Australia	Hard coal	2016	MHI	1 x M701G2	530 MW	90 % CO_2 capture

Appendix F

Results from the simulations which are not included in the main result chapter.

Inlet Humidification & condenser heat balance

The need for humidification, introduced by the heat requirement of the CO_2-capture unit, results in large heat recovery differences for the HTW options with CO_2-capture.

Without inlet humidification for the IGTC air blown HTW with ECO_2 CO_2-capture the result is the T-Q diagram shown in Figure 57.

Heat demand in flue gas with ECO2 & Dryer - without inlet humidification
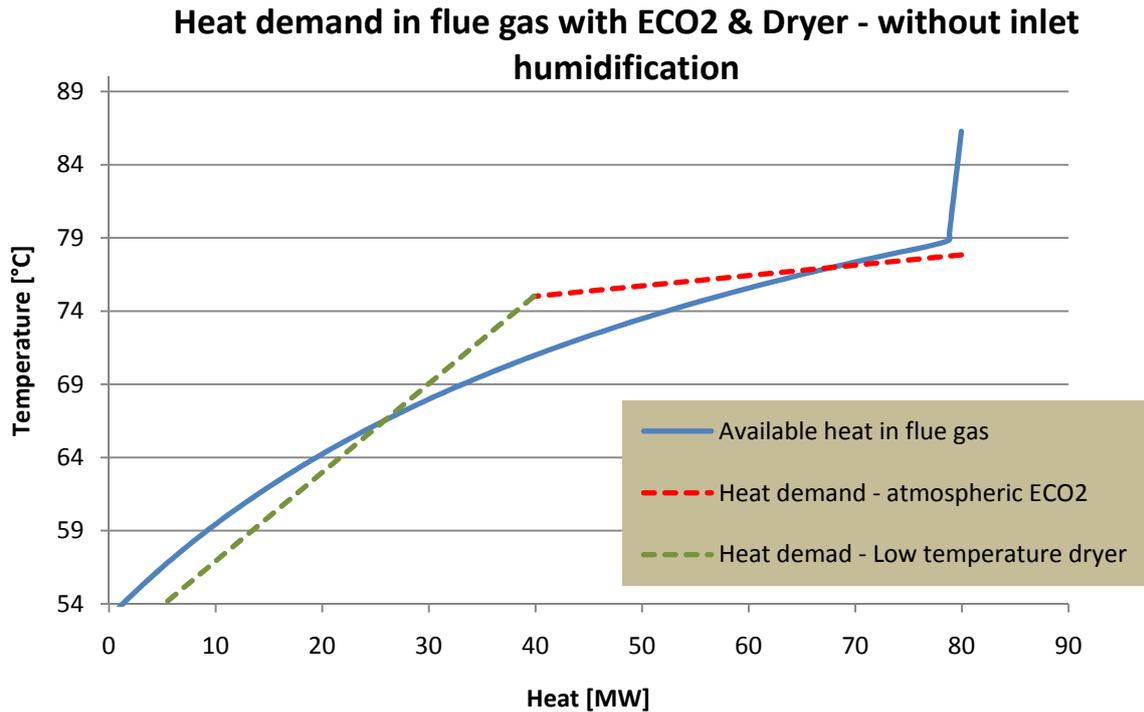
Figure 57. Heat deficit with the ECO$_2$ CO$_2$-capture unit - without inlet humidification

The heat deficit without inlet humidification is clear. The plant is not able to supply the required heat for the ancillary units at the required temperatures. Without the CO_2-capture unit the dashed green line is moved all the way to the right side of the diagram, and the diagram would be identical to a T-Q diagram for the HTW option without CO_2-capture, where only a dryer is the only heat load in the flue gas condenser. The sharp downward characteristic of the curve in the very beginning reflects a de-superheating zone before the flue gas reaches its dew point. This sensible heat is added to the flue gas between the flue gas scrubber and the condenser by a flue gas fan.

The heat demand needed from the flue gas condenser (FGC1, shown in Figure 10) to the ECO_2 unit is determined by the heat deficit after having been supplied with heat from the ash boiler, the excess steam from the HRSG (indirectly affected by inlet humidification) and heat from intercooling of the CO_2-compression unit. With no inlet humidification there is only a certain amount of HP steam available (5 MW), which is the excess 3.3 kg/s in the HTW option without CO_2-capture. The ash boiler has a fixed about of char which it can incinerate (about 8.4 MW), furthermore the heat available from CO_2-compression inter-cooling is also limited, about 1.5 MW.

By humidifying the inlet air more heat is made available in the flue gas condenser. The resulting T-Q diagrams with inlet humidification are shown in Figure 58, Figure 59 and Figure 60. The black dashed lines represent the heat demand of the inlet humidifier, and indicate the temperature of which the heat can be supplied to the inlet humidifier. The solid black line represents the inlet humidification temperature.

Heat demand in flue gas with MEA, Dryer and Humidifaction
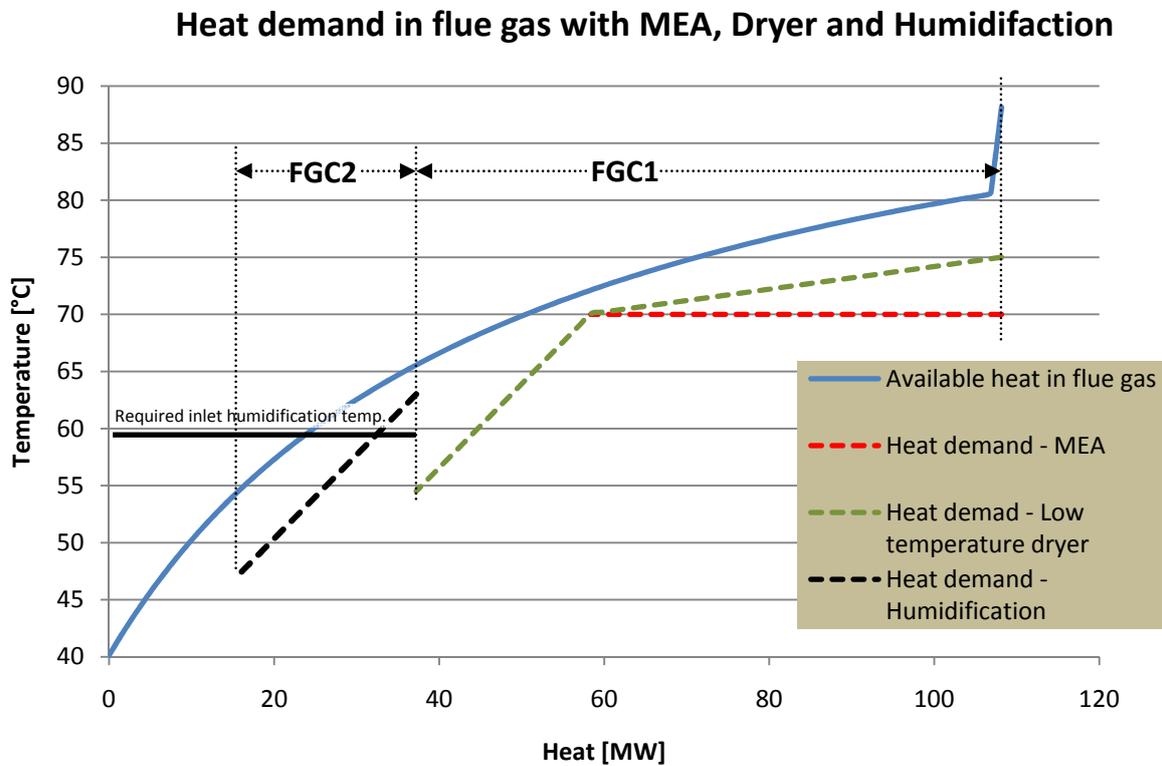
Figure 58. Simplified T-Q diagram of the IGTC HTW with the MEA CO$_2$-capture option

In the MEA option the high temperature portion of the dryer duty, 70°C to 75°C, is heat exchanged in parallel to the steam evaporator operated to supply 0.3 bar steam to the MEA unit. The composite curve is used to visualise the parallel configuration of the dryer and the steam boiler, i.e. the heat demands of the MEA and high temperature portion of the dryer are added to obtain a total demand of 50 MW between 88°C and 72°C together, not individually.

Figure 59. Simplified T-Q diagram of the IGTC HTW with the ECO$_2$ CO$_2$-capture option
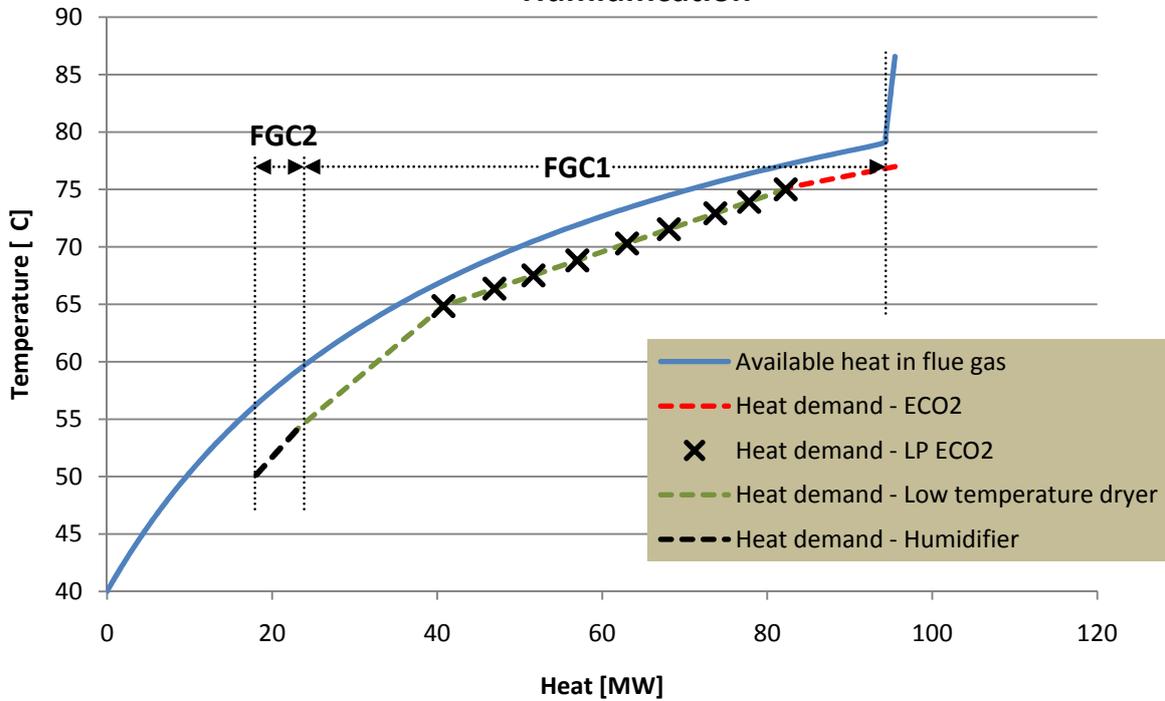
Figure 60. Simplified T-Q diagram of the IGTC HTW with the ECO$_2$ combination of an atmospheric and a sub-atmospheric stripper

In a similar fashion to the MEA case, the low pressure ECO_2 stripper (LP ECO_2) and the high temperature range of the low temperature dryer are heat exchanged with the flue gas in parallel. The absence of a line representing the inlet humidification temperature in the ECO_2 combination figure is simply due to that the temperature required (29°C) is so much lower than the temperatures available in the flue gas.

The heat demand characteristics for the MEA, ECO_2 and ECO_2 combination are due to the assumptions regarding each CO_2-capture process. The MEA requires heat at a low temperature, but instead requires a lot, resulting in a very large inlet humidification need. The ECO_2 combination may open for the possibility to carefully utilize the flue gas heat to a very high degree, fitting the heat demand to the flue gas characteristics.

Table 16 show the inlet temperatures and the temperatures of the flue gas prior to FGC2, and heat and mass transfer to the inlet humidifier. As can be see there is a very small temperature difference for the MEA option. Condensing the flue gas would result in a water system temperature of about 66°C, at most, and this water needs to humidify the inlet to ≈59°C. Preliminary calculations within Vattenfall, using an excel-based humidification programme, have shown that the low temperature difference makes it impossible to humidify the inlet air to the desired extent for the MEA option. For the ECO_2 option, the same kind of simulations has shown that the inlet humidification should be plausible.

Table 16. Inlet humidification temperatures for the IGTC air blown HTW options with CO_2-capture

	IGTC HTW air		
Humidification	MEA	ECO_2	ECO_2 Combi
Heat duty [MW]	31.3	18.9	5.0
Humidified air temp. [°C]	58.6	50.0	29.0
Flue gas temp., before FGC2 [°C]	68.5	71.4	62.2
Temperature difference [°C]	9.9	21.4	33.2

CO₂ compression characteristics

Due to the differences in delivery pressure to the CO_2-compression unit from the CO_2-capture unit the electricity demand for the CO_2-compression duty varies considerably for the different cases simulated. Information about the CO_2-capture and compression specifics for each simulation is found in Table 17.

Table 17. Electricity demand and key numbers for CO_2-capture and compression characteristics

	IGCC	IGTC Oxygen	IGTC HTW air				IGTC MHI air
	Selexol	Selexol	No CO_2-capture	MEA	ECO₂	ECO₂ Combi	No CO_2-capture
Fuel input [MW]	1038.2	286.9	184.3	185.8	185.2	184.5	192.6
CO_2-capture own consumption [MW]	21.1	5.1	N/A	1.4	2.2	2.2	N/A
CO_2-compression [MW]	25.8	6.0	N/A	9.6	6.9	7.5	N/A
Flue gas fan before CO_2 capture [MW]	N/A	N/A	N/A	0.7	0.7	0.7	N/A
CO_2 pressure after CO_2-capture unit [bar]	3.7	6.1	N/A	0.3	1.0	1.0/0.5	N/A
# of compression stages	5	5	N/A	7	7	7/8	N/A
CO_2 flow [kg/s]	108.5	30.1	N/A	19.1	19.1	12.1/6.9	N/A
Specific CO_2-compression duty [kJ/kg]	238	200	N/A	504	363	395	N/A

By dividing the amount of work needed to compress the CO_2 to the designed pressure of 110 bar with the mass flow of CO_2, a specific energy demand for compressing the CO_2 as a function of the delivery pressure from the CO_2-capture unit can be calculated. The result is shown in Figure 61.
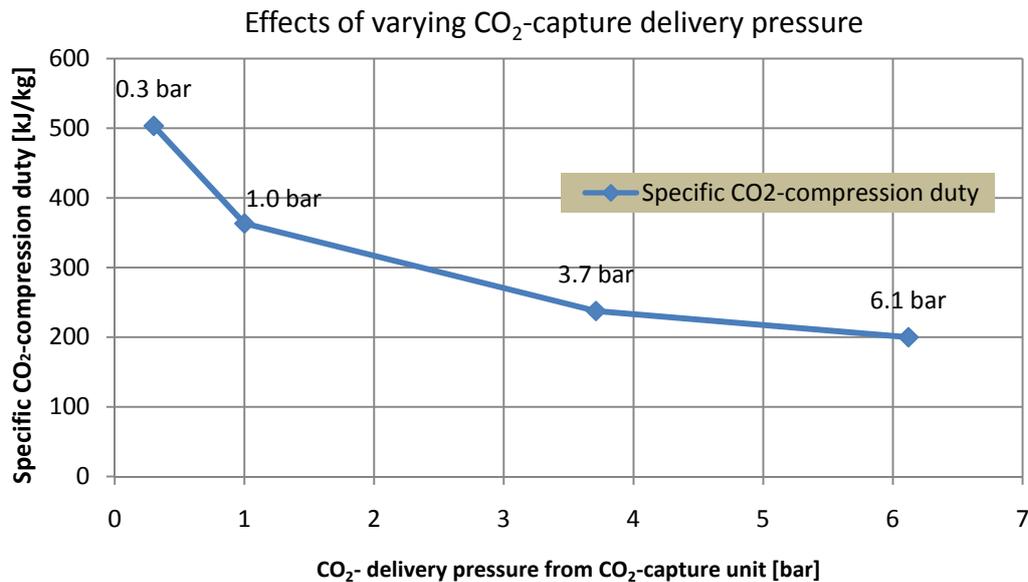
Figure 61. Effects of varying CO2-capture CO2 delivery pressure

It should be noted that only pure CO_2 is simulated to enter the compressor stages.

Syngas out of gasifier

Table 18. Syngas characteristics out of gasifier

Syngas out of gasifier –Composition (with N_2 dilution) in vol%							
Temperature [°C]	1441	1440	920	920	920	920	950
Pressure [bar]	42.5	77.0	71.8	69.0	65.9	64.9	65.0
LHV [MJ/kg]	9.8	9.4	5.0	5.5	4.9	4.9	5.9
H_2	23.7%	22.7%	14.1%	13.9%	13.8%	13.8%	16.5%
CO	54.9%	53.1%	18.9%	18.7%	18.6%	18.5%	34.9%
CO_2	5.4%	5.4%	9.3%	9.3%	9.4%	9.4%	2.1%
CH_4	0.0%	0.0%	3.9%	3.9%	3.8%	3.8%	0.0%
N_2	5.8%	8.4%	41.5%	41.7%	41.8%	41.9%	41.7%
O_2	0.1%	0.1%	0.0%	0.0%	0.0%	0.0%	0.0%
H_2O	8.6%	8.6%	10.7%	10.8%	10.8%	10.9%	3.7%
H_2S	0.7%	0.6%	0.4%	0.4%	0.4%	0.4%	0.4%
Ar	0.9%	0.9%	0.5%	0.5%	0.5%	0.5%	0.6%

Syngas into combustor

Table 19. Fuel characteristics into the combustor and sulphur level indications before the turbine and to the stack

				IGTC HTW air				IGTC MHI air
		IGCC	IGTC Oxygen	MEA	ECO_2	ECO_2 Combi	No CO_2-capture	No CO_2-capture
Fuel Into Combustor	Temperature [°C]	70	160	500	500	500	500	500
	Pressure [bar]	20.1	65.1	71.8	69.0	65.9	64.9	65.0
	LHV [MJ/kg]	7.5	22.5	3.8	3.8	3.8	3.8	4.6
	LHV [MJ/nm³]	5.1	7.9	3.6	3.6	3.6	3.6	4.3
	Wobbe index [MJ/m³]	6.9	14.8	4.1	4.1	4.1	4.1	4.9
	H_2S (vol%)	0.0%	0.0%	0.3%	0.3%	0.3%	0.3%	0.3%
	H_2S (mass%)	0.0%	0.0%	0.4%	0.4%	0.4%	0.4%	0.5%
Gas into the turbine	SO_2 (vol%)	0.0%	0.0%	0.1%	0.1%	0.1%	0.1%	0.1%
	SO_2 (mass%)	0.0%	0.0%	0.3%	0.3%	0.3%	0.3%	0.3%
Flue gas to stack	SO_2 ppm (vol)	12	29	21	21	21	18	18
	SO_2 ppm (mass)	28	69	48	48	47	39	39

Table 20. Syngas composition into the combustor

				IGTC HTW air				IGTC MHI air
		IGCC	IGTC Oxygen	MEA	ECO_2	ECO_2 Combi	No CO_2-capture	No CO_2-capture
Syngas into Combustor	H_2	48.2%	74.9%	10.0%	10.0%	9.9%	9.9%	11.9%
	CO	1.3%	2.0%	13.5%	13.4%	13.3%	13.3%	25.3%
	CO_2	2.4%	4.0%	6.6%	6.7%	6.7%	6.8%	1.5%
	CH_4	0.0%	0.0%	2.8%	2.8%	2.8%	2.8%	0.0%
	N_2	46.8%	8.5%	29.6%	29.8%	30.1%	30.2%	30.1%
	O_2	0.4%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%
	H_2O	0.2%	9.6%	36.9%	36.7%	36.5%	36.4%	30.4%
	H_2S	0.0%	0.0%	0.3%	0.3%	0.3%	0.3%	0.4%
	Ar	0.7%	1.0%	0.3%	0.3%	0.3%	0.3%	0.4%

Gas turbine characteristics

Table 21. Gas turbine characteristics and combustor flows

				IGTC HTW air				IGTC MHI air
		IGCC	IGTC Oxygen	MEA	ECO_2	ECO_2 Combi	No CO_2-capture	No CO_2-capture
Into combustor	Inlet humidification temperature [°C]	N/A	N/A	59	50	29	N/A	N/A
	Fuel flow [kg/s] (of which is water inj. [kg/s])	111.1 (0)	10.3 (2.1)	45.1 (10.6)	45.1 (10.5)	45.0 (10.3)	44.9 (10.2)	41.2 (9.3)
	Fuel flow [m^3/s]	9.9	0.7	1.8	1.9	1.9	2.0	1.8
	Air flow [kg/s]	468.5	77.6	66.8	63.2	59.2	57.9	59.3
	Steam injection [kg/s]	N/A	36.3	5.3	8.2	11.4	12.4	21.6
	Flue gas flow out of combustor [kg/s]	579.6	124.2	117.2	116.4	115.5	115.2	122.2
	Total LHV fuel input [MW]	836	230	172	171	170	170	189
Reduced mass flow	HP turbine [kg/s]	N/A	5.6	4.6	4.7	4.8	4.9	5.2
	LP turbine [kg/s]	N/A	17.2	13.8	14.2	14.8	15.0	16.0
Pressure	HP compressor exit [bar]	N/A	62.6	69.3	66.5	63.4	62.4	62.5
	LP compressor exit {bar]	N/A	19.9	22.7	21.5	20.2	19.8	19.8

Generalisation of the air blown IGTC with CO_2-capture

The air blown IGTC systems are humidified in order to make enough heat available to operate the post-combustion CO_2-capture process, as discussed previously. The four heat sources available for a post-combustion CO_2-capture unit are, as previously shown in Figure 11:

- Heat from flue gas condenser
- Heat from inter cooling of the CO_2-compression
- Excess steam from the HRSG
- Steam from the ash boiler

The unknown factors to be able to completely generalize the results of the HTW gasifier with CO_2-capture are the heat available at certain temperatures in the flue gas condenser, FGC1, and the heat available from the inter cooling of the CO_2-compression.

To achieve a more reliable generalization the small amount of heat available from the CO_2-compression stage is discarded. As can be seen in Table 7 the heat from this source only offers about 1.5 MW at a temperature above 75°C (which is what the ECO_2 process requires with current assumptions).

By running multiple simulations the characteristics of the condenser can however be determined. Figure 62 shows the condensing curves for the air blown HTW case. Due to larger load on the economiser during low inlet humidification, there is less heat left to be utilized in the flue gas condenser as the inlet humidification is decreased.
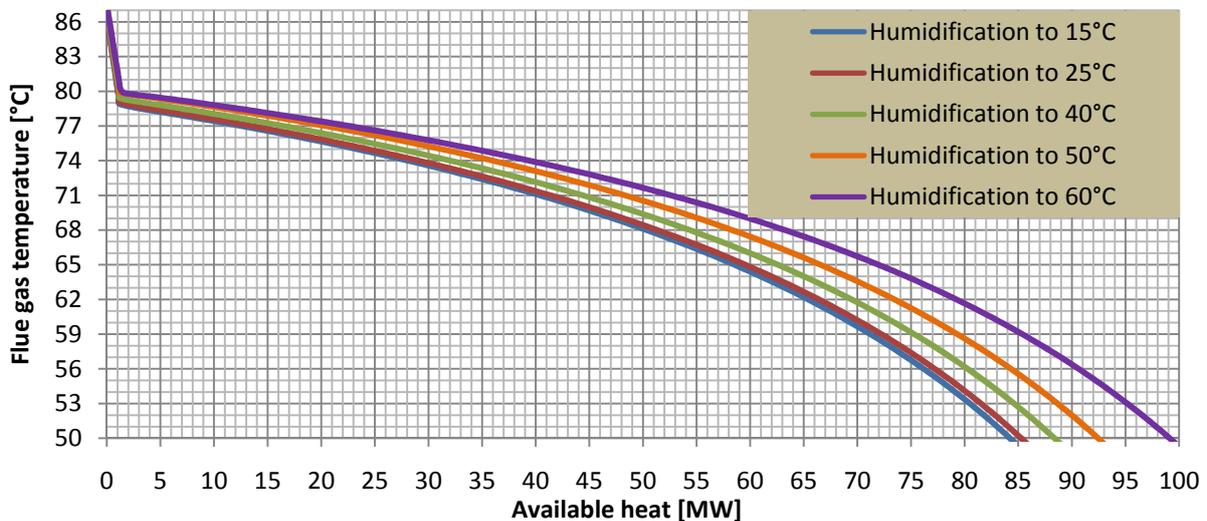
Figure 62. Available heat in the flue gas condenser depending on inlet humidification temperature

A larger size version of Figure 62 is shown as Figure 56 in Appendix C together with condensation curves for the intermediate temperatures.

As the inlet humidification increase, more excess steam is available from TopCycle unit as less steam can be provided to the combustion chamber. At the same time the steam produced from the ash boiler, which is proportional to the fuel input, is basically constant.

By knowing the inlet humidification temperature, the gross electrical efficiency can be estimated and then the net electrical efficiency can be readily calculated since the electrical losses are more or less directly proportional to the fuel input and the carbon capture rate. The efficiency in Table 22 is calculated with the specific electrical demand of the ECO_2 process, with the same characteristics as in previous simulations and the capture rate is assumed constant at 90%. The delivery pressure of the CO_2 from the CO_2-capture unit to the CO_2-compression stage is kept constant at 1 bar. Any change in the assumption of the delivery pressure would increase the CO_2-compression penalty. It should be noted that Table 22 is only valid with the low temperature dryer.

Table 22. The effects of inlet humidification of the TopCycle unit with ECO_2 post-combustion CO_2-capture and the low temperature dryer

Air inlet humidification temperature [°C]	Fuel Input [MW]	TopCycle output [MW]	Excess Steam from TopCycle [MW]	Steam from Ash boiler [MW]	Total Steam to CO_2 stripper [MW]	Net Power [MW]	Net efficiency
15	184.3	103.5	5.6	8.4	14.0	89.6	48.6%
20	184.4	102.7	7.3	8.4	15.6	88.7	48.1%
25	184.5	101.8	9.1	8.4	17.5	87.8	47.6%
30	184.5	100.8	11.2	8.4	19.6	86.8	47.0%
35	184.7	99.6	13.7	8.4	22.1	85.6	46.3%
40	184.8	98.2	16.7	8.4	25.1	84.1	45.5%
45	185.0	96.4	20.4	8.4	28.8	82.4	44.5%
50	185.2	94.3	25.0	8.4	33.4	80.2	43.3%
55	185.5	91.4	30.9	8.4	39.3	77.3	41.7%
60	185.9	87.5	38.6	8.4	47.0	73.4	39.5%

With a curve for the heat available from the flue gas condenser, and a table with the amount of steam the system is able to supply to the CO_2-stripper, the only missing information is at which humidification level the condenser can fulfil the temperature and heat requirement of CO_2-stripper.

If the standard ECO_2 CO_2-capture unit is considered, requiring the same amount of specific heat as before, 2.9 MJ/kg CO_2, the amount of humidification to make a feasible solution can be deduced from Figure 63 (which is zoomed in version of Figure 62) together with Table 23. In Table 23 the total amount of available steam is presented with the calculated required amount of heat for the ECO_2-unit. The difference between these two must then be supplied by the flue gas condenser, in order to fulfil the heat demand of the CO_2-capture unit. If the temperature requirement of the ECO_2 unit also is kept to the previous level of minimum 75°C and together with the 2 K pinch, Figure 63 shows that the feasible solution lies just above a humidification temperature of 50°C (by drawing a horizontal line at 75+2=77°C in Figure 63 and checking if the intersecting humidification line, ≈20 MW in the case of 50°C, matches the available condenser heat required in Table 23 which is 21.9 MW at humidification to 50°C). In the previous simulations, the ECO_2 unit together with TopCycle required a inlet air humidification temperature of 50°C and the slightly higher need for humidification in this case is due to the decision to discard the available heat from the CO_2-compression process.

Table 23. Required condenser heat for the ECO₂ process with a specific heat demand of 2.9 MJ/kg CO₂

Air inlet humidification temperature [°C]	Fuel input [MW]	Net power output [MW]	Net efficiency	Total available Steam [MW]	ECO₂ heat demand [MW]	Condenser heat needed [MW]
15	184.3	103.5	48.6%	14.0	55.0	41.0
20	184.4	102.7	48.1%	15.6	55.0	39.4
25	184.5	101.8	47.6%	17.5	55.0	37.5
30	184.5	100.8	47.0%	19.6	55.0	35.4
35	184.7	99.6	46.3%	22.1	55.1	33.0
40	184.8	98.2	45.5%	25.1	55.1	30.0
45	185.0	96.4	44.5%	28.8	55.2	26.4
50	185.2	94.3	43.3%	33.4	55.3	21.9
55	185.5	91.4	41.7%	39.3	55.3	16.1
60	185.9	87.5	39.5%	47.0	55.4	8.4
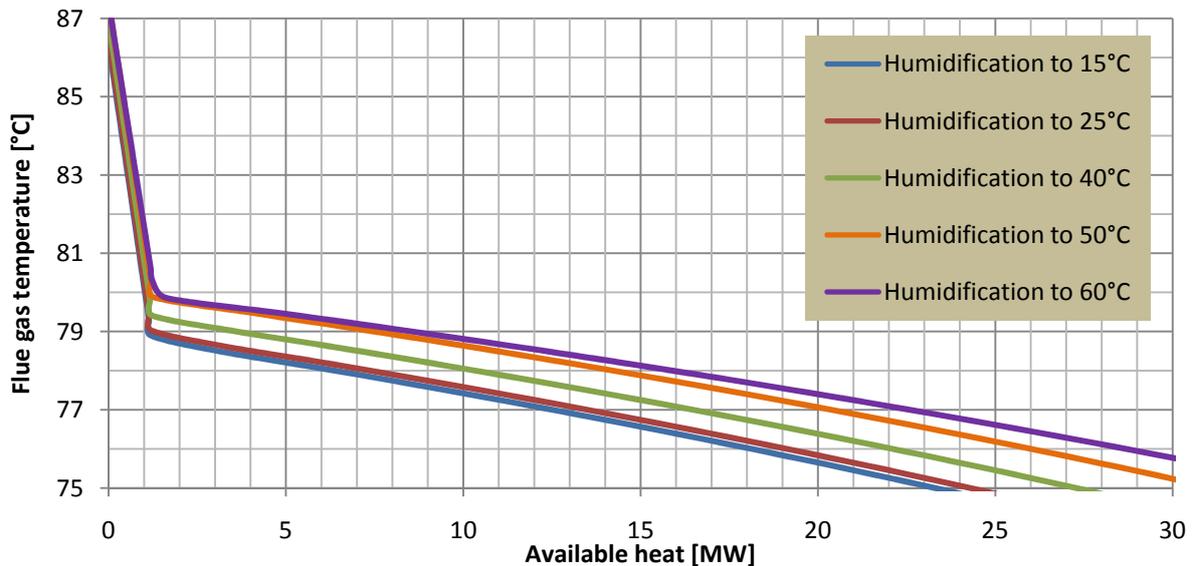
Figure 63. Available heat in the flue gas condenser depending on inlet humidification – zoomed in at 75-87°C flue gas temperature

In Table 23 can be seen that the resulting efficiency would be 43.3%, which is equal to the previously simulated HTW with ECO₂ with 90% carbon capture rate and low temperature dryer.

The next step in generalising is to change the amount of heat needed in the CO₂-stripper, from the previously used default value of 2.9 MJ/kg CO₂, but also changing the temperature at which this heat has to be supplied. By applying the same procedure as before, a table like the one shown above (Table 23) can be made for an arbitrary value of the specific heat requirement of the CO₂-stripper.

If the specific heat requirement of the stripper increases, so does the needed heat from the flue gas condenser. With an increase in inlet humidification more excess steam is produced from the TopCycle unit, as previously stated, and therefore less heat from the condenser is needed when

increasing the inlet humidification. This is represented in Figure 64 by the red dashed lines which each represent a certain specific amount of heat requirement for the CO_2-stripper.

The solid black lines in Figure 64 represent the heat available in the flue gas at a certain temperature, varying with the inlet humidification temperature. With more inlet humidification, the heat available at a certain temperature increases, which results in the positive gradient of the parallel black lines. Please note that there is a 2 K difference in flue gas temperature between each line.

The intersection of a red dashed line and a solid black line represents at which inlet humidification a feasible solution is achieved. For example if the CO_2-stripper requires 2.5 MJ/kg CO_2 at a temperature of 75°C with a 2 K pinch in the condenser, the heat available to the stripper is all the heat which follows the "77°C"-line. The lines intersect at around 45°C, which means the inlet air has to be humidified to 45°C in order to successfully operate the CO_2-stripper. With the inlet humidification known, the efficiency can be estimated from Figure 65, to about 44.5%. If the condenser had a pinch temperature of 0 K the inlet air would instead have to be humidified to about 38°C, resulting in an efficiency of 46%, an increase of 1.5%pt.

When the temperature of the flue gas in Figure 64 drops below 77°C the heat required for the low temperature dryer comes into play. The low temperature dryer is taken into account with identical characteristics as in previous simulations, requiring a temperature of 75°C with a 2 K pinch and supplying a water temperature of 54°C to the flue gas condenser. Therefore the distance between the black temperature lines decrease when the flue gas temperature drops below 77°C, there is simply less heat available beyond this point, due to the low temperature dryer.
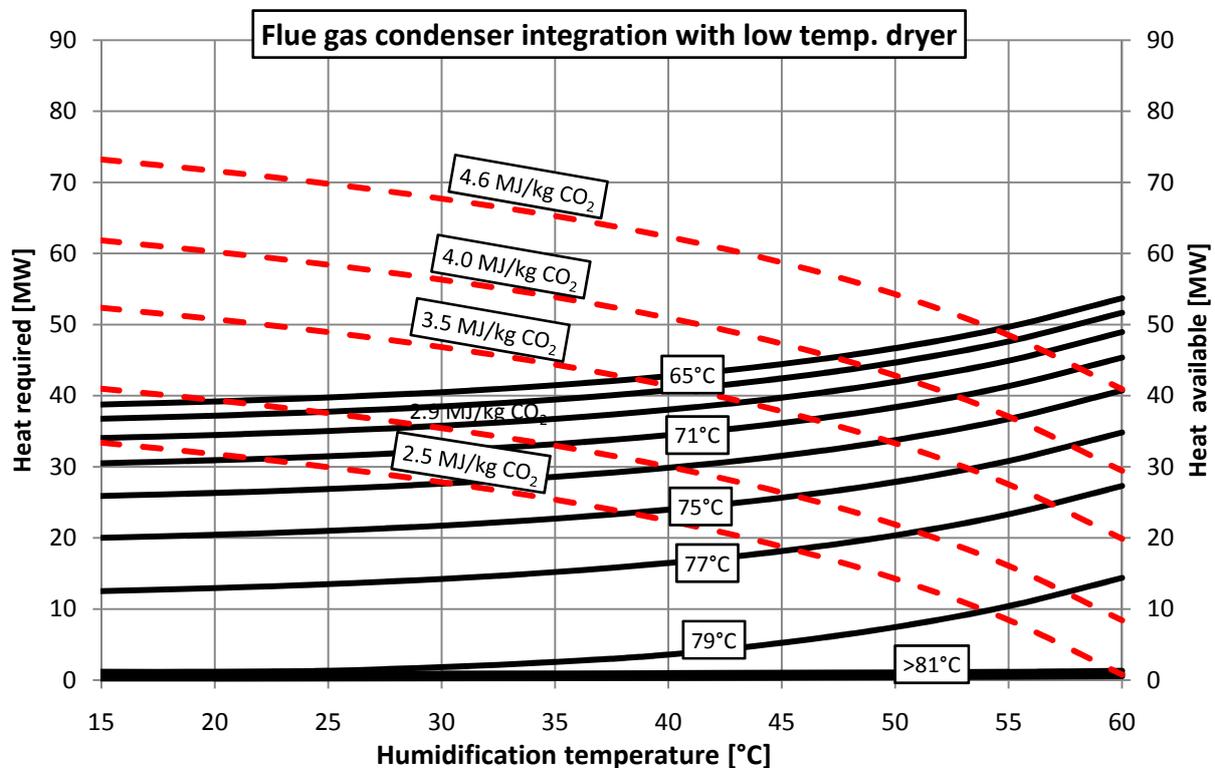
Figure 64. Flue gas condenser characteristic of air blown IGTC post-combustion CO_2-capture with low temperature dryer

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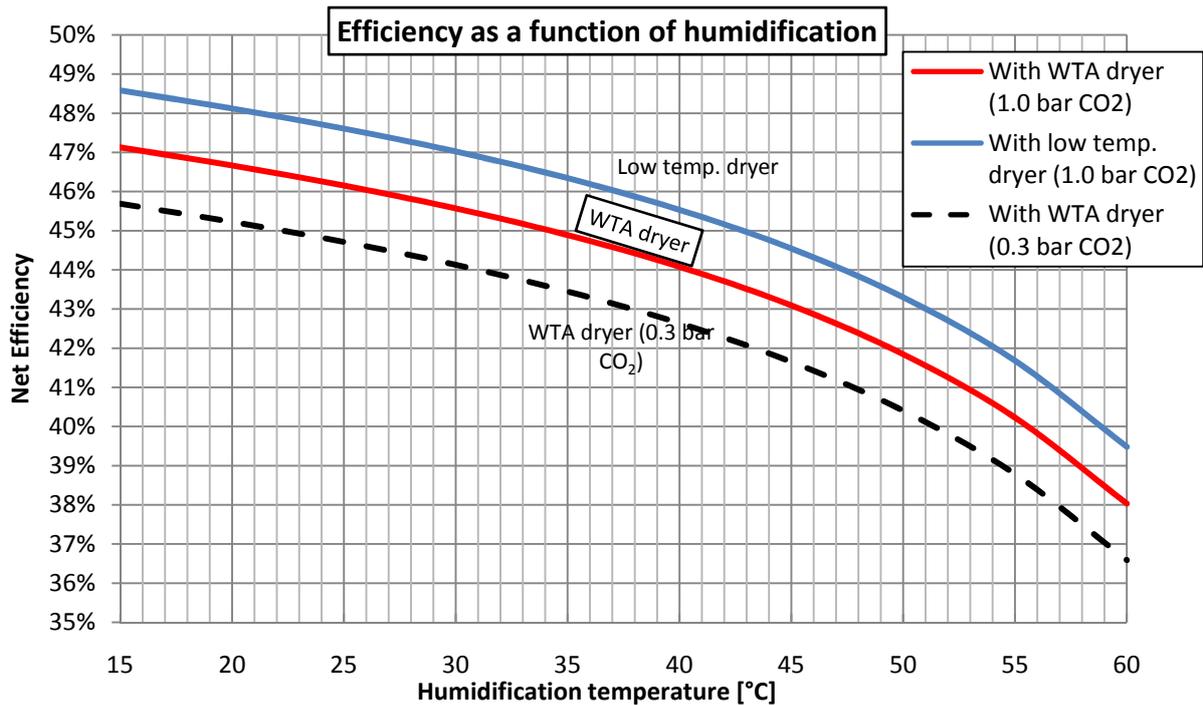
Figure 65. Net efficiency as a function of inlet humidification temperature

In a similar fashion, Figure 66 represents the characteristics of the post-combustion CO_2-capture IGTC system with the WTA dryer. The difference is that no flue gas heat is required to operate a low temperature dryer and that steam is needed to operate the WTA dryer. The result is that the red dashed lines, representing the required heat, are pushed to a higher starting position - while at the same time - more heat is available in the flue gas.

Integrated into Figure 65 is also the efficiency of the IGTC system with the WTA dryer as well as a special case, showing the impact of a lower pressure discharge of CO_2 from the CO_2-separation unit, leading to higher compression work.
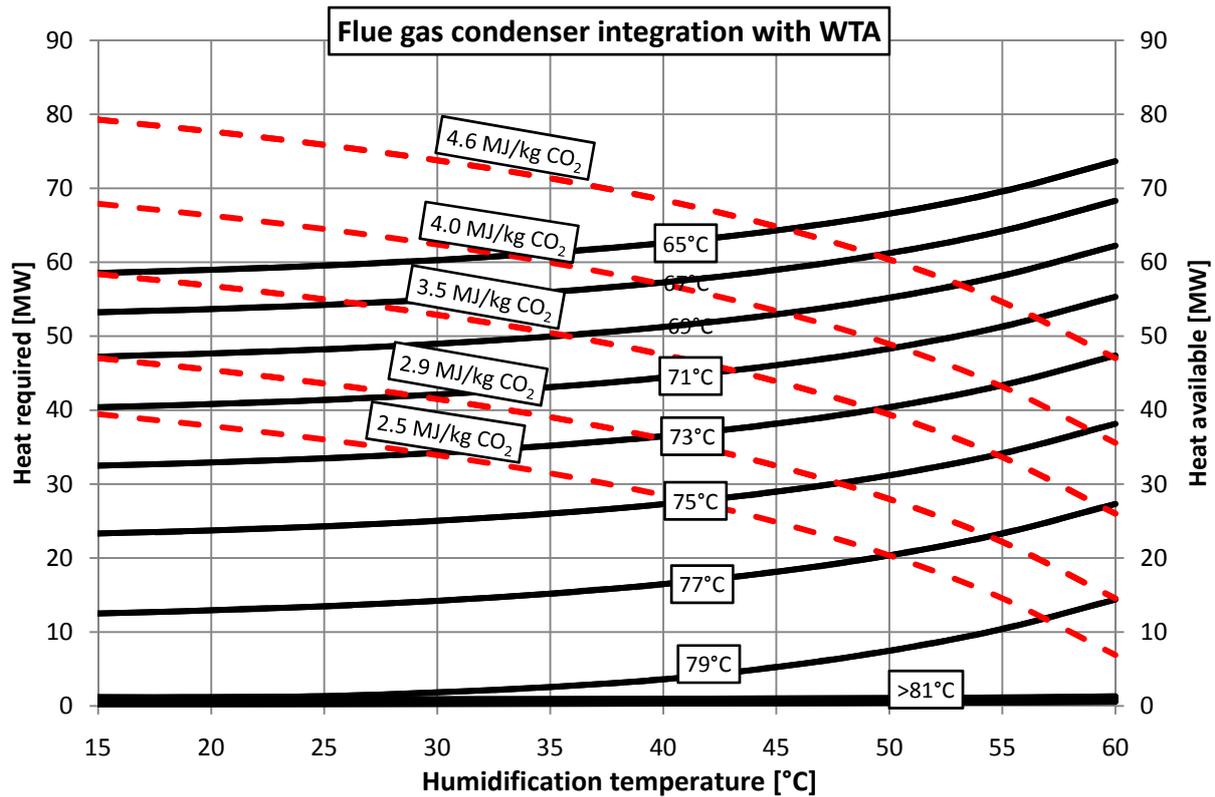
Figure 66. Flue gas condenser characteristic of air blown IGTC post-combustion CO$_2$-capture with WTA dryer

Larger copies of Figure 64, Figure 65 and Figure 66can be found in Appendix C.

A further complication of the heat integration is when the "ECO$_2$ Combination"-option, which has two low pressure strippers, is integrated with the WTA dryer. The combination would have a high total heat demand. But when switching to the WTA dryer, which has a much higher electricity demand but a lower heat demand, the humidification can be omitted and the system is still able to operate a dryer and CO$_2$-capture unit. This is due to the sub-atmospheric stripper which is able to extract usable heat from the flue gas at a very low temperature. The absence of the dryer enables higher CO$_2$-separation load on the sub-atmospheric ECO$_2$-stripper, unique to the "ECO$_2$ Combination"-option. The absence of inlet humidification counters the high electricity demand of the WTA dryer and the efficiency only drops by 0.2%, to an overall efficiency of 46.7%. The Ebsilon drawing for this system can be seen in Appendix A.