Perspectives on gasification systems to produce energy carriers and other chemicals with low CO_2 emissions

Techno-economic system analysis on current and advanced flexible thermo-chemical conversion of fossil fuels and biomass

Hans Meerman

Perspectives on gasification systems to produce energy carriers and other chemicals with low CO_2 emissions: Techno-economic system analysis on current and advanced flexible thermo-chemical conversion of fossil fuels and biomass

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Techno-economic system analysis on current and advanced
flexible thermo-chemical conversion of fossil fuels and biomass

Perspectief op vergassingssystemen voor het produceren van energiedragers en andere chemicaliën met lage CO_2 emissies

Techno-economische systeemanalyse van huidige en toekomstige flexibele
thermochemische omzetting van fossiele brandstoffen en biomassa

(met een samenvatting in het Nederlands)

Proefschrift

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Voor mijn ouders

We do not inherit the Earth from our ancestors, we borrow it from our children
native American proverb

Contents

1 Introduction

1.1 Background

Energy

Our modern society requires access to abundant, affordable, reliable and environmentally sound sources of energy. As population and living standards are increasing, so is our energy demand (see Figure 1.1). In 2009, global primary energy demand exceeded 500 EJ, representing a doubling in about 35 years. About 80% of this demand is covered by fossil fuels.[1,2,3] In its Energy Technology Perspectives 2010 report, the International Energy Agency (IEA) presented a business as usual scenario projecting an increase of the global energy demand to over 750 EJ in 2035.[2] Similar trends are forecasted by other studies.[1,2,4,5,6,7,8] Most of these studies indicate that fossil fuels will continue to play an important role for many decades to come.

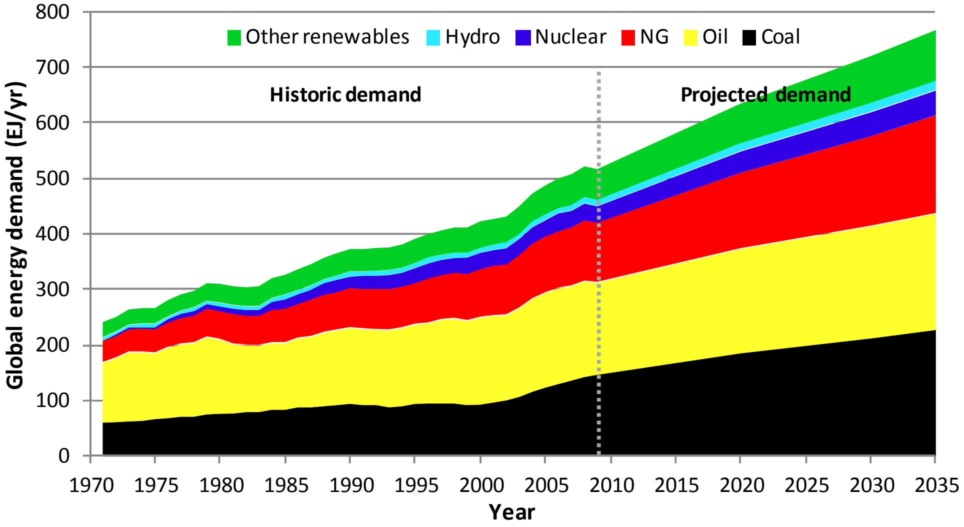

Figure 1.1 Global historic energy demand and future projection of the IEA business as usual scenario.[2,3]

Unfortunately, the use of fossil fuels has several important drawbacks. First, most of the conventional oil and gas reserves are located in a few world regions. Consequently, many countries are dependent on few other countries for their energy supplies.[2,4,9,10] Second, keeping up extraction from easily recoverable fossil fuel sources to the ever-increasing demand for fossil fuels proves difficult, causing impacts on the price of fossil fuels.[2] Third, the combustion of fossil fuels is responsible for the largest share of anthropogenic greenhouse gas (GHG) emissions into the atmosphere.[11] And last but not least, due to the depletion of easily recoverable fossil fuel sources, extraction of unconventional fossil fuel sources, such as tar sands and shale gas, and extraction in environmentally sensitive areas are expected to increase, resulting in higher GHG emissions and increased risks of severe environmental damage.[5,12]

Climate change

Emissions of GHG are linked to global warming[1], with CO_2 being responsible for more than 50% of the direct radiative forcing of GHG.[11] In 1992, the United Nations Framework Convention on Climate Change (UNFCCC) was established with the goal of achieving stabilisation of atmospheric greenhouse gas concentrations at a level that would prevent dangerous anthropogenic interference with the climate system.[13] By 2005, average global temperatures were already about 0.8°C higher than pre-industrial levels (see Figure 1.2).[11] The 4th Assessment Report of the International Panel on Climate Change (IPCC) projects that a scenario focussing on economic development and reliance on fossil fuels (A1-fossil intensive: A1FI) could result in an increase in the average temperature on Earth of 2.9-6.9°C above pre-industrial values (with a best estimate of 4.5°C) by 2100.[2] A temperature change of this magnitude would be unprecedented in human history and probably result in detrimental impacts on climate, biodiversity, food production and sea level rise. The same report also projects that in a scenario which assumes strong policies to mitigate GHG emissions, the increase in average temperatures could be limited to 1.6-3.4°C, with a best estimate of 2.3°C.[11] In 2009 and 2010, the Conference of the Parties (COP) to the UNFCCC agreed that the increase in global mean temperature should be kept below 2°C, compared to pre-industrial levels.[14,15]

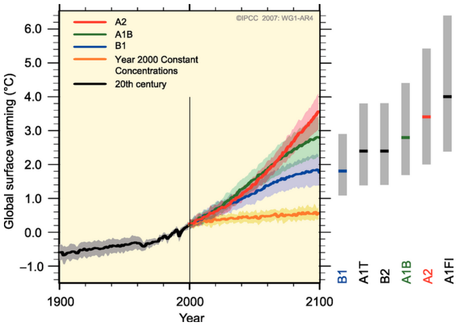

Figure 1.2 Historic climate change and projected climate change under different GHG mitigation policies.[11]

CO_2 emissions

Keeping the increase in global mean temperature to less than 2°C means that the atmospheric CO_2 concentration needs to stabilise most probably at 350-400 ppm(v) CO_2 (445-490 ppm(v) CO_2-equivalent).11] The current proven reserves of fossil fuels contain around 800 Gt carbon,[1] which could result in almost 3000 Gt of CO_2 emissions. This value is roughly the same amount of CO_2 currently present in the atmosphere. Maintaining the increase in global mean temperature below 2°C implies that only 1000-1625 Gt CO_2 can be emitted between 2000-2100, the so-called carbon budget.[16,17] Between 2000 and 2010, already 330 Gt CO_2 were emitted, leaving room for only 670-1295 Gt of anthropogenic CO_2 emissions in the coming decades. Without additional CO_2 abatement policies, this value is expected to be reached within a few decades.[2,4,6,16] Keeping within the indicated carbon

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[1] For more information on climate change and its effects we refer to the 4th assessment report of the IPCC.[11]

budget will require reductions in CO_2 emissions of most likely 50-80% in 2050 compared to the 2000-level.[11,15,18]

1.2 Mitigation strategies

There are several routes to reduce global CO_2 emissions. The most obvious one is to consume less primary energy. This, however, has proved to be difficult in the past. Keeping the global energy demand constant with an increasing population and increasing average standard of living is already a daunting challenge, as indicated by several studies.[2,4,5,6,7] Nevertheless, there is a large potential to increase the efficiency of energy conversion processes that needs to be targeted. The second option is to use fossil energy sources which are less carbon-intensive or to switch to renewable energy sources. This includes switching from fossil fuels to renewable and nuclear energy sources, but also from coal to natural gas. In the scenarios from e.g., the IEA, IPCC, BP, Greenpeace, WWF, and in the Global Energy Assessment (GEA), an increased utilisation of renewables is projected but at various degrees depending on, among others, assumptions related to technological improvements in the energy system and climate policies. A third option is to reduce the CO_2 emitted by the use of fossil fuels (and biomass) using carbon dioxide capture, transport and storage (CCS) technologies. All studies just mentioned project that none of the three options by themselves can reach the 2°C target, and that a portfolio of mitigation options has to be implemented in order to meet the GHG emission mitigation targets (see e.g., Figure 1.3).[2,4,5,6,7,8,19]

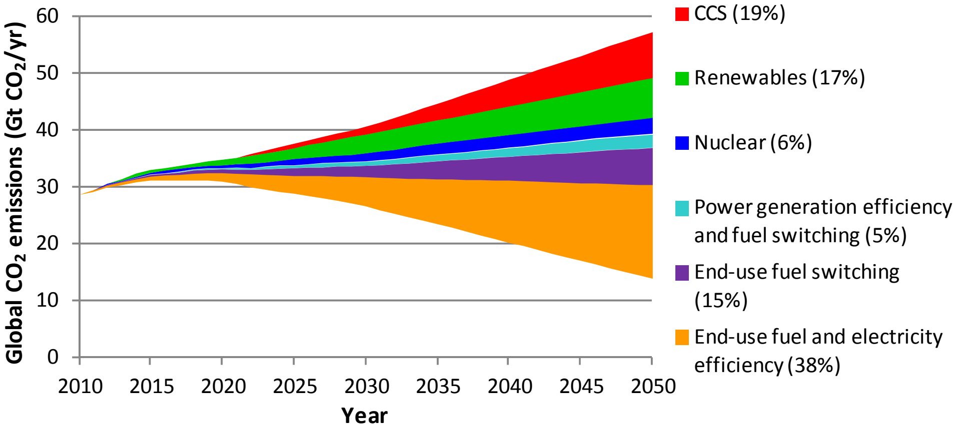

Figure 1.3 Contribution of different CO_2 mitigation strategies in the IEA Blue map scenario, aiming at stabilising atmospheric GHG emissions at 450 ppm(v) CO_2-eq.[4]

Sectors

Almost 85% of the 29 Gt CO_2 emitted by the combustion of fossil fuels in 2009, were produced by the power, transport and industry sectors combined (see Figure 1.4).[20] Each sector has its own unique characteristics and challenges and, therefore, needs specific strategies to mitigate GHG emissions.

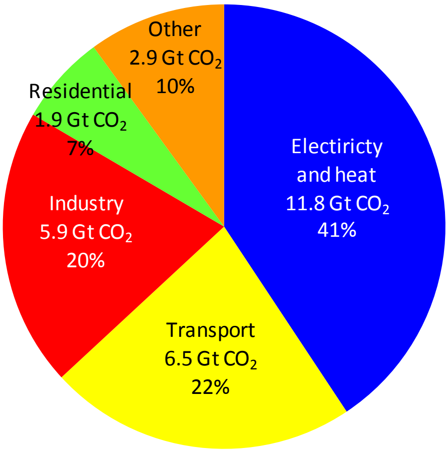

Figure 1.4 Global direct annual CO_2 emissions by sector in 2009.[20]

The *power sector* consumed 190 EJ of primary energy in 2009 of which almost 50% came from coal and another 25% from natural gas and oil.[2] Most CO_2 emissions from this sector stem from large, stationary sources, making this sector suitable for end-of-pipe solutions, such as CCS. Literature studies indicate that the most cost-effective CO_2 mitigation options for the short to mid term are fuel switching (co-firing biomass in coal-fired power plants or replacing coal-fired power plants by natural gas-fired power plants) and applying CCS.[2,4,5,6,7]

The *transport sector* consumed almost 100 EJ of primary energy in 2009, of which more than 90% came from oil.[2] The CO_2 emissions from this sector originate from numerous small, mobile sources, making end-of-pipe solutions unattractive.[34] Therefore, decarbonising this sector is only realistic via decarbonising the fuel. A main obstacle in any transition towards a new transportation fuel is the large interdependence between car, fuel and infrastructure, making any transition difficult and expensive if the current infrastructure and car park cannot be used.[21] The current infrastructure is designed for liquids with a high energy density, like liquefied petroleum gas (LPG), gasoline, diesel and Fischer-Tropsch liquids (FT-liquids are a mixture of linear hydrocarbons of different length). Literature indicates that a promising CO_2 mitigation option for the shorter to mid term is replacing conventional diesel and gasoline with liquid 2nd generation biofuels. When producing these fuels, CCS can be applied to reduce GHG emissions even further.[21,22,23,24,25]

The *industry sector* consumed almost 90 EJ of primary energy in 2009 of which 40% came from coal, 28% from natural gas and 20% from oil. Almost 75% of the direct CO_2 emissions from this sector originate mainly from the production of iron and steel, cement and chemicals.[4] These emissions come from large, stationary sources, making the industry sector suitable for CCS. For more information on CO_2 mitigation strategies in the industry sector, we refer to Kuramochi.[26]

Biomass and biofuels

Sustainable biomass is an attractive CO_2 mitigation strategy as it can be converted into carbon neutral building blocks for the chemical industry or into high energy density liquids for the transport sector.[7] Biomass is the only renewable carbon source for the chemical industry and is considered a key alternative for the transport sector in the short term.[2,4,7] In 2009, 10% of the global primary energy demand (about 50 EJ) was supplied by biomass, of which 11-12 EJ was being used in modern applications, like power plants and biofuels facilities.[7,27,28] For comparison, in the same year the total energy demand of the power and transport sectors combined was almost 300 EJ.[2] The deployment potential for sustainable biomass is uncertain, but considerable. Scenario and model analyses indicate that the technical potential of biomass energy resources may range between 50-500 EJ/yr in 2050[7,28,29,30] and the deployment potential could be in the order of 100-300 EJ/yr in 2050, depending on policies and market conditions.[7] These high deployment levels of biomass require the implementation of sustainability criteria to prevent, among others, competition with food or feed supplies, depletion of water sources, loss of biodiversity or excessive GHG emissions due to (indirect) land use change.[7,28]

In general, a distinction is made between 1st and 2nd generation biofuels. Examples of 1st generation biofuels are vegetable oils, ethanol from corn and sugarcane and biogas from organic waste, while 2nd generation biofuels include FT-liquids or ethanol from wood, grasses or residues from the forest and agricultural sectors.[7] FT-diesel has the advantage that it is almost similar to conventional diesel, allowing the use of the existing infrastructure and car park. Advantages of 2nd generation biofuels over 1st generation biofuels are the use of non-food/feed sources as feedstock and higher energy yields and (potential) savings in GHG emissions on a hectare basis.[28,31] Finally, biomass crops (including trees) for 2nd generation biofuels can be cultivated on marginal or degraded land, not only further reducing the competition with food supplies, but also offering an opportunity to restore degraded land.[32,33]

Carbon dioxide capture, transport and storage

Carbon dioxide capture, transport and storage (CCS) is the general term for processes that prevent the emission of CO_2 by long-term isolation of the CO_2 from the atmosphere.[34,35] It allows the use of fossil fuels but with significantly lower CO_2 emissions. For this reason CCS can be seen as a transitional mitigation option. It increases the available time to transform the current fossil fuel based energy infrastructure into an infrastructure based mainly on renewable sources. Moreover, when CCS is applied in combination with conversion of sustainably produced biomass, it can result in negative emissions, i.e., extracting and isolating CO_2 from the atmosphere and thus extending the carbon budget. The potential deployment of CCS depends heavily on technological improvements and climate policies. The IPCC special report on CCS indicates that the economic potential of CCS could be 220-2200 Gt CO_2 avoided cumulative by 2100, while the IEA Energy Technology

Perspectives projects a share of about 19% in 2050, equivalent to 18 Gt CO_2 emissions a year.[4,34] The European Commission's Energy Roadmap 2050 estimates that, depending on the scenario, CCS could account for 19-32% of the total mitigated CO_2 emissions in the power sector.[36] Although total mitigation costs are uncertain, it has been projected that large-scale deployment of CCS could reduce global mitigation costs by up to 40% as more expensive CO_2 mitigation options can be postponed or even avoided.[34,37,38,39,40] Furthermore applying CCS at biomass feedstocks could result in an additional reduction in global mitigation costs of about 40% relative to the option of using CCS at fossil feedstocks only.[38,39] However, to fully realise this potential requires that deployment of CCS takes off this decade. In 2011, there were only 8 operational integrated commercial scale CCS demonstration projects in the world[41] and, according to the IEA Technology roadmap on CCS, the number of projects needs to increase to around 100 projects by 2020 and to almost 3500 by 2050 if the full contribution of CCS to projected mitigation efforts is to be realised (see Figure 1.5).[37]

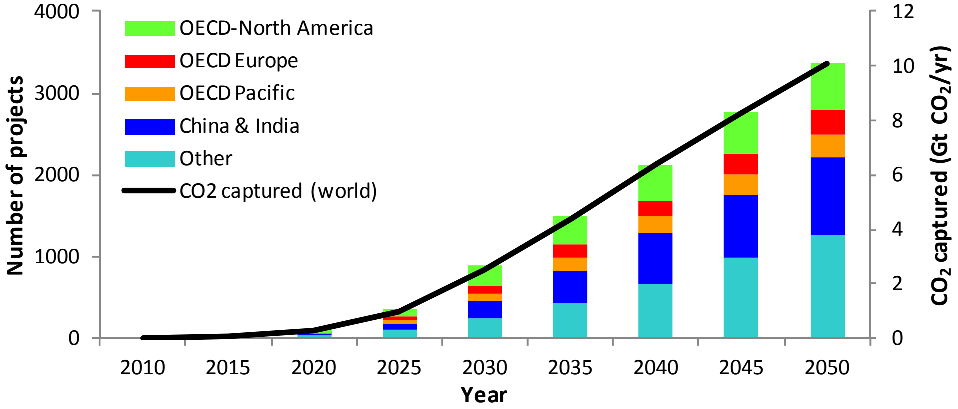

Figure 1.5 Roadmap for global deployment of CCS up till 2050 according to the IEA Technology roadmap on carbon capture and storage.[37]

The CCS chain is composed of three steps: capture, transport and storage. During the CO_2 capture step, CO_2 is extracted from industrial and energy-related sources e.g., natural gas, flue gas and syngas. This CO_2 is subsequently purified, dried and compressed. This step is generally the most expensive part of the CCS chain due to the high capital costs of the equipment and the energy use involved. Exceptions are some specific industrial processes, like ammonia or ethylene oxide production, which already produce a highly concentrated CO_2 gas stream. This reduces the CO_2 capture step to drying and compression of the CO_2.

In the transport step, the CO_2 is moved from the emission source to the storage location. CO_2 transport by pipeline is already applied in various settings, including enhanced oil recovery (EOR) and enhanced natural gas recovery (EGR). CO_2 is transported in supercritical phase to improve economics. In this phase CO_2 behaves like a liquid with a liquid-like density, while the viscosity and compressibility show a gas phase behaviour. The storage phase consists of the long-term isolation of CO_2 from the atmosphere via mineralisation or by injection of the CO_2 into underground geological locations, such as depleted gas or oil fields or saline aquifers.[5]

CO_2 capture is often divided into four categories: post-combustion capture, pre-combustion capture, oxyfuel combustion capture and capture at industrial processes.

- *Post-combustion capture* involves the extraction of CO_2 from flue gases which are produced by combusting fossil fuels or biomass with air. The flue gas is at atmospheric pressure and has a low CO_2 concentration, resulting in a low CO_2 partial pressure. Due to this low partial pressure, capture is currently achieved by chemical absorption using solvents. The regeneration of these solvents involves a temperature swing, which requires large amounts of energy. An advantage is that it can be applied at existing power plants (retrofit). Moreover, the CO_2 capture process does not interfere with normal operations, meaning that the availability of the facility is not affected.
- *Pre-combustion capture* is the extraction of CO_2 from an energy-rich gas stream prior to combustion. This prevents N_2 dilution and, if the facility operates at elevated pressure, allows the removal of the CO_2 at this elevated pressure. Because of the high partial pressure of CO_2, physical solvents can be used instead of chemical solvents to capture the CO_2. As regeneration of physical solvents can be achieved by pressure swing instead of temperature swing, this results in a lower energy consumption in the capture process and thus (in principle) lower CO_2 capture costs.
- *Oxyfuel combustion capture* entails the extraction of CO_2 from flue gas consisting mainly of CO_2 and steam as the fossil fuel or biomass is combusted with (almost) pure O_2. Removal of the steam by condensation results in a CO_2 purity of 80-98%.[34] In this process high temperatures in the combustion process can be reached resulting, in principle, in more efficient Brayton (for gas turbines) or Carnot (for boilers) cycles. Currently, metallurgical constrains prevent the utilisation of these high temperatures. Therefore, CO_2 or steam needs to be recycled to reduce the operating temperature. A main drawback of this option is the need for (currently expensive) pure O_2.
- For *capture from industrial processes* generally one of the above described options is applied, depending on the process conditions. In some industrial processes (e.g., ammonia production and natural gas processing) CO_2 is already extracted as part of the industrial process. Applying CCS at these sources, therefore, significantly reduces the additional costs for CO_2 separation. For more information on the possibilities and potential of CO_2 capture at industrial processes, we refer to Kuramochi.[26]

There are three main approaches to separate CO_2 from other gases, namely using sorbents, membranes or cryogenics.[34,35]

Separation by sorption means bringing the gas stream containing CO_2 in close contact with a sorbent (or solvent when in liquid phase). The CO_2 is absorbed into or adsorbed onto the sorbent. To release the CO_2, the sorbent is regenerated. This is commonly done by increasing the temperature for chemical sorbents or decreasing the pressure for physical sorbents. The main disadvantage of this process is the high energy consumption in the regeneration step, especially when chemical solvents are used.[34,35]

Cryogenic separation involves cooling the gas stream containing the CO_2 to below the boiling temperature of CO_2. Impurities can easily be removed from the liquid CO_2. Prior to cooling, the gas stream is dried to prevent ice formation in the heat exchangers. After compression, the CO_2 is ready for transport. An advantage of this technique is that a liquid CO_2 stream is produced, which is beneficial for certain transport options, such as transport by ship. The main disadvantage is that the entire gas stream needs to be cooled down, resulting in bulky equipment and a high energy consumption when separating CO_2 from diluted streams.[34,35]

Membrane separation involves passing the gas stream containing CO_2 along a (polymeric, metallic or ceramic) material which is selectively permeable for only CO_2 or permeable for all other relevant gases except CO_2. As only the permeable gases pass through the membrane, this results in a pure CO_2 stream. The driving force is the difference in partial pressure across the membrane, making high pressure streams more suitable than low pressure streams for membrane separation. The advantage of this technique is the low energy consumption when a pressurised gas stream is available. Although this process can also be applied at low pressure streams, it requires creating a vacuum, which drastically increases energy consumption. Furthermore, reliable and affordable CO_2 selective membranes are currently not commercially available and CO_2 removal by membrane separation has not yet been tested in large-scale industrial facilities.[34,35]

1.3 Integrated gasification poly-generation facilities

A key technology which can play a role in the development of a sustainable energy infrastructure is the production of fuels and other chemicals via gasification. Interest in gasification has increased significantly in the last few decades, resulting in an installed syngas capacity of around 70 GW_{th} in 2010. Projections from the industry indicate that this capacity could almost double by 2016 (Figure 1.6).[42] Gasification technology (partly combined with CCS) is seen as an interesting option for CO_2 mitigation. In the Blue energy scenario of the IEA, for instance, installed gasification capacity is projected to exceed 1500 $GW_{th\ syngas}$ in 2050[27] and the World Energy Technology Outlook 2050 H_2-case scenario projects that more than 900 $GW_{th\ syngas}$ is needed for hydrogen production alone.[43]

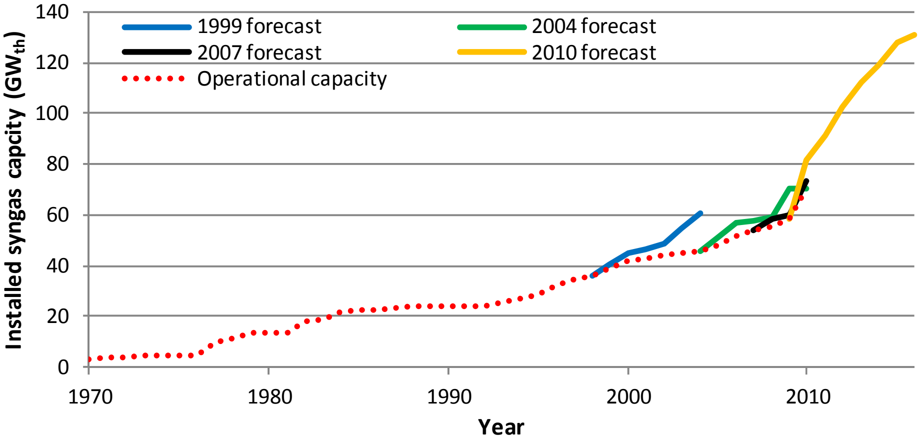

Figure 1.6 Worldwide installed syngas capacity (GW_{th}) according to Childress.[42,44,45,46]

The main reason behind the increasing interest in gasification is its ability to convert carbonaceous feedstocks, like coal, stranded natural gas and biomass, into high value products, e.g., electricity, transportation fuels and other chemicals.[42] Gasification offers additional benefits over the conventional production of energy carriers and chemicals, namely:

- The ability to use a variety of carbonaceous feedstocks without being dependent on a specific feedstock;[47]
- The ability to produce a variety of products (electricity, fuels, fertilizers and other chemicals) from the syngas, allowing flexible production and output;[47]
- Lower specific emissions of, among others, NO_x, sulphur compounds, particulate matter and heavy metals compared to conventional PC power plants and refineries;[47]
- Facilitating the more energy efficient pre-combustion CO_2 capture instead of post-combustion CO_2 capture, as the syngas has a high pressure and a high concentration of CO_2;
- The conversion of sustainable biomass feedstocks into carbon neutral transportation fuels which can be used in the current transport infrastructure;[2]
- The conversion of sustainable biomass feedstocks into carbon neutral raw materials for the chemical industry.[48]

Gasification facilities also have some disadvantages. For electricity production, integrated gasification combined cycle (IGCC) facilities are more complex and expensive than conventional natural gas combined cycle (NGCC) and pulverised coal (PC) power plants. When considering production of transportation fuels and other chemicals, the use of coal as feedstock in the absence of CCS can double specific CO_2 emissions compared to producing the chemical in conventional refineries using natural gas or crude oil.[22,23,24] Lastly, the commercial experience with gasification facilities is still limited. This, combined with the complex operation of the facilities, hinders its deployment.

Process description

An integrated gasification poly-generation (IG-PG) facility consists of various key processes. A schematic overview is given in Figure 1.7. First, a solid carbonaceous feedstock is fed into an entrained flow (EF) gasifier. The high operating temperature (>1500°C) of this type of gasifier breaks the feedstock down into syngas, consisting mainly of CO, CO_2, H_2 and H_2O. The required temperature is obtained by combusting part of the feedstock inside the gasifier. A sub-stoichiometric amount of (almost) pure O_2, produced in an air separation unit (ASU) functions as oxidant. Pure O_2 is used to prevent N_2 dilution of the syngas, increasing the overall conversion efficiency and reducing the size of downstream equipment.[47,49,50] Prior to the gas cleaning, the syngas is quenched to prevent fouling of the cooling equipment. In the gas cleaning sector, impurities like halogen gases and solid particles are removed using filters and wet scrubbers. Acidic compounds are removed in the acid gas removal section (AGR).

Depending on the desired production, the H_2:CO ratio needs to be adjusted. This is done in a water-gas shift (WGS) reactor which can be located upstream or downstream the AGR unit. After cleaning, the syngas can be converted into the final product. To improve economics and the overall conversion efficiency, steam is generated at various locations throughout the facility and used for electricity production.

Key characteristic of an IG-PG facility is that CO_2 needs to be extracted from the syngas to improve the conversion efficiency of the downstream chemical or transportation fuel conversion reactors. Using that concentrated CO_2 for CCS purposes results in low CO_2 removal costs, making IG-PG facilities interesting early options for CCS implementation.

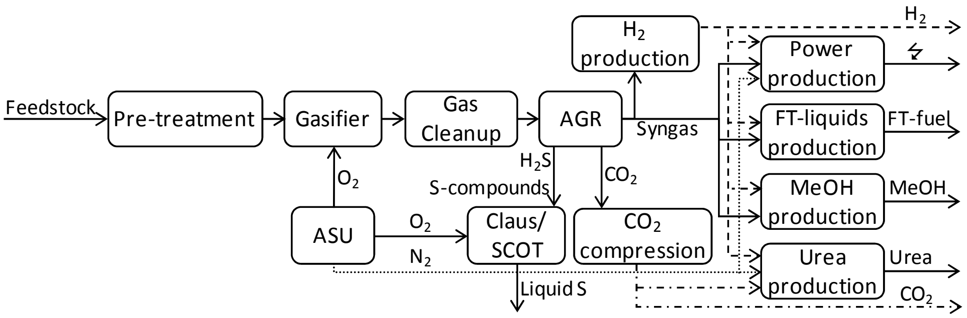

Figure 1.7 Simplified process layout of a flexible IG-PG facility with CO_2 capture using state-of-the-art technology. Waste, heat and recycle streams are not displayed.

Performance

Despite the many advantages of gasification facilities, the current globally installed gasification capacity is relatively low. A main reason is the high capital costs of gasification facilities, resulting in high production costs.[51] Literature studies indicate that an IGCC without CO_2 capture can produce electricity for 50-94 €/MWh with an efficiency of 40-45%$_{HHV}$.[52,53] If CCS is applied, the production costs are projected to increase to 64-132 €/MWh and efficiency could drop by as much as 11%$_{pt}$, depending on the technology used for gasification, CO_2 extraction and turbine; these results are similar for a conventional PC plant, both with and without CCS.[52,53] A recent study by the European Technology Platform for Zero Emission Fossil Fuel Power Plants (ZEP) reported the production costs of electricity of different types of Nth of a kind power plants, namely pulverised coal (PC), integrated gasification combined cycle (IGCC) and natural gas combined cycle (NGCC), both with and without CCS. Production costs of electricity production without applying CCS are 48 €/MWh for PC and 72 €/MWh for NGCC. If CCS is applied, the production costs of IGCC[2] and PC are similar (66-75 €/MWh vs. 67-73 €/MWh respectively), while NGCC-CCS results in significant higher electricity production costs (104 €/MWh). A study by the National Energy Technology Laboratory (NETL) found similar results for coal, but much lower production costs of electricity for NGCC mainly due to lower natural gas prices.[53,54]

Furthermore, model studies indicate that FT-liquids can be produced from coal in commercial scale plants for 6-17 €/GJ.[22,23,24,25,55,56] The lower ranges are mainly a result of low coal prices (in the range of 1 €/GJ). For comparison, oil-derived transportation fuels can be produced for 13 €/GJ if crude oil price are 72 €/bbl (which was the average crude oil price in 2011).[65] As mentioned before, CO_2 needs to be extracted from the syngas to enable high efficiencies in the conversion reactors. This results in CO_2 avoidance costs as low as 17 €/t CO_2.[57] IG-PG facilities can also use biomass. However, literature studies suggest that production costs of electricity and FT-liquids could double due to higher feedstock costs, lower efficiencies and smaller scale installations.[22,23,24] The combined use of coal and biomass in gasification facilities equipped with CCS has been investigated by the Princeton Environmental Institute, but only for dedicated gasifiers for each type of feedstock.[23,24,25,57] They conclude that CCS can be applied at integrated gasification facilities producing FT-liquids (IG-FT) at relative low costs. The co-feeding of biomass results in higher production costs compared to a coal-fired IG-FT, mainly due to higher capital and feedstock costs. Their studies also indicate that applying CCS at a coal-fired IG-FT facility has the same effect to mitigating CO_2 emissions as replacing about 40% of the coal by biomass at an IG-FT facility without CCS.[57]

In the mid to long term, significant costs reductions are expected for gasification facilities due to technological learning. Studies by the NETL estimate that novel technologies and improved operating experience could decrease production costs of electricity in an IGCC by 33% and reduce the CO_2 avoidance costs to 13 €/t CO_2.[58,59] Production costs of FT-liquids without CCS could potentially drop to 9 €/GJ, equivalent to an oil price of 50 €/bbl.

Flexibility

The mentioned studies assume no flexibility in feedstock or production. However, market conditions can change considerably in the operational lifetime of a large-scale IG-PG facility. Examples are the emerging market of biomass pellets for energy, the development of the CO_2 credits price and the market of non-oil based transportation fuels as well as the mature markets of coal and crude oil which has experienced strong price fluctuations in

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[2] In this study the reference plants is always a PC power plant, therefore data for IGCC without CCS are not reported.

the last years (see Figure 1.8).[60,61,62,63,64,65] Besides long-term uncertainties, there are also short-term variations, such as the daily variation in the electricity price,[66] which can impact the economics of IG-PG facilities, especially as large-scale storage of electricity is currently economically not feasible.

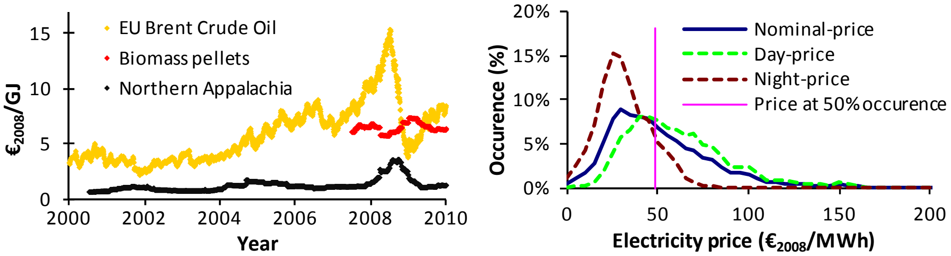

Figure 1.8 Historic prices of primary energy carriers (left) and distribution of historic electricity prices (right).[60,61,64,65,66]

As IG-PG facilities can handle a variety of feedstocks and can produce multiple products, these facilities allow reaction to and exploitation of fluctuations in market conditions. An example is exploiting the daily electricity variation existing in North West-Europe by producing electricity during peak hours and transportation fuels or chemicals during off-peak hours. Flexibility of the facility does, however, come with the disadvantage of higher capital investment and O&M costs and, depending on the type of flexibility, lower efficiencies compared to non-flexible IG-PG facilities. Currently, a detailed analysis of whether the advantages of flexibility offset the disadvantages has not been performed yet, despite the fact that flexibility is mentioned to be a strong argument in favour of IG-PG facilities.[44,67]

1.4 Summarising

There are significant potentials for both biomass and CCS to reduce GHG emissions. However, the current capacity of both CO_2 mitigation options is low. As stated previously, biomass for modern applications only comprises 2% of our global primary energy demand[7,27,28] and CCS is only being applied in eight commercial-scale projects.[41] Important reasons for the low penetration for biomass are the currently high feedstock prices, the uncertainty regarding availability due to the immaturity of the biomass for the energy market,[7] resulting in relatively small-scale biomass-fed conversion facilities. For CCS, important reasons are the high costs to avoid CO_2 emissions relative to the current CO_2 credit price or CO_2 tax and the uncertainty in future climate policy.

Gasification can play an important role to increase the deployment of both mitigation options. The installed gasification capacity is, however, only around 70 GW_{th}. An important cause is the high production costs. Although technological innovation resulting in large reductions in capital costs and improvement in overall efficiency are foreseen in the near future, achieving this improved performance requires that the installed gasification capacity increases significantly.

Flexibility might facilitate the deployment of gasification facilities for two main reasons. First, the ability of IG-PG facilities to be feedstock flexible allows the construction of biomass-fed facilities at coal-fed scale, and thereby takes advantage of economies of scale, but without the risk imposed by an inadequate supply of biomass. Second, flexibility lowers the investment risk as both feedstock and production can be altered during the operational lifetime of the facility. An additional advantage of gasification facilities is that CO_2 is often removed when chemicals are produced. This should allow the implementation of CCS at relatively low costs. Currently, there is a need for better understanding what the possibilities and limitations of flexibility at gasification facilities are and under which (economic) conditions flexible facilities are more advantageous than their static counterparts. These aspects are at the core of this thesis.

A promising (natural gas-fired) gasification application for the deployment of CCS in the short term is steam methane reforming (SMR) at industrial facilities. CO_2 avoidance costs at SMR facilities are projected to be lower compared to the avoidance costs at power plants due to the presence of a pressurised CO_2 rich gas stream and the possible availability of waste heat streams, as SMR facilities are generally located inside large industrial complexes, like refineries. Although literature studies on applying CCS at SMR facilities have been performed, the impact of applying CCS at existing facilities (retrofit) and the effect of using waste heat on the CO_2 avoidance costs at SMR facilities have not yet been studied.

1.5 Thesis objective and outline

The main objective of this thesis is to determine the techno-economic potential of commercial scale gasification systems producing energy carriers and other chemicals with low CO_2 emissions and to assess how and when flexibility improves the overall performance of these gasification systems.

The three main research questions are:

RQ. I What is the impact of flexible operation of state-of-the-art gasification facilities on the technical and economic performance of integrated gasification polygeneration facilities?

RQ. II What is the potential improvement in technical and economic performance of integrated gasification facilities in the coming decades?

RQ. III What are the CO_2 avoidance costs of gasification facilities, including SMR, in the short to long term?

Furthermore, literature indicates that performance of IG-PG facilities can improve in the future.[58,59] However, the rate of this improvement and the drivers behind it is still poorly understood. In this thesis the improvement potential is analysed using two approaches:

- A bottom-up analysis, giving insights into the extent of the potential cost reductions and how they can be realised. This approach involves identifying promising technologies, determining their current status and potential

development in the future and designing new process layouts of gasification facilities which take optimal advantage of new technologies;
- A top-down analysis making use of analogue learning curve studies, gives insights into how cost reductions could develop over time. This approach combines the required growth in installed capacity with scenario projections on the deployment of gasification facilities.

By comparing the two approaches a more robust prediction is obtained on how, when and to what extent production costs in gasification facilities can decrease in the future.

Table 1.1 gives an overview of the chapters of this thesis in which these research questions are addressed.

Table 1.1 Structure of the thesis.

	RQ. I	RQ. II	RQ. III
Chapter 2: *Performance of simulated flexible integrated gasification polygeneration facilities. Part A: A technical-energetic assessment*	●		
Chapter 3: *Performance of simulated flexible integrated gasification polygeneration facilities. Part B: Economic evaluation*	●		●
Chapter 4: *Technical and economic prospects of coal- and biomass-fired IG-PG facilities equipped with CCS over time*		●	●
Chapter 5: *Future technological and economic performance of IGCC and FT production facilities with and without CO_2 capture: combining component-based learning curve and bottom-up analysis*		●	●
Chapter 6: *Techno-economic assessment of CO_2 capture at steam methane reforming facilities using commercially available technology*			●

Chapter 2 aims to determine the technical impacts of flexibility in large-scale IG-PG facilities. The chapter provides an overview of the major components in an IG-PG facility. For each component the possibilities, limitations and effects of flexible operation is investigated. This information is used to develop a component-based computer simulation model. This model yields the mass and energy balances needed to determine the technical performance of IG-PG facilities and the impact flexibility has on overall IG-PG facility performance. The chapter also investigates the impact on the CO_2 fraction that can be captured in a flexible IG-PG facility. The flexibility comprises both feedstock (biomass pellets, torrefied biomass pellets and coal) and product (electricity, FT-liquids, methanol and urea).

Chapter 3 focuses on the impact of flexibility on the economics of IG-PG facilities. The results of the technical model developed in Chapter 2 are combined with an economic model to calculate production costs using the net present value (NPV) method. The resulting production costs are used to identify under which conditions flexible IG-PG facilities are economically more attractive than static IG-PG facilities. The chapter also investigates the effect a CO_2 credit price has on the profitability of the use of biomass versus coal as feedstock and on venting CO_2 versus applying CCS.

Chapter 4 assesses the potential improvement in overall performance of static IG-PG facilities producing electricity or transportation fuels, with or without CCS from either coal or biomass using a bottom-up analysis. For each component, possible improvements in existing technologies, as well as advanced alternative technologies are identified. The improvements and advanced technologies are categorized in three different time periods based on their expected commercialisation period. For each time period, a process layout of the IG-PG facility is made using the technologies available for that time period. These process layouts are used in the technical and economic models developed in Chapters 2 and 3. This results in insights into the potential reduction in production costs and into the extent the different improvements could potentially contribute to that reduction.

Chapter 5 aims to assess the potential improvement in overall performance of static IG-PG facilities producing electricity or transportation fuel, with or without CCS from either coal or biomass using a top-down approach. Learning curve analysis is applied in this chapter based on current and projected installed capacities of major IG-PG components. Based on these results, production costs for electricity and FT-liquids are calculated, as well as the impact on CO_2 emissions, up till 2050.The results are compared with the results of Chapter 4.

Chapter 6 aims to determine the optimal state-of-the-art CO_2 capture system for modern steam methane reforming (SMR) facilities using process simulations. A step-by-step methodology is developed to identify the optimal CO_2 capture system for an SMR facility, with the aim of reducing the energy penalty and minimising CO_2 avoidance costs, without compromising H_2 purity or SMR availability. The chapter assesses the location of the CO_2 capture unit within the SMR process, the CO_2 capture solvent and the dimensions of the absorber and regenerator.

Chapter 7 summarises the objectives, approach and key findings of this thesis. Besides the main conclusions, recommendations for further research are given.

2 Performance of simulated flexible integrated gasification polygeneration facilities. Part A: A technical-energetic assessment

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Abstract

This article investigates technical possibilities and performances of flexible integrated gasification polygeneration (IG-PG) facilities equipped with CO_2 capture for the near future. These facilities can produce electricity during peak hours, while switching to the production of chemicals during off-peak hours.

Several simulations were performed to investigate the influence of substituting feedstock and production on IG-PG facility output, load and efficiency. These simulations were done using a detailed AspenPlus simulation model of a Shell entrained flow gasifier combined with conversion facilities. In this model carbon-rich feedstocks (oil residues, coal and biomass) were converted to a variety of products (H_2, electricity, FT-liquids, methanol and urea) using state-of-the-art technology. The size of the gasifier was limited to the equivalent of 2000 MW_{th} Il #6 coal input.

Overall efficiency of the simulated non-flexible configurations to convert pure coal or pure wood pellets to electricity ($40\%_{HHV}$ vs. $38\%_{HHV}$), FT-liquids ($60\%_{HHV}$ vs. $55\%_{HHV}$), methanol ($53\%_{HHV}$ vs. $49\%_{HHV}$) or urea ($51\%_{HHV}$ vs. $47\%_{HHV}$) are in good agreement with the literature. Using torrefied wood pellets instead of pure wood pellets reduces the penalty drop in efficiency compared to coal. Moreover, torrefied wood pellets have superior energetic density, handling and feeding compared to wood pellets.

In this analysis, the H_2:CO ratio of the sweet syngas was fixed to match FT-liquids criterion. As a result, overall CO_2 capture rates are low, around 56-65%, depending on the feedstock used. Still, especially with FT-liquids and methanol production, CO_2 emissions at the facility are significantly reduced; less than 20% of the carbon feedstock entering the facility is emitted with the flue gas. Applying biomass and CO_2 capture shows great opportunities to produce CO_2-neutral electricity or chemicals. When the biomass fraction exceeds 40% on an energy basis, production is CO_2-neutral, independent of what is produced.

Biomass can be co-fed up till 50% on an energy basis. Higher fractions cause significant fouling on cooling equipment. A small part-load penalty is observed during the substitution of coal by biomass. When changing from pure coal to pure wood pellets, the power case suffers a 2.5% efficiency drop, while all three chemical cases have an efficiency drop of less than 1%. At the same time total output is reduced to 67-69%, mainly because of the lower energy density of biomass. By over-dimensioning the gasifier and gas cleanup and optimisation section this drop can be eliminated.

The syngas can be tailored to the desired composition regardless of the used feedstock. Therefore, the chemical conversion sections only have to cope with a reduction in syngas flow and not with a change in syngas composition. Altering production between chemicals and electricity is possible, although the load of the conversion sections should remain between 40% and 100% to prevent operational problems. This gives a high degree of flexibility.

Complete substitution between chemical and power production while using the same feedstock is possible for the methanol and urea cases. The FT-liquids case is restricted to 60-100% load of the chemical conversion section to prevent that the gas turbine load is reduced below 40%.

The economic aspects of flexible IG-PG facilities are addressed in part B.

2.1 Introduction

In 2008 global fossil fuel consumption was around 410 EJ. About 40% of this came from crude oil.[68,69] These fossil fuels power the economies of today. However, their use has several drawbacks. Fossil fuel resources are finite and their consumption results in large CO_2 emissions (unless carbon capture and storage is applied) as well as other environmental effects. Moreover, their consumption may result in high import dependence which may threaten supply security, for example in the European Union. Lastly, large-scale implementation of solar and wind power - so-called intermittent renewables - requires large-scale back-up electricity generation capacity. The output of back-up facilities must be easily adjustable in case solar and wind electricity production changes.[70] Also, the operating hours of these back-up facilities will be relatively low compared to base power plants.

A technology that can partly help to overcome these drawbacks is gasification. Through gasification (renewable) biomass, coal and oil residues can be converted into syngas, which can subsequently be converted into electricity, transportation fuels and other chemicals (Figure 2.1). Natural gas can be converted into syngas using reforming.

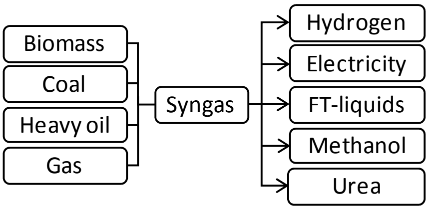

Figure 2.1 Conversion of carbon rich feedstocks to different products.

To reduce CO_2 emissions, CO_2 can be extracted from process and flue gases, compressed to 30-110 bar, transported and stored in underground fields (generally known as CCS). In this way, electricity and hydrogen can be produced with minimal CO_2 emissions. CCS at IG-PG facilities can be accomplished at reduced costs.[24,71] The reason for this is twofold. First, the chemical conversion processes already require removal of CO_2 from the syngas. Second, the higher partial pressure of the CO_2 makes pre-combustion capture using physical solvents possible. This lowers the energy requirements for the CO_2 capture and subsequent compression. By combining gasification and CCS it is possible to drastically reduce global CO_2 emissions without drastic changes to the current infrastructure. Furthermore, co-feeding biomass allows the production and use of liquid hydrocarbon transportation fuels with no net CO_2 emissions.[22,72] At large biomass fractions even net negative CO_2 emissions can be achieved.

An attractive feature of gasification is its potential flexibility. Especially the following two combinations of flexible integrated gasification polygeneration (IG-PG) facilities are interesting:

(1) Coal substitution by biomass can facilitate the use of biomass as feedstock without being dependent on a constant biomass supply.

(2) Chemical plants producing mid-load electricity can reduce the need of dedicated natural gas fired mid-load power plants. Mid-load plants are operated when electricity demand exceeds the base load. As intermittent renewables electricity generation capacity is ever-increasing, mid-load power plants will increasingly have to act as back-up power plants for these intermittent renewables. This may create opportunities for flexible large scale IG-PG facilities.

Multi-feedstock polygeneration (XtY)[3] system integration has seen a lot of attention in literature. A brief overview of studies highlighting results relevant for this study is given below.

2.1.1 General

Cummer and Brown investigated the major ancillary components, i.e. pre-treatment and gas cleaning, in integrated gasification combined cycle (IGCC) facilities.[73] They investigated whether components specifically designed for coal could also handle biomass. They also determined the stage of development of ancillary equipment for biomass fired IGCC. Their conclusion is that most of the ancillary equipment is not yet commercially available.

2.1.1.1 Electricity production

Electricity production using coal-fired IGCC combined with CCS has received quite some attention. A recent study by NETL indicates that, based on state-of-the-art (SOTA) Shell gasification technology, a 636 MW_e IGCC can achieve an overall efficiency of $41\%_{HHV}$, while emitting 750 kg CO_2/MWh. Equipped with CO_2 capture, plant efficiency would at present drop to $32\%_{HHV}$ and CO_2 emissions to 90 kg CO_2/MWh.[74] Similar values are presented by the IEA-GHG with $40\%_{HHV}$ efficiency for Shell IGCC without CO_2 capture and $32\%_{HHV}$ with CO_2 capture.[49,50]

2.1.1.2 Fischer-Tropsch fuels production

Although there is some variation in final efficiency among the literature, reported XtL conversion efficiencies lay between $46\%_{HHV}$ and $56\%_{HHV}$. Boerrigter calculated an overall conversion efficiency of $56\%_{HHV}$ for a non-capture CtL facility.[75] Larson et al., calculated an overall conversion efficiency for a non-capture CtL facility of $49\%_{HHV}$ (33% FT + 17% power). Introducing CO_2 capture and compression lowers the power output by $3\%_{pt}$. Converting a mixture of switch grass and coal to FT-liquids with CO_2 capture could be performed with an overall conversion efficiency of $47\%_{HHV}$.[76] This efficiency was confirmed by a study of Hamelinck et al. They calculated that biomass can be converted to FT-liquids with CO_2 capture at $46\%_{HHV}$ efficiency (33% FT + 13% power).[77]

2.1.1.3 Methanol production

A CtM study based on the liquid phase methanol (LPMeOH^{TM}) demonstration process at the Eastman Chemical Company in Kingsport was performed by Heydorn and Diamond

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[3] Facilities or systems where feedstocks are gasified and converted into products are referred to as XtY systems. The X is often substituted if a specific feedstock is used: biomass (BtY), torrefied biomass (TtY) or coal (CtY). The Y is often substituted if a specific output is produced: electricity (XtP), FT-liquids (XtL), methanol (XtM) or urea (XtU).

showing a CtM efficiency of 60%$_{HHV}$ - assuming a gasification efficiency of 80%$_{HHV}$ and a syngas energy to electricity efficiency of 40%$_{HHV}$.[78] Two different CtM configurations were analysed by Larson and Tingjin.[79] They calculated an efficiency of 47%$_{HHV}$ for once-through and 63%$_{HHV}$ for syngas recycle. A BtM efficiency was calculated by Williams et al., at 61%$_{HHV}$.[80] Hamelinck and Faaij calculated a BtM efficiency of 51%$_{HHV}$ using a fluidised bed system.[81]

2.1.1.4 Urea production

Urea synthesis via gasification has received little interest in open literature. Neelis et al., found an energy consumption of 21 GJ$_{LHV}$/t urea, resulting in a CtU energy conversion efficiency of 44%$_{LHV}$.[82]

2.1.2 Objectives

All these studies do not investigate variation in feedstock or production during the operation of the facility. A recent study from IEA GHG has made a first assessment of the flexibility between hydrogen and electricity production, but only at two different production ratios.[49] To our knowledge that is the only study which investigated the possibilities of flexible IG-PG facilities and the impact of flexibility on overall efficiencies. This study aims to fill this gap. The objectives of this study are therefore:

- First, to identify which bottlenecks occur when altering feedstock or production and how these bottlenecks can be resolved;
- Second, to analyse system behaviour and changes in the overall efficiency and in the energy, mass and carbon balances when altering feedstock or production;
- Third, to analyse the effect of flexibility on the CO_2 balance and emissions and compare the balance and emissions with those from dedicated XtY facilities.

In this study the focus is on the performance of SOTA commercial technologies relevant for XtY systems. To study these systems, an AspenPlus flowsheet process model of a flexible IG-PG facility was build to generate energy and mass balances for different IG-PG facilities. To allow accurate modelling, data on the technical bottlenecks as minimal load constraints and part-load behaviour of the various components were collected.

The structure of this article is as follows: Section 2 describes the commodities, including the different feedstock characteristics. In Section 3 the different components, including feedstock pre-treatment, syngas cleaning and syngas conversion, and their operating conditions are described. Section 4 gives the process model and plant configurations. In Section 5 the case studies are defined. Section 6 gives the simulation results, including overall performance and the effect of flexibility on overall IG-PG performance. In Section 7 a discussion of the results and used assumptions is given. Finally, Section 8 contains the conclusions.

In this study, units are in SI-units and heating values are in higher heating value (HHV), unless stated otherwise.

2.2 Commodities

2.2.1 Feedstocks

For this study three representative solid carbon-containing feedstocks were selected. Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal. Their properties are given in Table 2.1. Short rotation trees, of which Eucalyptus is one of them, is a potentially important biomass source.[83,84,85] The biomass is directly pelletised or torrefied and then pelletised as this drastically improves biomass properties, like heating value and moisture content (Table 2.2). This results, among others, to a more efficient transportation and handling of the biomass.[22,84,86,87] Illinois #6 coal is a Bituminous coal type often used in and commonly used as reference coal.[74,79,88,89]

Table 2.1 Feedstock parameters used in this study.

Composition	Unit (1)	EP[90]	TOPS[90]	Il #6 coal[74]
Heating value	MJ_{HHV}/kg a.r.	17.29	20.51	27.14
Moisture	wt% a.r.	10.00	5.00	11.12
Volatile Matter	wt% dry	86.60	75.90	44.50
Fixed Carbon	wt% dry	12.83	24.10	44.59
Ash	wt% dry	0.50	1.34	10.91
C	wt% d.a.f.	49.77	54.63	80.50
H	wt% d.a.f.	5.80	5.67	5.68
O	wt% d.a.f.	44.20	39.45	8.70
N	wt% d.a.f.	0.14	0.22	1.58
S	wt% d.a.f.	0.03	0.02	3.17
Cl	wt% d.a.f.	0.06	0.01	0.37

(1) The composition is on as received (a.r.), moisture free (dry) or moisture and ash free (d.a.f.) basis.

2.2.2 End-products

Syngas can be converted into a wide variety of end-products. The main end-products evaluated in this study are electricity, FT-liquids, methanol and urea. An important intermediate product is hydrogen.

2.2.3 By-products

Two important by-products of XtY facilities are slag and elemental sulphur. Slag is an inert glass-like material consisting mainly of minerals. It is formed as the ash in the feedstock melts and is subsequently quenched. Slag is, among others, used as concrete additive or road surface coating compound.[91] The amount and quality of the slag depends on the used feedstock. Coal generally produces more slag than woody biomass. Almost all sulphur in the feedstock is recovered as elemental sulphur. It is used in a wide variety of processes, including vulcanisation, bleaching and the production of fertilizers and viscose.[92] Coal generally contains more sulphur than biomass.

2.3 Process description

In this section a technical description and part-load performance of the different components used in a flexible IG-PG facility is presented. For all components only commercial available technologies were investigated. The key processes are displayed in Figure 2.2.

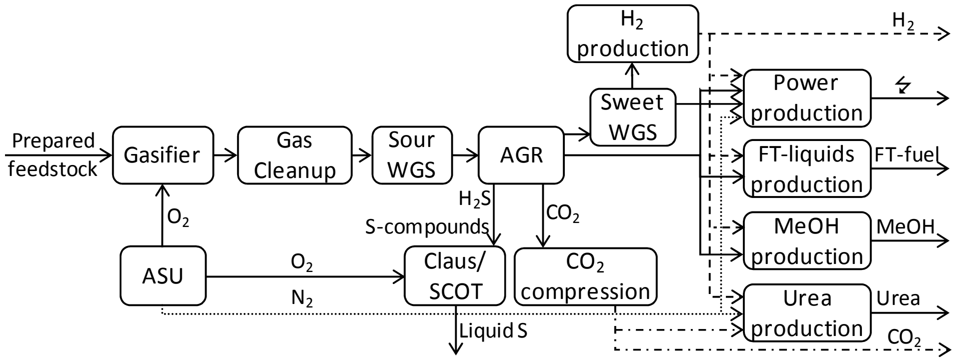


Figure 2.2 Simplified process layout of a flexible IG-PG facility using SOTA commercially available technology. Waste, heat and recycle streams are not displayed.

2.3.1 *Plant flexibility*

A flexible IG-PG facility must be designed to cope with alterations in feedstock, syngas flow and heat streams. Among others, it must be able to adjust the H_2:CO ratio. Physical limitations restrict the flexibility of the conversion processes. For example, separators are usually unable to operate properly below 40% volume load.[93,94] Also, gas turbines have difficulties maintaining the desired flue gas exit temperature below this load.[95] In the subsequent paragraphs the load restrictions and part-load behaviour of the various components are discussed.

2.3.2 *Pre-treatment*

The gasifier has several minimal requirements concerning the feedstock (see §2.3.5). To fulfil these requirements, the feedstock, especially biomass, needs pre-treatment, like pelletising and torrefaction combined with pelletising. The effects of these techniques are given in Table 2.2. For comparison the average coal characteristics are also given.

The pre-treatment can be performed at the harvesting, central gathering or factory location.[22,86] It can also be divided over these locations. In this study it was assumed that the biomass is already torrefied or pelletised when arriving at the IG-PG facility. The most important pre-treatment techniques are described below.

Table 2.2 Overview of chemical properties of different fuel types.

Property	Il. #6 coal[79] (Bituminous)	Fresh Wood[90]	Wood pellets[96]	TOPS[97]
Heating value (MJ_{HHV}/kg a.r.)	27	10-11	18-20	20-24
Mass density of bulk material (kg/m^3)	600	350	500-650	850
Energy density of bulk material (GJ/m^3)	16-17[97]	3-4	9-13	17-21
Moisture content (wt %)	11	50	6-10	1-5
C/H ratio (wt) (1)	13-16	~ 8	~ 8	9-10[87]
C/O ratio (wt) (2)	7-10	1	1	1-2[87]
Flammability (3)	o	++	+++	++[98]
Hydrophobicity	+	-	-	+

(1) Gasifying a feedstock with a higher C/H ratio results in more CO and less H_2.

(2) A lower C/O ratio results in a lower heating value (HHV) of the fuel. In a gasifier it also results in lower external O_2 consumption.

(3) The flammability of coal was set on average. Wood is easier to combust, but its high moisture content lowers flammability slightly. During pelletisation, this moisture is removed. TOPS has a low moisture content, but during the torrefaction process it lost most of its volatile compounds, thereby lowering its flammability.

2.3.2.1 *Torrefaction*

Torrefaction transforms biomass into a brittle, moisture-poor, hydrophobic solid with an increased energy density compared to the original biomass.[86,99] In short: the biomass becomes more coal-like. Another benefit is better storage properties as deterioration is much slower than raw biomass. The torrefaction process consists of four stages: heating, drying, torrefaction and cooling. During the torrefaction stage, the biomass is heated in a moving bed reactor to 200-300°C for several minutes. H_2O, CO_2 and other volatile compounds evaporate, resulting in an increased C/O ratio of the solid product. Besides this solid, a low caloric gas is produced. On average, the biomass will lose 30% of its mass and 5-10% of its energy, depending on the biomass used and the original moisture content. The low caloric gas is combusted and used as heating source. Additional biomass is combusted to supply the remaining heat demand, resulting in an almost completely self-sufficient system.[86,99] Torrefaction is a batch process requiring storage when applied in a continuous process, such as an IG-PG facility. Storage also fits to part-load operation of the facility. The torrefaction process is operated at full load, regardless of the IG-PG facility load. When the storage bins are filled, the torrefaction process is shutdown. It is restarted when the feedstock in the storage bins comes below a certain threshold value.

2.3.2.2 *Pelletising*

Pelletising increases the energy density of biomass feedstocks. This, among others, reduces transportation costs. During pelletising the biomass is highly compressed. If the biomass is torrefied, pelletising is mandatory except if the torrefied product is used immediately. When using raw biomass, the pelletising process consists of four stages: drying and grinding, steam conditioning (to soften the fibres), pressurisation and cooling. To prevent pellet disintegration, glue is needed. When pelletising at 150°C the lignin in the biomass - which softens above 100°C - is used as glue.[86] The energy density of the biomass increases by about a factor of 2 (Table 2.2). Electricity consumption of pelletising is 90-160 MJ_e/t input.[86]

Combining torrefaction with pelletising results in a factor 3 increase in bulk energy density compared to raw biomass (Table 2.2) and a halving of the electricity consumption of the pelletising process. It is best to pelletise the biomass directly after the torrefaction step. This reduces the pelletising process to pressurisation and cooling only. Like torrefaction, pelletising is a batch process. Therefore, the same part-load behaviour applies.

2.3.2.3 Sizing

Sizing reduces the particle size of solid feedstocks, making transporting and feeding the feedstock more efficient and easier. It also improves drying, torrefaction and pelletising. Shell EF gasifiers demand particles smaller than 0.1 mm for coal and smaller than 1 mm for biomass. Biomass particles can be larger as they are more reactive than coal.[100] Hard brittle feedstocks (TOPS, coal) are ground using crushers. More fibrous feedstocks (raw or pelletised biomass) are ground using a hammermill. Grinding biomass to 1mm and coal to 0.1 mm requires an energy consumption of 0.01-0.02 kW_e/kW_{th} - grinding biomass to 0.1 mm would require 0.08 kW_e/kW_{th}.[100] Sizing has good part-load behaviour. By reducing the speed of the conveyer belt and the sizing equipment, part-load operation can be achieved without lower efficiencies.

2.3.2.4 Drying

The high moisture content in raw biomass makes it unsuitable for gasification - fresh wood has a typical moisture content of around 50 wt%.[90] Drying is preferred before pelletising or long distance transport. Active drying of biomass below 10 wt% moisture is not practical as drying efficiency drops dramatically below that value.[101] Heat demand depends among others on initial and final moisture content, particle size and hydrophobicity of the feedstock. This heat can be supplied by waste heat of other processes. Drying has good part-load behaviour. Just like sizing the speed of the conveyer belt and the amount of heat added to the feedstock can be reduced. In this study the coal was dried to 2.5 wt% using waste heat of the IG-PG facility. Biomass did not receive further drying as its moisture content was already below 10 wt%.

2.3.3 Pressurising and feeding

Previous research indicated that pressurised gasifiers have significant advantages over atmospheric gasifiers. The most important advantages are smaller downstream equipment and having a gas stream at elevated pressure.[77] As most chemical conversion processes and the gas turbine operate at higher pressure, the latter advantage significantly lowers compression requirements. However, using pressurised gasifiers also requires pressurised feeding. For solid feedstocks, the pressurising and feeding systems depend on the particle density and size. Coal is usually fed by a lock-hopper system with pneumatic feed. Fresh (or pelletised) biomass needs a different feeding system as it is too fibrous. This can lead to blockage of the lock-hopper feeding system. Due to the larger particle size of biomass, a hydraulic piston system with screw feeding can be used. The torrefaction process destroys the fibres, allowing TOPS to use both feeding systems. However, screw feeding is more efficient than pneumatic feeding.[100] As both systems are batch systems, they are equipped with pressurised storage bins, allowing homogeneous and continuous feeding. As pressurising agent, CO_2 was used. Energy consumption depends on the used process and is needed for CO_2 compression and the hydraulic piston feeder system. In this study separate feeding system for the biomass and coal were used. Part-load operation only affects the frequency at which the storage bins are refilled. The effect of feeding the different feedstocks is given in Table 2.3.[100]

Table 2.3 Feedstock pressurisation characteristics.[100]

		Il. #6 coal (1)	**TOPS (1)**	**EP (1)**
CO_2 consumption	Nm^3 CO_2/t feedstock	3.7	0.2	0.2
of which:	to gasifier	2.1	0.1	0.1
	to atmosphere	1.6	0.1	0.1
Electrical consumption	kWh/t feedstock	46.7	42.5	42.5

(1) Based on a lock-hopper system with pneumatic feeding for coal and a hydraulic piston system with screw feeding for biomass.

2.3.4 Air separation unit (ASU)

The use of pure oxygen (95%) instead of air leads to a lower syngas volume and higher syngas energy density. There are three different air separation techniques available to extract oxygen from air: (vacuum) pressure swing adsorption, membrane separation, and cryogenic separation.[102] Large scale membrane separation shows high potential but is still in the demonstration phase. Of the other options, cryogenic separation is currently the most economical option when high purity (>95%) and throughput (>2.5 t/h) are required.[103,104] Energy savings can be accomplished by integrating the ASU with the gas turbine. An integrated ASU receives (part of) its air already partly compressed from the gas turbine. Although this increases efficiency, plant flexibility and reliability is reduced due to the dependence of the ASU on the gas turbine.[93,105] Also, other modern techniques, like adjustable rotor blades, decrease these integration gains significantly. Therefore, in this study a non-integrated ASU is used. Power consumption of the ASU is 0.20 kWh/kg O_2 produced.[106] The pure oxygen is compressed in several steps to 120% of gasifier pressure at an isentropic efficiency of 75%.[107] Between the compression steps the oxygen is cooled to 40°C. Large scale cryogenic ASU have good part-load behaviour till 50% load. Specific energy consumption is almost constant between 85% and 100% load.[93]

2.3.5 Gasifier

The gasifier converts the feedstocks into a gaseous mixture of mainly CO and H_2. There are several types of gasifiers available. They differ in operating temperature, feedstock state (slurry or dry), type of oxidant (air or oxygen) and gasifier configuration. As a high quality syngas is required, an oxygen-blown EF gasifier is used as they typically operate at high temperature (>1500°C) and pressure (20-70 bar).[91] The most used EF gasifiers are the Shell and General Electric (GE) (formerly Texaco) gasifiers.[46] Although both gasifiers are commercially available, only the Shell gasifier has had extensive biomass co-feeding runs.[108] Also, the Shell gasifier is the most fuel flexible of the two. This is mainly caused by the difference in dry and slurry feeding. Slurry for the slurry-fed GE gasifier contains roughly 40% water (excluding the moisture in the feedstock), which must be evaporated. Therefore the feedstock needs a high heating value. The feedstock of a dry-fed Shell gasifier contains much less water. This results in less water evaporation, making the Shell gasifier suitable for feedstock with a low heating value, like biomass.[109] Therefore in this study a Shell EF gasifier is used. Typical operating conditions are given in Table 2.4.

Currently, gasification facilities using a dry-fed Shell EF gasifier have been designed to co-feed biomass up to 30% on an energy basis.[108] Up to 50% might be possible for these gasifiers, but this requires considerable adjustment of various processes, resulting in a reduction of overall efficiency, as well as very clean woody biomass and high energy coal, like anthracite. The different make-up of the ash in biomass - especially its chloride content - and its lower energy value are the main reasons for this limitation.

Table 2.4 Feedstock boundaries.

Characteristic	Unit	Shell[110]
Particle Size (Coal-Biomass)	mm	< 0.1-1
Typical water content[111]	wt%	~ 2
Fuel state	-	Solid
Ash fusion temperature	°C	< 1500
Ash content	wt% dry	8-40
Sulphur content	wt% dry	< 7
Heating value	MJ_{HHV}/kg dry	>17
Maximum pressure[112] (1)	bar	40

(1) Shell is developing gasifiers that operate at pressures above 40 bar. These gasifiers have not yet used biomass.

The chosen gasifier is a 40 bar, oxygen-blown, slagging EF Shell gasifier with a syngas exit temperature of 1500°C. To reach the required high temperature a sub-stoichiometric amount of oxygen is added, which reacts with the fuel. In the gasifier all oxygen and almost all carbon (>99.5%) reacts. Therefore, the syngas composition can be described by the following equations:

$$CO + H_2O \leftrightarrows CO_2 + H_2 \qquad \Delta H^\circ_{298K} = -41.2 \text{ kJ/mol}^{[91]} \qquad (2\text{-}1)$$

$$CH_4 + H_2O \leftrightarrows CO + 3\,H_2 \qquad \Delta H^\circ_{298K} = +206 \text{ kJ/mol}^{[91]} \qquad (2\text{-}2)$$

Besides these main components, the syngas contains impurities like methane, N- and S-compounds, unconverted carbon particles, and slag are present. The outer wall of the gasifier is cooled with water to preserve the gasifier and the slag layer. This cooling extracts 4% of the feedstocks HHV.[113] This energy is used to generate steam. When syngas is cooled, contaminants (e.g., halides, carbon particles) in the syngas become sticky between 900-1100°C. When using pure biomass as feedstock this stickiness can occur at temperatures as low as 500°C.[113] This prevents the use of high fractions of biomass. To prevent fouling of cooling equipment, the syngas is quenched to 850°C.[114] As the syngas is shifted downstream a water quench is used. This lowers the steam consumption of the downstream water-gas shift (WGS) reactor.

Two important parameters to evaluate gasifier performance are carbon conversion and cold gas efficiency (CGE).

$$\text{Carbon conversion efficiency (\%)} = \left(1 - \frac{\text{Unconverted carbon (kmol/h)}}{\text{Carbon in feedstock (kmol/h)}}\right) * 100 \qquad (2\text{-}3)$$

$$\text{Cold gas efficiency (\%)} = \frac{\text{HHV}_{\text{syngas}}\ (\text{MW}_{\text{th}})}{\text{HHV}_{\text{feedstock}}\ (\text{MW}_{\text{th}})}*100 \tag{2-4}$$

The carbon conversion efficiency is a measure for feedstock conversion. It is the fraction of carbon converted to gaseous compounds like CO, CO_2, CH_4 and HCN. Shell gasifiers can reach carbon conversion efficiency > 99.7%. The remaining carbon ends up in the slag or fly ash. The CGE is a measure of gasifier energetic efficiency. The CGE depends on the used feedstock and its moisture content. Modern dry-fed EF gasifiers have a CGE ranging between 80% and 83% for coal.[79,115,116] The gasifier is run at constant space-time volume. This means that a constant volume of syngas is leaving the gasifier each second, regardless of the used feedstock. By over-dimensioning the gasifier, the chemical energy of the syngas exiting the gasifier can be kept constant, even when using large fractions of biomass. When in part-load, the high temperatures must still be reached. But the relative heat loss will be higher than the previously mentioned 4%. Therefore, a larger fraction of the feedstock must be combusted, resulting in a lower CGE. Also, a sufficient amount of slag must be produced to maintain the slag layer.[94]

2.3.6 Gas cleanup and syngas composition optimising

In the gas cleanup and composition optimising section contaminants are removed and the H_2:CO ratio of the syngas is adjusted to the desired value. The extend of purification and the desired H_2:CO ratio depend on the application. Table 2.5 gives the contaminant restrictions for the most demanding application of the syngas conversion processes - FT-liquids production - used in this study.

Table 2.5 Maximum allowable concentrations of main syngas impurities.

Impurity	FT-reactor[117]
Total S	< 1 ppmV
Total halides	< 10 ppbV
Total N	< 1 ppmV
Dust (d_{max}=10µm)	Almost completely removed
Tar	T_{dew} < 200°C

A gas cleaning and composition optimising train is described below in the order in which the syngas passes through. A downside of this train is the considerable variation in syngas temperature. The required heat exchangers are not described.

2.3.6.1 Cyclone

A cyclone or rotating particle separator (RPS) uses centrifugal and Coriolis forces to remove solid and condensed particles.[118] An average cyclone removes 90% of particles greater than 5 µm, while experiencing a pressure drop of 0.01 bar. With identical gas residence times a RPS removes 90% of particles greater than 0.5 µm.[119] The separation efficiency can be increased by increasing the residence time in the cyclone.[113,120] Operational data revealed that a combination of cyclone, filter and wet scrubber removes almost all particles from the syngas.[108,114] In this study it is assumed that the cyclone removes all solid particles. The cyclone has no energy consumption. Part-load behaviour affects cyclones in two ways. A reduction in inlet particle concentration lowers cyclone removal efficiency. However, this is offset by the lower initial particle concentration. The result is a more or less constant outlet particle concentration. A reduction in gas velocity lowers cyclone removal efficiency by about 2.5%point per 10% load reduction.[121]

2.3.6.2 Candle filter

A filter also removes particles from the syngas. By cooling the syngas to 300°C before the filter, alkali particles condense and are also removed from the syngas. A commonly used filter is the candle filter due to its fail-safe in the event of a filter breach. Regeneration is performed by back-flashing with an inert gas. This is commonly N_2 but in a CCS facility CO_2 can be used. Part-load behaviour has no negative effect on filter behaviour. Electricity is needed for the regular back-flashing. This consumption is negligible and not taken into account in this study.

2.3.6.3 Wet scrubber

A wet scrubber removes the remaining alkali and (small) particles by spraying the syngas with water. The water is collected at the bottom of the scrubber and is purified and recycled. Due to the high pressure, the temperature of the water can be around 175°C. Syngas will leave the scrubber saturated. During part-load operation, the water flow is reduced. This keeps the cleaning effectiveness constant. The reduced water flow lowers energy consumption in the subsequent purification and recycling steps. Electricity is needed for the water pumps. This consumption is negligible and not taken into account. Minimal part load operation is 30%.[122]

2.3.6.4 Water-gas shift (WGS) reactor

Each syngas conversion process has an optimal syngas composition (Table 2.6). The WGS reactor adjusts the H_2:CO ratio of the syngas using reaction 2-1. There are three types of WGS reactors: sour, high temperature sweet and low temperature sweet. The sour WGS reactor needs S-compounds in the syngas as the catalyst needs to be sulphided, while the sweet WGS reactors can only operate in absence of S-compounds. Therefore sour WGS reactors are located upstream the acid gas removal (AGR), while sweet WGS reactors are located downstream the AGR. Sour WGS reactors use Co/Mo catalysts at temperatures between 300-500°C.[91] Besides the WGS reaction, S-compounds are converted to H_2S and HCN to NH_3.[123] High temperature sweet WGS reactors (300-500°C) use Cu-promoted, Fe/Cr or Co/Mo catalysts.[91,113] At these temperatures, around 97% of the remaining CO is converted. Low temperature sweet WGS reactors (180-270°C) use Cu/Zn-Al oxide based catalysts. These catalysts are very sensitive to sulphur poisoning and demand very low S-concentrations (<1 ppm(v)). To maximise hydrogen production additional steam is added. In this study a sour WGS located upstream the AGR and a sweet WGS located downstream the AGR were used. In sour WGS reactors part-load operation or very clean biomass can lower the sulphur flow resulting in reduced performance of the catalyst. Besides this, part-load operation of WGS reactors has no negative effect on performance.

Table 2.6 Optimal syngas compositions.

Product	Variables	Opt. value
H_2	H_2/CO	∞
Electricity	H_2/CO	N.A.
FT-liquids	H_2/CO	0.6-2
Methanol	$H_2/(2*CO+3*CO_2)$	1.3-1.4
NH_3	H_2/N_2	3
Urea	NH_3/CO_2	2.95

2.3.6.5 AGR

During the acid gas removal (AGR) S-compounds and CO_2 are separately removed from the syngas. When operated at low temperatures (<100°C), the AGR also removes almost all H_2O from the syngas. Currently several commercial techniques can be used to extract the acid gases.[124] In this study the sour syngas contains a high CO_2 partial pressure. Also, deep sulphur removal is needed as downstream catalysts are very sensitive to sulphur poisoning. This makes the Rectisol process the preferred AGR technique. Its characteristics are displayed in Table 2.7. The Rectisol process consists of 2 absorbers, two flash drums, heat exchangers, coolers and compressors. Besides the sweet syngas, a pure CO_2 stream and a sulphur rich stream is produced. During operation the solvent slowly degenerates. Therefore, methanol is constantly added. The AGR uses 137 MJ_e/t CO_2 and 90 MJ_{th}/t CO_2 of low pressure steam.[75,125] When operating at part-load, two options are possible. Either the AGR continues to run at full capacity. This will improve capture efficiency, but will also increase energy consumption per kg CO_2 captured. Or, the recirculation flow of the solvent is reduced. This way capture efficiency and energy consumption per CO_2 captured remain almost constant.

Table 2.7 Overview of process conditions and performance of Rectisol.[126]

Technology	Final conc. (ppm S)	Operating temp. (°C)	Product
Rectisol	0.1-1	-30 to -60	H_2S

2.3.6.6 Claus plant

S-compounds are converted to elemental sulphur in the Claus process. Here, 1/3 of the H_2S is combusted to SO_2 (reaction 2-5) and any remaining hydrocarbons and ammonia are also combusted. Next are three catalytic beds where the remaining H_2S reacts with SO_2 to form elemental sulphur (reaction 2-6).

$$2\ H_2S + 3\ O_2 \leftrightarrows 2\ SO_2 + 2\ H_2O \qquad \Delta H^\circ_{298K} = -519\ \text{kJ/mol}\ H_2S^{[127]} \qquad (2\text{-}5)$$
$$2\ H_2S + SO_2 \leftrightarrows 3\ S + 2\ H_2O \qquad \Delta H^\circ_{298K} = -74\ \text{kJ/mol}\ H_2S^{[127]} \qquad (2\text{-}6)$$

After each bed the elemental sulphur is removed from the gas stream. Each catalytic bed operates at a lower temperature then the previous bed to maximise elemental sulphur formation. However, thermodynamic equilibrium prevents complete conversion to elemental sulphur at acceptable reaction rates. With three well tuned beds, recovery rates of 96-97% can be achieved.

The modelled Claus is oxygen-fired. This reduces the minimal sulphur concentration in the gas stream from 25 mol% for air-fired to 10 mol%.[91] The oxygen consumption of the Claus plant is around 2% of the oxygen demand of a gasification plant using Il #6 coal. The Claus process uses 190 MJ_e/t sulphur.[128] Part-load operation or substituting coal by biomass can result in a too low sulphur concentration in the gas stream. If total flow drops too low, the burner flame can become unstable or the temperature in the burner too low. To prevent this it might be necessary to recycle sulphur or add natural gas.[91]

2.3.6.7 SCOT plant

The Claus off gas contains too much sulphur to comply with Western air pollution regulations. Therefore, a Shell Claus Offgas Treatment (SCOT) plant is used to recycle the sulphur from the off gas back to the Claus burner. The Claus off gas is fed into a catalytic bed, where all sulphur is converted to H_2S using hydrogen at elevated temperatures. The hot gas stream is cooled and H_2S is removed from the gas using an amine adsorption process. The H_2S is recompressed and recycled to the Claus unit. In this way >99.8% sulphur recovery rates can be achieved. The SCOT uses 100 MJ_e/t sulphur.[128] The SCOT plant is less sensitive to part-load operation than the Claus process. Therefore, a SCOT plant will have no problem with part-load operation.

2.3.7 Syngas conversion

After the syngas is cleaned, it can be distributed to the H_2-PSA and the syngas conversion sections. By altering the syngas flows to the chemical conversion sections and the H_2-PSA, production can be altered.

2.3.7.1 H_2 extraction

To produce pure hydrogen (a part of) the sweet syngas is shifted in a low temperature WGS reactor to increase the hydrogen concentration. Next, the hydrogen is extracted from the gas stream. The extraction process can use temperature or pressure swing adsorption (TSA or PSA) or membrane separation. The current standard is PSA.[92] The syngas passes through sorbents at which only hydrogen hardly interacts. All other components in the syngas bind to the sorbents. To regenerate the sorbents, the pressure is reduced, releasing the bonded contaminants.[92] As this is a batch process, several PSA columns are operated in parallel. The pure hydrogen can be used for ammonia production, H_2-cracking, SCOT reducing agent, adjusting the syngas H_2/CO ratio or as a valuable end-product. Part-load operation will reduce the frequency of regenerating the PSA columns.[129]

2.3.7.2 Electricity

Electricity is generated by combusting syngas and combustible waste streams in a gas turbine. The syngas is diluted first with N_2 and, if necessary, with steam to lower the energy density of the syngas to the desired value. Excess air is injected to provide the necessary oxygen. The addition of N_2 and air also lowers the flame temperature, reducing NO_x formation. After combustion, the hot flue gas is used for steam generation. In this study an Alstom GT26 turbine was used. This model turbine is from 1993 and has more than 2 million hours of experience, making it suitable for IGCC application. The GT26 was selected as it has superior part-load behaviour because it can reduce the air flow down to 40% (Figure 2.3).[130]

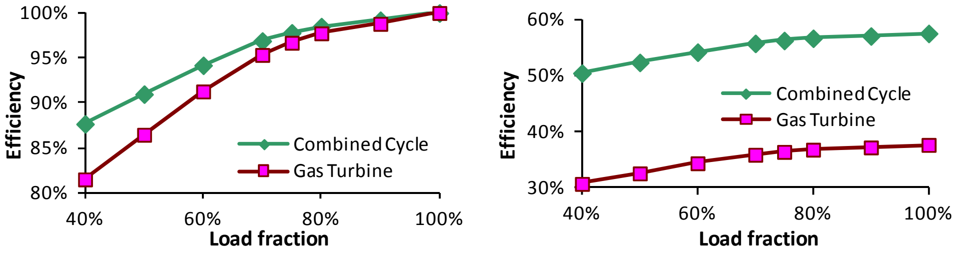


Figure 2.3 Part-load efficiency gas turbine and combined cycle.[95,130]

The air compressor efficiency was kept constant. To compensate for part-load behaviour the isentropic efficiency of the expander was adjusted using the following formula:

$$\eta_{Isentropic} = 0.65 + 0.3577*x - 0.1757*x^2 \qquad (2\text{-}7)$$

where x = ratio of the mass flow of the flue gas compared to the nominal load.

Calibration was done using data from Alstom.[130,131] Part-load behaviour was based on part-load behaviour of the combined cycle (Figure 2.3), assuming that the steam cycle compensates a third of the efficiency loss of the gas turbine.[95]

2.3.7.2.1 Heat recovery steam generation (HRSG)

In an IG-PG facility there are several processes - including coolers, syngas conversion reactors and compressors - that give off excess heat. Similarly, there are several processes that demand heat. All these heat streams are combined in the heat recovery steam generation (HRSG) system. Any remaining excess heat is used for electricity generation by steam turbines.[132] The HRSG in this model uses three steam levels: low pressure (LP: 12.5 bar and maximal 368°C), intermediate pressure (IP: 40 bar and maximal 550°C) and high pressure (HP: 125 bar and maximal 550°C). To improve the efficiency of the HRSG, the higher pressure steam is pre-heated with low temperature excess heat. Excess heat up to 200°C is used to pre-heat internal used steam and LP steam. Excess heat between 200°C and 450°C is used to pre-heat IP and HP steam and to heat LP steam and internal used steam to the desired temperatures. Excess heat above 450°C is used to generate IP steam. Only the heat from the first flue gas heat exchanger is used for HP steam generation. In part-load operation the water flows are reduced to maintain the desired temperatures of the steam.

2.3.7.3 FT-liquids synthesis

The FT-section converts syngas to linear alkanes and 1-alkenes. The process consists of an FT-reactor, heat exchangers, flash drums, distillation columns and H_2 cracker. The H_2 cracker is needed to upgrade the FT-liquids to transportation fuels that can be used in current car engines. During the FT-reaction about 20% of the chemical energy is converted into heat.[77] This heat is used for IP steam generation. The main FT-reaction (reaction 2.8) is catalysed by Fe, Ru or Co.[133] The result is a hydrocarbon mixture with a composition that can roughly be described using a modified Anderson-Schulz-Flory distribution (see equations 2-9, 2-10 and 2-11).[77,134]

$$2n\ H_2 + n\ CO \rightarrow -(-CH_2)_n- + n\ H_2O \qquad \Delta H = -165\ kJ/mol^{[133]} \tag{2-8}$$

$$S_{C5+} = 1.7 - 0.0024 * T - 0.088 \frac{[H_2]}{[CO]} + 0.18 * \left([H_2] + [CO]\right) + 0.0078 * p \tag{2-9}$$

$$\alpha \approx 0.75 - 0.373 * \sqrt{-\log(S_{C5+})} + 0.25 * S_{C5+} \tag{2-10}$$

$$\alpha_{Cn} = \alpha^{n-1} * (1-\alpha) \tag{2-11}$$

where S_{C5+} = fraction hydrocarbons containing at least five carbons
T = temperature (K)
$[H_2]$ and [CO] = concentration of hydrogen and CO in the gas mixture
p = pressure of the gas mixture (bar)
α = chain growth probability
α_{Cn} = fraction of hydrocarbons with n carbon atoms in FT-liquids mixture

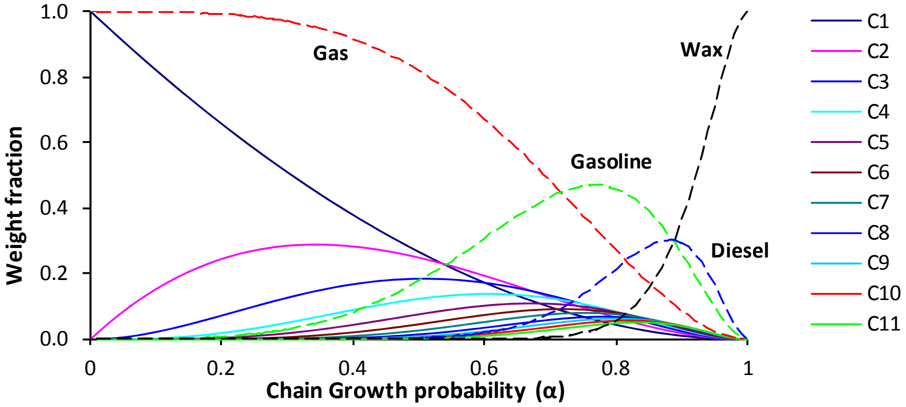


Figure 2.4 Effect of α on FT-product distribution.

α increases with higher temperature, lower pressure, lower H_2/CO ratio, lower inert partial pressure or using Ru- or Co- instead of Fe-based catalyst. Maximum gasoline production is achieved at α= 0.76 and maximum diesel production at α= 0.88, but at these α large amounts of (too) light alkenes are produced (Figure 2.4). Therefore, an α> 0.95 is commonly chosen, followed by H_2-cracking of the wax to yield gasoline and diesel.

2.3.7.3.1 FT-liquids upgrading

The FT-synthesis yields the following products: gases (C_1-C_4), naphtha or gasoline (C_5-C_{11}), diesel (C_{12}-C_{20}), light wax (C_{21}-C_{30}) and heavy wax (C_{31+}).[135] The gases are recycled to the gas turbine. To maximise economics - the desired products are gasoline and diesel - product upgrading is needed. This upgrading is similar to crude oil upgrading and FT-oil can be mixed with crude oil and used in existing refineries. The wax is led into a hydrocracker. Reaction rates of the hydrocracking depend on the size of the hydrocarbon as bigger hydrocarbons react faster than smaller ones. By using this principle, the alkane distribution can be tuned. The last step is distillation, resulting in naphtha (gasoline) and diesel.[136] The FT-liquids synthesis and upgrading has good part-load behaviour till 40% load, although small adjustments to temperature and pressure might be necessary.[137]

2.3.7.4 Methanol synthesis

Current methanol production uses the Cu/Zn/Al catalyst promoted reaction of H_2 with CO and small amounts of CO_2 (reactions 2.12 and 2.13) in a fixed bed or slurry reactor.[127] During the methanol reaction 15% of the chemical energy is converted into heat.

$2\ H_2 + CO \leftrightarrows CH_3OH\ (L)$ $\Delta H°_{298K} = -128$ kJ/mol[65] (2.12)
$3\ H_2 + CO_2 \leftrightarrows CH_3OH\ (L) + H_2O\ (L)$ $\Delta H°_{298K} = -131$ kJ/mol[65] (2.13)

A slurry reactor is used in this study as it has higher efficiencies - due to improved temperature control.[78] In the slurry reactor, the syngas diffuses from the gas phase through the liquid phase to the solid catalyst, where the reaction occurs. The primary function of the liquid is to regulate the temperature. Purification of the methanol is done using a flash drum and three distillation columns. This leads to >99.5% pure methanol.[136] Unreacted reactants can be recycled back to the methanol reactor to obtain high overall conversion rates.[123] In this study all unreacted reactants were fed into the gas turbine. The methanol synthesis and upgrading has good part-load behaviour till 40% load.[78]

2.3.7.5 NH_3 synthesis

Ammonia is synthesised by reacting H_2 and N_2 (reaction 2.14) at high pressure and temperature (100-250 bar, 350-550°C) using Fe-oxide catalysts. Due to thermodynamic limitations, high conversion rates can only be obtained by recycling the unused gas.[123,127]

$N_2 + 3H_2 \leftrightarrows 2\ NH_3$ $\Delta H°_{298K} = -45.9$ kJ/mol NH_3[127] (2.14)

2.3.7.6 Urea synthesis

Ammonia is converted into urea by reacting it with CO_2 (reaction 2.15).[138] The CO_2 is supplied by the AGR and needs an additional compression step to reach the desired pressure.

$2\ NH_3 + CO_2 \leftrightarrows (H_2N)_2CO\ (s) + H_2O\ (l)$ $\Delta H°_{298K} = -138$ kJ/mol[77] (2.15)

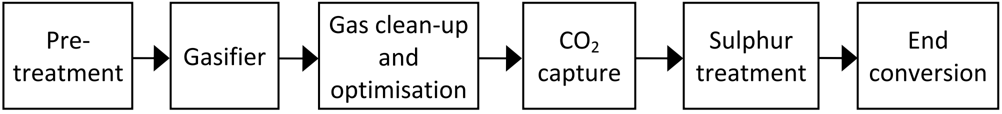
During the urea synthesis 26% of the chemical energy of the H_2 is converted into heat. The two most important urea synthesis processes are the Stamicarbon and the Snamprogetti process. Both processes use a two-step reaction. In this study the Stamicarbon process was used. It operates at 140 bar and 185°C with a NH_3/CO_2 ratio of 2.95. This results in 60% CO_2 conversion (or 41% NH_3).[139] High conversion rates are obtained by recycling the reactants. After decompressing the urea, its temperature must be kept below 130°C to prevent decomposition of the urea. The ammonia and urea synthesis and purification has good part-load behaviour till 40% load. A potential problem is biuret formation. At part-load the biuret concentration increases. At high concentration - typical >1 wt% - the biuret needs to be removed as it is toxic to plants. The specific energy consumption will slightly increase when operating at part-load conditions.[140]

2.4 Methodology: AspenPlus flowsheet model

2.4.1 Modelled components

The IG-PG facilities were modelled in the flowsheeting program AspenPlus. For each process there are, as indicated above, several alternatives. A short overview is given in Figure 2.5. The processes used in the modelled flexible IG-PG facility are depicted in bold.

Figure 2.5 Types of processes used in gasification.



Pre-treatment	Gasifier	Gas clean-up and optimisation	CO_2 capture	Sulphur treatment	End conversion
Drying	Moving bed	**Cyclone**	**Physical**	**Claus**	**MeOH/**DME
Grinding	Fluidised bed	Tar Cracker	Chemical	SuperClaus	**NH_3/Urea**
Pelletising	**Entrained flow**	**Filter**	Membrane	**SCOT**	**FT-liquids**
Torrefaction	Supercritical H_2O	**WGS**	Sorbents	LTGT	**Electricity**
Pyrolysis		Membrane			**Hydrogen**
					Polymers

The selected processes were modelled using the parameters described in Table 2.9. An overview of the model lay-out is given in Figure 2.6. The feedstock was gasified and converted to syngas. The syngas was cleaned and optimised (H_2:CO ratio). The sweet syngas was distributed over the different production sections. The offgas of the sweet WGS and FT-liquids, methanol and urea conversion sections were combusted in the gas turbine. Both FT-liquids and methanol sections were operated in a once-through mode. The urea section was operated in a recycle mode - both for the ammonia and urea synthesis. The pressure of the CO_2 for CCS is 110 bar, but the urea reactor operates at 140 bar. Therefore, the CO_2 for urea production underwent an additional compression step. To allow production flexibility, two distribution nodes distributed the gas flows. One node contained balanced syngas (H_2:CO ratio = 2.3) and was located between the guard bed and the conversion sections, bypassing the sweet WGS and H_2-PSA. The other node contained pure H_2 and was located between the H_2-PSA and the conversion sections. By altering the flow through these nodes the H_2:CO ratio was fine-tuned for the conversion processes. Especially the gas turbine behaved different at part-load. Therefore the isentropic efficiency of its expander was adjusted according to the load factor (Table 2.9). To maintain the desired pressure and temperature of the steam, the flow rates through the HRSG and steam turbines were adjusted. Regarding feedstock flexibility, the gasifier was the key determinant. Its gas generating capacity was kept constant for all cases.

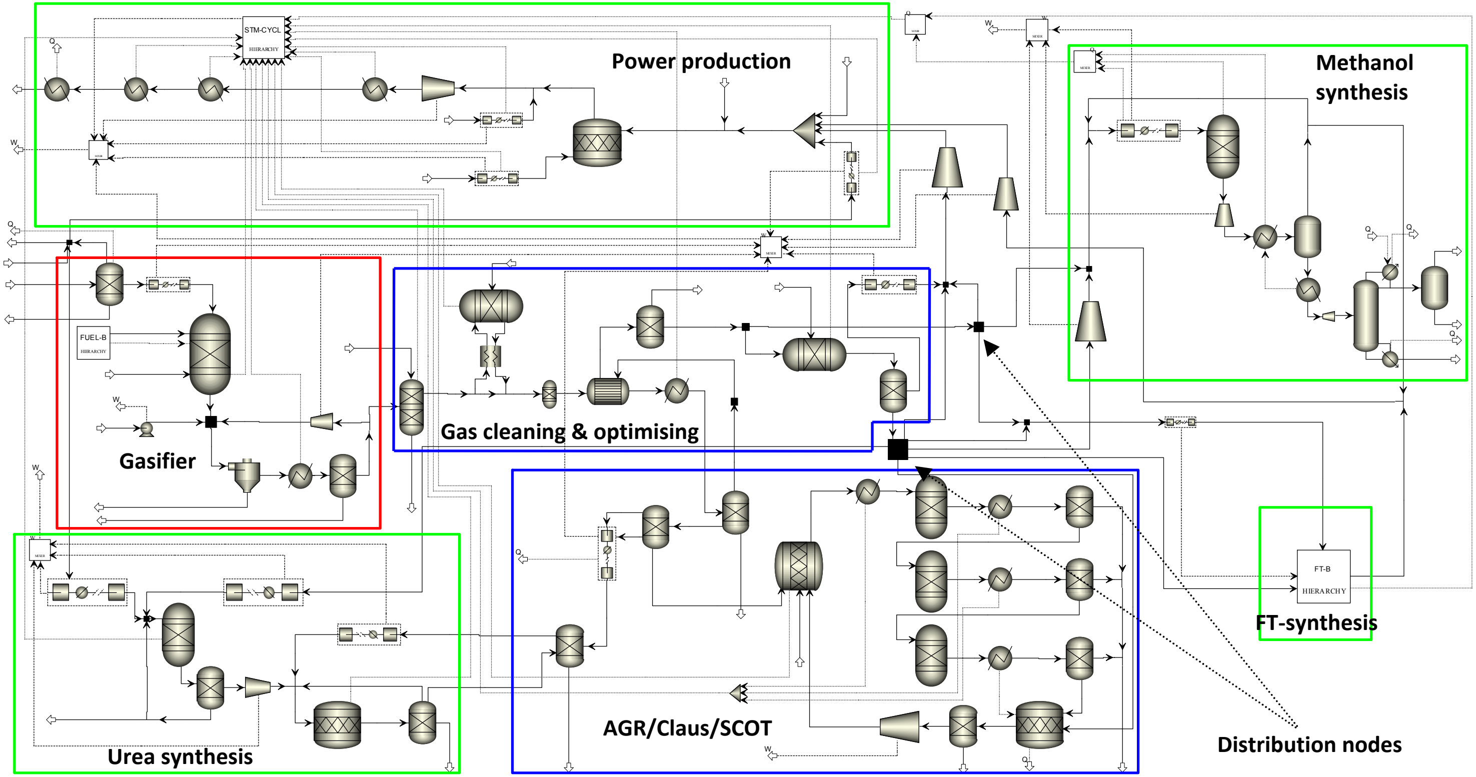


Figure 2.6 Schematic overview of the AspenPlus flowsheet of the IG-PG facility.
The solid lines are mass flows, the dashed lines are heat flows and the striped lines are electricity flows.

The thermodynamic models used were an important aspect of the Aspen model to accurately predict the behaviour of the different components. Due to different process conditions, different thermodynamic models were used (Table 2.9). The standard conditions used in this study are given in Table 2.8.

Table 2.8 Standard conditions

Parameter	Unit	Value
Temperature	°C	25
Pressure	bar	1.013
Air composition	mol% N_2	0.770
	mol% O_2	0.207
	mol% H_2O	0.0136
	mol% Ar	$9.17*10^{-3}$
	mol% CO_2	$2.97*10^{-4}$

2.4.2 Parameters

The process parameters of the modelled equipment were based on literature data, expert interviews and operational data of a commercially operated, Shell-gasifier based coal-fired power plant.[141] The parameters are given in Table 2.9.

Table 2.9 Overview AspenPlus model parameters. (1)

	Pre-treatment and gasifier
Drying	The fuel is dried to 10, 5 and 2 wt% moisture for EP, TOPS and coal respectively using excess heat from the flue gas. (2)
Milling	Grinding has an electric consumption of 0.015 kW_e/kW_{th}. (3)
Pressurising and Feeding	2.1 and 0.1 Nm^3 CO_2 needed to raise pressure from atmospheric to 48 bar at the inlet of the gasifier and feed the coal and biomass into the gasifier. (4)
ASU	Separated air into pure N_2 (100 mol% N_2) and oxygen (95 mol% O_2, 3.65 mol% argon and 1.35 mol% N_2). Electric consumption is 0.20 kWh/kg O_2. (5)
EF Gasifier	RGibbs reactor. T = 1500°C, p_{out} = 40 bar. Thermodynamic equilibrium. Carbon conversion = 99.5%. 4% HHV input heat loss. A moderator (H_2O; 42 bar, 246°C) is added only with the pure coal cases. (6)
Wet Quench	Mixer. T_{out} = 850°C, Δp = -2 bar. (7)
	Gas Cleaning
Cyclone	Removes all unconverted carbon particles. Δp = -0.01 bar.
Syngas Cooler	Cools syngas from 850°C to 230°C. Δp = -1 bar.
Candle Filter	Removes all carbon. Δp = -0.5 bar.
Wet Scrubber	8 Stage column. Syngas fed from the bottom. Pressurised water fed from the top. Water flow adjusted to remove 95% of HCl. Δp = -1 bar. (8)
Sour WGS	Isotherm RGibbs reactor. T = 250°C, final H_2:CO ratio = 2.3. Δp = -1.5 bar. H_2O:CO ratio in shift > 1.65. (9)
AGR	See AGR section below.
Guard Bed	Removes any remaining H_2S and SO_2. T = 190°C.
Sweet WGS	RGibbs reactor. T = 185°C, final H_2:CO ratio = 100. Δp = -1.5 bar.
H_2-PSA	Separates 90% of the H_2, $p_{exit\ H2}$ = 24.9 bar, $p_{exit\ offgas}$ = 0.7 bar. (10)
	AGR
Rectisol Process	A simplified version of the Rectisol process. Separation occurs based on literature references. Δp = -6.6 bar. 99.9% of the CO_2 is captured. Electric consumption is 137 MJ_e/t CO_2. Heat demand is 90 MJ_{th}/t CO_2. (11)

Claus Burner	Combusts ⅓ of H_2S to SO_2. All other combustible components are combusted. T = 1300°C. Gas is cooled to 330°C. (12)
Claus Conversion	3 Series of catalytic bed, cooler and gas/liquid separators. Catalytic beds are at 330°C, 240°C and 200°C. Δp = -1 bar for each series. Electric consumption is 190 MJ_e/t S. (12)
SCOT	Catalytic reactor, separator and compressor. SO_2 is hydrogenated to H_2S. H_2S is separated, repressurised and mixed with Claus burner feed. Electric consumption is 100 MJ_e/t S. (12)
	Power Production (13)
Syngas Dilution	Pressurised N_2 from the ASU is added to lower syngas LHV to 4.3 MJ/Nm^3. If N_2 supply is insufficient, pressurised steam is added.
Gas Turbine	Adiabatic combustion at 24.9 bar. Turbine inlet temperature (TIT) = 1375°C. Flue gas expanded to 1.5 bar. Isentropic efficiency expander = 0.65 + 0.3577*x - 0.1757*x^2, where x is the mass ratio of the flue gas compared to design specification. At full load isentropic efficiency is 83.2%. Oxygen excess is 10%. (14)
HRSG	Flue gas is cooled in 3 steps to 150°C. Remaining heat used as drying agent. A minimal temperature difference of 10°C is maintained at each heat exchanger.
Steam Turbines	Steam conditions are: HP = 125 bar, 550°C; IP = 42 bar, 550°C; LP = 12.5 bar, 368°C. Isentropic efficiencies: HP = 85%, IP = 93% and LP = 89%. Condenser at 15°C and 0.025 bar. (13)
Condenser	Condenses the steam to liquid water. The released heat is vented T_{exit} = 15°C. (13)
	FT-liquids production
FT-Reactor	T_{exit} = 230°C, p_{exit} = 59 bar. Uses modified Anderson-Schulz-Flory distribution. $CO_{conv.}$ = 90%, offgas to power. (15)
FT Flash	T_{exit} = 5°C. p_{exit} = 54.6 bar.
FT Decanter	T_{exit} = 55°C. Δp = -1 bar.
FT Cracker	T_{exit} = 250°C. Δp = -1 bar, 10% stoichiometric excess H_2.
	Methanol Production (16)
Compressor	3 Stages. p_{exit} = 75 bar. Intercooled to 150-250°C. Inter-stage Δp = 2% of pressure.
MeOH Reactor	T_{exit} = 265°C. p_{exit} = 69 bar, $T_{thermodynamic\ eq.}$ = 250°C.
MeOH Flash 1	T_{exit} = 40°C. p_{exit} = 35 bar. Offgas to power.
MeOH Distiller	10 Stage distiller. p_{exit} = 1 bar. Reflux ratio = 10.
MeOH Flash 2	T_{exit} = 84°C. p_{exit} = 3 bar.
	Urea production (17)
Compressors	1 Stage compressor. p_{exit} = 138 bar.
NH_3 Reactor	T_{exit} = 300°C. p_{exit} = 100 bar. Thermodynamic equilibrium.
NH_3 Separator	4 stage for H_2 and 2 stage for N_2. p_{exit} = 100 bar. Intercooled to 150°C.Inter-stage Δp is = 2% of pressure.
NH3 Recycle	90% is recycled. 10% is bled.
Urea Reactor	T_{exit} = 185°C. p_{exit} = 140 bar. $CO_{2\ conv.}$ = 60%. (18)
	Auxiliary equipment
Compressors	$Eff_{isentropic}$ = 0.83. For oxygen this is 0.74. For CO_2 this is 0.8. $Eff_{mechanic}$ = 1. (19)
Water Pumps	Eff_{pump} = 92.5%. $Eff_{mechanic}$ = 97%

(1) The used thermodynamic models were: Solids for pre-treatment; PR-BM for gasifier, gas cleaning optimising, methanol section and power section; SR-Polar for AGR; SRK for FT-section; RK-Soave for urea section; and STMNBS2 for the steam section. Standard conditions are 25°C and 1.013 bar.

(2) During pelletising the moisture content of the pellets is reduced to 6-9 wt%. Assuming that the pellets re-adsorb a little moist, it was expected that EP arrive with a moisture content of 10 wt%. Their hydrophilic nature prevents further - efficient - drying. TOPS have an initial

moisture content of 5 wt%. Due to their coal-like hydrophobic nature, they are dried to 2 wt%. Coal is dried to 2 wt%.[86,101]

(3) Coal particles were ground to 0.1 mm, biomass particles were ground to 1.0 mm. The bigger particle size, but more energy intensive grinding of biomass results in more or less similar energy consumption per kW_{th} compared to coal.[100]

(4) The values were based on the assumption that particles with a size of 0.1 mm for coal and 1.0 mm for TOPS and biomass are used. An additional 1.6 and 0.1 Nm^3 CO_2 was eventually vented to prevent syngas from escaping.[124]

(5) A standard non-integrated cryogenic ASU was simulated. A conservative energy consumption of 0.2 kWh/kg O_2 was used.[106]

(6) 4% of feedstock HHV is removed due to cooling. This is on the high side for Shell EF gasifiers. The high temperature justifies thermodynamic equilibrium. Literature data and expert interview indicates Teq.= 1500°C should be used.[142] The amount of moderator was varied to maximise overall output in the XtL and XtM cases. Only if pure coal was used, a moderator was needed. Even with 20% biomass, adding a moderator reduced overall output.

(7) Standard quench temperature is 900°C. As biomass increases fouling problems, 850°C was used.[114]

(8) The wet scrubber conditions were based on data from Chiesa et al.[88]

(9) The H_2:CO ratio was based on the FT process as it requires the largest CO fraction. The ratio will be obtained after the AGR. The H_2O:CO ratio was set at 1.65 instead of the conventional 2.0 based on expert interview.[143]

(10) It was assumed that the pure H_2 has no pressure drop. The H_2-PSA behaviour was based on Shell data.[144]

(11) The Rectisol process conditions were based on data from Kreutz et al.[24]

(12) Based on the theoretical optimal performance of a Claus unit. The high temperature was needed for NH_3 combustion. The beds operate at thermodynamic equilibrium.[128]

(13) Based on open literature, NUON experimental data and expert interview.[77,141,142,143] Pressure losses in the steam cycle due to part-load operation were ignored.

(14) Based on Alstom data and expert interview at KEMA.[95,130]

(15) FT-conversion based on previous literature. The Anderson-Schulz-Flory distribution was adjusted for increased C_1-C_4 production.[77]

(16) Methanol conversion based on data from E. Heydorn and W. Diamond.[78]

(17) The urea section was modelled based on AspenTech corporate flowsheets.[145]

(18) The CO_2 needed for the urea synthesis was first compressed to 110 bar together with the CO_2 for CCS. It was subsequently compressed to 140 bar required for the urea synthesis.

(19) Different compressor characteristics were used. To compress air for the gas turbine a 4 stage compressor with an exit pressure of 101% of gas turbine pressure and no intercooling was simulated. To compress syngas after the cleaning and optimising section a 2 stage compressor with an intercooler to 100°C and a 2% pressure drop was simulated. To compress CO_2 to 110 bar a 5 stage compressor with intercoolers to 40°C was simulated.

2.5 Case studies

To determine how plant efficiencies and output change due to flexibility, several case studies have been simulated. The following limitations were enforced for all cases:

- Three types of feedstock are used: Illinois #6 coal, TOPS and EP. Three main coal:biomass fractions are used: (1) pure coal; (2) 50/50 energy% coal:biomass (the current technical limit); and (3) pure biomass;

- Gasifier capacity is kept constant at a gas generating capacity of 136 Nm^3/s, equivalent to 2000 MW_{th} Illinois #6 coal input[4];
- Syngas conversion processes offgas is directed to the gas turbine to generate electricity;
- The syngas H_2:CO ratio after the AGR is kept constant at 2.3, regardless of the used feedstock. This is done by adjusting the by-pass ratio of the sour WGS reactor;
- H_2 production in the sweet WGS is kept to a minimum;
- The energy density of the syngas entering the gas turbine is kept constant by dilution with N_2 and, if necessary, steam;
- All cases capture and compress CO_2, regardless of feedstock or production;
- Reactors and separators have - as described above - a minimal load demand of 40%. The gas turbine is calibrated till 40% load.

2.5.1.1 Reference cases

In this case study all components were optimally sized for the specific feedstock. There is no feedstock or production flexibility.

This gives a reference point for the flexibility cases. The following simulations were performed:

- XtP: maximal electricity production;
- XtL: maximal FT-liquids production;
- XtM: maximal methanol production;
- XtU: maximal urea production.

2.5.1.2 Effect of feedstock substitution

In this case study coal was gradually replaced by either Eucalyptus pellets (EP) or torrefied wood pellets (TOPS). As more coal is being replaced by biomass, the total chemical energy of the syngas drops, i.e. the syngas energy density decreases. This affects chemical and electricity output. The following simulations were performed:

- 1-P: maximal electricity production while increasing the biomass fraction;
- 1-L: maximal FT-liquids production while increasing the biomass fraction;
- 1-M: maximal methanol production while increasing the biomass fraction;
- 1-U: maximal urea production while increasing the biomass fraction.

2.5.1.3 Effect of end-product substitution

In this case study the chemical output was reduced in favour of electricity output by diverting syngas form the chemical production section to the power section. The feedstock remains identical. The following simulations were investigated:

- 2-L: switching between electricity and FT-liquid production;
- 2-M: switching between electricity and methanol production;
- 2-U: switching between electricity and urea production.

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[4] Previous research indicated that gasification facilities under 1000 MW_{th} suffer greatly from reduced economies of scale.[75] To ensure that both the chemical conversion and the power sections are sufficiently sized, a base size of 2000 MW_{th} coal input was taken.

2.5.1.4 *Effect of changing production by feedstock substitution*

In this case study biomass was replaced by coal to increase electricity output. It was assumed that the base case was biomass conversion to chemicals. This can occur when biomass is less expensive than coal, by e.g., biomass subsidy or CO_2 tax. When electricity prices rise, biomass was replaced by coal. The additional energy in the syngas was directed to an over-dimensioned gas turbine, resulting in increased electricity output. The chemical production was kept at a constant level. This way the chemical production section always operated at optimal performance and capacity. The following simulations were investigated:

- 3-L: producing FT-liquids at a constant level and increasing electricity production by substituting biomass with coal;
- 3-M: producing methanol at a constant level and increasing electricity production by substituting biomass with coal;
- 3-U: producing urea at a constant level and increasing electricity production by substituting biomass with coal.

2.6 Results

The assumption that the gas volume flow of the gasifier is the limiting factor has a direct impact on the input of the gasifier and thus the gas flows throughout the entire IG-PG facility[5]. The differences between feeding biomass or coal are displayed in Table 2.10.

Table 2.10 General parameters gasifier.

	Unit	EP	TOPS	Coal
Feedstock	MW_{th} fuel	1,464	1,696	2,000
input	t dry fuel/h	274	283	236
Oxygen	kg O_2/kg dry fuel	0.61	0.59	0.88
consumption	kg O_2/GJ_{th} fuel	32	28	29
CGE	%	70	76	82
Raw syngas composition exiting gasifier				
H_2O	(mol% wet)	25	14	1
H_2	(mol% dry)	27	29	30
CO	(mol% dry)	54	61	66
CO_2	(mol% dry)	17	9	0
Other	(mol% dry)	2	2	3

Total chemical energy input drops with increasing coal:biomass fraction. Compared to pure coal the energy drops to 85% for pure TOPS and 73% for pure EP. Thermal input at different coal:biomass fractions is displayed in Figure 2.7. As a moderator is added only with the pure coal case, the thermal input at pure coal is lower than expected when looking at the thermal input of the other coal:biomass fractions. The feedstock mass input in the gasifier increased to 120% for pure TOPS and 116% for pure EP compared to pure coal.

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[5] A possible alternative is to over-dimension the gasifier to maintain a constant chemical energy input. In that case the gasifier is operated at part-load when using coal and at full load when using biomass.

However, as the specific oxygen consumption of the biomass is much lower (0.6 t O_2/t dry biomass) than that of coal (0.9 t O_2/t dry coal), absolute oxygen consumption is reduced to 81% when biomass is used instead of coal. The lower specific oxygen consumption is a result of the higher oxygen-atom fraction in the biomass.

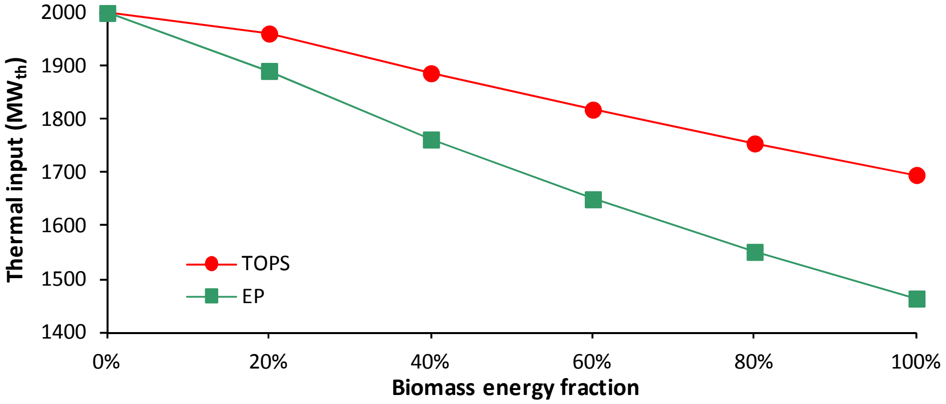

Figure 2.7 Effect biomass substitution on thermal input while maintaining constant gas volume generation in the gasifier.

The lower energy density of biomass results in a higher fraction of the feedstock that is combusted. This leads to higher CO_2 and H_2O concentrations in the raw syngas and lower cold gas efficiencies (CGE) of the gasifier.

Another difference between biomass and coal is ash composition and input. Slag (molten ash) production drops by a factor 7 with pure TOPS and a factor 19 with pure EP compared to pure coal. When slag production is too low, the protective slag layer at the inside of the gasifier may disappear. This can damage the gasifier and reduce availability. Also, the ash composition is different. Biomass ash usually has a higher melting point. Depending on the melting temperature of the ash, a slagging agent - which lowers the melting point of the ash -must be added to ensure a good melt of the ash.

Biomass, but not TOPS, is much more fibrous than coal, requiring a hammer mill instead of crusher. However, biomass can use the more efficient hydraulic piston with screw feeding system. It is therefore appropriate to use a dedicated pre-treatment train for EP and a different train for coal, both with their own pressurised storage bins. TOPS can be used in either feeding system, although the biomass feeding system is more efficient (Table 2.3).

The higher the biomass fraction, the higher the moisture content in the raw syngas. This reduces water consumption of the wet scrubber as the cooling - and resulting condensation of the moisture - already results in the water flows needed to remove the halides.

As known, coal is sulphur-rich, while TOPS and EP are sulphur-poor. This results in a reduction in sulphur content in the syngas when switching from coal to biomass. Also, the lower CGE results in a higher CO_2 content: coal derived syngas entering the AGR contains

5300 ppm H_2S and 21% CO_2; TOPS derived syngas contains <100 ppm H_2S and 23% CO_2; EP derived syngas contains <100 ppm H_2S and 25% CO_2. The low sulphur flows at high biomass fractions causes problems in the Claus unit. Therefore, sulphur needs to be recycled.

By altering the recirculation rates of the Rectisol, the sweet syngas leaving the AGR is almost similar in composition, regardless of the feedstock (Table 2.11). Total sweet syngas flow does differ significantly. Therefore, downstream conversion sections only have to cope with a change in gas flow and not with a change in gas composition.

Table 2.11 Sweet syngas characteristics.

	Unit	EP	TOPS	Coal
H_2O	mol% wet	1	1	0
H_2	mol% dry	68	68	68
CO	mol% dry	29	30	30
CO_2	mol% dry	0	0	0
Other	mol% dry	2	2	2
Energy density	MJ_{HHV}/Nm^3	12	12	12
Gas volume flow	Nm^3/s	42	51	64

Modern gas turbines are tuned for a specific gas composition and flow. Flexibility means that they must handle different syngas compositions and flows. Of special interest are the syngas H_2:CO ratio, energy density and flow. The H_2:CO ratio varies significant when altering methanol production. For FT-liquids and urea production, this variation is limited (Table 2.12).

Table 2.12 Gas Turbine characteristics when changing feedstock from pure biomass to pure coal.

		Electricity	FT-liquids	Methanol	Urea
H_2:CO ratio syngas to GT					
Reference	Coal, Tops and EP	2.3	4.3	12.6	2.8
Constant chemical production	TOPS → Coal	N.A.	2.9 → 4.3	7.6 → 12.7	2.6 → 2.8
	EP → Coal	N.A.	2.6 → 4.2	5.9 → 12.6	2.5 → 2.8
Energy density syngas to GT before N_2 dilution (MJ_{LHV}/Nm^3)					
Reference	Coal, Tops and EP	11	11	9	8
Constant chemical production	TOPS → Coal and EP → Coal	N.A.	11	9	8
Energy flow syngas to GT (MW_{LHV})					
Reference	Coal	1428	259	690	721
	TOPS	1148	203	550	580
	EP	908	161	436	459
Constant chemical production	TOPS → Coal	N.A.	203 → 483	550 → 824	580 → 854
	EP → Coal	N.A.	161 → 681	436 → 944	459 → 979

2.6.1 Reference cases

The overall energy efficiency and carbon balances of the extreme cases - pure biomass or pure coal - without substitution are given in Table 2.13. Using biomass results in slightly lower efficiencies compared to using coal. When using TOPS this difference is even smaller than when using EP. Overall efficiencies of producing methanol are relatively low. The main reason is the mediocre conversion of the liquid phase methanol synthesis process[6]. This, combined with a once-through concept, results in a relatively low methanol and high electricity production. The low conversion efficiency of the electricity production results in a low overall efficiency for the XtM facilities.

Table 2.13 Overall energy efficiencies of XtY facilities.

Energy Efficiency		**Coal**	**TOPS (1)**	**EP**
Electricity (XtP)	**Power**	**40%**	**39% (35%)**	**38%**
FT-Liquids (XtL)	FT	49%	47% (42%)	43%
	Power	10%	11% (10%)	12%
	Total	**60%**	**58% (53%)**	**55%**
Methanol (XtM)	MeOH	33%	31% (28%)	29%
	Power	21%	21% (19%)	21%
	Total	**53%**	**52% (47%)**	**49%**
Urea (XtU)	Urea	29%	28% (25%)	25%
	Power	22%	22% (20%)	22%
	Total	**51%**	**50% (45%)**	**47%**

(1) During torrefaction roughly 10% of the biomass energy is lost. The values in bracket include this penalty.

The allocation of carbon is displayed in Figure 2.8. As the syngas is shifted to a specific H_2:CO ratio, regardless of production, the amount of CO_2 captured is identical for each feedstock. This also limits the capture rate. The IG-PG facility is not optimised for CO_2 capture, but for flexibility. As a result, between 56% and 65% of the carbon is captured. Only with urea production, a part of the captured CO_2 is used and gets embedded in the product. For the FT-liquids and methanol production the carbon in the syngas is used and not the captured CO_2. Therefore, with urea production less CO_2 needs to be sequestered. The fraction of carbon that is captured increases from coal < TOPS < EP. The reason for this is that a higher fraction of the feedstock is combusted in the gasifier when using biomass as feedstock. CO_2 emissions vary depending on feedstock used and commodity produced. This is because a part of the carbon ends up in the chemical end-products for the FT-liquids and methanol cases. Therefore, the XtP and XtU cases have the largest emissions. As the syngas to FT-liquids conversion is higher than syngas to methanol, the XtL cases have the lowest emissions.

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[6] The LPMeOH process was selected as the improved heat management allows the use of CO-rich syngas. It also improves the conversion compared to the conventional process, although conversion remains mediocre. Finally, the LPMeOH process has been tested with variable production.

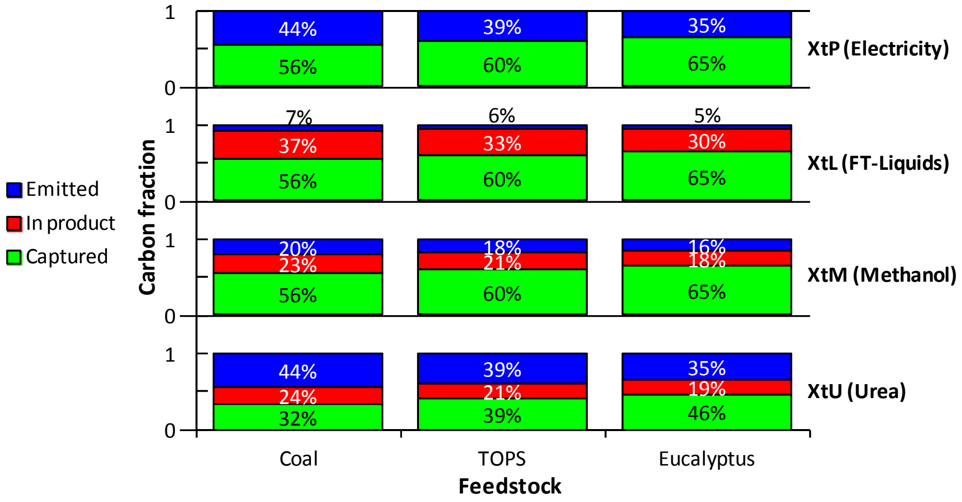

Figure 2.8 Carbon distribution. The numbers display the carbon fractions.

2.6.1.1 *Electricity*

In the following section the results of electricity production with and without flexibility are given. Also, the effects of feedstock substitution are presented.

2.6.1.1.1 *Reference*

Total output drops from 800 MW_e when using pure coal to 668 MW_e (-16%) when using pure TOPS and 550 MW_e process, although conversion remains mediocre. Finally, the LPMeOH process has been tested with variable production. (-31%) when using pure EP (Figure 2.9). This is caused mainly by the lower thermal input when using biomass: thermal input is 2000 MW_{th} for pure coal, but 1696 MW_{th} for pure TOPS and 1464 MW_{th} for pure EP. Total efficiency is hardly affected by the different feedstocks. It drops from 40% for pure coal to 39% for pure TOPS and 38% for pure EP. The net CO_2 emissions of the IGCC with CO_2 capture drop from 343 kg CO_2/MWh for pure coal to -506 kg CO_2/MWh for pure TOPS and -589 kg CO_2/MWh for pure EP. CO_2-neutral electricity is produced at biomass energy fractions of 41% (48 wt%) for TOPS and 39% (50 wt%) for EP. Although the gasifier has a larger feedstock mass input when using biomass compared to coal, total carbon input is lower for the biomass cases as the carbon fraction in coal is much higher. The result is a slight reduction in absolute amount of CO_2 captured. The summarised mass and energy balances are given in Table 2.14-Table 2.16.

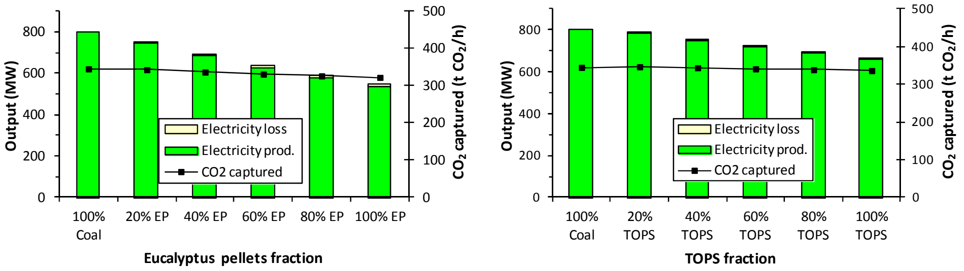

Figure 2.9 Effect different feedstocks on power production.

2.6.1.1.2 Feedstock substitution

When the facility is sized for pure coal, but biomass is used, the gas turbine operates at part-load. However, the inlet temperature can be maintained at the desired level of 1375°C. Although the efficiency of the gas turbine is reduced, total electricity output remains almost constant (Figure 2.9). The largest difference occurs at the substitution of coal for EP. The reduction in electric output compared to the EP-IGCC base case is 13 MW_e or 2%. There are three main reasons why this reduction is so small. First, a less efficient gas turbine results in a hotter flue gas. This improves the electricity production from the steam cycle. Second, the GT can reduce the air flow, thereby keeping the flue gas at a high temperature. Third, the gas turbine produces a smaller fraction of the total output than a standard NGCC. This is due to the additional heat streams inherent to an IGCC.

Table 2.14 Summary mass balances XtP reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen Gasifier	t/h	168	183	200	208	199	181	168
Syngas to Power	t/h	173	190	211	216	201	166	139
Air to GT	t/h	3.37k	3.71k	4.1k	4.19k	3.89k	3.2k	2.67k
Flue gas	t/h	3.77k	4.14k	4.58k	4.69k	4.34k	3.58k	2.98k
CO_2 emitted	t CO_2/h	220	242	268	274	254	209	174

Table 2.15 Summary energy balances XtP reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total internal power use	MW_e	-146	-156	-168	-172	-164	-148	-135
Net power production	MW_e	668	722	785	800	749	637	550
Total efficiency	$\%_{HHV}$	39.4	39.7	40.0	40.0	39.6	38.6	37.6

Table 2.16 Summary carbon balances XtP reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	60	66	73	75	69	57	48
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in flue gas	%	39	41	43	44	42	39	35
Net CO_2 emission	t C/h	-92	-32	38	75	34	-35	-88
CO_2 emissions	kg CO_2/MWh	-504	162	177	343	167	-200	-589
CO_2 emissions	kg CO_2/GJ_e	-141	-45	49	95	47	-56	-164

2.6.1.2 FT-liquids

In the following section the results of FT-liquids production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

2.6.1.2.1 Reference

Overall plant output (FT-liquids and electricity) drops from 1198 MW using pure coal to 990 MW with pure TOPS and 804 MW with pure EP (Figure 2.10). Besides this drop in output, also the ratio FT-liquids:electricity changes; a low ratio means a relatively low chemical and high electricity output. Increasing the biomass fraction reduces this ratio. This effect is larger for EP compared to TOPS. Total efficiency drops from 60% for pure coal to 58% for pure TOPS and 55% for pure EP. Net CO_2 emissions drop from 44 t CO_2/h with pure coal to -523 t CO_2/h with pure TOPS and -469 t CO_2/h with pure EP. CO_2-neutral production is achieved at co-feeding rates of 7% TOPS or 6% EP on an energy basis. This does not include the carbon bound in the FT-liquids. If the carbon in the FT-liquids counts as GHG emission, CO_2-neutral FT-liquids production needs the same biomass fraction as electricity production (41% for TOPS and 39% for EP). The summarised mass and energy balances are given in Table 2.17-Table 2.19.

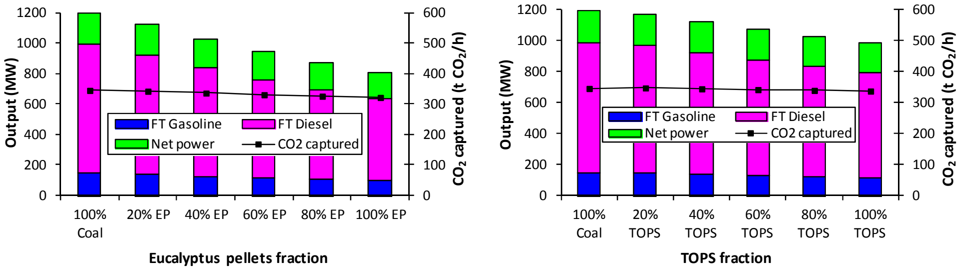

Figure 2.10 Effect different feedstocks on FT-liquids production.

Table 2.17 Summary mass balances XtL reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen Gasifier	t/h	168	183	200	208	199	181	168
Syngas to Power	t/h	3	3	3	3	3	2	2
Air to GT	t/h	0.61k	0.67k	0.75k	0.77k	0.71k	0.58k	0.48k
Flue gas	t/h	0.68k	0.75k	0.83k	0.87k	0.79k	0.65k	0.54k
FT-liquids	t/h	61	67	74	75	70	58	48
CO_2 emitted	t CO_2/h	31	34	38	40	36	30	25

Table 2.18 Summary energy balances XtL reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total Internal power use	MW_e	-143	-152	-163	-167	-160	-144	-132
Net power production	MW_e	192	198	206	210	200	185	173
FT-Gasoline production	MW_{HHV}	118	130	144	147	137	113	94
FT-Diesel production	MW_{HHV}	681	748	825	841	784	645	537
Total efficiency	$\%_{HHV}$	58.4	59.2	59.9	59.9	59.3	57.1	54.9

Table 2.19 Summary carbon balances XtL reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in FT-liquids	t C/h	51	56	62	63	59	48	40
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	9	10	11	11	10	8	7
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in products	%	33	35	37	37	36	33	30
Fraction C in flue gas	%	6	6	6	7	6	6	5
Net CO_2 emission	t C/h	-143	-88	-24	12	-24	-83	-128
CO_2 emissions chemical (1)	kg CO_2/GJ	-206	-124	-46	-8	-48	-135	-234

(1) Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

2.6.1.2.2 Feedstock substitution

If coal is substituted by biomass, the ratio FT-liquids:electricity is kept constant and part-load effects are included, the worst case - substituting pure coal for pure EP - has a 3 MW_e drop in electric output compared to the pure EP reference BtL case.

The effect of constant FT-liquid production and substituting biomass for coal is displayed in Figure 2.11 and Table 2.20. The TOPS cases are displayed on the left and the EP cases on the right. FT-liquid production is set at either maximum FT-liquid production at pure biomass or at a 50/50 coal:biomass ratio on an energy basis. The reference CtL case is displayed in the middle of both graphs. As can been seen, power production is increased when coal is used. Also, CO_2 emissions are higher. Compared to the reference cases, CO_2 capture rates are not affected by the substitution. This is because the AGR is adjusted to capture a specific fraction (Table 2.9). As FT-liquid production is more efficient than electricity production, efficiency drops when using coal. The modelled gas turbine has good part-load behaviour down to 40%. Only the pure EP case demands a lower load.

Table 2.20 Effect coal:biomass substitution with constant FT-liquid production. (1)

	Substitution degree	FT-prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions (t CO_2/h)	Total efficiency	GT load
TOPS	0% → 50%	899	263 → 197	63 → 37	58% → 59%	64%
	0% → 100%	799	323 → 184	87 → 32	56% → 58%	43%
EP	0% → 50%	794	326 → 181	88 → 32	56% → 57%	42%
	0% → 100%	631	423 → 163	126 → 23	53% → 54%	24%

(1) Before the arrow represents feeding only coal. After the arrow represents (partly) substitution of coal by biomass.

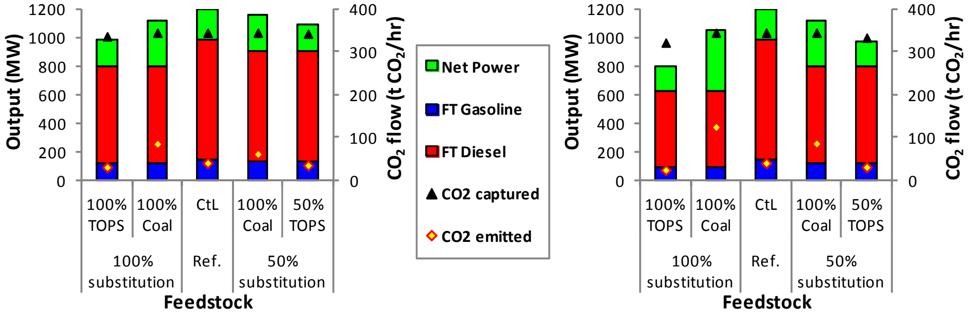

Figure 2.11 Impact of feedstock substitution with constant FT-liquids production.

2.6.1.2.3 *Production substitution*

With pure coal as feedstock and switching from FT-liquids production to electricity production reduces overall conversion efficiency from 59% to 40%. The difference between electricity output in full FT-production and the CtL base case is 18 MW_e. For this extreme case, the gas turbine operates at 19% part-load, which is very inefficient. The amount of CO_2 captured remains constant but CO_2 emissions increase as less CO_2 is bound in the FT-liquids. This increase is from 42 t CO_2/h to 274 t CO_2/h. The relation between FT-liquids:electricity ratio and output is displayed in Figure 2.12. Similar results have been found for TOPS and EP.

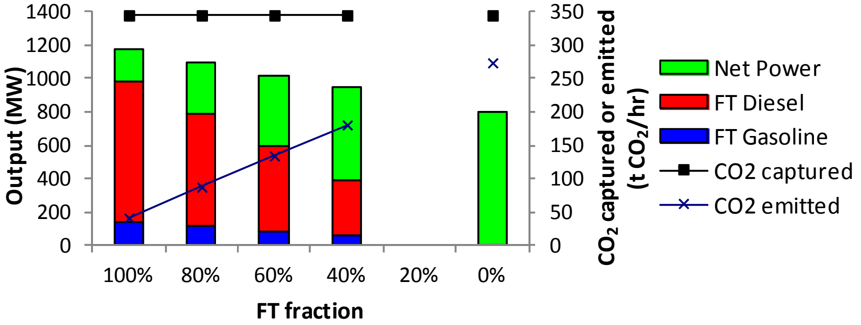

Figure 2.12 Impact substituting FT-production for power production.

2.6.1.3 Methanol

In the following section the results of methanol production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

2.6.1.3.1 Reference

Overall plant output drops from 1066 MW using pure coal to 886 MW with pure TOPS and 724 MW with pure EP (Figure 2.13). The ratio methanol:electricity decreases with increasing biomass fraction. The effect is larger for EP than for TOPS. Total efficiency drops from 53% for pure coal to 52% for pure TOPS and 49% for pure EP. Net CO_2 emissions drop from 133 t CO_2/h with pure coal to -452 t CO_2/h with pure TOPS and -414 t CO_2/h with pure EP. This does not include the carbon bound in the methanol. If the carbon in the methanol counts as GHG emission, CO_2-neutral methanol production needs the same biomass fraction as electricity production (41% for TOPS and 39% for EP). If the carbon in the methanol is not considered as GHG emission, CO_2-neutral production is achieved at co-feeding rates of 20% TOPS or 19% EP on an energy basis. The summarised mass and energy balances are given in Table 2.21-Table 2.23.

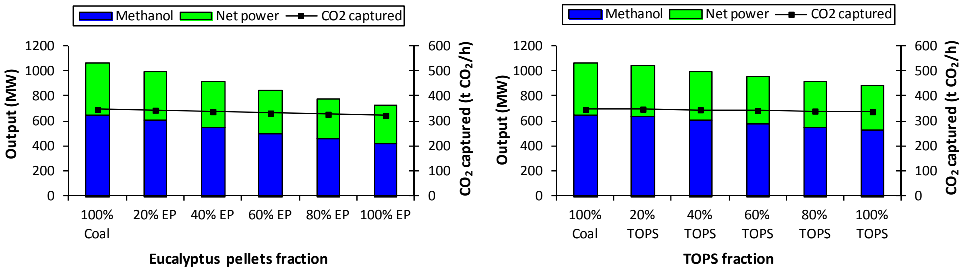

Figure 2.13 Effect different feedstocks on methanol production.

Table 2.21 Summary mass balances XtM reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen Gasifier	t/h	168	183	200	208	199	181	168
Syngas to Power	t/h	83	91	101	104	96	79	66
Air to GT	t/h	1.6k	1.77k	1.95k	2.01k	1.85k	1.53k	1.27k
Flue gas	t/h	1.77k	1.95k	2.16k	2.22k	2.04k	1.68k	1.4k
Methanol	t/h	84	92	102	103	96	80	67
CO_2 emitted	t CO_2/h	99	108	120	124	114	94	78

Table 2.22 Summary energy balances XtM reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total Internal power use	MW_e	-140	-149	-160	-163	-156	-141	-130
Net power production	MW_e	357	380	407	415	391	342	304
MeOH production	MW_{HHV}	529	580	639	651	607	502	420
Total efficiency	$\%_{HHV}$	52.3	52.8	53.3	53.3	52.8	51.1	49.5

Table 2.23 Summary carbon balances XtM reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in methanol	t C/h	32	35	38	39	36	30	25
Carbon captured	t C/h	92	93	95	94	94	90	88
Carbon emitted	t C/h	27	30	33	34	32	26	22
Fraction C captured	%	60	58	56	56	57	61	65
Fraction C in products	%	21	22	23	23	22	20	18
Fraction C in flue gas	%	18	19	20	20	19	18	16
Net CO_2 emission	t C/h	-123	-66	-0	36	-2	-64	-113
CO_2 emissions chemical (1)	kg CO_2/GJ	-305	-182	-64	-7	-67	-198	-345

(1) Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

2.6.1.3.2 Feedstock substitution

If coal is substituted by biomass, the ratio methanol:electricity is kept constant and part-load effects are included, the worst case - substituting pure coal for pure EP - has a 6 MW_e drop in electric output compared to the pure EP reference BtM case.

The effect of constant methanol production and substituting biomass for coal is displayed in Figure 2.14 and Table 2.24. Similar effects as those occurring with the FT-liquids cases can be seen, although less severe. The smaller impact is a result of the poor conversion of the LPMeOH synthesis[7]. This, combined with a once-through concept, results in large amounts of unreacted syngas directed to the gas turbine. This results in an already significant load for the gas turbine. Compared to the FT-liquids cases, the additional syngas to the gas turbine - caused by the substitution of biomass for coal - absolutely similar, but relatively much smaller.

Table 2.24 Effect coal:biomass substitution with constant methanol production. (1)

	Substitution degree	MeOH prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions (t CO_2/h)	Total efficiency	GT load
TOPS	0% → 50%	594	449 → 384	139 → 113	52% → 53%	82%
	0% → 100%	529	487 → 350	154 → 100	51% → 52%	66%
EP	0% → 50%	526	489 → 346	154 → 100	51% → 51%	66%
	0% → 100%	420	552 → 290	178 → 79	49% → 49%	46%

(1) Before the arrow represents feeding only coal; after the arrow represents (partly) substitution of coal by biomass.

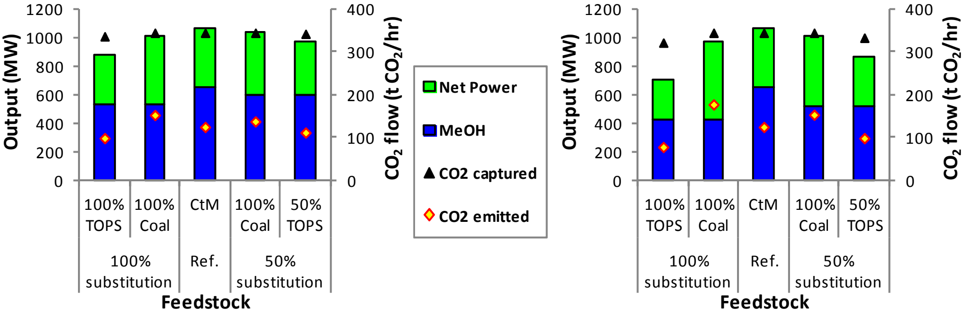

Figure 2.14 Impact of feedstock substitution with constant methanol production.

2.6.1.3.3 Production substitution

Substituting production from methanol to electricity gives similar results as with the substitution of FT-liquids and electricity. Overall efficiency drops from 52% for methanol production to 40% for only electricity production. The difference between electricity output in full methanol production and the CtM base case is 20 MW_e. For this extreme case, the gas turbine operates at 48% part-load. CO_2 emissions increase from 126 t CO_2/h to 274 t CO_2/h. The relation between methanol:electricity ratio and output is displayed in Figure 2.15.

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[7] This process was selected for its superior heat management and part-load performance and its lower capital costs compared to the conventional methanol synthesis process.

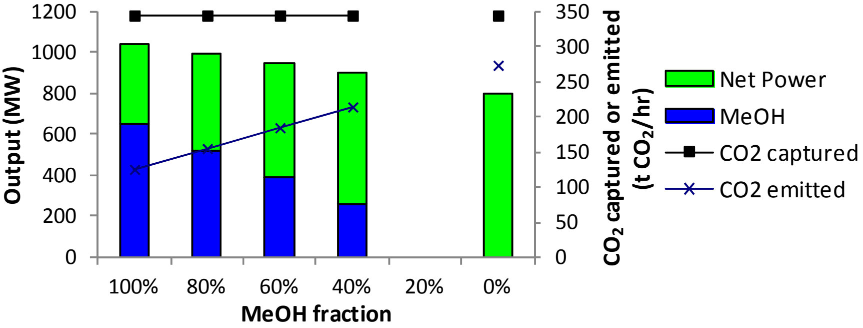

Figure 2.15 Impact substituting methanol production for power production.

2.6.1.4 Urea

In the following section the results of urea production with and without flexibility are given. Also, the effects of feedstock and production substitution are presented.

2.6.1.4.1 Reference

Overall plant output drops from 1015 MW using pure coal to 841 MW with pure TOPS and 686 MW with pure EP (Figure 2.16). The ratio urea:electricity decreases with increasing biomass fraction. The effect is larger for EP than for TOPS. Total efficiency drops from 51% for pure coal to 50% for pure TOPS and 47% for pure EP. Net CO_2 emissions drop from 421 t CO_2/h with pure coal to -220 t CO_2/h with pure TOPS and -229 t CO_2/h with pure EP. This does not include the carbon bound in the urea. If the carbon in the urea counts as GHG emission, CO_2-neutral urea production needs the same biomass fraction as electricity production (41% for TOPS and 39% for EP). If the carbon in the urea is not considered as GHG emission, CO_2-neutral production is achieved at co-feeding rates of 62% TOPS or 58% EP on an energy basis. The summarised mass and energy balances are given in Table 2.25-Table 2.27.

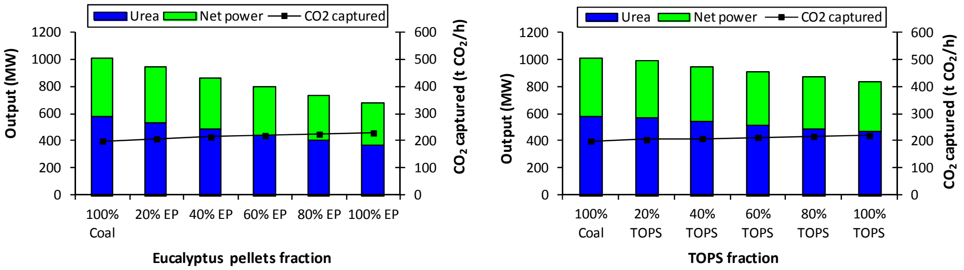

Figure 2.16 Effect different feedstocks on urea production.

Table 2.25 Summary mass balances XtU reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass	t/h	283	182	65		71	186	274
Coal	t/h		86	185	236	178	78	
Oxygen Gasifier	t/h	168	183	200	208	199	181	168
Syngas to Power	t/h	220	242	269	275	255	211	176
Air to GT	t/h	1.72k	1.89k	2.09k	2.14k	1.98k	1.63k	1.36k
Flue gas	t/h	1.94k	2.14k	2.36k	2.42k	2.24k	1.85k	1.54k
Urea	t/h	160	176	195	200	185	152	127
CO_2 emitted	t CO_2/h	218	240	265	271	251	207	172

Table 2.26 Summary energy balances XtU reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Biomass input	MW_{HHV}	1696	1091	392		378	990	1464
Coal input	MW_{HHV}		727	1569	2000	1513	660	
Total Internal power use	MW_e	-156	-167	-180	-184	-176	-157	-143
Net power production	MW_e	372	397	425	431	408	356	316
Net power production	MW_e	468	515	569	583	540	444	370
Total efficiency	$\%_{HHV}$	49.6	50.1	50.7	50.7	50.1	48.5	46.9

Table 2.27 Summary carbon balances XtU reference cases.

	Unit	TOPS			Coal	EP		
Biomass fraction	%	100	60	20	0	20	60	100
Carbon input from biomass	t C/h	152	98	35		35	92	136
Carbon input from coal	t C/h		62	133	169	128	56	
Carbon in urea	t C/h	32	35	39	40	37	30	25
Carbon captured	t C/h	60	58	56	54	57	60	63
Carbon emitted	t C/h	60	66	73	75	69	57	47
Fraction C captured	%	39	36	33	32	35	41	46
Fraction C in products	%	21	22	23	24	23	21	19
Fraction C in flue gas	%	39	41	43	44	42	38	35
Net CO_2 emission	t C/h	-92	-32	38	75	34	-35	-88
CO_2 emissions chemical (1)	kg CO_2/GJ	-277	-137	-4	60	-8	-157	-324

(1) Exported electricity is credited for 95 kg CO_2/GJ_e. This is subtracted from the carbon emissions.

2.6.1.4.2 Feedstock substitution

If coal is substituted by biomass, the ratio urea:electricity is kept constant and part-load effects are included, the worst case - substituting pure coal for pure EP - has a 7 MW_e drop in electric output compared to the pure EP reference BtU case.

The effect of constant urea production and substituting biomass for coal is displayed in Figure 2.17 and Table 2.28. Urea production does not differ that significantly. The reason for this is that in the base cases significant amounts of syngas are already directed to the gas turbine to maintain a sufficient energy density in the gas turbine. In all cases the gas turbine operates above the 40% load limit.

Table 2.28 Effect coal:biomass substitution with constant urea production. (1)

	Substitution degree	Urea prod. (MW_{th})	Power prod. (MW_e)	CO_2 emissions (t CO_2/h)	Total efficiency	GT load
TOPS	0% → 50%	532	463 → 401	273 → 247	50% → 50%	84%
	0% → 100%	468	500 → 366	274 → 219	49% → 49%	68%
EP	0% → 50%	471	502 → 361	274 → 218	49% → 49%	68%
	0% → 100%	370	566 → 302	274 → 174	47% → 46%	47%

(1) Before the arrow represents feeding only coal; after the arrow represents (partly) substitution of coal by biomass.

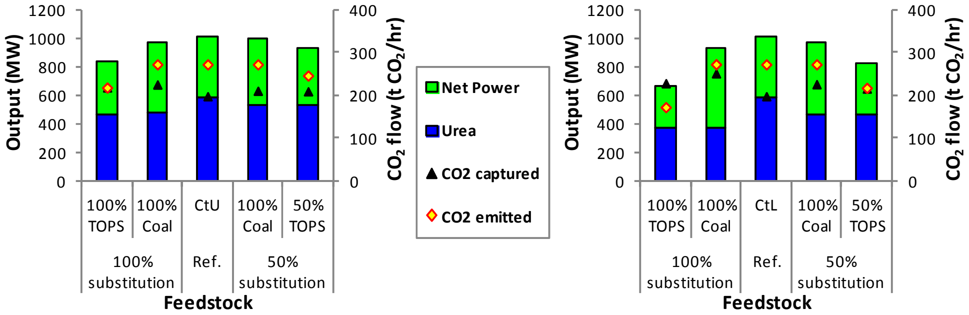

Figure 2.17 Impact of feedstock substitution with constant urea production.

2.6.1.4.3 *Production substitution*

When substituting urea production for electricity production overall efficiency drops from 50% for urea production to 40% for only electricity production. The difference between electricity output in full urea production and the CtU base case is 19 MW_e. For this extreme case, the gas turbine operates at 52% part-load. CO_2 emissions remain constant at 274 t CO_2/h. But, as less captured CO_2 is used for urea production, the amount of CO_2 that needs to be stored increases from 199 t CO_2/h to 345 t CO_2/h. The relation between urea:electricity ratio and output is displayed in Figure 2.18.

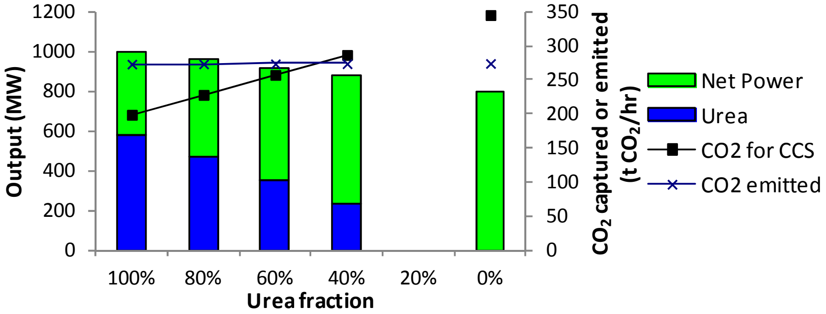

Figure 2.18 Impact substituting urea production for power production.

2.7 Discussion

2.7.1 Model validation

A drawback of our modelling studies is that results have not been tested. To validate the model two different parameters were compared with public available literature sources. The first parameter is the CGE of the gasifier. The obtained CGE for coal (81%) matches the value reported in literature (79-83%).[79,115,116] For EP and TOPS no CGE values using EF gasifiers were found in open literature.

The second parameter is overall plant efficiencies. For coal and EP these are slightly higher than open literature values (see Table 2.29). Values for TOPS using EF gasifiers were not found. Possible explanations are that the modelled IG-PG facility only captures part of the CO_2 - thereby reducing the CO_2 capture penalty - and the use of a different gas turbine than commonly used in the literature sources. Sometimes a different coal or biomass type is used. This can also result in different overall efficiencies. A possible explanation of the difference between the XtU efficiencies is that the literature value is based on an overall conversion efficiency, while the model value is based on the performance of the different individual components.

Table 2.29 Overall system efficiency comparison between model and literature.

	Literature			**Aspen model**		
	Coal	**TOPS**	**EP**	**Coal**	**TOPS**	**EP**
XtP	32-40 (1)	N.A.	36[24]	40	39	38
XtL	57[146]	N.A.	53[24]	60	58	55
XtM	51[147] (LHV)	N.A.	N.A.	53	52	49
XtU	44[82]	N.A.	N.A.	51	50	47

(1) The first value includes CO_2 capture. The second value excludes CO_2 capture.

2.7.2 Fuel flexibility

Switching from coal to TOPS or EP has some important impacts on feedstock volumes, oxygen consumption and syngas composition and volume:

- The difference in hardness and fibrousness of coal and biomass can cause feeding problems when using one feeding system. To prevent these problems and increase flexibility separate feeding systems and pressurised bins for feedstock storage should be used. This also allows the use of more efficient grinding and pressurising techniques.
- At this time no EF gasifier exists that allows co-firing biomass with a higher fraction than 50% on a mass basis, equivalent to 30% on an energy basis. Main bottlenecks are the low heating value of the biomass and the fouling (see comment below). It is questionable whether the current generation EF gasifiers can co-fire high biomass fractions, but, especially with TOPS, 50% co-firing on an energy basis is technically possible, although it might be uneconomical.
- Oxygen consumption drops by 10%. This holds both for substituting coal for EP or TOPS. Current generation ASU have no problems with this load reduction.
- The differences in coal slag and biomass slag behaviour can create problems. Especially the reduction in slag amount and the increase in slag melting

temperature are problematic. Too little slag can result in a loss of protective slag layer in the gasifier. Adding flux material and recycling slag, by crushing and adding slag can prevent this. It does mean constant monitoring of the slag layer.

- The differences in syngas composition (Table 2.10) can be adjusted and levelled. Therefore, the chemical conversion sections only have to cope with a difference in syngas flow and not in syngas composition.
- The Claus unit will give problems as pure biomass will reduce the sulphur flow to 1% compared to coal. This is much lower than the minimal requirement of the Claus unit. To prevent operational problems, sulphur may need to be recycled. This will slightly lower the total energy efficiency for wood and TOPS.
- Not modelled is the fouling in the syngas cooler. Although the fouling is negligible compared to the main mass and energy flows, the cumulative effect of prolonged fouling is important as this decreases the efficiency and availability of the syngas cooler. The higher chloride concentration in biomass will increase fouling compared to coal.[114] To limit this problem biomass fractions are currently restricted to maximum 50 wt% and the quench temperature is lowered from 900°C to 850°C.
- Changing the gas turbine fuel composition will result in different NO_x emissions. These emissions have not been modelled, but must be kept low enough to fulfil environmental legislation.

2.7.3 Model assumptions

Several assumptions were made to model flexible IG-PG facilities (see §2.5). This led to the configuration displayed in Figure 2.6, but other configurations are also possible:

- In this study the limiting factor for the scale of the facility was the gasifier. It dictated the feedstock input and thus mass and energy balances throughout the facility. Commonly, a facility with gas turbines makes the gas turbines the limiting factor - the facility is dimensioned on the gas turbine. This ensures that the gas turbines are always operated at optimal efficiency. By making the gasifier the limiting factor, the gasifier and gas cleaning train, which are the most capital-intensive part of an IG-PG facility, were optimally used. Also, by having a constant gas stream exiting the gasifier, the effects of flexibility are much clearer and more transparent.
- Alternatively, if the gasifier is the limiting factor, natural gas can be co-fed in the gas turbine to keep the gas turbine at optimal efficiency.
- The H_2:CO ratio was kept constant at 2.3 in the sweet syngas. This was done to keep the gas cleaning and optimising operating as constant as possible. Normally, CCS facilities will try to capture as much CO_2 as possible. Allowing the H_2:CO ratio in the sweet syngas to change can reduce CO_2 emissions to almost zero.
- The model used a sour and a sweet shift, with the AGR in between. Alternatively the train could be: S-capture, sweet shift, CO_2 capture.

2.8 Conclusion

In this study the technical possibilities and impacts on overall efficiency and carbon balance of introducing flexibility to IG-PG facilities was investigated. An IG-PG facility consists of numerous specific components. Of each component the technical performance at full and part-load was investigated. This data was combined with different IG-PG facility configurations in an Aspen- Tech process flowsheet. This allowed the calculation of the different case studies by altering the used feedstock or configuration. By analysing the resulting mass and energy balances the effects of flexibility on IG-PG facilities were determined.

Electricity with CCS can be produced at 40%$_{HHV}$ efficiency. Using TOPS or EP lowers this to 39%$_{HHV}$ and 38%$_{HHV}$ respectively. Of the carbon in the feedstock 56-65%, for coal and EP respectively, is captured by CO_2 removal. This value is much lower than the normally reported 90% capture for IGCC-CCS facilities, due to a much higher CO-concentration in the sweet syngas as the facility is designed for flexibility. The CO is not captured by the acid gas removal. It is combusted in the gas turbine and the carbon is emitted with the flue gas as CO_2. When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 13 MW$_e$ or 2% of overall efficiency compared to the output when using a static IG-PG facility due to part-load penalties.

FT-liquids (and electricity) with CCS can be produced at 55-60%$_{HHV}$ efficiency, with the lower value corresponding to pure EP as feedstock and the higher value to pure coal. Again, between 56% and 65% of the carbon in the feedstock is captured by CO_2 removal. But, an additional 37-30% of the carbon ends up in the FT-liquid. The carbon fraction emitted with the flue gas is 5-7% for pure coal and pure EP respectively. When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 3 MW$_e$ or < 1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties. Variable feedstock with constant chemical production results in a too small gas turbine load for the case where pure EP is substituted by pure coal. The load is reduced from 100% with coal to only 24% with EP. The smallest load reduction is achieved with 50/50 TOPS. Here the gas turbine is reduced to 64% load. Complete substitution of FT-liquids production for electricity production is not straight forward as gas turbine load is reduced to 18%. Possible solutions are adding natural gas as make-up fuel or having multiple smaller gas turbines. Even when FT-liquids production is varied between 60% and 100%, the gas turbine load still decreases to 36%.

Methanol production (and electricity) with CCS has similar results as FT-liquids, although less extreme. Overall efficiency is between 49%$_{HHV}$ and 53%$_{HHV}$. Total carbon emissions are between 16% and 20%. When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 6 MW$_e$ or <1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties. Variable feedstock with constant chemical production does not exceed part-load limits. In the XtM cases the gas turbine already has a higher load compared to the XtL cases. Therefore, although the absolute difference in gas turbine duty is similar to the XtL cases, the gas turbine load is reduced to only 46% when substituting coal for biomass. Complete substitution of methanol production for electricity production is possible. Gas turbine load is reduced to 48%.

Urea production (and electricity) with CCS has overall efficiencies of 47-51%$_{HHV}$. Total carbon emissions are between 35% and 44%. For urea production only carbon in the captured CO_2 is used. The remaining carbon - mostly in the form of CO - is directed to the gas turbine, combusted and emitted. When using pure EP in a full feedstock flexible IG-PG facility lowers total output by 7 MW_e or <1% of overall efficiency compared to the output of a static IG-PG facility due to part-load penalties. Similar to the methanol cases, variable feedstock with constant chemical production is possible. In the most extreme case gas turbine load is reduced to 47%. The normal XtU cases have an even greater gas turbine duty than the XtM cases. Therefore, constant feedstock and complete production substitution between urea and electricity is also possible. The gas turbine load is reduced to 52% when electricity production is substituted by urea production.

IG-PG facilities are very suitable for flexible operation, even though there are technical bottlenecks, significant variation in feedstock and production is possible without drastic reductions in overall efficiencies. The found bottlenecks are:

- To prevent frequent starting-up and shutting down of equipment a minimal 40% load should be maintained in the chemical conversion sections.
- The energy density of biomass is much lower than that of coal. If the gasifier is the limiting factor this results in a significant reduction in energetic input, and thus output, in the facility. This drop is 15% for TOPS and 27% for EP.
- Biomass has a different elemental composition than coal, especially concerning sulphur, chloride and ash content. This can give problems in the gasifier, syngas cooler and Claus unit. Possible solutions are adding flux material, recycling of slag and recycling of sulphur.
- The current maximum co-feeding fraction of biomass is 50% on an energy basis. At these high biomass fractions, the used biomass and coal must be of high quality. This means low moisture content, low ash content and high heating value.
- Producing maximum urea lowers the heating value of the gas stream fed into the gas turbine to below design specification. To maintain a sufficient energy rich gas stream, some of the syngas is directed to the gas turbine instead of the urea synthesis, lowering maximum urea production.
- As the IG-PG facility is designed for multiple chemicals production, optimisation is low. If the IG-PG configuration is limited to only one chemical combined with electricity, optimisation can improve overall efficiency by several %points.

In conclusion:
IG-PG facilities have a large potential for flexibility, even with current technology. But, there is a trade-off between flexibility and overall efficiency. Significant variations, both in feedstock and in production are possible with only limited impact on total efficiency. Overall IG-PG facility efficiency drops from 40% to 38% for power; 60% to 55% for FT; 53% to 49% for MeOH and 51% to 47% for urea production when substituting pure coal for pure EP.

When operating the gasifier at maximum capacity, the biggest effect is the reduction in input, and thus output, when switching from coal to biomass. Torrefaction of biomass to TOPS reduces both the drop in plant efficiency and the drop in input and thus output compared to using raw biomass.

The CO_2 emissions are higher than in a dedicated IGCC-CCS facility. This is primarily a direct result of the used assumption that the H_2:CO ratio of the sweet syngas is always equal to the most demanding chemical conversion section, the FT-conversion. Optimisation of the facility could increase CO_2 capture rates to levels comparable of dedicated static IG-PG facilities. But this could decrease flexibility. Especially when only one chemical production is considered, CO_2 capture rates could be improved.

When combining biomass with CCS, CO_2-neutral production is achieved at biomass fraction around 40% on an energy basis. Higher biomass fractions can result in net negative CO_2 emissions.

Topics for further research are the effect of prolonged part-load operation on component availability and efficiency. It should also be noted that the results given in this study are based on computer simulations. As no flexible IG-PG facilities exist, the results cannot be compared with experimental values. Finally, the possibilities and effects of co-feeding large fractions of biomass (up to 100%) have not been tested in EF gasifiers.

Flexibility is probably more important for chemical-electricity substitution than for chemical-chemical substitution. The economics of flexible IG-PG facilities will be investigated in part B.

2.9 Acknowledgements

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2.10 Supporting information

Table 2.30 Energy balance of static IG-PG facility producing electricity.

		Unit	TOPS					Coal	EP				
Electricity	**Biomass fraction**		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass input	**MW_{HHV}**	**1696**	**1404**	**1091**	**755**	**392**		**378**	**705**	**990**	**1241**	**1464**
		MW_{LHV}	1600	1325	1029	712	370		353	659	925	1160	1368
	Coal input	**MW_{HHV}**		**351**	**727**	**1132**	**1569**	**2000**	**1513**	**1057**	**660**	**310**	
		MW_{LHV}		335	694	1080	1497	1908	1443	1009	630	296	
	Internal Power use												
	Feeding system	MW_e	-25	-26	-27	-28	-29	-30	-28	-26	-25	-23	-22
	ASU	MW_e	-34	-35	-37	-38	-40	-42	-40	-38	-36	-35	-34
	O_2 compressor	MW_e	-20	-21	-22	-23	-24	-25	-24	-23	-22	-21	-20
	Rectisol system	MW_e	-12	-12	-12	-12	-12	-12	-12	-12	-12	-11	-11
	CO_2 Compressor	MW_e	-26	-27	-27	-27	-27	-27	-27	-26	-26	-26	-25
	N_2 Compressor	MW_e	-28	-29	-31	-32	-34	-35	-32	-29	-26	-24	-22
	Steam cycle pumps	MW_e	-1.2	-1.3	-1.3	-1.4	-1.4	-1.5	-1.4	-1.3	-1.2	-1.1	-1.0
	Total internal power use	**MW_e**	**-146**	**-151**	**-156**	**-162**	**-168**	**-172**	**-164**	**-155**	**-148**	**-141**	**-135**
	Production												
	Gas Turbine production	MW_e	485	508	533	560	590	604	559	507	461	420	384
	Steam Turbines production	MW_e	329	337	345	353	363	368	354	338	324	311	301
	Gross power production	MW_e	815	845	878	914	952	971	913	844	784	732	685
	Net power production	**MW_e**	**668**	**694**	**722**	**752**	**785**	**800**	**749**	**689**	**637**	**591**	**550**
	Net Power production eff.	$\%_{HHV}$	39%	40%	40%	40%	40%	40%	40%	39%	39%	38%	38%
		$\%_{LHV}$	42%	42%	42%	42%	42%	42%	42%	41%	41%	41%	40%
	Total efficiency	**$\%_{HHV}$**	**39%**	**40%**	**40%**	**40%**	**40%**	**40%**	**40%**	**39%**	**39%**	**38%**	**38%**
		$\%_{LHV}$	42%	42%	42%	42%	42%	42%	42%	41%	41%	41%	40%

Table 2.31 Mass balance of static IG-PG facility producing electricity.

		Unit	TOPS					Coal	EP				
Electricity	Biomass fraction		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass	**kg/h**	**283k**	**234k**	**182k**	**126k**	**65k**		**71k**	**132k**	**186k**	**233k**	**274k**
	Coal	**kg/h**		**41k**	**86k**	**133k**	**185k**	**236k**	**178k**	**125k**	**78k**	**37k**	
	Gasifier												
	Oxygen Gasifier	kg/h	168k	175k	183k	191k	200k	208k	199k	190k	181k	174k	168k
	Slag produced	kg/h	4k	8k	12k	16k	21k	26k	20k	14k	9k	5k	1k
	Raw syngas produced	kg/h	452k	448k	444k	439k	434k	431k	440k	449k	457k	464k	471k
	Unconverted C	kg/h	762	779	798	818	840	846	815	774	738	707	679
	Rectisol												
	Sweet syngas	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	151k	139k
	CO_2 removed	kg CO_2/h	337k	339k	342k	344k	347k	345k	343k	337k	331k	326k	322k
	Acid gas	kg/h	1,255	2,598	4,039	5,589	7,265	8,900	7,004	5,223	3,670	2,304	1,091
	Sulphur	kg/h	58	1,215	2,455	3,789	5,228	6,647	5,049	3,554	2,249	1,101	82
	Balanced syngas distribution												
	Power section	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	152k	139k
	Power section												
	Syngas	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	152k	139k
	Nitrogen	kg/h	223k	233k	244k	256k	270k	277k	255k	231k	210k	191k	174k
	Air	kg/h	3371k	3533k	3706k	3892k	4096k	4192k	3887k	3521k	3201k	2919k	2669k
	Flue gas	kg/h	3766k	3947k	4141k	4349k	4577k	4685k	4343k	3934k	3576k	3262k	2982k
	CO_2 emitted	kg CO_2/h	219k	229k	240k	253k	266k	272k	252k	228k	208k	189k	173k

Table 2.32 Energy balance of static IG-PG facility producing FT-Liquids.

FT-Liquids		**Unit**	**TOPS**					**Coal**	**EP**				
	Biomass fraction		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass input	**MW_{HHV}**	**1696**	**1404**	**1091**	**755**	**392**		**378**	**705**	**990**	**1241**	**1464**
		MW_{LHV}	1600	1325	1029	712	370		353	659	925	1160	1368
	Coal input	**MW_{HHV}**		**351**	**727**	**1132**	**1569**	**2000**	**1513**	**1057**	**660**	**310**	
		MW_{LHV}		335	694	1080	1497	1908	1443	1009	630	296	
	Internal Power use												
	Feeding system	MW_e	-25	-26	-27	-28	-29	-30	-28	-26	-25	-23	-22
	ASU	MW_e	-34	-35	-37	-38	-40	-42	-40	-38	-36	-35	-34
	O_2 compressor	MW_e	-20	-21	-22	-23	-24	-25	-24	-23	-22	-21	-20
	Rectisol system	MW_e	-12	-12	-12	-12	-12	-12	-12	-12	-12	-11	-11
	CO_2 Compressor	MW_e	-26	-27	-27	-27	-27	-27	-27	-26	-26	-26	-25
	N_2 Compressor	MW_e	-5	-5	-5	-5	-6	-6	-5	-5	-4	-4	-4
	Steam cycle pumps	MW_e	-0.8	-0.8	-0.9	-0.9	-0.9	-1.0	-0.9	-0.8	-0.8	-0.7	-0.7
	FT system	MW_e	-20	-21	-22	-23	-24	-25	-23	-21	-19	-17	-16
	Total internal power use	**MW_e**	**-143**	**-147**	**-152**	**-158**	**-163**	**-167**	**-160**	**-151**	**-144**	**-138**	**-132**
	Production												
	Gas Turbine production	MW_e	88	93	97	102	108	113	102	93	84	77	70
	Steam Turbines production	MW_e	246	249	253	257	262	264	258	251	245	239	235
	Gross power production	MW_e	334	342	351	360	370	377	360	343	329	316	305
	Net power production	**MW_e**	**192**	**195**	**198**	**202**	**206**	**210**	**200**	**192**	**185**	**178**	**173**
	FT-Gasoline	MW_{HHV}	118	124	130	136	144	147	137	124	113	103	94
		MW_{LHV}	109	114	120	126	133	136	126	115	104	95	87
	FT-Diesel	MW_{HHV}	681	713	748	785	825	841	784	710	645	588	537
		MW_{LHV}	633	663	695	730	767	782	728	660	600	547	499
	Total Chemical production	**MW_{HHV}**	**799**	**837**	**878**	**922**	**969**	**988**	**920**	**833**	**758**	**691**	**631**
		MW_{LHV}	742	777	815	856	900	918	855	774	704	642	587

	Net Power production eff.	$\%_{HHV}$	11%	11%	11%	11%	11%	10%	11%	11%	11%	11%	12%
		$\%_{LHV}$	12%	12%	12%	11%	11%	11%	11%	12%	12%	12%	13%
	Chemical production eff.	$\%_{HHV}$	47%	48%	48%	49%	49%	49%	49%	47%	46%	45%	43%
		$\%_{LHV}$	46%	47%	47%	48%	48%	48%	48%	46%	45%	44%	43%
	Total efficiency	**$\%_{HHV}$**	**58%**	**59%**	**59%**	**60%**	**60%**	**60%**	**59%**	**58%**	**57%**	**56%**	**55%**
		$\%_{LHV}$	58%	59%	59%	59%	59%	59%	59%	58%	57%	56%	56%

Table 2.33 Mass balance of static IG-PG facility producing FT-Liquids.

		Unit	TOPS					Coal	EP				
FT-Liquids	Biomass fraction		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass	**kg/h**	**283k**	**234k**	**182k**	**126k**	**65k**		**71k**	**132k**	**186k**	**233k**	**274k**
	Coal	**kg/h**		**41k**	**86k**	**133k**	**185k**	**236k**	**178k**	**125k**	**78k**	**37k**	
	Gasifier												
	Oxygen Gasifier	kg/h	168k	175k	183k	191k	200k	208k	199k	190k	181k	174k	168k
	Slag produced	kg/h	4k	8k	12k	16k	21k	26k	20k	14k	9k	5k	1k
	Raw syngas produced	kg/h	452k	448k	444k	439k	434k	431k	440k	449k	457k	464k	471k
	Unconverted C	kg/h	762	779	798	818	840	846	815	774	738	707	679
	Rectisol												
	Sweet syngas	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	151k	139k
	CO_2 removed	kg CO_2/h	337k	339k	342k	344k	347k	345k	343k	337k	331k	326k	322k
	Acid gas	kg/h	1,255	2,598	4,039	5,589	7,265	8,900	7,004	5,223	3,670	2,304	1,091
	Sulphur	kg/h	58	1,215	2,455	3,789	5,228	6,647	5,049	3,554	2,249	1,101	82
	Balanced syngas distribution												
	Power section	kg/h	3k	3k	3k	3k	3k	3k	3k	3k	2k	2k	2k
	FT section	kg/h	171k	179k	188k	198k	209k	214k	199k	180k	164k	150k	137k
	H_2 distribution												
	Power section	kg/h	46	48	51	54	60	62	57	51	46	42	38
	FT section	kg/h	243	255	267	280	292	299	277	250	228	208	190

Power section												
Syngas	kg/h	3k	3k	3k	3k	3k	3k	3k	3k	2k	2k	2k
Offgas	kg/h	30k	32k	33k	36k	38k	40k	36k	33k	30k	28k	25k
Nitrogen	kg/h	38k	40k	41k	43k	46k	48k	43k	38k	35k	32k	29k
Air	kg/h	610k	639k	671k	706k	745k	774k	706k	639k	580k	529k	484k
Flue gas	kg/h	680k	713k	749k	788k	832k	865k	788k	713k	648k	591k	540k
CO_2 emitted	kg CO_2/h	31k	32k	34k	36k	38k	40k	36k	32k	30k	27k	25k
Chemical production												
FT-Light	kg/h	9,432	9,895	10,392	10,926	11,503	11,794	10,947	9,931	9,044	8,264	7,570
FT-Heavy	kg/h	51,390	53,843	56,470	59,289	62,302	63,495	59,158	53,580	48,698	44,395	40,571

Table 2.34 Energy balance of static IG-PG facility producing methanol.

		Unit	TOPS					Coal	EP				
Methanol	**Biomass fraction**		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass input	**MW_{HHV}**	**1696**	**1404**	**1091**	**755**	**392**		**378**	**705**	**990**	**1241**	**1464**
		MW_{LHV}	1600	1325	1029	712	370		353	659	925	1160	1368
	Coal input	**MW_{HHV}**		**351**	**727**	**1132**	**1569**	**2000**	**1513**	**1057**	**660**	**310**	
		MW_{LHV}		335	694	1080	1497	1908	1443	1009	630	296	
	Internal Power use												
	Feeding system	MW_e	-25	-26	-27	-28	-29	-30	-28	-26	-25	-23	-22
	ASU	MW_e	-34	-35	-37	-38	-40	-42	-40	-38	-36	-35	-34
	O_2 compressor	MW_e	-20	-21	-22	-23	-24	-25	-24	-23	-22	-21	-20
	Rectisol system	MW_e	-12	-12	-12	-12	-12	-12	-12	-12	-12	-11	-11
	CO_2 Compressor	MW_e	-26	-27	-27	-27	-27	-27	-27	-26	-26	-26	-25
	N_2 Compressor	MW_e	-5	-6	-6	-6	-7	-7	-6	-6	-5	-5	-4
	Air Gas turbine compressors	MW_e	-250	-262	-275	-289	-304	-313	-289	-261	-238	-217	-198
	Steam cycle pumps	MW_e	-1.0	-1.0	-1.0	-1.1	-1.1	-1.2	-1.1	-1.0	-0.9	-0.9	-0.8
	Total internal power use	**MW_e**	**-140**	**-144**	**-149**	**-154**	**-160**	**-163**	**-156**	**-148**	**-141**	**-135**	**-130**
	Production												
	Gas Turbine production	MW_e	234	245	257	270	285	293	270	244	222	203	185
	Steam Turbines production	MW_e	263	267	272	277	282	285	277	268	261	254	248
	Gross power production	MW_e	496	512	529	547	567	578	547	513	483	457	433
	Net power production	**MW_e**	**357**	**368**	**380**	**393**	**407**	**415**	**391**	**365**	**342**	**321**	**304**
	Methanol	MW_{HHV}	529	554	580	609	639	651	607	551	502	459	420
		MW_{LHV}	465	487	510	535	561	572	534	484	441	403	369
	Total Chemical production	**MW_{HHV}**	**529**	**554**	**580**	**609**	**639**	**651**	**607**	**551**	**502**	**459**	**420**
		MW_{LHV}	465	487	510	535	561	572	534	484	441	403	369

Net Power production eff.	$\%_{HHV}$	21%	21%	21%	21%	21%	21%	21%	21%	21%	21%	21%
	$\%_{LHV}$	22%	22%	22%	22%	22%	22%	22%	22%	22%	22%	22%
Chemical production eff.	$\%_{HHV}$	31%	32%	32%	32%	33%	33%	32%	31%	30%	30%	29%
	$\%_{LHV}$	29%	29%	30%	30%	30%	30%	30%	29%	28%	28%	27%
Total efficiency	**$\%_{HHV}$**	**52%**	**53%**	**53%**	**53%**	**53%**	**53%**	**53%**	**52%**	**51%**	**50%**	**49%**
	$\%_{LHV}$	51%	51%	52%	52%	52%	52%	51%	51%	50%	50%	49%

Table 2.35 Mass balance of static IG-PG facility producing methanol.

		Unit	TOPS					Coal	EP				
Methanol	Biomass fraction		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass	**kg/h**	**283k**	**234k**	**182k**	**126k**	**65k**		**71k**	**132k**	**186k**	**233k**	**274k**
	Coal	**kg/h**		**41k**	**86k**	**133k**	**185k**	**236k**	**178k**	**125k**	**78k**	**37k**	
	Gasifier												
	Oxygen Gasifier	kg/h	168k	175k	183k	191k	200k	208k	199k	190k	181k	174k	168k
	Slag produced	kg/h	4k	8k	12k	16k	21k	26k	20k	14k	9k	5k	1k
	Raw syngas produced	kg/h	452k	448k	444k	439k	434k	431k	440k	449k	457k	464k	471k
	Unconverted C	kg/h	762	779	798	818	840	846	815	774	738	707	679
	Rectisol												
	Sweet syngas	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	151k	139k
	CO_2 removed	kg CO_2/h	337k	339k	342k	344k	347k	345k	343k	337k	331k	326k	322k
	Acid gas	kg/h	1,255	2,598	4,039	5,589	7,265	8,900	7,004	5,223	3,670	2,304	1,091
	Sulphur	kg/h	58	1,215	2,455	3,789	5,228	6,647	5,049	3,554	2,249	1,101	82
	Balanced syngas distribution												
	Power section	kg/h	83k	87k	91k	96k	101k	104k	96k	87k	79k	72k	66k
	MeOH section	kg/h	126k	132k	139k	146k	154k	158k	146k	133k	121k	110k	101k
	H_2 distribution												
	Power section	kg/h	108	113	119	125	143	201	125	113	99	90	82
	MeOH section	kg/h	9,386	9,837	10,321	10,841	11,397	11,626	10,825	9,802	8,914	8,128	7,430

Power section												
Syngas	kg/h	83k	87k	91k	96k	101k	104k	96k	87k	79k	72k	66k
Offgas	kg/h	38k	40k	42k	45k	47k	49k	45k	41k	37k	34k	31k
Nitrogen	kg/h	44k	46k	48k	50k	52k	55k	49k	44k	40k	37k	33k
Air	kg/h	1604k	1682k	1766k	1856k	1955k	2011k	1854k	1679k	1526k	1392k	1273k
Flue gas	kg/h	1769k	1855k	1947k	2046k	2155k	2219k	2045k	1852k	1683k	1535k	1404k
CO_2 emitted	kg CO_2/h	99k	103k	108k	114k	120k	124k	114k	103k	94k	85k	78k
Chemical production												
MeOH	kg/h	84k	88k	92k	97k	102k	103k	96k	88k	80k	73k	67k

Table 2.36 Energy balance of static IG-PG facility producing urea.

		Unit	**TOPS**					**Coal**	**EP**				
Urea	**Biomass fraction**		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass input	**MW_{HHV}**	**1696**	**1404**	**1091**	**755**	**392**		**378**	**705**	**990**	**1241**	**1464**
		MW_{LHV}	1600	1325	1029	712	370		353	659	925	1160	1368
	Coal input	**MW_{HHV}**		**351**	**727**	**1132**	**1569**	**2000**	**1513**	**1057**	**660**	**310**	
		MW_{LHV}		335	694	1080	1497	1908	1443	1009	630	296	
	Internal Power use												
	Feeding system	MW_e	-25	-26	-27	-28	-29	-30	-28	-26	-25	-23	-22
	ASU	MW_e	-34	-35	-37	-38	-40	-42	-40	-38	-36	-35	-34
	O_2 compressor	MW_e	-20	-21	-22	-23	-24	-25	-24	-23	-22	-21	-20
	Rectisol system	MW_e	-12	-12	-12	-12	-12	-12	-12	-12	-12	-11	-11
	CO_2 Compressor	MW_e	-26	-27	-27	-27	-27	-27	-27	-26	-26	-26	-25
	N_2 Compressor	MW_e	0	0	0	0	0	0	0	0	0	0	0
	Air Gas turbine compressors	MW_e	-267	-280	-294	-309	-325	-333	-308	-279	-254	-232	-212
	Steam cycle pumps	MW_e	-1.0	-1.1	-1.1	-1.2	-1.2	-1.3	-1.2	-1.1	-1.0	-0.9	-0.9
	Total internal power use	**MW_e**	**-156**	**-162**	**-167**	**-173**	**-180**	**-184**	**-176**	**-166**	**-157**	**-150**	**-143**
	Production												
	Gas Turbine production	MW_e	251	263	276	290	305	313	290	263	239	218	199
	Steam Turbines production	MW_e	278	282	288	293	300	303	294	284	275	267	260
	Gross power production	MW_e	529	546	564	584	605	615	584	546	513	484	459
	Net power production	**MW_e**	**372**	**384**	**397**	**410**	**425**	**431**	**408**	**380**	**356**	**335**	**316**
	Urea	MW_{HHV}	468	491	515	541	569	583	540	489	444	405	370
		MW_{LHV}	457	479	503	528	556	569	527	477	434	396	362
	Total Chemical production	**MW_{HHV}**	**468**	**491**	**515**	**541**	**569**	**583**	**540**	**489**	**444**	**405**	**370**
		MW_{LHV}	457	479	503	528	556	569	527	477	434	396	362

	Net Power production eff.	$\%_{HHV}$	22%	22%	22%	22%	22%	22%	22%	22%	22%	22%	22%
		$\%_{LHV}$	23%	23%	23%	23%	23%	23%	23%	23%	23%	23%	23%
	Chemical production eff.	$\%_{HHV}$	28%	28%	28%	29%	29%	29%	29%	28%	27%	26%	25%
		$\%_{LHV}$	29%	29%	29%	29%	30%	30%	29%	29%	28%	27%	26%
	Total efficiency	**$\%_{HHV}$**	**50%**	**50%**	**50%**	**50%**	**51%**	**51%**	**50%**	**49%**	**49%**	**48%**	**47%**
		$\%_{LHV}$	52%	52%	52%	52%	53%	52%	52%	51%	51%	50%	50%

Table 2.37 Mass balance of static IG-PG facility producing urea.

		Unit	**TOPS**					**Coal**	**EP**				
Urea	Biomass fraction		100%	80%	60%	40%	20%	0%	20%	40%	60%	80%	100%
	Biomass	**kg/h**	**283k**	**234k**	**182k**	**126k**	**65k**		**71k**	**132k**	**186k**	**233k**	**274k**
	Coal	**kg/h**		**41k**	**86k**	**133k**	**185k**	**236k**	**178k**	**125k**	**78k**	**37k**	
							Gasifier						
	Oxygen Gasifier	kg/h	168k	175k	183k	191k	200k	208k	199k	190k	181k	174k	168k
	Slag produced	kg/h	4k	8k	12k	16k	21k	26k	20k	14k	9k	5k	1k
	Raw syngas produced	kg/h	452k	448k	444k	439k	434k	431k	440k	449k	457k	464k	471k
	Unconverted C	kg/h	762	779	798	818	840	846	815	774	738	707	679
							Rectisol						
	Sweet syngas	kg/h	173k	181k	190k	200k	211k	216k	201k	182k	166k	151k	139k
	CO_2 removed	kg CO_2/h	337k	339k	342k	344k	347k	345k	343k	337k	331k	326k	322k
	Acid gas	kg/h	1,255	2,598	4,039	5,589	7,265	8,900	7,004	5,223	3,670	2,304	1,091
	Sulphur	kg/h	58	1,215	2,455	3,789	5,228	6,647	5,049	3,554	2,249	1,101	82
							Balanced syngas distribution						
	Power section	kg/h	220k	231k	242k	255k	269k	275k	255k	232k	211k	192k	176k
							H_2 distribution						
	Urea section	kg/h	17k	18k	18k	19k	20k	21k	19k	18k	16k	15k	13k

Power section												
Syngas	kg/h	220k	231k	242k	255k	269k	275k	255k	232k	211k	192k	176k
Offgas	kg/h	4k	4k	4k	4k	5k	5k	4k	4k	4k	3k	3k
Nitrogen	kg/h	3k	3k	3k	3k	3k	3k	2k	2k	1k	1k	1k
Air	kg/h	1717k	1799k	1887k	1983k	2086k	2135k	1980k	1794k	1631k	1487k	1360k
Flue gas	kg/h	1943k	2037k	2137k	2245k	2361k	2418k	2242k	2031k	1846k	1684k	1540k
CO_2 emitted	kg CO_2/h	218k	228k	240k	252k	265k	271k	251k	228k	207k	189k	172k
Chemical production												
Urea	kg/h	160k	168k	176k	185k	195k	200k	185k	167k	152k	139k	127k

3 Performance of simulated flexible integrated gasification polygeneration facilities. Part B: Economic evaluation.

Hans Meerman, Andrea Ramírez, Wim Turkenburg & André Faaij
Renewable and Sustainable Energy Reviews. In press.

Abstract

This paper investigates the economics of integrated gasification polygeneration (IG-PG) facilities and assesses under which market conditions flexible facilities outperform static facilities. In this study, the facilities use Eucalyptus wood pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal as feedstock to produce electricity, FT-liquids, methanol and urea. All facilities incorporate CCS.

The findings show production costs from static IG-PG facilities ranging between 12-21 €/GJ using coal, 19-33 €/GJ using TOPS and 22-38 €/GJ using EP, which is above the average market prices. IG-PG facilities can become competitive if capital costs drop by 10-27% for coal based facilities. Biomass based facilities will need lower biomass pellet prices or higher CO_2 credit prices. Biomass becomes competitive with coal at a CO_2 credit price of 50-55 €/t CO_2.

Variations in feedstock, CO_2 credit and electricity prices can be offset by operating a feedstock flexible IG-PG facility, which can switch between coal and TOPS, thereby altering its electricity production. The additional investment is around 0.5% of the capital costs of a dedicated coal based IG-PG facility. At 30 €/t CO_2, TOPS will be the preferred feedstock for 95% of the time at a feedstock price of 5.7 €/GJ. At these conditions, FT-liquids (gasoline/diesel) can be produced for 15.8 €/GJ (116 $/bbl). Historic records show price variations between 5.7-7.3 €/GJ for biomass pellet, 1.0-5.6 €/GJ for coal and 0-32 €/t CO_2. Within these price ranges, coal is generally the preferred feedstock, but occasionally biomass is preferred. Lower biomass prices will increase the frequency of switching feedstock preference from coal to biomass, raising the desire for flexibility.

Of the three investigated chemicals, an IG-PG facility producing FT-liquids benefits the most from flexibility. Our study suggests that if the uncertainty in commodity prices is high, a small additional investment can make flexible IG-PG facilities attractive.

3.1 Introduction

A significant reduction of global CO_2 emissions will require decarbonisation of both the transport and power sector. State-of-the-art (SOTA) flexible integrated gasification poly-generation (IG-PG) facilities[8] equipped with CO_2 capture can potentially decarbonise both sectors and play a role in the development of a sustainable energy infrastructure. In the first place, they can produce CO_2 neutral transportation fuels. Secondly, they can act as back-up power plants. Thirdly, flexible IG-PG facilities can potentially offer attractive economics by responding to fluctuating market conditions, such as the daily variation in electricity price or the seasonal price variation in biomass pellet price.

Large scale IG-PG facilities are built to operate for several decades. During that time frame, market conditions can change considerably. This holds especially for the emerging markets of biomass pellets for energy,[60,61] CO_2 credits[62] and non-oil based transportation fuels[63] as well as the mature markets of coal[64] and electricity.[66] In addition, in the coming decades shifts are likely to occur in the production and use of main transportation fuels, where gasoline and diesel could be replaced by methanol, hydrogen or electricity. Besides long-term uncertainties, there are also short-term variations which could have an impact on the profitability of IG-PG facilities. One example is the daily variation of the electricity price.[66] These uncertainties may make flexible IG-PG facilities valuable in the (future) development of a sustainable energy infrastructure.

Although the economics of IG-PG facilities have been extensively analysed,[24,25,49,50,72,74,77,78,79,80,81,148] only a few studies have investigated the impact of flexibility on these facilities. Carapellucci et al.[149] for instance, investigated the technical and economic performance of flexible methanol/electricity production, while a study made by the IEA GHG investigated the technical and economic performance of flexible hydrogen/electricity production.[49,50] In a previous study,[150] we investigated the technical performance and limitations of various processes in SOTA flexible IG-PG facilities. The investigated facilities use commercially available technologies only. The previous study focuses on the performance of individual components as well as the entire facility when feedstock or product mix is altered. This includes de-rating, over-dimensioning of key components, like the gasifier, and minimal load constraints. Results show that, from a technical point of view, both feedstock and product flexibility are possible, although with certain limitations. For instance, the volume of the syngas exiting the gasifier remains constant, regardless of the used feedstock. Compared to coal, biomass produces more gas per energy input, resulting in a reduction in total energetic input when switching from coal to biomass. Over-dimensioning the gasifier is possible but increases capital costs. The minimal load of important process equipment - e.g., chemical synthesis reactors, distillation columns and gas turbines - was found to be 40%. Also, co-feeding of biomass

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[8] In IG-PG facilities solid feedstocks are converted into synthesis gas and subsequently into electricity, chemicals or fuels. These facilities are referred to as XtY systems. The X is substituted if a specific feedstock is used: biomass (BtY), torrefied biomass (TtY) or coal (CtY). The Y is substituted if a specific output is produced: electricity (XtP), FT-liquids (XtL), methanol (XtM) or urea (XtU).

has been maximised at 50% on an energy basis[9]. In addition, the results indicated that part-load operation due to feedstock substitution hardly affects overall facility efficiency.

Compared to static[10] facilities, flexible facilities can respond to and exploit fluctuating market conditions, expectedly resulting in overall lower feedstock costs and higher product revenues. However, flexible facilities also have higher capital investment and O&M costs and, depending on operation mode, some efficiency losses (see Meerman et al.[150]). Currently, it is not known whether the advantages of flexibility offset its disadvantages, despite that flexibility is argued to be a strong argument in favour of IG-PG facilities.

The goal of the present study was to determine if, and under which economic conditions, flexible design and operation could lead to favourable economic performance of SOTA IG-PG facilities. This analysis focused on the use of Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal as potential feedstocks. The feedstocks were converted into electricity, FT-liquids, methanol and urea. To reduce CO_2 emissions, pre-combustion CO_2 capture technology was included in all cases. In this study it was assumed that the facilities begin operation in 2015. Therefore, only commercially available technologies were considered.

The structure of this article is as follows: Section 2 describes the methodology, including the technical component-based AspenPlus model, the case studies where different aspects of flexibility are analysed, the scenarios which represent different economic climates, and the component capital costs. In section 3 the economic parameters are given. Section 4 contains the results. In section 5 the discussion as well as key conclusions are provided. In this study all units are in SI-units, heating values are expressed in higher heating value (HHV) and costs in $€_{2008}$, unless stated otherwise.

3.2 Methodology

The required technical data of the facilities, like component dimensions, efficiencies and mass and energy balances, were derived from the technical assessment done in part A of this analysis.[150] It was assumed that the IG-PG facilities use SOTA technology and are operated between 2015-2035. The point of departure for the economic data used in this study was a study made by FosterWheeler[49] published in 2007 (see §3.3.2 for more details). The economic data were obtained using information from expert interviews and open literature sources and updated to $€_{2008}$. By combining the capital costs with current commodity prices, first production costs of electricity, FT-fuels, methanol and urea, produced by static IG-PG facilities, were calculated. Next, these production costs were

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[9] This limitation results from the different ash and alkali composition and quantity in biomass compared to coal. At high biomass ratios, the different ash amount and composition causes, among other, reduction of the protective slag layer and fouling of heat exchangers.

[10] Static facilities have little to no ability to change their input or output. Feedstock must meet stringent specifications and production of fuels, chemicals or electricity cannot be altered.

compared with those of flexible IG-PG facilities. For this purpose, production costs of 6 different operating strategies (cases) in 5 different scenarios were assessed, taking into account capital costs data and - variation in - historical prices of coal, biomass pellet and CO_2 credits. A schematic overview of the methodology is displayed in Figure 3.1. In the rest of this section an overview of each step is presented.

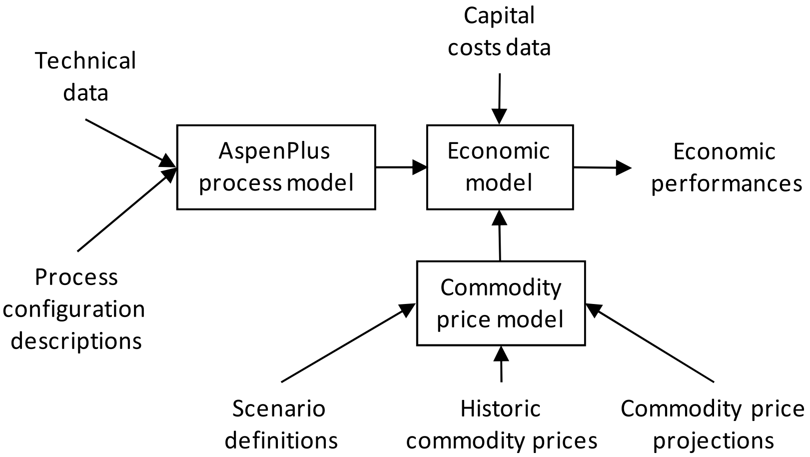

Figure 3.1 Schematic overview of the methodological steps included in this study.

3.2.1 Technical data and AspenPlus process model

When the feedstock or product mix of an IG-PG facility is altered, mass and energy flows and efficiencies throughout the different components change. The components in a flexible IG-PG facility (Figure 3.2) must be able to handle this and were therefore dimensioned to the largest possible required capacity. This, however, means that, at times, certain processes would operate under part-load conditions, resulting in de-rating and efficiency losses. These effects increase with decreasing load levels, up to a point where the component can no longer operate properly. A second effect is that the capital costs of components are higher than those of static facilities. To calculate the mass and energy flows and the required size of each component, a previous developed AspenPlus simulation model on a component level was used.[150] The most important assumptions and results of the model are given below. We refer to Meerman et al.,[150] for further detailed information.

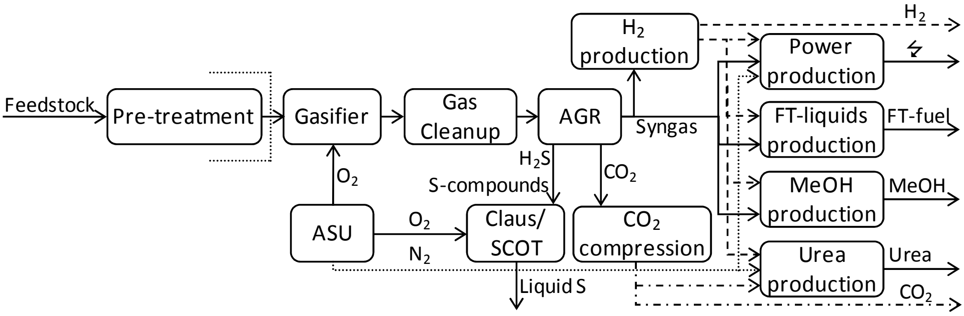

Figure 3.2 Simplified layout of a flexible IG-PG facility process. Waste, energy and heat streams are not displayed.[150]

Key characteristics of the model are:

- Inclusion of three different types of feedstock: Illinois #6 coal, Eucalyptus wood pellets (EP) and torrefied wood pellets (TOPS);
- The IG-PG facility is based on a 2000 MW_{th} coal input-equivalent Shell gasifier. This means that, based on the model results, thermal input for 100% TOPS is 1696 MW_{th} and for 100% EP only 1464 MW_{th}.[150] When biomass is co-fed, the thermal input lays between these two extremes;
- Three different feedstock ratios: 100% coal, 50/50% biomass/coal (on an energy basis) and 100% biomass; both for TOPS and EP. The 50 $energy_{HHV}$% biomass was considered the current technical maximum for co-feeding biomass in the gasifier.[150] Although 100% biomass input is currently technically not possible, this option was also assessed, to determine the effects on component size and overall plant efficiency and output;
- Four different outputs as main product: electricity, FT-liquids (gasoline/diesel), methanol and urea. When producing the chemicals or fuels as main product, electricity is produced as by-product. Not considered is the simultaneous production of two or more chemicals/fuels. Based on Meerman et al., the minimal load of the main processes, like the air separation unit, water-gas shift (WGS) reactor, chemical conversion reactors, gas and steam turbines, is 40% when considering production flexibility;[150]
- As all IG-PG facilities already separate CO_2 after gas cleaning, the only additional step for CO_2 capture is compression. However, since the H_2:CO ratio is kept constant at 2.3[11], only 56% (for coal) to 65% (for EP) of the total carbon is captured (see Figure 3.3). In this study all investigated IG-PG configurations were assumed to operate with carbon capture and storage (CCS).

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[11] This value was selected as it is the lowest H_2:CO ratio needed for the chemical section, resulting in less variation in the gas cleanup section.[150]

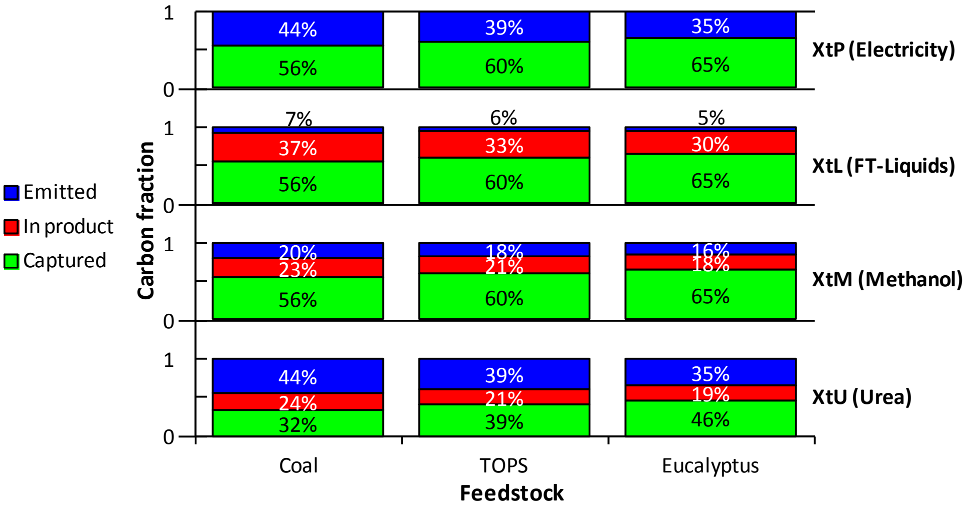

Figure 3.3 Calculated carbon fractions as captured, emitted and embedded in chemical/fuel output.[150]

Plant efficiencies of static IG-PG facilities following from the model analysed in Meerman et al.,[150] are shown in Table 3.1.

Table 3.1 Overall energy efficiencies (HHV) of static XtY facilities.

Energy efficiency		Coal	TOPS	EP
Electricity (XtP)	**Power**	**40%**	**39%**	**38%**
FT-Liquids (XtL)	FT	49%	47%	43%
	Power	10%	11%	12%
	Total	**60%**	**58%**	**55%**
Methanol (XtM)	MeOH	33%	31%	29%
	Power	21%	21%	21%
	Total	**53%**	**52%**	**49%**
Urea (XtU)	Urea	29%	28%	25%
	Power	22%	22%	22%
	Total	**51%**	**50%**	**47%**

3.2.2 Case studies

In this article 6 cases were studied to provide insight into the economic performance of flexible IG-PG facilities. In all cases the technical model, described in part A of this analysis, was used.[150]

Case 1. Static IG-PG facilities

This case study served as reference case. The static IG-PG facilities had fixed feedstock and production mixtures. All equipment was optimised to these characteristics. As a result, the facilities were almost non-flexible or would display strong de-rating impacts when operated at part-load conditions.

Case 2. Variation of feedstock: Capital costs penalty.

This case study served to investigate the increase in production costs when a feedstock flexible designed IG-PG facility is operated as if it is a static IG-PG facility by comparing the production costs with an IG-PG facility which is optimised for one feedstock only.

Case 3. Variation of feedstock: Constant chemical production and variable electricity production by feedstock substitution.

This case study was used to investigate under which economic conditions (feedstock and CO_2 credit prices) exploiting the fluctuations in electricity price by adjusting the feedstock biomass/coal ratio pays off. Using coal instead of biomass as feedstock leads to a larger energetic flow of the syngas[12].[150] As the chemical production was kept constant, an increase in syngas energy results in an increase in electricity production. This additional electricity was sold at spot market prices.[66] Under the right economic conditions, this can lead to the use of biomass during off-peak hours and the use of coal during peak hours. To determine whether coal or biomass is the most economic option, a feedstock equivalent price (FEP, see §3.2.5) was used.

Case 4. Variation of feedstock: Impact of short-term feedstock price variation.

This case study was used to investigate the impact of short-term feedstock price variation on the economics of chemical/mid-load[13] power plants. The main difference with case 3 is that the feedstock (TOPS and coal) prices were varied on a weekly basis to simulate a variable market. The NPV was compared with those of a static facility using only coal as feedstock.

Case 5. Variation of production: Producing mainly chemicals/fuels during off-peak hours and mainly electricity during peak hours.

This case study was used to investigate whether and under which economic conditions operating an IG-PG facility as a chemical/mid-load[13] power plant pays off. It was assumed that the IG-PG facilities used a fixed biomass/coal feedstock mixture. During peak hours electricity production was maximised and during off-peak hours chemical/fuel production was maximised. For both extremes a 40% minimal load restriction was maintained.[150] Operating the IG-PG facility this way ensured that the gasifier, syngas cleaning and acid gas removal (AGR) sections remained at nominal load and only the chemical/fuel and power sections varied in load. Production was switched from chemical/fuel to electricity when the electricity market price increased above the production equivalent price (PEP), see §3.2.5. The chemical/fuel (FT-liquids,

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[12] This is caused by the higher energy density of coal compared to biomass and the assumption that the gasifier has a fixed gas generation capacity. The thermal input of coal is $2000MW_{th}$. This drops to $1696MW_{th}$ for TOPS and $1464\ MW_{th}$ for EP.

[13] Mid-load or intermediate power plants are deployed after the main base (nuclear and coal) power plants and renewable (wind and solar) electricity generation capacity, but before the peak (gas) power plants. This roughly means that during working days mid-load power plants are operated at full capacity between 08:00u-20:00u. Outside this time-window the power plants are operating at considerably lower capacity.

methanol and urea) price was varied on a weekly basis to simulate fluctuations in product prices. The average NPV were compared to the NPV of a static facility with the same chemical/fuel as main product and operating in the same market.

Case 6. Retrofitting FT producing IG-PG facilities to methanol or H_2 producing IG-PG facilities.

This case study was used to evaluate the economic implications of retrofitting an IG-PG facility during the economic lifetime of the facility (2015-2045 with the retrofit in 2035). The preferred fuel mix for transportation may change in the coming decades from gasoline/diesel to methanol, hydrogen or electricity. Therefore, in this case study an IG-PG facility producing FT-fuels was retrofitted after 20 years to produce another transportation fuel (methanol) for another 10 years. The economics of this retrofit were compared to an IG-PG facility producing methanol for 30 years.

As indicated, in cases 3 till 5, the economic performance of IG-PG facilities under market conditions with fluctuating commodity prices was calculated. For this analysis, the economic lifetime of the facilities was divided into individual hours. For each hour, the most economic feedstock or production type was determined.

Prices were varied, depending on the case study. In case 3, only the electricity price was varied. Case 4 had fluctuating electricity and feedstock prices. The chemical/fuel prices behaved according to the scenarios. The analysis was performed using four different CO_2 credit prices, based on the results of case 3. In case 5, the chemical/fuel and electricity prices were fluctuating. The feedstock and CO_2 credit prices behaved according to the scenarios.

The method of variation was identical for all case studies. In cases 3 and 4, base load electricity production was sold for 15.7 €/GJ (Table 3.3). The additional electricity was sold according to the historical Dutch day-hourly market price between 2004-2008.[66] This data was repeated 5 times to obtain electricity prices spanning 20 years. In case 5, also the base load electricity production was sold for the day-hourly prices. Random values were selected for the feedstock, chemical/fuel or CO_2 price variations. These random values were based on the historical data using a Gauss distribution, have a minimum of zero and were altered every 168 hours. The analysis was repeated 200 times, resulting in the average economic performance of IG-PG facilities under short-term market variations. Case 3 was performed for the *Current Situation* scenario only, while cases 4 and 5 were performed for all scenarios.

3.2.3 Scenarios

Five scenarios were used to explore the impact of possible future global trends in primary energy carrier prices and policy developments, e.g., CO_2 price. The scenarios contain different price ranges and dynamics for biomass pellet, coal, CO_2 credit and CO_2 transport and storage costs. The values used were based on historical market prices and projections of the Netherlands Environmental Assessment Agency (PBL),[151,152] European Commission,[43] IEA[69] and studies commissioned by Greenpeace.[153,154] The commodity price model is discussed in more detail in §3.3.1.

Scenario 1. Current Situation scenario

In this scenario current production costs were calculated using the commodity price levels at the beginning of 2010, which were 6.3 €/GJ for biomass pellets,[1,2] 1.5 €/GJ for coal,[64] 10.1 €/GJ for gasoline/diesel,[65] 11.0 €/GJ for methanol[63] and 19.0 €/GJ for urea.[155] See §3.3.1 for the justification.

Scenario 2. Business as Usual scenario (B as U)

This scenario assumed modest CO_2 mitigation policies, resulting in a slight increase in CO_2 credits price. It was assumed that the limited incentive for GHG reduction will result in a limited substitution of coal by biomass, leading to a minor increase in biomass pellet prices and a minor decrease in coal prices.[69,151,152] Without a strong incentive for CCS, no large-scale CO_2 transport infrastructure will be build, limiting the CO_2 transport network to smaller single source-single sink pipelines. Therefore, CO_2 transport and storage costs remained constant at 15.0 €/t CO_2 for the whole time period (2015-2035).[156]

Scenario 3. Direct Action scenario (DA)

This scenario assumed stringent CO_2 policies, resulting in a strong increase in the price of CO_2 credits. This leads to a gradual transition from a fossil-based energy infrastructure towards a renewable-based energy infrastructure. The rate at which biomass pellet prices increased and coal prices decreased were higher compared to the *Business as Usual* scenario. The higher CO_2 credit price will result in the development of a large-scale CO_2 transportation network and, consequently, lower CO_2 transport and storage costs.

Scenario 4. Green scenario

This scenario assumed much higher coal prices and much lower biomass pellet prices compared to the previous scenarios. Literature studies show that biomass pellet prices could drop to 3.0 €/GJ due to technical learning, e.g., improved cultivation techniques, mature market and efficient local pre-treatment.[32,85,157,158] In 2008, coal prices exceeded 5.0 €/GJ in NW-Europe.[1] In this scenario it was assumed that biomass pellet prices drop gradually to 3.0 €/GJ, while coal prices rise gradually to 3.0 €/GJ. One option to reduce net CO_2 emissions is to replace coal by (sustainable) biomass. The CO_2 mitigation penalty of this option consists mainly of the price difference between coal and biomass. Due to the converging of biomass and coal prices, reducing net CO_2 emissions using this option becomes less expensive. Therefore, in this scenario it was assumed that CO_2 credit prices remained constant. Both the CO_2 credit price and CO_2 transport and storage costs were based on the *Current Situation* scenario.

Scenario 5. Delayed Climate Policy scenario (DCP)

This scenario assumed that a trigger point occurred in 2025. Up to 2025 limited CO_2 policies were assumed. After 2025, very stringent CO_2 policies are implemented. Dependency of fossil fuels will remain high up to the trigger point. The stringent CO_2 policies will result in a rapid transition from a fossil-based energy infrastructure towards a sustainability-based energy infrastructure. Up to 2025, feedstock and CO_2 credit prices behaved as in the *Business as Usual* scenario. After 2025 the rate at which commodity prices and CO_2 credits changed was doubled. Starting 2025, a large-scale CO_2 transport network is rapidly deployed, resulting in a reduction of CO_2 transport and storage costs.

3.2.4 Total Capital Investment

Total capital investments (TCI) were calculated using the Factored Estimation method. In this method, the cost of each major component is estimated using data from open literature sources and expert interviews. These cost data correspond to a specific scale and is adjusted to the required scale. An installation factor, representing e.g., infrastructure, overhead, engineering and contingencies, is added to the scaled component capital costs. The TCI is the summation of the scaled component costs, including installation factor, of the individual components. The used component capital costs, scaling factors and installation factor are given in §3.3.2.

The Factored Estimation method has an inherent uncertainty of approximately 30%.[159] Besides the inherent uncertainty, the method has two limitations. One limitation is the volatility of the equipment prices. In the last years, capital costs varied significantly, increasing rapidly from 2005, peaking late 2008 and decreasing thereafter.[160] A study made by CERA on the capital costs of US- and EU-based power plants, excluding nuclear, shows a steady increase from 2000 onwards. Capital costs peaked in the EU beginning 2008 (index = 180, base year is 2000), while in the U.S. this peak occurred beginning 2009 (index = 188, base year is 2000). Since that peak, capital costs dropped by 12% (EU) and 7% (US).[161] Early 2010, prices were back at mid 2006 (EU) or mid 2007 (US) level (Figure 3.4). This volatility resulted in benchmarking studies requiring updating in relatively short time spans. For instance, FosterWheeler (FW) updated its analysis of the cost effectiveness of IGCC-CCS published in July 2007[49] already in August 2008.[50] The updated study has an increase in capital costs of 20%.

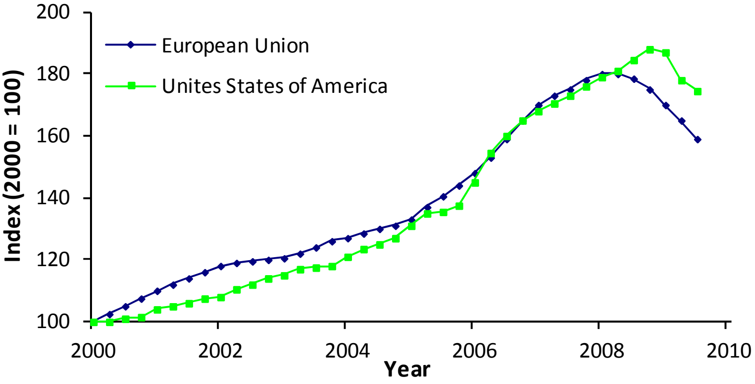

Figure 3.4 Relative capital costs for non-nuclear power plants.[161]

The second limitation is the uncertainty of the component capital costs. This is highlighted by comparing the FW studies[49,50] with a study performed by NETL[74]. Both studies calculate the production costs of a coal-fired IGCC, albeit with slightly different base year, scale and technology. After correcting for those factors, specific capital costs in the NETL study are still 24% higher than those in the FW study.

Capital investment for a given component depends on its scale. The required scale was provided by the technical model (see §3.2.1). To scale the capital investment from a known scale and cost, equation 3-1 was used. If the required scale exceeds the maximum scale of a component, multiple identical components in parallel are needed. Installing identical components in parallel gives a small cost reduction.[25] This reduction was calculated using equation 3-2.

$$y_s = y_0 * \left(\frac{x_s}{x_0} \right)^S \tag{3-1}$$

$$y = y_s * n^m \tag{3-2}$$

where y, y_s and y_0 = actual, scaled and initial costs respectively
x_s and x_0 = actual and initial scale respectively
S = scaling factor
n = desired number of units
m = multiplication exponent (= 0.9)

3.2.5 Economic model

The economics of the different case studies and scenarios were evaluated by calculating the production costs of the main product using the Net Present Value (NPV) method,[162] see equation 3-3. By setting the NPV to zero the production cost of the main product was calculated. Note that in the economic analysis only stored CO_2 was exempted from CO_2 credits. Carbon locked in chemical/fuel products were counted as if it was emitted.

$$NPV = -I + \sum_{i=o}^{L} \frac{B_i - C_i}{(1+r)^i} \tag{3-3}$$

where B and C = annual revenues (benefits) and costs respectively
r = discount rate
L = plant economic lifetime (years)
I = total capital investment cost
i = year

Common parameters used in all case studies and scenarios are displayed in Table 3.2. Based on other studies, a construction time of three years was assumed and capital costs were evenly divided over these years.[24,50,74,79]

Table 3.2 Economic assumptions IG-PG facilities.

Parameter	Unit	Value
Location	-	NW-Europe
Construction time	Year	3
Plant economic lifetime (L)	Year	20
Discount rate (r)	%	10
Operating fraction	%	90
O&M costs	% of cap. cost	4

In Cases 3-5 a feedstock equivalent price (FEP) price and a production equivalent price (PEP) were calculated to determine the break-even point in feedstock and production price respectively. FEP is the electricity price at which the feedstock should be switched from biomass to coal. Note that the production of chemicals is kept constant; the additional energy is used for electricity production only. PEP is the electricity price at which production should be switched from chemical to electricity. The FEP and PEP were calculated according to the following equations:

$$FEP=\frac{\left(\Delta E_{Bio}*P_{Bio}+\Delta E_{Coal}*P_{Coal}\right)-\left(\Delta E_{Slag}*P_{Slag}+\Delta E_{Sulphur}*P_{Sulphur}\right)+\Delta E_{CO_2}*P_{CO_2}}{\Delta E_{Electricity}} \quad (3\text{-}4)$$

$$PEP=\frac{\Delta E_{Chemical}*P_{Chemical}-\Delta E_{CO_2}*P_{CO_2}}{-\Delta E_{Electricity}} \quad (3\text{-}5)$$

where FEP = feedstock equivalent price (€/GJ)
PEP = production equivalent price (€/GJ)
ΔE_x = difference in input or output of commodity x (energy flow (GJ/yr) for feedstock, main product and electricity or mass flow (kt/yr) for by-products and CO_2)
P_x = price of commodity x (€/GJ for feedstock, main product and electricity or €/kt for by-products and CO_2)

3.3 Economic Data

3.3.1 Commodity prices

To estimate the feedstock and CO_2 credit prices during the lifespan of the facilities (2015-2035) for each scenario, historical data were combined with price projections from other studies (see supplementary data). For coal and CO_2 credits the WEO[69] projections were used. For biomass pellets, projections from the PBL[152] were used as WEO data were missing. Prices used are displayed in Figures 3.5-3.7. For the end-products no future projections were made. Instead, the prices as of beginning 2010 as well as the minimum and maximum values of the historical prices were used (Figures 3.8-3.10).

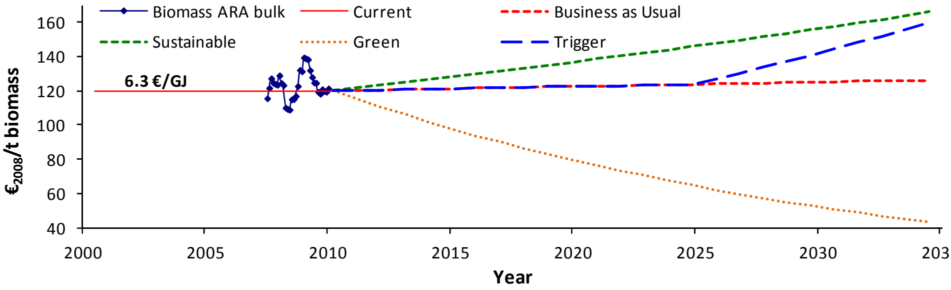

Figure 3.5 Historical (left) and projected (right) bulk biomass pellet prices for the different scenarios.[60,61]

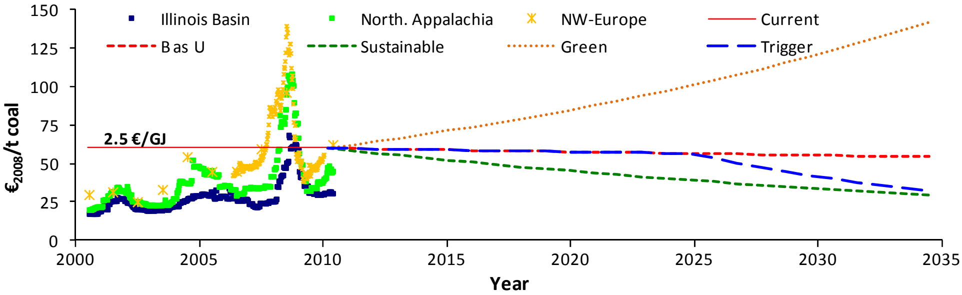

Figure 3.6 Historical (left) and projected (right) coal prices for the different scenarios. The historical coal prices were on a weekly basis. The NW-Europe prices prior to May 2006 were on a yearly basis.[1,64]

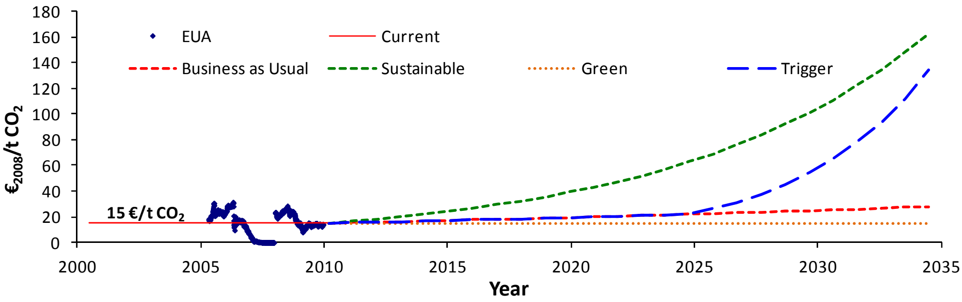

Figure 3.7 Historical (left) and projected (right) European Union Allowance (CO_2 emission rights) spot prices for the different scenarios.[62]

Biomass pellet prices are volatile. The market is relatively small and immature. Therefore, a moderate change in demand can lead to a large variation in price. It is expected that, as the biomass pellet market grows, price spikes will become rarer and less severe. The small and immature market makes long term projections of biomass pellet prices uncertain and significant differences between future projections have been found. As already indicated in §3.2.3, biomass pellet prices were expected to drop due to technical learning in the long term.[32,85,157]

Coal prices have been relatively stable between 2000 and 2007. In 2007 prices rose sharp due to a large increase in demand. Prices fell again due to the economic crisis. It is expected that prices will rise significantly again when the global economy recovers. NW-Europe coal is delivered at ARA (Amsterdam-Rotterdam-Antwerp) and is, on average, 1.5 times more expansive than Northern Appalachia coal. Contrary to biomass pellets, where technical learning could reduce prices considerable, coal production is already well developed. Therefore, no strong reductions in production costs due to learning are expected. At the same time, strong increases in production costs due to depletion of (cheap) mining sites are expected. It is, therefore, likely that coal prices will increase in the long term. When this will happen is outside the scope of this study. The impact of increasing coal prices was investigated in the *Green* scenario.

The CO_2 credit price collapsed during the height of the economic recession. Since then prices have stabilised at 15 €/t CO_2. Mulder et al., analysed the CO_2 credit market and concludes that CO_2 credit prices will remain around this price due to an abundance of CO_2 emission rights. However, the CO_2 credit market is an artificial market. The number of available CO_2 emission rights emitted by the European Union can directly influence CO_2 credit prices.[163]

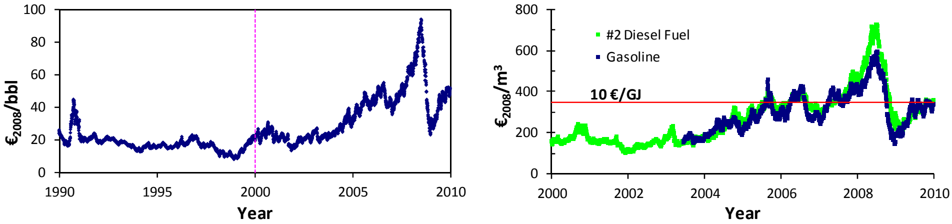

Figure 3.8 Historical Brent crude oil (left) and oil products (right) prices, excluding duties and taxes.[65]

The IEA World Energy Outlook 2010 current policy scenario assumes that oil prices will more than double between 2009-2025 and then start to stabilise at almost 100 $€_{2008}$/bbl (135 $\$_{2009}$/bbl) in 2035, exceeding the 2008-spike. However, forecasting the oil price is difficult as uncertain factors, e.g., geo-political stability and world economy, have a direct influence on the oil price.

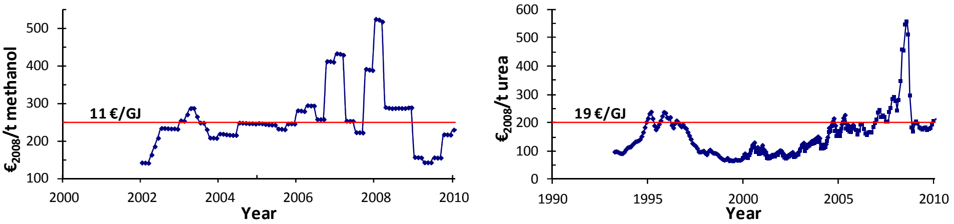

Figure 3.9 Historical methanol[63] (left) and urea[155] (right) prices.

Methanol prices were relatively stable at 200 €/t between 2002-2006. The past several years saw some large price spikes. These spikes were caused both by an increase in demand and a shutdown of production facilities due to maintenance.[63] Since early 2010 methanol prices continued to rise to over 300 €/t, while the market remained volatile due to a small buffer between supply and demand.

For the past decades urea prices were fluctuating between 50-200 €/t. In the last few years urea prices spiked to over 500 €/t. The World Bank expects that up till 2020 urea prices will fluctuate around 175 €/t.[155] The International Fertilizer Industry Association indicates that in the coming years production capacity will increase more rapidly than supply, resulting in more stable and possibly lower prices.[164]

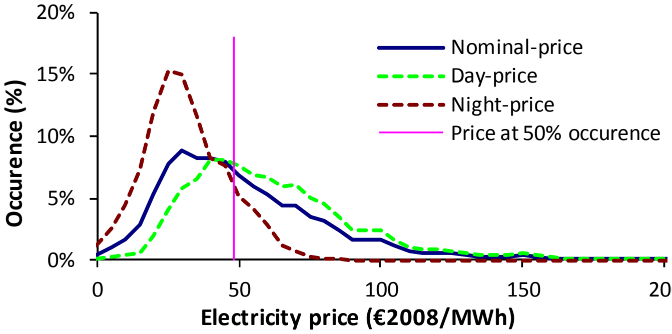

Figure 3.10 Electricity price distribution in the Netherlands during 2004-2008.[66] The vertical line in the graph marks half of the occurrences.

Electricity prices are fluctuating constantly. In NW-Europe, in general a distinction can be made between day (peak) and night (off-peak) prices. As can be seen in Figure 3.10, peak prices are significantly higher. Future electricity prices will depend on feedstock prices, status of the economy, penetration rate of renewable as well as other energy sources, technological developments, political decisions and weather conditions. As an example, the announced shutdown of German nuclear power plants can influence the production costs of electricity in NW-Europe.

3.3.1.1 Commodity price projections

Price projections used in this study are summarized in Table 3.3. Note that the product prices from early 2010 were taken as reference.

Table 3.3 Commodity prices.

			Scenarios					
	Parameter	Unit	Current	Business as Usual	Sustainable	Green	Trigger	
							2015-2025	2025-2035
Feedstock Prices	EP	€/GJ	6.3 (1)	+ 0.2 %/yr	+ 1.3 %/yr	-3.2%/yr	+ 0.2 %/yr	+ 2.6 %/yr
	TOPS	€/GJ	6.3 (1)	+ 0.2 %/yr	+ 1.3 %/yr	-3.2%/yr	+ 0.2 %/yr	+ 2.6 %/yr
	Coal	€/GJ	2.25 (2)	- 0.4 %/yr	- 2.8 %/yr	+2.8%/yr	- 0.4 %/yr	- 5.6 %/yr
CO_2	CO_2 credits	€/t CO_2	15.0 (3)	+ 2.5 %/yr	+ 10 %/yr	15.0	+ 2.5 %/yr	+ 20 %/yr
	CO_2 transport & storage costs	€/t CO_2	10.0 (4)	15.0 (4)	10.0-15.0 (5)	10.0	15.0 (5)	10.0 (5)

Reference Product Prices			
Current product prices	FT-fuel	€/GJ	10.1 (3-21) (6)
	Methanol	€/GJ	11.0 (6-23) (7)
	Urea	€/GJ	19.0 (6-53) (8)
	Electricity	€/GJ	15.7 (0-290) (9)
	Slag	€/t	0
	Sulphur	€/t	100

(1) Biomass pellet prices were 120 €/t ARA (Amsterdam-Rotterdam-Antwerp) beginning 2010. Assuming an energy density of 19 GJ/t, this resulted in 6.3 €/GJ. TOPS are not produced commercially today. Although torrefaction results in additional costs, it was assumed that the reduction in transportation costs compensates this.[86] Therefore, it was assumed that the price for TOPS is also 6.3 €/GJ. Note that future improvements in torrefaction technology can result in delivery prices of TOPS being lower than those of EP.[86] During 2007-2010 biomass pellet prices fluctuated between 5.7-7.3 €/GJ.[60,61]

(2) Coal prices were 56 €/t ARA beginning 2010. Assuming an energy density of 25 GJ/t, this resulted in 2.25 €/GJ. During 2000-2010 coal prices fluctuated between 1.0-5.6 €/GJ.[1]

(3) CO_2 credit prices were fluctuating around 15 €/t CO_2 beginning 2010. Between 2005-2010 CO_2 credit prices fluctuated between 0-32 €/t CO_2.[62] Note that for the economic analysis, carbon locked in the chemical/fuel products are considered emissions and must be compensated by CO_2 credits.

(4) In the *Business as Usual* scenario it was assumed that no large-scale CO_2 transport infrastructure will be realised. Therefore, the CO_2 transport and storage price remained at 15 €/t.[156]

(5) In the *Direct Action* scenario it was assumed that a large-scale CO_2 transportation infrastructure would be realised. Costs dropped by 4% per year from 15 €/t CO_2 in 2015 to 10 €/t CO_2 in 2025. After that, CO_2 transport and storage costs remained at 10 €/t CO_2. In the *Delayed Climate Policy* scenario it was assumed that CO_2 transport and storage costs remained at 15 €/t CO_2 up till 2025, similar to the *Business as Usual* scenario. In 2025 rapid deployment of a large-scale infrastructure will bring CO_2 transport and storage costs down to 10 €/t CO_2 for the period 2025-2035.[156]

(6) Gasoline and diesel prices excluding duties and taxes were 350 €/m^3 in NW Europe beginning 2010. Assuming an energy density of 45 GJ/t and a mass density of 770 kg/m^3, this resulted in 10.1 €/GJ. During 2000-2010 gasoline and diesel prices fluctuated between 3-21 €/GJ. During that time, Brent oil prices fluctuated between 13-94 €/bbl (2-15 €/GJ).[65]

(7) Methanol prices were fluctuating around 250 €/t beginning 2010. Assuming an energy density of 22.7 GJ/t, this resulted in 11.0 €/GJ. During 2002-2010 methanol prices fluctuated between 6-23 €/GJ.[63]

(8) Urea prices were 200 €/t beginning 2010. Assuming an energy density of 10.5 GJ/t, this resulted in 19 €/GJ. During 1993-2010 urea prices fluctuated between 6-53 €/GJ.[155]

(9) Electricity prices were based on the average Dutch day-hourly market price between 2004-2008. The observed trends were considered representative for NW-Europe. During that period the electricity price varied between 0-1050 €/MWh (0-290 €/GJ), with an average price of 57 €/MWh (15.7 €/GJ).[66]

3.3.2 Capital costs data

Several literature sources were used to obtain component capital cost data. The main characteristics of these studies are given below. The selected data and argumentation is given in Table 3.4 and its footnotes.

- IEA GHG study *Co-production of hydrogen and electricity by coal gasification with CO_2 capture*, performed by FosterWheeler (FW) in 2007[49] and the economic update performed in 2008.[50]

This study calculates, among other, electricity production costs of IGCC facilities both with and without CCS. A Shell EF gasifier, Selexol AGR removal, Claus/SCOT sulphur removal and 9FA GE gas turbines are considered. Australian bituminous coal is used as feedstock. The CO_2 in the CCS facility is compressed to 110 bar. The main economic parameters are 10% discount rate, 25 years economic lifetime and a coal price of 1.2 $€_{2007}/GJ_{LHV}$. Capital costs are calculated on a component level, but only given on an aggregated level (gasifier, gas cleanup, power island, etc.). The 2008-study has identical technical configuration and technical performance as the 2007-study. The only difference is that the capital costs and coal price have been updated to reflect the rise in feedstock and construction prices.

- NETL study *Cost and Performance Baseline for Fossil Energy Plants* in 2007.[74]

This study calculates electricity production costs of various fossil fuel power plants. This includes a Shell EF IGCC power plant with and without CO_2 capture. The used technology is a Shell EF gasifier, Sulfinol AGR removal (for vent) or Selexol AGR removal (for CO_2 capture), Claus/SCOT sulphur removal and MS7001FB gas turbines. Illinois #6 coal is used as feedstock. The CO_2 in the CCS facility is compressed to 153 bar. The main economic parameters are 10% discount rate, 30 years economic lifetime and a coal price of 1.2 $€_{2007}/GJ_{HHV}$. Capital costs are calculated on a component level for the U.S. Midwest.

- Several studies from Princeton Environmental Institute by Williams, Larson et al.[24,25,72,79,80]

These studies investigate the conversion of biomass and coal to transportation fuels using a bottom-up approach on the component level using AspenPlus. Studied are the conversions of coal to FT-liquids and methanol, biomass to FT-liquids, and a mixture of coal and biomass to FT-liquids. In the process configurations considered, dedicated gasifiers are assumed for biomass and coal conversion to syngas (fluidised bad gasifier for biomass, GE entrained flow gasifier for coal). Capital costs are calculated on a component level, using a database which was updated in 2007.

- Hamelinck et al., *Future prospects for production of methanol and hydrogen from biomass* in 2002 and *Production of FT transportation fuels from biomass; technical options, process analysis and optimisation, and development potential* in 2004.[77,81]

These studies calculate production costs and efficiencies of biomass to methanol, hydrogen and FT-liquids for different process configurations. The scale investigated ranges between 80MW_{th} and 2000 MW_{th} input. No entrained flow gasifiers were evaluated in these studies. As in the previous studies, capital costs are calculated on a component level. Although the biomass to methanol and biomass to hydrogen process configurations considered differ from our study, many components are the same.

The cost data of the 2007 FW study have been used as point of departure, as beginning 2010 the capital costs for power plants were back at their 2007-level.[161] When large differences occurred between the cost data of the different sources, an expert judgement was made in selecting which source to use. The exact component cost data and motivation to select the data are given in Table 3.5 and its footnotes.

Table 3.4 Summary of key results on conversion system performance from the considered studies.

Feedstock	Product	Energy input (MW_{HHV})	CO_2 capture	Efficiency (chem. + elec.)	Production costs	Source
Coal	Electricity	1884	No	40%	52 €$_{2007}$/MWh	[49]
Coal	Electricity	2053	Yes	32%	72 €$_{2007}$/MWh	[49]
Coal	Electricity	1884	No	40%	70 €$_{2008}$/MWh	[50]
Coal	Electricity	2053	Yes	32%	94 €$_{2008}$/MWh	[50]
Coal	Electricity	1547	No	41%	59 \$$_{2007}$/MWh	[74]
Coal	Electricity	1616	Yes	32%	78 \$$_{2007}$/MWh	[74]
Coal	FT-liquids	7624	No	33% + 17%	8.2 \$$_{2007}$/GJ	[72]
Coal	FT-liquids	7624	Yes	33% + 14%	10.5 \$$_{2007}$/GJ	[72]
Switch grass	FT-liquids	661	No	45% + 5%	28.5 \$$_{2007}$/GJ	[72]
Switch grass	FT-liquids	661	Yes	45% + 4%	30.5 \$$_{2007}$/GJ	[72]
Coal + Switch grass	FT-liquids	2099 + 661	Yes	32% + 15%	17.1 \$$_{2007}$/GJ	[72]
Coal	Methanol	2870	No	25% + 22%	8.0 \$$_{2002}$/GJ	[79]
Coal	Methanol	2870	Yes	25% + 20%	9.9 \$$_{2002}$/GJ	[79]
Biomass	Methanol	1000	No	57% + 0%	7.6 \$$_{2001}$/GJ	[81]
Biomass	Hydrogen	1000	No	70% - 5%	7.0 \$$_{2001}$/GJ	[81]
Biomass	FT-liquids	1000	No	42% + 4%	14.6 €$_{2002}$/GJ	[77]

In Table 3.5, the base costs are the bare equipment costs. These were increased by the installation factor (IF), which consisted of direct (instrumentation and control, buildings, grid connection, site preparation, civil works, electronics and piping) and indirect (engineering, building interest, contingency, fees, overhead, profits and start-up costs) costs. The direct costs decreased with increasing scale, using a scaling factor of -0.18.[77] When no values for the direct and indirect costs are available, 33% direct and 50% indirect costs were assumed at the scale used in the original literature source.[77] It was assumed that the IG-PG facilities consist of two main trains (see Figure 3.11), with four parallel feedstock pre-treatment set-ups. Depending on the production, up to 3 parallel chemical conversion trains were used.

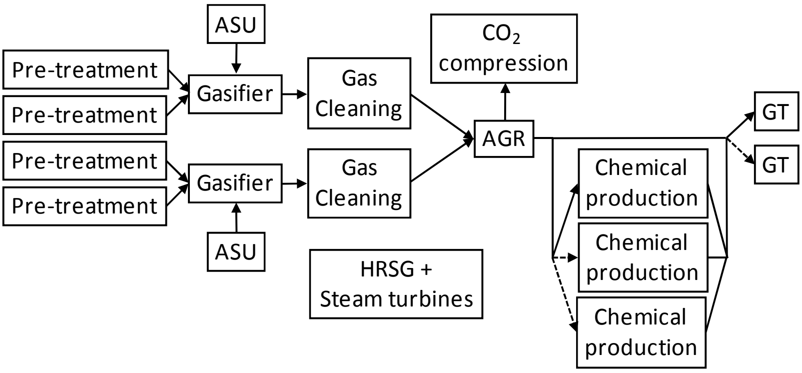

Figure 3.11 Schematic representation of train configuration of IG-PG facility

Table 3.5 Selected capital costs of the components of the IG-PG facilities investigated.

Component	Base Cost (M€$_{2008}$)	Base scale	Scaling factor	Scale unit	Direct costs	Indirect costs	# units (1)
Pre-treatment and feeding							
Biomass (2)	10	65	0.77	tonne a.r./hr	27%	25%	0-4
Coal (3)	41	273	0.65	tonne a.r./hr	53%	14%	0-4
Gasification							
Air separating unit (95% O_2) (4)	112	5149	0.8	tonne output/d	37%	16%	2
Shell EF gasifier (5)	139	2053	0.66	MW_{HHV} coal eq.	73%	24%	2
Gas Cleaning							
Candle filter (6)	2	12	0.65	gas m^3/s	33%	50%	2
Wet Scrubber (6)	3	12	0.7	gas m^3/s	33%	50%	2
Heat exchanger (6)	9	138	0.6	MW_{th}	31%	40%	2
Sour WGS reactor (7)	14	8819	0.65	H_2+CO kmol/hr	0%	81%	2
Guard bed (8)	0	0.6	1	gas Nm^3/s	0%	200%	1
Sweet WGS reactor (7)	14	8819	0.65	H_2+CO kmol/hr	0%	81%	1
H_2 PSA (8)	12	16616	0.65	H_2 kmol/hr	100%	28%	1
Acid Gas Removal							
Rectisol AGR removal (9)	120	554	0.7	CO_2 tonne/hr	33%	50%	1
Claus/SCOT (10)	16	56	0.67	S tonne/d	33%	14%	1
Power Isle							
Gas Turbine (11)	42	266	0.75	Net MW_e	16%	27%	1-2
Steam Turbines and Cycle (12)	38	275	0.67	Gross MW_e	16%	27%	1
HRSG (13)	34	355	1	MW_{th} exch.	16%	27%	1
FT Isle							
FT Slurry Reactor (14)	21	131	0.72	MW_{HHV} FT	33%	50%	0-3
Product Upgrading (14)	132	220	0.7	FT tonne/hr	33%	50%	0-3
MeOH Isle							
MeOH Reactor (15)	13	42	0.65	MeOH tonne/hr	15%	50%	0-1
Product Upgrading (15)	1	17	0.65	MeOH tonne/hr	15%	50%	0-1
Urea Isle							
Ammonia Production (16)	150	2000	0.7	NH_3 tonne/d	33%	50%	0-1
Urea Reactor + upgrading (16)	150	3500	0.7	urea tonne/d	33%	50%	0-1
CCS							
CO_2 Compression (17)	6	13	1	MW_e	16%	32%	1

Miscellaneous							
Syngas Compressor (17)	4	10	1	MW_e	16%	32%	Varies

(1) The number of units depends on the used configuration. A total of four feeding trains are used, divided over biomass and coal. The division depends on the biomass/coal ratio. The gasifier and gas cleaning sections have two trains. Before the AGR, the syngas streams combine. Downstream, only one train was used. The maximum output of each gas turbine was fixed at 302 MW_e, meaning 2 gas turbines are needed in the coal to liquid case and all power cases. Currently the GT-26 is rated at 289 MW_e.[165]

(2) This includes the storage, handling and feeding of the biomass pellets. A scaling factor of 0.77[24,77] is reported. [24] reports, corrected for feeding at 40 bar, 14 $M\$_{2007}$ for 65 t as received (a.r.) wood/hr and an IF of 59%. [77] reports 11 $M€_{2002}$ for 34 t a.r. wood/hr and an IF of 100%. This translates to 17 M€ and 47 M€ respectively for 76 t a.r. wood/hr (≈366 MW_{HHV} biomass). The value of [24] was taken as they are for a Nth facility, while the data from [77] is for a first-of-a-kind facility.

(3) This includes the storage, handling and feeding of the coal. Scaling factors between 0.63[50] and 0.65[74] are reported. A scaling factor of 0.65 was used. [50] reports 40 $M€_{2007}$ for 273 t a.r. coal/hr and an IF of 74%. [74] reports 118 $M\$_{2006}$ for 215 t a.r. coal/hr and an IF of 73%. This translates to 71 M€ and 73 M€ respectively for 66 t a.r. coal/hr (≈ 500 MW_{HHV} coal). The value of [50] was taken.

(4) This includes the air separation unit and the oxygen compression to 40 bar. Scaling factors of 0.5,[24] 0.75[77] and 0.8[50] are reported. A scaling factor of 0.8 was used. [50] reports 110 $M€_{2007}$ for 5150 t/d and an IF of 53%. [74] reports 144 $M\$_{2007}$ for 4070 t/d and an IF of 20%. [77] reports, corrected for 40 bar, 27 $M€_{2002}$ for 576 t/d and an IF of 30% for the ASU and 21 $M€_{2002}$ for 13 MW_e and an IF of 72% for the oxygen compressor. This translates to 170 M€, 146 M€ and 205 M€ respectively for 2493 t/d. The value of [50] was taken.

(5) Scaling factors of 0.66,[50] 0.67[24] and 0.7[22] are reported. A scaling factor of 0.66 was used. [50] reports 137 $M€_{2007}$ for 2053 MW_{HHV} and an IF of 98%. [74] reports 128 $M\$_{2007}$ for 1618 MW_{HHV} and an IF of 114% [22] reports 120 $M€_{2005}$ for 600 MW_{HHV} and an IF of 10%. This translates to 276 M€, 222 M€ and 199 M€ respectively for 1000 MW_{HHV}. The value of [50] was taken as it is the most up-to-date and in €.

(6) Data was taken from [77] and indexed to $€_{2008}$.

(7) Scaling factors of 0.65[77] and 0.67[24] are reported. A scaling factor of 0.65 was used. [74] reports 8.8 $M\$_{2006}$ for 1,533 MW_{HHV} coal input (≈ 15,500 kmol (H_2+CO)/hr) and an IF of 84%. [77] reports 12 $M€_{2002}$ for 8,819 kmol (CO+ H_2)/hr and an IF of 81%. This translates to 6 M€ and 19 M€ respectively for 11,500 kmol (H_2+CO)/hr (≈ 1,000 MW_{th} coal input). The value of [77] was taken as the value of [74] was deemed too low.

(8) Scaling factors of 0.65,[50] 0.7[77] and 0.74[25] are reported. A scaling factor of 0.65 was used. [50] reports 12 $M€_{2007}$ for 16,616 kmol H_2/hr and an IF of 128%. [25] reports 7 $M€_{2002}$ for 1,058 kmol H_2/hr, including BOP and indirect costs. [77] reports 33 $M€_{2002}$ for 9.600 kmol H_2/hr and an IF of 69%. This translates to 17 M€, 17 M€ and 49 M€ respectively for 6,500 kmol H_2/hr. The value of [50] was taken as it is the most up-to-date.

(9) Scaling factors of 0.63,[24] and 0.7[77] are reported. A scaling factor of 0.7 was used. [74] and [24] use a Selexol based AGR. Using Rectisol instead of Selexol increases costs by 75%.[50] [50] reports 119 $M€_{2007}$ for 554 t CO_2/hr and an IF of 100%. [74] reports, after correction, 104 $M\$_{2006}$ for 453 t CO_2/hr and an IF of 189%. [24] reports 52 $M\$_{2007}$ for 200,000 Nm^3/hr (≈ 82.5 t CO_2/hr at a CO_2 concentration of 21 mol%) and an IF of 59%. [77] reports, after correction, 55 $M€_{2002}$, for 436 t CO_2/hr assuming a IF of 100%. This translates to 180 M€, 181 M€, 148 M€ and 108 M€ respectively for 345 t CO_2/hr. The value of [50] was taken as it is the most up-to-date and in €. A direct cost factor of 33% at a scale of 554 t CO_2/hr and an indirect cost factor of 50% was assumed.

(10) A scaling factor of 0.67[74,24] is reported. The Claus and Scot units receive the acid gas stream after Rectisol processing. This stream has a higher sulphur content than after Selexol processing. This reduces costs by 46%.[50] [50] reports 21 M€$_{2007}$ for 56 t S/d and a IF of 14%. [74] reports, after correction, 5 M\$$_{2007}$ for 129 t S/d and a IF of 228%. [24] reports 25 M\$$_{2007}$ for 137 t S/d and a IF of 52%. This translates to 49 M€, 13 M€ and 29 M€ respectively for 160 t S/d. The value of [50] was taken as it is the most up-to-date and in €.

(11) Scaling factors of 0.7[166] and 0.75[24] are reported. A scaling factor of 0.75 is used as, according to Kreutz et al.,[24] this gave the best fit based on the Gas Turbine World's 2003 Handbook. [74] reports 44 M\$$_{2006}$ for 232 MW$_e$ and a IF of 42%. [166] reports 17 M€$_{2003}$ for 26 MW$_e$ and an IF of 100%. [24] reports 56 M\$$_{2007}$ for 266 MW$_e$ and a IF of 47%. Using the scaling factor this translates to 61 M€, 243 M€, and 70 M€ respectively for 360 MW$_e$. Compared to quoted costs in the GTW 07/08,[165] with bare equipment costs of the gas and steam sections of a 400 MW$_e$ combined cycle power plant around 136 M€, the value from Hamelinck is far too high. This is most likely a result from the low initial scale and the use of first generation data. The value from Princeton is taken as it is more up-to-date than the NETL data.

(12) Scaling factors of 0.66,[74] 0.67[24] and 0.7[166] are reported. A scaling factor of 0.67 was selected. [74] reports 33 M\$$_{2006}$ for 230 MW$_e$ and a IF of 65%. [166] reports 5 M€$_{2003}$ for 10 MW$_e$ and a IF of 100%. [24] reports 51 M\$$_{2007}$ for 275 MW$_e$ and a IF of 47%. Using the scaling factor this translates to 52 M€, 114 M€, and 61 M€ respectively for 360 MW$_e$. The value from Hamelinck appears too high, most probably due to the use of first generation data. The value from Princeton is taken as it is more up-to-date than the NETL data.

(13) This is often incorporated with the steam or gas turbines. As the IG-PG facility has more heat streams than a normal IGCC, the costs of the HRSG are calculated separate. A scaling factor of 1.0[24] is reported. According to [24] generating 355 MW$_{th}$ costs 46 M€.

(14) For the FT-synthesis a slurry reactor at 60 bar was used. Scaling factors of 0.72[77] and 0.75[24] are reported. The lowest scaling factor was selected. [24] reports 24 M\$$_{2007}$ for 103 MW$_{HHV}$ FT and a IF of 67%. [77] reports, after correcting for 60 bar, 18 M€$_{2002}$ for 131 MW$_{HHV}$ FT and a IF of 100%. Using the scaling factor this translates to 73 M€ and 92 M€ respectively for 400 MW$_{HHV}$ FT. The highest value was selected. The FT product upgrading data is based on [77], reporting a scaling factor of 0.7 and 233 M€$_{2002}$, including BOP and indirect costs, for 286 m^3 FT product. This translates to 37 M€ for 31 t FT/hr (≈400 MW$_{HHV}$).

(15) Scaling factors of 0.65[79] and 0.72[81] are reported for a once-through LPMeOH synthesis reactor and 0.291[79] and 0.7[81] for methanol separation and purification. The value from [79] for the purification appears too low for a stand-alone facility. Therefore, a scaling factor of 0.7 were selected. [79] reports 13.6 M\$$_{2002}$ for 42 t MeOH/hr and a IF of 73% for the synthesis and 1.1 M\$$_{2002}$ for 16.8 t MeOH/hr and a IF of 73% for the purification. [81] reports 3.5 M\$$_{2001}$ for 87.5 t MeOH/hr and a IF of 110% for the synthesis and 15 M\$$_{2001}$ for 87.5 t MeOH/hr and a IF of 110% for the purification. Using these scaling factors, this translates to 40 M€, and 35 M€ respectively for the synthesis and purification of 100 t MeOH/hr (≈630 MW$_{HHV}$). The values from [79] were selected as they result in the highest capital costs.

(16) Production of 3500 t/d urea from natural gas has a capital investment of around 1000 M€. This is divided into 400 M€ for ammonia production, 300 M€ for urea production and 300 M€ for off-sites.[167] Producing ammonia from syngas removes the need of desulphurisation, reforming and shifting, but still requires an ammonia reactor, separation and recycling. Therefore, it is assumed that ammonia production from syngas has a capital costs 300 M€. The urea production capital costs are kept at 300 M€.

(17) Compressor costs are identical for all gas compressors, regardless of the content of the gas. The exceptions are the supercritical CO_2 compressor and the O_2 compressor. The O_2 compressor is, due to fire-hazards, non-lubricated, adding 40% to the costs.[77] The supercritical CO_2 compressor must handle a phase change during compression. Scaling factors of 0.67,[24] 0.85[77] and 1.0[144] are reported. A scaling factor of 1.0 was used as this data was supplied by a

petro-chemical company. [24] reports 6 M$\$_{2007}$ for 10 MW_e and a IF of 52%. [77] reports 13 M€$_{2002}$ for 13 MW_e and a IF of 72%. This translates to 15 M€ and 51 M€ respectively for 27 MW_e. Contrary to a Selexol system, the CO_2 stream coming from a Rectisol unit does not need drying. Capital costs data of the CO_2 compressor based on a Selexol system is lowered by -36% to account for the absence of drying section.[25] [74] reports, after removing the drying section, 11 M$\$_{2006}$ for 28 MW_e and a IF of 109%. [24] reports 8 M$\$_{2007}$ for 13 MW_e and a IF of 48%. [50] reports 42 M€$_{2007}$ for 52 MW_e and a IF of 14%. Using the scaling factor this translates to 16 M€, 17 M€ and 26 M€ respectively for 27 MW_e. The highest value is taken. The data of [24] for the normal compressors falls within the costs estimated for the CO_2 compressor and is therefore selected. It is therefore assumed that their value for the CO_2 compressor is also the best.

3.4 Results

In this section, results are presented per case study. This is followed by a sensitivity analysis of the *Current Situation* scenario. Detailed breakdowns of costs, revenues and component capital costs are given in the supplementary data.

3.4.1 Static IG-PG facilities

Production costs of the main product of static IG-PG facilities under the *Current Situation* scenario varies between 9-38 €/GJ, depending on feedstock and product (see Figure 3.13). Capital cost breakdowns of the different cases are given in Figure 3.12. Detailed breakdowns of annual costs, revenues and capital costs are given in the supplementary data.

When producing electricity, FT-liquids or methanol, the gasifier represents the largest share of the capital costs (24-33%), regardless of the feedstock used. When producing urea, the urea section represents the largest share of the capital costs (33-36%), followed by the gasifier (18-22%). In all cases the pre-treatment, gas cleaning and gas turbine equipment contribute less (combined 11-22%).

Under *Current Situation* scenario conditions, none of the IG-PG facilities have production costs under the average market price of their products. The pure coal cases show significantly lower production costs than the biomass cases, mainly due to lower feedstocks costs and, to a lesser extent, a higher output which results in lower relative capital costs. Coal based production costs are 18 €/GJ for electricity; 13 €/GJ for FT-liquids; 12 €/GJ for methanol and 21 €/GJ for urea. Using TOPS results in lower production costs compared to using EP, with the largest difference occurring when producing urea (Δ=5.0 €/GJ). Using TOPS results in production costs of 26 €/GJ for electricity; 19 €/GJ for FT-liquids; 22 €/GJ for methanol and 33 €/GJ for urea. The market price of the products has occasionally exceeded these values in the past years. The use of EP results in production costs of 29 €/GJ for electricity; 22 €/GJ for FT-liquids; 26 €/GJ for methanol and 38 €/GJ for urea. In the last decade, only electricity and urea had market prices reaching this value. Note that for the CO_2 credits, carbon locked in the products was treated as emissions. If the locked carbon would be treated as sequested carbon, production costs of the chemicals would be around 1 €/GJ lower than the values given above.

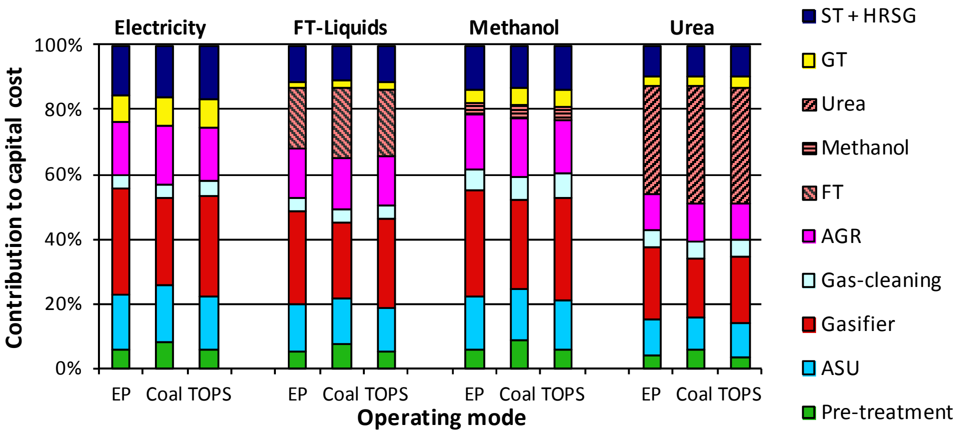

Figure 3.12 Capital cost breakdowns static IG-PG facilities. The facilities use only one feedstock.

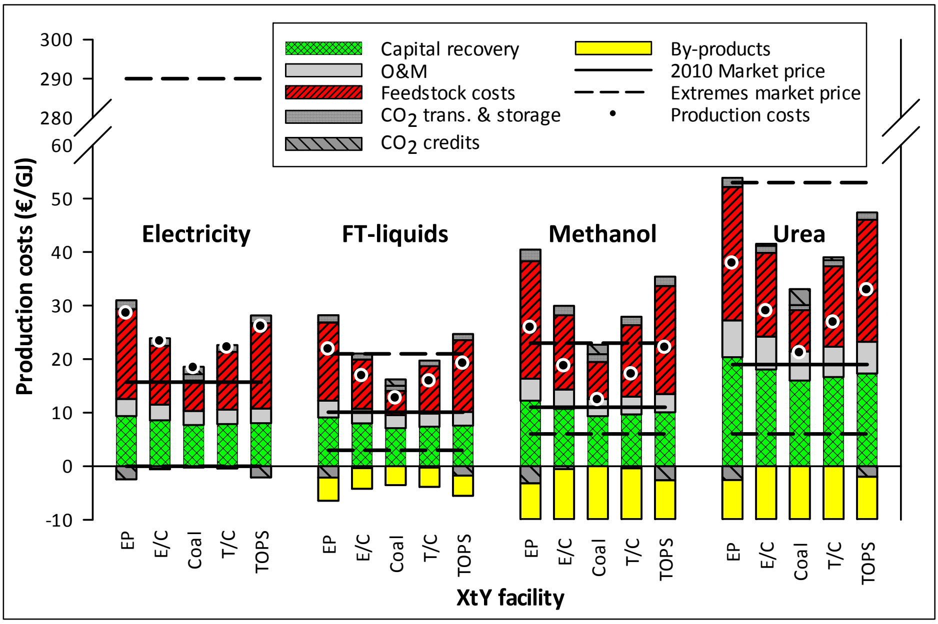

Figure 3.13 Breakdown production cost of static IG-PG facilities under *Current Situation* scenario: 2.25 €/GJ coal; 6.3 €/GJ biomass pellet; 10 €/t CO_2 for transport and storage; 15 €/t CO_2 credit; 15.7 €/GJ co-produced electricity; 2000 MW_{th} coal eq. input; 4% O&M costs; 10% discount rate; 20 year lifetime. The dashed lines show the minimum and maximum market price of the produced commodity that occurred during 2000-2010.

When producing chemicals, there is the option to capture CO_2 and venting it to the atmosphere. At a CO_2 credit price of 15 €/t CO_2, this slightly increases production costs (+ 0.1 €/GJ) when producing FT-liquids or methanol, as there is no need to compress the CO_2. When producing urea, CO_2 is needed as a feedstock and must be compressed. Consequently, venting the CO_2 increases production costs by 0.5 €/GJ.

Table 3.6 Impact scenarios on production costs of static IG-PG facilities (€/GJ).

	Production costs under different scenarios					
	Scenario	EP	EP/Coal	Coal	TOPS/Coal	TOPS
X to Electricity (Market price: 15.7 €/GJ Range: 0-290 €/GJ)	Current	28.5	23.3	18.4	22.2	26.0
	B as U	28.9	23.9	19.2	22.8	26.4
	Sustainable	25.6	23.1	20.6	22.2	23.8
	Green	23.5	22.0	20.5	20.9	21.2
	Trigger	27.7	23.6	19.6	22.5	25.5
X to FT-Liquids (Market price: 10.1 €/GJ Range: 3-21 €/GJ)	Current	21.8	16.9	12.7	15.9	19.2
	B as U	22.1	17.4	13.3	16.4	19.5
	Sustainable	19.3	16.7	14.4	15.9	17.3
	Green	17.4	15.8	14.4	14.8	15.1
	Trigger	21.1	17.1	13.7	16.1	18.7
X to Methanol (Market price: 11.0 €/GJ Range: 6-23 €/GJ)	Current	25.9	18.7	12.4	17.1	22.2
	B as U	26.4	19.5	13.4	17.9	22.6
	Sustainable	22.1	18.4	15.1	17.2	19.3
	Green	19.2	17.0	15.1	15.5	16.1
	Trigger	24.8	19.0	13.9	17.6	21.4
X to Urea (Market price: 19.0 €/GJ Range: 6-53 €/GJ)	Current	37.9	29.0	21.1	26.8	32.9
	B as U	38.4	29.8	22.3	27.7	33.4
	Sustainable	36.0	31.0	26.6	29.3	32.1
	Green	30.3	27.1	24.1	25.0	26.0
	Trigger	37.5	30.1	23.6	28.1	32.8

Table 3.6 presents the impact of the different scenarios on the calculated production costs. When using coal as feedstock, production costs are lowest under *Current Situation* scenario conditions, being 12-21/GJ. Highest production costs are obtained in the *Direct Action* scenario, ranging between 14-27 €/GJ. The main reason is the higher CO_2 credit prices, increasing the production cost by 2.8-7.3 €/GJ, which is not compensated by the reduction in feedstock costs (-2.1- -1.2 €/GJ). The increase in CO_2 transport and storage costs is small, raising production costs by 0.2-0.3 €/GJ. The *Business as Usual* and *Delayed Climate Policy* scenarios show similar trends as the *Direct Action* scenario, although the absolute variation due to decreasing coal price and increasing CO_2 credit and CO_2 transport and storage prices varies. The only differences between the *Green* and *Current Situation* scenarios are the feedstock costs. In the *Green* scenario, higher coal prices increase production costs by 1.8-3.0 €/GJ.

When using biomass as feedstock, production costs are lowest when using TOPS, being 1.8-5.0 €/GJ lower than when using EP. Therefore, only the results related to TOPS are discussed here. Lowest production costs are obtained under the *Green* scenario (15 €/GJ for FT-liquids till 26 €/GJ for urea production) due to the lower biomass pellet costs. Highest production costs are obtained under the *Business as Usual* scenario, ranging between 19 €/GJ for FT-liquids and 33 €/GJ for urea production. Production costs are affected in three ways. First, increasing TOPS prices raise production costs by 0.3-0.5 €/GJ.

Second, higher CO_2 credit prices decrease production costs by 0.6-0.9 €/GJ as, contrary to the coal cases, stored CO_2 count as a revenue and emitted CO_2 or CO_2 locked in the chemicals/fuels are not charged. Third, the higher CO_2 transport and storage costs increase production costs by 0.6-0.9 €/GJ. The net effect is a small increase in production costs. The *Direct Action* and *Delayed Climate Policy* scenarios show similar trends, but the higher CO_2 credit prices yield sufficient revenues to obtain a net reduction in production costs.

FT-liquids can be produced from TOPS for 19 €/GJ under *Current Situation* conditions. This roughly corresponds to an oil price of 105 €/bbl[14] (155 \$/bbl). However, with lower biomass pellet prices, as in the *Green* scenario, production costs decrease considerable. The 15 €/GJ production costs in the *Green* scenario is break-even with an oil price of 81 €/bbl (118 \$/bbl). In the past years oil prices already exceeded this value (see Figure 3.8).

Finally, results also indicate that higher CO_2 transport and storage prices have a limited effect (between 0.9-3.2 €/GJ for all scenarios) on production costs (see Figure 3.13 for the *Current Situation* scenario). As expected, the highest values are found in the *Business as Usual* scenario, while the *Current Situation* and *Green* scenarios show the lowest values.

3.4.2 Variation of feedstock

Feedstock flexible IG-PG facilities have higher capital costs and slightly lower overall efficiencies compared to static IG-PG facilities. However, they can respond to fluctuations in feedstock prices. We investigated whether feedstock flexibility pays off. First, the increase in capital costs and the impact on production costs of having a feedstock flexible IG-PG facility instead of a static IG-PG facility is presented. Optimal economic returns in relation to feedstock choice were calculated for different biomass pellet, coal and CO_2 credit prices, taking into account the variation in electricity price. Finally, a weekly variation in biomass pellet and coal prices was simulated at different CO_2 credit prices to assess the impact of short-term price variation on the economic performance of feedstock flexible and static IG-PG facilities.

3.4.2.1 Case 2. Capital costs penalty

Production costs of IG-PG facilities with static production, but with three levels of feedstock flexibility (0, 50 and 100%) were compared, providing insights into the penalty induced by incorporating this flexibility in an IG-PG facility. The relative increase in capital costs is given in Table 3.7 and the impact on production costs is given in Table 3.8. The results were calculated for the *Current Situation* scenario. However, as the scenarios do not impact capital costs, the found trends also apply for the other scenarios.

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[14] The break-even production costs of crude oil based gasoline/diesel consist of the crude oil price, a refinery margin and a carbon tax. A 6.0 €/bbl refinery margin was used, based on 46.5 $\$_{2002}/m^3$ diesel.[25] A 6.0 €/bbl CO_2 carbon tax was used, assuming 15 €/t CO_2. The following properties were used: an oil density of 845 kg/m^3; an oil carbon emission of 0.4 t CO_2/bbl, assuming a carbon mass content of crude oil of around 80%; an oil energy density of 6.12 GJ/bbl; 0.159 m^3 per bbl; and a currency conversion rate of 1.23 $\$_{2002}$ = 1.47 $\$_{2008}$ = 1.00 $€_{2008}$.

Table 3.7 Impact feedstock flexibility on capital costs compared to static design.

	Flexibility degree	Feedstock (X)					
Starting feedstock		EP	E/C	Coal	Coal	T/C	TOPS
Flexible towards		Coal	Coal	EP	TOPS	Coal	Coal
X to Electricity	0%	100%	100%	100%	100%	100%	100%
	50%		112%	103%	100%	106%	
	100%	125%	114%	105%	101%	107%	116%
X to FT-liquids	0%	100%	100%	100%	100%	100%	100%
	50%		113%	103%	100%	106%	
	100%	127%	115%	104%	100%	106%	116%
X to Methanol	0%	100%	100%	100%	100%	100%	100%
	50%		111%	103%	100%	106%	
	100%	124%	114%	105%	101%	106%	115%
X to Urea	0%	100%	100%	100%	100%	100%	100%
	50%		113%	102%	100%	106%	
	100%	128%	114%	103%	100%	107%	115%

Table 3.8 Impact on production costs of increased capital costs needed for feedstock flexibility under *Current Situation* scenario (€/GJ). (1)

	Flexibility degree	Feedstock (X)					
Starting feedstock		EP	E/C	Coal	Coal	T/C	TOPS
Flexible towards		Coal	Coal	EP	TOPS	Coal	Coal
X to Electricity	0%	28.5	23.3	18.4	18.4	22.2	26.0
(Market price	50%		24.7	18.7	18.4	22.9	
= 15.7 €/GJ)	100%	31.7	24.9	18.9	18.4	22.9	27.7
X to FT-liquids	0%	21.8	16.9	12.7	12.7	15.9	19.2
(Market price	50%		18.3	12.9	12.7	16.5	
= 10.1 €/GJ)	100%	25.1	18.5	13.1	12.7	16.5	20.8
X to Methanol	0%	25.9	18.7	12.4	12.4	17.1	22.2
(Market price	50%		20.3	12.7	12.4	17.9	
= 11.0 €/GJ)	100%	29.9	20.6	13.0	12.4	18.0	24.2
X to Urea	0%	37.9	29.0	21.1	21.1	26.8	32.9
(Market price	50%		32.1	21.6	21.2	28.3	
= 19.0 €/GJ)	100%	45.4	32.4	21.8	21.2	28.3	36.4

(1) *Current Situation* scenario conditions: 2.25 €/GJ coal; 6.3 €/GJ biomass pellet; 10 €/t CO_2 for transport and storage; 15 €/t CO_2 credit;15.7 €/GJ co-generated electricity; 2000 MW_{th} coal eq. input; 4% O&M costs; 10% discount rate; 20 year lifetime.

Results indicate that an IG-PG facility designed as feedstock flexible, but operated as a static facility has production costs up to 20% higher compared to a static designed IG-PG facility. The increase in costs depends both on the used feedstock and desired production. The lowest penalty is obtained when making a coal based power plant (CtP) suitable for TOPS, while making an EP based urea (EtU) facility suitable for coal results in the highest penalty. The main reason for the added costs when re-designing a biomass based facility to be able to process coal is the need for larger Claus/SCOT and chemical conversion installations.

A major difference between EP and TOPS is the possibility to feed TOPS using the coal feeding system. EP is too fibrous and always requires a dedicated and different feeding system. This increases capital costs when constructing an IG-PG facility capable of processing both biomass and coal.

3.4.2.2 Case 3. Constant chemical production and variable electricity production by feedstock substitution

A way to exploit the daily electricity variation is using different feedstocks depending on the electricity price. For this scenario it was assumed that switching between coal and biomass is instantaneously. In a previous study,[150] it was found that the amount of gas generated in the EF gasifier must be kept constant. Due to differences in the heating value of biomass and coal, this means that the energetic value of the syngas is highest when using coal and lowest when using EP. This effect can be used to operate an IG-PG facility as a mid-load[13] power plant. At low electricity prices, the facility will use biomass with maximal chemical production. When electricity prices increase, the feedstock is switched from biomass to coal, but chemical production is kept constant. The additional chemical energy in the syngas is used for extra electricity production. However, this principle is interesting only when feedstock prices per unit of energy, including CO_2 credit costs, are lower for biomass pellets than for coal. The *Current Situation* scenario has a low price for coal and CO_2 credits and a high price for biomass pellets. At these conditions, coal is more favourable than biomass pellets, regardless of the electricity price. However, with increasing CO_2 credit price, biomass pellets became more competitive with coal and eventually replaced coal (see Figure 3.14).

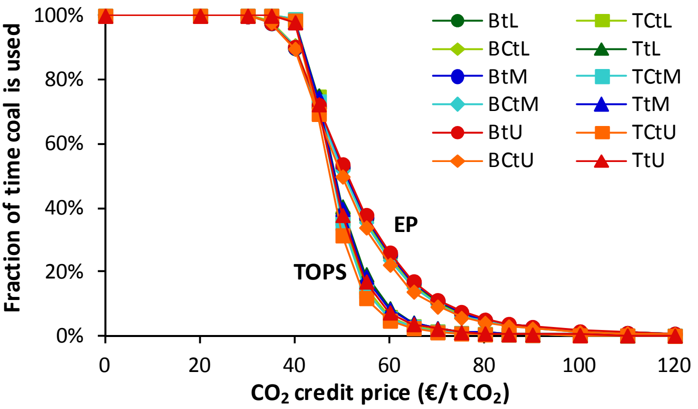

Figure 3.14 Impact CO_2 credit price on feedstock preference under *Current Situation* scenario: 6.3 €/GJ biomass pellets and 2.25 €/GJ coal.

EP becomes attractive at a CO_2 credit price lower than TOPS (30 €/t CO_2 compared to 40 €/t CO_2). However, full feedstock substitution is reached at a lower CO_2 credit price for TOPS (70 €/t CO_2) than for EP (100 €/t CO_2). For this analysis the electricity price was varied according to the Dutch day-hourly market prices between 2004-2008. If the electricity price is kept constant at 15.7 €/GJ, a sharp transition occurs around 54 €/t CO_2 for EP and 50 €/t CO_2 for TOPS, at which complete substitution of coal to biomass is observed.

The same effect of gradual substitution of coal by biomass is observed with increasing coal price or decreasing biomass pellet price. Analogous to the CO_2 credit price impact, EP begins replacing coal earlier than TOPS, but TOPS replaces coal faster and achieves complete substitution at lower coal prices or higher biomass pellet prices than EP (see Figure 3.15).

Figure 3.16 shows the relation between varying biomass pellet, coal and CO_2 credit prices for a flexible TOPS/coal to FT-liquids (XtL) facility with constant FT-liquids output. The base load electricity was sold for a fixed price (15.7 €/GJ) and the additional electricity was sold conform market spot prices. Biomass is the preferred feedstock in front of the two vertical planes, while coal is preferred behind them. Between the two planes feedstock preference is shared between biomass and coal, making this the space where feedstock flexibility is desired. The CO_2 credit price influenced the feedstock preference by tilting the planes, making biomass more economical at higher CO_2 credit prices. The difference between low and high biomass fractions is constant (2.2 €/GJ biomass pellet, 1.9 €/GJ coal and 22 €/t CO_2 respectively).

At 2010-market prices of 6.3 €/GJ biomass pellets, 2.25 €/GJ coal and 15 €/t CO_2, coal is the preferred feedstock. However, the historical data shows price variations between 5.7-7.3 €/GJ biomass pellet, 1.0-5.6 €/GJ coal and 0-32 €/t CO_2 (Figures 4-6), represented by the dark blue box in Figure 3.16. Although coal is still mostly the preferred feedstock within these ranges, within the area ABCD[15] in Figure 3.16, both biomass and coal are preferred and a feedstock flexible facility is desired.

The graph clearly shows that a higher CO_2 credit price allows the use of more expensive biomass. At 15 €/t CO_2 and 2.7 €/GJ coal, a price of 4.2 €/GJ TOPS results in a biomass preference of 95% of the time. If the CO_2 credit price increases to 30 €/t CO_2, the TOPS price can increase to 5.7 €/GJ TOPS, while still being the preferred feedstock for 95% of the time. At these conditions, FT-liquids can be produced for 15.8 €/GJ, equivalent to 116 $/bbl oil[14] if the oil is also taxed for 30 €/t CO_2.

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[15] Coordinates: A (7.3;5.6;32); B (7.3;3.9;32); C (5.7;2.6;32); D (5.7;4.4;32).

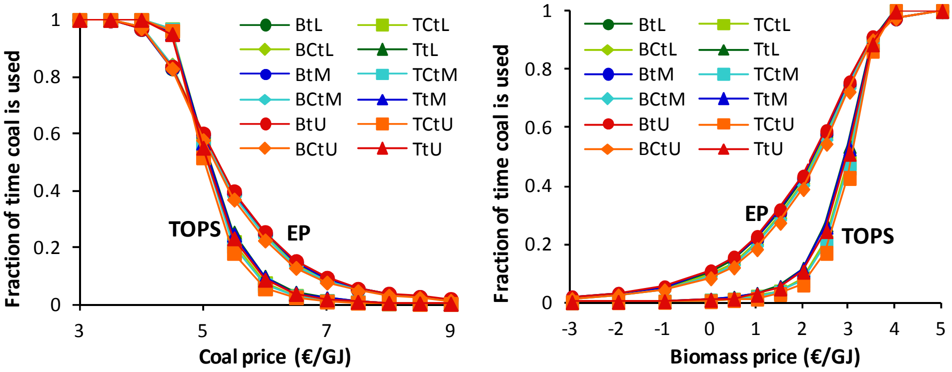

Figure 3.15 Impact biomass pellet and coal prices on feedstock preference under *Current Situation* scenario: 6.3 €/GJ biomass pellets (left), 2.25 €/GJ coal (right) and 15 €/t CO_2.

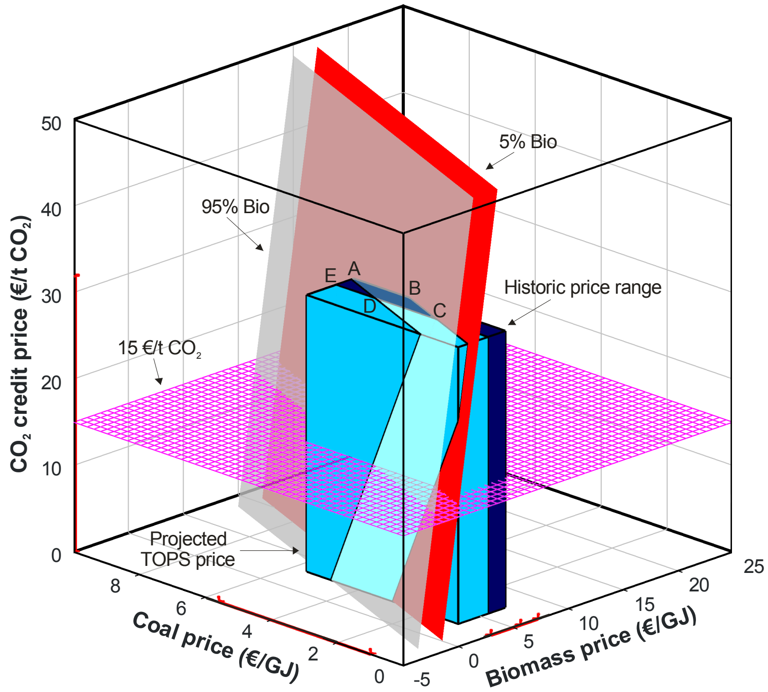

Figure 3.16 The effect of biomass pellets, coal and CO_2 credit price on feedstock preference. The horizontal plane illustrates a CO_2 credit price of 15 €/t CO_2. In front of the light (grey) shaded vertical plane biomass is the preferred feedstock for more than 95% of the time, while behind the dark (red) shaded plane biomass is preferred less than 5% of the time. Between the two planes, both biomass and coal are preferred. Note that the dark blue block shows the price ranges that have occurred. The light blue block illustrates a drop in biomass pellet price to 3.0 €/GJ.

3.4.2.3 Case 4. Impact of short-term feedstock price variation

Feedstock costs account for 23% (for coal) to 57% (for biomass) of total production costs under *Current Situation* scenario conditions. In this case study the effects of short-term variable feedstock prices on the economics of feedstock flexible IG-PG facilities were analysed. The flexible facilities adjusted, depending on the feedstock prices, their feedstock between either a 50/50 TOPS/coal mixture and pure coal (semi-feedstock flexible), or between pure TOPS and pure coal (full feedstock flexible). Only TOPS was used for the analysis as the results from the previous case studies showed that TOPS based facilities have better economics than EP based facilities.

Based on the results of case 3, four different CO_2 credit prices were examined. The reference price of 15 €/t CO_2; 40 €/t CO_2 reflected the value where TOPS is sometimes the preferred feedstock; 50 €/t CO_2 is the value where TOPS is the preferred feedstock for about 50% of the time; and at 65 €/t CO_2, TOPS is almost always the preferred feedstock (Figure 3.14).

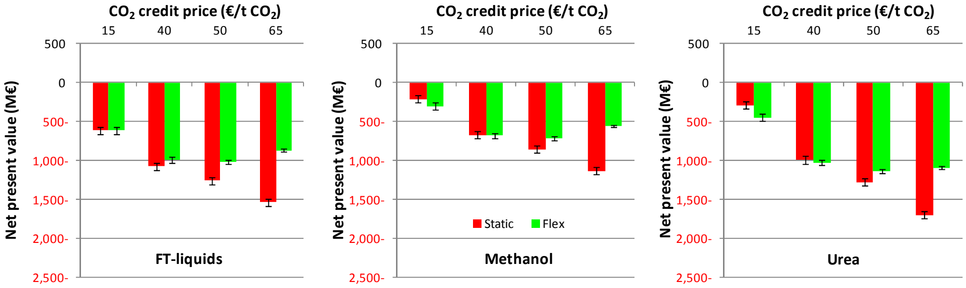

Figure 3.17 NPV of static coal facilities (left columns) and full feedstock flexible TOPS/coal facilities (right columns) with constant chemical production under *Current Situation* scenario conditions: feedstock prices based on historic data; 10 €/t CO_2 for transport and storage; 15.7 €/GJ base load electricity; 10.1 €/GJ FT-liquids; 11.0 €/GJ methanol; 19.0 €/GJ urea; 2000 MW_{th} coal eq. input; 4% O&M costs; 10% discount rate; 20 year lifetime.

Results for the *Current Situation* scenario (Figure 3.17) show that at a CO_2 credit price of 15 €/t CO_2, the economic performances of the static coal facilities are better than those of the feedstock flexible facilities. Increasing the CO_2 credit price reduces the NPV of both type of facilities. At higher CO_2 credit prices (40 €/t CO_2), the flexible facilities can, contrary to the static facility, switch their feedstock to TOPS. This reduces the decrease in NPV and, at even higher CO_2 credit prices (>50 €/t CO_2), improves the NPV of the flexible facilities. At a CO_2 credit price of 50 €/t CO_2, all three feedstock flexible facilities perform better than their static counterparts. The biomass share increases from almost zero at 15 €/t CO_2 to 25% at 40 €/t CO_2, 50% at 50 €/t CO_2, and reaches 85% at 65 €/t CO_2.

To determine the impact of changing TOPS and coal prices on the economics of IG-PG facilities, the same analysis was performed for the other scenarios. The biggest differences were observed under the *Green* and *Direct Action* scenarios (see Figure 3.18).

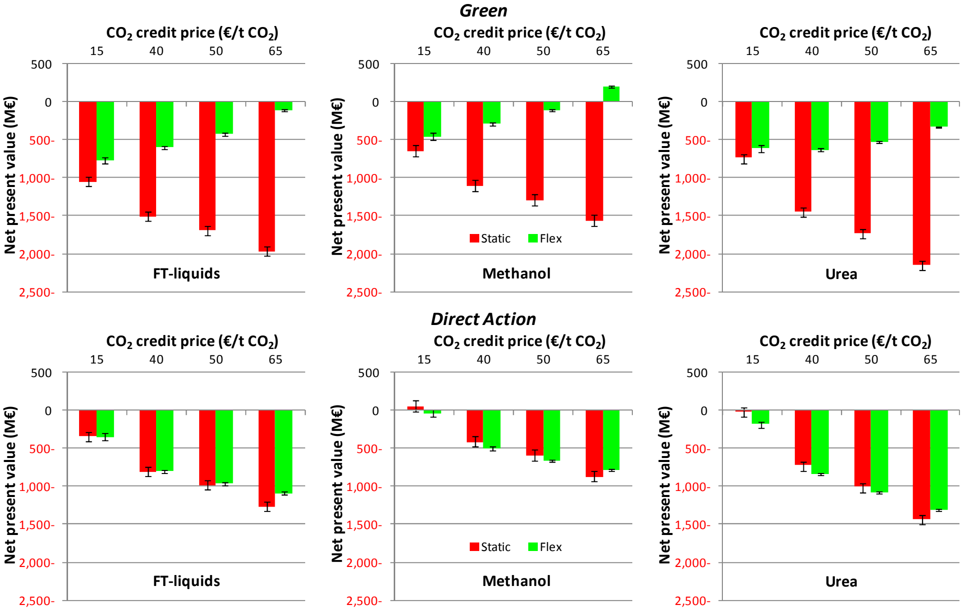

Figure 3.18 NPV of static coal facility (left columns) and full feedstock flexible TOPS/coal facility (right columns) with constant chemical production. Upper figures used *Green* scenario conditions. Lower figures used *Direct Action* scenario conditions. Feedstock prices were varied based on historic data.

The static facilities have a lower NPV in the *Green* scenario and a higher NPV in the *Direct Action* scenario. This is a direct consequence of the different coal prices. The flexible facilities show a different behaviour. At low CO_2 credit prices, coal is predominantly used. Therefore, the NPVs under the *Direct Action* scenario, with its lower coal prices, are better compared to those under the *Current Situation* and *Green* scenarios. With increasing CO_2 credit prices, TOPS is used more often. Therefore, the TOPS price becomes more important and the *Green* scenario has the best NPVs. This already happens at a CO_2 credit price of 40 €/t CO_2.

The prices of TOPS and coal also determine at which CO_2 credit price the switch is made from coal to TOPS. Over the lifetime of the facility, the *Green* scenario already has a 50% biomass share at 15 €/t CO_2. The *Current Situation* scenario needs a CO_2 credit price of 50 €/t CO_2 to reach this share, while the *Direct Action* scenario has a biomass share of only 30% at 65 €/t CO_2.

Another difference between the *Current Situation* and *Green* scenarios is when the NPV of the flexible facilities surpasses those of the static facilities. In the *Current Situation* scenario, this happens at a low CO_2 credit price (≈ 15 €/t CO_2) for the static and feedstock flexible FT-liquids facilities. The methanol and urea facilities need a CO_2 credit price of 40 €/t CO_2. In the *Green* scenario, all flexible facilities have a better NPV than the static facilities, regardless of the CO_2 credit price. For the *Direct Action* scenario, only at a CO_2

credit price of 65 €/t CO_2 are the NPV of the flexible facilities better than those of the static facilities.

Also investigated was how often coal was replaced by biomass and vice versa. By keeping the feedstock substitution frequency as low as possible, the facility can be operated much smoother, resulting in, among others, reduced O&M costs and increased operating time. In the *Current Situation* scenario, the average substitution frequency reached a peak of 1.2 times a day at a CO_2 credit price of 50 €/t CO.

The simulations were also performed for the *Business as Usual*, *Direct Action*, and *Delayed Climate Policy* scenarios. Their results are in between those of the *Current Situation* and *Green* scenarios and are not shown here.

3.4.3 Variation of production

Besides exploiting the daily price variation of electricity by changing the feedstock, it is also possible to change production between chemicals/fuels and electricity. To minimise efficiency losses, both the power and chemical sections had a minimal load factor of 40%, resulting in only minor efficiency losses ($\Delta\eta<1\%$)[150]. We investigated whether flexibility in production pays off. The NPV method was used to compare the economics of the static and production flexible facilities.

3.4.3.1 Case 5. Producing mainly chemicals/fuels during off-peak hours and mainly electricity during peak hours

In this case, the facility was used as a mid-load[13] power plant. Production was switched between chemical/fuel and electricity. A variable market price for the chemical/fuel was assumed using a Gauss distribution based on the historic mean and standard deviation. The electricity was sold according to the fluctuating Dutch day-hourly market prices. The feedstock, CO_2 credit and CO_2 transport and storage costs were based on the scenarios. The resulting NPVs over the lifetime of the facility for the *Current Situation* scenario are given in Figure 3.19 left.

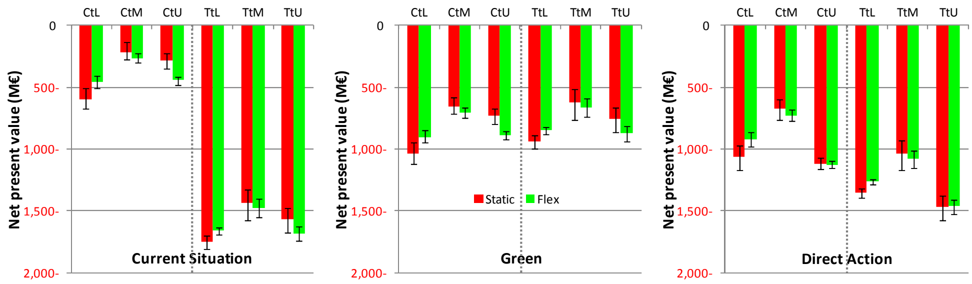

Figure 3.19 NPV of static and production flexible TOPS/coal facilities under *Current Situation* (left), *Green* (middle) and *Direct Action* (right) scenario. Chemical/fuel prices were varied based on historic data. Electricity prices were based on the day-hourly spot market. Other commodity prices were based on the historic data and scenarios.

The flexible facilities have, to satisfy the 40% minimal load constraint, a smaller chemical section and a larger electricity section compared to the static facilities. This results for the FT-liquids facilities in an increase in capital and O&M costs (+1-2%) but also in an increase in fuel and electricity sales (+3%). The net effect is a higher NPV compared to the static facilities (ΔNPV = 138 M€ for coal to FT-liquids (CtL) and 92 M€ for TOPS to FT-liquids (TtL)). The methanol facilities show a similar behaviour, except that the increase in revenues is almost zero (+0%), while the capital costs increase by 3-4%, resulting in a lower NPV for the flexible facilities (ΔNPV = -53 for coal to methanol (CtM) and -36 M€ for TOPS to methanol (TtM)). For the urea facilities it is the other way round. The capital costs decrease (-5%), but also the revenues from the urea and electricity (-8%), resulting in a lower NPV compared to the static facilities (ΔNPV = -159 for coal to urea (CtU) and -114 M€ for TOPS to urea (TtU)).

The CO_2 credit price has no influence on when to operate the flexible FT-liquids (XtL) and methanol (XtM) facilities as a power plant, as the sum of CO_2 emissions and carbon locked in the products remains constant, regardless production mode. As captured CO_2 is used for urea production, the amount of CO_2 emitted and stored by flexible urea facilities is dependent on the production mode and here a reduction in chemical production was observed at higher CO_2 credit prices. This is also the only influence of the different scenarios on the trends found in this case study.

Similar to case 4, also investigated was how often the production mode was switched between chemical/fuel and electricity. The lower the switching frequency, the smoother the facility can be operated, resulting in, among others, reduced O&M costs and increased operating time. In the *Current Situation* scenario, the average switching frequency was 1.8 times a day for the methanol facilities, 1.7 times for the FT-liquids facilities and 1.1 times a day for the urea facilities.

The ability to adjust production made the flexible facilities less vulnerable to fluctuating market prices. The flexible facilities have a variation in NPV which is 40% smaller then the variation of the static facilities. For instance, the CtL facility has an uncertainty of 173 M€ for the static facility, but for the flexible facility this is only 100 M€. This holds for all cases, regardless the feedstock or production mix.

3.4.4 Retrofit

We also analysed the impact of a retrofit during the lifetime of an IG-PG facility on the economic performance. The examined retrofit facility was a FT-liquids facility, producing FT-fuels for 20 years, which was then retrofitted to methanol production, producing methanol for another 10 years. The first step was to calculate the production costs of a static XtL facility operating for 30 years. This was the price at which the FT-liquids of the retrofit facility were sold. Next, the methanol production costs of the retrofit facility were calculated to obtain an NPV of zero. These production costs were compared to the production costs of a static XtM facility operating for 30 years. For the analysis, it was assumed that the retrofit was instantaneously and that excess equipment was sold for 5% of their original value. This analysis was done only for the *Current Situation* scenario.

Table 3.9 Production costs (€/GJ) of static IG-PG facilities operating 30 years under the *Current Situation* scenario. (1)

	Product	EP	EP/Coal	Coal	TOPS/Coal	Coal
XtL facility	FT-fuels	20.9	16.1	12.0	15.2	18.4
XtM facility	MeOH	22.7	16.1	10.2	14.8	19.6
Retrofit facility	MeOH	25.6	18.7	12.7	17.5	22.3

(1) *Current Situation* scenario conditions: 2.25 €/GJ coal; 6.3 €/GJ biomass pellet; 10 €/t CO_2 for transport and storage; 15.7 €/GJ co-generated electricity; 2000 MW_{th} coal eq. input; 4% O&M costs; 10% discount rate.

Because the newly installed equipment for methanol synthesis was depreciated over 10 years only, instead of the 30 years for the XtM facility, this leads to an increase in methanol production costs of around 2.7 €/GJ (≈60 €/t MeOH). Such an increase in methanol market price occurred several times between 2000-2009, where 13% of the time the methanol market prices were at least 2.7 €/GJ above the average market price, all of which occurred in the period 2005-2010. Therefore, depending on product price development, such retrofits of IG-PG facilities can be attractive.

3.4.5 Sensitivity analysis

A sensitivity analysis was performed on a coal to FT-liquids, a TOPS to FT-liquids, and a feedstock flexible TOPS/coal to FT-liquids facility. The latter facility could switch feedstock between pure coal and pure TOPS and had a constant chemical production. For the sensitivity analysis, the *Current Situation* scenario was used. The varied parameters and their ranges are given in Table 3.10. The assumptions and uncertainties regarding the technical model of IG-PG facilities, e.g., gasifier limitation, fixed H_2:CO ratio and model validation, are discussed in Meerman et al., and are not discussed here.[150] Due to the different feedstocks and operation mode, the amount of FT-liquids produced differed per facility. The static coal-fired facility produced 28 PJ_{HHV} FT-liquids per year, while the static TOPS-fired and flexible TOPS/coal-fired facilities produced 23 PJ_{HHV} FT-liquids per year.

Table 3.10 Input sensitivity analysis

Parameter	Unit	Normal	Range
Capital costs (1)	%	100	50% - 150%
Discount rate	%	10	5% - 15%
CO_2 credit price	€/t CO_2	15	0 - 100
Coal price (2)	€/GJ	2.25	1.0 - 6.3
Biomass pellet price (3)	€/GJ	6.3	3.0 - 7.3
Electricity price (4)	€/GJ	15.7	0 - 78.7

(1) The specific capital costs of the three main sources used for capital costs data varies between 76-120%. A slightly larger range was used due to the volatility and uncertainty of capital costs data.

(2) In the past years, coal prices varied between 25-140 €/t, or 1.0-5.6 €/GJ. The upper range is extended to 6.3 €/GJ to see the effect of a coal price which is similar to the biomass pellet price.

(3) In the past years, biomass pellet prices varied between 109-139 €/t, or 5.7-7.3 €/GJ. Several studies point out that biomass pellet prices may drop to 3.0 €/GJ in the future, due to e.g., technical learning, development of the market and introducing best practice agriculture.[30-32]

(4) In the past years, the Dutch day-hourly market price of electricity varied between 0.01-292 €/GJ. An upper limit of five times the current electricity price was taken.

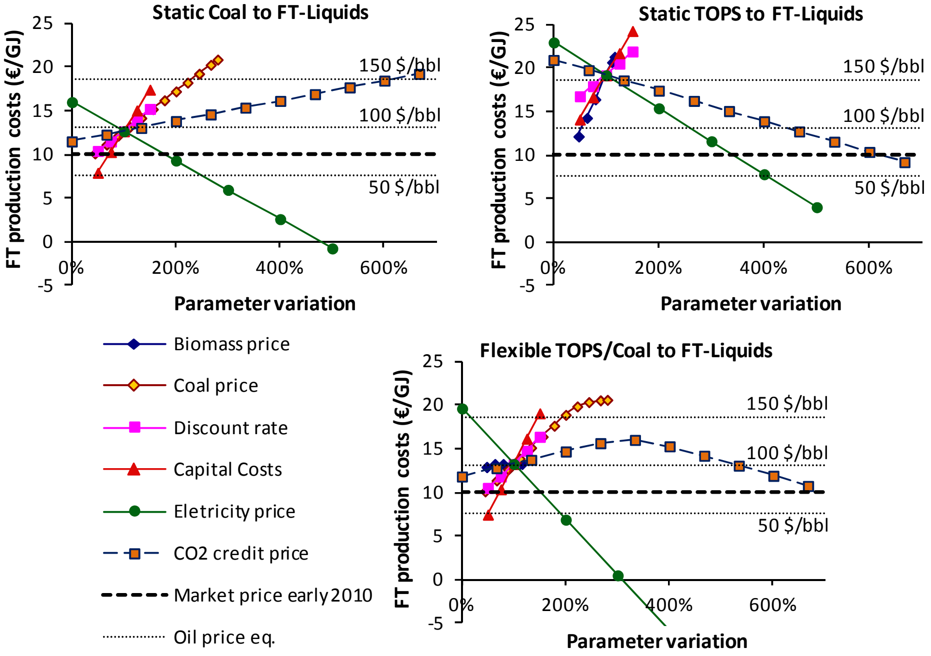

Figure 3.20 Sensitivity analysis of static coal (upper left), static TOPS (upper right) and feedstock flexible TOPS/coal (bottom) to FT facility.

Results (Figure 3.20) show that in all three cases, production costs of the FT-liquids are very sensitive to the capital costs. The electricity price also has a large impact, especially for the flexible facility. This was expected as the flexible facility produces much more electricity than the static facilities.

The feedstock prices show a linear relation with the production cost for the static facilities. This does not apply for the flexible facility. A lower biomass pellet price has, at first, almost no effect on FT-liquids production costs as it is still more profitable to use almost only coal. Only at low biomass pellet prices it becomes profitable to use biomass. At that point (<4 €/GJ biomass pellet), FT-liquids production cost starts to decrease. For the coal prices the opposite holds. When the coal price increases, coal initially remains the most economical feedstock, resulting in increasing production costs. Only at higher coal prices does biomass becomes more economical. At that point (>4.5 €/GJ coal), the increase in production costs slows down, until only biomass is used, at which point (>7 €/GJ coal) production costs remain constant.

As expected, CO_2 credit price has a strong increasing effect on FT-liquids production cost when using coal (+0.8 €/GJ per 10 €/t CO_2) and a decreasing effect when using TOPS (-1.2 €/GJ per 10 €/t CO_2). Production cost of the flexible facility initially increases with higher CO_2 credit price, but above 50 €/t CO_2, production costs drop again as biomass replaces coal as the preferred feedstock.

3.5 Discussion & Conclusions

This study focuses on the economic implications of flexible operation of state-of-the-art gasification facilities, thereby complementing the technical assessment performed in part A of this analysis. The main advantage of flexible IG-PG facilities over their static counterparts is their ability to react to variations in commodity prices and thus to maximise profits. Historical data clearly show strong fluctuations in the commodity prices. It is expected that these fluctuations will continue in the future and may become even more extreme due to increasing mismatches between supply and demand. Designing a coal-fired facility also allowing the processing of TOPS would increase capital costs by 0.5%. The combination of small additional investments and high uncertainty in commodity prices makes flexible IG-PG facilities attractive. Replacing coal for biomass, even when it results in lower electricity production, becomes attractive when the CO_2 credit price exceeds 40 €/t CO_2, the coal price exceeds 4 €/GJ or the biomass pellet price is below 4 €/GJ. In the last decade, coal prices have already exceeded this price. Furthermore, if changing market conditions are insufficient to make flexible IG-PG facilities more economical than their static counterparts, the ability of the flexible facilities to function as mid-load power plant may justify their deployment as they could prevent the construction of dedicated mid-load power plants.

Recently, a few studies assessing the techno-economic performance of gasification facilities have become available. These studies indicate that electricity could be produced from a gasification facility equipped with CCS for 18-21 €/GJ[53,54] if coal is used and for 37 €/GJ[53] if biomass is used as feedstock. In this study, these production costs were determined at 18 and 29 €/GJ respectively. The higher production cost of converting biomass to electricity in the NETL study is most likely a result of the assumed relative small scale of the facility. The literature studies also indicate FT-liquids can be produced for 8-9 €/GJ[55,57] if coal is used and for 19 €/GJ[57] if biomass is used. In this study, production costs of respectively 13 and 22 €/GJ were found. The lower feedstock prices used in these literature studies (1.0-1.6 €/GJ for coal and 3.5 €/GJ for biomass vs. 2.25 €/GJ for coal and 6.3 €/GJ for biomass used in this study) are the main reason of the lower production costs.

Static IG-PG facilities have production costs between 12 €/GJ (coal to methanol) and 38 €/GJ (EP to urea) under *Current Situation* scenario conditions. Using coal results in the lowest production costs, while the use of EP results in the highest production costs. This is mainly due to lower feedstock costs and higher product output related to coal. None of the static facilities have production costs under the market value of the products. When looking at the biomass fired facilities, the lowest production costs are under the *Green* scenario, mainly due to the lower biomass pellet price. The *Direct Action* scenario also has relatively low production costs as here the increase in CO_2 credit price offsets the increase in biomass pellet price.

At the current prices of 2.25 €/GJ coal, 6.3 €/GJ TOPS and 15 €/t CO_2, FT-liquids can be produced for 13 €/GJ from coal, equivalent to 96 $/bbl, and 19 €/GJ from TOPS, equivalent to 155 \$/bbl. To match the profitability of a coal based FT-liquids facility, TOPS prices need to drop to about 3.3 €/GJ. However, when comparing the profitability with oil-based gasoline/diesel, the break-even point at an oil price of 144 \$/bbl - 2008's peak - is already reached at 5.7 €/GJ TOPS.

The share of the capital costs on the production costs of static IG-PG facilities is 28-38% if biomass is used and 41-48% if coal is used. If a coal-fired facility is designed to also process EP, capital costs increase by 3-5%, which is higher than the 0.5% estimated for TOPS. The main reason for this difference is that TOPS can use the existing coal feeding system, while EP requires a dedicated feeding system. In the IG-PG facilities, the gasifier is the most capital intensive equipment, except for the urea facilities, where the urea section is the most capital intensive. As the capital costs make up a large share of the production costs, IG-PG facilities are very sensitive to changes in capital costs. To make static coal based facilities profitable, capital costs need to drop by 10-27%. These values are within the uncertainty of the used method. However, this drop may also be obtained by technological learning.[168] For biomass, the required reductions in capital costs are 60-96%. This is well above the methodological uncertainty and to make biomass based facilities profitable will, in addition to cost reductions due to learning, most likely also require lower biomass pellet prices or higher CO_2 credit prices.

The large difference in production costs when comparing coal-fired facilities with biomass-fired facilities is mainly the feedstock price. Under the market conditions of early 2010, feedstock prices account for 23-30% of the production costs in coal-fired facilities, while in biomass-fired facilities feedstock prices account for 46-57% of the production costs. A CO_2 credit price can be used to compensate this difference. The break-even CO_2 credit price for biomass pellets and coal for static facilities was calculated at around 50 €/t CO_2 for TOPS and 55 €/t CO_2 for EP. When looking at feedstock flexible facilities and considering the fluctuations in electricity price, biomass pellets are sometimes already the most preferable option at a CO_2 credit price of around 40 €/t CO_2. At a CO_2 credit price above 60 €/t CO_2 biomass pellets are calculated to be almost always the preferred feedstock.

The biomass pellet prices used in this study reflect current market prices. However, several studies indicate that technological learning, mature markets and local pre-treatment could drop biomass pellet prices considerably, e.g., to 3-4 €/GJ in the medium term.[32,85,157,158] At this price, coal is the favoured feedstock for about 50% of the time. At a CO_2 credit price of 30 €/t CO_2 for TOPS and 50 €/t CO_2 for EP biomass will replace coal almost completely as the preferred feedstock.

This study indicates that CO_2 capture and storage adds 2.8-6.4 €/GJ to the production costs: 1.3-3.3 €/GJ for capture, 0.5-1.4 €/GJ for compression and 0.9-2.1 €/GJ for transport and storage. As CO_2 capture is mandatory for chemical production, the CO_2 compression, transport and storage costs are already offset by a CO_2 credit price of 15 €/t CO_2. It should be noted, however, that a CO_2 transport and storage costs of 10 €/t CO_2 was assumed. The actual costs could be higher depending on, for instance, reservoir availability and characteristics, pipeline availability and distance between IG-PG facility and reservoir site.[169]

Changing market conditions have a large impact on the economics of IG-PG facilities. With the current market conditions, a production flexible FT-liquids facility is already more profitable than its static counterpart by exploiting the hourly variation in electricity price. For facilities which are feedstock flexible, the CO_2 credit price needs to increase to above 40 €/t CO_2 before flexibility pays off. The economics of the flexible facilities depend heavily on the feedstock and CO_2 credit prices. The larger these prices vary, the better flexible facilities perform compared to static facilities. In this scenario it was assumed that switching between coal and biomass is simultaneously. Although this is technically possible, the facility must be designed for it by, for instance, having separate feeding trains for the coal and biomass.[170]

Finally, the results indicate that retrofitting an IG-PG facility producing FT-liquids towards methanol production increases methanol production costs by around 2.7 €/GJ compared to an IG-PG facility that continuously produces methanol. However, if the dominant transportation fuel switches from gasoline and diesel to methanol, the methanol market price could well rise above this value.

It should be noted that innovative processes that increase overall efficiency and decrease capital costs of the facilities are currently being developed, like membrane processes, hot gas cleaning and fuel cells.[168,171,172] Further research is required to determine how these processes may affect plant economics of both static and flexible IG-PG facilities.

The technical analysis, performed in part A, showed that flexible IG-PG can technically operate as mid-load power plants.[150] Integrating these facilities to the national power grid could result in added benefits. The extent of these benefits requires further research.

3.6 Acknowledgement

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3.7 Supplementary data

3.7.1 Commodity price projections in literature

Several organisations have made projections on future energy prices under different political climates. The price assumptions for these projections are displayed in Table 3.11. The data in the table were used as basis for the price behaviour of the feedstocks and CO_2 of the different scenarios. Note that the different reference scenarios (WETO-ref, WEO-ref and PBL-ref) all assume a continuation of current policy and technological progress.[43,69,151,152,153,154]

The sustainable scenarios of IEA (WEO-450), PBL (PBL-450 and PBL-Delay) and Greenpeace (ER) have CO_2-eq. emissions drop to 450 ppm CO_2-eq. in 2050. The difference between the PBL-450 and delay scenario is that in the 450 scenario measures are taken in 2020, while in the delay scenario measures are not taken till 2030. The WETO-H_2 has two scenarios where atmospheric CO_2 concentration reach 460 ppmv in 2050, a normal scenario (carbon constraint case) and a scenario with increased emphasis on H_2 (hydrogen case).[43,69,151,152,153,154]

For the CO_2 credit price the rate was rounded to 2.5%/yr for the *Business as Usual* scenario and 10%/yr for the *Direct Action* scenario. The *Delayed Climate Policy* scenario was altered to 2.5%/yr up till 2025 and 20%/yr from 2025 onwards.

Table 3.11 Feedstock and CO_2 price assumptions used in various studies ($€_{2008}$/GJ or $€_{2008}$/t CO_2).[43,69,151,152,153,154]

Commodity (1)	Scenario (2)	2008	2010	2015	2020	2025	2030	2040	2050	%/yr (3)
Natural Gas	WETO-Ref		3.1		5.6		7.3		12.4	3.5%
	WETO-H_2		3.2		5.6		7.3		11.3	3.2%
	WETO-CCC		3.2		5.6		7.3		11.3	3.2%
	ER		7.4		11.3		13.1	15.0	17.7	2.2%
	WEO-Ref	6.7		6.7	7.8	8.4	9.0			1.4%
	WEO-450	6.7		6.7	7.1	7.1	7.1			0.3%
	PBL-Ref	5.2	3.8	3.7	3.7	3.8	3.9	4.3	4.8	-0.2%
	PBL-450	5.2	3.8	3.7	3.7	3.8	3.8	4.1	4.6	-0.3%
	PBL-Delay	5.2	3.8	3.7	3.7	3.8	3.9	4.2	4.8	-0.2%
Oil	WETO-Ref		4.7		6.6		8.2		13.5	2.7%
	WETO-H_2		4.9		6.8		8.4		12.3	2.3%
	WETO-CCC		4.9		6.8		8.4		12.3	2.3%
	ER		9.6	12.3	14.5		16.7	16.7	16.7	1.4%
	WEO-Ref	10.8		9.6	11.1	11.9	12.8			0.8%
	WEO-450	10.8		9.6	10.0	10.0	10.0			-0.3%
	PBL-Ref	10.9	7.2	6.1	6.2	6.2	6.2	6.6	6.9	-1.1%
	PBL-450	10.9	7.2	6.1	6.2	6.2	6.0	6.1	6.0	-1.4%
	PBL-Delay	10.9	7.2	6.1	6.2	6.2	6.1	6.3	6.2	-1.3%

Coal	ER		3.0	2.9	3.4		3.6	4.0	4.3	0.9%
	WEO-Ref	3.0		2.3	2.6	2.7	2.8			-0.4%
	WEO-450	3.0		2.2	2.0	1.8	1.6			-2.8%
	PBL-Ref	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	-0.1%
	PBL-450	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	-0.1%
	PBL-Delay	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.5	-0.1%
Biomass	ER		5.2	5.6	6.3		6.8	7.0	7.1	0.8%
	PBL-Ref	3.7	3.6	3.6	3.6	3.5	3.5	3.6	4.0	0.2%
	PBL-450	3.7	3.6	3.6	3.6	3.9	4.7	6.6	6.2	1.3%
	PBL-Delay	3.7	3.6	3.6	3.6	3.6	4.2	5.9	6.2	1.3%
CO_2	ER		6.8		13.6		20.4	27.2	34.0	4.1%
	WEO-Ref				29.3		36.7			2.3%
	WEO-450				34.0		74.8			8.2%
	PBL-Ref (4)	0	0	0	0	0	0	0	0	0%
	PBL-450 (4)	0.4	0.7	6.8	67.2	110.2	117.3	114.6	158.7	15.0%
	PBL-Delay (4)	1.5	2.6	12.1	23.7	43.9	45.7	97.2	260.7	13.0%
Electricity (5)	PBL-Ref	15.8	15.7	15.2	14.8	14.3	14.1	13.6	13.2	-0.4%
	PBL-450	16.0	15.9	15.8	19.5	22.3	22.4	17.9	15.8	0.0%
	PBL-Delay	16.0	15.9	16.2	16.8	17.6	17.7	17.9	17.2	0.2%

(1) The following conversion factors were used: 1t coal = 27 GJ; 1 bbl (equivalent) = 6.12 GJ; 1 t C = 3.66 t CO_2.

(2) WETO-ref, -H_2 and -CCC are respectively the reference, H_2 economy and carbon constraint case scenarios of the World Energy Technology Outlook 2050; ER is the scenario of the Energy (r)evolution from Greenpeace; WEO-Ref and -450 are respectively the reference and 450 ppm scenarios of the World Energy Outlook 2009. PBL-Ref, -450 and -Delay are respectively the reference, 450 ppm and 450 ppm with delayed action of the Netherlands Environmental Assessment Agency (PBL). See text above for more details.

(3) This is the average annual price change of the commodity over the longest possible time-span.

(4) The CO_2 price is calculated based on the measures needed to reduce CO_2 emissions. It is, therefore, not a prediction of how CO_2 prices will evolve, but of which price is needed to reduce CO_2 emissions sufficiently in an ideal world. As the reference has no CO_2 target, there is no CO_2 price.

(5) The electricity price is the production costs. Therefore, it includes the CO_2 tax of fossil fuels.

3.7.2 *Detailed technical and economic data of the static IG-PG facilities*

Table 3.12 Technical and economic data static IG-PG facilities under the *Current Situation* scenario. (1)

		Electricity					FT-liquids					Methanol					Urea				
Feedstock (2)		EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS
Total Capital Costs	M€	1,245	1,369	1,489	1,405	1,296	1,391	1,534	1,693	1,599	1,461	1,241	1,360	1,472	1,391	1,286	1,820	2,033	2,253	2,123	1,962
Feedstock consumption																					
EP	PJ/yr	42	24				42	24				42	24				42	24			
TOPS	PJ/yr				26	48				26	48				26	48				26	48
Coal	PJ/yr		24	57	26			24	57	26			24	57	26			24	57	26	
Coal based C	kt C/yr		568		617			568	1,333	617			568	1,333	617			568	1,333	617	
Biomass based C	kt C/yr	1,071	623		656	1,201	1,071	623		656	1,201	1,071	623		656	1,201	1,071	623		656	1,201
Production																					
Chemical/fuel	PJ/yr						18	23	28	26	23	12	15	18	17	15	11	13	17	15	13
Slag	kt/yr	11	93	203	110	30	11	93	203	110	30	11	93	203	110	30	11	93	203	110	30
Sulphur	kt/yr	1	23	52	25	0	1	23	52	25	0	1	23	52	25	0	1	23	52	25	0
Electricity	PJ/yr	16	19	23	21	19	5	5	6	6	5	9	10	12	11	10	9	10	12	11	11
CO_2 emitted	kt C/yr	375	472	590	534	473	55	70	90	79	70	170	215	271	243	215	373	470	588	532	472
CO_2 in product	kt C/yr						317	398	495	451	400	197	247	306	279	248	200	251	315	285	253
CO_2 for storage	kt C/yr	693	718	743	738	725	693	718	743	738	725	693	718	743	738	725	494	467	428	453	473
Costs																					
Capital Recovery	M€	1,245	1,369	1,489	1,405	1,296	1,391	1,534	1,693	1,599	1,461	1,241	1,360	1,472	1,391	1,286	1,820	2,033	2,253	2,123	1,962
O&M	M€	424	466	507	478	441	474	522	577	545	498	423	463	501	474	438	620	692	767	723	668
Feedstock costs	M€	2,228	1,761	1,087	1,913	2,582	2,228	1,761	1,087	1,913	2,582	2,228	1,761	1,087	1,913	2,582	2,228	1,761	1,087	1,913	2,582
CO_2 credits @ 15 €/t CO_2	M€	-324	-70	276	-56	-339	-324	-70	276	-56	-339	-324	-70	276	-56	-339	-231	47	424	77	-221
CO_2 trans.& storage	M€	216	224	232	230	226	216	224	232	230	226	216	224	232	230	226	154	146	134	141	147

Revenues																						
Main product @ 15 €/t CO_2		M€	3,789	3,731	3,546	3,949	4,205	427	361	260	383	474	345	260	134	267	373	429	357	246	372	470
Electricity co-generated		M€						3,328	3,236	3,022	3,448	3,697	2,628	2,376	1,945	2,462	2,834	3,390	3,261	2,979	3,422	3,720
Sulphur		M€	1	19	45	21	0	1	19	45	21	0	1	19	45	21	0	1	19	45	21	0
CO_2 credits		kt C	1,245	1,369	1,489	1,405	1,296	-1,016	-554	90	-577	-1,132	-900	-409	271	-413	-986	-698	-153	588	-124	-729
CO_2 credits		kt CO_2						-3,722	-2,029	331	-2,114	-4,147	-3,300	-1,498	992	-1,514	-3,614	-2,556	-562	2,155	-454	-2,673
Production costs [3]	**@ 0 €/t CO_2**	**€/GJ**	**31**	**24**	**17**	**23**	**28**	**24**	**17**	**11**	**16**	**21**	**29**	**19**	**11**	**18**	**25**	**40**	**29**	**18**	**26**	**35**
	@ 15 €/t CO_2	**€/GJ**	**29**	**23**	**18**	**22**	**26**	**22**	**17**	**13**	**16**	**19**	**26**	**19**	**12**	**17**	**22**	**38**	**29**	**21**	**27**	**33**
	@ 50 €/t CO_2	**€/GJ**	**23**	**22**	**22**	**21**	**21**	**17**	**16**	**15**	**15**	**15**	**18**	**17**	**16**	**16**	**16**	**32**	**30**	**28**	**28**	**28**
	@ 100 €/t CO_2	**€/GJ**	**15**	**21**	**26**	**20**	**14**	**10**	**15**	**19**	**14**	**9**	**8**	**16**	**22**	**15**	**7**	**23**	**31**	**38**	**30**	**22**

(1) *Current Situation* scenario conditions: 2.25 €/GJ coal; 6.3 €/GJ biomass pellet; 10 €/t CO_2 for transport and storage; 15.7 €/GJ co-generated electricity; 2000MW_{th} coal eq. input; 10% discount rate; 20 year lifetime.

(2) The used feedstocks are Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal. If multiple feedstocks are used, their ratio is 50/50 on an energy basis.

(3) The production costs refer to the product listed at the top of the table. If electricity is co-generated, it was sold for the price listed in Table 3.3. The production costs were calculated at three different CO_2 taxes. These taxes are related to the net CO_2 emissions of the IG-PG facility.

Table 3.13 Capital cost breakdown static IG-PG facilities.

	Electricity					**FT-liquids**					**Methanol**					**Urea**				
Feedstock (1)	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS	EP	EP/ Coal	Coal	TOPS/ Coal	TOPS
Pre-treatment																				
Biomass	65	39		37	64	65	39		37	64	65	39		37	64	65	39		37	64
Coal		54	111	57			54	111	57			54	111	57			54	111	57	
Gasification																				
Air separating unit (95% O_2)	164	176	192	178	164	164	176	192	178	164	164	176	192	178	164	164	176	192	178	164
O_2 compressor	16	18	20	18	16	16	18	20	18	16	16	18	20	18	16	16	18	20	18	16
N_2 compressor	13	16	20	18	16	2	3	4	3	3	3	3	4	4	3	0	0	0	0	0
Shell EF gasifier	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368	368

Gas Cleaning																				
Candle filter	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6	6
Wet Scrubber	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9	9
High Temp HX	4	5	5	5	5	4	5	5	5	5	4	5	5	5	5	4	5	5	5	5
Sour shift	26	30	35	33	31	26	30	35	33	31	26	30	35	33	31	26	30	35	33	31
Guard bed	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Sweet shift	0	0	0	0	0	2	2	2	2	2	15	18	21	19	18	22	26	30	28	26
H_2 PSA	0	0	0	0	0	2	3	3	3	3	14	16	18	18	16	20	22	25	24	23
AGR																				
Rectisol AGR removal	169	173	176	176	174	169	173	176	176	174	169	173	176	176	174	169	173	176	176	174
Claus/SCOT	3	27	48	29	2	3	27	48	29	2	3	27	48	29	2	3	27	48	29	2
Power Isle																				
Gas Turbine	90	106	125	116	106	23	27	33	30	28	47	55	65	61	55	56	66	77	72	66
HRSG	118	131	148	140	131	95	103	113	108	102	101	111	123	116	110	108	119	133	126	119
Steam Turbines & cycle	59	63	67	65	62	50	52	54	53	52	52	54	57	55	54	54	56	59	58	56
FT Isle																				
FT Slurry Reactor						139	163	207	194	164										
Product Upgrading						91	106	124	116	107										
MeOH Isle																				
MeOH Reactor											38	43	49	47	43					
Product Upgrading											6	6	7	7	7					
Urea Isle																				
Ammonia Reactor																274	318	369	345	319
Urea Reactor																274	318	369	345	319
CCS																				
CO_2 Compression & drying	18	19	19	19	19	18	19	19	19	19	18	19	19	19	19	18	19	19	19	19
Total	1,128	1,241	1,351	1,273	1,174	1,253	1,384	1,530	1,443	1,317	1,124	1,233	1,336	1,261	1,165	1,649	1,843	2,042	1,924	1,778

(1) The used feedstocks are Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and Illinois #6 coal. If multiple feedstocks are used, their ratio is 50/50 on an energy basis.

4 Technical and economic prospects of coal- and biomass-fired IG-PG facilities equipped with CCS over time.

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Submitted

Abstract

This study analyses the impacts of technological improvements and increased operating experience on the techno-economic performance of integrated gasification polygeneration facilities (IG-PG). The IG-PG facilities investigated produce electricity (IGCC) or FT-liquids (IG-FT).

Results suggest that a state-of-the-art (SOTA) coal-fired IGCC without CO_2 capture has electricity production costs of 17 €/GJ with the potential to decrease to 11 €/GJ in the long term. Specific direct CO_2 emissions may drop from about 0.71 to 0.59 kg CO_2/kWh. If CO_2 is captured, production costs may increase to 23 €/GJ, with the potential to drop to 14 €/GJ in the long term. As a result, CO_2 avoidance costs would decrease from 35 €/t CO_2 to 18 €/t CO_2. The efficiency penalty due to CCS may decrease from 8.8%pt to 2.9%$_{pt}$.

CO_2 emissions can also be reduced by using torrefied biomass (TOPS) instead of coal. Production costs of a SOTA TOPS-fired IGCC without CO_2 capture are 25 €/GJ. In the long term, this may drop to 19 €/GJ, resulting in CO_2 avoidance costs of 45 €/t CO_2. If TOPS is available at 3 €/GJ, production costs could drop to 12 €/GJ and avoidance costs to 7 €/t CO_2. The greatest reduction in anthropogenic CO_2 emissions is obtained by using biomass combined with carbon capture and storage (CCS). A SOTA TOPS-fired IGCC with CCS has, depending on the biomass price, production costs of 25-35 €/GJ with CO_2 avoidance costs of 19-40 €/t CO_2. These values may decrease to 15 €/GJ and 12 €/t CO_2 avoided in the long term. As carbon from biomass is captured, specific direct CO_2 emissions are negative and estimated at -0.93 kg CO_2/kWh for SOTA and -0.59 kg CO_2/kWh in the long term.

New technologies in IG-FT facilities have a slightly smaller impact on production costs. In the long term, production costs of FT-liquids from coal may drop from 13 €/GJ to 9 €/GJ if CO_2 is vented and from 15 €/GJ to 10 €/GJ if CCS is applied. The use of TOPS results in 15-23 €/GJ (Vent) and 17-24 €/GJ (CCS) for SOTA facilities. These production costs may drop to 11-18 €/GJ (Vent) and 12-19 €/GJ (CCS) in the long term. Contrary to the IGCC cases, the coal-fired IG-FT facility shows the lowest CO_2 avoidance costs. The CO_2 emission of coal to FT-liquids with CCS is, however, similar to gasoline/diesel production from crude oil.

4.1 Introduction

To significantly reduce global CO_2 emissions decarbonisation of both the transport and power sector is required.[34] Integrated gasification poly-generation (IG-PG) facilities producing electricity and FT-liquids can potentially decarbonise both sectors by applying carbon capture and storage (CCS) and using biomass as feedstock. Being able to use biomass as well as coal means that these facilities can play a role in the transition towards a renewables based energy infrastructure.

In previous research we examined the economic and technical potential of state-of-the-art (SOTA) IG-PG facilities.[150,173] Our results show that coal and biomass can be converted into electricity at 38-40% efficiency and FT-liquids at 55-60% efficiency. Using torrefied wood pellets (TOPS) results in improved technical and economic performance compared to conventional wood pellets. Also, it was shown that with SOTA technology neither electricity nor FT-liquids can be produced competitively. Advanced technologies and technological learning can, however, bring production costs down. This may make IG-PG facilities profitable in the longer term. Therefore, in this study the impact of potential technological and operational improvements on the technical and economic performance of IG-PG facilities is assessed.

The impact of future technologies on the performance of existing processes has been studied before. For instance, NETL performed a detailed analysis on the potential technological and operational improvements of an IGCC, both with and without CO_2 capture. They calculated that an IGCC using 2002-technology[16] and without CCS has electricity production costs of 66 €/MWh. Advances in technology and availability may reduce these costs to 44 €/MWh, among others because of an increase of the energy conversion efficiency from $35.4\%_{HHV}$ to $46.1\%_{HHV}$. They also suggest that, incorporating solid oxide fuel cells (SOFCs) increase the efficiency considerable ($58.8\%_{HHV}$), but also increase production costs (45 €/MWh).[59] The same analysis was performed for a slurry-fed, CO_2 capture IGCC using 2002-technology[17]. Technological and operational improvements may reduce production costs from 81 €/MWh to 52 €/MWh, while increasing the efficiency from $30.4\%_{HHV}$ to $40.0\%_{HHV}$. Incorporating SOFCs result in production costs of 53 €/MWh at an efficiency of $56.3\%_{HHV}$.[58] The reference system in these studies is very similar to the reference system used in our study. The NETL studies, however, use a slurry-fed gasifier, which is unsuitable for biomass as the caloric value of biomass is too low. In our study, a dry-fed gasifier is used instead.

The potential improvement in FT-liquids production has been studied by Tijmensen et al., indicating hot gas cleaning in a dedicated biomass-fired IG-FT could reduce production

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[16] The analysed IGCC consists of a single-stage slurry feed gasifier with radiant-only gas cooler, Selexol acid gas removal, 7FA syngas turbine, conventional three-pressure level steam cycle and cryogenic ASU. The availability is 75%.

[17] Compared to the non-capture IGCC, the reference CO_2-capture IGCC includes a WGS reactor and CO_2 compression section. The AGR was replaced by a 2-stage Selexol system and the 7FA gas turbine was modified to run on a H_2-rich mixture. The availability was increased to 80%.

costs by 10%.[174] Eerhart et al., calculate that future improvements, like hot gas cleaning, diesel selective FT-catalyst and more efficient gas and steam turbines, could reduce FT-liquids production costs by 20% when using coal and 16% when using biomass as feedstock.[172]

Although the previous studies investigated the potential of new technologies for coal-fired IGCC and IG-FT, the effect of firing TOPS in coal-based facilities has not yet been investigated. Also, the studies were not explicit about a time period at which the various technologies are expected to become commercially available. Our study aims to fill this gap by identifying the potential improvements for these facilities for different time periods and compare the results when using different feedstocks. This allows for a quantification of the possible performance improvements over time. In addition, the effect of carbon capture and storage on overall performance is analysed. For this purpose, an overview of the potential of present and future technologies was made. Based on this overview, four process layouts, for current (2010), short (2015), mid (2030) and long (2050) term, were made.

4.2 Integrated poly-generation gasification facilities

4.2.1 Process description

In an IG-PG facility (Figure 4.1), a solid carbon-containing feedstock is fed into an entrained flow (EF) gasifier. The high operating temperatures (>1500°C) result in a syngas consisting mainly of CO, CO_2, H_2 and H_2O. The required heat is supplied by combusting part of the feedstock by adding a sub-stoichiometric amount of oxygen, supplied by an air separation unit (ASU). Pure oxygen instead of air is used to obtain the required high temperatures, to increase overall efficiency and to reduce the size of downstream equipment.[50,47] The syngas is cooled and cleaned of contaminates. Depending on the desired product, the H_2:CO ratio of the syngas is adjusted in a water-gas shift (WGS) reactor. This can be done before or after the acid gas removal (AGR). When producing electricity, the syngas is fed into a gas turbine and combusted. When producing FT-liquids, the syngas is fed into a FT-reactor. The liquids are purified and any off-gas is fed into a gas turbine and combusted. To increase the overall economics of an IG-PG, steam is generated at various locations and used for electricity production in steam turbines. To lower the CO_2 emissions of the facility, CO_2 can be captured at the AGR, compressed and subsequently stored in underground geological reservoirs.

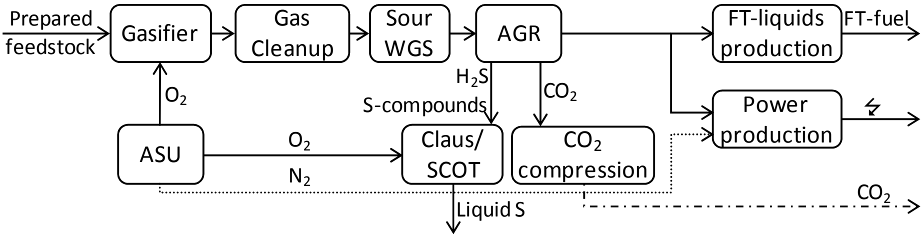

Figure 4.1 Simplified process layout of a flexible IG-PG facility using SOTA technology. Waste, heat and recycle streams are not displayed.[150]

Detailed information of the individual components, i.e., ASU, gas cleanup, gas and steam turbines, can be found in the Appendix, which includes a description of the technologies, their current development status, operating conditions as well as other relevant process improvements.

4.3 Configurations

Based on the commercially available technologies described in the Appendix, plant configurations for the current IGCC and IG-FT facilities, both with CO_2 capture and storage (CCS) and without (Vent), were selected. The time period in which new technologies are expected to become commercially available was selected based on the following criteria:

- Short term (2015-2020): technologies that are currently being tested in large-scale pilot projects;
- Mid term (2020-2035): technologies that have been successfully tested in laboratories and/or that are being tested in small scale pilot projects;
- Long term (2035-2050): technologies that are currently under development at lab scale or are at proof-of-concept stage.

Table 4.1 Key components and process configurations used for the different time periods.

	Current	**Short term**	**Mid term**	**Long term - GT**	**Long term - SOFC**
Feeding	Lock hopper	Lock hopper	Solid feed pump	Solid feed pump	Solid feed pump
Oxygen production	Cryogenic ASU	Cryogenic ASU +	ITM	ITM	ITM
Quench					
- IGCC-Vent	Syngas	Syngas	Syngas	Syngas	Syngas
- IGCC-CCS / IG-FT	Water	Water	Water	Water	Water
Gas cleaning					
- IGCC-Vent	Selexol Claus	Selexol + Claus	TDS & DSRP	TDS & DSRP	TDS & DSRP
- IGCC-CCS	WGS Selexol Claus	WGS Selexol + Claus	SEWGS TDS & DSRP	SEWGS TDS & DSRP	TDS & DSRP
- IG-FT	WGS Rectisol Claus	WGS Rectisol + Claus	Adv. WGS Rectisol ++ Claus	Adv. WGS Rectisol ++ Claus	N.A.
CO_2 compression	Conventional	Conventional	Shock wave	Shock wave	Shock wave
Syngas conversion	Adv. F-GT	F^+-GT	G-GT	H-GT	SOFC & H-GT
HRSG gasifier	IP steam	IP steam	IP steam	HP steam	HP steam
FT-liquids synthesis	Cobalt-based catalyst	Cobalt-based catalyst	Cobalt-based catalyst	Diesel selective catalyst	N.A.

ASU: air separating unit; ITM: ion transfer membrane; TDS: transport desulphurisation; DSRP: direct sulphur recovery plant; WGS: water-gas shift; SEWGS: sorption enhanced water-gas shift; GT: gas turbine; SOFC: solid oxide fuel cell; HRSG: heat recovery steam generation; IP: intermediate pressure; HP: high pressure; N.A: not applicable.

When multiple technologies are available for the same process, the technology with the lowest production costs of the main product was selected. The resulting configurations (see Table 4.1 and Figure 4.2) were modelled in a component-based chemical AspenPlus simulation model.[150,173] This model calculates the relevant mass- and energy balances and, combined with an Excel-based economic model, allows to calculate the production costs.[173]

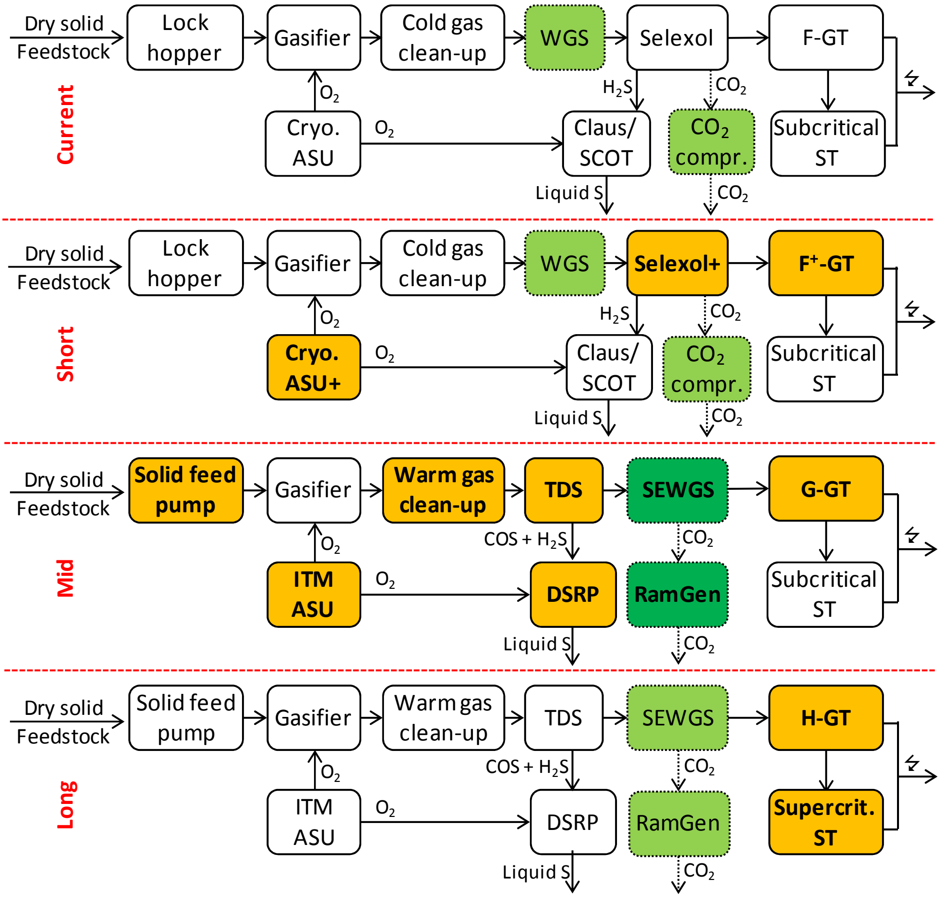

Figure 4.2 Schematic overview of IGCC for the different time periods. New technologies are given in bold. Technologies only required when CO_2 is captured are in green and have a dashed border.

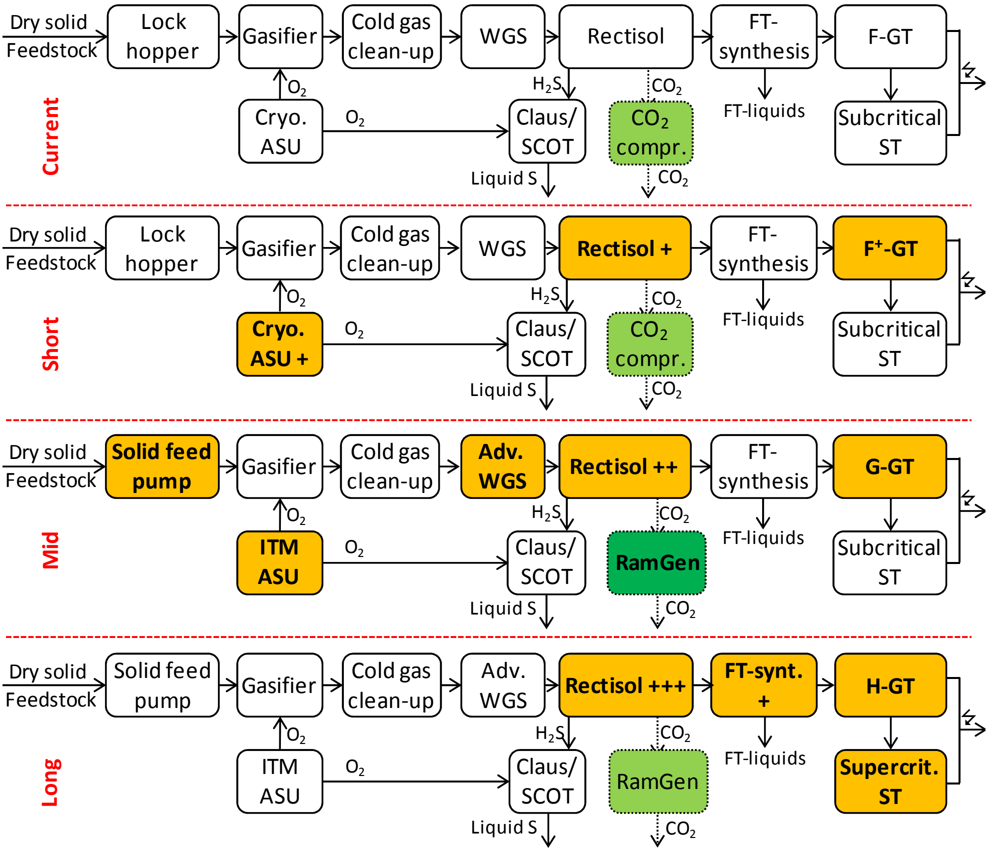

Figure 4.3 Schematic overview of IG-FT for the different time periods. New technologies are given in bold. Technologies only required when CO_2 is captured are in green and have a dashed border.

Below, short descriptions of the configurations in each time period are presented here as well as changes made with respect to the previous time period. For more details see the Appendix as well as Meerman et al., Eerhart et al., Knoope et al., and NETL.[58,59,150,171,172,173] The technical performance of the technologies is summarised in Table 4.2.

Current configurations consist of cryogenic ASU, lock-hopper feeding system, dry-fed Shell EF gasifier, candle filter, wet scrubber, WGS reactor (for CO_2 capture or FT-liquids production), solvent based AGR (Selexol for IGCC or Rectisol for IG-FT), Claus/SCOT, FT-reactor with conventional FT-catalysts (only for IG-FT) and SOTA gas and three pressure steam turbines. If CO_2 is captured, the IG-PG facilities also contain a conventional CO_2 compressor.

In the short term, only gradual improvements to already existing technologies are expected. The improved technologies are cryogenic ASU, solvent-based AGR and the gas turbine.

In the mid term several new technologies can be introduced which require alterations in the overall process configurations compared to SOTA. Common to both facilities is the replacement of the cryogenic ASU with an ion transfer membrane ASU. The lock-hopper is replaced by a solid feed pump and the gas turbine is upgraded to a high efficiency design. If CO_2 is compressed, the CO_2 compressor is replaced by a RamGen compressor. IGCC facilities can be equipped with hot gas cleaning and transport desulphurisation. The sulphur compounds are converted into elemental sulphur using the direct sulphur removal process, thereby eliminating the need for the Claus and SCOT installations. In the case of IGCC-CCS, the syngas is shifted using SEWGS after the sulphur compounds are removed. SEWGS also removes CO_2 from the syngas. A problem is that the H_2:CO ratio cannot be manipulated while still obtaining a low CO_2 concentration in the syngas. As the FT-reactor requires a certain H_2:CO ratio as well as a low CO_2 concentration, SEWGS cannot be used. Therefore, the IG-FT facilities will still rely on a separate WGS and CO_2 removal units. The WGS is upgraded to reduce the steam consumption and the Rectisol AGR is improved, resulting in reduced energy consumption.

In the long term, the operation conditions of the steam cycle could change from subcritical to supercritical. Also, the syngas cooler is expected to be upgraded to produce high pressure steam instead of intermediate pressure steam. The Rectisol based AGR in the IG-FT facilities is improved even further. The catalyst in the FT-reactors is replaced by a diesel selective catalyst with a different chain growth probability (α) depending on the length of the hydrocarbon. It is assumed that this has no effect on the reactor size and costs. The gas turbine is improved even further.

SOFCs are expected to become available in the long term. Previous studies indicate that incorporating SOFCs will improve efficiency, but also increase capital costs.[59,60] IGCC-SOFC facilities have been compared to the conventional long term IGCC facilities to determine which configuration results in the lowest production costs. The IG-FT facilities will not be equipped with SOFCs as the off-gas contains a high fraction of heavier hydrocarbons (C>1), which are not converted in SOFCs.

Table 4.2 Technical parameters of the selected technologies.

Process	Component	Parameter	Unit	Current	Short	Mid	Long
Feeding	Lock hoppers (1)	Specific energy consumption	kJ_e/MJ_{HHV} coal	15	15	15	15
	Solid feed pump (1)	Specific energy consumption	kJ_e/MJ_{HHV} coal			19.5	19.5
		Inert gas cons.	%			12	12
ASU	Cryogenic (2)	Specific energy consumption	kWh/t O_2	222	200		
	ITM (3)	Specific energy consumption	kWh/t O_2			162	162
WGS	SEWGS (4)	Specific energy consumption	MJ_e/t CO_2			0	0
AGR	Physical separation (5)	Specific energy consumption	MJ_e/t CO_2	214	126	77	47
			MJ_{th}/t CO_2	90	90	90	90
			MJ_e/kg S	2.35	1.39		
	TDS + DSRP (6)	Specific energy consumption	MJ_e/t S			8	8
Gas turbine	Gas turbine (7)	TIT	°C	1325	1375	1500	1700
		Pressure ratio	-	33.5	39	61	63
		Compressor $\eta_{Isentropic}$	%	92	92	92	92
		Expander $\eta_{Isentropic}$	%	82.2	83.2	84.2	85.2
	SOFC (8)	$\eta_{electric}$	%				63
		Fuel conversion efficiency	%				85
Steam turbines	Steam turbine	Pressure levels	bar	125/ 42/12.5	125/ 42/12.5	125/ 42/12.5	275/ 42/12.5
FT-synthesis	FT-catalyst (9)	Chain propagation value (α)	-	0.91	0.91	0.91	0.95 C_1-C_4 0.90 C_5-C_8 0.50 C_9-C_{19} 0.20 C_{20+}
CCS	CO_2 compressor (10)	Specific energy consumption	kWh/t CO_2	94	94	86	79
	Shock wave (11)	Specific energy consumption	kWh/t CO_2			116	116
Other	Overall plant (12)	Availability	%	75	80	85	90

(1) The specific energy consumption of the pre-treatment is 1.5% of the coal energy$_{HHV}$ input.[150,171] As the gasifier is already dry-fed, the solid feed pump will increase specific energy consumption by 30%, while capital costs decrease by 67%.[59,171,175] Inert gas consumption will also drop by 88%.

(2) Better heat integration, gas turbine integration and more efficient air compressor will reduce specific energy consumption. It is assumed that the historic trend of the last 40 years of a 2% per year reduction in specific energy consumption will continue for the next 10 years.[106,176]

(3) In the long term, the cryogenic ASU will be replaced by the ITM ASU. Specific energy consumption of the ITM is expected to be 81% of the current cryogenic technology.[177]

(4) Integration of the compression and decompression steps removed the need for a compressor. This results in no energy consumption of the SEWGS.[178]

(5) Specific energy consumption of removing CO_2 using Selexol is 214 MJ_e/t CO_2 and 90 MJ_{th} LP steam/t CO_2. When using Rectisol, this changes to 110 MJ_e/t CO_2 and

54 MJ_{th} LP steam/t CO_2. If only sulphur is removed, specific energy consumption of the Selexol unit is 2.4 MJ_e/kg S. Developments in chemical and physical solvents are expected to lower specific energy consumption by increasing loading capacities and decreasing heat capacity and heat of reaction. According to Peeters et al., regeneration energy may drop by 41% in the short term and 64% in the mid term.[179] Although this is for chemical solvents, it is assumed that the same trend will apply for physical solvents.

(6) Energy consumption of the transport desulphurisation (TDS) and direct sulphur recovery plant (DSRP) is assumed to be 8 MJ_e/t S.[59]

(7) The turbine inlet temperature (TIT) is expected to increase from 1325°C now to 1375°C in the short term, to 1500°C in the mid term and to 1700°C in the long term. The pressure ratio is selected to keep the temperature of the flue gas exiting the gas turbine 50°C above the highest temperature in the steam cycle.[180] Between 1992-2002, every 1$\%_{pt}$ increase in combined cycle efficiency due to higher turbine inlet temperatures was, according to Siemens, accompanied by a 0.37$\%_{pt}$ increase in combined cycle efficiency due to improved compressor and turbine efficiency.[181] It is assumed that this ratio also holds for other time periods.

(8) The SOFC is expected to have a fuel conversion efficiency of 85%. Of the chemical energy released, 63% is converted into electricity. The remaining energy is released as heat.[58,59,182]

(9) Although the FT-catalyst is constantly improving and new reactor designs are being developed, it is assumed that major improvements will only occur in the long term. At that point the FT-synthesis will be optimised for diesel production by having a different chain growth probability α, depending on the length of the hydrocarbon chain.[77,183,184,185,186]

(10) Compressing CO_2 from a Rectisol based AGR to 110 bar requires 94 kWh/t CO_2. When the AGR is Selexol, the compression energy is only 60 kWh/t CO_2 as the CO_2 exiting the AGR has a higher pressure. It is expected that improvements in CO_2 compressor technology will reduce specific compression energy by 16%.[187,188] It is assumed that half of this decrease is obtained in the mid term and the other half in the long term.

(11) The Ramgen shock wave compressor can obtain a much higher pressure ratio than conventional compressors. Therefore, fewer compression steps are needed and the waste heat obtained by cooling the CO_2 between the compression steps can more efficiently be used. The drawback is that specific compression energy is increased by 24%.[189]

(12) Newly build facilities are expected to have an availability of 80%. Improved operating experience and new technologies will gradually increase availability to 90%.[47,58,59,190] To account for the technological learning between the current and short term, the current availability was set at 75%.

4.4 Economic analysis

The economics of the different IG-PG facilities were evaluated by calculating the production costs of the main product using the Net Present Value (NPV) method (equation 4-1).[162] Note that temporary stored carbon in the chemical products still counts as emitted CO_2. To include transport and storage of CO_2, a fixed price per t CO_2 was taken. CO_2 avoidance costs were calculated according to equation 4-2. All cost data are given in $€_{2008}$. Common technical and economic parameters are presented in Table 4.3.

$$P_{MP}\ (€/GJ) = \frac{\alpha * I + O\&M + \text{Feedstock} - \sum\left(F_{SPx} * P_{SPx}\right)}{F_{MP}} \qquad (4\text{-}1)$$

$$CO_2\ \text{avoidance costs}\ (€/t\ CO_2) = \frac{P_{MP} - P_{MP\,ref}}{\left(E_{ref}/F_{MP\,ref}\right) - \left(E/F_{MP}\right)} \qquad (4\text{-}2)$$

where α = Capital recovery factor (yr^{-1}), calculated by $r/(1-(1+r)^{-L})$;
r = Discount rate;
L = Economic lifetime (yr);
I = Total capital investments of the facility (M€);
O&M = Operating and maintenance costs (M€/yr);
Feedstock = Coal or TOPS cost (M€/yr);
F_{SPx} = Annual flow side-product x (GJ/yr or kt/yr);
P_{SPx} = Market price of side-product x (€/GJ or €/kt);
F_{MP} = Annual flow of main product (GJ/yr);
P_{MP} = Production costs of main product (€/GJ);
E = Net CO_2 emission, including carbon in chemical products (t CO_2/yr);
Ref = The reference system, namely a coal-fired IG-PG without CCS.

Table 4.3 Technical and economic assumptions IG-PG facilities.

Parameter	**Unit**	**Value**
Location	-	NW-Europe
Construction time (1)	Year	3
Plant economic lifetime	Year	20
Discount rate	%	10
Plant size	MW_{HHV} coal eq.	1000
O&M costs (2)	% of cap. cost	4
TOPS costs (3)	€/GJ	3.0-6.3
Coal costs (3)	€/GJ	2.25
CO_2 trans. & storage costs (4)	€/t CO_2	10
Ref. electricity price (5)	€/GJ	15.7
Sulphur price	€/t S	100
Slag price	€/t slag	0
CO_2 credits (6)	€/t CO_2	0

(1) Based on literature, a construction time of three years was assumed and capital costs were evenly divided over these years.[24,50,74,79]

(2) The O&M costs are assumed to be 4% for all components except if stated differently in literature, see Table 4.4.

(3) Feedstock costs were 2.25 €/GJ for coal and 6.3 €/GJ for biomass pellets beginning 2010. Although TOPS are not produced commercially today, it was assumed that they have the same price as biomass pellets as the increase in production costs is compensated by reduction in transportation costs. Literature studies show that TOPS prices could drop to 3 €/GJ TOPS.[32,85,86,157,158] We refer to Meerman et al., for more information.[173]

(4) According to the Zero Emission Platform, transport to and storage in depleted gas or oil fields of CO_2 will cost between 2-15.7 €/t CO_2. When storing offshore, the CO_2 transport and storage costs increase to 5.5-20 €/t CO_2 for depleted gas or oil fields.[54] Based on expert interview, the CO_2 transport and storage costs were set at 10 €/t CO_2.[191]

(5) The reference electricity price is based on the average Dutch day-hourly market price between 2004-2008. The observed trends were considered representative for NW-Europe. During that period the electricity price varied between 0-1050 €/MWh (0-290 €/GJ), with an average price of 57 €/MWh (15.7 €/GJ).[66]

(6) In this study CO_2 avoidance costs are calculated. Therefore, no CO_2 credit price was used.

Table 4.4 Capital costs of the components of the IG-PG facilities investigated.

Component		Base scale	Scale unit	Scaling factor	Base Cost (M€$_{2008}$)	Direct costs (1)	Indirect costs (1)	O&M costs (1)	# units
Pre-treatment and feeding									
Lock hoppers	Biomass (2)	65	tonne a.r./hr	0.77	10	27%	25%	4%	2
	Coal (2)	273	tonne a.r./hr	0.65	41	53%	14%	4%	2
Solid feed pump	Biomass (3)	65	tonne a.r./hr	0.77	3	27%	25%	4%	2
	Coal (2)	273	tonne a.r./hr	0.65	13	53%	14%	4%	2
Gasification									
ASU	Cryogenic (2)	5149	tonne output/d	0.8	112	37%	16%	4%	1
	ITM (4)	5149	tonne output/d	0.8	75	37%	16%	4%	1
Shell EF gasifier (2)		2053	MW_{HHV} coal eq.	0.66	139	73%	24%	4%	1
Gas Cleaning									
Candle filter (2)		12	gas m^3/s	0.65	2	33%	50%	4%	0-1
Wet Scrubber (2)		12	gas m^3/s	0.7	3	33%	50%	4%	0-1
Heat exchanger (2)		138	MW_{th}	0.6	9	31%	40%	4%	0-1
WGS reactor (2)		8819	H_2+CO kmol/hr	0.65	14	0%	81%	4%	0-1
Guard bed (2)		0.6	gas Nm^3/s	1	0.003	0%	200%	4%	1
H_2 PSA (2)		16616	H_2 kmol/hr	0.65	12	100%	28%	4%	0-1
Hot gas cleaning (5)		1417	MW_{HHV} coal eq.	0.67	50	21% combined		10%	0-1
SEWGS (6)		1000	MW_{HHV} coal eq.	0.67	28	33%	50%	4%	0-1
Acid Gas Removal									
Selexol AGR removal (7)		554	CO_2 tonne/hr	0.7	34-69	33%	50%	4%	0-1
Rectisol AGR removal (7)		554	CO_2 tonne/hr	0.7	120	33%	50%	4%	0-1
Claus + SCOT (7)		56	S tonne/d	0.67	16-29	33%	14%	4%	0-1
TDS + DSRP (5)		This is included in the hot gas cleaning costs.							
Power Isle									
Syngas combustion	Gas Turbine (2)	266	Net MW_e	0.75	42	16%	27%	4%	1
	SOFC (8)	1	Gross MW_e	1	0.5	0%	20%	9.2%	Varies
Steam Turbines and Cycle (2)		275	Gross MW_e	0.67	38	16%	27%	4%	1
HRSG (2)		355	MW_{th} exch.	1	34	16%	27%	4%	1
FT Isle									
FT Slurry Reactor (2)		131	MW_{HHV} FT	0.72	21	33%	50%	4%	0-1
Product Upgrading (2)		220	FT tonne/hr	0.7	132	33%	50%	4%	0-1
Compression									
Syngas Compr. (2)		10	MW_e	1	4	16%	32%	4%	Varies
CO_2 Compr.	Conventional (7)	13	MW_e	1	5-6	16%	32%	4%	1
	Shock wave (9)	16	MW_e	1	3-4	16%	32%	4%	1

(1) The capital investment of a component was calculated by taking the scaled base costs of the component. The direct cost percentage relates to the scaled base costs. The indirect cost percentage was taken over the summation of base and direct costs. The sum of base, direct and indirect costs is the capital investment of that component. The O&M fraction is based on the capital investment of the component. Note that the direct costs decrease with increasing scale, with a scaling factor of -0.18.[173]

(2) Component costs data and scale as well as installation and scaling factors were taken from Meerman et al.,[173] and are based on literature data.[22,24,25,50,74,77,144,165,166] It was assumed that existing technologies do not become more efficient unless specifically said so, e.g., the FT-reactor will not become smaller over time.

(3) The solid feed pump is expected to reduce capital costs of feeding system by 67% compared to a lock hopper system. Installation factors are assumed to remain constant.[59,175]

(4) The ITM is expected to reduce capital costs of the ASU by 33% compared to a cryogenic ASU. Installation factors are assumed to remain constant.[192,193,194]

(5) Hot gas cleaning refers to a cleaning train where the syngas temperature stays above 250°C. The conventional IGCC cold gas cleaning train, including wet scrubber, AGR and Claus/SCOT is replaced with a hot gas cleaning, including absorbent beds, TDS and DSRP. According to the NETL, base capital costs of the gas cleaning train of a 1409 MW_{th} input IGCC will drop by 26% to 71 $M\$_{2007}$. The O&M costs of the gas cleaning train will increase to 10%.[59]

(6) A SEWGS is expected to be 20-30% cheaper than the WGS and Selexol based AGR it will replace.[178]

(7) Capital costs of the AGR as well as the Claus and SCOT plants depend on whether Selexol or Rectisol is used and whether CO_2 needs to be removed. Compared to a Selexol based AGR removing only sulphur, a Selexol based AGR extracting both CO_2 and sulphur is twice as expensive and a Rectisol based AGR is 3.5 times more expensive. The sulphur treatment after Rectisol is 47% cheaper than after Selexol.[50] If CO_2 is captured using Selexol, water needs to be removed during compression. If Rectisol is used, no additional water removal is required. The lower pressure at which CO_2 comes available with Rectisol, however, results in a more expensive CO_2 compressor when using Rectisol than when using Selexol. The CO_2 compressor after Selexol is 75% of the capital costs of a CO_2 compressor after Rectisol.[50]

(8) Capital costs of the SOFC are based on a study by the NETL and amount 500 k€/MW_e.[58] The fuel stacks need to be replaced every 40.000 operating hours and costs 26.5% of the initial total capital costs.[59] With a 90% availability, this results in an O&M costs of 9.2%.

(9) The Ramgen shock wave compressor can obtain a much higher pressure ratio than conventional compressors, resulting in fewer compression steps. This leads to a decrease in capital costs of 25-38% compared to conventional compressors.[189] In this study, a cost reduction of 30% was assumed.

Total capital investments (TCI) were calculated using the Factored Estimation method. In this method, the cost of each major component is estimated using data from open literature sources and expert interviews. These cost data correspond to a specific scale and are adjusted using a scaling factor. An installation factor, representing e.g., infrastructure, overhead, engineering and contingencies, is added to the scaled component capital costs. The TCI is the summation of the scaled costs. The used component capital costs, scaling factors and installation factors are given in Table 4.4. A detailed explanation of the cost improvements of new technologies is given in the Appendix.

4.5 Results

Overall efficiency, production costs and CO_2 avoidance costs for electricity and FT-liquids production were calculated for the different time periods. Detailed mass, energy and economic balances can be found in the supplementary data.

Our results indicate that for a coal-fired IGCC without CCS, the overall energy conversion efficiency, presented in HHV, increases from 44% to 59% in the long term (Figure 4.4). The

CCS penalty remains around 9%$_{pt}$, except when SOFCs are used. In that case, the CCS penalty is only 4%$_{pt}$. Production of FT-liquids can be done at 61% efficiency. This increases to 65% in the long term. CO_2 is always removed from the syngas when FT-liquids are produced as this improves the overall syngas conversion into FT-liquids. Therefore, the CCS penalty consists of only the compression of CO_2 and is around 2%$_{pt}$. Biomass-fired IG-PG facilities have almost similar efficiencies as their coal-fired counterparts.

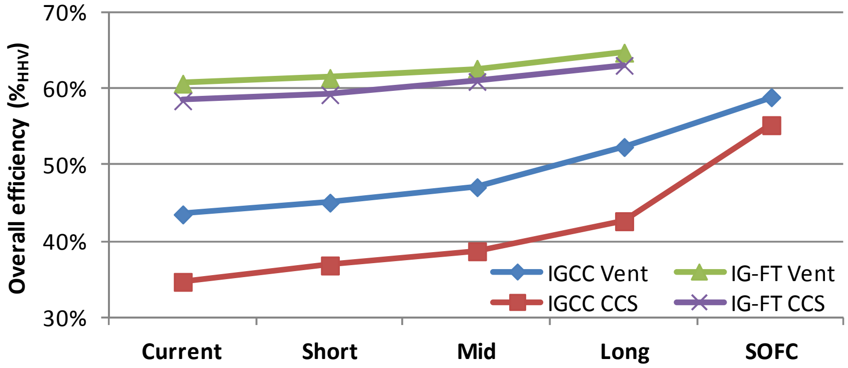

Figure 4.4 Overall efficiency of a coal-fired IG-PG facility in the different time periods.

Calculated production costs are presented in Figure 4.5. Results show electricity production for a coal-fired SOTA IGCC-Vent at 16.6 €/GJ (60 €/MWh), decreasing to 11.1 €/GJ (40 €/MWh) in the long term. If CCS is applied, production costs decrease from 23.1 €/GJ (83 €/MWh) to 15.0 €/GJ (54 €/MWh). If TOPS is used as feedstock, production costs decrease from 25.4 €/GJ (92 €/MWh) to 18.5 €/GJ (67 €/MWh) without CCS and from 34.9 €/GJ (126 €/MWh) to 23.9 €/GJ (86 €/MWh) with CCS. If, however, TOPS is obtained at e.g., 3 €/GJ instead of 6.3 €/GJ, production costs may decrease to 12.2 €/GJ (44 €/MWh) for Vent and 16.4 €/GJ (59 €/MWh) with CCS in the long term.

Converting coal to FT-liquids in a SOTA IG-FT results in production costs of 13.3 €/GJ, which decrease to 9.1 €/GJ in the long term. If CO_2 is captured, these values change to 14.6 €/GJ to 10.1 €/GJ in the long term. If TOPS is used as feedstock, production costs drop from 22.7 €/GJ to 18.1 €/GJ without CCS and from 24.5 €/GJ to 19.3 €/GJ with CCS when the TOPS price is 6.3 €/GJ. If TOPS prices reduce to 3 €/GJ, FT-liquids production costs drop by 29-38%.

Deployment of SOFC shows a mixed picture when looking at production costs. For the Vent facilities, the conventional gas turbine appears more economical, while the CCS facilities have lower production costs if SOFC are used (14.1 €/GJ or 51 €/MWh for Vent and 21.2 €/GJ or 76 €/MWh for CCS).

When comparing the production costs with the market price of electricity (15.7 €/GJ or 57 €/MWh) and transportation fuels (10.1 €/GJ, equivalent to an oil price of 85 $/bbl), results indicate that a coal-based IGCC Vent has production costs which drop below the market price in the short term. IG-FT facilities produce FT-liquids at competitive costs in the long term only. Increases in oil price, however, may bring this time period forward.

Specific direct CO_2 emissions may drop from 712 g CO_2/kWh for a coal-fired SOTA IGCC Vent to 29 g CO_2/kWh if CCS is applied. The use of TOPS results in no emissions without CO_2 capture and -934 g CO_2/kWh for the combination biomass and CCS.

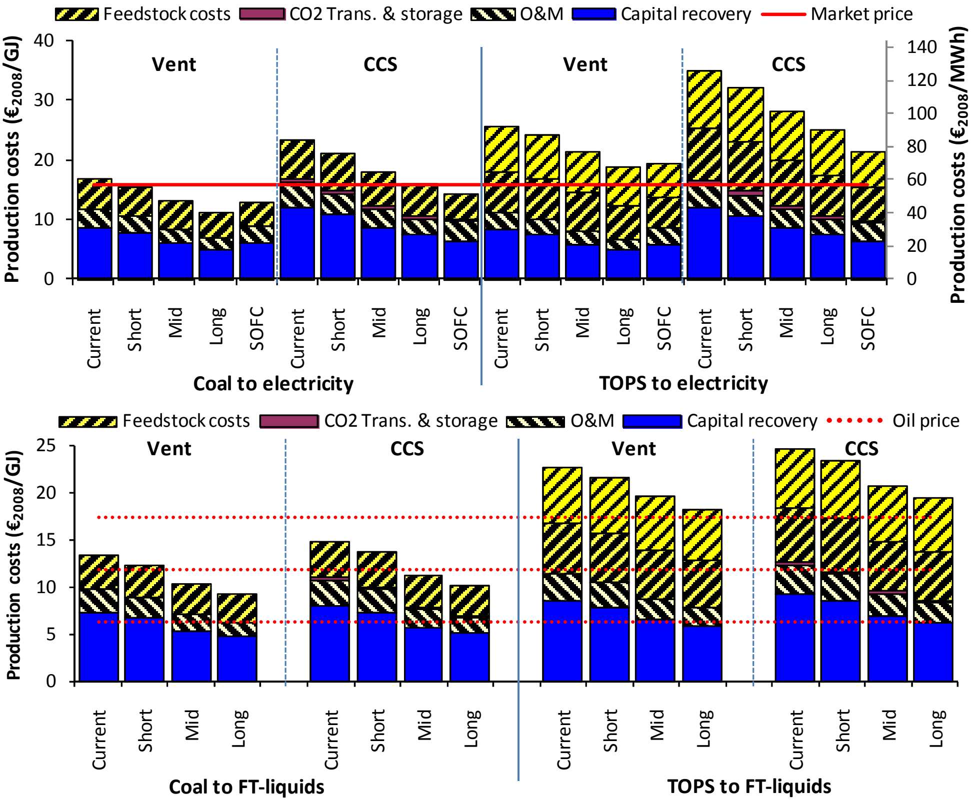

Figure 4.5 Production costs of electricity (above) and FT-liquids (below). The lighter upper part of the feedstock bar is the addition in production costs when using the high value for the TOPS price.

A capital costs breakdown of the coal-fired IG-PG facilities is given in Figure 4.6. The costs breakdown of the TOPS-fired facilities is given in the supplementary data. The largest reduction in production costs occurs between the short and mid term. This was to be expected as in this time period several new technologies, like ITM, solid feed pump and hot gas cleaning, are implemented in our study. The largest reductions in capital costs occur at the gas cleaning and oxygen production sections.

Net CO_2 emissions can be reduced by substituting coal for biomass and by capturing CO_2. The cost per tonne avoided CO_2 emissions depends on the option and the production mode (see Figure 4.7). Advanced technologies reduce the CCS penalty by the combination of increased overall efficiency and more efficient CO_2 capture and compression. Similar to the production costs, the largest reduction in CO_2 avoidance costs is obtained between the short and mid term. Our findings also indicate that for IG-FT facilities the lowest CO_2 avoidance cost are obtained when capturing CO_2 from coal, while the use of TOPS of 6.3 €/GJ without CCS results in the highest CO_2 avoidance costs. For IGCC, the picture is more complex. At high TOPS prices (6.3 €/GJ), CO_2 capture with coal has the lowest CO_2 avoidance costs. When TOPS prices drop, this changes to TOPS with CCS and at even lower TOPS prices (3 €/GJ) to TOPS without CCS.

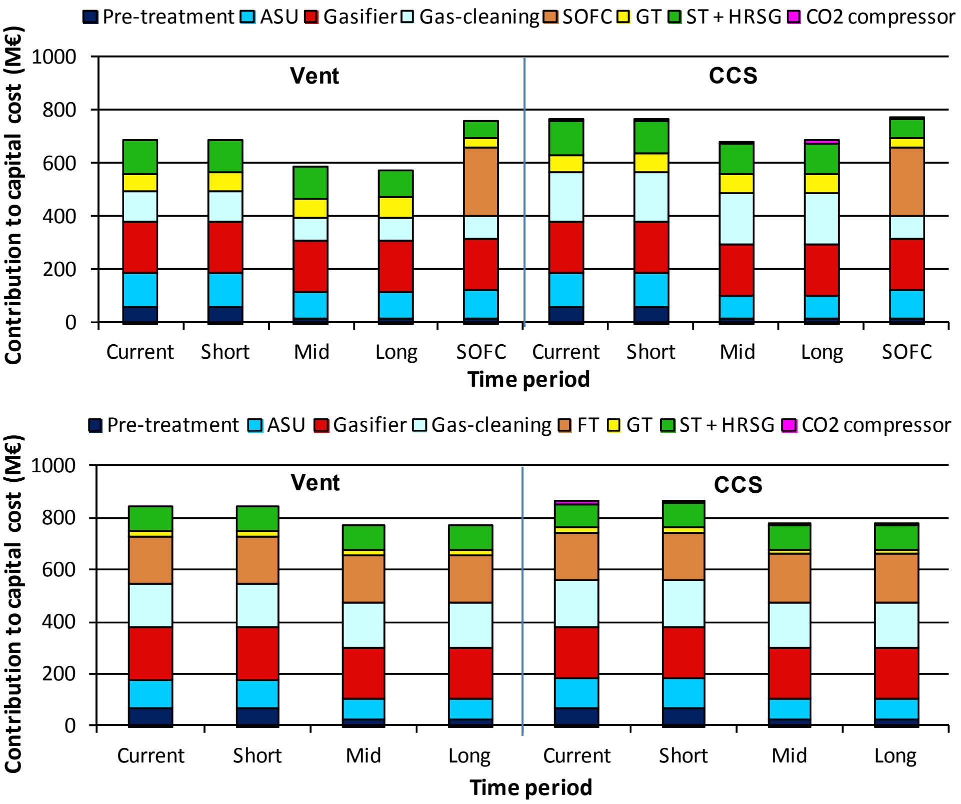

Figure 4.6 Breakdown capital costs coal-fired IGCC (above) and IG-FT (below) in M€.

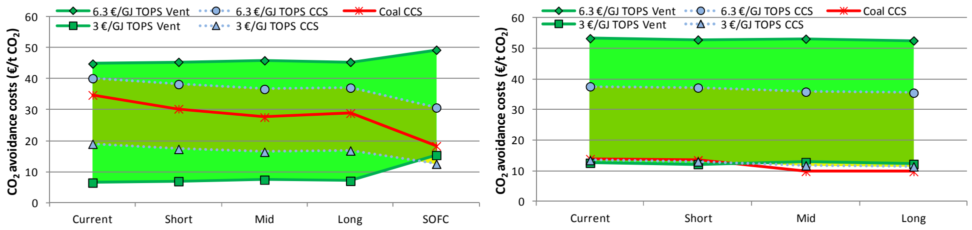

Figure 4.7 CO_2 avoidance costs compared to coal-fired IGCC or IG-FT without CO_2 capture in the same time period. The shaded area gives the uncertainty related with the TOPS price.

4.5.1 *Sensitivity*

In this section the effect of using a different gasifier, different SOFC price and of a CO_2 credit price on production costs is investigated.

The high pressure gasifier was not expected to become commercially available before 2050 and, as such, was not included in the long term configurations. Using this high pressure gasifier could result in production costs of 9.8 €/GJ for the IGCC-Vent (-11%) and 8.8 €/GJ (-3%) for the IG-FT-Vent.

In this study IGCC-CCS with SOFCs results in lower production costs than with only a gas turbine. The assumed SOFC module costs is, however, low. SOFC investment costs can increase to 0.62 M€/MW_e when using coal and to 0.88 M€/MW_e when using biomass before it becomes more attractive to operate the IGCC-CCS without SOFCs.

In the economic analysis, it was assumed that there is no CO_2 tax. When applying such a tax, facilities equipped with CCS will become more attractive, as will the use of biomass (see Figure 4.8). As expected, the IG-FT facilities with CCS break even with the Vent facilities at a lower CO_2 credit price than IGCC facilities, due to the lower CO_2 capture penalty. If TOPS can be obtained at 3 €/GJ, it becomes attractive to use TOPS instead of coal at low CO_2 price (≈10 €/t CO_2).

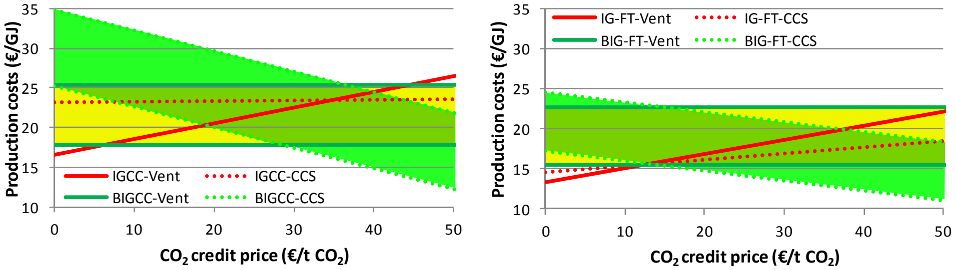

Figure 4.8 Effect CO_2 credit price on production costs of SOTA IG-PG facilities. The shaded area is the range due to low and high TOPS prices.

4.6 Discussion

A breakdown of the individual contribution of the new technologies was estimated for coal-fired IGCC (see Figure 4.9). Increasing the availability from 75% to 90% may lower the electricity production costs of an IGCC with CO_2 capture by 2.7 €/GJ. Better gas and steam turbines give a reduction of 2.6 €/GJ. The other technologies combined reduce production costs by 5.2 €/GJ. The upgrade of producing HP instead of IP in the gasifier and syngas coolers results in a reduction in production costs of 0.3 €/GJ. The shock wave compression may increase the production costs when applied in a SOTA IG-PG facility. When combined with hot gas cleaning, however, the shock wave compressor does result in a reduction in production costs. The main difference is the higher pressure of the CO_2 after a Selexol-based AGR compared to the CO_2 pressure after a SEWGS. The shock wave compressor is much more effective at the lower pressure. As some technologies have identical impacts, e.g., reduced temperature variations or less steam consumption, the total reduction of the individual technologies is larger than the combined reduction. This difference is 1.8 €/GJ and grouped in Figure 4.9 under synergies.

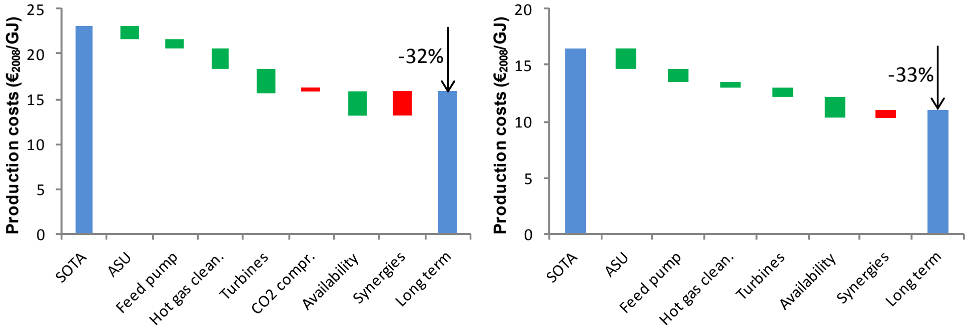

Figure 4.9 Individual contribution of improvements in a coal-fired IGCC with CCS (left) and without CCS (right).

The SOFC does lower production costs of the facilities equipped with CCS. However, the captured CO_2 is more contaminated than CO_2 captured using a conventional AGR or SEWGS. This could increase the CO_2 cost for transport and underground storage, thereby annulling the purpose of the CO_2 capture or adding requirements for cleaning facilities.

4.7 Conclusions

Advanced technologies may reduce production costs of a coal-fired IGCC without CO_2 capture from 16.6 €/GJ to 11.1 €/GJ (60 €/MWh to 40 €/MWh). When CO_2 is captured, it is found that production costs are lowered from 23.1 €/GJ to 14.1 €/GJ (83 €/MWh to 51 €/MWh) using SOFC. This would result in IGCC becoming profitable in the short term if CCS is not applied and in the long term if CCS is applied. When TOPS are used as feedstock, production costs are currently calculated at 25.4 €/GJ (92 €/MWh) without CCS and 34.9 €/GJ (126 €/MWh) with CCS, dropping to respectively 18.5 €/GJ and 21.2 €/GJ (67 €/MWh and 76 €/MWh) in the long term. New technologies alone do not lower production costs of TOPS-fired IGCC under the current average electricity market value of 15.7 €/GJ (56 €/MWh). If, as several studies indicate, TOPS prices drop to 3 €/GJ, production costs would decrease to 12.2 €/GJ (44 €/MWh) without CCS and 15.2 €/GJ (55 €/MWh) with CCS in the long term. In this case, production costs would drop under the current market price.

New technologies in IG-FT facilities are found to have a slightly smaller impact on the production costs. When using coal, production costs decrease from 13.3 €/GJ to 9.1 €/GJ if CO_2 is vented and from 14.6 €/GJ to 10.1 €/GJ if CO_2 is captured and stored. The use of TOPS would result in 22.7 €/GJ and 18.1 €/GJ without CCS and 24.5 €/GJ and 19.3 €/GJ with CCS for respectively now and in the long term. Here, lower biomass feedstock costs of 3 €/GJ results in production costs of 11.2 €/GJ without CCS and 12.4 €/GJ with CCS.

Specific CO_2 emissions can be reduced by capturing CO_2 or by substituting coal by TOPS. If both options are applied, net negative emissions can be obtained. The lowest CO_2

avoidance costs are found when CO_2 is captured from a coal-fired IG-FT facility (9.8 €/t CO_2 in the long term).

The main improvement in plant economics is obtained by increasing the availability of the IG-PG facility from 75% to 90%, closely followed by improvements in gas and steam turbines. Together, they are responsible for almost two-third of the reduction in production costs of a coal-fired IGCC without CCS.

It is concluded that IG-PG facilities can make a major contribution to the transition towards a sustainable energy infrastructure by utilising TOPS and CCS. The perspective that this analysis predicts suggests that in the future IG-PG facilities can produce the electricity and FT-liquids at competitive costs. Considerable R&D efforts and market deployment are necessary to achieve such a situation. In the transition period CO_2 credit prices of 23-45 €/t CO_2 or oil prices of 95-150 $/bbl can tick the balance for competitive operation on shorter term.

4.8 Acknowledgement

This research is part of the CATO2 and CAPTECH programmes. CATO2 is the Dutch national R&D programme for CO_2 capture, transport and storage in which a consortium of nearly 40 partners cooperate. CAPTECH is a cooperation between the Dutch Ministry of Economic Affairs (EOS-LT programme), Energy research Centre of the Netherlands (ECN); KEMA Nederland B.V.; Procede; Shell Global Solutions International B.V.; TNO Science and Industry; Utrecht Centre for Energy research (UCE); and Utrecht University (UU). For more information: www.co2-cato.nl and www.co2-captech.nl.

4.9 Supporting information

4.9.1 Production and CO_2 avoidance costs formula

Production costs of the main products (electricity for IGCC and FT-liquids for IG-FT) were calculated according to equation 4-3. CO_2 avoidance costs were calculated according to equation 4-4.

$$P_{MP}\ (€/GJ) = \frac{\alpha * I + O\&M + Feedstock - \sum\left(F_{SPx} * P_{SPx}\right)}{F_{MP}} \tag{4-3}$$

$$CO_2\ \text{avoidance costs}\ (€/t\ CO_2) = \frac{P_{MP} - P_{MP\ ref}}{\left(E_{ref}/F_{MP\ ref}\right) - \left(E/F_{MP}\right)} \tag{4-4}$$

where α = Capital recovery factor (yr^{-1}), calculated by $r/(1-(1+r)^{-L})$;
r = Discount rate;
L = Economic lifetime (year);
I = Total capital investments of the facility (M€);
O&M = Operating and maintenance costs (M€/yr);
Feedstock = Coal or TOPS cost (M€/yr);
F_{SPx} = Annual flow side-product x (GJ/yr or kt/yr);
P_{SPx} = Market price of side-product x (€/GJ or €/kt);
F_{MP} = Annual flow of main product (GJ/yr);
P_{MP} = Production costs of main product (€/GJ);
E = Net CO_2 emission, including carbon in chemical products (t CO_2/yr);
Ref = The reference system, namely a coal-fired IG-PG without CCS.

4.9.2 Capital cost breakdown

A breakdown of the capital costs of the biomass-fired IG-PG facilities is given in Figure 4.10.

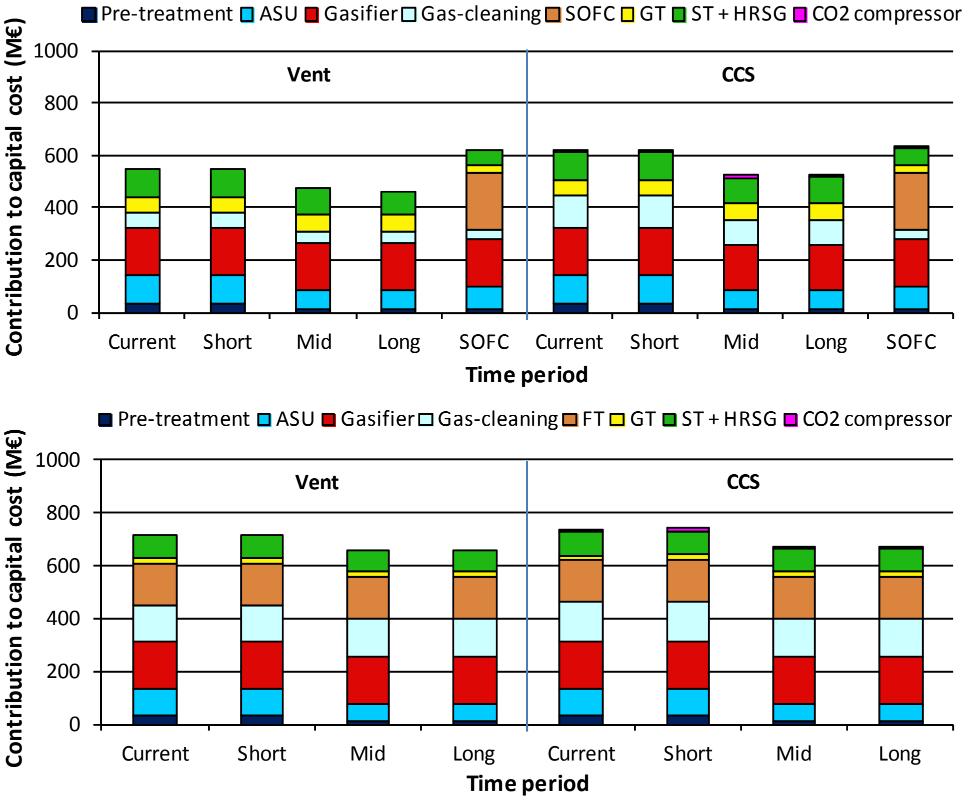

Figure 4.10 Breakdown capital costs TOPS-fired IGCC (above) and IG-FT (below) in M€.

4.9.3 Detailed mass, energy and economic results

The economic performance of IG-PG facilities are given in Table 4.5 for IGCC and Table 4.6 for IG-FT. Mass balances are given in Table 4.7 (IGCC) and Table 4.8 (IG-FT). Table 4.9 (IGCC) and Table 4.10 (IG-FT) contain the energy balance and capital cost breakdown is given in Table 4.11 for IGCC and in Table 4.12 for IG-FT.

Table 4.5 Economic performance IGCC.

		Coal (2.25 €/GJ)										**TOPS (3 €/GJ)**									
		VENT					**CCS**					**VENT**					**CCS**				
		Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC
Annual costs																					
Cap. recovery	M€/yr	89	89	76	74	98	100	100	82	82	101	71	71	61	60	81	80	81	68	68	82
O&M	M€/yr	30	30	29	29	51	34	34	31	32	52	24	24	23	22	40	27	27	25	25	41
Feedstock	M€/yr	53	57	60	64	64	53	57	60	64	64	60	64	68	72	72	60	64	68	72	72
CO_2 T. & S.	M€/yr						5	6	6	7	7						5	5	6	6	6
Annual revenues																					
Electricity	M€/yr	170	174	163	164	210	190	194	177	182	220	155	160	152	154	193	173	177	166	172	202
Sulphur	M€/yr	2	2	2	3	3	2	2	2	3	3	0	0	0	0	0	0	0	0	0	0
Prod. costs	**€/GJ**	17	15	13	11	13	23	21	17	15	14	18	17	14	12.2	13.6	25	23	19	16.4	15.2
	€/MWh	60	55	47	40	45	83	75	61	54	51	64	60	51	44	49	91	82	67	59	55
CO_2 avoid. cost	€/t CO_2						35	30	23	24	18	6	7	7	7	15	19	17	14	14	12

Table 4.6 Economic performance IG-FT.

		Coal (2.25 €/GJ)								TOPS (3 €/GJ)							
		VENT				CCS				VENT				CCS			
		Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long
Annual costs																	
Capital recovery	M€/yr	109	109	99	99	112	112	101	101	93	93	86	86	96	96	87	87
O&M	M€/yr	37	37	34	34	38	38	34	34	32	32	29	29	33	33	30	30
Feedstock	M€/yr	53	57	60	64	53	57	60	64	60	64	68	72	60	64	68	72
CO_2 transport & storage	M€/yr					3	3	4	4					3	3	3	4
Annual revenues																	
FT-liquids	M€/yr	153	150	135	131	169	167	148	144	143	141	133	130	159	158	146	143
Electricity	M€/yr	44	50	56	64	36	42	49	56	42	47	50	57	33	38	43	50
Sulphur	M€/yr	2	2	2	3	2	2	2	3	0	0	0	0	0	0	0	0
Production costs	**€/GJ**	13	12	10	9	15	14	11	10	15	14	12	11	17	16	14	12
CO_2 avoid. cost	€/t CO_2					14	13	10	10	13	12	13	12	13	13	12	11

Table 4.7 Mass balance IGCC.

		Coal										TOPS									
		VENT					CCS					VENT					CCS				
		Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC
Input																					
Coal / TOPS	kt/yr	872	930	988	1,046	1,046	872	930	988	1,046	1,046	978	1,043	1,108	1,173	1,173	978	1,043	1,108	1,173	1,173
Bio-based C	kt C/yr											501	534	567	601	601	501	534	567	601	601
Coal-based C	kt C/yr	556	593	630	667	667	556	593	630	667	667										
Output																					
Slag	kt/yr	85	90	96	101	101	85	90	96	101	101	12	13	14	15	15	12	13	14	15	15
Sulphur	kt/yr	22	23	25	26	26	22	23	25	26	26	0	0	0	0	0	0	0	0	0	0
Carbon balance																					
C emitted	kt C/yr	556	593	630	667	667	18	19	4	4	3	501	534	567	601	601	16	17	3	3	3
C stored	kt C/yr						538	574	625	662	663						485	517	564	598	598
Net carbon emission	kt CO_2/yr	2,036	2,172	2,307	2,443	2,443	66	70	15	16	12	0	0	0	0	0	-1,777	-1,895	-2,068	-2,190	-2,190
	kg CO_2/GJ	198	191	183	164	146	8	8	1	1	1	0	0	0	0	0	-260	-244	-231	-209	-165
	g CO_2/kWh	712	687	658	592	526	29	27	5	5	3	0	0	0	0	0	-934	-879	-832	-752	-593

Table 4.8 Mass balance IG-FT.

		Coal								TOPS							
		VENT				CCS				VENT				CCS			
		Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long
Input																	
Coal / TOPS	kt/yr	872	930	988	1,046	872	930	988	1,046	978	1,043	1,108	1,173	978	1,043	1,108	1,173
Bio-based C	kt C/yr									501	534	567	601	501	534	567	601
Coal-based C	kt C/yr	556	593	630	667	556	593	630	667								
Output																	
FT-liquids	kt/yr	544	580	620	668	542	578	621	668	471	502	540	582	469	500	540	581
Slag	kt/yr	85	90	96	101	85	90	96	101	12	13	14	15	12	13	14	15
Sulphur	kt/yr	22	23	25	26	22	23	25	26	0	0	0	0	0	0	0	0
Carbon balance																	
C emitted	kt C/yr	352	376	396	415	39	42	45	44	338	360	379	397	31	33	36	34
C in product	kt C/yr	204	217	234	252	203	217	234	252	163	173	189	203	162	173	188	203
C stored	kt C/yr					313	334	350	371					307	328	343	363
Net C emissions	kt CO_2/yr	2,036	2,172	2,307	2,443	888	947	1,024	1,084	0	0	0	0	-1,127	-1,202	-1,257	-1,331

Table 4.9 Energy balance IGCC.

		Coal										TOPS									
		VENT					CCS					VENT					CCS				
		Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC
Input																					
Coal / TOPS	MW$_{th}$	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	848	848	848	848	848	848	848	848	848	848
	PJ/yr	23.7	25.2	26.8	28.4	28.4	23.7	25.2	26.8	28.4	28.4	20.1	21.4	22.7	24.1	24.1	20.1	21.4	22.7	24.1	24.1
Output																					
Electricity	PJ/yr	10.3	11.4	12.6	14.9	16.7	8.2	9.3	10.4	12.1	15.7	8.7	9.6	10.7	12.6	14.2	6.8	7.8	8.9	10.5	13.3
Gross elec. prod.	MW$_e$	507	519	543	602	652	452	464	485	527	660	425	436	451	503	553	380	389	416	454	561
Net elec. prod.	**MW$_e$**	**435**	**451**	**471**	**524**	**589**	**347**	**369**	**387**	**427**	**552**	**369**	**381**	**398**	**444**	**501**	**289**	**308**	**334**	**370**	**468**
Net efficiency	**%$_{HHV}$**	**43.5**	**45.1**	**47.1**	**52.4**	**58.9**	**34.7**	**36.9**	**38.7**	**42.7**	**55.2**	**43.5**	**44.9**	**46.9**	**52.3**	**59.1**	**34.1**	**36.3**	**39.4**	**43.6**	**55.2**
Auxiliary cons.																					
Pre-treatment	MW$_e$	15.0	15.0	19.5	19.5	19.5	15.0	15.0	19.5	19.5	19.5	12.7	12.7	16.5	16.5	16.5	12.7	12.7	16.5	16.5	16.5
ASU	MW$_e$	52.9	48.5	41.9	42.1	30.9	49.0	44.7	27.6	27.6	30.9	41.8	38.3	27.4	27.2	24.8	39.7	36.2	21.7	21.7	24.8
AGR	MW$_e$	0.7	0.4	0.7	0.7	0.7	19.1	11.3	0.7	0.7	0.7	0.0	0.0	0.0	0.0	0.0	17.6	10.4	0.0	0.0	0.0
CO$_2$ compr.	MW$_e$						19.2	19.2	34.9	34.9	44.5						17.6	17.7	31.5	31.5	40.2
Misc.	MW$_e$	2.6	4.6	9.6	16.3	12.1	3.0	5.5	15.4	17.9	12.1	2.2	4.0	9.4	15.5	10.9	2.4	4.4	12.5	14.7	10.9
Total	**MW$_e$**	**71.2**	**68.5**	**71.7**	**78.6**	**63.2**	**105.3**	**95.7**	**98.0**	**100.5**	**107.8**	**56.8**	**55.0**	**53.3**	**59.2**	**52.2**	**90.0**	**81.4**	**82.2**	**84.4**	**92.4**

Table 4.10 Energy balance IG-FT.

		Coal								TOPS							
		VENT				CCS				VENT				CCS			
		Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long
Input																	
Coal / TOPS	MW_{th}	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	848	848	848	848	848	848	848	848
	PJ/yr	23.7	25.2	26.8	28.4	23.7	25.2	26.8	28.4	20.1	21.4	22.7	24.1	20.1	21.4	22.7	24.1
Output																	
Electricity	PJ/yr	2.8	3.2	3.6	4.0	2.3	2.7	3.1	3.5	2.6	3.0	3.2	3.6	2.1	2.4	2.7	3.1
Gasoline	PJ/yr	1.8	2.0	2.0	3.7	1.7	1.8	2.0	3.7	1.5	1.6	1.6	3.0	1.4	1.4	1.6	3.0
Diesel	PJ/yr	9.7	10.3	11.2	10.6	9.8	10.5	11.3	10.6	7.7	8.2	9.1	8.6	7.8	8.3	9.1	8.5
Gross elec. prod.	MW_e	187	189	194	204	186	188	197	207	168	170	169	179	167	169	173	182
Net elec. prod.	**MW_e**	**119**	**127**	**133**	**142**	**97**	**105**	**115**	**125**	**112**	**118**	**119**	**128**	**90**	**97**	**101**	**111**
Net efficiency	**$\%_{HHV}$**	**60.7**	**61.4**	**62.6**	**64.7**	**58.4**	**59.2**	**60.9**	**63.0**	**59.1**	**59.9**	**60.9**	**63.2**	**56.3**	**57.2**	**58.9**	**61.0**
Auxiliary cons.																	
Pre-treatment	MW_e	15.0	15.0	19.5	19.5	15.0	15.0	19.5	19.5	12.7	12.7	16.5	16.5	12.7	12.7	16.5	16.5
ASU	MW_e	35.1	31.7	26.8	26.5	37.1	33.5	26.9	26.7	28.0	25.2	21.5	21.4	30.4	27.4	21.8	21.6
AGR	MW_e	5.3	3.1	1.9	1.2	6.0	3.5	1.9	1.2	5.2	3.1	1.9	1.1	6.0	3.6	1.9	1.2
CO_2 compr.	MW_e					18.4	18.5	19.9	19.9					18.6	18.6	19.5	19.2
Misc.	MW_e	12.7	13.0	13.4	14.4	12.2	12.5	13.4	14.3	10.3	10.3	10.9	11.8	9.7	9.9	11.5	12.5
Total	**MW_e**	**68.1**	**62.8**	**61.6**	**61.5**	**88.8**	**83.0**	**81.6**	**81.5**	**56.2**	**51.3**	**50.9**	**50.9**	**77.4**	**72.2**	**71.3**	**70.9**

Table 4.11 Capital cost breakdown IGCC (M€).

	Coal										TOPS									
	VENT					CCS					VENT					CCS				
	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC	Current	Short	Mid	Long	SOFC
Pre-treatment	60	60	20	20	20	60	60	20	20	20	34	34	11	11	11	34	34	11	11	11
ASU	127	127	93	94	101	125	126	81	82	101	107	108	74	74	86	107	108	70	70	86
Gasifier	197	197	197	197	197	197	197	197	197	197	179	179	179	179	179	179	179	179	179	179
Gas cleaning	113	113	82	82	82	188	188	138	138	82	58	58	41	41	42	127	127	92	92	40
SOFC	0	0	0	0	263	0	0	0	0	263	0	0	0	0	215	0	0	0	0	215
GT	67	69	74	79	33	62	65	70	74	33	57	59	64	69	31	53	55	61	64	31
ST + HRSG	125	123	119	101	63	127	125	112	114	68	113	112	103	85	56	111	109	100	102	61
CO_2 compr.						10	10	10	10	13						9	9	9	9	12
Total	**688**	**689**	**586**	**573**	**759**	**769**	**770**	**629**	**635**	**777**	**549**	**550**	**473**	**460**	**621**	**620**	**621**	**523**	**528**	**636**

Table 4.12 Capital cost breakdown IG-FT (M€).

	Coal								TOPS							
	VENT				CCS				VENT				CCS			
	Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long	Current	Short	Mid	Long
Pre-treatment	60	60	20	20	60	60	20	20	34	34	11	11	34	34	11	11
ASU	116	116	82	82	119	119	82	82	99	99	69	69	101	102	70	70
Gasifier	197	197	197	197	197	197	197	197	179	179	179	179	179	179	179	179
Gas cleaning	173	173	172	170	183	183	174	171	141	141	140	138	152	152	141	139
FT	183	183	185	188	183	183	185	188	158	158	160	163	157	157	160	162
GT	19	20	21	21	19	19	21	21	16	17	18	18	16	16	18	17
ST + HRSG	93	93	90	90	94	94	92	92	86	86	83	84	87	87	85	85
CO_2 compr.					13	13	8	8					13	13	8	8
Total	**842**	**842**	**767**	**768**	**867**	**867**	**779**	**779**	**714**	**714**	**661**	**662**	**741**	**742**	**673**	**673**

5 Future technological and economic performance of IGCC and FT production facilities with and without CO_2 capture: combining component based learning curve and bottom-up analysis.

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Submitted

Abstract

This study aims to investigate the technological and economic prospects of integrated gasification facilities for power (IGCC) and Fischer-Tropsch (FT) liquid production with and without CCS over time. For this purpose, a component-based experience curve about technological learning was constructed and applied to identify the potential performance improvement of integrated gasification facilities. The results of the experience curve were compared with a bottom-up engineering analysis conducted in previous work.[173]

Results indicate that substantial cost reductions and performance improvements are possible, especially for IGCC with CCS (IGCC-CCS). The costs of electricity production (COE) may decrease from 82 €$_{2008}$/MWh at present to 50 €$_{2008}$/MWh in the long term (2050), if solid oxide fuel cells (SOFCs) become commercially available. Without SOFC, the COE is projected to be slightly higher, namely 56 €$_{2008}$/MWh by 2050. This indicated cost decrease can only be realized if installed IGCC capacity increases to over 600 GW_e and installed CCS capacity to over 2,000 GW_e equivalent. IGCC plants without CCS (IGCC-VENT) also have considerable learning potential, with COE projected to decrease from the current 60 to 40 €$_{2008}$/MWh in the long term. Furthermore, in the short term, the COE of IGCC-VENT could be competitive with the current market price of 57 €$_{2008}$/MWh. Initial governmental support is, however, needed to realize the first 20 GW_e.

Comparing IGCC-VENT with IGCC-CCS leads to CO_2 capture costs of 32 €$_{2008}$/t CO_2 in the current situation and to 16 €$_{2008}$/t CO_2 in 2050 if SOFCs become competitive. If IGCC-CCS replace pulverized coal power plants or natural gas combined cycles, the CO_2 mitigation costs are currently 44 and 92 €$_{2008}$/t CO_2, respectively. However, these may decrease over time to 7-16 and 4-23 €$_{2008}$/t CO_2, respectively, in 2050.

The use of torrefied wood pellets (TOPS) as feedstock results in an increase in COE, both currently and in the future. However, if the CO_2 tax exceeds 34 €$_{2008}$/t CO_2, state-of-the-art (SOTA) TOPS-fired IGCC-CCS could result in a lower COE compared to SOTA coal-fired IGCC-CCS. If the TOPS price decreases from 6.3 €$_{2008}$/GJ to 3.0 €$_{2008}$/GJ, by 2050, TOPS-fired IGCC without and with CCS could result in lower COE compared to coal-fired IGCC at a CO_2 tax of 6 and 16 €$_{2008}$/t $_{CO2}$, respectively.

Currently, production of FT-liquids is already competitive at an oil price of 77 €$_{2008}$/bbl for FT-liquids produced without CCS and at 83 €$_{2008}$/bbl with CCS, resulting in CO_2 capture costs of only 11 €$_{2008}$/t CO_2. By 2050, production costs of FT-liquids could drop to 9.3 €$_{2008}$/GJ without CCS and 10.0 €$_{2008}$/GJ with CCS, reducing the CO_2 capture costs to

7 €$_{2008}$/t CO_2. To realize this cost reduction, an installed capacity of about 430 GW$_{th\ FT}$ would be needed.

Producing FT-liquids from coal and applying CCS could result in comparable CO_2 emissions (77 kg CO_2/GJ$_{th\ FT}$) as oil-derived transportation fuels (80 kg CO_2/GJ$_{th\ FT}$), while TOPS-based FT-liquids would have no CO_2 emissions. Producing FT-liquids from TOPS and applying CCS could even result in negative CO_2 emissions (-123 kg CO_2/GJ$_{th\ FT}$). However, with the current TOPS price of 6.3 €$_{2008}$/GJ, TOPS-based FT-liquids are more expensive and CO_2 mitigation costs are 123 €$_{2008}$/t CO_2 without CCS and 56 €$_{2008}$/t CO_2 with CCS compared to oil-derived transportation fuels. By 2050, coal-based FT-liquids could be produced at lower costs than oil-derived transportation fuel at an oil price exceeding 47 €$_{2008}$/bbl. Additionally, if TOPS prices decrease to 3 €$_{2008}$/GJ and the CO_2 price exceeds 11 €$_{2008}$/t CO_2, TOPS-based FT-liquids could be more competitive than coal-based FT-liquids by 2050.

The bottom-up analysis and the component-based experience curve analysis gave comparable trends regarding the potential development of efficiency improvement, capital costs and production costs reductions for a scenario with a strong growth in IGCC, FT and CCS capacity. However, when zooming in on the capital cost reduction, some minor differences arose, but these could largely be explained. The main advantage of combining the two approaches is that it becomes clear how cost reductions can be achieved and what kind of capacity development and time frame may be required to reach the projected improvements.

5.1 Introduction

In the last two decades, gasification of coal and biomass has become one of the most promising technologies, especially in combination with carbon capture and storage (CCS), to satisfy future electricity and liquid fuel demand.[24,42,44,45,46,55,195,196] There are two main reasons for this development.

Firstly, integrated gasification poly-generation (IG-PG) facilities can handle on the one hand multiple kinds of feedstock, which makes it possible to adjust, for instance, the biomass / coal ratio to feedstock prices. On the other hand can be adapted to produce multiple kinds of products, like chemicals, synfuels and electricity, where the output mix can be adjusted depending on the products prices.[150] On the short term, Fisher Tropsch (FT) liquids are a promising synfuel because they can be used directly as a substitute for fossil fuel based gasoline and diesel,[24,197] thereby decreasing the dependency on crude oil. However, if coal-based FT-liquids are produced without CCS, the overall life cycle CO_2 emissions will roughly double compared to crude oil-derived liquids.[22,24,172] FT-liquids with CO_2 emissions comparable to conventional diesel can be produced if CCS is applied. Zero to negative CO_2 emissions can be realized if biomass is (co-) gasified.[24,172,174]

Secondly, the costs of electricity (COE) are expected to be lower for IGCC-CCS than for PC-CCS.[71,168,198,199] This would be an advantage because equipping power plants with CCS is expected to become an important technology to reach the target of maximum 2°C increase in average temperature on earth compared to pre-industrial levels.[200] Nevertheless, some recent studies indicate similar or even slightly higher COE for IGCC-CCS than for PC-CCS.[54,201] Additionally, Damen et al., indicate that there are larger uncertainties in the costs and performance of IGCC than of PC.[71]

Despite the advantages that IG-PG offer, the current installed capacity of this technology (with or without CCS) is low. Nowadays, less than 5 GW_e of IGCC and around 28 GW_{th} of FT-synthesis plants are in operation worldwide.[202] The main reason for the low installed capacity is the high capital costs combined with moderate variable costs.[203] For coal-based FT-liquids, the production costs are estimated at 6.3-17 €/GJ_{HHV} without CCS and 7.6-14 €/GJ_{HHV} with CCS (see Table 5.1).[22,23,24,55,204] These production costs are close to the market price of oil-derived fuels, which was on average about 13 €/GJ_{HHV} in 2011[18].[65] The production costs of electricity for IGCC without CCS (IGCC-VENT) are

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[18] In 2011, the crude oil price was on average 72 €$_{2008}$/bbl.[65] Using a conversion factor of 6.12 GJ_{HHV}/bbl, this translates to 11.8 €$_{2008}$/GJ_{HHV}.[204] The crude oil has to be refined to obtain gasoline or diesel. The costs of this are estimated at 38 €$_{2008}$/m^3 diesel.[25] With a volume of 0.159 m^3/bbl, this results in a refinery margin of 6.2 €$_{2008}$/bbl or 1 €$_{2008}$/GJ. Hence, at an oil price of 72 €$_{2008}$/bbl, the costs for fuel derived liquids are estimated at 12.8 €$_{2008}$/GJ_{HHV}.

In the same period, the actual gasoline (based on regular New York harbour conventional gasoline) and diesel (based on New York harbour ultra-low sulphur No. 2 diesel) spot prices were on average 1.9 and 2.0 €$_{2008}$/gallon, respectively, in the USA.[313] By using an energy density of 0.133 GJ_{HHV} per gallon gasoline and 0.143 GJ_{HHV} per gallon diesel,[77] the costs are 14.1 and 14.2 €$_{2008}$/GJ_{HHV} for gasoline and diesel, respectively.

estimated at 23-101 €$_{2008}$/MWh and at 36-141 €$_{2008}$/MWh if CCS is applied (Table 5.2).[50,54,74,199,201,204,205,206] This is considerably higher than the current market price for electricity, which is around 57 €/MWh[19].[66] Hence, production costs have to go down before IGCC (with CCS) will be implemented on a large scale.

Future cost developments for IGCC have been explored using experience curves by Van den Broek et al., and Rubin et al.[168,198] The experience curve concept states that with each doubling of installed capacity, production costs decline with a certain factor. Rubin et al., indicate a capital cost reduction potential of 12-30% to 1.5-1.9 k€$_{2008}$/kW$_e$ for IGCC-CCS after 100 GW$_e$ experience.[198] Van den Broek et al., estimate that the capital costs will reduce by almost 50%, in the period 2001-2050, to 0.9-1.0 for IGCC without CCS (hereafter IGCC-VENT) and 1.3-1.4 k€$_{2008}$/kW$_e$ for IGCC-CCS, respectively. In the same period, efficiencies are projected to increase with 10-12%-points to 48-49%$_{HHV}$ for IGCC-VENT and 44%$_{HHV}$ for IGCC-CCS.[168] For these projections, they use the capacity growth scenarios of the World Energy Technology Outlook 2050.[43]

Future cost reduction can also be estimated by performing a bottom-up engineering study which takes an in-depth look into the technologies and research activities going on in the field.[207,208] This has already been done for IGCC-VENT by Klara[59] and for IGCC-CCS by Gerdes.[58] In the longer term[20], they project that the efficiency would increase significantly to 46%$_{HHV}$ and 40%$_{HHV}$ and specific capital costs would decrease to 1.0 and 1.2 k€$_{2008}$/kW$_e$ for IGCC-VENT and IGCC-CCS, respectively. If fuel cells become commercially available, the capital costs may increase slightly to 1.1 and 1.3 k€$_{2008}$/kW$_e$ but the efficiency would increase to 59%$_{HHV}$ and 56%$_{HHV}$ for IGCC-VENT and IGCC-CCS, respectively.[58,59] Also, Meerman et al., investigate the cost reduction and performance improvements of gasification plants with CCS for the production of IG-PG with a bottom-up analysis. Their results indicate that the efficiency of IGCC-VENT may increase from the current 44% to 59% in the long term, if solid oxide fuel cells can be applied competitive. Similarly, the efficiency for IGCC-CCS is projected to increase from 35% to 55% in the long term. Consequently, the cost of electricity could decrease from 60 €/MWh at present to 40 €/MWh in the long term for IGCC-VENT, with a constant coal price of 2.25 €/GJ. For IGCC-CCS, the production costs may decrease from 82 €/MWh to 50 €/MWh. For coal-based FT-liquids, they predict a cost decrease from 14 to 9 €/GJ without CCS and from 15 to 10 €/GJ with CCS.[204]

The fuel cost derived from the oil price is somewhat lower than the actual costs of gasoline and diesel. Although the latter is of course the best estimation for the costs of gasoline and diesel, the oil price is used for evaluation of the potential of FT-liquids because future scenarios predict the oil rather than fuel prices. The discrepancy between oil and fuel prices can be caused by an underestimation of the refinery margin, a shortage of supply or speculation. Nevertheless, in this study the refinery margin of 1 €$_{2008}$/GJ is used, which may lead to a conservative estimation of the fuel price.

[19] Electricity prices are based on the average Dutch daily market price between 2004-2005. During that period, the price varied between 0 and 1,050 €$_{2008}$/MWh with an average of 57 €$_{2008}$/MWh.[66]

[20] No specific time frame is mentioned for the different cases.

The economic and performance development of FT-liquids has also been analyzed with a bottom-up study by Hamelinck et al.,[77] Eerhart[172] and Wit et al.,[158] Hamelinck et al., indicate that the costs of biomass-based FT-liquids without CCS may decrease from 18 €/GJ_{HHV} at present to 10 €/GJ_{HHV} in 2035, owing to lower biomass prices (3.4 vs. 2.3 €/GJ_{HHV}), economies of scale (400 vs. 2,000 MW_{th}), and a selective diesel catalyst.[77] Eerhart projects a cost decrease from the current 11 to 9.6 €/GJ_{LHV} in 2030 for biomass-based FT-liquids with CCS, mainly due to a better temperature profile of the plant and a selective catalyst. He assumes that the cost of biomass remains at 2.6 €/GJ_{LHV}. For coal-based FT-liquids, Eerhart estimate that production costs could decrease from 8.4 to 6.7 €/GJ_{LHV} at a constant coal price of 1.2 €/GJ_{LHV}.[172] Wit et al., analyze the costs of biomass-based FT-liquids by using a bottom-up approach for scale-dependent learning and an experience curve approach for scale-independent learning[21]. They estimate that the costs can decrease from 21 $€_{2008}$/GJ_{HHV} in 2010 to around 8.5-16 $€_{2008}$/GJ in 2030, mainly due to a larger scale of the plants (150 MW_{th} vs. 3,200 MW_{th}) and 20-35% lower biomass prices.[158] These bottom-up studies, however, do not indicate the growth of installed capacity that is required, and in the case of Gerdes[58] and Klara[59] there is also no indication of the timeframe needed before these projections are expected to be realized.

A comparison of the results of experience curves studies and bottom-up analyses is difficult due to the lack of details about the time frame of mainly the bottom-up analyses. Nevertheless, the bottom-up analysis has the advantage that insights are gained into how the cost reductions could be realized. In contrast, the experience curve approach gives insights into the capacity growth needed before a certain production cost or performance level is reached.[207] The two approaches validate each other results and reduce thereby the risk of over-optimistic (or over-pessimistic) estimations on the cost reduction and performance improvement that can be realized in a specific time frame.[207] Therefore, this study combines a top-down experience curve analysis with a bottom-up study for FT-liquids and IGCC development with and without CCS. The bottom-up study has been described in depth in previous work[204] whereas the component-based experience curve analysis is executed in this study.

This main research question in this study is: ***What are the technological and economic prospects of IG-PG (for power and FT-liquid production) with and without CCS and how may the production costs develop over time?*** Additionally, the methodological objective of this study is to combine a top-down experience curve analysis with a bottom-up study to assess if both methods give comparable results.

The prospects are analyzed over time, starting from the current (2010) situation to the short (2020), medium (2030) and long term (2050). In this study, bituminous coal Illinois #6 and torrefied wood pellets (TOPS) are used as feedstock. For this type of coal, the historical costs are examined and future costs projections are made in Meerman et al.[204]

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[21] Scale dependent learning are the reductions attached to economies of scale only, whereas scale-independent learning are the improvements through technological process development, like the application of advanced materials, increased conversion efficiencies etc.

TOPS are selected because they are considered to be one of the most promising biomass commodities due to their low moisture content and high energy density.[86,209]

The methodology is explained in Section 2. Subsequently, in Section 3, background knowledge, economic and technological data are given for the current situation and the future projections. In Section 4, the results of the experience curve approach are given and are compared with the results of the bottom-up analysis. Finally, a discussion and conclusion are given in Sections 5 and 6.

Table 5.1: Performances and costs mentioned in literature for coal-based FT plants with once-through configuration.

Technology	Technology	Size (1) (MW_{HHV})	Efficiency (2) ($\%_{HHV}$)	Carbon capture ratio (%)	Capital costs ($M€_{2008}$/ $MW_{HHV, installed}$)	Production costs of FT-liquids (3) ($€_{2008}/GJ_{HHV}$)	Source
FT-VENT	SOTA (4)	1,578	51 $36\%_{FT} + 15\%_{e}$	61 (5)	1.3	6.5	Williams et al., 2006[23]
FT-VENT	N^th of a kind plant	3,726	49 $33\%_{FT} + 17\%_{e}$	51(5)	0.95	7.6	Kreutz et al., 2008[24]
FT-VENT	SOTA (2005)	414	59 (6) $49\%_{FT} + 10\%_{e}$	n.a.	1.4	17	Vliet et al., 2009[22]
FT-VENT	Plant of 2020	2,070	59 (6) $52\%_{FT} + 7\%_{e}$	n.a.	0.89	12	Vliet et al., 2009[22]
FT-VENT	SOTA (7)	4,452 (8)	53 $41\%_{FT} + 13\%_{e}$	50 (5)	0.94	6.3	Mantripragada and Rubin, 2011[55]
FT-VENT	SOTA	607	61 $49\%_{FT} + 12\%_{e}$	56 (5)	1.5	13	Meerman et al., 2012[204]
FT-CCS	SOTA (4)	1,544	48 $35\%_{FT} + 13\%_{e}$	61	1.4	7.6	Williams et al., 2006[23]
FT-CCS	N^th of a kind plant	3,558	47 $33\%_{FT} + 14\%_{e}$	51	1.0	8.4	Kreutz et al., 2008[24]
FT-CCS	Plant of 2020	2,070	57 (6) $52\%_{FT} + 5\%_{e}$	n.a.	0.91	12	Vliet et al., 2009[22]
FT-CCS	SOTA (7)	4,377 (8)	52 $41\%_{FT} + 11\%_{e}$	50	1.4	12	Mantripragada and Rubin, 2011[55]
FT-CCS	SOTA	589	67 $52\%_{FT} + 15\%_{e}$	56	1.6	14	Meerman et al., 2012[204]

(1) Size is related to the output and is the sum of the installed capacity for FT-liquids and electricity production.

(2) In the table, three efficiencies are given: total, fuel and electrical efficiency. Due to rounding differences the sum of the electrical and fuel efficiency can deviate from the total efficiency.

(3) The production costs of FT-liquids include revenues from ash and sulphur and exclude CO_2 tax and cost for CO_2 transport and storage.

(4) Williams et al., state that the coal-based technologies are based on "commercially-ready technological components".[23] This is assumed to be state-of-the-art (SOTA) technology.

(5) The CO_2 is captured because this enhances the FT-synthesis reaction. However, the CO_2 is not compressed and stored but vented into the atmosphere.

(6) The efficiencies of van Vliet et al., were expressed in LHV and no conversion between LHV and HHV is mentioned in the article.[22] However, it is stated that the feedstock is bituminous coal. Therefore, a typical LHV/HHV ratio for bituminous coal of 1.035 was used.

(7) In the study of Mantripragada and Rubin it is not mentioned whether the technology is based on SOTA components.[55] However, this is assumed to be the case based on the technology description.

(8) The FT-liquid output is 50,000 barrels/day and the net electricity output is 910 MW_e for FT-VENT and 835 MW_e for FT-CCS.[55] A barrel is assumed to be 6.12 GJ_{HHV}. This leads to a FT-liquid output is 4,452 and 4,377 MW_{HHV}, respectively.

Table 5.2: Performances and costs mentioned in literature for coal-based IGCC based on dry fed gasification technology.

Technology	Technology	Size (MW_e)	Efficiency ($\%_{HHV}$)	Carbon capture ratio (%)	Capital costs ($k€_{2008}/kW_e$)	COE (1) ($€_{2008}/MWh$)	Source
IGCC-VENT (2)	Near term com. available	413	45.7	n.a.	1.6	23	Parsons et al., 2002[199]
IGCC-VENT	SOTA	636	41.1	n.a.	1.4	50	DoE and NETL, 2007[74] (3)
IGCC-VENT	SOTA	762	40.0	n.a.	2.1	70	IEA GHG, 2008[50]
IGCC-VENT	1st of a kind plant	870	43.9 (42.9-44.9)	n.a.	2.9 (2.6-3.3)	94 (86-101) (4)	Mott MacDonald, 2010[201]
IGCC-VENT	Nth of a kind plant (±2020)	870	45.0 (44.0-46.0)	n.a.	2.2 (1.9-2.4)	66 (59-73) (4)	Mott MacDonald, 2010[201]
IGCC-VENT	SOTA	629	42.1	n.a.	2.0	53	DoE and NETL, 2010[205] (3)
IGCC-VENT	Nth of a kind plant	636	41.1	n.a.	1.9	62	GCCSI, 2011[206]
IGCC-VENT	SOTA	880	44.0	n.a.	1.7	60	Meerman et al., 2012[204]
IGCC-CCS (2)	Near term com. available	351	38.7	89	2.6	36	Parsons et al., 2002[199]
IGCC-CCS	SOTA	518	32.0	91	1.9	73	DoE and NETL, 2007[74] (3)
IGCC-CCS	SOTA	656	32.0	85	3.0	94	IEA GHG, 2008[50]
IGCC-CCS	1st of a kind plant (±2013)	870	36.0 (34.0-37.5)	90	3.5 (3.2-3.8)	132 (120-141) (4)	Mott MacDonald, 2010[201]
IGCC-CCS	Nth of a kind plant (±2020)	870	35.1 (33.2-36.6)	90	2.7 (2.4-2.9)	92 (83-102) (4)	Mott MacDonald, 2010[201]
IGCC-CCS	SOTA	497	31.2	90	2.8	78	DoE and NETL, 2010[205] (3)
IGCC-CCS	Nth of a kind plant	517	32.0	n.a.	2.5	78	GCCSI, 2011[206]
IGCC-CCS	Nth of a kind early commercial power plant (2020)	870	37.0 (35.0-39.0)	90	2.8 (2.6-3.0)	72 (68-77)	ZEP, 2011[54]
IGCC-CCS	SOTA	735	38.0	95	2.4	81	Meerman et al., 2012[204]

(1) COE excludes CO_2 transport and storage costs as well as a CO_2 tax but includes revenues from ash and sulphur.

(2) Multiple cases are described in Parsons et al. The case with the Shell gasifier, CGCU and G-gas turbine is included in the table, because this reflects current technology and is based on a dry-fed gasifier.

(3) DoE and NETL, 2010[205] is the updated version of DoE and NETL, 2007.[74] In the updated version, improved modelling algorithms and additional vendor information are used and coal prices changed from 1.1 to 1.0 €/GJ.

(4) In the Mott MacDonald cases, the COE range is mostly caused by differences in fuel price, starting year and discount rate rather than differences in capital costs.

5.2 Methodology

The technological and economic prospects for IG-PG facilities with and without CCS are analyzed for two feedstock cases, namely 100% coal and 100% TOPS. Although 50% biomass is currently the technical limit for entrained flow gasifiers,[150] it is expected that 100% TOPS will be feasible in the future.

In this study, all efficiency figures are HHV-based and all economic data are expressed in $€_{2008}$, unless stated otherwise. Capital cost data in US dollars were first converted to $\$_{2008}$ with the IHS CERA American Power Capital Costs Index (PCCI)[161] and subsequently to $€_{2008}$. For capital costs in Euros, the IHS CERA European Power Capital Costs Index (EPCCI) was used.[161] For feedstock, electricity prices and production costs, the average inflation rate of the European Union or the USA was used.[210,211] The exchange rates used in this study were 1.00 $€_{2008}$= 0.68 $\$_{2008}$ = 0.91 $£_{2008}$.

In this study, CCS is limited to capture and compression (to 110 bar) of CO_2. Transportation and storage is not taken into account.

5.2.1 Gasification technology

A schematic overview of an IG-PG facility based on state-of-the-art (SOTA) technology is given in Figure 5.1. In the pre-treatment section, the feedstock is ground to small particles and supplied to a dry-fed entrained flow gasifier, where it is mixed with steam and a sub-stoichiometric amount of pure oxygen produced in the cryogenic air separation unit (ASU). Under influence of high temperature and pressure, a mixture of mainly CO, H_2, H_2O and CO_2 is formed, which is called raw syngas. The ratio between the chemical energy in the syngas and in the feedstock is called the cold gas efficiency (CGE).

After the gasifier, the raw syngas is cooled from approximately 1600°C to 300°C, to make it suitable for the cleaning process. In this process (see dotted bold box in Figure 5.1), particulates are removed from the syngas by the candle filter, alkali compounds by the wet scrubber and sulphur compounds in the acid gas removal (AGR) section. In the AGR, the sulphur compounds, which are mainly in the form of H_2S, dissolve in a liquid solvent. After regeneration, the sulphur compounds are converted to elementary sulphur in the Claus plant. The Claus plant, however, cannot recover the sulphur percentage which is required by environmental legislation in many countries. Therefore, a Shell Claus offgas treating (SCOT) plant is installed. In this plant, all remaining sulphur is converted back to H_2S which is subsequently removed from the stream and recycled to the Claus unit.

If CO_2 capture is applied or the syngas is used for FT-liquid production, the syngas is led to a water gas shift (WGS) reactor before the AGR. In the WGS, (part of) the CO reacts with H_2O to CO_2 and H_2. The CO_2 is removed after the sulphur compounds in a second AGR, and subsequently compressed to 110 bar for transport.

The clean syngas can then be led to a combined cycle to produce electricity or a FT-synthesis reactor where the CO reacts with H_2 to (long) carbon chains. Subsequently, the formed chains are cracked to gasoline and diesel. During the synthesis and cracking process some short carbon chains ($<C_4$) are formed. These can be recycled to the

FT-synthesis reactor or can be used in a combined cycle. In this study, only the second configuration is considered, since it is reported as a more economical way of FT-liquids production.[24,55,57]

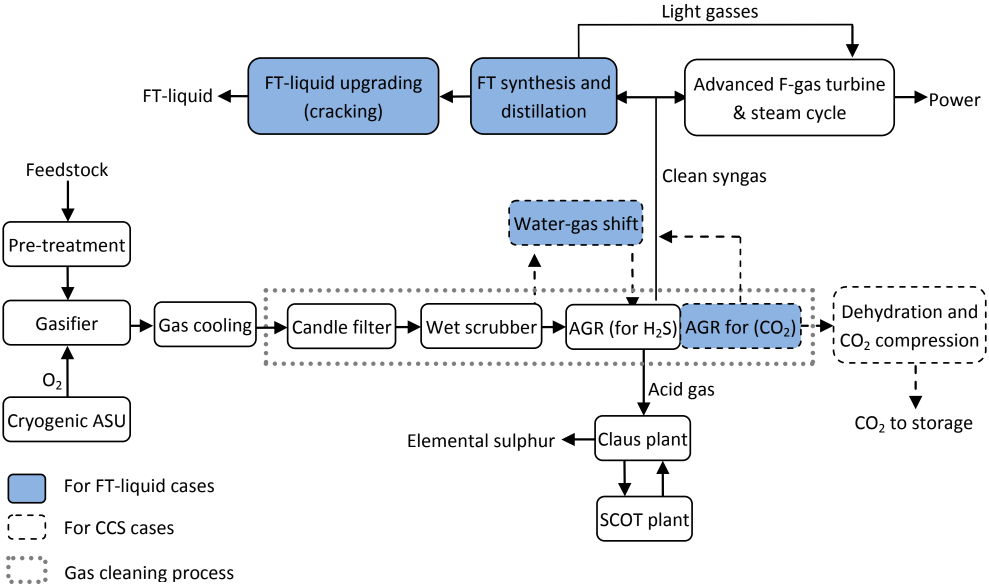

Figure 5.1: Simplified schematic overview of the IG-PG facility (with CCS) based on SOTA technology (based on IEA GHG, Kreutz et al., Meerman et al., and Rubin et al.).[24,50,150,198]

5.2.2 Cost and performance projections with experience curves

Potential cost reductions are often investigated with the so-called experience or learning curve approach (for more information and examples see Junginger et al.,[207,212] and NEEDS.[208] In this approach the relation between costs development and cumulative production of a specific product or technology is analyzed.[212] Most experience curves show that the cost per unit decrease with a certain factor (1-progress ratio) after each doubling of installed capacity. Progress ratios (PR) are technology specific and have to be derived.

FT-liquid production plants and IGCC are still in an early stage of commercialization, as the installed capacities of 28 GW_{th} and 5 GW_e show.[202] Therefore, not enough empirical cost and cumulative production data are available to derive a PR for the complete gasification plant.[168] However, most components are not specific for gasification plants but are also installed in other facilities. For these components, an experience curve can be derived. The experience curves of these separate components can be combined to estimate the costs of the overall facility, see equation 5-1.[213] This so-called component-based experience curve approach is, for instance, used by Van den Broek et al., and Rubin et al., for power plants with CO_2 capture.[168,198]

$$C(x_t)=\left\{C(x_{0,1})*\left(\frac{x_{t,1}}{x_{0,1}}\right)^{b_1}\right\}+\left\{C(x_{0,2})*\left(\frac{x_{t,2}}{x_{0,2}}\right)^{b_2}\right\}+...+\left\{C(x_{0,n})*\left(\frac{x_{t,n}}{x_{0,n}}\right)^{b_n}\right\} \quad (5\text{-}1)$$

where C = the production cost per production unit;
x = the cumulative installed capacity;
the subscripts t and 0 indicate time t and initial time, respectively;
the subscripts 1 to n indicate the different components;
b = the experience index which is technology specific and has to be derived ($\mathrm{PR} = 2^b$).

In order to make future costs projections with a component-based experience curve analysis, information is needed on the PR, the current costs and installed capacity, and projections of future capacities for each component.

5.2.2.1 Progress ratios

Progress ratios are reported in literature for some components, for instance for coal pre-treatment, oxygen production and steam turbines. However, this is not the case for all components, especially for the components specific for gasification plants, such as for the sulphur and CO_2 capture units and syngas turbines. For these components, the PR of analogue processes are used as approximation. For example, the PR of the AGR section is based on the flue gas desulfurization (FGD) process of conventional coal power plants,[198,207] because both are scrubbers for cleaning purposes.

Besides an average PR, also a low and a high PR are used to reflect uncertainty. If an uncertainty range is mentioned in literature (for the analogue technology), this is used. Otherwise, a range of 10% is used (i.e., ± $5\%_{\text{relatief}}$ the average value). The PR is assumed to have an upper limit of 1.0, because it is unlikely that no learning take place when more capacity is built[22]. Nevertheless, components can become more expensive over time, despite the learning effect, due to increasing labour and material prices.

5.2.2.2 Current and future installed capacities

Similar components are installed in multiple facilities, for example, coal pre-treatment capacity is installed in PC power plants as well as in coal-based gasification plants.

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[22] In the past, a few examples can be found where the costs increased, rather than decreased, after a few doublings of installed capacity. However, often a clear reason is behind this cost increase. For instance, the capital cost of nuclear power plants increased significantly over time mainly due to higher safety requirements and increasing complexity.[314]
Although a less mature technology, a cost increase is also observed by offshore wind turbines. This is mainly caused by the less economic favourable location of the turbines (further away from the coast line and in deeper waters), which increased the installation costs and the connection costs to the grid.[207] The capital costs of offshore wind parks near the coast line has decreased over time.[207]

Consequently, the total installed capacity of coal pre-treatment is the sum of the installed capacity of PC power plants and coal-based gasification plants (corrected to the same unit). The analogous technologies for all components are listed in Table 5.3. The installed capacities of these related technologies are used for estimating the current as well as the future installed capacity for each component.

Table 5.3: Overview of the technologies which determine the installed capacity of the components.

	PC		NGCC (1)		IGCC		FT		OGP (2)		Industry (3)	
	CCS	Vent	CCS	Vent	CCS	Vent	CCS	Vent	CCS	Vent	CCS	Vent
Coal pre-treatment	x	x	-	-	x (4)	x (4)	x (4)	x (4)	x (4)	x (4)	x	x (5)
ASU	x (5)	-	-	-	x	x	x	x	x	x	x (6)	x (5)
Gasifier	-	-	-	-	x	x	x	x	x	x	-	-
Gas cleaning	-	-	-	-	x	x	x	x	x	x	-	-
Pre-combustion CO_2 capture	-	-	-	-	x	-	x	x (7)	x	x (7)	-	-
CO_2 compression	x	-	x	-	x	-	x	-	x	-	x	x (5)
Syngas turbine	-	-	x (8)	x (8)	x	x	x	x	-	-	-	-
Steam cycle	x	x	x	x	x	x	x	x	-	-	x (5)	x (5)
FT-synthesis	-	-	-	-	-	-	x	x	-	-	-	-

(1) NGCC stands for natural gas combined cycle.

(2) OGP stands for other gasification plants, meaning gasification facilities other than IGCC or IG-FT. These are mainly gasification plants for the production of chemicals and gaseous fuel.

(3) Industries exclude gasification plants, as these are already included in the OGP. Industries that are attractive for CCS are mainly situated in cement, iron and steel production, and gas processing.

(4) Only 55% of the current gasification plants are based on coal.[202] For further information see footnote 1 of Table 5.8.

(5) Several industrial plants, besides gasification plants, are equipped with a coal pre-treatment unit, ASU or steam cycles. Additional, some CO_2 compressors are installed in the food and chemical industry. For industrial plants it is often not specified if they have a coal pre-treatment unit, ASU, steam cycle or CO_2 compressor. Therefore, these capacities are ignored in the experience curve analysis for the current as well as for the future capacity except for the CO_2 compressor were the current capacity is set on 10 GW_e after Rubin et al.[198] If CCS would be installed on a large scale, the additional future capacity of CO_2 compressors in the industry would have a negligible effect.

For the coal pre-treatment unit, the influence of ignoring the industrial capacity on the projection of the total capital costs would be limited due to the very large current installed capacity (1,536 GW_e) combined with the high PR for coal pre-treatment (0.91-0.97) and the low share of this unit in the total capital costs (7-9%). Also for the steam cycle, the influence is assumed to be small due to the very large current installed capacity (1,957 GW_e).

For the ASU, the underestimation of current and future installed capacity could play a more significant role, mainly due to the low installed capacity (0.28 Mt/d) combined with a modest PR (0.85-0.95) and the higher share in the total capital costs (14-18%). For the influence of this assumption, see section 5.5.

(6) In the future, part of the pulverized coal power and industrial plants with CCS will be based on oxyfuel capture technology. These plants will have an ASU. For further information see footnote 2 of Table 5.8.

(7) In FT-VENT and OGP-VENT, the CO_2 is removed from the syngas to enhance the performance and economics of the overall facility.[24]

(8) Because a natural gas-fired gas turbine is not the same as a syngas turbine, only half of the installed capacity of NGCC will be included in the installed capacity of syngas turbines. For further information see footnote 4 of Table 5.8.

The capacities of the technologies are not always reported in the same unit. To sum them up, they have to be converted to a common unit or expressed as an equivalent. For instance, the current installed capacity for gasification plants is expressed in MW_{th} of the syngas whereas the future projections for IGCC are given in MW_e. To express them both in MW_{th} of the syngas, the MW_e is divided with the electrical efficiency and multiplied with the cold gas efficiency (CGE), see Figure 5.2 for all relations.

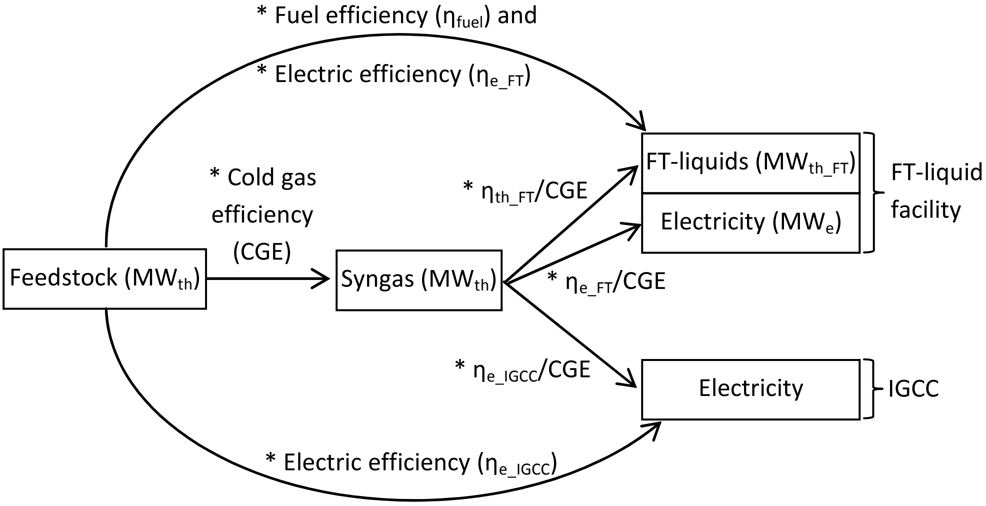

Figure 5.2: Relation between different capacities and efficiencies.

Future projections about the role of different technologies for supplying the future energy demand are presented in a range of studies, see for an overview Fischedick et al.[214] These studies differ with respect to the future energy demand, the allowable CO_2 emissions for a given maximum CO_2 concentration in the atmosphere, and the share of different technologies (CCS, fossil fuels, renewables, nuclear) in the future energy mix.[214] In this study, the future projections of the World Energy Outlook 2010,[27] the Technology roadmap for Carbon Capture and Storage[37] and the Energy Technology Perspectives Study 2008[215] and 2010[4] are used because these studies have the necessary level of detail. In all these future projections, two scenarios are presented, namely the

Baseline / Current Policy scenario and the BLUE / 450 scenario[23]. The Baseline scenario reflects developments that could occur with only the current energy and climate policies in place, whereas in the BLUE scenario additional measurements are taken to limit atmospheric greenhouse gas concentrations to 450 ppm CO_2-equivalent to limit, with a reasonable possibility, the average global temperature increase to 2°C compared to pre-industrial levels.[4]

5.2.2.3 Performance development

In this study, the experience curve approach is not only used for capital cost development but also to estimate performance development. Van den Broek et al., project the performance of IGCC with an experience curve based on the energy loss, i.e., 1-η.[168] However, it is considered better to include the maximum theoretical efficiency, to ensure that thermodynamic limits are not crossed and that the improvement rate will slow down when approaching the maximum efficiency. Since 30% of the exergy is lost during combustion,[216] the experience curve is constructed based on 70%-efficiency for IGCC-VENT and IGCC-CCS[24]. For FT-liquids, the maximum theoretical efficiency is somewhat higher than for IGCC because part of the chemical energy in the feedstock is not combusted but directly converted to chemical energy of the synfuel produced. A maximum efficiency of 85% is assumed for FT synthesis[25]. Furthermore, it is assumed that

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[23] The Energy Technology Perspective Study and the World Energy Outlook are related to each other. The latter is somewhat more extensive but only describes the scenarios till 2035. The Energy Technology Perspective Study extends this to 2050. The Technology Roadmap focuses specific on CCS development from 2010 to 2050.

[24] IGCC-CCS has in theory a slightly lower thermodynamic maximum efficiency than IGCC-VENT. However, for simplicity, the same maximum theoretical efficiency is assumed for VENT and CCS configurations. This leads to an overestimation of the maximum efficiency of about 2.5%-points due to compression process.

The compression energy is calculated with $nRT_0 * \ln\left(\frac{P_{outlet}}{P_{inlet}}\right)$, where n is molar amount, R the gas constant and T_0 the environmental temperature which is assumed to be 283 K.[315] This formula can only be used for an ideal isothermal compression process, a temperature-independent heat capacity and ideal gas behaviour. With this formula, it is calculated that compressing CO_2 from 2.5 bar to 100 bar requires 197 kJ/kg CO_2. Assuming 100% capture efficiency, this corresponds to 461 kJ/kg Illinois #6 coal or around 1.7% of the HHV of the coal. If, however, the inlet pressure is 1.0 bar then the minimal compression energy increases to 246 kJ/kg CO_2 or 575 kJ/kg coal, which is 2.1% of the HHV of the coal.

In SOTA CCS process designs, part of the chemical energy in the syngas is converted to heat in the WGS reactor and energy is used for CO_2 capture. However, the energy used for CO_2 capture can be in principle zero, if membranes are used.[315] Consequently, the only unavoidable energy penalty is the compression which would cost around $2\%_{pt}$, depending on the inlet and outlet pressure.

[25] The maximum efficiency is approached by assuming 100 units input, wherefrom 49 units are converted to FT-liquids, assuming no loss in this conversion step. The remaining 51 units are combusted with a maximum efficiency of 70%, leading to a maximum electricity

the current conversion efficiency of synfuels will not improve over time, and all improvements are caused by an improved electrical efficiency. This simplification is made to avoid allocation issues between the two efficiencies.

5.2.2.4 Pre-learning phase

Several studies have observed that learning often does not start by the first doublings in capacity due to shortfalls in performance and reliability, uncertainties in scaling-up pilot and prototype data and problems arising during full-scale construction and operation.[168,198,207,217] Consequently, in the early phase of technology deployment, the costs are often increasing and the performance may decrease. The duration of this so-called pre-learning phase depends mainly on the complexity of the technology.[198,217] For instance, the costs of natural gas combined cycles increased by about 20% until 14 GW_e was installed, after which the specific capital costs started to decline.[222] A similar trend was observed for flue gas desulfurization (FGD) and selective catalytic reduction (SCR) systems, where the capital costs first more than doubled before they declined. For FGD, costs only started to decrease after an installed capacity of 5 GW_e and for SCR after full commercialization. For power plants with CCS, Rubin et al., estimate the pre-learning phase between 3 and 10 GW_e.[198]

Although all key components studied here are already beyond the pre-learning phase, this is not the case for the overall IGCC. Therefore, it is assumed that the efficiency of IGCC remains constant until the pre-learning phase is crossed. After this phase, the performance is projected to increase according to the experience curve approach with an initial installed capacity of x_t rather than x_0 (see Figure 5.3). The analyzes of the effect of a different pre-learning phase is addressed in §5.5.

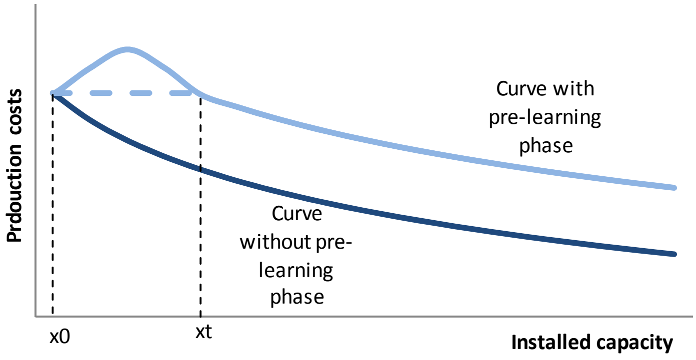

Figure 5.3: The consequences of a pre-learning phase on the experience curve.

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output of 36 units electricity. Consequently, the maximum efficiency is 85%. In reality, the process is more complicated because energy is lost by converting the feedstock to syngas and heat is generated by converting the syngas to FT-liquids.

5.2.3 Cost and performance projections by means of a bottom-up analysis

The main aspects of the bottom-up analysis are described in Meerman et al.,[204] The bottom-up analysis consisted of three steps:

First, the technical and economic performance of a SOTA coal-based 1,000 MW_{th} gasification facility was calculated. An AspenPlus simulation model on a component level was constructed which calculates energy and mass balances. For the costs estimation of each component, data from open literature sources were used. These cost data were corrected to the required scale that followed from the process simulations. The total capital investment is the sum of the scaled component costs, including installation factors, representing e.g., infrastructure, overhead, engineering and contingencies.[150,173,204]

Table 5.4: Overview of the main technologies incorporated in the several process designs.[204] (1)

		Current	**Short term**	**Mid term**	**Long term GT (2)**	**Long term SOFC (2)**
Feeding		Lock hopper	Lock hopper	Solid feed pump	Solid feed pump	Solid feed pump
Oxygen production		Cryogenic ASU	Cryogenic ASU +	ITM	ITM	ITM
Quench	**IGCC-Vent**	Syngas	Syngas	Syngas	Syngas	Syngas
	IGCC-CCS **IG-FT**	Water	Water	Water	Water	Water
Gas Cleaning	**IGCC-Vent**	Selexol Claus	Selexol + Claus	TDS & DSRP	TDS & DSRP	TDS & DSRP
	IGCC-CCS	WGS Selexol Claus	WGS Selexol + Claus	SEWGS TDS & DSRP	SEWGS TDS & DSRP	TDS & DSRP
	IG-FT	WGS Rectisol Claus	WGS Rectisol + Claus	Adv. WGS Rectisol ++ Claus	Adv. WGS Rectisol ++ Claus	N.A.
CO_2 compression		Conventional	Conventional	Shock wave	Shock wave	Shock wave
(Syn)gas combustion		Adv. F-GT	F^+-GT	G-GT	H-GT	SOFC & H-GT
HRSG gasifier		IP steam	IP steam	IP steam	HP steam	HP steam
FT-liquids synthesis		Cobalt-based catalyst	Cobalt-based catalyst	Cobalt-based catalyst	Diesel selective catalyst	N.A.

(1) Incremental improvement of technologies with respect to energetic performance and/or lower costs is indicated by a +. Explanations for the abbreviations used in the table are: ASU: air separating unit; ITM: ion transfer membrane; WGS: water-gas shift; AGR: acid gas removal; SRT: sulphur recovery treatment; TDS: transport desulfurization; DSRP: direct sulphur recovery plant; SEWGS: sorption enhanced water-gas shift; GT: gas turbine; SOFC: solid oxide fuel cell; HRSG: heat recovery steam generation; IP: intermediate pressure; HP: high pressure

(2) To incorporate the uncertainty on the availability of fuel cells, two process designs are developed for IGCC in the long term. One with solid oxide fuel cells (IGCC-SOFC) and one which remains based on advanced syngas turbines (IGCC-GT). No fuel cell case was analyzed for FT-synthesis as the energy gain is very limited due to the low energy content in the gas after FT-synthesis while fuel cells increase capital costs significantly.

Second, promising technologies were identified for the short (2020), medium (2030) and long term (2050) by conducting an extensive literature review and interviews with experts. It is assumed that in the short term only minor improvements are feasible. These will slightly increase performance and reduce capital costs compared to the SOTA configuration. In the medium term, technologies are expected to become commercially available which are currently in the demonstration and pilot stage. Technologies which are currently only available on a laboratory scale are expected to become commercially available in the long term. The technologies with the most positive impact on the production costs are integrated in process designs, taking into account the time frame when the technologies are expected to become available. An overview of the selected technologies for different process designs is given in Table 5.4.

Third, the AspenPlus model developed for the SOTA design was adapted to reflect the selected future process designs. The fuel input of the future concepts was kept constant at 1,000 MW_{th} coal-equivalent. Capital costs were adjusted to describe the costs of the new technologies considered.[204] Subsequently, the production costs for the future concepts were calculated.

Additional information about the economic and technical background of the technologies, process designs, and the modelling can be found in Meerman et al.[204]

5.2.4 Comparison of experience curve based with the bottom-up analysis

The results of the bottom-up and the experience curve analyses were compared with respect to capital costs, efficiency improvement, CO_2 mitigation costs and (levelized) production costs of FT-liquids or electricity (see equation 5-2).

The CO_2 mitigation costs were calculated based on the related technology without CO_2 capture in the same timeframe. Additionally, the CO_2 mitigation costs for TOPS-based FT-liquids were also calculated by taking oil-derived liquids as reference because these would be replaced. Thereby was assumed that all CO_2 emissions are allocated to FT-liquids. Coal-based liquids have higher or comparable CO_2 emissions than oil-derived liquids, and therefore no CO_2 mitigation costs were calculated with oil-derived liquids as reference.

For electricity production, the CO_2 mitigation costs were also calculated with NGCC and PC power plants as reference systems because these two technologies could be replaced if IGCC-CCS are build.[71,218] Also PC and NGCC will improve over time when more capacity is build. Therefore, the future projections for PC and NGCC of Van den Broek et al., were used.[168] To avoid that the results are influenced by differences in discount rate, costs of coal and capacity factor, the projected capital costs and efficiency and the data stated in Table 5.5, were used to calculate the COE and CO_2 emissions of NGCC and PC. These were, subsequently, compared with the projections for COE and CO_2 emissions of coal-based IGCC-CCS and TOPS-based IGCC.

$$LC=\frac{I^{*}(\alpha+OM)+F-(E^{*}EP)}{P} \tag{5-2}$$

where LC = the levelised production costs of electricity (€/MWh) or FT-liquids (€/GJ);
I = the total investment costs in €;
α = the capital recovery factor, which depends on lifetime L (yr) and discount rate r (%) via $\alpha = \frac{r}{1-(1+r)^{-L}}$;
OM = the annual operation and maintenance costs, which is a percentage of total investment costs;
F = the annual feedstock costs (€/yr);
E = the net annual electricity production in the case of FT-liquid production (MWh/yr);
EP = the electricity price (€/MWh);
P = the annual net production of electricity for IGCC (MWh/yr) or synfuels for FT-liquids (GJ_{HHV}/yr).

The overall economic assumptions for this study are given in Table 5.5. These assumptions remain constant over time to focus on the influence of technology development on costs of FT-liquids or electricity produced. However, to incorporate that TOPS production is still at an initial stage of development and learning could reduce biomass prices considerably,[32,85,157,158] two TOPS prices were used throughout this study, namely 3.0 and 6.3 €/GJ.

Table 5.5: Economic assumptions and prices of commodities in this study

Parameter	**Unit**	**Value**
Discount rate	%	10
Capacity factor	%	75
Construction time	Year	3 (1)
Economic lifetime	Year	20 (2)
O&M cost	% of capital costs	4.0 (3)
Coal price	$€_{2008}/GJ_{HHV}$	2.25 (4)
TOPS price	$€_{2008}/GJ_{HHV}$	3.0 (5) | 6.3 (5)
Electricity price	$€_{2008}$/MWh	57 (6)
Sulphur refund	$€_{2008}$/t S	0 (7)
Oil price	$€_{2008}$/barrel	72 (8)
Oil based fuels	$€_{2008}/GJ_{HHV}$	13 (8)
CO_2 emissions oil-derived fuels	kg CO_2/GJ_{fuel}	80 (9)
Carbon content coal	t C/t coal (a.r.)	0.64
Carbon content TOPS	t C/t TOPS (a.r.)	0.51
Carbon content natural gas	t C/t natural gas	0.75
Natural gas price	$€_{2008}/GJ_{HHV}$	6.0

(1) Based on other studies, a construction time of three years was assumed.[24,50,74,204]

(2) The economic lifetime for IGCC varies between 20-45 years[54,59,199,201,215] and for FT synthesis plants between 20-30 years.[23,24] In this study, an economic lifetime of 20 year was chosen to get a conservative estimation of the production costs.

(3) There are two studies that give O&M cost expressed in % of total capital costs for IGCC, namely 2.2%[199] and 3.5-5%.[71] For FT synthesis an O&M percentage of 4.0% is commonly used.[23,24,219]

(4) In the beginning of 2010, coal prices were 56 €$_{2008}$/t, corresponding to 2.25 €$_{2008}$/GJ by assuming an energy density of 25 GJ/t.[1] In this study, the coal price was not expected to decrease because technology of coal winning is already well developed. Additionally, depletion of (cheap) mining sites could increase production costs. Consequently, it is likely that coal prices will increase, rather than decrease, in the future. Nevertheless, to focus on technology development, the coal price is assumed to remain constant over time.

(5) No market prices for TOPS are known because it is currently not produced on a large scale. Uslu et al., indicate that the extra production costs for TOPS compared to normal wood pellets are earned back by the lower transportation costs, leading to similar (or even lower) prices for TOPS than for wood pellets.[86] Therefore, the price of wood pellets was used in this study as an approximation for the TOPS price. In 2008, wood pellet prices were around 120 €$_{2008}$/tonne, which is 6.3 €$_{2008}$/GJ, with an energy density of 19 GJ/t.

    However, several sources indicate technical learning (e.g., improved cultivation techniques, efficient local pre-treatment and economies of scale) may reduce biomass prices in the future.[32,85,157,158] Additionally, it is expected that as the biomass pellet market grows, large price variations become less severe and less frequently.[173] Therefore, also a scenario was analyzed with a TOPS price of 3.0 €/GJ. We refer to Meerman et al., for more information.[173]

(6) Electricity prices are based on the average Dutch spot prices between 2004-2008. During that period, the price varied between 0 and 1,050 €/MWh with an average of 57 €/MWh.[66]

(7) Current practice is that the pure sulphur coming out of the AGR is sold for about 100 €/t S.[204] However, in this study this additional revenue was not taken into account. This leads to approximately 1% higher production costs for coal-based products and of about 0.01% for TOPS-based products.

(8) The oil price in 2011 was on average 72 €$_{2008}$/bbl, with a maximum of 83 €$_{2008}$/bbl and a minimum of 62 €$_{2008}$/bbl.[65] However, in the years before 2011 the oil price varied more strongly. In the period 2008-2011, the oil price varied between 24-96 €$_{2008}$/bbl,[65] which would be a better uncertainty range for the future. By using a refinery margin of 1.0 €$_{2008}$/GJ (see footnote 18), the costs for diesel and gasoline are estimated at 12.8 (4.8-16.7) €$_{2008}$/GJ$_{HHV}$ at an oil price of 72 (24-96) €$_{2008}$/bbl**.**

(9) CO_2 emissions of oil-derived liquids can be divided in two part. First, the CO_2 emissions emitted at the refineries where crude oil is converted into gasoline and diesel. Vliet et al., estimate these on 8.6 kg CO_2-eq per GJ diesel produced.[22] It is assumed that this number is also valid for gasoline. Second, CO_2 emissions are emitted by the combustion of the fuel. These are estimated at 69.3 and 74.1 kg CO_2/GJ for gasoline and diesel, respectively.[220] Hence, in total the CO_2 emissions are 77.9 kg CO_2/GJ$_{gasoline}$ and 82.7 kg CO_2/GJ$_{diesel}$. By assuming that 50% is diesel and 50% is gasoline, the average CO_2 emissions are estimated at 80 kg CO_2/GJ$_{fuel}$.

5.3 Data for the current situation and future projections

5.3.1 Current costs and efficiencies

In literature reported estimations of the costs and performance of SOTA IG-PG plants with and without CCS differ considerably, see Table 5.1 and 5.2. This makes the starting situation ambiguous which increases the uncertainty in future projections. One of the main explanations for the large cost differences are variations in material costs, which increased rapidly from 2005, peaked in the second half of 2008 and decreased again after that.[161]

In this study, the cost and performance estimates of Meerman et al.,[204] were used, because, to our knowledge, this is the only source that incorporates IGCC and FT-liquid synthesis plants and projects the performance in respect to different time frames. For IGCC, the capital costs, efficiency as well as the COE projected by Meerman et al., are in the range projected by other studies. For FT-liquids production, however, the capital costs (1.5-1.6 M€/MW$_e$) and the overall efficiency (59-61%) are higher than from other sources (0.89-1.4 M€/MW$_e$ and 49-59% for FT-VENT and 0.91-1.4 M€/MW$_e$ and 47-57% for FT-CCS). An explanation for this can be the better heat integration of Meerman et al., which increases the electrical efficiency but also increases the capital costs. Additionally, the costs of synfuels with CO_2 capture of Meerman et al., (14 €/GJ) are higher than for the other sources (6.0-12 €/GJ). This can be largely explained by the higher assumed coal price of 2.25 €/GJ compared with 0.8-2.1 €/GJ of the other studies.

5.3.2 Progress ratios and installed capacities

In Table 5.6, progress ratios (PR) are given for the different components based on similar or analogue technologies. Furthermore, a low and a high PR is given to quantify uncertainty. Following Rubin et al., the PR of CO_2 capture, sulphur removal and recovery is based on flue gas desulfurization (FGD).[198] For the gasifier, Rubin et al., use the PR of liquefied natural gas (LNG) production.[198] However, LNG production consists mainly of gas cleaning and cooling with intermediate compression. This process has little similarities with gasifiers. Another analogue technology which could be used as approximation for gasifiers is the PC boiler. PC boilers consist, like gasifiers, mainly of steel, operate at high temperatures and have coal as input. Besides these similarities, there are also significant differences, such as the temperature level, the direction of the incoming flows and the different type of burners in a PC boiler. Nevertheless, PC boilers were used as approximation for gasifiers in this study due to the lack of better alternatives. For FT-liquid synthesis and cracking the PR for biomass-based FT-liquid production was used.[174]

Besides a PR for the capital costs of the different components, a PR is also expected for the electrical efficiency for IGCC and FT facilities. In this study, we used a PR for efficiency development based on the observed values for NGCC and PC power plants.[168,221]

Overviews of the installed capacities for relevant technologies and for each component with the Baseline and the BLUE scenario are given for different time frames in Table 5.7 and Table 5.8. In the Baseline scenario, no substantial additional capacity for IGCC or CCS is installed over time. However, there is significant growth in fossil fuels based FT-liquids up to 430 GW$_{th\ FT}$ in 2050, which are expected to become profitable due to a high oil price. In the BLUE scenario, there is a strong focus on the deployment and implementation of CCS and biomass-based technologies, like FT-synthesis based on biomass equipped with CCS.

Table 5.6: Overview of the progress ratios for the different components and efficiency

Technology	Progress Ratios			Based on	Source
	Low	Average	High		
Pre-treatment (1)	0.91	0.94	0.97	PC boiler	IEA GHG, 2006[217]; Yeh and Rubin, 2007[221]
ASU (2)	0.86	0.90	0.95	Oxygen production	Rubin et al., 2007[198]; IEA GHG, 2006[217]; Junginger et al., 2010[207]
Gasifier (3)	0.91	0.94	0.97	PC boiler	IEA GHG, 2006[217]; Yeh and Rubin, 2007[221]
Sulphur removal & recovery (4)	0.84	0.88	0.92	Flue gas desulfurization (FGD)	Junginger et al., 2010[207]; Rubin et al., 2007[198]
Other cleaning equipment (5)	0.82	0.88	0.94	Selective catalytic reduction (SCR) & FGD	IEA GHG, 2006[217]; Junginger et al., 2010[207]; Rubin et al., 2007[198]
Syngas turbine (6)	0.75	0.90	1.00	Combined cycle	Colpier and Cornland, 2002[222]
HRSG & Steam turbine (6)	0.75	0.90	1.00	Combined cycle	Colpier and Cornland, 2002[222]
CO_2 capture system (4)	0.84	0.88	0.92	FGD	Rubin et al., 2007[198]; Junginger et al., 2010[207]
CO_2 compression (7)	0.95	1.00	1.00	CO_2 compression	IEA GHG, 2006[217]; Junginger et al., 2010[207]
FT synthesis and upgrading (8)	0.86	0.90	0.98	Biomass-based FT-liquids production	Tijmensen et al., 2002[174]; Wit et al., 2010[158]
Efficiency electricity generation (9)	0.93	0.97	1.00	Efficiency of NGCC and PC	Yeh and Rubin, 2007[221]; Van den Broek et al., 2009[168]

(1) No PR could be found for coal pre-treatment in literature. Most often the coal pre-treatment is included in the costs of the PC boiler. Therefore, the PR of PC boilers is used. This is estimated at 0.94 by Yeh and Rubin, without giving an uncertainty range.[221] IEA GHG finds a PR of 0.947 with an uncertainty range of 0.91-0.97.[217] The mean of both values was used for the average scenario, while the lower and upper value were used for the low and high PR scenario, respectively.

(2) Rubin et al., and Junginger et al., indicate a PR of 0.90 without an uncertainty range.[198,207] The IEA GHG reports a PR for O_2 production of 0.90 (0.85-0.95). The values of the IEA GHG were used in this study.

(3) The PR used by Rubin et al., for the gasifier is 0.86 (0.79-0.93) based on liquefied natural gas (LNG) production.[198] However, as explained in the text, pulverized boilers may be a better analogue technology. Yeh and Rubin give a PR of 0.94 for PC boiler but they do not mention an uncertainty range.[221] IEA GHG finds a PR of 0.947 and an uncertainty range of 0.91-0.97.[217] The average of the value of IEA GHG and Yeh and Rubin was used for the average scenario, while upper and lower value were used for the low and high PR scenario, respectively.

(4) Rubin et al., use FGD as analogue technology for sulphur removal and recovery.[198] The PR of FGD is estimated at 0.89 and no uncertainty range is mentioned. Junginger et al., indicate a PR of 0.87 for sulphur removal and recovery.[207] The average of these two was used for the average PR scenario and, as indicated in the section on methodology, a 10% range is used to incorporate uncertainty.

Currently, sulphur capture and CO_2 capture are based on similar technologies. Therefore, FGD was also used as analogue technology for CO_2 capture.

(5) Rubin et al., use FGD and SCR as analogue technologies for other cleaning equipment, the PR for FGD is 0.89 and for SCR 0.88.[198] They do not mention an uncertainty range. IEA GHG gives a PR of 0.88 (0.82-0.94) for FGD/SCR.[217] Junginger et al., estimate the PR for FGD/SCR at 0.87.[207] The IEA GHG values were used here, because this is the only one which gives an uncertainty range, in which the values from the other studies are incorporated.

(6) The PR for combined cycles varied over the years. In the period up to 1991, the PR was more than 1.0 meaning that prices increased over time.[222] From 1991-1997, the experience curve showed a strong cost decline with a PR of approximately 0.75.[222] According to Colpier and Cornland, it is unlikely that this strong cost reduction will continue into the future. Therefore, they suggest to use for future projections a PR of 0.90.[222] No uncertainty range is mentioned in the source. In this study, the lowest value (0.75) mentioned was used for the low PR scenario and the PR for the high PR scenario was set to 1.0.

(7) A PR of 1.0 is derived for CO_2 compression with an uncertainty range of 0.95 to 1.0 in IEA GHG.[217] Also Junginger et al., state a PR of 1.0 for CO_2 compression.[207]

(8) Wit et al.,[158] divide learning for biomass-based FT-liquids production in two parts, a scale-dependent and a scale-independent learning component. The scale dependent component is analyzed with $\frac{Costs\ 2}{Costs\ 1} = \left(\frac{Scale\ 2}{Scale\ 1}\right)^R$, where R is the scale exponent which is 0.85 for facilities larger than 900 MW_{th}. This will lead to a specific cost decrease of 20% if the scale is doubled, which will typically happen every 3-5 years until a certain scale is reached. The scale independent progress ratio is estimated at 0.98. So, if the scale of the facility will not change then the costs will decrease with 2% if the total installed capacity doubles.[158]
Tijmensen et al., estimate the PR at 0.90 for biomass-based FT-synthesis.[174] This value was used as average PR, while the low PR was assumed to be 5% lower than the average (0.86). As high PR, the scale independent PR of 0.98 of Wit et al., was used.

(9) The average PR of NGCC and PC was used as approximation for electrical efficiency development. The PR for PC plants is estimated at 0.98 (0.98-1.00) by Yeh and Rubin.[221] For NGCC, the PR is estimated at 0.95 (0.93-0.98) by Van den Broek et al.[168] For the PR range of IGCC and electricity generation for FT-liquids production, the highest and lowest value of NGCC and PC were used in this study to incorporate uncertainty.

The number of doublings in capacity and the PR are key for determining the impact of learning in the coming decades of a specific technology. For example, the ASU and steam turbines both have a PR of 0.90 in the BLUE-average PR-scenario, but the capacity of ASU doubles 6.4 times and the steam turbines 0.8 times, indicating more learning for ASU than for steam turbines.

As explained in §5.2, costs and performances do often not improve by the first doublings but after the pre-learning phase.[168,198,207] Rubin et al., estimate the pre-learning phase of IGCC-CCS at 7 GW_e.[198] In our study, a pre-learning phase of 7 GW_e was assumed for the capital cost as well as the efficiency development of all components of IGCC-CCS, IGCC-VENT and FT-synthesis[26]. Consequently, the efficiency of IGCC will remain at the same level, until the installed capacity is above 7 GW_e.

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[26] For the pre-learning phase of the ASU, gasifier and FT-synthesis, the 7 GW_e has to be converted to Mt O_2/d, $GW_{th\ syngas}$ and $GW_{th\ FT}$, respectively. For the ASU, 7.5 t O_2/d is required for each MW_e of coal-based IGCC installed.[226] Consequently, the pre-learning

Table 5.7: Current and projected installed capacities for the technologies in two scenarios.

		2010	2020		2030		2040		2050	
Technologies & Components	**Unit**	**Current**	**Base**	**BLUE**	**Base**	**BLUE**	**Base**	**BLUE**	**Base**	**BLUE**
IGCC-VENT (1),(2)	GW$_e$	4.2	4.2	13	4.2	33	4.2	50	4.2	332
• **Coal-based**	GW$_e$	4.2	4.2	13	4.2	33	4.2	46	4.2	283
• **Biomass**	GW$_e$					0.4		4.3		49
IGCC-CCS (1),(2)	GW$_e$			8.0		71		196		288
• **Coal-based**	GW$_e$			8.0		71		194		271
• **Biomass**	GW$_e$					0.1		1.4		16
PC-VENT (3)	GW$_e$	1,510	2,180	2,060	2,743	2,196	3,393	2,135	4,198	2,031
PC-CCS (2),(3)	GW$_e$	0.1	0.3	8.0	0.6	71	0.8	317	1.0	472
• **Oxyfuel**	GW$_e$					42		165		242
• **Post-comb.**	GW$_e$	0.1	0.3	8.0	0.6	29	0.8	152	1.0	229
NGCC-VENT (3)	GW$_e$	1,230	1,667	1,661	2,027	1,841	2,520	1,924	3,132	2,090
NGCC-CCS (3),(4)	GW$_e$	0.1	0.3	3.5	0.6	48	0.8	221	1.0	345
• **Oxyfuel**	GW$_e$					22		109		171
• **Post-comb.**	GW$_e$	0.1	0.3	3.5	0.6	26	0.8	112	1.0	174
Biomass CHP-CCS (2)	GW$_e$	0.1	0.1	0.5	0.1	2.2	0.1	9.5	0.1	36
Industry-CCS (5)	GW$_e$			27		87		279		894
• **Oxyfuel**	GW$_e$			14		44		140		447
• **Post-comb.**	GW$_e$			14		44		140		447
• **Pre-comb.**	GW$_e$			1.0		3.7		14		52
OGP-CCS (5),(6)	GW$_{th}$			2.2		7.7		28		102
OGP-VENT (6)	GW$_{th}$	43	69	67	112	104	179	151	289	187
FT-VENT (7)	GW$_{th\ FT}$	28	66	32	136	42	283	71	439	110
• **CtL**	GW$_{th\ FT}$	19	38	21	76	24	151	33	227	42
• **GtL**	GW$_{th\ FT}$	9.4	28	11.3	44	11.3	88	15	132	19
• **BtL**	GW$_{th\ FT}$	0.1	0.1	0.1	17	6.8	43	23	80	49
FT-CCS (7)	GW$_{th\ FT}$			15		54		170		325
• **CtL**	GW$_{th\ FT}$			7.6		20		55		91
• **GtL**	GW$_{th\ FT}$			7.6		7.6		23		38
• **BtL**	GW$_{th\ FT}$					27		92		197

(1) The current capacity for IGCC was taken from the Gasification World Database. This database contains data on all kinds of gasification plants based on coal or other fossil fuels.[202] No projections are made for IGCC in the Energy Technologies Perspective Study 2010, the Technology roadmap or in the World Energy Outlook 2010. Therefore, the projections for coal-based and biomass-based IGCC were derived from the Energy Technology Perspective Study 2008.[215] In the Baseline scenario, IGCC technology is projected to fail to develop. In the

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phase of the ASU is 0.05 Mt O_2/day. For the gasifier, the 7 GW$_e$ has to be divided by the electric efficiency of IGCC and multiplied with the CGE. This results in a pre-learning phase of (7/ 0.44 * 0.82=) 13 GW$_{th\ syngas}$. For the FT-synthesis, the 7 GW$_e$ has to be divided by the electric efficiency and multiplied with the fuel efficiency. This results in a pre-learning phase of (7/0.44*0.49=) 8 GW$_{th\ FT}$.

BLUE scenario, IGCC technology is expected to be competitive by 2030. Consequently, more than 100 GW_e coal-based IGCC is projected to be installed in 2030 and over 550 GW_e in 2050.[215] Furthermore, the installed capacity of biomass-based IGCC could increase to 0.5 GW_e in 2030 and 65 GW_e in 2050.[215] The figures for 2020 and 2040 were estimated by assuming constant growth rates.

The development of petcoke and petroleum-based IGCC was assumed to be negligible.

(2) The current capacity of IGCC-CCS was assumed to be zero. In the Baseline scenario, CCS is expected to be hardly implemented. In 2035, only 1.5% of total electricity generation is projected to apply CCS,[27] which is used for enhanced oil recovery (EOR) or demonstration purposes.[4] However, the share of IGCC-CCS in these demonstration plants and for EOR purposes was assumed to be negligible.

Two studies make future capacity estimation for coal-based electricity generation with CCS for the BLUE scenario. The IEA Roadmap study projects coal-based CCS capacity of 19 and 743 GW_e in 2020 and 2050, respectively.[37] The World Energy Outlook study projects an increase in coal-based CCS capacity of 13 and 408 GW_e in the period 2010-2020 and 2021-2035, respectively.[27] In this study, the future CCS capacity for 2020 was set at the average of the two studies. For 2050, the value of the Roadmap capacity was used. The capacities in 2030 and 2040 were estimated using two constant growth rates, one based on the period 2020-2035 and the other on the period 2035-2050.

Coal-based power production with CCS can be based on oxyfuel, post-combustion or pre-combustion capture, where the former two will be PC and the latter IGCC. Oxyfuel based power plants are expected to be introduced from 2020 onwards, dominating over post-combustion capture within 15 years (in 2035).[4] Furthermore, it is stated in the Energy Technology Perspective 2010 study that IGCC-CCS predominates over post-combustion PC-CCS due to slightly lower costs. This is opposite to the conclusions of the Roadmap study, which state that in the near-term CCS will mainly be based on post-combustion capture because it is the most established technology.[37] Over time, the share of other capture technologies will increase.[37] To combine these two insights it was assumed that in the first 20 years pre-combustion based IGCC-CCS and post-combustion PC-CCS will equally share the projected CCS capacity. From 2030 onwards, when oxyfuel technology is implemented, the share of pre- and post-combustion will diminish. Consequently, it was assumed that one third of the coal-based CCS capacity will be pre-combustion IGCC, one third post-combustion PC and one third oxyfuel PC.

Besides coal-based capacity, the IEA roadmap estimate that 52 GW_e of biomass-based power generation will be equipped with CCS in the BLUE scenario in 2050. No figures are given in the Roadmap study or in other studies, for the current situation or the development over time. Therefore, it was assumed that the biomass capacity with CCS will grow linearly from the current estimated capacity of 0.1 to 52 GW_e in 2050. The biomass-based CCS capacity will mainly be installed in combined heat and power plants (CHP),[37] where CCS was assumed to be solely based on post-combustion technology. However, part of the CCS capacity will probably also be installed at biomass-based IGCC. It was assumed that 25% of the biomass-based IGCC capacity will be equipped with CCS from 2030 onwards.

(3) For the BLUE scenario, the capacity additions for coal-fired and gas-fired power plants are given with and without CCS for the period 2010-2020 and 2021-2035 in the World Energy Outlook 2010.[27] For instance, the World Energy Outlook projects, for the BLUE scenario, capacity additions of 575 and 438 GW_e of coal-based power plants (PC and coal-based IGCC with and without CCS) in the period 2010-2020 and 2021-2035, respectively. The capacities for 2030, 2040 and 2050 were estimated by interpolation and extrapolation, by assuming a constant growth rate based on the period 2020-2035.

For the Baseline scenario, no capacity additions are given but the electric installed capacity for coal-fired and gas-fired power plants is given in the appendix from the World Energy Outlook

2010.[27] Using the electric installed capacity instead of capacity additions ignores the plants that are replaced. However, this amount will be small compared to the number of new plants installed. The electric installed coal-based capacity in the Baseline scenario is 2,184, 2,748 and 3,056 GW_e in 2020, 2030 and 2035, respectively.[27] The figures for 2040 and 2050 were estimated by assuming a constant growth rate based on the period 2030-2035.

For both future scenarios, all gas capacity was assumed to be NGCC whereas coal-based capacity can be IGCC as well as (sub-, super- or ultra-critical) PC. The coal-based IGCC capacity is projected by the Energy Technology Perspective 2008 study (see note 1) and the rest was assumed to be PC.

(4) In the Baseline scenario, CCS will only be implemented for EOR or for demonstration purposes.[4] However, the share of NGCC-CCS in this scenario was assumed to be negligible.

For the BLUE scenario, two studies make future capacity estimation for gas-based electricity generation with CCS, which was assumed to be all NGCC-CCS. The IEA Roadmap study predicts gas-based CCS capacity of 3 and 345 GW_e in 2020 and 2050, respectively.[37] The World Energy Outlook study predicts an addition in NGCC-CCS capacity of 4 and 173 GW_e in the period 2010-2020 and 2021-2035, respectively.[27] As with the coal-based CCS capacity, in this study, the average of both studies was taken for the capacity in 2020 and the Roadmap capacity was used for 2050. Furthermore, the capacities for 2030 and 2040 were estimated using two constant growth rates, one based on the period 2020-2035 and one on the period 2035-2050.

Pre-combustion can be used in NGCC-CCS but it is more expensive compared to post-combustion and oxyfuel.[223] Furthermore, Kanniche et al.,[223] and ZEP[54] state that oxyfuel for NGCC is more expensive than post-combustion and is not as far developed. However, if oxyfuel is successively implemented in PC plants and hydrogen turbines are developed for IGCC-CCS, it is not a large step to include oxyfuel technologies in NGCC-CCS. In the ZEP report, it is pointed out that oxyfuel may have a small cost advantage compared to post-combustion with second generation CCS technologies.[54] Therefore, it was assumed that from 2030 onwards the new installed capacity for NGCC-CCS will use for 50% post-combustion and for 50% oxyfuel.

(5) In the Baseline scenario, CCS fails to develop commercially in the power as well as in the industrial sector.[4] In the BLUE scenario, CCS is installed in the industry and this may abate 167 and 4,508 Mt CO_2 per year in 2020 and 2050, respectively.[37] As in the power sector 22 GW_e CCS capacity is projected to be installed in 2020, which will abate 131 Mt CO_2 per year according to the Roadmap study,[37] this results in an equivalent of 28 GW_e (= 22 GW_e/131 Mt CO_2/yr *167 Mt CO_2/yr) which would be installed in the industry in 2020. Similar, the industrial CCS capacity in 2050 is estimated to be 946 GW_e-equivalent. 7% and 11% from the abated CO_2 is projected to originate from chemical production in 2020 and 2050, respectively.[37] It was assumed that half of this capacity will be realized in gasification plants and will be based on pre-combustion capture whereas the other half, as well as other sectors (cement, iron & steel, pulp & paper and natural gas processing) will be divided equally between post-combustion and oxy-fuel. Hence, an equivalent of 1.0 and 52 GW_e CCS capacity based on pre-combustion technology is installed in the industry in 2020 and 2050, respectively. The figures for 2030 and 2040 were found by assuming a constant growth rate.

(6) The current installed capacities in MW_{th} of the syngas of other gasification plants (OGP), their main product not being electricity or FT-liquids, were taken from the Gasification World Database.[202] All these plants were assumed not to be equipped with CCS.

For the future projections, historical growth rates since 1990 were used for capacity projections of gaseous fuel (2.6%) and chemicals (5.0%). Part of these future OGP plants will be equipped with CCS, namely the chemical industry plants using pre-combustion CO_2 capture, see note 4. The amount of OGP-CCS derived in note 4 were, however, expressed in equivalent of power plants in GW_e. These have to be converted to GW_{th} of syngas before they can be added to the current installed capacities. For this conversion, the current CGE and the projected electric efficiency of IGCC-CCS, using an average PR, were used. This gives an OGP-CCS capacity of 2.2 and 102 GW_{th} in 2020 and 2050, respectively.

(7) The current installed capacities of coal-to-liquids (CtL) and gas-to-liquids (GtL) were taken from the Gasification World Database in GW_{th}.[202] The projected future capacities for CtL and GtL are stated in the World Energy Outlook in million barrels per day of oil equivalent. Other fossil fuels based FT-liquid capacities (from petcoke, petroleum residues etc.) were assumed to be negligible at present as well as in the future for both scenarios. The million barrels per day were converted to GJ_{HHV} (by multiplying with 6.12 GJ_{HHV}/barrel)[224] and subsequently to $GW_{th\ FT}$ with the projected FT efficiency (with an average PR) and an average capacity factor of 75%.
For biomass-to-liquids (BtL), no production facility is included in the Gasification World Database.[202] However, there is one small biomass-to-liquid pilot plant of 1 MW_{th} and one demonstration plant of 45 MW_{th}, both situated in Germany.[225] Besides these plants other small scale BtL plants are probably installed. Therefore, the current production capacity was estimated at 0.1 $GW_{th\ FT}$. For BtL, the installed capacity is not stated in the World Energy Outlook, as it falls together with conventional technologies and other advanced technologies such as bio-liquids from algae. However, it is stated that advanced biofuels will develop from 2020 onwards. So, the production capacity was estimated to remain at 0.1 $GW_{th\ FT}$ until 2020. Furthermore, it is stated in the World Energy Outlook that 17% of the cumulative investment in biofuel production facilities will be spend to advanced biodiesel production facilities and 5% to bio-based aviation fuel. It was assumed that 20% of the amount of biofuels produced in 2050 will be based on gasification. Due to the time needed for the replacement of conventional to advanced biofuels, a linear increase to 20% was assumed.
In the World Energy Outlook, no figures are given on the shares of CtL, GtL and BtL that are equipped with CCS.[27] However, it is stated that it is a relatively cheap capture option. Therefore, it was assumed that 80% of the new installed FT-liquids capacity (regardless the feedstock) will be equipped with CCS in the BLUE scenario. In the Baseline Scenario, CCS fails to develop commercially, hence no CCS capacity was assumed to be installed at FT-liquids facilities.

Table 5.8: Current and project future installed capacities for the different component in both future scenarios.

Technologies & Components	Unit	2010	2020		2030		2040		2050	
		Current	Base	BLUE	Base	BLUE	Base	BLUE	Base	BLUE
Coal pre-treatment (1)	GW_e	1,537	2,231	2,129	2,841	2,442	3,579	2,830	4,483	3,277
ASU (2)	Mt/d	0.3	0.6	0.9	1.1	3.7	2.1	12	3.2	25
Gasifier and cleaning section (3)	GW_{th}	86	176	177	336	468	650	1,054	1,023	2,128
Syngas turbine (4)	GW_e	401	551	558	684	704	880	926	1,117	1,344
Steam turbine & HRSG (5)	GW_e	1,950	2,789	2,674	3,493	2,997	4,341	3,355	5,388	3,698
CO_2 capture (6)	GW_e	16	39	41	80	143	160	362	254	593
CO_2 compression (7)	GW_e	3.0	8.0	52	13	289	18	1,049	23	2,168
FT-synthesis	$GW_{th\ FT}$	21	58	40	129	89	275	234	431	427

(1) Coal pre-treatment capacity is installed in PC, CTL, coal-based IGCC and OGP. About 55% of all gasification plants are currently running on coal.[202] This percentage was used for the future projections of OGP. To go from GW_{th} of the syngas for the chemical industry and $GW_{th\ FT}$ for the fuel production to GW_e, the current CGE and the projected electric efficiency for the IGCC-VENT with an average PR was used for all cases.

(2) The capacity for ASU is expressed in Mt O_2 per day. To calculate the capacity of the ASU, an oxygen consumption of 7.5 and 20 t O_2/d per MW_e installed is used for coal-based IGCC and

oxyfuel plants, respectively.[226] These values were assumed to be also valid for industrial oxyfuel plants and biomass-based IGCC.

For oxyfuel based NGCC, no oxygen consumption is given in literature. In this study, an oxygen consumption of 13.4 t O_2/d per MW_e was used, assuming two molecules of O_2 for each natural gas (methane) molecule, a heating content of 54 MJ/kg, an overall electric efficiency of 50% and a 5% excess in oxygen.

To estimate the ASU capacity for FT-liquids and OGP, the capacities in GW_{th} of the syngas were converted to an equivalent IGCC-CCS in GW_e with the projected efficiency development with an average PR. These capacities were multiplied with the oxygen consumption of IGCC plants.

Besides in gasification and oxyfuel plants, ASU capacity is also installed in some other industrial plants, nowadays and in the future. However, it is unknown how large these capacities are. Therefore, these were ignored in the experience curve analysis. The influence of this assumption was analyzed in section 5.5.

(3) The capacity for gasifiers and cleaning is expressed in GW_{th}. The capacity of IGCC-VENT and IGCC-CCS were converted to GW_{th} with the current CGE and the projected efficiency with an average PR. Similarly, the capacity of FT-liquid production in $GW_{th\ FT}$ was converted with the current CGE and projected fuel efficiency with an average PR.

Besides in gasification plants, (part of the) cleaning equipment is also installed in some other industrial plants. However, it is unknown how large this capacity is. Therefore, it was ignored in the experience curve analysis.

(4) The installed capacity of syngas turbine is determined by IGCC, FT-synthesis and partly by natural gas-fired gas turbines. On the one hand, syngas turbines have somewhat other characteristics than natural gas-fired gas turbines.[227] On the other hand, there are learning spillovers between natural gas-fired gas turbines and syngas turbines.[168,198] In this study, these two effects were balanced by assuming that the initial capacity of syngas turbines was the installed capacity of syngas turbines plus the half of the installed capacity of gas turbines in NGCC plants.

Note that the given installed electric capacity of NGCC, IGCC and FT-liquids is generated partly by (syn)gas turbines and partly by steam turbines. In NGCC, 65% of the electricity generated comes from the gas turbines, while steam turbines generate 36% of the electricity. The excess of 1% is consumed in the power plant.[228] For IGCC, 74% is generated by (syn)gas turbines, 43% by steam turbines and 17% is consumed internally.[228]

For FT-liquids, the installed capacity of syngas turbines was found by dividing the $GW_{th\ FT}$ capacity with the fuel conversion efficiency and multiplying with the electric efficiency for FT-liquids production without CCS projected with the average PR scenarios. Subsequently, the same (syn)gas - steam turbine division was used as for IGCC.

(5) The future installed capacity of steam turbines is based on FT-liquid production, NGCC, IGCC and PC. For the conversion of installed capacity of the overall technology to installed capacity of steam turbines for NGCC and IGCC, see note 4. With PC, all capacity is generated with steam turbines.

(6) The currently installed capacity of pre-combustion CO_2 capture is mainly located in the production process of chemicals and the production of FT-fuels to enhance the kinetics of the reactions and improve the quality of the FT-liquids. Not all carbon is removed, as part of the carbon ends up in the products. It was assumed that on average 60% of the carbon normally captured for an IGCC-CCS is removed for the FT and OGP processes, nowadays and in the future. The FT capacity in $GW_{th\ FT}$ was converted to GW_{th} fuel input with the projected fuel efficiency assuming an average PR. Subsequently, the capacities in GW_{th} input were converted to GW_e with the projected efficiencies with an average PR of IGCC-CCS. The resulting current capacity for pre-combustion CO_2 capture (16 GW_e) is somewhat larger as the capacity estimated by Rubin et al., which is 10 GW_e.[198]

(7) Currently, CO_2 compression capacity is mainly installed in the food and chemical industry. Rubin et al., estimate this capacity at 10 GW_e and this figure was also used in this study.[198] In the Baseline scenario, it is stated that some CCS capacity will be installed for EOR. Furthermore, some additional CO_2 compression capacity will be installed in the industry. Therefore, it was assumed that the CO_2 compression capacity increases linearly from 10 to 30 GW_e in 2050.
In the BLUE scenario, the CO_2 compression capacity is the sum of the capacity of NGCC-CCS, PC-CCS, IGCC-CCS, industry-CCS, OGP-CCS and FT-CCS. The latter two capacities are similar to the pre-combustion CO_2 capacity, explained in footnote 6 of this table. In the BLUE scenario some additional CO_2 compression capacity will be installed for other purposes than CCS. However, there are no figures for this and was therefore ignored. Nevertheless, this additional CO_2 compression capacity was expected to be small compared to the projected installed capacity for CCS in the BLUE scenario.

5.4 Results of the experience curve analysis

This section presents the results of the experience curve approach for IGCC and FT-liquid production for coal and TOPS-based gasification. These results are compared to the results of the bottom-up analysis conducted by Meerman et al.[204] For IGCC, the projected efficiency, capital costs, COE and CO_2 mitigation cost development for IGCC are discussed in §5.4.1-§5.4.3. For FT-liquids, the main results are given in §5.4.4. Detailed tables with values for all time periods can be found in §5.7.

5.4.1 Efficiency development of IGCC

The efficiency development for IGCC projected by the bottom-up analysis as well as by the experience curve approach is given in Figure 5.4. Note that most scenarios from the experience curve approach remains on the current efficiency level of 35% and 44% for IGCC-VENT and IGCC-CCS, respectively. There are two main reasons for this. Firstly, in the Baseline scenario there is almost no IGCC capacity constructed and consequently the pre-learning phase of 7 GW_e is not crossed. Secondly, the PR in the high PR scenario is estimated at 1.0 indicating that the efficiency will not improve with increasing installed capacity.

The bottom-up assessments for the efficiency of IGCC-CCS and IGCC-VENT are more or less comparable with the BLUE-average PR line (see Figure 5.4), which reflects a strong growth in IGCC, FT and CCS capacity combined with an average PR. In the long term, the efficiency of IGCC-VENT with gas turbines from the bottom-up assessment is somewhat higher (52%) than expected from the experience curve (49%). There are two main reasons for the increase in efficiency for IGCC-VENT from the medium to long term in the bottom-up assessment. Firstly, the assumption that heat recovered in the gasifier and syngas cooler will be used for the generation of high instead of intermediate pressure steam. Secondly, the steam cycle is upgraded in the long term from sub- to supercritical. IGCC-VENT profits much more from this upgrade than IGCC-CCS because more high pressure steam is available.

For the SOFC cases the efficiencies estimated with the bottom-up assessment, namely 59% for IGCC-VENT and 55% for IGCC-CCS, are outside the range projected by the experience curve analysis. However, the ranges projected by the experience curve analysis are only valid if the process remains based on similar technologies. This is not the case with fuel cells, as they offer a different way of power generation than syngas turbines. The high efficiencies projected by the bottom-up analysis are comparable with the efficiencies quoted in literature for IGCC with SOFCs, namely 59.5% without CCS[59] and 56.3% with CCS.[58]

If TOPS is gasified instead of coal, the efficiency trends will not change. By assuming a similar growth and PR, the coal case will always give a slightly higher efficiency than the 100% TOPS case, due to the lower SOTA efficiency of TOPS-based IGCC. However, in the BLUE-low PR scenario, the efficiency penalty of using TOPS instead of coal in an IGCC-CCS decreases with 0.5$\%_{pt}$ in the period 2010-2050.

5.4.2 Capital costs of IGCC

The capital cost development for coal and TOPS-based IGCC from the experience curve approach is shown in Figure 5.5. Both show a similar trend, namely a cost decrease of 10 to 58% for TOPS-based and of 11 to 58% for coal-based IGCC. In 2020, the projections of coal-based IGCC of the bottom-up analysis are comparable with the estimations of the Baseline-Average PR scenario. Note that although there is almost no capacity growth in IGCC in this scenario, the costs still decrease due to growth in gasification plants for the production of FT-liquids and chemicals. From 2020 onwards, the estimations of the bottom-up analysis follow more closely the results from the BLUE-Average PR scenario. The reason for this is that many new technologies, like ITM, solid feed pump and SEWGS, are expected to be implemented in the medium term, which could reduce the capital costs, significantly.

For IGCC-VENT and IGCC-CCS, capital costs of the SOFC and GT configurations in the long term are comparable and within the range projected by the experience curve analysis. For IGCC-VENT, there is a cost advantage for the GT configuration of 0.2 k€/kW$_e$, which results in capital costs of 1.2 k€/kW$_e$ for IGCC-VENT-GT. However, if CCS is applied, the SOFC configuration could have a cost advantage of 0.2 k€/kW$_e$, resulting in capital costs of 1.6 k€/kW$_e$ for IGCC-SOFC-CCS. Overall, the projected range in capital costs with the experience curve approach is considerably larger for IGCC-CCS than for IGCC-VENT, namely 1.0-2.1 k€/kW$_e$ against 0.9-1.6 k€/kW$_e$ in 2050.

The contribution of the different components to the capital cost decrease is given in Figure 5.6, for the component-based experience curve with the BLUE-average PR scenario as well as the bottom-up analysis. Both approaches project that the largest capital costs reduction can be realized in the power section, gasifier and ASU, complemented with the CO_2 capture unit for IGCC-CCS. The bottom-up assessment indicates also a cost reduction in the coal pre-treatment, which is absent in the experience curve approach, due to limited amount of doublings in the installed capacity. Possible explanations are discussed in §5.5.

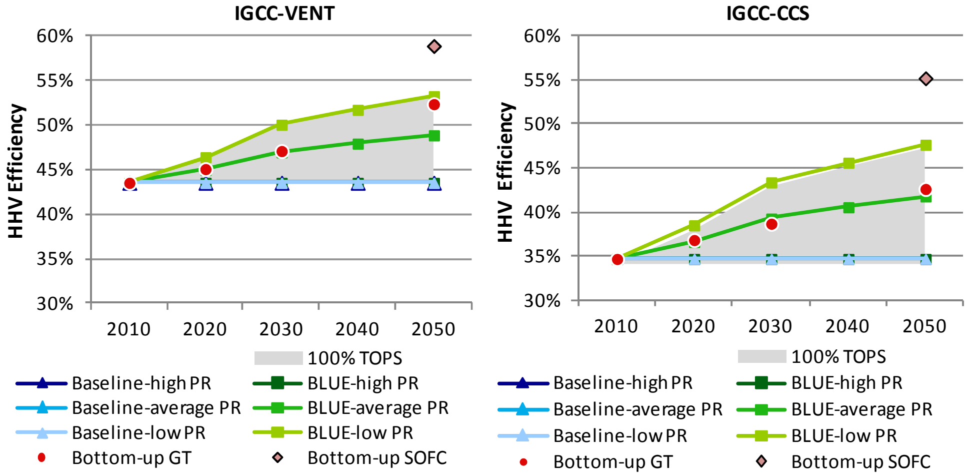

Figure 5.4: Efficiency development for IGCC-VENT (left) and IGCC-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based IGCC. The maximum and minimum values of the 100% TOPS case are depicted by a shadow for the experience curve approach.

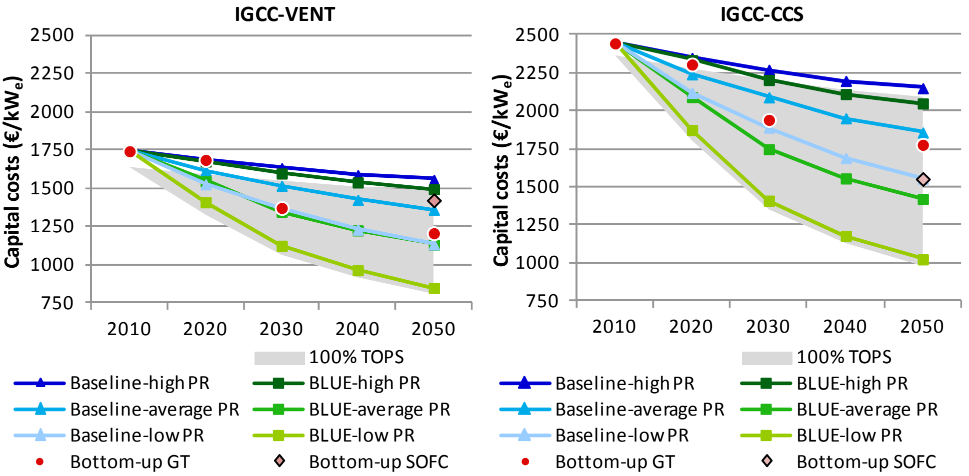

Figure 5.5: Capital cost development for coal-based IGCC-VENT (left) and IGCC-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based IGCC. The maximum and minimum values of the 100% TOPS case are depicted by a shadow for the experience curve approach.

If SOFC technology becomes commercially available, the costs for the power section will increase slightly due to the expensive fuel cells. However, the total specific costs compared with the SOTA configuration decrease to about 1.4 and 1.6 k€/kW$_e$ for IGCC-VENT and IGCC-CCS, respectively. Overall, the differences in projected capital cost reduction are small between the bottom-up and the experience curve analysis.

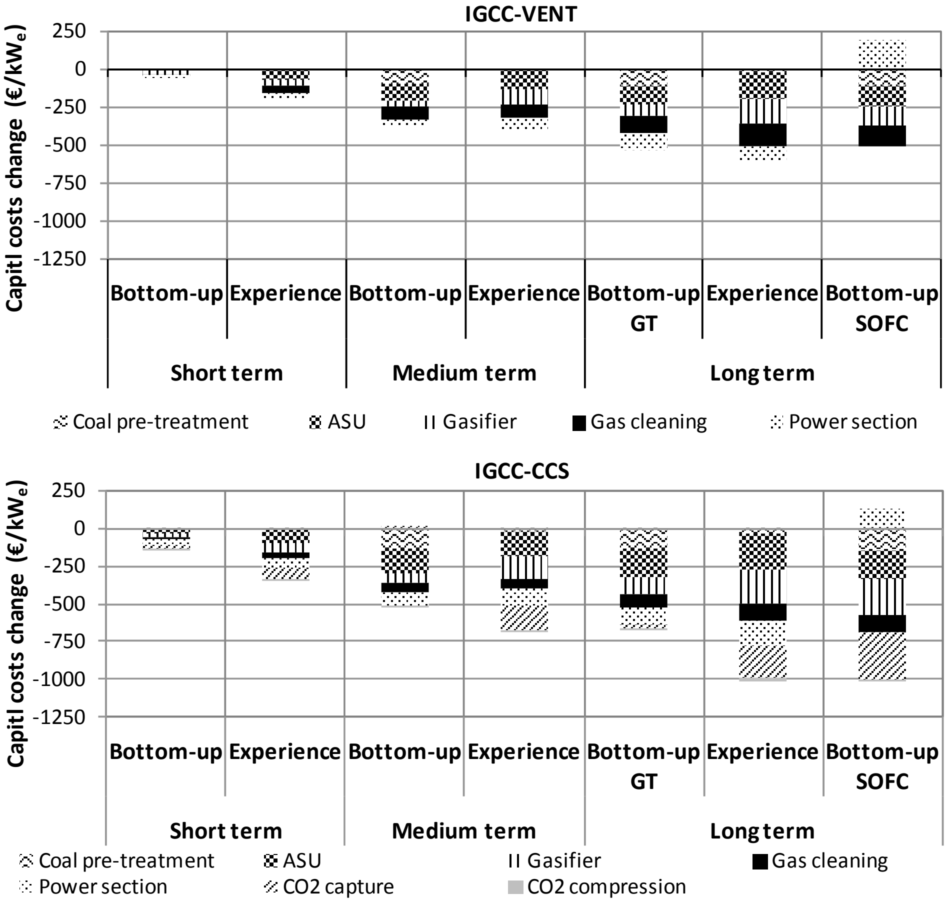

Figure 5.6: Contribution of different components to the capital cost decrease predicted by the component-based experience curve with the BLUE-average PR scenario and the bottom-up analysis for IGCC-VENT (upper) and IGCC-CCS (lower).

5.4.3 Trends in the COE and CO_2 mitigation costs for IGCC

Figure 5.7 shows the development of COE for the experience curve approach, as well as for the bottom-up analysis, by using the projected efficiency, capital costs and O&M costs. The bottom-up results correspond mostly to the projections of the BLUE-average PR. Although the efficiencies for the SOFC cases are considerably higher than the projections of the experience curve, the COE are within the range projected by the experience curve analysis due to the short life time and high replacement costs of fuel cells. If fuel cells become commercially available in 2050, the COE for IGCC-CCS could decrease to 50 €/MWh, while if IGCC remain based on advanced gas turbines, the COE will be 56 €/MWh. For IGCC-VENT, SOFCs will not be cost effective with a COE of 46 €/MWh compared to 40 €/MWh with advanced gas turbines.

The PR for the COE can be estimated at 0.96 (0.93-0.99) and 0.94 (0.91-0.98) for coal-based IGCC-VENT and IGCC-CCS, respectively, by using the COE development and the estimated future installed capacities over time of the BLUE scenario. The PR outside the brackets is based on projections of the BLUE-average PR scenario, which was the scenario matching best with the bottom-up projections, while the range inside the brackets is the minimum and maximum PR distracted from the data.

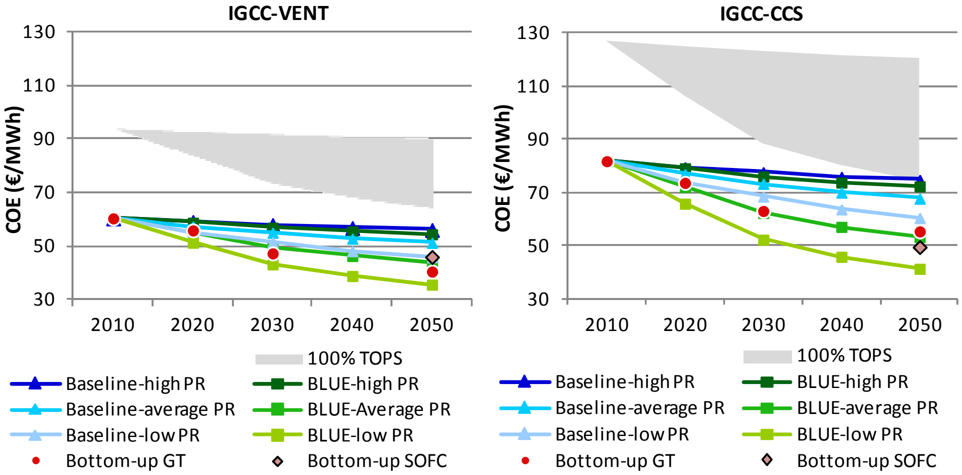

Figure 5.7: COE development for coal-based IGCC-VENT (left) and IGCC-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based IGCC. The maximum and minimum values of the 100% TOPS case with a TOPS price of 6.3 €/GJ are depicted by a shadow for the experience curve approach.

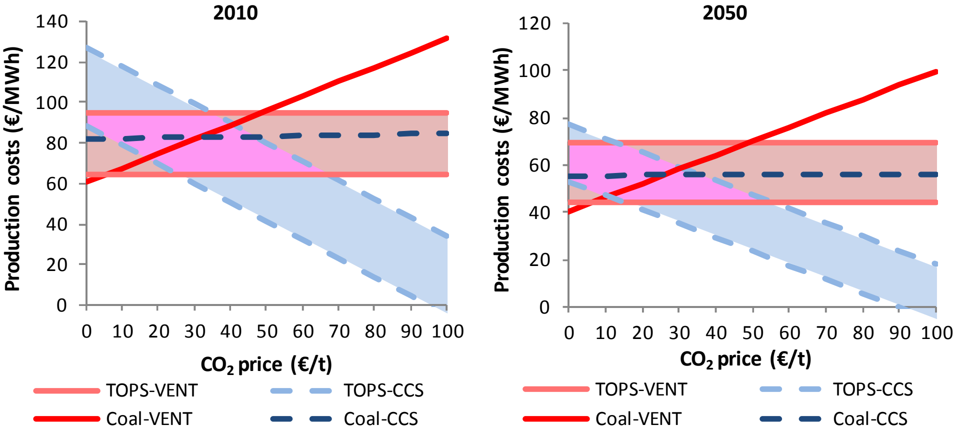

Figure 5.8: Effect of a CO_2 price on the COE for the current situation (left) and the situation in 2050 (right). The shaded area in the graph is the range due to low (3 €/GJ) and high TOPS prices (6.3 €/GJ).

In the case of TOPS-based gasification, the results indicate a higher COE for all time periods compared to coal-based gasification. The main reason for this is the higher assumed feedstock price of 6.3 or 3.0 €/GJ compared to 2.25 €/GJ for coal. For instance, the production costs are currently 92 €/MWh without CCS and 123 €/MWh with CCS . In

2050, they could decrease to respectively 69 €/MWh and 75 €/MWh with a constant TOPS price of 6.3 €/GJ and to 49 €/MWh and 53 €/MWh with a TOPS price of 3 €/GJ. Adding a CO_2 price would change the production costs, see Figure 5.8. With a TOPS price of 6.3 €/GJ, a CO_2 tax of 34 €/t CO_2 is currently required to achieve comparable COE as coal-based IGCC-CCS. If the TOPS price drop to 3.0 €/GJ in 2050, a CO_2 tax of 6 and 16 €/t CO_2 is required to make TOPS-based IGCC-VENT or TOPS-based IGCC-CCS the optimal option, respectively.

5.4.3.1 CO_2 emissions and CO_2 mitigation costs

Specific CO_2 emissions are currently 712 kg CO_2/MWh for coal-based IGCC-VENT and could be 29 kg CO_2/MWh for IGCC-CCS. These figures are projected to become 582-712 kg/MWh and 21-29 kg CO_2/MWh, respectively, in 2050 using the experience curve approach. With the bottom-up approach, a higher reduction in CO_2 emissions is predicted for IGCC-CCS, to 3-5 kg/MWh mainly due to the higher CO_2 capture ratio.

In Figure 5.9, the CO_2 mitigation costs are given with respect to IGCC-VENT, PC-VENT and NGCC-VENT. Calculated CO_2 mitigation costs strongly depend on the chosen reference system. With IGCC-VENT as reference, the CO_2 capture costs for coal-based IGCC-CCS are projected to decrease over time from the current 32 €/t $CO_{2,}$ to 27 €/t CO_2 in the short term, 25 €/t CO_2 in the medium term and 16 €/t CO_2 in the long term if fuel cells are commercially available[27]. If fuel cells do not become commercially available, the CO_2 capture costs could be 26 €/t CO_2 in the long term.

With PC-VENT as reference, the current CO_2 mitigation costs are considerably higher namely 44 €/t CO_2 and they are project to decrease to 7-16 €/t CO_2 in 2050. For TOPS-based IGCC, the current CO_2 mitigation costs are slightly higher, namely 43 €/t CO_2 with CCS and 55 €/t CO_2 without CCS. These could drop, with a TOPS price of 3 €/GJ, to 1-23 and -7-22 €/t CO_2 in 2050, respectively. The future CO_2 mitigation costs are lower than the CO_2 mitigation costs calculated with IGCC-VENT as reference, due to the smaller improvement projected for PC-VENT by Van den Broek et al.[168]

With NGCC-VENT as reference, the CO_2 mitigation costs are currently 92 €/t CO_2 and 4-23 €/t CO_2 in 2050. Interestingly, the current CO_2 mitigation costs are lower for TOPS-based (55 €/t CO_2) than for coal-based IGCC-CCS (92 €/t CO_2), even with a high TOPS price of 6.3 €/GJ.

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[27] The CO_2 mitigation costs in the long term are based on IGCC-VENT-GT as a reference, because this configuration has lower production costs than IGCC-VENT-SOFC. If the latter was taken as a reference, CO_2 mitigation costs would be 7 €/t CO_2 instead of 16 €/t $CO_{2.}$

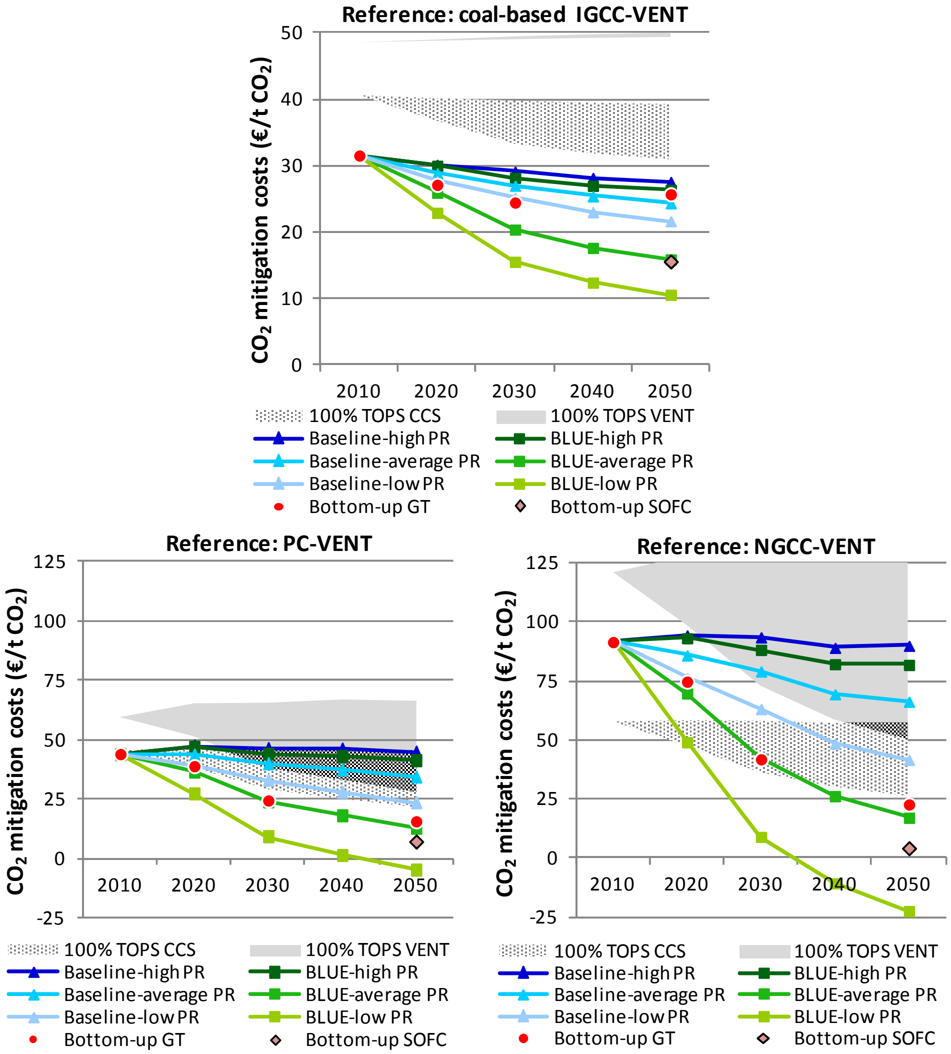

Figure 5.9: CO_2 mitigation cost development for coal and TOPS-based IGCC (with a TOPS price of 6.3 €/GJ) compared to a coal-based IGCC-VENT (upper), a PC without CO_2 capture (left) and a NGCC without CO_2 capture (right).

5.4.4 FT-liquids production

The projected development of the costs of FT-liquids and CO_2 mitigation costs are presented in Figure 5.10 and Figure 5.11, respectively. Results indicate similar trends for the BLUE and Baseline scenario for a given PR scenario. This can be explained by the fact that the total installed capacity of FT-liquids is comparable in the two scenarios, namely 435 and 439 $GW_{th\ FT}$ in 2050. However, in the BLUE scenario the majority (75%) of the FT facilities are equipped with CCS and half of the capacity is based on biomass, while in the Baseline scenario FT facilities are mainly based on fossil fuel and no FT-CCS capacity is installed. Consequently, policies to stimulate CCS do not influence the cost and efficiency development of FT-synthesis facilities.

The projections for production costs of FT-liquids in the bottom-up assessments are comparable to the projection costs following an average PR scenario. The production costs for FT-VENT are projected to decrease with 31% in the period 2010-2050, from 13.5 to 9.3 €/GJ. This means that the break-even oil price drops from the current 77 €/bbl to 51 €/bbl in 2050. If CCS is applied, the current production costs increase to 14.6 €/GJ now, and to 10.0 €/GJ in 2050. These costs are competitive with an oil price of 83 and 55 €/bbl, respectively. The small difference in production costs combined with the large difference in CO_2 emissions between FT-VENT (177 kg $CO_2/GJ_{th\ FT}$) and FT-CCS (77 kg $CO_2/GJ_{th\ FT}$) results in CO_2 mitigation costs of 11 €/t CO_2 in the current situation and 7 €/t CO_2 in 2050 (see Figure 5.11).

With the production costs development and the estimated installed capacities over time, the PR for the production costs of FT-VENT and FT-CCS can be estimated at 0.90 (0.82-0.97) and 0.91 (0.83-0.97), respectively. The figure outside the brackets refers to the PR based on projections of the average PR scenario, which was the scenario matching best with the bottom-up projections, while the range inside the brackets is the minimum and maximum PR distracted from the data.

Similar to IGCC, the production costs are considerably higher for TOPS-based FT-liquids compared to coal-based FT-liquids, while CO_2 emissions are significantly lower, namely 0 kg $CO_2/GJ_{th\ FT}$ for FT-VENT and -123 kg CO_2/GJ for FT-CSS. Currently, production costs are 22.7 €/GJ (= 133 €/bbl) for FT-VENT and 24.2 €/GJ (= 142 €/bbl) for FT-CCS with a TOPS price of 6.3 €/GJ. These are projected to decrease with a constant feedstock price to 18.1 €/GJ (= 105 €/bbl) and 19.0 €/GJ (= 110 €/bbl) in 2050, respectively, due to technological learning. This results in CO_2 mitigation costs of 52 €/t CO_2 if no CCS is applied and about 34 €/t CO_2 if CCS is applied compared to a coal-based FT-VENT as reference (see Figure 5.11). However, if TOPS prices decrease to 3 €/GJ, production costs will decrease to 11.2 (= 62 €/bbl) and 12.1 €/GJ (= 68 €/bbl) in 2050 for FT-VENT and FT-CCS, respectively. At the same time, the CO_2 mitigation costs will decrease to 11 and 10 €/t CO_2 with coal-based IGCC-VENT as reference.

If the CO_2 mitigation costs are calculated by taking oil-derived liquids as reference, the CO_2 mitigation cost are currently 56 and 123 €/t CO_2 for TOPS-based liquids with and without CCS, respectively. These are projected to decrease to 17-103 and 12-48 €/t CO_2 in 2050 with a TOPS price of 6.3 €/GJ. In 2030, the production costs for FT-VENT, with a TOPS price of 3 €/GJ, are projected to be lower than for oil-derived liquids with an oil price of 73 €/bbl, leading to negative CO_2 mitigation costs. For FT-CCS, this is the case in 2050.

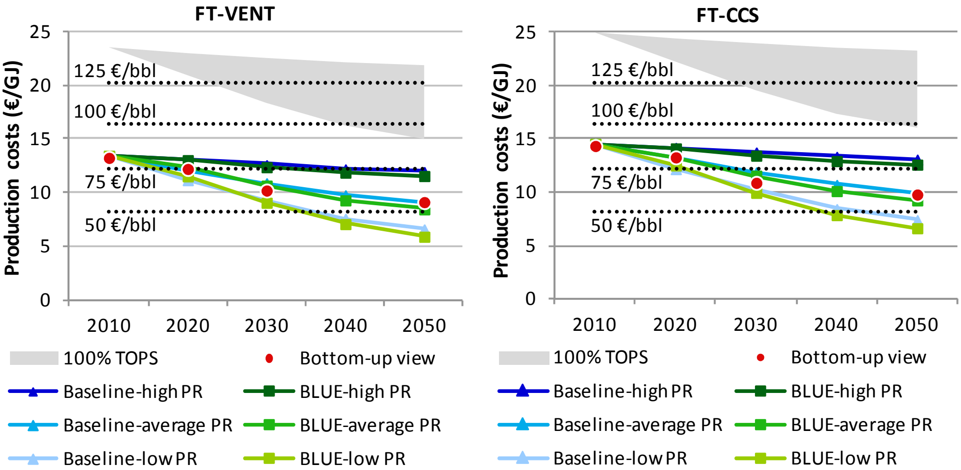

Figure 5.10: COE development for coal-based FT-VENT (left) and FT-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based FT. The maximum and minimum values of the 100% TOPS case with a TOPS price of 6.3 €/GJ are depicted by a shadow for the experience curve approach.

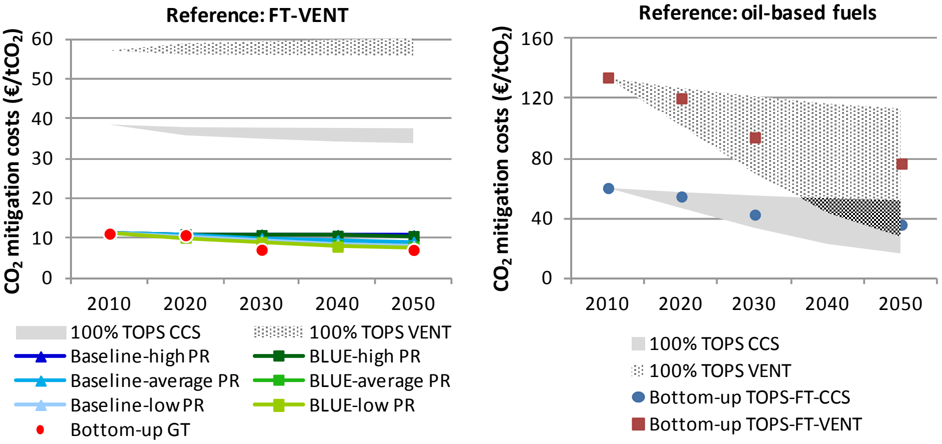

Figure 5.11: CO_2 mitigation costs of coal-based and TOPS-based FT-liquids with a TOPS price of 6.3 €/GJ calculated with a coal-based FT-VENT as reference (left) and oil-derived liquids as reference using an oil price of 73 €/bbl (right).

Figure 5.12 shows the influence of CO_2 price on the production costs of oil-based liquids, TOPS-based and coal-based FT-liquids for the current situation as well as for 2050. Currently, TOPS-based FT-liquids are economically the best alternative with a CO_2 price exceeding 48 €/t CO_2 and an oil price above 87 €/bbl for a feedstock price of 6.3 €/GJ. If TOPS can be obtained at 3 €/GJ in 2050, it becomes attractive to use TOPS instead of coal and to equip the FT production facility with CCS, with a CO_2 price of about 11 €/t CO_2. This could become economically the optimal alternative with an oil price above 58 €/bbl in 2050. If TOPS prices remain at 6.3 €/GJ in 2050, the CO_2 price has to increase to 48 €/t CO_2

and the oil price to 57 €/bbl, before TOPS-based FT-liquids (with CCS) could be economically the best alternative.

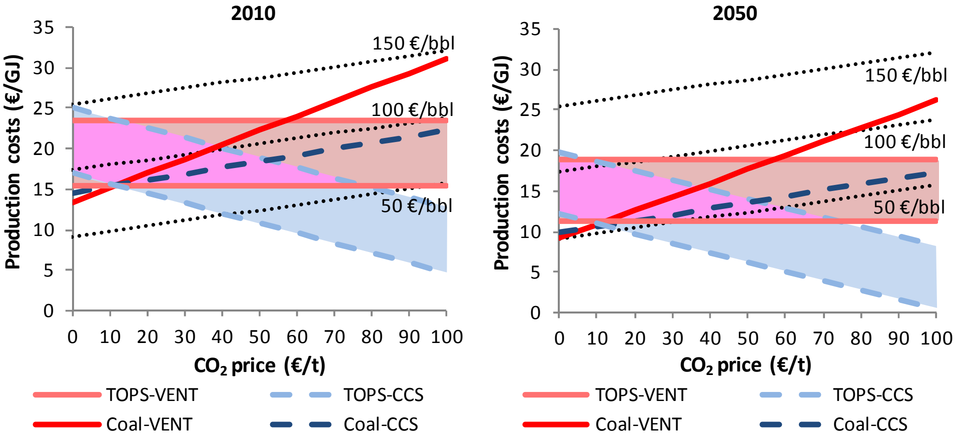

Figure 5.12: Effect of a CO_2 price on the production costs of FT-liquids for the current situation (left) and the situation in 2050 (right). The shaded area in the graph is the range due to low (3 €/GJ) and high TOPS prices (6.3 €/GJ).

5.5 Discussion

In this study, it is assumed that the PR values found for technological learning in an early stage of technology development also hold when further deployment is taking place. However, these values may change when mature markets for the implementation of the technology have been achieved.

The learning potentials indicated in this study are comparable to other cost and performance projections made for IG-PG. For instance, this study projects efficiencies of 52% for IGCC-VENT and 43% for IGCC-CCS in the long term without fuel cells, whereas other sources project efficiencies of 46-49% for IGCC-VENT and 40-44% for IGCC-CCS.[58,59,168] Accompanying capital costs are estimated at 1.2 and 1.8 k€/kW$_e$ for IGCC-VENT and IGCC-CCS, respectively, in this study, whereas other studies estimate them slightly lower, at 0.9-1.0 and 1.2-1.4 k€/kW$_e$.[58,59,168]

Also for FT-liquids similar learning potentials were found. Eerhart projects a production cost decrease of 13% from SOTA to 2030, with a constant biomass price. Here, TOPS-based FT-liquids are projected to decrease with 14% until 2030 with a constant TOPS price of 6.3 €/GJ. For coal-based FT-liquids, Eerhart projects a production cost decrease of 20% whereas this study estimates a 23% cost reduction till 2030.[172]

The learning potentials are influenced by the current installed capacities of the individual components, especially for components with a low installed capacity like the ASU and CO_2 compression. These are, however, difficult to validate. It is also uncertain how reliable our assumption is concerning the installed capacity of syngas turbines, which assume that 50% of the installed capacity of natural gas turbines is added to the installed capacity of syngas turbines. However, analyzing the consequences of this assumption showed that it influenced the results only to a limited extent, namely the long term COE would increase with maximum 5% if only the installed capacity of syngas turbines is incorporated in the experience curve analysis.

To incorporate the observation that learning does often not start from the first doublings onwards, a pre-learning phase of 7 GW_e was assumed in this study. Without a pre-learning phase, the capital costs are higher in the short term but lower in the long term. The reason for this is that after the pre-learning phase is crossed, capital costs are decreasing rapidly. The efficiency, at the other hand, is decreasing with an increasing pre-learning phase for all time periods. Hence, with a longer pre-learning the projected COE would be higher in the short term, whereas, in the long term the COE are projected to be lower with an increasing pre-learning phase until about 50 GW_e. For an even longer pre-learning phase, the COE are projected to increase in the long term.

Over time, the bottom up analysis and the experience curve approach show similar results with respect to efficiency, capital costs, production costs and CO_2 mitigation costs development. The bottom-up analysis indicates that new technologies like syngas turbines, H_2 membranes, ITM and SOFCs have to become commercially available to reach the projected production costs. Further research and development is required to solve some critical problems identified with these technologies. For example, some actors question whether SOFCs will become commercially available on the necessary scale at competitive costs.[229,230] Additionally, the experience curve approach indicates that a strong growth is required in IGCC, FT-synthesis and CCS capacity. For instance, CCS capacity should increase to over 2,000 GW_e to avoid 10 t CO_2/yr in 2050. It will be a challenge to reach this, although other scenario studies project the possibility of even larger deployment rates of CCS, up to 37.5 Gt CO_2/yr in 2050.[34] Simultaneously, IGCC capacity should increase from the current 4 to 620 GW_e and FT-synthesis capacity from 28 $GW_{th\ FT}$ at present to 430 $GW_{th\ FT}$ in 2050. Nowadays, IG-PG facilities and CCS equipment are not (yet) planned on a large scale.

Although the overall results obtained are comparable for both approaches, some differences occur when zooming in on the calculated capital costs reductions. One of the most significant differences is the improvement estimated in the coal pre-treatment unit in the bottom-up analysis, mainly due to the introduction of the solid feed pump. This could not be observed in the experience curve approach, mainly due to the large installed capacity of coal pre-treatment in PC power plants and industry at present. However, it can be argued that the coal pre-treatment in gasification plants is different than in PC power plants, due to the higher pressure and smaller required particles of the former one. Therefore, expensive lock hoppers are used in dry-fed gasification plants which are not used in PC power plants. If only the capacity of coal-based gasification plants was used in the experience curve approach, the capital costs of the coal pre-treatment unit would be reduced with 19% instead of 9%. Nevertheless, the projected capital reduction of 67% in the pre-treatment unit due to the introduction of the solid feed pump is not reached. This could mean that the solid feed pump has to be considered as a radical innovation with its own experience curve or that the predicted cost reduction of the solid feed pump is too optimistic.

For some components, a somewhat higher capital reduction is predicted by the experience curve analysis than by the bottom-up study. This is, for instance, the case for the ASU and gasifier. The current installed capacity of the ASU is underestimated because the current capacity in the industry (with the exception of gasification plants) is neglected. A doubling in the currently installed capacity would result in a 10% increase in capital costs of the ASU and a 1% increase in total capital costs in 2050 with the BLUE-average-PR scenario. This could explain the observed difference.

For the gasifier, no improvement option was implemented in the bottom-up analysis, but there was one option identified, namely the high pressure gasifier.[204] This type of gasifier was not included in the process designs because the commercialization date was expected to be after 2050. However, if the high pressure gasifier would be implemented in process design of the long term, capital costs of the gasifier may be reduced with 50% and the efficiency would improve with 0.3$\%_{pt}$, leading to a 3-11% reduction in production costs.[204] This could more than explain the observed cost difference of the gasifier between the two approaches.

5.6 Conclusion

This study aimed to investigate the technological and economic prospects of IG-PG (for power and FT-liquid production) with and without CCS over time. Therefore, a component-based experience curve analysis was executed, and the results were compared with a bottom-up analysis conducted in previous work.

The COE from IGCC-VENT are currently calculated at 60 €/MWh, which is slightly higher than the market price (57 €/MWh). It is projected that the COE could reduce to 56 €/MWh in 2020, mainly because of a higher conversion efficiency. However, to reach this COE, an increase in installed IGCC capacity is required from the current 4 to about 20 GW_e. If CCS is applied, the current production cost is significantly higher (82 €/MWh). Although the production costs could reduce significantly over time, they are expected to remain above the current market price (of 57 €/MWh) until 2050. In 2050, costs are projected at 50 €/MWh if SOFCs are becoming commercially available and at 56 €/MWh if an IGCC remains based on (advanced) gas turbines. Hence, SOFCs are projected to be advantageous for IGCC-CCS, but if SOFCs do not become competitive, still a large costs reduction can be expected.

Comparing IGCC-VENT with IGCC-CCS, leads to CO_2 capture costs of 32 €/t CO_2 in the current situation and to 16 €/t CO_2 in 2050, if SOFCs become commercially available. If IGCC-CCS is replacing PC-VENT or NGCC-VENT, current CO_2 mitigation costs are 44 and 92 €/t CO_2, respectively. However, also these are projected to decrease over time to 7 and 4 €/t CO_2 in 2050. These costs are expected to be reached with an IGCC capacity exceeding 600 GW_e, a FT capacity of around 430 $GW_{th\ FT}$ and more than 2,000 GW_e equivalent of CCS capacity.

When TOPS is used as feedstock, the COE are currently 92 €/MWh without CCS and 123 €/MWh with CCS, and they could decrease to respectively 69 €/MWh and 75 €/MWh in 2050 with a constant TOPS price of 6.3 €/GJ. With a TOPS price of 3 €/GJ, COE could be 49 and 53 €/MWh in the long term. This means that, in the long term, TOPS-based IGCC-VENT can become economically attractive with a CO_2 tax of 6 €/t CO_2, and TOPS-based IGCC-CCS becomes the most competitive option with a CO_2 tax of 16 €/t CO_2.

Currently, production costs of coal-based FT-liquids without CCS are 13.5 $€_{2008}$/GJ, which is competitive to oil based fuels at an oil price of 77 $€_{2008}$/bbl. If CCS is applied the costs will increase to 14.6 €/GJ, which is competitive at a CO_2 price of 11 $€_{2008}$/t CO_2 (excluding the costs for transportation and storage). Due to the low costs for capturing CO_2, it is expected that FT-synthesis plants could become one of the first applications for CCS. Over time, learning could reduce production costs about 30% to 9.3 $€_{2008}$/GJ for FT-liquids without CCS and to 10.0 $€_{2008}$/GJ for FT-liquids with CCS in 2050, reducing the CO_2 capture costs to 7 €/t CO_2. To reach the projected cost reductions, FT-synthesis capacity should grow, however, from the current 28 $GW_{th\ FT}$ to around 430 $GW_{th\ FT}$ in 2050. It is likely that this will happen without any additional CCS policy.

With CCS, the life cycle CO_2 emissions of coal-based FT-liquids (77 kg/$GJ_{th\ FT}$) and oil-derived liquids (80 kg/$GJ_{th\ FT}$) are comparable. Reduction in CO_2 emissions can be realized by using TOPS as feedstock. Then CO_2 emissions are zero without CCS and they become even negative with CCS, namely -123 kg CO_2/$GJ_{th\ FT}$. However, with the current TOPS price of 6.3 €/GJ, FT-liquids are at present more expensive than coal-based FT-liquids with production costs of 22.7 €/GJ without CCS and 24.2 €/GJ with CCS. These could drop to respectively 18.1 €/GJ and 19.0 €/GJ in 2050. If TOPS prices reduce to 3 €/GJ, production costs could decrease to 11.2 and 12.1 €/GJ in 2050. This means that they are economically the best alternative for liquid fuels with an oil price above 58 €/bbl and a CO_2 price exceeding 11 €/t CO_2.

To summarize, our results indicated that gasification technology can be a promising technology in satisfying future liquid fuel demand and electricity demand. However, to stimulate IGCC-VENT development initial (financial) support is needed up to an installed capacity of about 20 GW_e while FT-liquids will already be competitive with an oil price of 77 $/bbl. A modest CO_2 price of about 11€/t CO_2 would stimulate the development of FT-CCS, which promote learning in CCS components wherefrom other (gasification) facilities with CCS would benefit.

With respect to the applied methodology, it can be concluded that the bottom-up analysis and the component-based experience curve approach are giving comparable trends regarding efficiency improvement, capital costs and production costs reductions for a scenario with a strong growth in IGCC, FT and CCS capacity. The main advantage of combining the two approaches is that it becomes explicitly clear where the main costs reductions can be realized, namely in the syngas turbines and ASU, combined with an indication about the time frame and capacity development required to reach the projected improvements. When zooming in on specific costs reductions, some differences could be observed, like the costs reduction of coal pre-treatment and ASU, but these differences are limited and could overall be explained.

Overall, combinations of bottom-up analysis and (component-based) experience curve approach may result in better validated results and projections on the future performance of energy technologies.

5.7 Supporting information

Table 5.9: Cost and performance projections of coal-based IGCC-VENT and IGCC-CCS for the period 2010-2050. (1)

Technology	Variable	Unit	Scenario	2010	2020	2030	2040	2050	%-change in 2010-2050
IGCC-VENT	**Capital costs**	€/kW$_e$	**Base**	1,742	1,615 (1,526 - 1,686)	1,517 (1,370 - 1,638)	1,425 (1,231 - 1,592)	1,360 (1,134 - 1,561)	-22% (-10% - -35%)
			BLUE	1,742	1,546 (1,409 - 1,675)	1,348 (1,125 - 1,597)	1,226 (966 - 1,538)	1,132 (850 - 1,494)	-35% (-14% - -51%)
			Bottom-up	1,742	1,686	1,372		1,207 - 1,422	-18% - -31%
	Overall efficiency	%$_{HHV}$	**Base**	44%	44% (44% - 44%)	44% (44% - 44%)	44% (44% - 44%)	44% (44% - 44%)	0% (0%-0%)
			BLUE	44%	45% (44% - 46%)	47% (44% - 50%)	48% (44% - 52%)	49% (44% - 53%)	12% (0% - 22%)
			Bottom-up	44%	45%	47%		52% - 59%	20% - 35%
	CGE	%$_{HHV}$	**Bottom-up**	82%	82%	82%		82%	0%
	COE	€/MWh	**Base**	60	55 (51 - 59)	55 (51 - 58)	53 (48 - 57)	51 (46 - 56)	-15% (-7% - -24%)
			BLUE	60	57 (55 - 59)	50 (43 - 57)	46 (39 - 55)	44 (36 - 54)	-28% (-10% - -41%)
			Bottom-up	60	56	47		46 - 40	-33% - -24%
	CO_2 emissions	kg/MWh	**Base**	712	712 (712 - 712)	712 (712 - 712)	712 (712 - 712)	712 (712 - 712)	0% (0%-0%)
			BLUE	712	689 (712 - 668)	660 (712 - 619)	646 (712 - 599)	634 (712 - 582)	-11% (0% - -18%)
			Bottom-up	712	687	658		592 - 526	-17% - -26%

IGCC-CCS	**Capital costs**	€/kW$_e$	**Base**	2,445	2,241 (2,119 - 2,349)	2,090 (1,891 - 2,271)	1,952 (1,690 - 2,198)	1,858 (1,555 - 2,150)	-24% (-12% - -36%)
			BLUE	2,445	2,092 (1,875 - 2,334)	1,748 (1,410 - 2,204)	1,557 (1,179 - 2,111)	1,422 (1,024 - 2,050)	-42% (-16% - -58%)
			Bottom-up	2,445	2,306	1,941		1,553 - 1,778	-27% - -36%
	Efficiency	%$_{HHV}$	**Base**	35%	35% (35% - 35%)	35% (35% - 35%)	35% (35% - 35%)	35% (35% - 35%)	0% (0% - 0%)
			BLUE	35%	37% (35% - 39%)	39% (35% - 43%)	41% (35% - 46%)	42% (35% - 48%)	20% (37% - 0%)
			Bottom-up	35%	37%	39%		43% - 55%	59% - 23%
	COE	€/MWh	**Base**	82	77 (74 - 80)	73 (69 - 78)	70 (64 - 76)	68 (61 - 75)	-17% (-9% - -26%)
			BLUE	82	72 (66 - 79)	63 (52 - 76)	57 (46 - 74)	53 (42 - 72)	-35% (-12% - -49%)
			Bottom-up	82	74	63		50 - 56	-32% - -39%
	CO_2 emissions	kg/MWh	**Base**	29	29 (29 - 29)	29 (29 - 29)	29 (29 - 29)	29 (29 - 29)	0% (0% - 0%)
			BLUE	29	27 (26 - 29)	25 (23 - 29)	25 (22 - 29)	24 (21 - 29)	-17% (-27% - 0%)
			Bottom-up	29	27	5		5 - 3	-90% - -83%

(1) In each row for the Baseline and BLUE scenario, the first result is based on the average PR and the values below are based on the high and low PR.

Table 5.10: Cost projections of coal-based FT-VENT and FT-CCS for the period 2010-2050. (1)

Technology	Variable	Unit	Scenario	2010	2020	2030	2040	2050	%-change in 2010-2050
FT-VENT	**Capital costs**	M€	**Base**	929	861 822 - 902	792 722 - 871	731 637 - 842	692 584 - 824	-25% -11% - -37%
			BLUE	929	864 827 - 899	767 691 - 851	692 592 - 814	647 534 - 791	-30% -15% - -43%
			Bottom-up	929	929	846		847	-9%
	Overall efficiency	$\%_{HHV}$	**Bottom-up**	61%	61%	63%		65%	7%
	Fuel efficiency	$\%_{HHV}$	**Bottom-up**	49%	49%	49%		50%	4%
	Electric efficiency	$\%_{HHV}$	**Bottom-up**	12%	13%	13%		14%	20%
	CGE	$\%_{HHV}$	**Bottom-up**	82%	82%	82%		82%	0%
	Cost of synfuel	€/GJ_{fuel}	**Base**	13.5	12.1 (11.2 - 13)	10.9 (9.3 - 12.7)	9.8 (7.6 - 12)	9.1 (6.6 - 12)	-32% (-11% - -51%)
			BLUE	13.5	12.3 (11.5 - 13)	10.7 (9.1 - 12.4)	9.3 (7.1 - 12)	8.5 (5.9 - 12)	-37% (-14%- -56%)
			Bottom-up	13.3	12.2	10.2		9.1	-31%
	CO_2 emissions	kg/GJ	**Base**	177	177 (177 - 177)	177 (177 - 177)	177 (177 - 177)	177 (177 - 177)	0% (0% - 0%)
			BLUE	177	177 (177 - 177)	177 (177 - 177)	177 (177 - 177)	177 (177 - 177)	0% (0% - 0%)
			Bottom-up	177	177	177		177	0%
	Capital recovery	€/GJ_{fuel}	**Base**	9.5	8.8 (8.4 - 9.2)	8.1 (7.4 - 8.9)	7.4 (6.5 - 8.6)	7.1 (5.9 - 8.4)	-25% (-11% - -37%)
			BLUE	9.5	8.8 (8.4 - 9.2)	7.8 (7 - 8.7)	7.1 (6 - 8.3)	6.6 (5.4 - 8.1)	-30% (-15% - -43%)
			Bottom-up	9.5	8.9	7.5		6.9	-27%

FT-CCS	**Capital costs**	M€	**Base**	957	889 (849 - 929)	711 (644 - 794)	641 (552 - 760)	598 (498 - 738)	-37% (-23% - -48%)
			BLUE	957	886 (851 - 926)	741 (676 - 814)	683 (597 - 787)	647 (548 - 769)	-32% (-20% - -43%)
			Bottom-up	957	957	859		860	-10%
	Overall efficiency	$\%_{HHV}$	**Bottom-up**	58%	59%	61%		63%	8%
	Fuel efficiency	$\%_{HHV}$	**Bottom-up**	49%	49%	49%		51%	4%
	Electric efficiency	$\%_{HHV}$	**Bottom-up**	10%	11%	12%		13%	28%
	Cost of synfuel	€/GJ_{fuel}	**Base**	14.6	13.2 (12.2 - 14.2)	11.6 (10 - 13.5)	10.1 (8 - 12.9)	9.3 (6.6 - 12.6)	-36% (-13%- -54%)
			BLUE	14.6	13.3 (12.5 - 14.1)	11.9 (10.2 - 13.8)	10.7 (8.4 - 13.3)	10 (7.4 - 13.1)	-31% (-49%- -10%)
			Bottom-up	14.4	13.3	10.9		9.8	-32%
	CO_2 emissions	kg/GJ	**Base**	77	77 (77 - 77)	77 (77 - 77)	77 (77 - 77)	77 (77 - 77)	0% (0% - 0%)
			BLUE	77	77 (77 - 77)	77 (77 - 77)	77 (77 - 77)	77 (77 - 77)	0% (0% - 0%)
			Bottom-up	77	77	77		76	-2%
	Capital recovery	€/GJ_{fuel}	**Base**	9.8	9.1 (8.7 - 9)	8 (7.2 - 9)	7.2 (6.2 - 9)	6.7 (5.6 - 8)	-31% (-15% - -43%)
			BLUE	9.8	9 (8.7 - 9)	8 (7.2 - 9)	7.2 (6.2 - 9)	6.7 (5.6 - 8)	-31% (-15% - -43%)
			Bottom-up	9.8	9.2	7.6		7	-28%

(1) The results for the Baseline and BLUE scenario are based on the average PR, the results for the high and low PR are given in brackets.

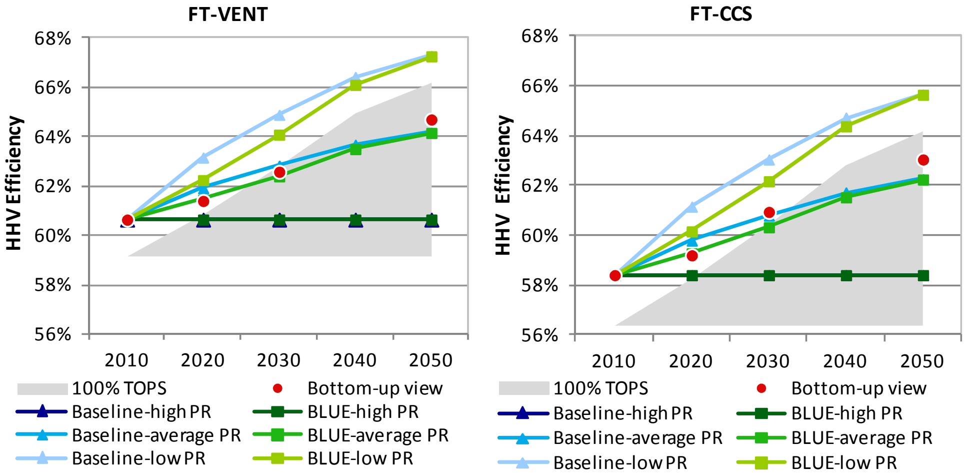

Figure 5.13: Efficiency development for coal-based FT-VENT (left) and FT-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based FT. The maximum and minimum values of the 100% TOPS case are inserted with a shadow for the experience curve approach.

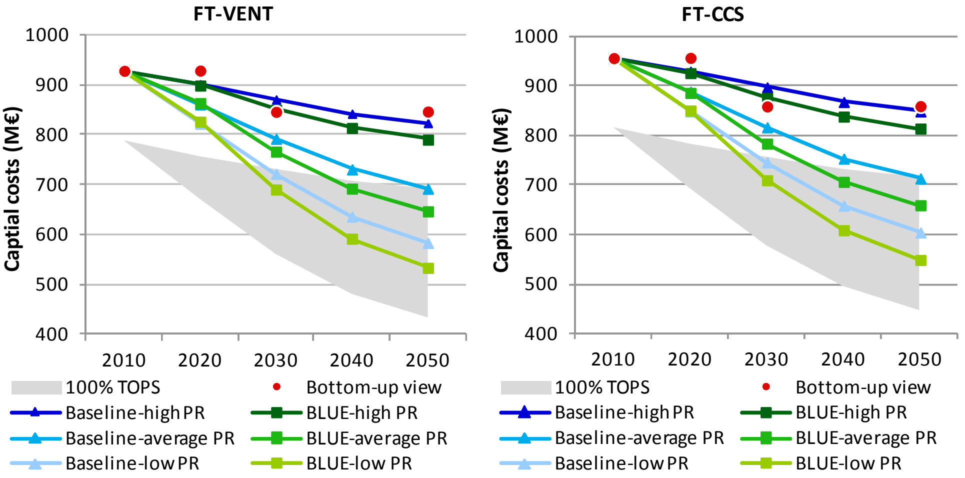

Figure 5.14: Capital cost development for coal-based FT-VENT (left) and FT-CCS (right). The lines show the results of the experience curve approach while the dots and diamond show the results of the bottom-up analysis for coal-based FT. The maximum and minimum values of the 100% TOPS case are inserted with a shadow for the experience curve approach.

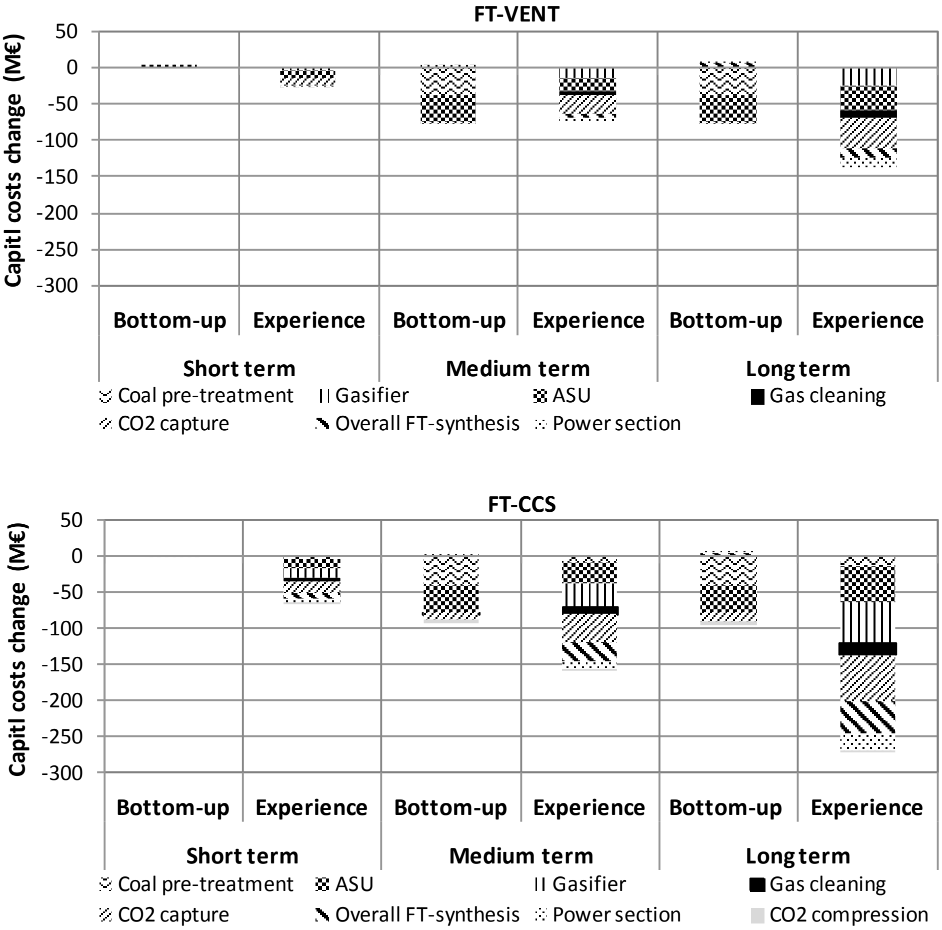

Figure 5.15: Contribution of different components to the capital cost decrease predicted by the component based experience curve with the BLUE-high PR scenario and the bottom-up analysis for FT-VENT (upper) and FT-CCS (lower).

6 Techno-economic assessment of CO_2 capture at steam methane reforming facilities using commercially available technology

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Abstract

This study aimed to identify the optimal techno-economic configuration of CO_2 capture at steam methane reforming facilities using currently available technologies by means of process simulations. Results indicate that the optimal system is CO_2 capture with ADIP-X located between the water-gas shift and pressure swing adsorption units.

Process simulations of this system configuration showed a CO_2 emission reduction of 60% at 41 €/t CO_2 avoidance. This is at the lower end of the range reported in open literature for CO_2 capture at refineries (26-82 €/t CO_2) and below the avoidance costs for CO_2 capture at natural gas-fired power plants (44-93 €/t CO_2). CO_2 avoidance costs are dominated by the natural gas consumption, responsible for up to 66% of total costs. Using imported steam and electricity can reduce CO_2 avoidance costs by 45%.

Addition of small amounts of piperazine to aqueous MDEA solutions results in up to 70% smaller absorbers or 10% lower reboiler heat duty. Optimising the whole capture process instead of individual units resulted in lower piperazine concentrations than the common industrial practice (3 wt% vs. 5 wt%). Finally, keeping the solvent rate constant when operating the capture unit below its design load resulted in a lower specific energy for CO_2 capture than when the solvent rate was downscaled with the syngas flow.

6.1 Introduction

To keep global warming under 2°C, atmospheric CO_2 concentration needs to stabilise most probably at a level below 450 ppm(v) CO_2-equivalent. A number of studies indicate that carbon dioxide capture and storage (CCS) can be an important option to reduce CO_2 emissions and to stabilise atmospheric CO_2 concentrations at acceptable levels.[11,18,29,34,215] Research on CO_2 capture has mainly focussed on the power sector. The industrial sector has received less attention, despite its significant emissions, about 6.7 Gt CO_2/yr globally in 2005.[18,231] A recent study by the IEA indicates a significant potential role of CCS in the industrial sector, capturing 2 Gt CO_2 (one-fifth of the total amount of CO_2 captured) in 2050.[4] A breakdown of the current industrial CO_2 emissions is presented in Figure 6.1.

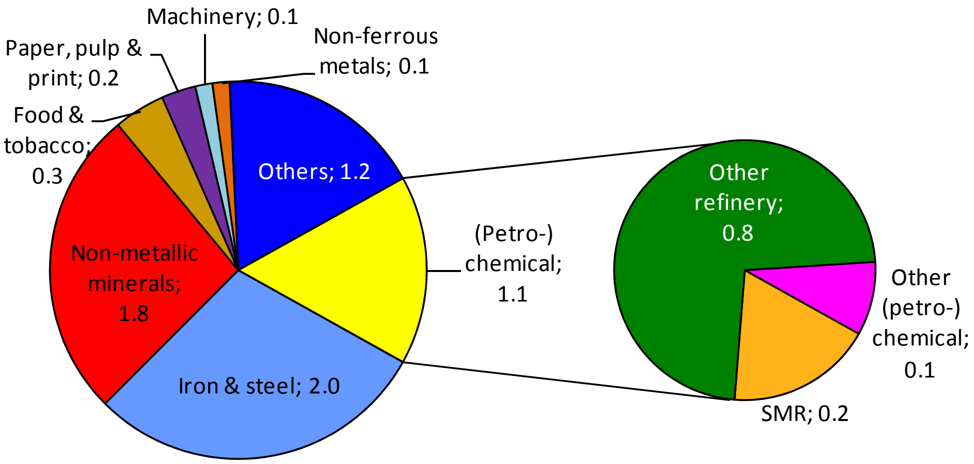

Figure 6.1 Direct global CO_2 emissions of the industrial sector in Gt/yr CO_2 in 2005.[18]

The (petro)chemical sector was responsible for about 16% of industrial CO_2 emissions in 2005.[18] A large share originated from steam methane reforming (SMR) facilities. Global annual H_2 production in 2005 is estimated at 65 Mt, of which 48% was from SMR facilities. SMR facilities emit on average 7 kg CO_2/kg H_2,[18] resulting in global CO_2 emissions of about 220 Mt CO_2 in 2005. Compared to total global CO_2 emissions, the share of SMR facilities is small (3%).[18] This share is, however, expected to increase if H_2 is (partly) used to decarbonise the transportation sector. The IEA estimates that, if H_2 fuel cell vehicles are successfully introduced, global H_2 demand for transportation could be as high as 275 Mt H_2/yr in 2050,[18] resulting in about 2 Gt CO_2/yr if produced from natural gas using SMR facilities. CCS can then play a key role in decreasing these emissions.

SMR facilities are attractive to capture CO_2 for several reasons:

- SMR facilities are large-scale stationary emitters of CO_2;
- a pressurised CO_2 rich gas stream is produced, making CO_2 capture potentially more efficient and less costly;
- CO_2 capture can be done using commercially available technology[34];

- SMR facilities are often located in close proximity to waste heat streams. If the CO_2 capture unit requires heat, waste heat integration may reduce the penalty and costs of CO_2 capture.

Several research groups have investigated CO_2 capture at SMR facilities. Simbeck indicates that CO_2 capture from the furnace of a SMR facility is possible using an amine-based solvent (MEA).[232] Both the IEA GHG and Damen et al., point out that capture of CO_2 after the water-gas shift (WGS) reactor may be achieved at low CO_2 avoidance cost.[71,233] Damen et al., also indicate that CO_2 capture at SMR facilities is technically feasible in the short term and that the technical and economic implications are rather limited.[71] The reported CO_2 avoidance costs at SMR facilities varies significantly in the literature, ranging between 26 and 82 €/t CO_2 captured, although the higher value includes not only capturing CO_2 from SMR facilities but also from other sources at refineries.[71,234] It is expected that CO_2 avoidance costs at SMR facilities will be on the lower side of the range because of the presence of high pressure, high concentration CO_2 streams. For comparison, the costs of post-combustion, amine-based capture at gas-fired power stations in OECD countries is estimated at 58 €/t CO_2 avoided, with estimates ranging between 44 and 93 €/t CO_2.[235]

Investment decisions are, however, not solely based on cost values. The impact of CO_2 capture on SMR facility performance is also important, as the H_2 purity, SMR availability, and SMR conversion efficiency must remain high. In the open literature, the relation between cost, configuration and performance has not been explored yet. The goal of this study is therefore threefold: (1) to identify and assess optimal techno-economic SMR-CO_2 capture configurations using commercially available technology; (2) to assess the economics of importing the required electricity and steam instead of on-site production; and (3) to evaluate the effect of part-load operation of the CO_2 capture unit on the techno-economic performance of the SMR installation. These findings will provide insights into the technical potential of CO_2 capture at existing SMR facilities, the effect of CO_2 capture on SMR facility performance, and the potential cost of reducing CO_2 emissions at these facilities.

The structure of this article is as follows: Section 2 gives technical background on SMR and CO_2 capture. Section 3 describes the methodology used. Section 4 shows technical and economical results, while Section 5 contains the conclusions and recommendations. Note that in this study, all units are in SI-units, heating values are in higher heating value (HHV) and monetary values are in €$_{2008}$, unless otherwise stated.

6.2 Background

Refineries are used to convert primary fossil energy carries (natural gas, coal and oil) into secondary energy carriers, products and basic building blocks used by the chemical industry for further processing (e.g., alkenes, benzenes and carbon monoxide). The main processes are distillation, hydrogenation and cracking. H_2 is consumed during, among others, hydrogenation, hydrocracking and hydrodesulphurisation. Most of the required H_2 is produced via SMR or gasification. As large volumes of H_2 are required, the SMR facilities are generally large-scale, providing economies of scale when applying CCS.[18]

6.2.1 Steam methane reforming

In a SMR facility (see Figure 6.2), natural gas is converted into H_2. This is done by reacting the natural gas with steam in a reformer (630-1000°C and 5-40 bar) to mainly H_2, CO and CO_2, as described by reactions 6-1 and 6-2. To maximise H_2 production, more steam is added in a low temperature water-gas shift (WGS) reactor to react with CO according to reaction 6-3.

$$CH_4 + H_2O \leftrightarrows CO + 3\,H_2 \qquad \Delta H^\circ_{298K} = +\,206 \text{ kJ/mol} \qquad (6\text{-}1)$$
$$CH_4 + 2\,H_2O \leftrightarrows CO_2 + 4\,H_2 \qquad \Delta H^\circ_{298K} = +\,165 \text{ kJ/mol} \qquad (6\text{-}2)$$
$$CO + H_2O \leftrightarrows CO_2 + H_2 \qquad \Delta H^\circ_{298K} = -\ \ 41 \text{ kJ/mol} \qquad (6\text{-}3)$$

The gas stream, containing mainly H_2 and CO_2, is fed into a H_2-pressure swing adsorption (H_2-PSA) unit, where the gas is separated into a H_2-rich stream (containing >99.99 vol% pure H_2) and a low-pressure waste stream (containing CO_2, steam and traces of methane, CO and H_2). Natural gas is added to this stream, which is combusted to heat the reformer.

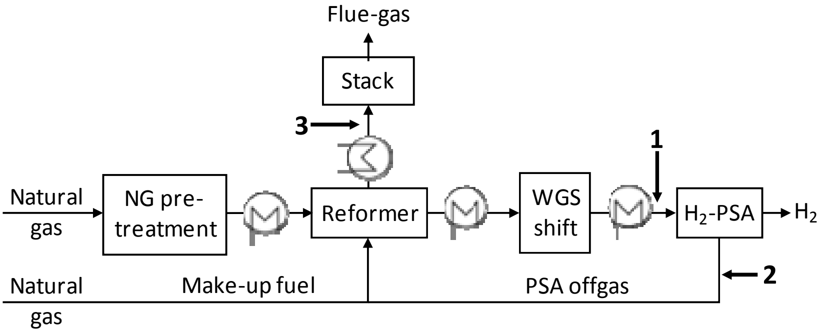

Figure 6.2 Schematic overview SMR unit. Numbers indicate possible CO_2 capture locations, see §6.3.1.[144]

6.2.2 Carbon dioxide capture and storage

Carbon dioxide capture and storage is the general term for processes that extract, purify, compress, transport and store CO_2. This article focuses on the first part of the CCS chain, consisting of separation, purification and compression of CO_2. There are four techniques to separate and purify CO_2, namely gas absorption, gas adsorption, membrane separation and cryogenic separation.[34,35] Gas adsorption has the disadvantage that the main impurities (CO and CH_4) have almost identical adsorption forces as CO_2, requiring large and expensive separation columns. Membrane separation is currently not commercially available. Cryogenic separation has a high energy penalty. Therefore, this study will focus on gas absorption, as this technique is commercially available and widely used in the industry. For more information on CO_2 capture techniques, see the IPCC special report on Carbon dioxide Capture and Storage[34] or the Advanced Carbon Dioxide Capture R&D Program: Technology Update.[35]

6.2.2.1 CO_2 capture solvents

Gas absorption systems to capture CO_2 can use physical or chemical solvents. Physical solvents are considered suitable for high CO_2 partial pressure gas streams as their loading capacity increases proportional to the partial pressure.[236] Chemical solvents are more suitable for low CO_2 partial pressures as, in this case, the loading capacity is much higher compared to the physical solvents, see Figure 6.3.[35,237] The higher the solvent loading capacity, the lower the solvent recirculation rate, generally resulting in lower capital and operational costs and reduced energy consumption. Several promising, commercially available solvents and their operating conditions are given in Table 6.1. The technical performance of these solvents was evaluated in this study.

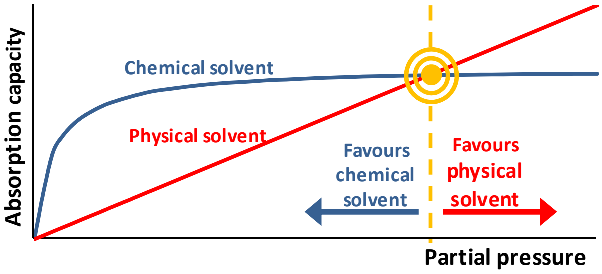

Figure 6.3 Impact CO_2 partial pressure on loading capacity for chemical and physical solvents.[35]

Table 6.1 Overview of commercially available absorption processes.[47,126,238,239,240,241,242]

Absorption process	**Solvent** (1)	**Process conditions** (2)
	Physical Solvent	
Rectisol	Methanol	-60 - -30°C >20 bar
Selexol	Dimethyl ethers of polyethylene glycol (DMPEG)	0 - 40°C 20 - 30 bar
	Chemical Solvent	
Organic (Amine-based)		
MEA	2.5 N 2-aminoethanol and chemical inhibitors in water	40 - 60°C, ambient-intermediate pressures
MDEA	2 N Bis(2-hydroxyethyl)methylamine in water	Wide range
Inorganic		
Benfield and Catacarb	Potassium carbonate and (Lurgi or Catacarb) catalysts in water	70 - 120°C 22 - 70 bar
	Mixed Solvents	
Sulfinol-D and Sulfinol-X	Mixture of Diisopropanolamine (DIPA) or MDEA, tetrahydrothiophene (DIPAM) and diethylamine or piperazine in water	>5 bar
ADIP	Mixture of 2-4 N DIPA and 2 N MDEA in water	35 - 40°C >1 bar
ADIP-X	Mixture of MDEA and piperazine in water	35 - 45°C 2 - 27 bar

(1) N stands for the molar amount of alkaline sites per litre of solution.
(2) Common process conditions when the CO_2 is captured.

6.3 Approach

In this study we used a gate-to-gate approach, i.e., CO_2 compression was included, but transport and storage of the CO_2 was not. To reduce transportation costs, CO_2 is transported in supercritical state. Its critical point is 73.8 bar and 31.1°C. In this study, the CO_2 is compressed to 110 bar to ensure that the CO_2 stays in supercritical phase during transportation.[243,244] As CO_2 capture requires both heat and electricity, an on-site dedicated combined heat and power (CHP) plant was included in the study. However, the CO_2 of the CHP plant was not captured as this would require most probably a second capture unit due to the differences in CO_2 concentration and pressure between the gas stream at the capture unit in the SMR facility and the flue gas of the CHP unit. Literature indicates that an additional (small) CO_2 capture unit is very costly.[26]

The techno-economic impacts of CO_2 capture depend on the chosen CO_2 capture system. To select the optimal CO_2 capture system, four main aspects were analysed: CO_2 capture position within the SMR process; solvent; configuration; and optimisation. This was followed by calculating the techno-economic impact of CO_2 capture using the selected CO_2 capture system. The technical performance was evaluated using the Procede Process Simulator (PPS), which is a rate-based steady-state simulator specifically designed for acid gas treating processes, developed by Procede Process Simulations B.V.[245,246] The simulator uses an in-house database for thermodynamics, physicochemical properties, mass transfer rates, kinetics, etc., to satisfy the vapour-liquid equilibrium (VLE) and mass and energy balances. The model was experimentally validated for the process conditions used in this study.[245,246] The CO_2 compression energy requirement and the capital costs for CO_2 capture were calculated using a TNO Excel-based model, consisting of an in-house database based on vendor quotes and literature and indexed using a European chemical engineering index.[247,248] Capital costs were calculated on a component basis and include direct costs (e.g., instrumentations, piping, electrical equipment and land) and indirect costs (e.g., land improvement, engineering, supporting buildings and contingencies). It excludes interest losses due to construction time.

6.3.1 CO_2 capture location

The first step was to analyse possible locations for the CO_2 capture unit in the SMR process. The analysis focused on two gas stream properties:

- First, to identify which bottlenecks occur when altering feedstock or production and how these bottlenecks can be resolved;
- Presence of contaminants (mainly O_2 and NO_x), degrading the solvent;
- CO_2 partial pressure, as a higher CO_2 partial pressure increases the driving force and reduces the size of the CO_2 capture unit.

From literature and interviews with experts[28], three potential CO_2 capture locations (see the arrows in Figure 6.2) were identified: location 1 is between the WGS reactor and the H_2-PSA unit, having the highest partial pressure of CO_2; location 2 is at the H_2-PSA waste stream, having the highest concentration of CO_2; and location 3 is at the flue gas of the reformer, having the highest mass flow of CO_2. Table 6.2 shows the relevant characteristics of the gas stream at these three locations.

Table 6.2 Overview location characteristics. (1)

	Contaminants presence	Total pressure (bar)	CO_2 concentration (mol%)	CO_2 partial pressure (bar)
Location 1	Low	20 - 31	15 - 35	3 - 11
Location 2	Low	0.14 - 2	30 - 60	0.04 - 1.2
Location 3	High	~ 1	5 - 10	~ 0.05 - 0.1

(1) SMR facilities can have different configurations and operating conditions. Also, the processed natural gas can differ in composition. This results in a large range in the CO_2 partial pressure at the three investigated CO_2 capture locations.

The presence of contaminants like O_2 and NO_x will drastically increase solvent deterioration, increasing operational costs.[249] Therefore, location 3 was rejected, even though almost all produced CO_2 could be captured. The main difference between locations 1 and 2 is the CO_2 partial pressure. The higher this pressure, the easier CO_2 capture occurs.[35] Therefore, the CO_2 capture location selected in this study is between the WGS reactor and the H_2-PSA unit (location 1).

6.3.2 Solvent selection

The second step was to select the solvent. The location of the CO_2 capture unit determines the composition and conditions of the gas stream fed to the CO_2 capture unit. A reference case (M-2) was selected based on typical industrial SMR process data published in Ullmann's Encyclopedia.[250] In order to evaluate the effect of syngas conditions on the CO_2 capture performance, several alternative cases were also investigated. They differ in syngas flow, composition and pressure. In the Middle cases, which includes the reference case, the CO_2 concentration is 20 $\text{mol\%}_{dry}$. This is altered to 15 $\text{mol\%}_{dry}$ in the Low cases and to 25 $\text{mol\%}_{dry}$ in the High cases. The remaining gases were scaled accordingly while preserving their original ratios. In all cases it was assumed that the syngas is saturated. The characteristics of the different cases are given in Table 6.3 and Table 6.4. The syngas flow rate was varied between 1000 kmol/h for the -1 cases, 5000 kmol/h for the -2 cases, and 10,000 kmol/h for the -3 cases.

Table 6.3 Syngas composition (mol% dry).

Case	**CO_2**	**H_2**	**CO**	**CH_4**	**N_2**	**H_2O**
Low	15.0	79.7	0.2	4.9	0.2	Saturated
Middle	20.0	75.0	0.2	4.6	0.2	Saturated
High	25.0	70.3	0.2	4.3	0.2	Saturated

---

[28] Experts from consultancy company Procede, research organisation TNO and a large petro-chemical company have been consulted throughout this article.

Table 6.4 Characteristics of the cases evaluated in this study.
The reference case (M-2) is highlighted in bold.

Case	Pressure	CO_2 conc.	CO_2 partial pressure	Syngas flow rate	CO_2 flow	H_2 flow	
	Bar	Mol%$_{dry}$	Bar	kmol/hr	kt/yr (1)	MW$_{HHV}$	kt/yr (1)
L-1	20	15	3.0	1,000	52	63	13
L-2				5,000	259	315	63
L-3				10,000	519	630	126
M-1	**25**	**20**	**5.0**	1,000	69	59	12
M-2 (2)				**5,000**	**346**	**297**	**59**
M-3				10,000	692	593	119
H-1	30	25	7.5	1,000	87	56	11
H-2				5,000	433	278	56
H-3				10,000	865	556	111

(1) An availability of 7,884 hours per year (90%) was assumed.
(2) Reference case. Data based on Ullmann's Encyclopedia.[250]

The most promising solvent was selected based on the characteristics of the reference case (M-2). For each solvent the following criteria were evaluated:

- Energy requirement - the amount of energy required to capture, purify and compress 1 kg CO_2;
- Loading capacity - the amount of CO_2 that can be absorbed by 1 mol solvent;
- Corrosiveness - the ability of the solvent to corrode the metallic equipment.;
- Vapour pressure - a low vapour pressure reduces the amount of solvent that ends up in the gas stream;
- Chemical stability - the ability of the solvent to resist chemical degradation.

Note that quantitative performance data for the solvents studied are only available for operating conditions which are different from those used in the current study.[241] Therefore, the data available was extrapolated to the used operating conditions and complemented by consulting industry experts, resulting in the qualitative comparison given in Table 6.5.

Both the physical and chemical solvents have a clear trade-off between loading capacity per time unit and energy requirement. Solvents with a low regeneration energy requirement also have a low loading capacity per time unit and solvents with a high loading capacity per time unit also have a high energy requirement. By mixing different solvents, it is possible to obtain solvents with both a low energy requirement and a high loading capacity per time unit.

Based on the analysis, the most promising solvents (high loading capacity per time unit and chemical stability and low regeneration energy requirement, corrosiveness and vapour pressure) were Sulfinol-X and ADIP-X. As there is much more information and expertise about ADIP-X than Sulfinol-X, ADIP-X was selected for this study.

Table 6.5 Evaluation of CO_2 capture solvents when applied in the reference case (M-2).[241,251] (1)

		Energy requirement	Loading capacity per time unit	Corrosive-ness	Vapour pressure	Chemical stability
Physical solvents	Rectisol	+	-	+	++	N.A.
	Selexol	+	-	+	++	N.A.
Chemical solvents	MEA	---	+++	--	-	-
	MDEA	+++	-	++	+	++
	Potassium carbonate	-	+++	---	+++	+++
Mixed solvents	MEA-MDEA	+	++	O	O	O
	Sulfinol-D	++	+	++	+	++
	Sulfinol-X	+	+++	+++	++	+++
	ADIP	++	O	++	+	++
	ADIP-X	+	+++	+++	++	+++

(1) Good (+++) > neutral (O) > bad (---) performance; (N.A.) not applicable.

6.3.3 CO_2 capture unit configuration

The third step was to select the CO_2 capture unit configuration. Three configurations were analysed:

- First, to identify which bottlenecks occur when altering feedstock or production and how these bottlenecks can be resolved;
- Absorber-stripper configuration. This is the simplest configuration. The CO_2 is captured in the absorber using a lean solvent. The now rich solvent is fed into the stripper. By increasing the temperature or lowering the pressure, the CO_2 is removed from the solvent. The lean solvent is cooled, pressurised and fed back into the absorber;
- Absorber-flash-stripper configuration. This configuration is identical to the previous one, except for an additional flash unit between the absorber and the stripper to remove dissolved hydrocarbons from the solvent;
- Absorber-still-stripper configuration. This is the most complex configuration. The solvent is regenerated in two steps. The rich solvent leaving the absorber is fed into a still, a combination of a heater and a flash unit. A fraction of the CO_2 evaporates from the solvent and is collected. The now semi-lean solvent is split into two parts. One part is fed directly into the absorber, at several meters below the top. The other part is fed into the stripper, where the solvent is completely regenerated and then fed into the top of the absorber.

Table 6.6 CO_2 capture configuration characteristics at reference case conditions. (1)

Technology	Simplicity	O&M cost	Capital costs	CO_2 purity	Maturity
Absorber-stripper	+	0	+++	++	+++
Absorber-flash-stripper	0	0	+	+++	+++
Absorber-still-stripper	---	++	0	++	++

(1) Results are based on interviews with industry experts working in the (petro)chemical industry. Good (+++) > neutral (O) > bad (---) performance.

All three configurations are common practice in the (petro)chemical industry.[251] The characteristics of the three configurations were qualitatively assessed using the judgements of industry experts, specialised in refineries (see Table 6.6).

The absorber-stripper configuration is the simplest one, resulting in low capital costs, average O&M costs and sufficient CO_2 purity. The addition of a flash improves the purity of the CO_2 stream. However, as the CO_2 hardly contains any hydrocarbons, the increase in purity is minor. Adding a flash unit increases capital costs and decreases simplicity. The addition of a still reduces O&M costs, but capital costs increase and simplicity decreases significantly. For this study, the absorber-stripper configuration was selected because of its low capital costs, simplicity, robustness and the widely available knowledge of this configuration. A schematic process diagram is given in Figure 6.4.

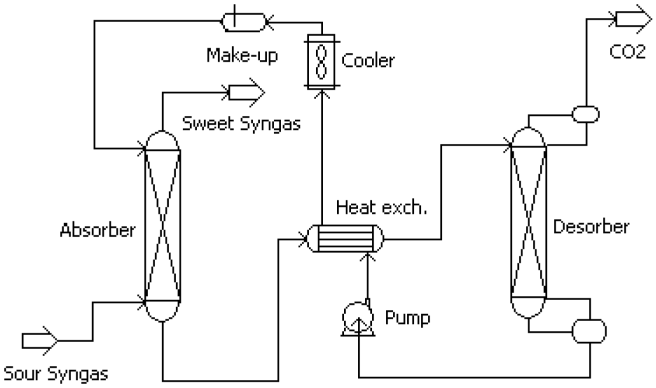

Figure 6.4 Schematic overview absorber-stripper configuration.

6.3.4 *Optimisation of CO_2 capture configuration conditions*

The fourth step was to optimise capture configuration conditions. Three parameters were optimised as they have the largest influence on the capture unit performance and capital costs:

- The absorber and regenerator pressure - this influences equipment size as well as energy requirement and regeneration temperature;
- The capture solvent flow - this also influences equipment size, energy requirement and regeneration temperature;
- The capture solvent composition - this refers to the concentrations of amine and accelerator with water being the third component and influences, among others, reaction rates and thus equipment size as well as energy requirement.

ADIP-X is a mixture of an amine and an accelerator. In our study it was assumed to be MDEA and piperazine. Based on expert interviews, a solvent composition of 45 wt% MDEA, 5 wt% piperazine and 50 wt% water was selected, typical for the (petro)chemical industry. The diameters of the absorber and the regenerator could be calculated based on the solvent and gas flows as the selected packing (Sulzer MellaPak Metal 250Y) dictated a constant flooding factor of 0.8.[252] Flooding is a condition where (a tray in) the column cannot carry both the upward gas flow and the downward liquid flow. At a flooding factor of 1 (the flooding point) the gas flow is preventing the entire liquid flow from flowing downwards, resulting in a build-up of liquid inside the column and reducing column efficiency. When the flooding factor exceeds 1, column efficiency is reduced even further and the column can trip.

6.3.4.1 Absorber and regenerator pressure

The absorber pressure was set at the syngas pressure to keep the syngas at the highest possible pressure without recompression. Based on simulation results, the regenerator pressure was set at 2 bar. Although a higher regeneration pressure would lower the energy required for compressing the CO_2, it would also increase the reboiler temperature, which in turn would increase amine degradation.[253] Simulation results showed that increasing the regenerator pressure from 2 bar to 20 bar raised the reboiler temperature from 109°C to 159°C.

6.3.4.2 Effect of solvent flow

The optimal solvent rate was defined as the minimal rate at which the reboiler temperature was under 110°C and was determined by running several simulations of the CO_2 capture process using the reference case (M-2) with different solvent flow rates (Figure 6.5).

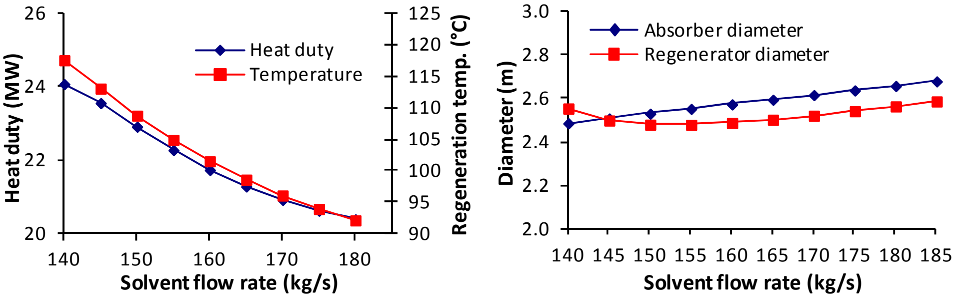

Figure 6.5 Impact of solvent rate on reboiler heat duty and temperature (left) and on absorber and regenerator diameter (right).

As shown in Figure 6.5, increasing the solvent flow rate increased the diameter of the absorber as the volume of solvent passing per second increased. Results indicated that at flow rates lower than 140 kg/s, the system had difficulties capturing 90% of the CO_2 and the lean solvent had to be almost completely CO_2-free. Increasing the solvent rate increased the richness of the lean solvent. This resulted in reduced regeneration duty and lower boiler temperature. The lower temperature resulted in less water evaporation and, therefore, a lower gas flow in the regenerator, allowing for a smaller column. At reference case conditions, the regenerator diameter reached a minimum at 150 kg/s. At this flow, both columns had a diameter of about 2.5 m. Increasing the solvent flow rate above the 150 kg/s increased the regenerator diameter again, due to the increase in solvent volume.

6.3.4.3 Effect of piperazine concentration

The impact of the piperazine concentration on the performance of the CO_2 capture unit was investigated to determine the effect of using an accelerator compared to using one amine only. The accelerator serves as a transfer agent, reacting with CO_2 and delivering the adsorbed CO_2 to the MDEA, speeding up the formation of MDEA-CO_2. For detailed information on the mechanism and reactions, we refer to Derks.[254] The reference case (M-2) was taken as starting point. The piperazine concentration was varied between 0 and 5 wt%. The MDEA concentration was kept constant at 45 wt% and the solvent flow at 150 kg/s. Altering the piperazine concentration had an impact on the performance of the capture unit. This could be countered by altering the heat duty or the height of the absorber. Therefore, two different series of simulations were performed.

In the first series, the heat duty was kept constant (see Figure 6.6, left). Increasing the piperazine concentration decreased the absorber height. But when piperazine concentrations exceeded 3 wt% the absorber height started to increase. Increasing the piperazine concentration from 0 to 3 wt% reduced the required absorber height by 70%. Two effects occurred when increasing the piperazine concentration. Firstly, the additional piperazine increased the overall mass transfer of the CO_2 absorption. Secondly, the richness of the lean solvent increased as reboiler temperature decreased, reducing the mass transfer rate of the CO_2-absorption. At low piperazine concentrations the effect of the additional piperazine was large, while the effect of the solvent loading was small. This resulted in faster overall mass transfer rates, allowing a smaller absorber. However, above 3 wt% piperazine the decrease in mass transfer rates, due to increasing richness of the solvent, became dominant and the overall mass transfer rate of the CO_2 absorption decreased. This resulted in a slightly larger absorber.

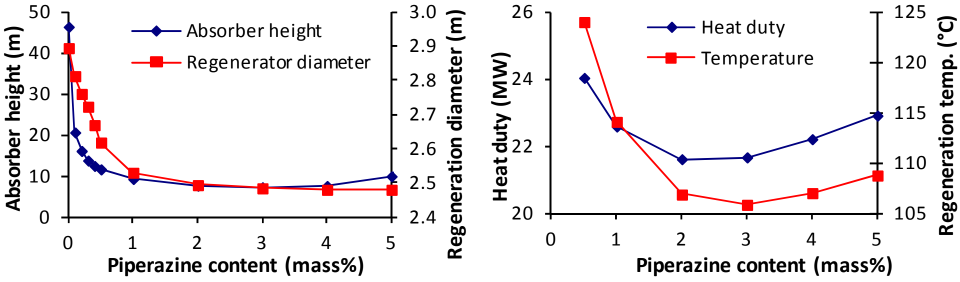

Figure 6.6 Effect of piperazine concentration on absorber and regenerator dimensions with constant regenerator duty (left) and on heat duty and regeneration temperature with constant absorber dimensions (right).

In the second series, the absorber height was kept constant (see Figure 6.6, right). The same two opposing effects on the mass transfer rates occurred as with the first series. At low piperazine concentrations, mass transfer rates were so slow that the lean solvent had to be virtually CO_2 free in order to achieve a capture rate of 95%. This was only achieved by increasing the temperature in the regenerator, which increased total energy demand. Similar to the first series, around a piperazine concentration of 3 wt% there was a minimum in temperature and heat duty. Increasing the piperazine concentration from

0.5 to 3 wt% lowered the required reboiler heat duty by 10% and the reboiler temperature from 124°C to 106°C. This could allow the use of low pressure steam (<120°C) as heat source for the reboiler.

6.3.5 *Overview of technical parameters*

Based on the previous results, the selected CO_2 capture unit has the parameters shown in Table 6.7. These parameters were used to calculate the technical and economic performance of the different case studies. The analysis on piperazine concentration showed that the optimal piperazine concentration was 3 wt%. In practice, however, the industry uses a 5 wt% concentration. To allow comparison between our study and industrial analyses, this study also used a 5 wt% piperazine concentration.

Table 6.7 Technical parameters CO_2 capture unit.

		Unit	Value		
Syngas	Flow rate	kmol/h	1.000	5.000	10.000
Capture solvent	Circulation flow rate	kg/s	30	150	300
	MDEA conc. (1)	wt%	45		
	Piperazine conc. (1)	wt%	5		
Absorber	Flooding factor (2)	-	0.8		
	Height (3)	m	10		
	Pressure (4)	bar	25		
Regenerator	Flooding factor (2)	-	0.8		
	Height (3)	m	5		
	Pressure (4)	bar	2		
CO_2 compressor	Specific compr. energy (5)	$kJ_e/kg\ CO_2$	302		
CO_2 capture unit	CO_2 capture target	%	95		
SMR Non-capture	Specific CO_2 emission (6)	$kg\ CO_2/GJ\ H_2$	62		
H_2-PSA	H_2 recovery rate	%	90		

(1) Based on expert interviews at an engineering consultancy company.[245,246]
(2) Based on the used Sulzer MellaPak Metal 250Y packing.[252,255]
(3) Based on the smallest height at which absorber or regenerator efficiency was hardly affected.
(4) Absorber pressure is set similar to the syngas pressure.
(5) Result from the TNO model used.
(6) Besides natural gas for the reformer, additional natural gas (make-up fuel) is added to supply the necessary heat for the reformer. In this study, it was assumed that the non-capture SMR has an H_2 efficiency of $80\%_{HHV}$ ($75\%_{LHV}$).[233,250] Assuming natural gas is purely CH_4, this translates into 62 kg CO_2/GJ H_2.

6.3.6 *Techno-economic performance of CO_2 capture at SMR*

Once the optimal CO_2 capture system was selected, the technical and economic impacts of capturing CO_2 at SMR facilities could be examined. As mentioned in Section 6.3.2, Table 6.3 and Table 6.4, a reference case and eight alternatives were analysed. These alternatives reflect combinations of different gas flow rates, CO_2 concentration and total pressure compared to the reference case. For each case, the following parameters were evaluated:

- First, to identify which bottlenecks occur when altering feedstock or production and how these bottlenecks can be resolved;
- Energy consumption; both heat and electricity, which will be supplied by an on-site CHP;
- Equipment size; absorber, regenerator, heat exchanger, pump and compressor;
- Capital cost; total investment costs of the CO_2 capture, CO_2 compression and CHP units;
- CO_2 capture and avoidance cost per t CO_2 and H_2-CO_2 capture production cost per GJ H_2 (the H_2 production costs including CO_2 capture).

The energy consumption for capturing CO_2 and the equipment size were calculated using PPS.[245,246] The CO_2 compression energy requirements and the capital costs for the capture and compression equipment were determined by TNO.[247,248] The capital costs of the CHP were based on vendor quotes in the Catalog of CHP technologies.[256] The CO_2 capture and avoidance costs were calculated using equations 6-4 and 6-5, while the H_2-CO_2 capture production costs were calculated using equation 6-6. No construction time for the CO_2 capture equipment was considered.

$$CO_2 \text{ capture costs} = \frac{\alpha * I + O\&M + Fuel}{F_{CO_2 \text{ captured}}} \tag{6-4}$$

$$CO_2 \text{ avoidance costs} = \frac{\alpha * I + O\&M + Fuel}{F_{CO_2 \text{ captured}} - F_{CO_2 \text{ CHP}}} \tag{6-5}$$

$$H_2\text{-CC production costs} = H_2 \text{ production costs} + \frac{\alpha * I + O\&M + Fuel}{F_{H_2 \text{ produced}}} \tag{6-6}$$

where α = capital recovery factor (yr^{-1}), calculated by $r/(1-(1 + r)^{-L})$;
r = discount rate;
L = economic lifetime (year);
I = capital costs (M€) of the CO_2 capture, compression and CHP units;
O&M = operating and maintenance costs (M€/yr) of the CO_2 capture, compression and CHP units;
fuel = cost of natural gas required to supply power and heat (M€/yr);
$F_{CO2 \text{ captured}}$ = annual flow of CO_2 that is captured (t CO_2/yr);
$F_{CO2 \text{ CHP}}$ = annual flow of CO_2 emitted by the CHP unit (t CO_2/yr);
$F_{H2 \text{ produced}}$ = annual flow of H_2 that is produced (GJ H_2/yr);
H_2-CC production costs = production costs of H_2 at a SMR facility equipped with CO_2 capture (€/GJ H_2)
H_2 production costs = production costs of H_2 at a SMR facility without CO_2 capture (€/GJ H_2)

The economic parameters are given in Table 6.8. Note that all monetary values are in $€_{2008}$.

Table 6.8 Summary of main economic parameters used in this study.

Parameter	**Unit**	**Value**
Plant lifetime	Years	20
Availability	%	90
O&M costs	% of cap. costs	4
Discount rate	%	10
Overall CHP efficiency (1)	%	80
Natural Gas (NG) 2008-price (2)	€/GJ	6.90

(1) The overall CHP efficiency is kept constant at 80%, regardless of the power-to-heat ratio.
(2) Data based on World bank commodity markets.[155]

6.4 Techno-economic results

6.4.1 CO_2 capture unit performance case studies

The technical performances of the different case studies were calculated using PPS. The technical performance of the CO_2 capture unit is given in Table 6.9 for each case.

Table 6.9 Technical results of the different case studies.
L = 20 bar, 15 mol% CO_2; M = 25 bar, 20 mol% CO_2; H = 30 bar, 25 mol% CO_2.
1 = 1,000 kmol syngas/hr; 2 = 5,000 kmol syngas/hr and 3 = 10,000 kmol syngas/hr.

	Case	**Unit**	**L-1**	**L-2**	**L-3**	**M-1**	**M-2**	**M-3**	**H-1**	**H-2**	**H-3**
	CO_2 captured	kg CO_2/s	1.7	8.7	17	2.3	**12**	23	2.9	15	29
Absorber	Diameter	m	1.1	2.5	3.5	1.1	**2.5**	3.6	1.1	2.6	3.6
	Base area	m^2	1.0	4.9	9.9	1.0	**5.0**	10	1.0	5.1	10
	Spec. diameter	m/(kg CO_2/s)	0.6	0.3	0.2	0.5	**0.2**	0.2	0.4	0.2	0.1
	Spec. base area	m^2/(kg CO_2/s)	0.57	0.57	0.57	0.43	**0.43**	0.43	0.35	0.35	0.35
Regenerator	Diameter	m	1.0	2.3	3.3	1.1	**2.5**	3.5	1.4	3.1	4.4
	Base area	m^2	0.9	4.3	8.6	1.0	**4.8**	9.7	1.5	7.6	15.3
	Spec. diameter	m/(kg CO_2/s)	0.6	0.3	0.2	0.5	**0.2**	0.2	0.5	0.2	0.2
	Spec. base area	m^2/(kg CO_2/s)	0.5	0.5	0.5	0.4	**0.4**	0.4	0.5	0.5	0.5
	Reboiler temp.	°C	99	99	99	109	**109**	109	124	124	124
	Spec. heat duty	MJ/kg CO_2	1.96	1.96	1.96	1.97	**1.97**	1.97	2.13	2.13	2.13

It was found that the temperature to which the capture solvent was heated during regeneration increased with increasing CO_2 partial pressure, from 99°C at a p_{CO2} of 3 bar to 124°C at a p_{CO2} of 7.5 bar. As a result, the specific energy consumption of the CO_2 capture unit increased, from 1.96 MJ/kg CO_2 at a p_{CO2} of 3 bar to 2.13 MJ/kg CO_2 at a p_{CO2} of 7.5 bar. This was not an effect of the higher CO_2 partial pressure, but of the increased CO_2 flow. As the capture solvent needed to carry increasing amounts of CO_2, the lean solvent needed to have a lower concentration of CO_2. This was achieved by increasing the temperature of the solvent, resulting in a slight increase in specific energy consumption. The higher driving force due to the higher p_{CO2} could not compensate this effect.

The absorber base area showed a linear relation with the sour syngas flow, which was a direct result of the constant flooding factor of 0.8. Increasing p_{CO2} or syngas flow resulted in a more efficient absorber as the specific absorber diameter dropped from 0.60 m per kg/s for the L-1 case to 0.19 m per kg/s for the L-3 case and 0.15 m per kg/s for the H-3 case.

As with the absorber, the specific regenerator diameter decreased significantly from the L- to the M-cases. No scale benefits, however, appeared between the M- and H-cases, mainly because of the higher boiler temperature, resulting in significant water evaporation and thereby increasing the gas flow in the regenerator, resulting in an increase of the diameter of the regenerator. The results indicate that the specific regenerator duty and reboiler temperature are independent of syngas volume flow. This, however, is a result of the simulation programming as heat losses were not taken into account.

6.4.1.1 Energy and carbon balances

The energy- and CO_2 mass balances are given in Table 6.10. As the CO_2 from the make-up fuel and the CHP was not captured, net CO_2 emissions for the M-cases were reduced by only 60%. Heat requirement for solvent regeneration accounted for about 85% of total energy demand.

Table 6.10 Calculated energy and CO_2 mass balance SMR-CC unit.
L = 20 bar, 15 mol% CO_2; M = 25 bar, 20 mol% CO_2; H = 30 bar, 25 mol% CO_2.
1 = 1,000 kmol syngas/hr; 2 = 5,000 kmol syngas/hr and 3 = 10,000 kmol syngas/hr.

Case	H_2 prod.	CO_2 capture eff.	Heat and electricity			Energy penalty	CO_2 captured	Extra CO_2 from CHP	Net CO_2 avoided
			Capture		Compr.				
	MJ_{HHV}	(%)	(MJ_{th})	(MJ_e)	(MJ_e)	(MJ/GJ H_2)	(kt/yr)	(kt/yr)	(kt/yr)
L-1	57	95%	3	0.1	0.5	71	49	7	42
L-2	283	95%	17	0.3	2.6	71	247	35	212
L-3	567	95%	34	0.7	5.3	71	494	70	424
M-1	53	95%	5	0.1	0.7	101	66	9	56
M-2	**267**	**95%**	**23**	**0.5**	3.5	**101**	**330**	**47**	**282**
M-3	534	95%	46	0.9	7.0	101	659	94	565
H-1	50	95%	6	0.1	0.9	144	82	13	70
H-2	250	95%	31	0.6	4.4	144	412	63	349
H-3	500	95%	62	1.1	8.8	144	824	126	698

6.4.1.2 Economic performance

Based on the results from the process analysis, the costs of capturing CO_2 at a SMR facility were calculated. The results (Table 6.11 and Figure 6.7) show the importance of the energy consumption, being responsible for 23-25 €/t CO_2 avoided. The results also show the benefits of economies of scale for the capture and compression units. The share of capital and O&M costs of the capture and compressor equipment in the total avoidance costs was 47% in the M-1 case, but only 20% in the H-3 case. As the specific natural gas consumption did not change with scale, the natural gas share increased from 35% in the M-1 case to 66% in the H-3 case. The CHP costs also experienced economies of scale, but much smaller. The share of CHP in the overall costs decreased from 19% in the M-1 case to 15% in the H-3 case. The increase in H_2 production costs was between 0.9 and 2.7 €/GJ H_2 (see Table 6.11 and Figure 6.7).

Table 6.11 Economics of CO_2 capture at SMR facilities.

Case	**Unit**	**L-1**	**L-2**	**L-3**	**M-1**	**M-2**	**M-3**	**H-1**	**H-2**	**H-3**
Capital recovery	M€/yr	1.4	3.3	4.9	1.6	**3.8**	5.7	1.8	4.5	6.8
O&M costs	M€/yr	0.5	1.1	1.7	0.5	**1.3**	2.0	0.6	1.5	2.3
NG costs	M€/yr	1.0	4.9	9.8	1.3	**6.6**	13.2	1.8	8.8	17.6
Total costs	**M€/yr**	**2.8**	**9.3**	**16.3**	**3.4**	**11.7**	**20.8**	**4.2**	**14.8**	**26.8**
CO_2 capture costs (1)	€/t CO_2	57	37	33	52	**35**	32	51	36	32
CO_2 avoidance costs (1)	€/t CO_2	67	44	39	61	**41**	37	60	42	38
Additional H_2 production costs	€/GJ_{H2}	1.6	1.0	0.9	2.0	**1.4**	1.2	2.7	1.9	1.7

(1) This includes capture and compression to 110 bar, but excludes transport and storage.

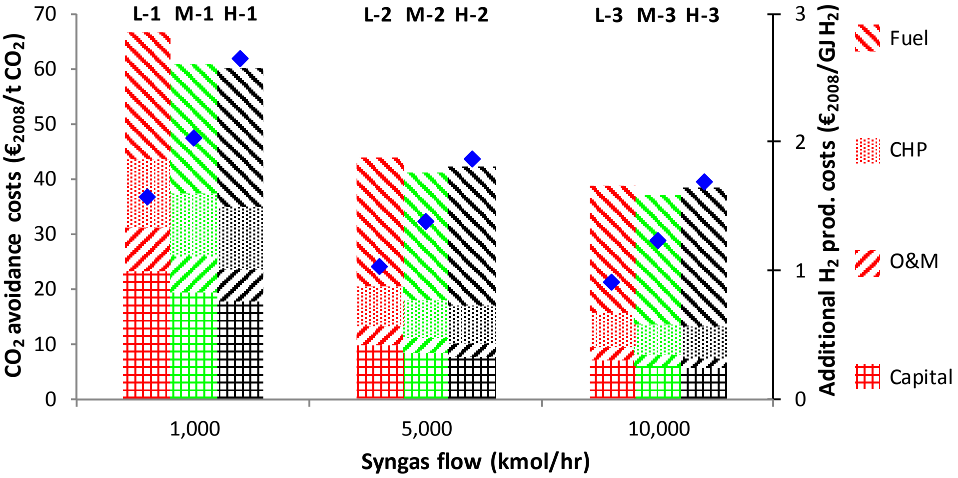

Figure 6.7 CO_2 capture costs for the different case studies. The bars refer to the CO_2 capture costs; the dots refer to the extra costs for H_2 prod. due to CO_2 capture. The capital recovery and O&M contain only the costs related to CO_2 capture and compression.

The lowest CO_2 capture and avoided costs occurred in the M-cases. The only exception was the H-1 case, where the CO_2 avoidance costs were slightly lower than those of the M-1 case (60 vs. 61 €/t CO_2 avoided).

Increasing the scale did reduce specific CO_2 capture costs, but this effect diminished with increasing scale. Between the -1 and -2 cases a 29-34% decrease in specific capture costs was observed, while between the -2 and -3 cases this reduction was only 9-12%. Operating the CO_2 capture unit at elevated pressures had a strong effect on the capital costs, reducing its contribution to 12-20% of the total capital costs of the capture, compressor and CHP units.

The used model assumed no heat losses and used a piperazine concentration of 5 wt%. The effect of these assumptions on the CO_2 avoidance costs of the reference case (M-2) was also calculated. Heat losses will increase costs as more natural gas is consumed, but also as a larger CHP is needed, increasing capital and O&M costs. It was found that for every 1% increase in heat demand, CO_2 avoidance costs increased by 0.3 €/t CO_2, raising annual costs by 68 k€. The effect of using the optimal piperazine concentration (3 wt% vs. the industrial used 5 wt%, see §6.3.4.3) was a reduction in heat requirement of 5%, resulting in 4% lower CO_2 avoidance costs. This translated to a 0.4 M€ decrease in annual costs.

6.4.2 Effect load factor on CO_2 capture unit performance

It may sometimes be necessary to operate the CO_2 capture unit at non-optimal load. Therefore, the effect of operating the unit at different loads was investigated. The reference case (M-2) was taken as starting point. The syngas flow was altered, while the absorber and regenerator diameter and height were kept constant. Two simulation series were performed. In the first series the solvent rate was kept constant, while in the second series the solvent rate was adjusted proportional to the syngas flow rate.

Keeping the solvent flow constant with increasing syngas flow raised the specific energy consumption of the capture unit (Figure 6.8, left). As the same amount of solvent has to transport more CO_2, reboiler temperature increased to reduce the loading of the lean solvent, thereby increasing the CO_2 capture capacity of the solvent. The result was a higher heat duty. When the solvent flow was adjusted proportional to the syngas flow (Figure 6.8, right), both the reboiler temperature and the specific energy consumption remained almost constant as the CO_2 concentration in the lean solvent remained almost constant.

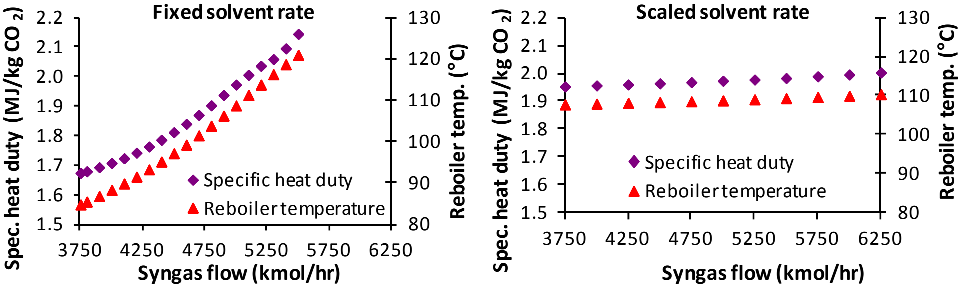

Figure 6.8 Effect of load operation on energy requirement for fixed capture solvent flow (150 kg/s) (left) and variable capture solvent flow (right).

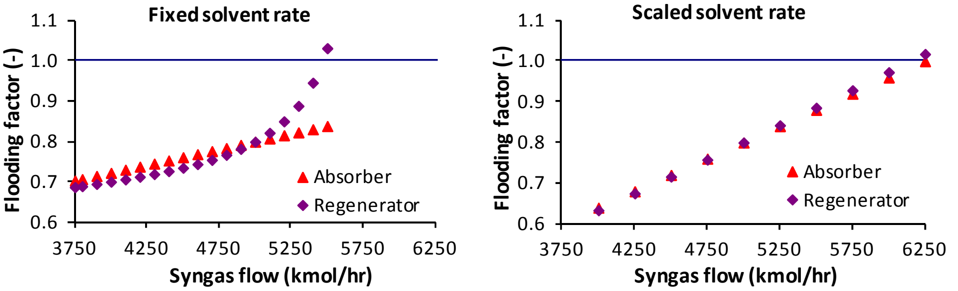

Figure 6.9 Effect of load operation on flooding factor of absorber and desorber for fixed capture solvent flow (150 kg/s) (left) and variable capture solvent flow (right).

Increasing the solvent rate, however, increased the flooding factor of both the absorber and the regenerator (Figure 6.9, right). At high flooding factors an increase in pressure drop over the column as well as a decrease in column efficiency will occur. In addition, when the flooding factor exceeds 1 (the flooding point), constant and stable column operation conditions can no longer be achieved. An increase in syngas and solvent flow by 20% resulted in a flooding factor of almost one in both columns. When keeping the solvent flow constant, this flooding factor was already reached in the regenerator at a 10% increase in syngas flow (Figure 6.9, left). This was a direct result of the higher reboiler temperature which increased the evaporation of water and thus the gas flow in the regenerator.

Altogether, this means that if the CO_2 capture unit operates under its design load, solvent flow should be kept constant. This results in the largest reduction in both reboiler temperature and duty; a 10% reduction in syngas flow reduced heat demand by 17% and lowered reboiler temperature by 12°C. If the CO_2 capture unit operates above its design load, solvent flow should be adjusted proportional to the syngas flow. Not only will this limit the increase in reboiler temperature and duty; a 10% increase in syngas flow increased heat demand by 11% and raised reboiler temperature by 1°C - it will also slow down the increase in flooding factor, improving the availability of the capture unit. It will be possible to operate the CO_2 capture unit at syngas flows exceeding the design flow with almost no increase in specific heat demand. This will, however, reduce the safety margin of the capture unit, which could lead to a reduction in availability. What this means for the economic performance of the capture facility is unclear as an analysis on how availability is affected was outside the scope of this study.

6.4.3 Sensitivity analysis

A sensitivity analysis was performed for the reference case (M- 2). The varied parameters are displayed in Table 6.12. The values were varied between 50% and 150% of their original value. The CHP efficiency was varied between 50 and 100%. Results indicated that the economic performance is most sensitive to the CHP efficiency and natural gas price (Figure 6.10).

Table 6.12 Input sensitivity analysis.

Parameter	Unit	Normal
CHP efficiency	%	80
Capital costs	M€	32.3
Discount factor	%	10
Natural gas price	€/GJ	7.33

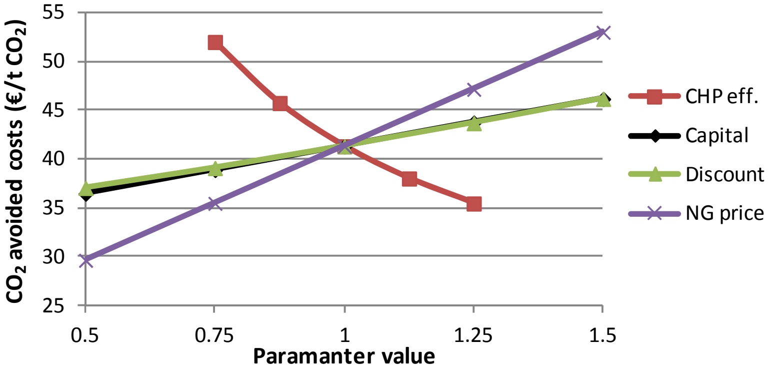

Figure 6.10 Sensitivity analysis reference case (M-2).

6.4.4 Utility import

An alternative to producing the needed electricity and steam is to import them. This would remove the need for the CHP. An emission factor of 370 kg CO_2/MWh (≈103 kg CO_2/GJ) was used for the imported electricity.[257] The steam was assumed to be excess steam. Therefore, no CO_2 emissions were allocated. Again, the reference case (M-2) was taken as example. The cost of electricity was assumed to be 15.7 €/GJ (≈57 €/MWh), which was the average hourly spot market price in the Netherlands during 2005-2008.[66] At this price, electricity consumption accounts for 5.6 €/t CO_2 avoided, although it is expected that large-scale industrial users can negotiate lower electricity prices. Results (Figure 6.11) show that steam import becomes break-even with an on-site CHP at 12.7 €/GJ. As regeneration took place at 109°C, the steam must be at around 120°C when arriving at the regenerator. Assuming saturated steam of 130°C and 10 €/t steam,[258] this corresponds to about 4 €/GJ. At that price, CO_2 avoidance costs were 24 €/t CO_2, a reduction of 43%. If the steam is available for free, CO_2 avoidance costs would drop to 15 €/t CO_2, a reduction of almost 63%.

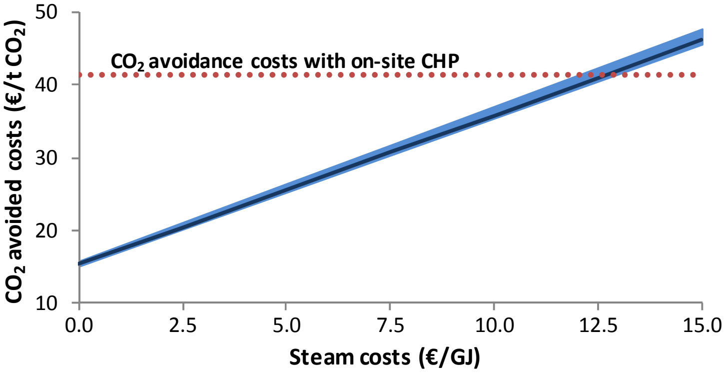

Figure 6.11 CO_2 avoidance costs as function of steam import price, including steam transport.

Electricity costs 15.7 €/GJ_e and has an emission factor of 375 kg CO_2/MWh (line). The shaded area corresponds to emissions of 300 kg CO_2/MWh (60% eff. NGCC) and 750 kg CO_2/MWh (40% eff. PC).

6.4.4.1 H_2-production costs

Industrial H_2 production costs are unknown, partly as most H_2 produced is used directly by the same company in the same refinery. H_2 production costs including CO_2 capture can be calculated as a function of H_2 production costs without CO_2 capture. For the reference case (M-2) the CO_2 capture costs were 1.4 €/GJ H_2. At H_2 production costs of 5-9 €/GJ, capturing CO_2 leads to 15-28% higher production costs. As shown in Figure 6.12, our results are comparable to CO_2 capture penalties published in the literature (between 0.4 and 2.6 €/GJ H_2), even though we assumed a retrofit instead of a SMR facility with integrated CO_2 capture.

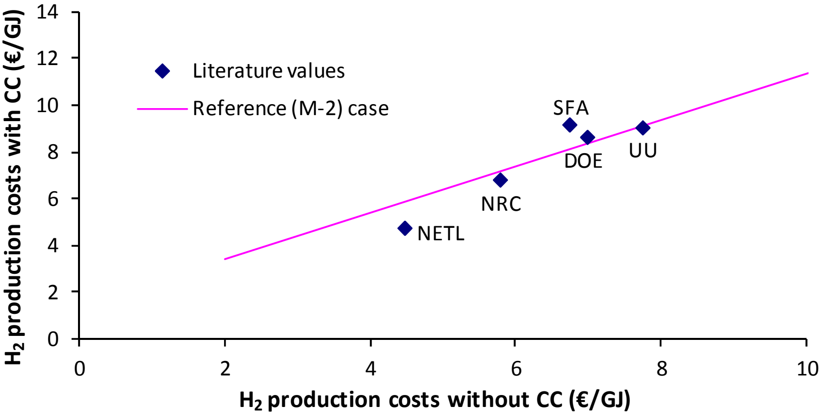

Figure 6.12 H_2-CC production costs as function of H_2 production costs without CO_2 capture, showing the results of this study (line) and literature values (dots).[232,259,260,261]

6.4.5 Conclusion

This study aimed to explore the relation between cost, configuration and performance of equipping a SMR facility with CO_2 capture by identifying the optimal techno-economic SMR-CO_2 capture configuration using currently available technologies. Based on the results, the following conclusions about the technology were drawn:

- The optimal CO_2 capture system currently available for SMR facilities is an ADIP-X based absorption process located between the WGS and PSA units. Implementing this capture process reduces specific CO_2 emissions compared to a conventional non-capture SMR facility by 60%. The remaining CO_2 emissions are from the CHP (10%) and the make-up fuel (30%);
- The addition of small amounts of piperazine (from 0% to 3 wt%) to a MDEA-based CO_2 capture process improves the techno-economic performance of the capture process by reducing the equipment size (absorber height reduced by 70%) or lowering the reboiler heat duty (-10%) and maximum temperature (-18°C);
- The optimal piperazine concentration found in this study (3 wt%) is lower than the widely used industrial concentration of 5 wt%, which is the optimal

concentration when looking at the regenerator only. Optimising the entire CO_2 capture unit, however, resulted in the optimal piperazine concentration of 3 wt%;
- Sometime it might be necessary to operate the CO_2 capture unit at a different load than the design load. Although this is possible, at syngas flows above the design load, the flooding point (flooding factor = 1) is quickly exceeded, resulting in decreased column efficiency or even tripping of the column. When operating above the design load, it is best to adjust the solvent flow proportional to the syngas flow. Not only does this results in lower specific energy consumption compared to keeping the solvent flow constant, it also allows for a larger deviation in the syngas flow (+20% vs. +10%). When operating at lower loads than the design loads, it is best to keep the solvent flow constant. This results in a lower specific energy consumption than when the solvent flow is downscaled (-15% vs. -1%).

Based on the combination of the technical and economic models, the following conclusions about the economics were drawn:

- CO_2 avoidance costs for retrofitting an existing SMR facility are estimated at 41 €/t CO_2, with uncertainties ranging between 37 and 67 €/t CO_2. This is at the lower end of the values reported in literature for CO_2 capture at refineries (26-82 €/t CO_2) and below the literature values for integrated CO_2 capture at a natural gas-fired power plant (44-93 €/t CO_2). Absorbing the CO_2 at elevated pressures could be a main reason why the CO_2 avoidance costs are at the lower end of the literature values. The high pressure results in smaller columns, drastically reducing capital costs;
- CO_2 capture will increase H_2 production costs by 1.4 €/GJ for the reference case. If H_2 production costs are between 5 and 9 €/GJ, CO_2 capture increases H_2 production costs by 15-28%;
- The CHP and especially the natural gas consumption play a key role in the CO_2 capture costs, having a share of 50-80%. Importing the required electricity and steam instead of on-site generation could reduce CO_2 avoidance costs by 43%. Due to the low power to heat ratio, another option is to import electricity and use a boiler to generate the steam. Due to their simple design, boilers are much cheaper than CHP installations, which should result in lower CO_2 avoidance costs.

The technical results are considered sound due to the combination of computer modelling and lab-scale experimental validation.[245,246] The model did not, however, include heat losses in the absorber or regenerator. The economic results are more indicative, being based on data supplied by vendor quotes and from open literature.

Based on the results three recommendations for further research are proposed. First, this study used two different computer models, a simulator from Procede Process Simulations B.V. and an economic model from TNO. To verify the results from these models, a pilot project is needed, even though the models have been calibrated using, among other, experimental results and historic data. Second, this study investigated retrofitting an existing facility for CO_2 capture. CO_2 avoidance costs should be considerably lower when designing a new H_2 SMR facility as, for instance, the CHP needed for the electricity and steam could be provided by a central CHP providing utility to the entire facility. Finally, in this study only commercially available technologies were considered. The impact of more advanced technologies, such as autothermal reforming or combined water-gas shift and CO_2 capture, could result in a substantial reduction of the CO_2 avoidance costs, as pointed out by Damen et al.[71] It is recommended that the potential costs reduction and expected availability of these new technologies are evaluated.

To conclude, in this study an optimal techno-economic CO_2 capture configuration for SMR facilities was identified. By using this configuration CO_2 emissions can be reduced by 61% at 41 €/t CO_2 avoided, making it more economical than CO_2 capture at gas-fired power plants. Applying this technique at the existing SMR facilities could reduce CO_2 emissions by 2.4 Gt CO_2 in the coming 20 years.

6.4.6 Acknowledgements

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7 Summary, conclusions and recommendations

7.1 Background

Our society demands abundant, affordable, reliable and environmentally sound sources of energy. As population and living standards are increasing, so is the global energy demand.[1,2,3,4,5] In the Energy Technology Perspectives 2010 report by the International Energy Agency (IEA), global energy demand is projected to exceed 750 EJ by 2035,[4] compared to 500 EJ in 2009.[1,2,3] Fossil fuels may by then supply more than two-third of the global energy demand. This large use of fossil fuels can have important drawbacks, including emission of anthropogenic greenhouse gases (GHG).[2,4,5,9,10,11] Combustion of fossil fuels resulted in 29 Gt emitted CO_2 in 2009. Almost 85% of these emissions came from the sectors power (11.8 Gt CO_2), transport (6.5 Gt CO_2) and industry (5.9 Gt CO_2).[20]

Recently, the Conference of the Parties (COP) to the United Nations Framework Convention on Climate Change (UNFCCC) agreed that global warming should not exceed 2°C above pre-industrial levels to prevent dangerous climate change.[13,14,15] When looking at CO_2, this requires most likely a reduction in CO_2 emissions of 50-80% by 2050.[11,15,18] To realise such emission reductions a broad portfolio of mitigation options is probably needed, including an enhanced improvement of the efficiency of energy conversion processes, a switch to less fossil carbon-intensive energy sources, and application of carbon dioxide capture and storage (CCS) technologies.[2,4,5,6,7]

There is a significant potential for both the application of biomass and CCS to reduce GHG emissions. However, current deployment of especially CCS is low. Gasification can play an important role in enhancing the development and deployment of both mitigation options. Gasification facilities can convert biomass into high value products at relatively low costs. CO_2 is often already separated in existing gasification facilities producing chemicals. Consequently, its capture can be achieved at significantly reduced CO_2 avoidance costs.

However, the current deployment rate of gasification facilities is low, mainly due to the high investment costs involved and rapidly changing commodity prices.[42,60,61,62,63,64,65,66] Production costs of commodities could, however, be reduced significantly by technological learning which could reduce the capital costs and improve the overall conversion efficiency of facilities in the short to longer term.[58,59] Achieving these improvements requires investments in research and development (R&D) and a significant increase in installed gasification capacities.

In the short term, the ability to use different feedstocks and/or to alter the production of commodities in a flexible manner might stimulate the deployment of gasification facilities for two reasons. First, the ability of integrated gasification poly-generation (IG-PG) facilities to be flexible in feedstock use allows the construction of large-scale facilities running (partly) on biomass, apart from fossil fuels, thereby reducing the CO_2 emissions and taking advantage of economies of scale, but without the risk of disturbances in the production process induced by potential disturbances in the supply of biomass feedstock. Second, the ability to respond to and exploit fluctuations in market price of feedstock or commodities reduces the investment risk. Currently, there is a need for better understanding what the possibilities and limitations of flexibility at gasification facilities are and under which (economic) conditions flexible facilities are more advantageous than their static counterparts. An in-depth analysis of these aspects is at the core of this thesis.

Literature indicates that the performance of IG-PG facilities can improve in the future. However, the rate of this improvement and the drivers behind it are at present poorly understood. In this thesis the potential improvement in gasification performance is assessed using a bottom-up and a top-down approach and subsequently comparing the results with each other.

A promising (natural gas-fired) gasification application in industrial facilities for the deployment of CCS in the short term is steam methane reforming (SMR). At the SMR facilities, the CO_2 avoidance costs are projected to be lower compared to the avoidance costs at power plants due to the presence of a pressurised CO_2 rich gas stream and the possible availability of waste heat streams, as SMR facilities are generally located inside large industrial complexes, like refineries. The impact of applying CCS at existing facilities (retrofit) and the effect of using waste heat on the CO_2 avoidance costs at SMR facilities have not yet been studied.

7.2 Objective and research questions

The main objective of this thesis is to determine the techno-economic potential of commercial scale gasification systems producing energy carriers and other chemicals with low CO_2 emissions and to assess how and when flexibility improves the overall performance of these gasification systems.

The three main research questions are:

RQ. I What is the impact of flexible operation of state-of-the-art gasification facilities on the technical and economic performance of integrated gasification polygeneration facilities?

RQ. II What is the potential improvement in technical and economic performance of integrated gasification facilities in the coming decades?

RQ. III What are the CO_2 avoidance costs of gasification facilities, including SMR, in the short to long term?

Table 7.1 gives an overview of the chapters of this thesis in which these research questions are addressed.

Table 7.1 Structure of the thesis.

	RQ. I	RQ. II	RQ. III
Chapter 2: *Performance of simulated flexible integrated gasification polygeneration facilities. Part A: A technical-energetic assessment*	●		
Chapter 3: *Performance of simulated flexible integrated gasification polygeneration facilities. Part B: Economic evaluation*	●		●
Chapter 4: *Technical and economic prospects of coal- and biomass-fired IG-PG facilities equipped with CCS over time*		●	●
Chapter 5: *Future technological and economic performance of IGCC and FT production facilities with and without CO_2 capture: combining component-based learning curve and bottom-up analysis*		●	●
Chapter 6: *Techno-economic assessment of CO_2 capture at steam methane reforming facilities using commercially available technology*			●

7.3 Methodology

In this thesis two types of gasification facilities are investigated: IG-PG facilities for solid feedstocks (Eucalyptus pellets (EP), torrefied wood pellets (TOPS) and coal) and SMR facilities for gaseous feedstocks (natural gas). To determine the technical performance of these facilities, two component-based flowsheet models were made using AspenPlus and Procede Process Simulator. The models use build-in, experimentally verified databases to determine amongst other the thermodynamics, physicochemical properties, mass transfer rates, and kinetics while satisfying vapour-liquid equilibrium (VLE) and mass and energy balances.

For the solid feedstocks a dry-fed, slagging entrained flow gasifier was selected as this type of gasifier can handle carbonaceous feedstocks ranging from raw biomass to anthracite coal. A steady-state component-based AspenPlus model of IG-PG facilities was made in a chemical process simulator consisting of, among others, an air separation unit (ASU), gasifier, scrubbers, water-gas shift reactors, acid gas removal, sulphur processors, turbines, chemical conversion reactors, distillation columns, heat exchangers and compressors. The model was calibrated using a combination of data from open literature, operational data from a commercial IGCC, and expert interviews.

Flexible operation was simulated by changing the feedstock or production of commodities and adjusting the efficiency, conversion rate and utility consumption of individual components based on data and descriptions in the open literature and expert interviews. As the kind of flexibility investigated in this thesis is not (yet) used in actual facilities, calibration using operational data was not possible. Important assumptions regarding flexibility were: keeping the gasifier at a constant capacity, resulting in optimal use of the most capital expensive components, and fixing the H_2:CO ratio after the acid gas removal, to operate the gas cleaning train as constant as possible. In our study, this ratio was subsequently adjusted for each conversion process by adding H_2 from the H_2-pressure swing adsorption (H_2-PSA) unit, needed to purify the H_2. In an optimised IG-PG facility the H_2:CO ratio will be fine-tuned to the desired ratio using the water-gas shift (WGS) reactor, eliminating the need for H_2 addition.

In this thesis, the potential improvement in the performance of IG-PG facilities is assessed by combining an engineering study of the expected performance of technologies based on technological information (bottom-up analysis) with a trend analysis of the potential performance of components as function of their current and projected installed capacity based on experiences in the past (top-down analysis), and comparing the results with the performance of a state-of-the-art (SOTA) IG-PG facility.

For the bottom-up analysis, an inventory of promising novel technologies was made. The investigated technologies include, among other, an ASU using ion transfer membranes, sorption enhanced water-gas shift (SEWGS), transport desulphurisation, a shock wave CO_2 compressor, solid oxide fuel cells (SOFC), and improvements in gas and steam turbines. For each technology the expected technical performance, investment costs and development status were estimated using literature sources and expert interviews. Next, process layouts were constructed for the short, mid and long term using technologies which are expected to be commercially available in the specific time period. These layouts were modelled in the previously mentioned AspenPlus model. Based on the simulation runs, projections of the technical and economic performance of IG-PG facilities for each time period and overviews of the contribution of individual technologies were obtained.

In the top-down analysis, experience curves about the performance of the main process components of an IG-PG facility were constructed and applied using literature sources and expert interviews. If no curves were available for a specific component, experience curves from analogue technologies were used. The experience curves were combined with estimates of the total installed capacity of IG-PG facilities over time, derived from scenario studies. The findings gave projections on how production costs may decrease over time. In this approach it is assumed that the present trend, obtained from experiences in the past, also holds for the development in the future. This assumption is plausible for technologies in the early stage of their deployment.[207]

For SMR facilities, the technical performance was determined based on data in the open literature, expert interviews and operational data from a commercial SMR facility. A rate-based model of the CO_2 capture at an SMR facility was made in the chemical component-based Procede Process Simulator which was specifically designed to model acid gas treating processes. By running the model with different flows and compositions of the syngas entering the CO_2 absorber, the effect of flexible operation of the CO_2 capture unit was investigated.

In this thesis, economic aspects of the investigated processes and facilities are calculated using the net present value (NPV) method. Capital costs were assessed using the factored estimation method. For each component, the capital costs were estimated using data from open literature and expert interviews. As these data correspond to a specific scale, they were adjusted to the scale investigated in our study using scaling factors. Also, an installation factor, representing e.g., infrastructure, overhead, engineering and contingencies, was added. Total investment costs were compiled by the summation of the individual scaled component costs, including the installation factor. The inherent uncertainty of this method is approximately 30%. Another important economic variable is feedstock prices, which were assumed to be 2.25 €/GJ for coal, 6.3 €/GJ for both Eucalyptus pellets (EP) and torrefied wood pellets (TOPS) and 6.9 €/GJ for natural gas in the base case. Due to the uncertain behaviour of biomass prices, a sensitivity analysis was performed in which the price for TOPS drops to 3 €/GJ in the long term. The CO_2 transport and storage costs were assumed to be 10 €/t CO_2.

7.4 Summary of the results

Chapter 2 addresses research question I by assessing the possibilities of substituting the feedstock or varying the production of commodities in large-scale SOTA IG-PG facilities and investigating the potential impact this kind of flexibility may have on the technical performance. For this purpose, a component-based AspenPlus model of an IG-PG facility equipped with CO_2 capture was developed. Flexible operation of the IG-PG facility was simulated in the model by varying the feedstock between biomass (EP or TOPS) and bituminous coal and the production of commodities between electricity, methanol, urea and Fischer-Tropsch liquids (FT-liquids). The results show that static coal-fired IG-PG facilities with CO_2 capture can have overall energetic efficiencies of about 40% when producing electricity and 51-60% when producing chemicals (e.g., transportation fuels). Using EP lowers this efficiency by 2-5%$_{pt}$ due to the lower heating value of the biomass. When using TOPS, the penalty to the overall conversion efficiency is reduced to 1-2%$_{pt}$. The lower efficiency is primarily a result of a reduced cold gas efficiency of the gasifier (the ratio of chemical energy in the syngas compared to the chemical energy in the feedstock) as a larger share of the biomass needs to be combusted to compensate for its lower heating value, which is mainly caused by the oxygen. This, however, also reduces the ASU demand by up to 20%, thereby limiting the penalty in the overall conversion efficiency.

The findings also indicate that feedstock flexibility of up to 50% on an energy basis and product flexibility of up to 60% is technically feasible. If the feedstock is changed from 100% coal to a 50/50 mixture of biomass and coal, high quality biomass and coal is required, especially concerning heating value and quantity and composition of the ash and halide content. At 50% substitution ratio, the drop in overall conversion efficiency due to part-load operation is around 1%$_{pt}$. The overall efficiency drop due to altering production between chemicals and electricity is also around 1%$_{pt}$. The restriction to keep the gasifier at fixed dimensions means that the amount of gas created in the gasifier must be remained constant. This, combined with the lower energy density of biomass compared to coal, implies that the energetic input into the gasifier is reduced by around 15% when the feedstock is changed from 100% coal to a 50/50 EP-coal mixture. Torrefaction of biomass to TOPS prior to gasification reduces this drop in energetic input to around 8%.

In flexible IG-PG facilities it was assumed that the syngas is shifted to a specific H_2:CO ratio, regardless of production. This limits the amount of carbon that can be extracted for

CCS to 56-65% of the total carbon input, depending on the feedstock used. It also means that when CCS is applied, a biomass fraction of 40% results in a CO_2 neutral production of commodities (excluding indirect GHG emissions).

In **Chapter 3** the economic impact of substituting feedstock or production in large-scale SOTA IG-PG facilities is analysed further, thereby addressing research questions I and III of this thesis. In our study the economic performance of static coal-fired IG-PG facilities was evaluated. With the previously mentioned economic assumptions, production costs were determined at 12 €/GJ for methanol, 13 €/GJ for FT-liquids, 18 €/GJ for electricity and 21 €/GJ for urea. At these prices, none of the products can be considered competitive looking at the market prices of early 2010. Coal-fired IG-PG facilities may become competitive if capital costs would decrease by 10-27% . The sensitivity analysis indicated that production costs are most sensitive to the capital costs, electricity price and feedstock price.

It was found that the use of TOPS instead of coal increases production costs by 7-12 €/GJ. To make TOPS an attractive alternative to coal requires that either CO_2 credit prices increase to 40 €/t CO_2, coal prices increase to 4 €/GJ, or TOPS prices drop to 4 €/GJ. The use of wood pellets (like EP) yield a larger gas flow with lower energy content, resulting in a large decrease in output. This will makes EP economically not viable as feedstock. The findings also indicate that when applying CCS, production costs may increase by 3-6 €/GJ. This increase can be allocated to CO_2 extraction (50%), compression (20%) and transport & storage (30%). As CO_2 is already separated when producing chemicals the penalty due to CO_2 capture is, in this case, significantly reduced.

In the last decade, considerable variation in the market prices of biomass pellets (5.7-7.3 €/GJ), coal (1.0-5.6 €/GJ), CO_2 credits (0-32 €/t CO_2), electricity (0-290 €/GJ), gasoline/diesel (3-32 €/GJ), methanol (6-23 €/GJ) and urea (6-53 €/GJ) have occurred. In our study, a strategy to exploit these variations in market prices has been identified. The strategy involves altering production between chemicals and electricity, while the type of feedstock is kept constant. To prevent the gas turbine from operating below a 40% load factor, in the flexible IG-PG facilities the reactors which convert the syngas into chemicals must be smaller than those of static facilities optimised for chemical production. A product flexible IG-PG facility has comparable capital costs as a static IG-PG facility. Although the gas and steam turbines in the flexible facility are larger and thus more expensive, this is compensated by the smaller chemical conversion reactors. Results from scenario runs, based on the historic fluctuation in market prices, indicate that if FT-liquids are produced, the flexible facilities are economically more attractive than static ones (ΔNPV = 138M€). If, however, methanol or urea is produced, the flexible facilities perform worse than the static facilities (ΔNPV = -53 and -159 M€ respectively). When using TOPS as feedstock similar trends were found, though with a smaller impact. The findings also show that the flexible facilities are less vulnerable to fluctuations in market prices as the variation in economic performance in the different simulation runs is around 40% lower compared to the static facilities. The sensitivity analysis also indicates that flexible IG-PG facilities are much less affected by large variations in feedstock or CO_2 credit prices.

To determine under what economic conditions a flexible IG-PG facility could be attractive, another concept exploiting market variations was investigated. In this concept chemical production is kept constant and electricity production is varied by switching the feedstock between coal and TOPS. The capital costs of a feedstock flexible IG-PG facility were found to be only around 0.5% higher compared to a static coal-fired IG-PG facility as only minor over-dimensioning of the pre-treatment and gas cleaning equipment is required. Which feedstock provides the largest profit was found to be a function of the market prices of coal, TOPS, CO_2 credits and electricity.

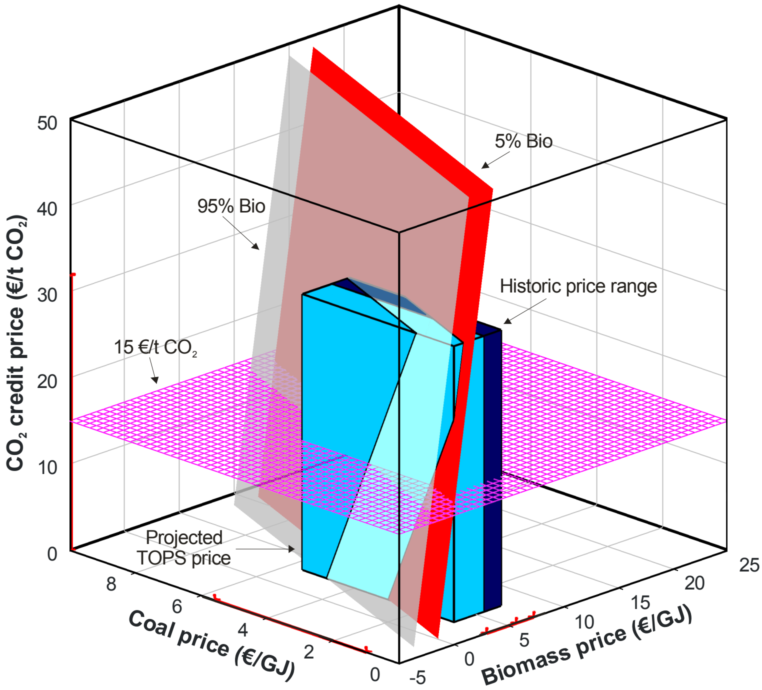


Figure 7.1 The effect of biomass pellets, coal and CO_2 credit price on feedstock preference. The horizontal plane illustrates a CO_2 credit price of 15 €/t CO_2. In front of the light (grey) shaded vertical plane biomass is the preferred feedstock for more than 95% of the time, while behind the dark (red) shaded plane biomass is preferred less than 5% of the time. Between the two planes, both biomass and coal are preferred. Note that the dark blue block shows the price ranges that have occurred. The light blue block illustrates a drop in biomass pellet price to 3.0 €/GJ.

At the market prices of early 2010, the most cost-effective feedstock (hereafter called preferred feedstock) turned out to be almost exclusively coal and a static IG-PG facility is the preferred technology. However, simulation runs also indicated that, due to the daily variation in electricity price, there is a price area of 2.2 €/GJ for coal, 1.9 €/GJ for TOPS and 22 €/t for the CO_2 credits where the preferred feedstock changes during the day (see Figure 7.1). It is expected that in the near future coal and CO_2 credit prices will increase, while biomass prices have the potential to decrease. This combination may result in market conditions corresponding to the area between the two diagonal planes shown in Figure 7.1. At these prices, optimal economic performance is achieved when coal is used during peak hours and biomass during off-peak hours, requiring a feedstock flexible IG-PG facility. If the rise in coal or CO_2 credit prices and/or the drop in TOPS prices continues, biomass will almost exclusively become the preferred feedstock and constructing a static IG-PG facility becomes (again) more interesting.

The effects of retrofitting an IG-PG facility producing FT-liquids were also assessed. The results indicate that converting an IG-FT facility towards methanol production may increase production costs of the methanol by < 3 €/GJ compared to an IG-MeOH facility as only a small part of the existing installation needs to be replaced. A change in the dominant transportation fuel from gasoline/diesel towards methanol could make this kind of retrofit attractive.

The results from this chapter indicate that when the uncertainty and variation in commodity prices are high, a small additional investment to make IG-PG facilities flexible in feedstock and production may make these facilities economically more attractive than static IG-PG facilities.

Chapter 4 addresses research questions II and III of this thesis by assessing the potential techno-economic improvement in the short to long term in overall performance of static IG-PG facilities producing electricity or transportation fuels from either coal or torrefied biomass using a bottom-up analysis. Contrary to Chapter 3, this analysis does not include a CO_2 credit price. Over a dozen promising novel technologies were identified as well as potential incremental improvements in seven conventional technologies. The identified novel technologies include solid feed pump, ASU using ion transport membranes (ITM), warm gas cleaning, transport desulphurisation, shock wave compression and solid oxide fuel cell (SOFC) and the conventional technologies include gas and steam turbines, solvent based acid gas removal and cryogenic ASU.

The results of our study indicate that for the short term only incremental improvements in the performance of cryogenic ASU, acid gas removal and gas and steam turbines are to be expected, resulting in an increase in overall conversion efficiency of $2\%_{pt}$ for IGCC and $1\%_{pt}$ for IG-FT. Capital costs are expected to remain constant. In the mid term the cryogenic ASU is expected to be replaced by an ion transfer membrane ASU, and the feedstock feeding system by a solid feed pump and the gas turbine is expected to have an additional incremental improvement. For IGCC facilities, the cold gas cleaning train is expected to be replaced by a more efficient warm gas cleaning train, including transport desulphurisation and the direct sulphur recovery process. If CCS is applied, the water-gas shift (WGS) reactor was found to be replaced by a sorption enhanced water-gas shift (SEWGS) reactor and the conventional CO_2 compressor by a shock wave compressor. In the IG-FT facility, the WGS reactor may be replaced by an advanced WGS which has reduced steam consumption and the energy consumption. The Rectisol-based acid gas removal is expected to have become even more efficient. Similar to the IGCC-CCS case, if CCS is applied, a shock wave compressor will be used. The overall conversion efficiency is expected to increase by $4\%_{pt}$ for IGCC and $2\text{-}3\%_{pt}$ for IG-FT compared to SOTA facilities. In the long term, the main improvements are expected to be the use of supercritical steam turbines and solid oxide fuel cells.

Our results also indicate that novel technologies and improved operating experience may reduce production costs of electricity at a coal-fired IGCC without CCS from the current 17 €/GJ to 11 €/GJ in the long term. These options may also reduce specific CO_2 emissions from 0.71 to 0.53 kg CO_2/kWh and increase net efficiencies from 44 to 59%. Applying CCS results in our study in production costs of electricity for a SOTA IGCC of 23 €/GJ with the potential to drop to 16 €/GJ in the long term. This would decrease CO_2 avoidance costs from 35 to 18 €/t CO_2. The efficiency penalty due to CCS remains around $9\%_{pt}$. In the long term, solid oxide fuel cells (SOFC's) are expected to become available. Although implementation of SOFC results in higher overall conversion efficiencies for IGCC, around 55% for both with and without CCS, the high capital costs makes SOFC only economically attractive in the case of IGCC-CCS, resulting in production costs of 14 €/GJ.

Current production costs of FT-liquids were determined at 13 €/GJ without CCS and 15 €/GJ with CCS. In the long term this could drop to 9 and 10 €/GJ respectively, resulting in CO_2 avoidance costs decreasing from 14 to 10 €/t CO_2 in the long term. The overall conversion efficiency was found to increase by $4\%_{pt}$ to 65% if the CO_2 is vented. Regardless of the time period, the energy penalty due to CCS remains around 2%pt as the CO_2 is already extracted and only needs compression. The findings also indicate that for an IGCC, the largest reductions in production costs are caused by the implementation of the ITM ASU and improved availability of the facility. For an IGCC-CCS better turbines, using warm gas cleaning and improved availability were found to have the largest potential to reduce the production costs.

CO_2 emissions can also be diminished by using torrefied biomass (TOPS) instead of coal. This would result in production costs of 25 €/GJ for electricity and 23 €/GJ for FT-liquids. Advanced technologies have the potential to reduce production costs to 19 and 18 €/GJ respectively. If biomass becomes available for 3 €/GJ, FT-liquids could be produced for 12 and 11 €/GJ respectively.

Finally, the combination of biomass and CCS can result in negative CO_2 emissions. As a result of technological learning, this option could have production costs of 15 €/GJ for electricity and 12 €/GJ for FT-liquids in the long term. Specific CO_2 emissions for electricity production are currently -0.93 kg CO_2/kWh and could change to -0.59 kg CO_2/kWh in the long term. This change is mainly caused by an increased energy conversion efficiency as this results in less CO_2 production (and thus less capture of CO_2) per kWh. In our study, the lowest CO_2 avoidance costs in the long term were obtained by biomass without CCS for electricity (6 €/t CO_2 avoided) and coal with CCS for FT-liquids (10 €/t CO_2).

In **Chapter 5** the assessment of the potential improvement of static IG-PG facilities was expanded by performing a top-down analysis. Progress ratios (PR) of each major component in IG-PG facilities were determined using data from open literature, expert interviews and PR values of analogue technologies. The growth in total installed capacity was based on the Baseline and Blue map scenarios of the International Energy Agency (IEA), as published in their Energy Technology Perspectives 2010. The Baseline scenario is characterised by almost no growth in CCS and IGCC and strong growth in FT. The Blue map is characterised by strong growth in CCS and IGCC and average growth in FT. Similar to Chapter 4, a CO_2 credit price was not included in the economic analysis.

When using the Blue map scenario and an average PR, our results indicate that production costs of electricity from coal could decrease from 17 to 12 €/GJ and efficiency could increase from 44 to 49% in the long term. Producing electricity with CCS might be achieved at 23 €/GJ in the short term and 15 €/GJ in the long term, resulting in CO_2 avoidance costs of 32 and 16 €/t CO_2 respectively. The efficiency could increase from 35 to 43%. Similar to the results from Chapter 4, SOFC was found to be only economically interesting for IGCC-CCS. The findings also indicate that the total installed IGCC capacity needs to increase significantly, reaching 600 GW_e by 2050 if these improvements are to be realised. Under current conditions, to make TOPS competitive with coal would require a CO_2 credit price of 47 €/t CO_2. In the long term, a combination of technological learning and reduction in biomass prices could make TOPS competitive with coal at a CO_2 credit price as low as 6 €/t CO_2.

FT-liquids can be produced from coal for 13.5 €/GJ, equivalent to a crude oil price of 77 €/bbl, and with an overall efficiency of 61%. Applying CCS may result in an overall conversion efficiency of 58% and a production costs increase to 14.6 €/GJ (equivalent to 83 €/bbl), corresponding to CO_2 avoidance costs of 11 €/t CO_2. In the long term, production costs of coal-based FT-liquids could drop to 8.5 €/GJ without CCS and 9.3 €/GJ with CCS and overall efficiency could increase to 64% and 62% respectively. Carbon neutral FT-liquids based on TOPS could be produced for 11 €/GJ in the long term. To realise these cost reductions, it is projected that the total installed FT-liquid capacity should increase from 28 $GW_{th\ FT}$ now to 430 $GW_{th\ FT}$ in the long term.

When examining which components drive the reduction in capital costs, the findings indicate that for an IGCC without CCS the largest reduction comes from the ASU, followed by the gasifier and the gas cleaning section. For an IGCC-CCS the reductions are mainly due to the CO_2 capture unit, ASU and gasifier. Capital costs of IG-FT decrease primarily due to the CO_2 capture unit, gasifier and ASU, regardless whether CCS is deployed.

Chapter 6 addresses research question III by identifying the optimal techno-economic configuration of a SOTA CO_2 capture system at a modern facility and with the aim of reducing the energy penalty associated with CO_2 capture and thereby minimising CO_2 avoidance costs. A step-by-step methodology was developed consisting of selecting the CO_2 capture technique; location of the CO_2 capture unit; CO_2 capture medium; configuration and optimisation of the CO_2 capture system. It was assumed that all required energy is supplied by a dedicated CHP unit.

Results indicate that only CO_2 capture by solvents is currently technically and economically viable for an industrial scale SMR facility. The optimal location of the CO_2 capture unit was determined to be between the water-gas shift (WGS) and H_2-pressure swing adsorption (PSA) units because of the high CO_2 partial pressure and the absence of combustion products. This location also has the added benefit of reducing the amount of CO_2 which needs to be captured in the H_2-PSA. Based on a qualitative comparison of different chemical, physical and mixed solvents, ADIP-X was deemed the optimal solvent. In this study the ADIP-X is a water-based solution containing 45 wt% MDEA and 5 wt% piperazine. An absorber-stripper configuration was selected because of its low capital costs, simplicity, robustness and the widely available knowledge of this configuration. The CO_2 capture system was optimised to allow regeneration using LP steam of 130°C. Implementing this system could reduce specific CO_2 emissions by 60% at 41 €/t CO_2 avoided, increasing H_2-production costs by 1.4 €/GJ. A large share of the CO_2 avoidance costs is due to the natural gas for the CHP and the CHP itself. Importing steam and electricity could reduce CO_2 avoidance costs to 24 €/t CO_2. Finally, the CO_2 capture system can handle variations in load. Simulation results indicate that at lower syngas flows the CO_2 capture solvent flow should be kept constant, while at higher syngas flows the solvent flow should be increased proportionally with the syngas flow.

7.5 Main findings and conclusions

The findings in Chapters 2 to 6 are used to formulate answers to the three research questions posed at the beginning of this thesis.

Research question I: *What is the impact of flexible operation of state-of-the-art gasification facilities on the technical and economic performance of integrated gasification polygeneration facilities?*

Conventional gasification configurations using SOTA technologies have the ability to use different feedstocks (coal, wood pellets (EP) and torrefied wood pellets (TOPS)) as well as to produce a varying product mix (electricity, FT-liquids, methanol, urea) without major modifications to the IG-PG facility. By using a dry-fed oxygen blown slagging entrained flow gasifier, biomass can replace coal by up to 50% on an energy basis. Without integration between the ASU and the gas turbine, chemical production can be reduced by 60% in favour of electricity production without significantly impacting the availability of the facility.

To allow such variations in feedstock use or production of commodities requires that certain components, e.g., gas cleaning train, chemical conversion reactors and turbines, are over-dimensioned. This means also that, depending on the used feedstock and desired production, these components operate at part-load conditions. Especially for the gas turbine this impacts its performance. If the gas turbine load is reduced by 60%, the efficiency of the gas turbine is reduced by almost 20%. The lower efficiency does mean that more heat is available for the steam turbine, thereby partly compensating for the efficiency loss in the gas turbine. Our studies indicate that the efficiency penalty due to variation in feedstock or in production is in both cases around $1\%_{pt}$. One of the main assumptions in this thesis is that the gasifier is not over-dimensioned. This is done to keep the load of the most expensive components of the IG-PG facility as high as possible. Consequently, when 100% coal is substituted by a 50/50 biomass/coal mixture, the energetic input drops by 15% when EP is used and 7.5% when TOPS is used. As the conversion efficiency remains almost constant, the output of the facility is also reduced.

To allow smooth adjustment of the production of an IG-PG facility, the H_2:CO ratio of the sweet syngas is fixed. This, however, decreases the amount of carbon (in the form of CO_2) that can be used for CCS. In the investigated IG-PG facilities only between 56-65% of the carbon can be extracted for CCS purposes.

Our results show that the ability to adjust feedstock or production in an IG-PG facility has a limited effect on plant economics, even though capital investment increases due to the over-dimensioning of equipment. A TOPS/coal flexible IG-PG facility has investment costs which are only around 0.5% higher than a static coal-fired IG-PG facility. The reduction in output has a much larger effect on the production costs. These drawbacks, however, can be compensated by exploiting the variations in market prices. A flexible IG-PG facility can also be attractive if markets change fundamentally, for instance due to the development of a biomass for energy market or due to a shift in dominant transportation fuel. Our findings indicate that the additional investment needed to allow adjustment of feedstock or production is cost-effective if the fluctuations in commodity prices are high.

Research question II: *What is the potential improvement in technical and economic performance of integration gasification facilities in the coming decades?*

There is a large potential to reduce production costs of IG-PG facilities due to novel technologies, increased operational experience and improvements to existing technologies. The largest expected effect on production costs results from gas turbines operating at higher pressure and temperature, replacement of the cryogenic ASU for an ionic transfer membrane ASU and higher availability. If CCS is applied, upgrading the cold gas cleanup to a warm gas cleanup also results in significant reductions in production costs.

Electricity production from a coal-fired IGCC facilities could be achieved at 40 €/MWh in the long term, compared to the current 60 €/MWh. Overall efficiency could increase from 44 to 52%. If CCS is applied, production costs could drop from 83 €/MWh to 51 €/MWh and the efficiency could increase from 35% to 55%. This would mean that, at the market prices of early 2010 of 56 €/MWh for electricity and 15 €/t CO_2 for CO_2 credits, coal-fired IGCC without CCS could become profitable in the short term and IGCC-CCS in the long term. Increases in the electricity or CO_2 credit price will shorten this time period.

Using 100% TOPS as feedstock increases the production costs to 92 €/MWh and 126 €/MWh, without and with CCS respectively. Besides the fact that production costs are much higher than the market prices, dry-fed slagging EF gasifiers can currently not handle 100% biomass streams. Production costs could decrease to 67 €/MWh for IGCC and 76 €/MWh for IGCC-CCS in the long term. If biomass prices drop from the current 6.3 €/GJ to 3.0 €/GJ, electricity from TOPS can be produced for 44 €/MWh without CCS and 55 €/MWh with CCS, which is only slightly more expensive then electricity production from coal.

Considerable cost reductions are also expected in the production of FT-liquids. Currently, converting coal into FT-liquids without CCS can be achieved at 13 €/GJ with an overall efficiency of 61%. In the long term, this might change to 9 €/GJ and 65%. Applying CCS results in production costs of 15 €/GJ now, with the potential to drop to 10 €/GJ in the long term. The efficiency penalty due to CCS is currently around 2$\%_{pt}$ and changes hardly in the future. Also with IG-FT, TOPS can be used as feedstock, which would result in a decrease in overall efficiencies of around 2$\%_{pt}$ compared to a coal-fired IG-FT. This penalty remains more or less identical, regardless of whether CCS is applied or not, in the investigated time periods. FT-liquids could be produced by a TOPS-fired SOTA IG-FT for 23 €/GJ without CCS and 24 €/GJ IG-FT with CCS. In the long term, a combination of technological learning and lower biomass feedstock costs could potentially reduce production costs to 11 €/GJ and 12 €/GJ. At these costs, bio-based FT-liquids would compete with crude oil-based gasoline/diesel at an oil price of 95-106 $/bbl (≈65-72 €/bbl).

An interesting alternative for the gas turbine for the long term is the solid oxide fuel cell (SOFC). For an IGCC without CCS, implementation of SOFC could increase the overall conversion efficiency by 7$\%_{pt}$. Production costs, however, are not reduced due to the projected high capital costs of SOFC. Implementation in an IGCC-CCS, however, increases overall efficiency by 13$\%_{pt}$. This increase does result in a reduction in production costs, despite the high capital costs of the SOFC. The main reason for the difference in efficiency gain is that SOFC works on H_2, eliminating the CCS penalty due to the WGS. Implementation of SOFC in an IGCC with CCS, but not in an IGCC without CCS, could result in the IGCC-SOFC-CCS facility having a higher conversion efficiency compared to the IGCC-Vent facility (55% vs. 52%), although production costs at the IGCC-Vent are considerably lower.

When comparing the results of the engineering study (bottom-up analysis) in Chapter 4 with those of the trend analysis using the learning curve approach (top-down analysis) in Chapter 5, similar projections regarding improvements in overall conversion efficiency and reductions in capital and production costs are obtained over time if average progress ratios and strong growth in IGCC, FT and CCS capacity are assumed. The break-down of the capital cost reduction shows minor differences between the two approaches, with the pre-treatment showing a smaller reduction and the ASU and gasifier showing larger reduction in capital costs in the top-down analysis compared to the bottom-up analysis.

Research question III: *What are the CO_2 avoidance costs of gasification facilities in the short to long term?*

Three CO_2 mitigation options have been investigated in this thesis, namely substituting coal for biomass, with TOPS being the most attractive, applying CCS, and the combination of the two. For each option the current mitigation costs as well as the potential development of these costs over time were investigated. Note that if CCS is applied, the costs associated with transport, storage and monitoring of the captured CO_2 were assumed to be 10 €/t CO_2.

In the short term, the lowest CO_2 avoidance costs were found by applying CCS. Results indicate that the implementation of CCS at a coal-fired IG-FT facility may result in mitigation costs of 14 €/t CO_2 avoidance. These costs may increase to 35 €/t CO_2 avoided for a coal-fired IGCC with CCS and 53 €/t CO_2 avoided if CCS is applied at a SMR facility producing H_2. The usage of waste heat can potentially lower the mitigation costs of the SMR facility to 34 €/t CO_2. Replacing coal with TOPS may increase mitigation costs to 53 and 45 €/t CO_2 avoided for IG-FT and IGCC respectively. The main reason for this increase is the high market price for biomass (6.3 €/GJ) compared to coal (2.25 €/GJ). The combination of using TOPS and applying CCS may result in mitigation costs of 38 €/t CO_2 avoided for IG-FT and 40 €/t CO_2 avoided for IGCC.

In the long term, the CO_2 avoidance costs at a coal-fired IG-FT facilities are calculated to change hardly as the CO_2 capture step only involves compression of the CO_2 and no learning in transport, storage and monitoring is assumed. At coal-fired IGCC facilities, however, the CO_2 avoidance costs could drop by 47% to 18 €/t CO_2. If CO_2 is mitigated by substituting coal for TOPS, the CO_2 avoidance costs are expected to remain almost constant at 45 €/t CO_2 over time. If, however, biomass prices drop to 3 €/GJ, mitigation costs would drop to 7 €/t CO_2 avoided, making TOPS-fired IGCC more attractive than coal-fired IGCC-CCS. The combination of technological learning in conversion processes and lower biomass price could reduce the CO_2 avoidance costs of a TOPS-fired IG-FT facility to almost the same level as a coal-fired IG-FT-CCS facility (12 €/t CO_2 vs. 10 €/t CO_2 respectively). The combination of TOPS and CCS could result in CO_2 avoidance costs of 12 €/t CO_2 for IGCC and 11 €/t CO_2 for IG-FT.

When comparing the results of SOTA gasification facilities with mitigation options at conventional SOTA power plants, it was found that a coal-fired IGCC-CCS has comparable CO_2 avoidance costs as a coal-fired PC-CCS. The SMR-CCS, however, has lower mitigation costs than a NGCC-CCS. If waste heat is available, the CO_2 avoidance costs could be more than 25% lower for the SMR-CCS.

7.6 Final remarks and recommendations for further research

Based on the results found in this thesis it can be concluded that there is a large potential for flexible gasification facilities as well as for gasification facilities applying CCS. This combination has the potential to allow gasification facilities to make a valuable contribution in the transition from a fossil-based energy infrastructure towards one based on renewables. Furthermore, as gasification is still a relative young technology, the potential for technological learning is considerable. However, achieving this learning requires R&D as well as the construction and operation of new gasification facilities.

To realise the full potential of gasification facilities and to help the transition from fossil fuels towards renewables while keeping human induced CO_2 emissions limited, the following roadmap can be envisioned. First, CCS can be implemented at gasification facilities which produce chemicals and at SMR facilities if waste heat is available. Second, IG-PG facilities can potentially act as large scale back-up power plants in a market with a high penetration of intermittent renewables, such as wind and solar power. The IG-PG facilities could also create a large market for energy crops.

When the transition from fossil fuel to renewables has accelerated, affecting market conditions of primary and secondary energy carriers, the flexibility of the gasification facilities can reduce investor risks as the facility can react relatively easy to these changes. The construction and operation of gasification facilities and the increasing implementation of CCS will drive technological learning, leading to lower production costs and lower CCS penalties. This, combined with potentially lower biomass prices caused by a more mature biomass-for-energy market, may result in the competitive production of carbon neutral, or even carbon negative, transportation fuels or electricity.

Based on the results found in the different chapters, the following recommendations for policy makers as well as topics for further research are identified.

Recommendations for policy makers:

- The ability of flexible IG-PG facilities to function as mid-load power plants and potentially even as peak-load power plants, means that integrating these facilities in the national power grid could result in added benefits. The extent of these benefits requires further research;
- Applying CCS at gasification facilities can be achieved at relative low costs, especially if chemicals are produced or if waste heat streams are available. By encouraging the deployment of CCS at these facilities, technological learning for CCS can be supported. A key pre-requirement in the deployment of CCS is a stable CO_2 credit market and/or the introduction of (strict) standards for the emission of CO_2 from industrial facilities as well as social acceptability of CO_2 storage and infrastructure development;
- The future development of sustainable biomass supplies and an effective biomass feedstock market is a key precondition for large scale use of biomass for energy (and chemical). This is as important for IG-PG facilities as for any other major biomass application;
- Upgrading biomass by torrefaction prior to injection into the gasifier significantly improves the technical and economic performance of the gasification facility. The technical feasibility of torrefaction of biomass at large-scale and its economic competitiveness still need to be demonstrated in the market;
- Niche applications for gasification facilities, such as in the conversion of municipal waste, should be identified and the research, development, demonstration and deployment (RDD&D) of gasification technology in these niches should be promoted to support future technology development and reduction of costs;
- Large potential improvements in the performance of gasification facilities have been identified. Realising these improvements requires that gasification capacity is increased considerable over time. The results of this thesis, however, also indicate that gasification facilities are not competitive in the short term. Therefore, it is recommended that policies which address the above mentioned points are implemented to accelerate the deployment of gasification facilities.

Recommendations for further research:

- The technical models developed and used in this research could not be calibrated using operational data from existing facilities, as there are currently no operational flexible IG-PG facilities nor is CO_2 captured at SMR facilities. Validation of the models is needed but requires measured data, e.g., from newly constructed demonstration plants;
- In this thesis, it is assumed that the PR values found for technological learning in an early stage of technology development also hold when further deployment is taking place. However, these values may change when mature markets for the implementation of the technology have been achieved. This needs further investigation;
- Comparing the economic results of different studies published in the scientific literature with each other is complicated as each study uses different capital costs. It is, therefore, recommended that a general database is constructed unifying capital costs data on a component basis;
- In a facility that can alter its feedstock or production, various components operate below their maximum capacity at least part of the time. The effect of prolonged part-load operation on the availability and efficiency of these components should be investigated and verified in practice;
- The flexible IG-PG facility is designed for production of multiple transport fuels, other chemicals and electricity. Optimising the facility for production of only one chemical (and electricity) can increase overall efficiency. If CCS is applied at the facility, optimisation could also increase the fraction of carbon (in the form of CO_2) that can be used for CCS. The potential effects of optimised flexible gasification facilities should be investigated further;
- Integrating the ASU with the gas turbine could increase overall efficiency by several percent points. In the IG-PG facility modelled in this thesis, however, the ASU and gas turbine were not integrated as this integration poses limitations of the load variations of the gas turbine. It is recommended that, when the understanding of this integration increases, it is investigated whether the integration between ASU and gas turbine can be realised without limiting the ability of IG-PG facilities to alter feedstock use or production of commodities;
- In this thesis, technical flowsheet models were used to calculate the main mass and energy streams. Trace elements, such as particulate matter, halogen gas, heavy metals and NO_x were not investigated into detail. Especially with flexible operation of the gas turbine, NO_x formation, but also the production and emissions of other contaminants, should be investigated in more detail to ensure that the flexible facility complies with environmental legislation;
- The bottom-up analysis in this thesis indicates that new technologies like syngas turbines, H_2 membranes, ITM and SOFC have to become commercially available to achieve the projected production costs. This requires additional research and development to solve some critical problems identified with these technologies;
- For determining the potential improvement in performance of gasification facilities in the future, in this thesis estimates were made of the expected technical performance and costs of technologies which have not yet been tested on a commercial scale, such as membrane separation and SOFC. This results in

uncertainties in the projections. It is advised that up-to-date analyses of the techno-economic performance of gasification facilities are repeated in the future. This should result in more accurate results;

- Although the techno-economic performance of 100% biomass feeding has been assessed, the possibilities and effects of co-feeding large fractions of biomass (up to 100%) have not yet been tested in EF gasifiers. Pilot scale experiments are desired to test the maximum biomass co-feeding fraction and to develop operating techniques to increase this fraction;
- The analysis of applying CCS at an SMR facility involves a retrofit of an existing facility. CO_2 avoidance costs are expected to decrease if the CO_2 capture process is an integral part of the design of a SMR facility. The impact of such design on the techno-economic performance of SMR-CCS should be investigated.

8 Samenvatting, conclusies en aanbevelingen

Achtergrond

Onze samenleving heeft energiebronnen nodig die overvloedig, betaalbaar, betrouwbaar en milieuvriendelijk zijn. Gelijk met de toename van de wereldbevolking en het stijgen van de gemiddelde levensstandaard is de wereldwijde vraag naar energie gestegen.[1,2,3,4,5] De International Energy Agency (IEA) gaat er in het rapport Energy Technology Perspectives 2010 vanuit dat de wereldwijde vraag naar energie stijgt van circa 500 EJ in 2009[1,2,3] tot meer dan 750 EJ in 2035.[4] Van deze vraag zou meer dan twee derde geleverd moeten worden door fossiele brandstoffen.

Het grootschalig gebruik van fossiele brandstoffen kan aanzienlijke nadelen hebben, waaronder de uitstoot van broeikasgassen (greenhouse gases: GHG).[2,4,5,9,10,11] In 2009 werd wereldwijd 29 Gt CO_2 uitgestoten door de verbranding van fossiele brandstoffen. Bijna 85% van deze emissies kwam van de sectoren elektriciteitsproductie (11,8 Gt CO_2), transport (6,5 Gt CO_2) en industrie (5,9 Gt CO_2).[20]

Recentelijk kwam de Conference of the Parties (COP) van de United Nations Framework Convention on Climate Change (UNFCCC) overeen dat, om gevaarlijke klimaatverandering te voorkomen, de gemiddelde temperatuur op aarde met niet meer dan 2°C mag stijgen boven het pre-industriële niveau.[13,14,15] Voor CO_2 betekent dit zeer waarschijnlijk dat de emissies in 2050 met 50-80% moeten zijn teruggebracht ten opzichte van de huidige uitstoot.[11,15,18] Dit is aanzienlijk. Om zulke reducties te realiseren is waarschijnlijk een portfolio van mitigatiemaatregelen nodig, waaronder het verbeteren van de efficiëntie van energieconversies, het overschakelen naar energiebronnen die minder (fossiel) koolstofintensief zijn, en het toepassen van afvangst, transport en opslag van koolstofdioxide (carbon dioxide capture, transport and storage: CCS).[2,4,5,6,7]

Er is een aanzienlijk potentieel om zowel biomassa op een moderne wijze als energiebron te gebruiken als CCS toe te passen om aldus de uitstoot van broeikasgassen te beperken. De huidige toepassing van deze opties bevindt zich echter op een laag niveau. Dit geldt zeker voor CCS.

Toepassing van vergassingstechnologie kan een belangrijke rol spelen in het bevorderen van de ontwikkeling van beide mitigatieopties. Vergassingsinstallaties kunnen naar verwachting tegen relatief lage kosten biomassa omzetten in waardevolle (eind)producten. In vergassingsinstallaties waar chemicaliën geproduceerd worden, wordt CO_2 vaak al afgescheiden. Het afvangen en ondergronds opslaan van deze CO_2 kan de kosten om de uitstoot van CO_2 te beperken aanzienlijk verminderen.

Het geïnstalleerde vermogen van vergassingsinstallaties dat wereldwijd is opgesteld is op dit moment laag. Dit komt vooral door de hoge investeringskosten van deze installaties en de onzekerheid in de prijzen van zowel grondstoffen als eindproducten op de wereldmarkt.[42,60,61,62,63,64,65,66] De productiekosten van de eindproducten kunnen door technologisch leren echter significant dalen. Door technologisch leren kunnen zowel de investeringskosten dalen als de conversierendementen verbeteren, op zowel de korte als langere termijn.[58,59] Het realiseren van deze verbeteringen vereist verdere investeringen in onderzoek en ontwikkeling (R&D) en een significante toename van de geïnstalleerde vergassingscapaciteit.

De mogelijkheid om in het productieproces verschillende grondstoffen te gebruiken en/of makkelijk een ander eindproduct te maken, kan ervoor zorgen dat vergassingstechnologie op korte termijn meer toegepast gaat worden. Daarvoor kunnen twee redenen worden genoemd. De eerste reden is dat de geïntegreerde wijze waarop zowel de vergassing van meerdere grondstoffen als de productie van meerdere eindproducten wordt gerealiseerd (integrated gasification poly-generation: IG-PG) het mogelijk maakt om grootschalige IG-PG installaties te bouwen die (gedeeltelijk) op biomassa draaien. Dit leidt tot schaalvoordelen en een lagere CO_2-uitstoot, maar vermindert ook de kans op langdurige storingen in het productieproces door mogelijke ontregelingen in de aanvoer van grondstoffen zoals biomassa. Ten tweede heeft de installatie de mogelijkheid om snel te reageren op fluctuaties in de marktprijs van grondstoffen en eindproducten. Dit kan tot verkleining van het risico voor investeerders leiden. Het is echter gewenst meer inzicht te krijgen in de mogelijkheden en beperkingen van vergroting van de flexibiliteit van vergassingsinstallaties. Ook is een vraag bij welke (economische) condities flexibele installaties winstgevender zijn dan hun statische tegenhangers. Een diepgaande analyse van deze aspecten vormt de kern van dit proefschrift.

De literatuur geeft aan dat de prestaties van IG-PG installaties in de toekomst kunnen verbeteren. Echter, op welke manier die verbetering tot stand komt en hoe snel dit zal gaan is op dit moment niet erg duidelijk. In dit proefschrift wordt getracht een beoordeling te geven van de potentiële verbeteringen in de prestaties van vergassingsinstallaties. Dit is gedaan door een zogenaamde bottom-up analyse en top-down analyse uit te voeren en de resultaten van beide analyses te vergelijken.

Een veelbelovende industriële toepassing van vergassing in combinatie met CCS op korte termijn is omzetting van aardgas tot syngas met behulp van stoom (steam methane reforming: SMR). De kosten om de uitstoot van CO_2 te vermijden zijn bij SMR-installaties naar verwachting lager dan bij elektriciteitscentrales. Dit komt door de aanwezigheid van een CO_2-rijke gasstroom op hoge druk en de mogelijke beschikbaarheid van restwarmte doordat SMR-installaties vaak onderdeel zijn van grote industriële complexen, zoals raffinaderijen. De invloed op de CO_2 vermijdingskosten van het toepassen van CCS bij bestaande vergassingsinstallaties (retrofit) en van het gebruik van restwarmte hierbij is bij ons weten in de literatuur nog niet geanalyseerd.

Doel en onderzoeksvragen

Het eerste hoofddoel van dit proefschrift is te onderzoeken wat het technisch-economische potentieel is van grootschalige vergassingsinstallaties die energiedragers en andere chemicaliën kunnen produceren met weinig CO_2-uitstoot. Het tweede hoofddoel is te bepalen onder welke omstandigheden en op welke manier flexibiliteit de prestaties van deze vergassingsinstallaties kan verbeteren.

Deze hoofddoelen hebben geleid tot de volgende drie onderzoeksvragen:

OV. I Welke invloed heeft het flexibel bedrijven van moderne vergassingsinstallaties op de technische en economische prestaties van IG-PG installaties?

OV. II Wat zijn de potentiële verbeteringen in de technische en economische prestaties van geïntegreerde vergassingsinstallaties in de komende decennia?

OV. III Wat zijn de CO_2 vermijdingskosten bij vergassingsinstallaties, inclusief SMR, op korte tot lange termijn?

Tabel 8.1 geeft een overzicht van de hoofdstukken in dit proefschrift en laat zien waar in deze hoofdstukken elk van de drie onderzoeksvragen (OV) wordt behandeld.

Tabel 8.1 Opbouw van het proefschrift.

	OV. I	OV. II	OV. III
Hoofdstuk 2: *Performance of simulated flexible integrated gasification polygeneration facilities. Part A: A technical-energetic assessment*	●		
Hoofdstuk 3: *Performance of simulated flexible integrated gasification polygeneration facilities. Part B: Economic evaluation*	●		●
Hoofdstuk 4: *Technical and economic prospects of coal- and biomass-fired IG-PG facilities equipped with CCS over time*		●	●
Hoofdstuk 5: *Future technological and economic performance of IGCC and FT production facilities with and without CO_2 capture: combining component-based learning curve and bottom-up analysis*		●	●
Hoofdstuk 6: *Techno-economic assessment of CO_2 capture at steam methane reforming facilities using commercially available technology*			●

Methodologie

In dit proefschrift worden twee verschillende typen vergassingsinstallaties onderzocht: IG-PG installaties voor de verwerking van vaste grondstoffen - te weten Eucalyptus pellets (EP), getorreficeerde houtpellets (TOPS) en kolen - en SMR-installaties voor de verwerking van gasvormige grondstoffen - te weten aardgas. Om de technische prestaties van deze installaties te bepalen zijn op componentniveau twee flowsheet modellen gemaakt gebruik makend van respectievelijk AspenPlus en de Process Simulator van Procede. De modellen gebruiken interne, experimenteel geverifieerde databestanden voor het bepalen van onder andere: de thermodynamica, fysisch-chemische eigenschappen, massa-overdrachtsnelheden en kinetiek met aandacht voor het gas-vloeistof evenwicht (VLE) en de massa- en energiebalansen.

Voor de verwerking van vaste grondstoffen wordt in dit proefschrift een z.g. slagging entrained flow gasifier met droge voeding gebruikt, omdat dit type vergasser koolstofhoudende grondstoffen variërend van biomassa tot antraciet kan verwerken. In AspenPlus is op componentniveau een evenwichtsmodel van de IG-PG installaties gemaakt met onder andere een luchtscheider (air separation unit: ASU), vergasser, gaswassers, water-gas shift (WGS) reactoren, verwijderaars van zuurgas, zwavelreactoren, turbines, chemische conversiereactoren, destillatiekolommen, warmtewisselaars en compressoren. Het model is gekalibreerd op basis van literatuurgegevens, gegevens van een operationele commerciële vergasser die is geïntegreerd met een stoom- en gasturbine voor het opwekken van elektriciteit (integrated gasification combined cycle: IGCC) en interviews met experts.

Het flexibel opereren van de IG-PG installatie is gesimuleerd door de voeding en/of productie te veranderen en bij elke afzonderlijk component de efficiëntie, conversie en het verbruik van gas, water en elektra aan te passen op basis van publicaties in de literatuur en interviews met experts. Omdat het type flexibiliteit dat in dit proefschrift onderzocht is (nog) niet wordt toegepast in commerciële installaties, was kalibratie met data van operationele installaties niet mogelijk.

Belangrijke aannames met betrekking tot het variëren van voeding en/of productie zijn een constante capaciteit van de vergasser, omdat hierdoor de kapitaalintensiefste componenten optimaal gebruikt worden, en een constante H_2:CO verhouding na de zuurgasverwijderaar teneinde het functioneren van de gasschoonmaak apparatuur op een zo constant mogelijk niveau te houden. In dit proefschrift wordt vervolgens de H_2:CO verhouding in ieder van de verschillende productieprocessen aangepast door toevoeging van H_2 afkomstig van de z.g. H_2-pressure swing adsorption reactor (H_2-PSA), waarin de H_2 wordt gezuiverd. In een geoptimaliseerde IG-PG installatie kan de H_2:CO ratio aangepast worden door de WGS reactor; toevoeging van extra H_2 is dan niet meer nodig.

In dit proefschrift zijn de potentiële verbeteringen in de prestaties van IG-PG installaties onderzocht door het uitvoeren van een engineeringstudie van de te verwachten verbeteringen van technologieën op basis van gedetailleerde technische informatie (bottom-up analyse) en een trendanalyse van de mogelijke prestaties van de componenten als functie van hun huidige en toekomstige geïnstalleerd vermogen gebaseerd op ervaringen uit het verleden (top-down analyse). Vervolgens zijn de resultaten van beide analyses gecombineerd en vergeleken met de huidige prestaties van IG-PG installaties.

Voor de bottom-up analyse is eerst een overzicht gemaakt van veelbelovende technologieën, waaronder een luchtscheider met ion transfer membranen (ITM), een sorption enhanced water-gas shift (SEWGS) reactor, een transportontzwavelingreactor, een schokgolfcompressor, een solid oxide fuel cell (SOFC), en verbeteringen in gas- en stoomturbines. Voor elke technologie zijn de verwachte technische prestaties, investeringskosten en ontwikkelingsfase geschat op basis van literatuurgegevens en interviews met experts. Vervolgens zijn proces lay-outs gemaakt voor de korte, middellange en lange termijn waarbij alleen die technologieën zijn gebruikt waarvan verwacht wordt dat ze op deze tijdschaal commercieel beschikbaar komen.

De lay-outs zijn gemodelleerd in het hiervoor genoemde AspenPlus model. Op basis van simulatieruns zijn voor elke tijdsperiode projecties gemaakt van de technische en economische prestaties van IG-PG installaties. Door dit onderzoek is ook inzicht in de individuele bijdrage van ieder van de technologieën verkregen.

Voor de top-down analyse zijn experience curves (ervaringscurven) van de belangrijkste componenten van een IG-PG installatie gemaakt op basis van literatuurgegevens en interviews met experts. Als er te weinig gegevens beschikbaar waren om op deze wijze een experience curve te maken, werd gebruik gemaakt van data van vergelijkbare componenten. De experience curves zijn vervolgens gecombineerd met schattingen over de geïnstalleerde capaciteit van IG-PG installaties in de toekomst, op basis van scenariostudies. Dit heeft geleid tot projecties hoe de productiekosten in de toekomst zouden kunnen dalen.

In deze analyse is aangenomen dat de huidige kostenontwikkeling, verkregen door ervaringen in het verleden, zich doorzet in de toekomst. Deze aanname is aannemelijk voor technologieën die in een vroeg stadium van hun ontwikkeling zitten.[207]

Voor SMR-installaties zijn de technische prestaties gebaseerd op literatuurgegevens, data van een operationele commerciële SMR-installatie, en interviews met experts. Een kinetisch model op componentniveau van het CO_2-afvangproces bij een SMR-installatie is gemaakt in de Procede Process Simulator. Deze simulator is specifiek voor het modelleren van processen waarin zure gassen zoals CO_2 worden verwijderd. Door het uitvoeren van meerdere simulatieruns met verschillende syngasdebieten en -composities is het effect van het flexibel opereren van de reactor voor het afvangen van CO_2 geanalyseerd.

In dit proefschrift zijn de economische aspecten van de onderzochte processen en installaties berekend met behulp van de netto contante waarde (net present value: NPV) methode. Kapitaalkosten zijn geschat met behulp van de z.g. factored estimation method. In deze methode worden de kapitaalkosten van ieder van de componenten van een installatie bepaald op basis van literatuurgegevens en interviews met experts. Deze kostendata zijn gerelateerd aan een bepaalde capaciteit. Waar nodig zijn deze data gecorrigeerd voor de capaciteit waarmee in dit proefschrift is gerekend met behulp van een schalingsfactor. Voor het berekenen van de totale kapitaalkosten is vervolgens een installatiefactor toegevoegd, voor onder andere infrastructuur, overhead, engineering en onvoorziene kosten. Aldus zijn de totale investeringskosten geschat. De inherente onzekerheid van deze methode is ongeveer 30%.

Een andere belangrijke economische variabele zijn de grondstofprijzen. In het basismodel zijn deze gesteld op 2,25 €/GJ voor kolen, 6,3 €/GJ voor Eucalyptus pellets (EP) en getorreficeerde houtpellets (TOPS) en 6,9 €/GJ voor aardgas. Gezien de onzekerheid in de biomassaprijzen is hiervoor een gevoeligheidsanalyse uitgevoerd waarin de TOPS prijs op lange termijn zakt tot 3 €/GJ. De CO_2 transport- en opslagkosten zijn gesteld op 10 €/t CO_2.

Samenvatting van de resultaten

Hoofdstuk 2 behandelt onderzoeksvraag I door middel van een analyse van de mogelijkheden van het vervangen van de voeding en het variëren van de productie in grootschalige, best beschikbare techniek (BBT) IG-PG installaties. Ook wordt de mogelijke invloed van deze flexibiliteit op de technische prestaties bestudeerd. Hiervoor is een op componenten gebaseerd AspenPlus model ontwikkeld van een IG-PG installatie uitgerust met CO_2-afvangapparatuur. Het flexibel bedrijven van de IG-PG installatie is gesimuleerd door de voeding te variëren tussen biomassa (EP of TOPS) en bitumineuze kolen, en de productie te variëren tussen elektriciteit, methanol, ureum en Fischer-Tropsch vloeistoffen (FT-vloeistoffen). De resultaten laten zien dat statische kolengestookte IG-PG installaties met CO_2-afvang een energetisch rendement kunnen hebben van ongeveer 40% bij de productie van elektriciteit en van 51-60% wanneer chemicaliën (bv. transportbrandstoffen) geproduceerd worden. Het gebruik van EP verlaagt deze rendementen met 2-5%$_{pt}$ als gevolg van de lagere energiedichtheid van de biomassa ten opzichte van kolen. Bij gebruik van TOPS is de daling van het energetisch rendement slechts 1-2%$_{pt}$. De lagere rendementen zijn voornamelijk het resultaat van een afname van de koudgasefficiëntie van de vergasser (de verhouding tussen de chemische energie van het syngas en de chemische energie van de voeding). De afname wordt veroorzaakt doordat bij biomassa een groter gedeelte van de voeding wordt verbrand ter compensatie van de lagere energiedichtheid. Deze lagere energiedichtheid wordt voornamelijk veroorzaakt doordat de biomassa een hoger gehalte aan zuurstofatomen heeft. Anderzijds vermindert dit de zuurstofvraag van de luchtscheider met mogelijk 20% wat de reductie in het energetisch rendement gedeeltelijk compenseert.

De resultaten duiden erop dat flexibiliteit in de voeding tot 50% (op energiebasis) technische mogelijk is en in het maken van producten tot 60%. Bij een verandering van de voeding van 100% kolen naar een 50/50 mengsel van biomassa en kolen is het noodzakelijk dat de kolen van goede kwaliteit zijn. Dit betreft voornamelijk de energiedichtheid en de concentratie en samenstelling van de mineralen en halogenen. Bij een substitutie van de voeding van 50% is de vermindering van het energetisch rendement door het opereren in deellast ongeveer 1%$_{pt}$. De terugval in rendement bij het veranderen van de productie van chemicaliën naar elektriciteit is ook ongeveer 1%$_{pt}$.

De aanname dat de capaciteit van de vergasser niet verandert, heeft tot gevolg dat de hoeveelheid gas die in de vergasser gevormd wordt constant blijft. Dit, gecombineerd met de lagere energiedichtheid van biomassa ten opzichte van kolen, zorgt ervoor dat de energetische invoer in de vergasser daalt met ongeveer 15% als de voeding verandert van 100% kolen naar een 50/50 mengsel van EP en kolen. Het gevormde gasmengsel bevat ook minder CO en H_2 en meer CO_2. Torrefactie van de biomassa naar TOPS vermindert de afname in energetische invoer naar ongeveer 8%.

Voor de flexibel opererende IG-PG installaties is aangenomen dat de H_2:CO verhouding van het syngas wordt aangepast naar een specifieke waarde, ongeacht de productie. Dit vermindert de hoeveelheid CO_2 die afgevangen kan worden voor CCS naar, afhankelijk van de gebruikte grondstof, 56-65% van de totale koolstofinvoer. Bij deze aanname is een biomassafractie van 40% nodig om, in combinatie met CCS, koolstofneutraal te produceren (exclusief indirecte broeikasgasemissies).

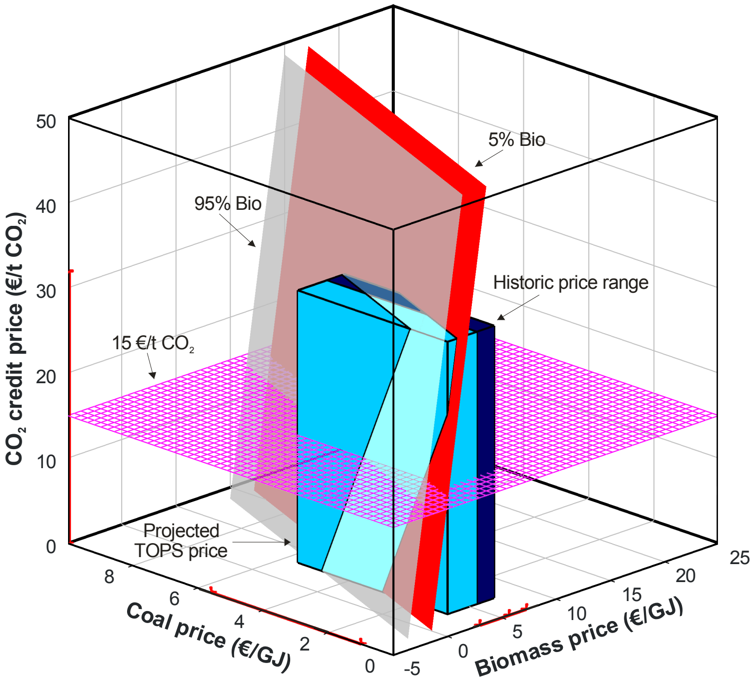
In **Hoofdstuk 3** worden de economische aspecten van flexibiliteit in zowel voeding als productie van grootschalige IG-PG installaties verder geanalyseerd. Hiermee behandelt dit hoofdstuk onderzoeksvragen I en III van dit proefschrift. In het onderzoek zijn de economische prestaties van statische, kolengestookte IG-PG installaties geëvalueerd. Bij de hiervoor genoemde economische aannames zijn de productiekosten berekend op 12 €/GJ voor methanol, 13 €/GJ voor FT-vloeistoffen, 18 €/GJ voor elektriciteit en 21 €/GJ voor ureum. Deze productiekosten zijn alle vier boven de marktprijs van begin 2010. De kolengestookte IG-PG installaties zouden concurrerend kunnen worden als de kapitaalkosten dalen met 10-27%. De gevoeligheidsanalyse laat zien dat de productiekosten het meest gevoelig zijn voor wijzigingen in de kapitaalkosten, de elektriciteitsprijs en de grondstoffenprijs.

Als gebruik wordt gemaakt van TOPS zouden de productiekosten met 7-12 €/GJ kunnen stijgen. TOPS wordt concurrerend met kolen als de CO_2 prijs stijgt tot 40 €/t CO_2, de kolenprijs stijgt tot 4 €/GJ of de TOPS-prijs daalt tot 4 €/GJ. Het gebruik van houtpellets (zoals EP) zorgt voor een groter gasvolume met een lage energiedichtheid wat leidt tot een sterke daling in de totale productie. Dit maakt EP economisch niet geschikt als grondstof.

De resultaten laten ook zien dat de productiekosten kunnen stijgen met 3-6 €/GJ door het toepassen van CCS. Deze toename komt door de CO_2-afvang (50%), compressie van CO_2 (20%) en het transport & opslag van CO_2 (30%). Doordat bij de productie van chemicaliën CO_2 al wordt afgevangen is de toename van de productiekosten door het toepassen van CCS in dit geval significant lager.

In het afgelopen decennium zijn er aanzienlijke variaties geweest in de (momentane) marktprijs van biomassapellets (5,7-7,3 €/GJ), kolen (1,0-5,6 €/GJ), CO_2 (0-32 €/t CO_2), elektriciteit (0-290 €/GJ), benzine/diesel (3-32 €/GJ), methanol (6-23 €/GJ) en ureum (6-53 €/GJ). In dit proefschrift is onderzocht hoe van deze fluctuaties op strategische wijze gebruik kan worden gemaakt. De strategie die is onderzocht betreft het veranderen van de productie tussen elektriciteit en chemicaliën, terwijl de voeding hetzelfde blijft. Om te voorkomen dat de gasturbine onder 40% deellast komt, zijn de reactoren die het syngas omzetten in de chemicaliën kleiner in de flexibele IG-PG installaties dan in de installaties die geoptimaliseerd zijn voor chemicaliënproductie. Een productieflexibele IG-PG installatie heeft vergelijkbare kapitaalkosten als een statische IG-PG installatie. De grotere, en dus duurdere, gas- en stoomturbines in de flexibele installatie worden gecompenseerd door de kleinere chemische reactoren. De resultaten van de scenario's, gebaseerd op de historische fluctuaties in de marktprijzen, duiden erop dat flexibele installaties die FT-vloeistoffen produceren economisch beter presteren dan de statische installaties (ΔNPV = 138M€). Als methanol of ureum geproduceerd wordt, presteren de flexibele installaties juist slechter (respectievelijk ΔNPV = -53 and -159 M€). Als TOPS de voeding is worden dezelfde trends gevonden, maar met een kleiner effect. De resultaten laten ook zien dat flexibele installaties minder gevoelig zijn voor fluctuaties in de marktprijzen. Het verschil in de economische prestaties tussen de verschillende simulatieruns is ongeveer 40% lager dan in het geval van de statische installaties. De gevoeligheidsanalyse laat zien dat de flexibele IG-PG installaties minder gevoelig zijn voor fluctuaties in grondstof- of CO_2-prijzen.

Om te bepalen onder welke economische condities een flexibele IG-PG installatie aantrekkelijk is, is een ander concept geanalyseerd dat ook gebruik maakt van fluctuaties in de marktprijzen. In dit concept wordt de chemische productie constant gehouden, maar wordt de elektriciteitsproductie aangepast als de voeding varieert tussen kolen en TOPS. De kapitaalkosten van een voedingflexibele IG-PG installatie zijn ongeveer 0.5% hoger dan die van een statische kolengestookte IG-PG installatie, omdat alleen een kleine overdimensionering van de componenten in de grondstofvoorbehandeling en de gasschoonmaak nodig is. Welke grondstof de meeste winst oplevert is een functie van de marktprijs van kolen, TOPS, CO_2 en elektriciteit.



Figuur 8.1 Het effect van de marktprijs van biomassa pellets, kolen en CO_2 op de preferente grondstof. Het horizontale vlak illustreert een CO_2 prijs van 15 €/t CO_2. Voor het lichtgrijze verticale vlak is biomassa de preferente grondstof voor meer dan 95% van de tijd. Achter het donkerrode verticale vlak is biomassa de preferente grondstof voor minder dan 5% van de tijd. Tussen de twee vlakken zijn zowel biomassa als kolen de preferente grondstof. Het donkerblauwe blok laat de historische extremen van de marktprijzen zien. Het lichtblauwe blok illustreert een daling in de prijs van biomassapellets tot 3,0 €/GJ.

Onder de marktcondities van begin 2010 is de meest kosteneffectieve grondstof (hierna preferente grondstof genoemd) bijna volledig kolen en een statische IG-PG installatie is de preferente technologie. Echter, de simulatieruns laten ook zien dat door de dagelijkse variatie in elektriciteitsprijs er een prijsgebied is ter grootte van 2,2 €/GJ voor kolen, 1,9 €/GJ voor TOPS en 22 €/t CO_2 waarin de preferente grondstof verandert tijdens de dag (zie Figuur 8.1). In de nabije toekomst worden hogere prijzen voor kolen en de uitstoot van CO_2 verwacht terwijl de biomassaprijzen zouden kunnen dalen. Deze combinatie kan resulteren in marktcondities die overeenkomen met het gebied tussen de twee diagonale vlakken in Figuur 8.1. Bij deze marktprijzen is het voor optimale economische prestaties nodig dat kolen worden gebruikt tijdens de piekuren en biomassapellets tijdens de daluren. Hiervoor is een flexibele IG-PG installatie vereist. Als de prijsstijging voor kolen en CO_2 doorzet en/of de TOPS-prijs blijft dalen, zal biomassa bijna volledig de preferente grondstof worden en is een statische IG-PG installatie economisch (weer) het meest interessant.

Het effect van het retrofitten van een IG-PG installatie die FT-brandstoffen produceert is ook geanalyseerd. De resultaten laten zien dat door het ombouwen van een IG-FT installatie naar methanolproductie de toename in de productiekosten van de methanol, vergeleken met een IG-MeOH installatie, onder de 3 €/GJ kan zijn. Deze toename is laag. Dit komt doordat maar een klein gedeelte van de oorspronkelijke installatie vervangen hoeft te worden. Een verschuiving in de dominante transportbrandstof van benzine/diesel naar methanol kan dit type retrofit aantrekkelijk maken.

De resultaten van dit hoofdstuk laten zien dat wanneer de onzekerheid en variatie in marktprijzen van grondstoffen en eindproducten groot zijn, met een kleine additionele investering om een IG-PG installatie flexibel te maken in zowel voeding als productie, deze installatie economisch aantrekkelijker kan worden dan een statische IG-PG installatie.

Hoofdstuk 4 behandelt onderzoeksvragen II en III van dit proefschrift door middel van een beoordeling van de potentiële technische en economische verbeteringen op korte tot lange termijn van de prestaties van IG-PG installaties die elektriciteit of transportbrandstoffen maken van kolen of getorreficeerde biomassa. De beoordeling berust op een bottom-up analyse. In tegenstelling tot Hoofdstuk 3 bevat deze analyse geen CO_2 prijs. Meer dan een dozijn veelbelovende nieuwe technologieën zijn geanalyseerd evenals potentiële verdere verbeteringen in zeven conventionele technologieën. Tot de nieuwe technologieën behoren een vaste voedingspomp, een luchtscheider op basis van het gebruik van z.g. ion transport membranes (ITM), warme gasschoonmaak, transportontzwaveling, schokgolfcompressie en de solid oxide fuel cell (SOFC). De conventionele technologieën omvatten onder andere de gas- en stoomturbines, zuurgasverwijderaar op basis van een oplosmiddel, en cryogene luchtscheider.

De resultaten van de analyse wijzen erop dat op korte termijn alleen verbeteringen in de cryogene luchtscheider, zuurgasverwijderaar en gas- en stoomturbines verwacht mogen worden. Deze verbeteringen kunnen zorgen voor een toename in het energetisch rendement van $2\%_{pt}$ voor IGCC en $1\%_{pt}$ voor IG-FT. De kapitaalkosten blijven waarschijnlijk gelijk.

Op middellange termijn is de verwachting dat de cryogene luchtscheider vervangen wordt door de ITM luchtscheider en dat het voedingsysteem van de vergasser vervangen wordt door de vaste voedingspomp. Ook worden er verdere verbeteringen van de gasturbines verwacht. Voor IGCC wordt de koude gasschoonmaak vervangen door de efficiëntere warme gasschoonmaak. Dit omvat onder ander transportontzwaveling en het toepassen van een proces om zwavel direct te herwinnen (direct sulphur recovery process).

Als CCS toegepast wordt, wordt de WGS reactor vervangen door de SEWGS en de conventionele CO_2 compressor door de schokgolfcompressor. In de IG-FT installatie wordt de WGS reactor vervangen door een geavanceerdere WGS reactor, welke een verminderd stoom- en energieverbruik heeft. De verwachting is dat de zuurgasverwijderaar op basis van Rectisol efficiënter wordt. Bij toepassing van CCS zal, net als bij de IGCC-CCS, schokgolfcompressie gebruikt worden. Door deze verbeteringen kan het energetisch rendement waarschijnlijk toenemen met $4\%_{pt}$ voor IGCC en $2\text{-}3\%_{pt}$ voor IG-FT vergeleken met de huidige installaties. Op lange termijn is de verwachting dat gebruik kan worden gemaakt van superkritische stoomturbines en van solid oxide fuel cells.

De resultaten laten zien dat door nieuwe technologieën en betere ervaring met het bedrijven van deze installaties verwacht kan worden dat de productiekosten van elektriciteit van een kolengestookte IGCC zonder CCS kunnen dalen van de huidige 17 €/GJ naar 11 €/GJ op lange termijn (bij gelijk blijven van de economische aannames). Tegelijkertijd zouden de specifieke CO_2 emissies dan dalen van 0,71 naar 0,53 $kgCO_2/kWh$ en de efficiëntie stijgen van 44 naar 59%. Bij toepassing van CCS bedragen de productiekosten van een BBT IGCC 23 €/GJ, welke zouden kunnen dalen naar 16 €/GJ op lange termijn. Hierdoor zouden de CO_2-afvangkosten dalen van 35 naar 18 €/t CO_2. De afname in het rendement door het toepassen van CCS blijft in alle onderzochte tijdsperiodes rond de $9\%_{pt}$. Het gebruik van SOFC wordt waarschijnlijk pas commercieel op lange termijn. Ondanks het feit dat het gebruik van SOFC resulteert in een hoger energetisch rendement van de IGCC - namelijk rond de 55%, zowel met als zonder CCS - zorgen de hoge kapitaalkosten ervoor dat SOFC alleen economisch aantrekkelijk is voor een IGCC-CCS. Deze IGCC-SOFC-CCS zou elektriciteit kunnen produceren voor 14 €/GJ.

De huidige productiekosten van FT-vloeistoffen zijn berekend op 13 €/GJ zonder CCS en 15 €/GJ met CCS. Op de lange termijn zouden deze kosten kunnen dalen naar respectievelijk 9 en 10 €/GJ, wat zou resulteren in een daling in de CO_2-afvangkosten van 14 naar 10 €/t CO_2. Het energetisch rendement voor de IG-FT zonder CCS kan op lange termijn toenemen met $4\%_{pt}$ tot 65%. De efficiëntiepenalty door het toepassen van CCS blijft in alle onderzochte tijdperiodes ongeveer $2\%_{pt}$. Dit komt voornamelijk doordat in een IG-FT installatie CO_2 al standaard wordt afgevangen en daardoor alleen nog compressie van de CO_2 nodig is indien CCS toegepast gaat worden. De resultaten laten ook zien dat voor een IGCC zonder CCS de grootste besparing in productiekosten komt door het gebruiken van de ITM luchtscheider en de verwachte hogere beschikbaarheid van de installatie. Voor een IGCC-CCS komt de daling in productiekosten voornamelijk door betere turbines, het toepassen van warme gasschoonmaak en een hogere beschikbaarheid.

CO_2 emissies kunnen ook gereduceerd worden door getorreficeerde biomassa (TOPS) in plaats van kolen te gebruiken. Dit zou in huidige installaties resulteren in productiekosten van 25 €/GJ voor elektriciteit en 23 €/GJ voor FT-vloeistoffen. Geavanceerde technieken en technologieën hebben de mogelijkheid deze kosten omlaag te brengen naar respectievelijk 19 en 18 €/GJ. Als biomassa beschikbaar komt voor 3 €/GJ dalen de productiekosten nog verder naar respectievelijk 12 en 11 €/GJ.

Ten slotte moet worden opgemerkt dat de combinatie biomassa en CCS kan resulteren in negatieve CO_2 emissies. Door technologisch leren zouden de productiekosten bij deze optie op lange termijn kunnen dalen naar 15 €/GJ voor elektriciteit en 12 €/GJ voor FT-vloeistoffen. De specifieke CO_2 emissie voor elektriciteitsproductie zou dan veranderen van -0,93 $kgCO_2/kWh$ naar -0,59 $kgCO_2/kWh$. Deze verandering komt voornamelijk door een hoger energetisch rendement wat resulteert in minder CO_2 productie (en dus minder CO_2-afvang) per kWh.

In deze analyse worden de laagste CO_2-afvangkosten voor de lange termijn gevonden bij gebruik van biomassa voor het maken van elektriciteit zonder CCS (6 €/t CO_2 vermeden) en bij gebruik van kolen voor het maken van FT-vloeistoffen met CCS (10 €/t CO_2 vermeden).

In **Hoofdstuk 5** wordt de beoordeling van de potentiële verbeteringen van statische IG-PG installaties voortgezet door middel van een top-down analyse. Hierbij zijn zogenaamde Progress ratios (PRs) bepaald van de ervaringscurven die gelden voor de hoofdcomponenten. Dit is gedaan aan de hand van literatuurgegevens, interviews met experts en het gebruik van PR-waarden van analoge technologieën.

De verwachte groei van het geïnstalleerd vermogen is in deze analyse gebaseerd op de Baseline en Blue map scenario's van de International Energy Agency (IEA). Deze scenario's zijn gepubliceerd in de Energy Technology Perspectives 2010. In het Baseline scenario vindt vrijwel geen groei plaats in de toepassing van zowel CCS als IGCC en een grote groei in het toepassen van IG-FT. Het Blue map scenario kent een sterke groei in de toepassing van zowel CCS als IGCC en een gematigde groei in de toepassing van IG-FT. Net zo als in Hoofdstuk 4 zit in de economische analyse van dit hoofdstuk geen prijs voor de uitstoot van CO_2.

De resultaten laten voor het Blue map scenario en bij een gemiddelde waarde van de PRs zien dat de productiekosten van elektriciteit uit kolen kunnen dalen van 17 €/GJ naar 12 €/GJ op de lange termijn. Het energetisch rendement neemt hierbij toe van 44 naar 49%. Als CCS wordt toegepast zouden de productiekosten van elektriciteit kunnen dalen van 23 nu naar 15 €/GJ op de lange termijn. Dit zou dan resulteren in CO_2-afvangkosten van respectievelijk 32 en 16 €/t CO_2, terwijl het energetisch rendement toeneemt van 35 naar 43%. Gelijk aan de resultaten gevonden in Hoofdstuk 4, is SOFC waarschijnlijk alleen economisch interessant voor IGCC-CCS. De resultaten laten ook zien dat, om de verbeteringen door technologisch leren mogelijk te maken, de geïnstalleerde capaciteit van IGCC significant moet toenemen naar bijna 600 GW_e op lange termijn. Om TOPS concurrerend te maken onder de huidige condities is een CO_2 prijs van 47 €/t CO_2 nodig.

Op lange termijn kan de combinatie van technologisch leren en lagere biomassaprijzen TOPS al concurrerend met kolen maken bij een CO_2 prijs van maar 6 €/t CO_2.

De productie van FT-vloeistoffen op basis van kolen is op dit moment mogelijk voor 13,5 €/GJ (equivalent aan een aardolieprijs van 77 € per vat) en met een energetisch rendement van 61%. Als CCS toegepast wordt daalt dit rendement naar 58% en stijgen de productiekosten naar 14,6 €/GJ (equivalent aan 83 € per vat). Dit betekent dat de CO_2-afvangkosten 11 €/t CO_2 zijn. Op lange termijn zouden de productiekosten kunnen dalen naar 8,5 €/GJ zonder CCS en 9,3 €/GJ met CCS. Het energetisch rendement zou dan respectievelijk 64% en 62% zijn. De productiekosten van CO_2-neutrale FT-vloeistoffen op basis van TOPS zouden kunnen dalen tot 11 €/GJ op lange termijn. Voor deze verbeteringen zou de geïnstalleerde capaciteit van FT-vloeistofinstallaties moeten toenemen van 28 $GW_{th\ FT}$ naar 430 $GW_{th\ FT}$.

De resultaten laten ook zien dat de grootste afname in kapitaalkosten bij een IGCC zonder CCS veroorzaakt wordt door verbeteringen van de luchtscheider, gevolgd door de vergasser en de gasschoonmaak. Bij een IGCC-CCS zijn het de CO_2-afvanginstallatie, luchtscheider en vergasser en bij IG-FT installaties voornamelijk de CO_2-afvanginstallatie, vergasser en luchtscheider, ongeacht of CCS wordt toegepast of niet.

Hoofdstuk 6 behandelt onderzoeksvraag III door te bepalen wat de optimale technisch-economische configuratie is voor een BBT CO_2-afvangsysteem bij een moderne SMR-installatie. Het doel hierbij is de energieconsumptie van de CO_2-afvang te beperken om aldus de CO_2-afvangkosten te minimaliseren. Een stappenplan is ontwikkeld welke bestaat uit het selecteren van respectievelijk de CO_2-afvangtechniek; de locatie van de CO_2-afvang; het CO_2-afvang medium; en de configuratie en optimalisatie van de CO_2-afvanginstallatie. In de analyse is aangenomen dat alle benodigde energie wordt geleverd door een specifieke warmtekrachtkoppelinginstallatie (WKK).

De resultaten duiden erop dat voor een industriële SMR-installatie op dit moment alleen CO_2-afvang met behulp van een oplosmiddel technisch en economisch haalbaar is. De optimale locatie van de CO_2-afvanginstallatie is tussen de WGS en H_2-PSA reactoren vanwege de aanwezigheid van een hoge partiële druk van CO_2 en de afwezigheid van verbrandingsproducten. Deze locatie heeft als extra voordeel dat er minder CO_2 in de H_2-PSA afgevangen hoeft te worden.

Verschillende chemische, fysische en gemengde oplosmiddelen zijn kwalitatief beoordeeld en ADIP-X kwam hierbij als beste oplosmiddel te voorschijn. In deze analyse is aangenomen dat ADIP-X een waterige oplossing is van 45 $\%_{massa}$ MDEA en 5 $\%_{massa}$ piperazine. Als configuratie werd voor een absorber-stripper gekozen vanwege zijn lage kapitaalkosten, eenvoud, betrouwbaarheid en de ruime beschikbaarheid van operationele kennis. De CO_2-afvanginstallatie is geoptimaliseerd om regeneratie met lage druk stoom van 130°C mogelijk te maken.

Door dit systeem toe te passen zou de specifieke CO_2 emissie van de SMR-installatie met 60% kunnen dalen tegen CO_2 vermijdingskosten van 41 €/t CO_2. Dit zou de H_2-prijs verhogen met 1,4 €/GJ. Een groot gedeelte van de CO_2-afvangkosten wordt veroorzaakt door het benodigde aardgas voor de WKK en door de kapitaalkosten van de WKK. Door het importeren van stoom en elektriciteit zouden de CO_2-afvangkosten kunnen dalen tot 24 €/t CO_2. Ten slotte moet worden opgemerkt dat het geselecteerde CO_2-afvangsysteem variaties in het syngas aan kan. Simulatieruns laten zien dat bij een verminderd syngasdebiet het debiet van het oplosmiddel van de CO_2-afvang constant moet worden gehouden, terwijl bij een hoger syngasdebiet het debiet van het oplosmiddel proportioneel mee moet schalen met het syngasdebiet.

Hoofdbevindingen en conclusies

De bevindingen in Hoofdstuk 2-6 zijn gebruikt om een antwoord te geven op de drie onderzoeksvragen die aan het begin van dit proefschrift geformuleerd zijn.

Onderzoeksvraag I: *Welke invloed heeft het flexibel bedrijven van moderne vergassingsinstallaties op de technische en economische prestaties van IG-PG installaties?*

Conventionele vergassingsconfiguraties die gebruik maken van BBT technologieën kunnen zowel hun voeding variëren (kolen, houtpellets (EP) en getorreficeerde houtpellets (TOPS)) als hun productie (elektriciteit, FT-vloeistoffen, methanol en ureum) zonder grote aanpassingen aan de IG-PG installatie. In een slagging entrained flow gasifier met droge voeding kan biomassa kolen vervangen tot 50% (op energiebasis). Zonder integratie tussen de luchtscheider en de gasturbine kan de productie van chemicaliën met 60% worden gereduceerd ten gunste van elektriciteitproductie zonder daarbij een significant effect op de beschikbaarheid van de installatie te hebben.

Om zulke variaties in voeding en productie mogelijk te maken is het noodzakelijk dat bepaalde componenten, zoals de gasschoonmaak, chemische conversiereactoren en turbines, overgedimensioneerd zijn. Dit betekent dat, afhankelijk van de voeding en productie, deze componenten op deellast opereren. Dit beïnvloedt de prestaties van die componenten, met name die van de gasturbine. Als de gasturbine op 60% deellast opereert daalt het omzettingsrendement van de gasturbine met bijna 20%. Dit lagere rendement heeft wel tot gevolg dat er meer warmte beschikbaar is voor de stoomturbine, waardoor een gedeelte van het lagere rendement van de gasturbine gecompenseerd wordt. De resultaten in dit proefschrift duiden erop dat de vermindering in energetisch rendement in IG-PG installaties door het variëren van de voeding of de productie in beide gevallen ongeveer $1\%_{pt}$ is.

Een van de hoofdaannames in dit proefschrift is dat de vergasser niet overgedimensioneerd is. Dit is gedaan om de deellast van de meest kapitaalintensieve componenten van de IG-PG installatie zo hoog mogelijk te houden. Een gevolg is dat als 100% kolen wordt vervangen door een 50/50 mengsel van biomassa en kolen de energetische invoer daalt met 15% als EP wordt gebruikt en met 7,5% als TOPS wordt gebruikt. Doordat het energetisch rendement bijna gelijk blijft, heeft dit een daling in de productie van de installatie tot gevolg.

Om een soepele aanpassing in de productie van een IG-PG installatie mogelijk te maken is de H_2:CO verhouding van het syngas na de zuurgasverwijdering constant gehouden. Dit vermindert echter de hoeveelheid koolstof die afgevangen kan worden (in de vorm van CO_2). In de onderzochte IG-PG installatie kan maar 56-65% van de koolstofinvoer door middel van CCS worden afgevangen.

De resultaten laten zien dat de mogelijkheid om de voeding of productie in een IG-PG installatie te veranderen een beperkt effect heeft op de economische prestaties van de installatie, ondanks het feit dat de kapitaalinvesteringen toenemen door het overdimensioneren van de componenten. Een TOPS/kolen flexibele IG-PG installatie heeft kapitaalkosten die slechts ongeveer 0,5% hoger zijn dan die van een statische kolengestookte IG-PG installatie. Een veel groter effect op de productiekosten heeft de verminderde productie. Deze nadelen kunnen echter gecompenseerd worden door gebruik te maken van fluctuatie in de marktprijzen van de grondstoffen en eindproducten. Een flexibele IG-PG installatie is aantrekkelijk als de markten fundamenteel veranderen, bijvoorbeeld door de ontwikkeling van een biomassamarkt voor energiedoeleinde of een verandering in de dominante transportbrandstof. Onze resultaten duiden erop dat de additionele investering die nodig is om variatie in voeding en productie mogelijk te maken kosteneffectief is indien de fluctuatie in de prijzen van de grondstoffen en eindproducten hoog is.

Onderzoeksvraag II: *Wat zijn de potentiële verbeteringen in de technische en economische prestaties van geïntegreerde vergassingsinstallaties in de komende decennia?*

Er is een groot potentieel om de productiekosten van IG-PG installaties te verminderen door gebruik te maken van nieuwe technologieën, vergroting van de operationele ervaring en verbeteringen in bestaande technologieën. Het grootste effect op de productiekosten wordt bereikt door het toepassen van gasturbines die op een hogere temperatuur opereren, het vervangen van de cryogene luchtscheider door een luchtscheider met ITM en het vergroten van de beschikbaarheid van de installatie. Als CCS wordt toegepast kan het vervangen van de koude gasschoonmaak door een warme gasschoonmaak ook een significante verlaging van de productiekosten tot gevolg hebben.

De productie van elektriciteit in een kolengestookte IGCC installatie zou op lange termijn mogelijk kunnen zijn voor 40 €/MWh vergeleken met de huidige prijs van 60 €/MWh. Het energetisch rendement zou dan stijgen van 44 naar 52%. Als CCS wordt toegepast zouden de productiekosten op lange termijn kunnen dalen van 83 naar 51 €/MWh en het energetisch rendement kunnen stijgen van 35 naar 55%. Dit betekent dat bij de marktprijzen van begin 2010 - te weten 56 €/MWh en 15 €/t CO_2 - een kolengestookte IGCC zonder CCS economisch rendabel wordt op de korte termijn en een IGCC met CCS op de lange termijn. Een toename van de elektriciteitprijs of de CO_2 prijs zal deze tijdsperiode verkorten.

Het gebruik van 100% TOPS als voeding verhoogt de productie naar 92 €/MWh zonder toepassing van CCS en naar 126 €/MWh met CCS. Naast het feit dat deze productiekosten veel hoger zijn dan de huidige marktprijs, is een slagging entrained flow gasifier met droge voeding op dit moment niet geschikt voor voeding met 100% biomassa. Productiekosten zouden op lange termijn kunnen dalen naar 67 €/MWh voor een IGCC zonder CCS en naar 76 €/MWh voor een IGCC met CCS. Als biomassa beschikbaar komt voor 3,0 €/GJ in plaats van de huidige 6,3 €/GJ dalen de productiekosten nog verder, naar respectievelijk 44 en 55 €/MWh. Deze waarden zijn maar weinig hoger dan de productiekosten van elektriciteit uit kolen.

Aanzienlijke kostenbesparingen worden ook verwacht voor de productie van FT-vloeistoffen. Op dit moment kunnen kolen worden omgezet in FT-vloeistoffen voor 13 €/GJ en met een energetisch rendement van 61% mits CCS niet wordt toegepast. Op lange termijn kunnen deze waarden veranderen naar 9 €/GJ en 65%. Het toepassen van CCS resulteert in huidige productiekosten van 15 €/GJ met een potentiële verlaging naar 10 €/GJ op lange termijn. Het lagere energetisch rendement door het toepassen van CCS is op dit moment ongeveer $2\%_{pt}$. Dit verandert in de toekomst nauwelijks.

Ook bij IG-FT installaties kan TOPS als voeding gebruikt worden. Het energetisch rendement is dan ongeveer $2\%_{pt}$ lager dan bij een kolengestookte IG-FT installatie. Dit verschil tussen TOPS en kolen blijft praktisch hetzelfde, ongeacht of CCS wordt toegepast en naar welke tijdsperiode gekeken wordt. De productiekosten voor een TOPS gestookte BBT IG-FT installatie zijn 23 €/GJ zonder CCS en 24 €/GJ met CCS. Op lange termijn kan een combinatie van technologisch leren en lagere biomassaprijzen de productiekosten laten dalen tot respectievelijk 11 en 12 €/GJ. Bij deze productiekosten is een TOPS gestookte IG-FT installatie concurrerend met op ruwe olie gebaseerde productie van benzine/diesel indien de olieprijs 95-106 $ per vat (≈ 65-72 € per vat) bedraagt.

Op lange termijn is een interessant alternatief voor de gasturbine de solid oxide fuel cell (SOFC). Voor een IGCC zonder CCS zou de implementatie van SOFC het energetisch rendement met $7\%_{pt}$ kunnen doen toenemen. De productiekosten zouden echter niet omlaag gaan door de verwachte hoge kapitaalkosten van SOFC. Bij toepassing van SOFC in een IGCC-CCS installatie zou het energetisch rendement van de installatie zelfs met $13\%_{pt}$ kunnen toenemen. Deze toename zal wel voor lagere productiekosten kunnen zorgen, ondanks de hoge kapitaalkosten van de SOFC. De belangrijkste reden voor het verschil in energetisch rendement is dat de SOFC H_2 nodig heeft. Dit elimineert de verlaging van het energetisch rendement veroorzaakt door de WGS als CCS wordt toegepast. Het gebruik van SOFC in een IGCC-CCS, maar niet in een IGCC zonder CCS (IGCC-Vent) betekent dat het energetisch rendement van een IGCC-SOFC-CCS installatie (55%) hoger is dan die van de IGCC-Vent (52%). Daar staat tegenover dat de productiekosten bij de IGCC-Vent aanzienlijk lager zijn.

Het vergelijken van de resultaten van de engineeringstudie (bottom-up analyse) van Hoofdstuk 4 met de trendanalyse op basis van ervaringscurven (top-down analyse) van Hoofdstuk 5 laat vergelijkbare projecties zien met betrekking tot verbeteringen in energetisch rendement, besparingen in kapitaalkosten en verwachte afname van de productiekosten, mits gemiddelde waarden voor de progress ratio's (PRs) en een sterke groei in het toepassen van IGCC, FT en CCS worden aangenomen. Een gedetailleerd overzicht van de besparingen in kapitaalkosten laat kleine verschillende zien tussen de twee aanpakken. Vergeleken met de bottom-up analyse laat de top-down analyse een kleinere besparing in kapitaalkosten zien voor de voedingvoorbehandeling en een grotere besparing voor de luchtscheider en vergasser.

Onderzoeksvraag III: *Wat zijn de CO_2 vermijdingskosten bij vergassingsinstallaties, inclusief SMR, op korte tot lange termijn?*

Drie opties voor het verminderen of vermijden van de uitstoot van CO_2 zijn in dit proefschrift onderzocht, namelijk het vervangen van kolen door biomassa, waarbij TOPS het aantrekkelijkst is, het toepassen van CCS bij het gebruik van steenkool of aardgas en het toepassen van biomassa gecombineerd met CCS. Voor elke optie zijn zowel de huidige als de toekomstige vermijdingskosten geanalyseerd. Een belangrijke aanname hierbij is dat, als CCS toegepast wordt, de transport- en opslagkosten van CO_2 gesteld kunnen worden op 10 €/t CO_2.

Op korte termijn resulteert het toepassen van CCS bij gebruik van kolen in de laagste CO_2 vermijdingskosten. De resultaten duiden erop dat het toepassen van CCS bij een kolengestookte IG-FT installatie mogelijk is tegen CO_2 vermijdingskosten van 14 €/t CO_2. Deze kosten nemen toe tot 35 €/t CO_2 voor een kolengestookte IGCC met CCS en tot 53 €/t CO_2 als CCS wordt toegepast bij een SMR-installatie voor H_2-productie. Bij gebruik van restwarmte kunnen de vermijdingskosten bij SMR-installatie dalen naar 34 €/t CO_2.

Het vervangen van kolen door TOPS zorgt voor vermijdingskosten van 53 en 45 €/t CO_2 voor respectievelijk IG-FT en IGCC. De belangrijkste reden voor deze toename in de vermijdingskosten is de relatief hoge biomassaprijs (6,3 €/GJ) ten opzichte van de kolenprijs (2,25 €/GJ). De combinatie biomassa en CCS kan resulteren in vermijdingskosten van 38 €/t CO_2 voor IG-FT en 40 €/t CO_2 voor IGCC.

Op lange termijn veranderen de CO_2 vermijdingskosten van een kolengestookte IG-FT installatie nauwelijks, aangezien de CO_2-afvangstap alleen de compressie van de CO_2 inhoudt en er aangenomen is dat er geen kostenverbeteringen optreden bij het transport, opslag en monitoren van de CO_2. Bij kolengestookte IGCC kunnen de CO_2 vermijdingskosten echter dalen met 47% naar 18 €/t CO_2. Als CO_2 vermeden wordt door het vervangen van kolen door TOPS blijven de CO_2 vermijdingskosten ongeveer 45 €/t CO_2, ongeacht de tijdsperiode. Als echter de biomassaprijzen dalen naar 3 €/GJ zouden de mitigatiekosten dalen naar 7 €/GJ. Dit betekent dat een TOPS gestookte IGCC aantrekkelijker wordt dan een kolengestookte IGCC-CCS. Bij de combinatie van technologisch leren en een lagere biomassaprijs kunnen de CO_2 vermijdingskosten van een TOPS gestookte IG-FT installatie op bijna hetzelfde niveau komen als een kolengestookte IG-FT-CCS (respectievelijk 12 €/t CO_2 versus 10 €/t CO_2). De combinatie TOPS en CCS resulteert dan in CO_2 vermijdingskosten van 12 €/t CO_2 voor IGCC en 11 €/t CO_2 voor IG-FT.

Wanneer de resultaten van de CO_2 vermijdingsopties van BBT vergassingsinstallaties worden vergeleken met die van conventionele BBT elektriciteitscentrales wordt duidelijk dat een kolengestookte IGCC-CCS vergelijkbare CO_2 vermijdingskosten heeft als een kolengestookte poederkoolcentrale met CCS. De SMR-CCS heeft echter lagere vermijdingskosten dan een aardgascentrale met CCS. Als restwarmte beschikbaar is zijn de CO_2 vermijdingskosten zelfs meer dan 25% lager bij de SMR-CCS-installatie.

Slotopmerkingen en aanbevelingen voor verder onderzoek

Op basis van de resultaten van dit proefschrift kan geconcludeerd worden dat er een groot potentieel is voor flexibele vergassingsinstallaties en voor vergassingsinstallaties uitgerust met CCS. Deze combinatie zou ervoor kunnen zorgen dat vergassingsinstallaties een waardevolle bijdrage kunnen leveren aan de transitie van een op fossiele energie gebaseerde energie-infrastructuur naar een die gebaseerd is op hernieuwbare energie. Bovendien, omdat vergassing nog steeds een relatief nieuwe technologie is, zijn er nog aanzienlijk verbeteringen door technologisch leren te verwachten. Hiervoor is echter wel R&D en het bouwen van nieuwe vergassingsinstallaties noodzakelijk.

Een stappenplan is bedacht om het volledige potentieel van vergassingsinstallaties te realiseren en om de transitie van fossiele brandstoffen naar hernieuwbare energiedragers te bevorderen terwijl tegelijkertijd de antropogene CO_2 emissies worden beperkt. Eerst kan CCS geïmplementeerd worden bij vergassingsinstallaties die chemicaliën produceren en bij SMR-installaties indien restwarmte aanwezig is. Vervolgens kunnen IG-PG installaties fungeren als grootschalige back-up elektriciteitscentrales in een energiemarkt met een hoge penetratiegraad van hernieuwbare energiebronnen, zoals wind- en zonne-energie. Deze IG-PG installaties kunnen tegelijkertijd een grootschalige markt voor energiegewassen creëren.

Wanneer de transitie van fossiele brandstoffen naar hernieuwbare energiedragers versnelt, en daardoor de marktcondities van primaire en secundaire energiedragers verandert, kan de flexibiliteit van vergassingsinstallaties het risico voor investeerders verminderen doordat de installaties relatief gemakkelijk kunnen reageren op die veranderingen. Het bouwen en bedrijven van vergassingsinstallaties en de toenemende implementatie van CCS zal het technologisch leren versnellen, wat zal leiden tot lagere productiekosten en lagere kosten met betrekking tot CCS. Dit, in combinatie met mogelijk lagere biomassaprijzen door een meer volwassen ontwikkeling van het gebruik van biomassa op de energiemarkt, kan resulteren in een concurrerende productie van koolstofneutrale of zelfs koolstofnegatieve transportbrandstoffen en elektriciteit.

Op basis van de resultaten van de verschillende hoofdstukken worden de volgende aanbevelingen voor beleidsmakers gedaan en suggesties gegeven voor verder onderzoek:

Aanbevelingen voor beleidsmakers

- De mogelijkheid om flexibele IG-PG installaties te gebruiken als middenlast elektriciteitscentrales en mogelijk zelfs als pieklast centrales betekent dat de integratie van deze installaties in het nationale elektriciteitsnet additionele voordelen kan hebben. Het bepalen van de omvang van deze voordelen vereist verder onderzoek.
- CCS kan bij vergassingsinstallaties voor relatief lage kosten worden toegepast, in het bijzonder als chemicaliën geproduceerd worden of als restwarmte

beschikbaar is. Door toepassing van CCS bij deze installaties te promoten kan het technologisch leren van CCS worden bevorderd. Belangrijke voorwaarden voor de implementatie van CCS is een stabiele CO_2 markt en/of de introductie van (strikte) standaarden voor de uitstoot van CO_2 van industriële installaties. Daarnaast is ook maatschappelijke steun voor de opslag van CO_2 en de ontwikkeling van een CO_2 infrastructuur van belang.

- De ontwikkeling van productie en gebruik van duurzame biomassa en van een effectieve biomassamarkt zijn belangrijke voorwaarden voor het grootschalig gebruik van biomassa voor energiedoeleinden (en het maken van chemicaliën). Dit is belangrijk voor zowel IG-PG installaties als andere grootschalige biomassatoepassingen.
- Het opwaarderen van biomassa door torrefactie voordat de biomassa in de vergasser wordt geïnjecteerd zorgt voor een significante verbetering van de technische en economische prestaties van de vergassingsinstallatie. De technische haalbaarheid van grootschalige torrefactie van biomassa en zijn economische concurrentievermogen moeten nog in de markt worden gedemonstreerd.
- Niche-toepassingen voor vergassingsinstallaties, zoals de verwerking van huishoudelijk afval, zouden moeten worden geanalyseerd. Onderzoek, ontwikkeling, demonstratie en toepassing van vergassingsinstallaties voor die niche-toepassingen moeten worden gestimuleerd om verdere technologische ontwikkeling en kostenreducties mogelijk te maken.
- Mogelijkheden voor grote verbeteringen in de prestaties van vergassingsinstallaties zijn geïdentificeerd. Om deze verbeteringen te realiseren is een aanzienlijke toename van de wereldwijde vergassingscapaciteit noodzakelijk. De resultaten in dit proefschrift laten echter zien dat vergassingsinstallaties op korte termijn niet economisch concurrerend zijn. Het wordt daarom aanbevolen dat er beleidsmaatregelen worden ontwikkeld en geïmplementeerd die de bovengenoemde punten adresseren en de implementatie van vergassingsinstallaties bevorderen.

Suggesties voor verder onderzoek

- De technische modellen die in dit onderzoek ontwikkeld en gebruikt zijn konden niet worden gekalibreerd aan de hand van gegevens van operationele installaties, omdat er thans geen IG-PG installatie of SMR-installatie met CO_2-afvang operationeel is. Validatie van de modellen is nodig maar dit vereist het beschikbaar komen van gegevens van onder meer nieuwe demonstratieprojecten.
- In dit proefschrift is aangenomen dat de PR-waarden die horen bij het technologisch leren van een technologie in een vroeg stadium van zijn ontwikkeling hetzelfde blijven als de technologie verder geïmplementeerd wordt. Deze waarden kunnen echter veranderen als de toepassing van de technologie een volwassen marktpositie heeft bereikt. Het effect hiervan op de PR-waarden vereist meer onderzoek.
- Het onderling vergelijken van economische resultaten die in verschillende wetenschappelijke studies worden gepresenteerd is gecompliceerd doordat in

iedere studie andere kapitaalkosten worden gebruikt. Het is daarom aan te raden dat op componentenniveau een databank wordt gemaakt met uniform te gebruiken kapitaalkosten.

- In een installatie waar de voeding of de productie aangepast kan worden opereren meerdere componenten voor een deel van de tijd op deellast. Het effect van langdurig opereren op deellast op de beschikbaarheid en efficiëntie van die componenten moet worden onderzocht en geverifieerd in de praktijk.
- De flexibele IG-PG installatie in dit proefschrift is ontworpen om meerdere transportbrandstoffen, chemicaliën en elektriciteit te kunnen produceren. Het optimaliseren van de installatie voor één chemisch product (en elektriciteit) kan het energetisch rendement mogelijk verhogen. Als CCS wordt toegepast kan deze optimalisatie ook de hoeveelheid koolstof (in de vorm van CO_2) die afgevangen wordt vergroten. De mogelijke effecten van optimalisatie van een IG-PG installatie zouden verder onderzocht moeten worden.
- Integratie van de luchtscheider met de gasturbine zou het energetisch rendement moeten kunnen verhogen met enkele procenten. In dit proefschrift zijn de luchtscheider en de gasturbine echter niet geïntegreerd omdat dit beperkingen oplegt aan het vermogen van de gasturbine om op deellast te draaien. Het is aan te bevelen dat, wanneer er meer inzicht is in deze integratie, er gekeken wordt naar de mogelijkheden van integratie tussen de luchtscheider en de gasturbine zonder dat hierbij de mogelijkheid van IG-PG installaties om hun voeding of productie te veranderen wordt beperkt.
- In dit proefschrift zijn technische flowsheet modellen gebruikt om de massa- en energiebalansen van de belangrijkste stoffen te berekenen. Sporenelementen zoals fijn stof, halogenen, zware metalen en NO_x zijn niet in detail bekeken. In het bijzonder bij het flexibel bedrijven van de gasturbine kan NO_x uitstoot een probleem worden. Ook de productie en de emissie van andere sporenelementen moeten nader onderzocht worden om er zeker van te zijn dat de flexibele installaties voldoen aan de milieuwetgeving.
- De engineeringanalyse in dit proefschrift duidt erop dat nieuwe technologieën zoals syngasturbines, H_2-membranen, ITM en SOFC commercieel beschikbaar moeten zijn om de in dit proefschrift geprojecteerde productiekosten te halen. Om dit mogelijk te maken is verder onderzoek en ontwikkeling nodig.
- Bij het bepalen van de mogelijke verbeteringen in de prestaties van vergassingsinstallaties is in dit proefschrift gebruik gemaakt van technische en economische projecties van technologieën die op commerciële schaal nog niet zijn getest, zoals membraanscheiding en SOFC. Dit resulteert in onzekerheid in de gevonden waarden. In de toekomst zouden up-to-date analyses van de technisch-economische prestaties van vergassingsinstallaties uitgevoerd moeten worden. Dit zou moeten leiden tot nauwkeuriger resultaten.
- In dit proefschrift zijn de technisch-economische prestaties van 100% biomassavoeding geanalyseerd ondanks het feit dat in de huidige vergassers de mogelijkheden en effecten van het meestoken van een zeer grote fractie biomassa (tot 100%) nog niet getest zijn. Pilot experimenten zijn gewenst om de maximale fractie biomassa die meegestookt kan worden te bepalen en om operationele technieken te ontwikkelen om deze fractie te vergroten.

- De analyse in dit proefschrift van het uitrusten van een SMR-installatie met CCS betreft een retrofit van een bestaande installatie. De CO_2 vermijdingskosten zijn waarschijnlijk een stuk lager als de CO_2-afvanginstallatie een integraal onderdeel bij de bouw van de SMR installatie is. De effecten hiervan op de technisch-economische prestaties van een SMR-CCS installatie zouden onderzocht moeten worden.

9 Appendix: Technical description of IG-PG facilities

9.1 Feedstock

The investigated facilities can process solid carbon-containing feedstock. The feedstocks used in this study are coal and torrefied biomass pellets (TOPS). Their properties are given in Table 9.1. Untreated biomass is excluded, as previous studies by Meerman et al.,[150,173] revealed that using TOPS results in a superior technical and economic IG-PG performance. The main reasons are the lower moisture content and higher heating value of TOPS compared to untreated biomass. Natural gas and (heavy) oil are also excluded from this study as their conversion into syngas requires a different gasifier (or reformer).

Table 9.1 Feedstock parameters used in this study.

	Unit (1)	TOPS[90]	Illinois #6 coal[74]
Heating value	MJ_{HHV}/kg a.r.	20.51	27.14
Moisture	wt% a.r.	5.00	11.12
Volatile Matter	wt% dry	75.90	44.50
Fixed Carbon	wt% dry	24.10	44.59
Ash	wt% dry	1.34	10.91
C	wt% d.a.f.	54.63	80.50
H	wt% d.a.f.	5.67	5.68
O	wt% d.a.f.	39.45	8.70
N	wt% d.a.f.	0.22	1.58
S	wt% d.a.f.	0.02	3.17
Cl	wt% d.a.f.	0.01	0.37

(1) The composition is on as received (a.r.), moisture free (dry) or moisture and ash free (d.a.f.) basis.

9.2 Feedstock pre-treatment

Feedstock pre-treatment in the IG-PG facility involves drying (coal only), sizing, pressurization and feeding of the feedstock into the gasifier. Feedstock pre-treatment increases overall efficiency, improves carbon conversion and reduces downstream compression requirements.

Part of the pre-treatment is lowering the moisture content of the feedstock, increasing the higher heating value of the fuel and reducing the amount of water which needs to be evaporated in the gasifier. The energy required for reducing the moisture content in the feedstock, however, increases with decreasing moisture content. Therefore, coal is generally dried to a moisture content of <5%.[142,188] Due to torrefaction, TOPS already has a low moisture content and needs no further drying. The heat from the flue gas can be used to (partly) supply the energy for drying.

The second main process of the pre-treatment is reducing the particle size of the feedstock. A small particle size is needed as the residence time of the feedstock in the gasifier is typical a few seconds only.[47,100] To ensure almost complete carbon conversion, the particle size must be less than 1 mm for TOPS and 0.1 mm for coal. Because of their brittle nature, both coal and TOPS can be sized using hammers.

The final step of the pre-treatment is to pressurise the feedstock as most modern IG-PG facilities are operated at elevated pressure. This not only reduces the equipment size of the downstream processes, it also reduces or even eliminates the need to compress the syngas for the downstream conversion processes.

9.2.1 Feedstock processing and feeding

The current practice to pressurise and feed solid dry feedstocks is by using lock hoppers. In this system, the feedstock is fed into large bins. By sealing the bins and adding an inert gas, commonly N_2 or CO_2, the pressure is increased. The pressurised feedstock is fed into storage bins. The gasifier is fed from these bins. The downside of this process is the use of large amounts of inert gas for the compression and the need for a pressurised storage tank to allow continuous feeding. A new technology is the solid feed pump, see Figure 9.1. Examples are the Stamet Posimetric and Pratt & Whitney Rocketdyne (PWR) pumps. The Stamet pump consists of two discs mounted on a shaft, forming a spool. The spool is placed inside a casing, forming a duct. A continuous flow of pulverised feedstock enters the pump at the top. As the feedstock rotates through the duct, it is transported from a low pressure environment to a high pressure environment. Tests showed the pump can compress 30 t/h coal from atmospheric pressure to 70 bar.

Tests so far only included pulverised coal. Feeding of biomass, which is much more fibrous, has not been tested yet. Feeding of TOPS, however, should not be more difficult than feeding coal as most fibres will be destroyed during the torrefaction process. The feeding rate can be adjusted by varying the speed of the pump.[262] The PWR pump consists of two caterpillar track-like belts. The feedstock is led in between the belts and forced through a narrow exit. It is claimed that the feedstocks can reach pressures of over 80 bar. The pump is being tested at present.[263]

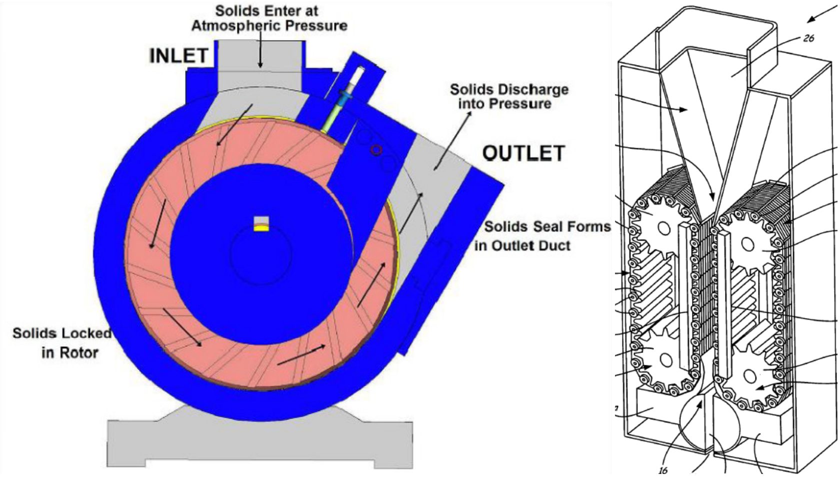


Figure 9.1 Solid feed pumps (Stamet Posimetric - left and Pratt & Whitney Rocketdyne - right).[262,263,264]

The solid feed pump biggest advantage is the ability to replace slurry feeding systems by a dry feeding system. As the investigated Shell gasifier is already a dry-feed type gasifier, the improvement in plant performance obtained by applying the solid feed pump is limited. Compared to the lock hopper system, capital costs of the solid feed pump are lowered by 67%, but specific energy consumption is about 30% higher. Inert gas entering the gasifier is reduced by 88%, increasing cold gas efficiency.[59,100,175] Another advantage of a solid feed pump is the ability to compress the feedstock to higher pressures than lock hoppers. This in itself has no added value, but combined with advances in gasification design it allows for high pressure gasification systems, increasing overall efficiency and potentially decreasing capital costs.[77,188,265]

9.3 Air separation

The energy required to heat the gasifier to over 1500°C is supplied by combusting part of the feedstock. To improve overall efficiency, pure oxygen ($\geq$ 95 mol%) instead of air is fed into the gasifier.[47,50] The current process of producing large amounts of high purity oxygen is with a cryogenic air separation unit (ASU). The first step is compressing air. Next, moisture and CO_2 are removed by condensation and the use of beds of silica gel and molecular sieves. This prevents frosting on the heat exchangers. The air is cooled to around -196°C. At this temperature the oxygen liquefies and is easily removed from the still gaseous nitrogen. Disadvantages of this process are the large capital costs and energy consumption. Most of the required energy is used in compressing the air. To reduce the energy consumption, a partial integration with the gas turbine can be applied. In this case, the gas turbine compresses (part of) the air. As the compressor of the gas turbine is more efficient than the compressor of the ASU[29], this results in a reduction in specific energy consumption. However, integration also increases the complexity of the facility, decreasing the availability of the facility. It is expected that economies of scale, improvement of the main air compressor and better integration with the gas turbine will reduce the specific energy consumption of the cryogenic ASU.[176,188,226] Over the past 40 years, specific energy consumption of (low pressure) cryogenic ASU decreased almost linearly by 2% per year (see Figure 9.2).[106,176] In this study, it is assumed that specific energy consumption will continue to decrease by 2% per year till 2020.

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[29] The ASU compressor is powered by electricity, while the GT compressor uses mechanical energy. The lack of conversion from mechanical energy to electricity is the main reason for the increased GT compressor efficiency.

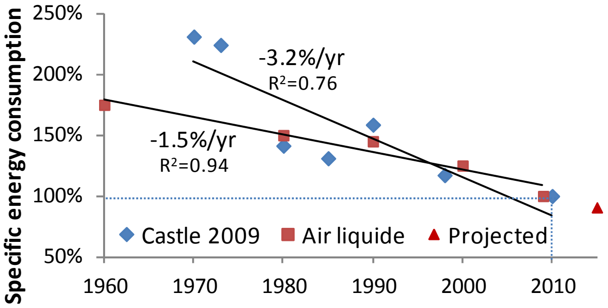

Figure 9.2 Specific energy requirement ASU over time.[106,176]

9.3.1 *Ion transport membrane separation*

A new air separation process is ion transport membrane (ITM) separation. This process uses the difference in permeabilities of the various components in air. The driving force is the difference in partial pressure between the two sides of the membrane. With ITM separation, air is compressed to 10-30 bar and heated to 800-900°C. The air is passed along a membrane which is selectively permeable to O^{2-} ions and electrons.

The oxygen ions migrate through the membrane and recombine to O_2 at low pressure at the permeate side (see reactions 9-1 and 9-2). The still pressurised retentate (non-permeate) consists of mainly nitrogen and can be used for the gas turbine. The membrane is selectively permeable to oxygen-ions due to the presence of oxygen vacancies within the non-porous ceramic membrane. Oxygen-ions can move through the membrane by hopping.[226] This hopping is thermally limited. Current materials exhibit severely reduced hopping activity at lower temperatures (<700°C). Below 450°C there is almost no hopping activity, making the membrane non-permeable to oxygen ions.[266,267,268] The main energy consumption of the ITM process is to compress the air and to maintain the membrane at high temperatures. Future energy savings can be made by improving the compressor efficiency and by developing low temperature ITM membranes.

$O_2 + 4e^- \rightarrow 2\,O^{2-}$ (at the cathode) (9-1)
$2\,O^{2-} \rightarrow O_2 + 4e^-$ (at the anode) (9-2)

Testing of a 5 t O_2/d test plant using 0.5 t O_2/d modules was completed in 2006. In 2010, 1 t O_2/d modules were successfully tested. Testing of a 100 t O_2/d pilot plant should begin in 2012-2013. Construction of a full scale ITM plant of 1500-2000 t O_2/d is scheduled in 2015. This is enough oxygen to supply a 250 MW_e IGCC.[177,269,270] Current R&D focuses on upscaling ITM as well as lowering membrane production costs, improving membrane performance under both oxidizing and reducing conditions, demonstrating the reliability of the membranes and optimising the integration of ITM in, among others, gasification facilities.[192,226,268] Initially it was expected that ITM would reduce both capital costs and specific energy consumption of the ASU by around 35%.[192] This has been downgraded to a 33% reduction in capital costs and a 19%[193,194] reduction in specific energy consumption.[177] Another benefit of the ITM is that almost 100% pure oxygen is produced.[59]

9.3.2 Chemical looping

Another air separation technique is chemical looping, where oxygen is supplied by a (solid) oxygen carrier. The carrier consists of a metal oxide, usually based on iron, nickel, copper, cobalt or magnesium. Chemical looping is a two-step process. In the first step, the metal carrier adsorbs oxygen from the air (see reaction 9-3). In the second step, the metal oxide reacts with the fuel and is reduced to the original metal carrier. The fuel is combusted at the same time (see reaction 9-4). The process can take place in a single reaction. The benefit is that there is no transport of metal particles.[271,272] This configuration is, however, a batch process. To create a continuous oxygen supply, as required for gasification facilities, multiple reactors must be installed in parallel, increasing capital costs.

$$\mathrm{Me_vO_w + O_2 \rightarrow Me_vO_{w+2}} \qquad \text{(air side)} \qquad (9\text{-}3)$$

$$\left(x + \frac{y}{4}\right)\mathrm{Me_vO_{w+2} + C_xH_y \rightarrow Me_vO_w} + x\,\mathrm{CO_2} + \left(\frac{y}{2}\right)\mathrm{H_2O} \qquad \text{(fuel side)} \qquad (9\text{-}4)$$

Chemical looping works best with gaseous fuels, although new techniques and metal carriers are being developed allowing the gasification of solid fuels. Reaction rates are, however, low. The relative low temperature (<1000°C) reduces reaction rates even further. Higher temperatures are not possible as the metal particles would sinter. A major problem is that chemical looping can produce H_2 only and not CO.[271,272] It is possible to produce CO via the WGS reaction, but this will lower efficiency. Alternatively, produced methane can be reformed into H_2 and CO, but this requires an additional reformer, increasing capital costs. Therefore, this technique is not considered in this study.

9.3.3 (Vacuum) pressure swing absorption

An already commercially applied oxygen separation technique is (vacuum) pressure swing absorption (V-PSA). In this process air is led over a bed with a nitrogen absorption material, e.g., zeolite molecular sieves. The nitrogen is absorbed out of the air and oxygen with 90-94% purity is obtained. The absorbed nitrogen is removed by reducing the pressure. To enable a continuous oxygen supply, several parallel beds are required. Currently, the production of large amounts of high purity oxygen is done more economical by the cryogenic process. The main reason for this is the lack of scalability of the PSA system. A higher capacity means more absorption material, having almost no advantage of economies of scale.[176] Major technological breakthroughs in PSA to make it competitive with cryogenic ASU are not expected. Therefore, this technology is not considered in this study.

9.4 Gasifier

The gasification market is currently dominated by coal-fed entrained flow (EF) gasifiers. EF gasifiers are pressurised, oxygen-blown gasifiers. The feedstock is either compressed with air (dry-fed) or mixed with water (slurry-fed). Dry-fed gasifiers are operated at around 40 bar, while slurry-fed gasifiers can reach 60 bar. A main advantage of a dry-fed EF gasifier is the ability to gasify feedstocks with a wide range of heating values, allowing the use of both coal and biomass. Therefore, in this study a dry-fed, oxygen blown Shell EF gasifier is used. Currently, EF gasifiers can co-gasify maximum 50% biomass. It is assumed that in the future, this limitation will be overcome as a result of improved materials and operating experience.

The syngas exiting the gasifier is quenched from 1500°C to 850- 900°C. This quench prevents particulates from sticking to the heat exchangers. The syngas is then further cooled to allow cleaning of the gas. Currently, this syngas cooler uses the heat to generate medium pressure steam. Even though temperatures are high enough to generate high pressure steam, this reduces availability due to operational problems. It is assumed that the generation of high pressure steam without sacrificing availability is possible in the long term.

Current R&D focuses, among others, on operating at increased pressure, improving availability, and on increasing the allowable biomass content. New gasification techniques such as supercritical water gasification and high pressure gasification are also investigated.

9.4.1 Supercritical water gasifier (SCWG)

A technology to efficiently convert (very) wet feedstocks is the supercritical water gasifier. By operating above the critical point of water (374°C and 221 bar) the heat of water vaporisation becomes zero. SCWG is performed at low temperatures (375-500°C) with a metallic catalyst, yielding a methane rich gas, or at high temperatures (500-750°C) with a non-metallic catalyst or without any catalyst, yielding a H_2 rich gas. This technology is still in the laboratory scale.[113,273,274] Furthermore, the greatest advantage is achieved when processing very wet feedstocks (>70% moisture content). The feedstocks investigated in this study have a low moisture content. Therefore, this technology is not considered in this study.

9.4.2 High pressure gasification

A high-pressure gasifier (compact gasifier) is being developed by Pratt & Whitney Rocketdyne Inc. (PWR). Its main advantage is the higher pressure of 80 bar. It is expected that the PWR gasifier volume could be reduced by 90% compared to conventional gasifiers. Other improvements are a novel injection system and new cooled refractory liner design. Using the PWR gasifier increases IGCC efficiency by $0.9\%_{pt}$ and reduces capital costs of the gasifier by 50% compared to a conventional Shell gasifier. Note that part of the improvement in overall efficiency is obtained by the feeding system.[275,276,277] The PWR gasifier uses a solid feed pump, while the Shell gasifier is fed via lock-hoppers. A 18 t/d gasifier successfully ran for 764 hours in 2011.[277] As the gasifier has not yet undergone large-scale and long-term testing, the expected improvements and commercialisation date are uncertain. These reasons, combined with the lack of data to accurately model the PWR gasifier, led to the exclusion of this gasifier for the different time periods. The PWR gasifier is, however, considered in the sensitivity analysis to determine the effect of the higher pressure on overall IG-PG facility performance.

9.5 Gas Cleaning

Before the raw syngas can be used for chemical, fuel or electricity production, it needs to be cleaned of contaminants, e.g., particulates, halides, acid compounds and trace amounts of heavy metals. The removal of the contaminants is done in the gas cleaning section and increases the efficiency of downstream syngas conversion equipment as well as protecting catalysts and heat exchangers from poisoning and fouling. It is also mandatory to comply with environmental legislation. The adjustment of the H_2:CO composition of the syngas is also performed in the gas cleaning section.

The gas cleaning in a modern entrained flow gasification facility starts after the syngas is cooled to 200°C in a syngas cooler. The first step is a filter to remove solid particles. The gas is scrubbed in a wet scrubber to remove alkali and ammonia compounds and any remaining fine particulates. The temperature of the wet scrubber is limited by the boiling point of water at the operating pressure. The gas is reheated to 250°C and passed through a WGS reactor to adjust the H_2:CO ratio. Depending on the AGR solvent, COS needs to be removed via a guard bed prior to the AGR. Another guard bed, operating around 200°C, is used to remove heavy metals, including mercury. The AGR operates at temperatures around 0°C. Here, the acid gasses, e.g., CO_2 and sulphur compounds, are removed. Depending on the syngas conversion process, the gas is fed into a second guard bed which ensures that the sulphur content in the sweet syngas is low enough for the syngas conversion equipment, e.g., chemical, fuel or electricity production. The final step is heating the syngas to the desired temperature needed for the syngas conversion process.

The different gas cleaning processes all target a specific contaminant and all operate at a specific temperature. The result is a rollercoaster-like temperature profile (Figure 9.3).[150,172] Smoothing out this profile by rearranging the cleaning equipment is difficult as most processes only operate properly when certain contaminants are already removed. With proper heat integration and steam generation, the impact of the variations in temperature can be minimised, but this increases capital and operational costs and overall efficiency is still reduced. Current gas cleaning research focuses on processes which operate at higher temperatures as well as processes which can remove several types of contaminants simultaneously.[47,58,59,172,193]

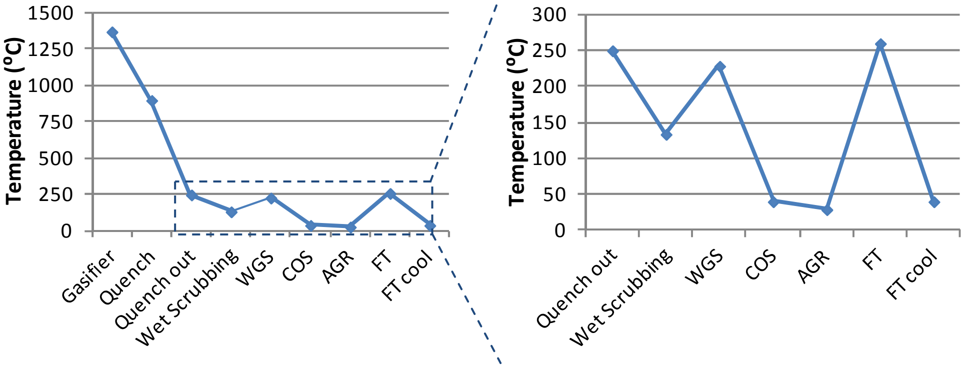

Figure 9.3 Temperature profile of a SOTA IG-FT gas cleaning train.[172]

9.5.1 Hot gas cleaning

A major obstacle in hot gas cleaning is that the entire gas cleaning train must be operated at elevated temperatures. The minimum temperature aimed for at hot gas cleaning is 250°C. This means that new processes have to be found for the wet scrubber, COS hydrolysis and AGR, needed for the removal of ammonia, chloride, sulphur and mercury compounds (and CO_2). Currently, as shown in Figure 9.3, the lowest temperature in the gas cleaning section occurs at the AGR process.[58,59,278,283]

According to a study published by the NETL, replacing a conventional cold gas cleaning train using Selexol-based AGR, Claus and SCOT with a hot gas cleaning train using absorber and catalytic beds for trace elements and TDS and DSRP for sulphur removal could decrease capital costs of the gas cleaning train by 26% and increase O&M costs of the gas cleaning train from 4% to 8%.[59] The NETL study does have a slurry-fed gasifier. In this study it is assumed that the found improvements also hold for dry-fed gasifiers. In the case of IGCC-CCS a hot gas cleaning train, including H_2-membranes, decrease capital costs of the gas cleaning train by 35% compared to the cold gas cleaning train. O&M costs of the gas cleaning train increase from 4% to 8%. Not taken into account is the higher pressure at which the CO_2 enters the compressor.[58]

9.5.1.1 *Trace element removal*

The syngas contains various trace elements. They are currently removed by a filter, wet scrubber, guard bed and AGR. In hot gas cleaning, catalytic and absorber bed will be used to remove these trace elements. Below are a short summary of the important trace elements and an overview of the beds. The difficulty is finding absorbents and catalysts which are not poisoned by the other compounds of the syngas. In this study it is assumed that proper absorbents and catalysts have been found in the mid term.

Alkali

To allow removal of alkali compounds at high temperatures, the getter is developed. Here, alkali vapour is chemically or physically absorbed by fixed (filter) beds containing granular inorganic solid sorbents, like bauxite, at temperatures between 600-900°C.[120,279,280]

Ammonia

Ammonia can be catalytically destroyed into H_2 and N_2. To ensure high reaction rates and full conversion at high pressure, the operating temperature is above 700°C. Although various catalysts can decompose ammonia, the presence of, among others, H_2, CO, H_2O and sulphur compounds severely limits the number of possible catalysts.[120,280,281,282] No experimental data were found where ammonia in syngas from an operational EF gasifier was successfully converted into H_2 and N_2.

Another technique to remove ammonia from the syngas is by adding an oxidiser.[282] Due to the presence of H_2 and CO, combined with high temperatures and pressures, this is deemed too risky and is not being considered in this study.

Chlorides

Hydrogen chloride can be removed using a packed bed of sodium bicarbonate or sodium carbonate at 500°C. The presence of sulphur does affect bed performance, although no significant impact was detected below 5000 ppm H_2S.[58,59,280]

Mercury

Mercury and other heavy metals are currently removed using non-regenerable absorber bed containing activated carbon upstream the AGR. Currently, these beds can operate at around 200°C.[47,283] It is assumed that for the mid term the operating temperature is increased to 250°C.

9.5.1.2 Acid Gas Removal and downstream sulphur treatment

The AGR removes sulphur compounds from the syngas. In the case of IGCC-CCS, IG-FT or when fuel cells are used, the AGR also removes CO_2. This additional function of the AGR has an impact on when the AGR can be improved and what the effect on overall facility performance and economics will be.

The current method to remove sulphurous compounds and/or CO_2 is using (chemical or physical) solvents. For IGCC facilities Selexol and MDEA are commonly used, while for IG-FT the more expensive Rectisol is used due to its better sulphur removal.[42,50,284] Drawbacks of the solvents are the high specific energy consumption and the low operating temperature. R&D focuses on increasing the loading capacity, thereby reducing solvent recirculation rate, and increasing the operating temperature.

Once the (gaseous) sulphurous compounds have been removed from the syngas, they are converted to elemental sulphur by catalytic reaction. This is done in the Claus plant. To comply with environmental legislation, the off-gas from the Claus plant is fed into the Shell Claus Off-gas Treatment (SCOT) plant, where the remaining sulphurous compounds are extracted from the gas and recycled to the Claus plant. This way, more than 99.5% of the sulphur present in the feedstock is converted into elemental sulphur. R&D focuses on removing the need for the SCOT plant and on combining the sulphur extraction and conversion to elemental sulphur into one reactor.

Advanced solvents

Solvents with superior absorption and regeneration properties are currently being investigated. However, performance data of these solvents is considered proprietary. Therefore, the estimation of Peeters et al. for post-combustion CO_2 capture solvents was used in this study. They expect that the regeneration energy of chemical solvents will reduce by 41% and 64% in the short and mid term respectively.[179] It is assumed that operating temperature and pressures as well as capital and O&M costs will remain constant.

Transport desulphurisation with direct sulphur recovery plant

In a transport desulphurisation (TDS) reactor H_2S and COS react with metal oxide sorbents (equations 9-5 and 9-6). The TDS process does not capture CO_2. If CO_2 is to be removed, an additional CO_2 capture unit is required. The loaded sorbents are fed into the regenerator where air is added. At 600°C the absorbed sulphur is converted into SO_2 and the sorbent is regenerated (equation 9-7). The regenerated sorbents are subsequently fed into the TDS reactor (see Figure 9.4). The SO_2 is converted in the direct sulphur recovery plant (DSRP) to elemental sulphur by reacting with H_2 or CO (equations 9-8 and 9-9), which is supplied by redirecting a small portion of the clean sweet syngas.[59,120,280,282,283]

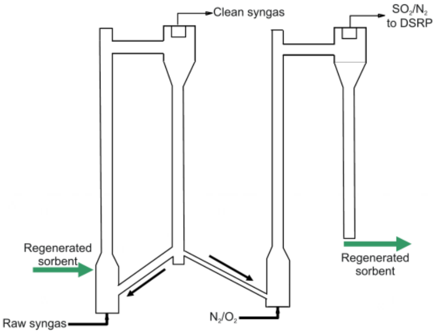

Figure 9.4 Schematic overview of transport desulphurisation.[283]

$H_2S + ZnO \leftrightarrows ZnS + H_2O$ (absorption TDS) (9-5)
$COS + ZnO \leftrightarrows ZnS + CO_2$ (absorption TDS) (9-6)
$2\ ZnS + 3\ O_2 \leftrightarrows 2\ ZnO + 2\ SO_2$ (regeneration TDS) (9-7)

$SO_2 + 2\ H_2 \leftrightarrows 2\ H_2O + S$ (DSRP) (9-8)
$SO_2 + 2\ CO \leftrightarrows 2\ CO_2 + S$ (DSRP) (9-9)

Different metal oxides can be used, but promising results have been obtained by using zinc oxides. A major drawback is the formation of carbon deposition from CO at lower temperatures (<450°C). Doping the zinc oxide with different metal oxides reduced carbon deposition, but not significantly enough.[281] In this study it is assumed that in the mid term, sorbents are available which can function at lower temperatures without significant carbon deposition.

Selective catalytic oxidation of hydrogen sulphide

The selective catalytic oxidation of hydrogen sulphide (SCOHS) process lets H_2S react with oxygen to elemental sulphur (see equation 9-10). This reaction is catalytically and occurs below 200°C. Compared to a conventional AGR and Claus unit, the SCOHS process could improve overall IGCC efficiency by $1.2\%_{pt}$, while capital costs of the AGR and sulphur processing is reduced by 62%.[285] Currently, this concept has been tested at a laboratory scale.[286] The main drawback of this process is the introduction of oxygen in the syngas. This makes the fire and explosion risks unacceptable. Therefore, this concept is not used in this study.

$2\ H_2S + O_2 \leftrightarrows 2\ S + 2\ H_2O$ (9-10)

Membranes
A promising method to remove CO_2 from the syngas is by using membranes. The currently investigated membranes are sensitive to sulphur. They have to be located downstream of the sulphur removal. The membranes can either be permeable for H_2 (H_2-membranes) or CO_2 (CO_2-membranes). Although both types are being investigated, most R&D is dedicated on H_2-membranes. Therefore, in this study only H_2-membranes are being considered. This does mean that the CO concentration must be a low as possible, making H_2-membranes only interesting for IGCC equipped with CCS. In case of the IG-FT facilities, a portion of the syngas could bypass the H_2-membrane. To obtain sufficient CO in the FT-reactor, however, would require such a large bypass that the CO_2 level in the FT-reactor becomes too high, leading to a large fraction of small hydrocarbons in the FT-product, which results in a large increase in production costs. Therefore, this option is not used in this study.

By combining H_2-membranes with TDS and DRSP, no solvent based AGR is required. This will eliminate the need for regeneration heat. An additional advantage is the higher pressure at which the CO_2 becomes available. Lastly, as the H_2 is removed from the CO_2, instead of the other way around, capture rates of almost 100% are possible. Main problems with the membranes are obtaining a high flux while avoiding leakages and the high production costs combined with a relative short lifespan. Currently, H_2-membranes have successfully been tested on lab scale. Upscaling, however, proves difficult. Production costs are estimated around 24 M€/(17 t H_2/hr) with an operational lifetime between 5-10 years.[58,77]

9.6 Water-gas shift

When producing chemicals/fuels or capturing CO_2, the $CO:H_2$ ratio of the syngas needs to be shifted. This is done in the water-gas shift (WGS) reactor (reaction 9-11). As this reaction is exothermic, thermodynamic equilibrium is shifted towards H_2 production at lower temperatures. At these temperatures, however, reaction rates are uneconomically slow, even with the use of catalysts. To increase reaction speeds, while obtaining almost full CO conversion, the combination of high temperature (300-500°C) and low temperature (200-270°C) WGS reactor are often used. Additional steam is added to increase the H_2 yield even further and prevent carbon deposition on the catalyst.

$$CO + H_2O \leftrightarrows CO_2 + H_2 \qquad \Delta H^\circ_{298K} = -41.2\ kJ/mol \qquad (9\text{-}11)$$

9.6.1 Sorption enhanced water-gas shift

Another method to shift the thermodynamic equilibrium towards H_2 production is by reducing the CO_2 concentration. This is done in the sorption enhanced water-gas shift (SEWGS) reactor. This reactor is filled not only with WGS catalysts, but also with hydrotalcite-based CO_2 sorbents. The catalyst and sorbent are mixed together. As CO reacts with H_2O to H_2 and CO_2, the CO_2 is almost immediately absorbed and removed from the gas phase. Using this method, almost full conversion can be obtained at temperatures over 400°C. The sorbents are regenerated using low pressure and steam. The result is a CO_2 stream which is, after drying, 98% pure. The drawback of the system is that the reactor needs to be regenerated. To ensure continuous operation, multiple reactors are required. Experimental results indicate 6 parallel reactors will allow this as well as reducing energy consumption by integrating the de- and repressurisation steps. Another drawback is that the WGS catalyst is currently sensitive to sulphur poisoning. The SEWGS reactor should therefore be located after the AGR. A sulphur tolerant sorbent is already developed, but the separate regeneration of H_2S and CO_2 still requires more research.[287] Current R&D also focuses on sulphur tolerant WGS catalysts and on reducing the specific energy consumption. It is expected that replacing a conventional WGS reactor and amine scrubber with the SEWGS system in an IGCC-CCS can increase overall efficiency by 1.1-3.8%, while reducing capital costs by 19%.[288,289] Currently, a 250 kg CO_2/hr SEWGS system, equivalent to 400 kWe, is being tested.[290]

9.6.2 Advanced water-gas shift

A major part of the energy penalty of the WGS reactor is the consumption of MP steam, a main result of the minimum H_2O:CO ratio to avoid carbon deposition. The advanced WGS process reduces this consumption by splitting the syngas in several fractions (Figure 9.5). A fraction of the syngas is injected in the first WGS reactor and sufficient steam is added. The shifted syngas is mixed with another fraction of the syngas and injected in the second WGS reactor. This time water is added. The heat of reaction is used to evaporate the water. This principle is repeated till all the syngas is shifted. The advantage is not only a reduced steam consumption, but also a lower total water consumption. The optimal number of WGS reactors depends on the desired CO conversion. Replacing a WGS with an advanced WGS could lower steam consumption by 54% and increase overall efficiency by $1\%_{\text{pt LHV}}$.[291,292]

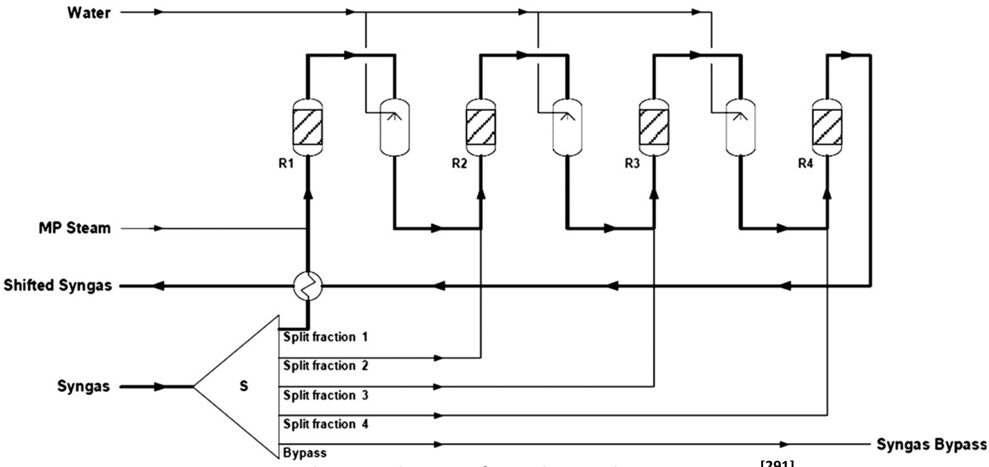

Figure 9.5 Schematic layout of an advanced WGS process.[291]

9.7 CO_2 compressor

Multistage compressors are used to compress the CO_2 to 110 bar. Between the stages the CO_2 is cooled to improve compression efficiency. The CO_2 compressor makes up around 2-3% of total capital costs and consumes about 2-4% of gross power production in a conventional coal-fired IGCC with CO_2 capture.[25,50,74,188] The relative high capital costs are caused by the use of stainless steel (as the water vapour in the CO_2 makes the gas corrosive) as well as the high number of stages. Electric consumption of conventional compressors can be reduced by increasing compression efficiency or by increasing the pressure at which the CO_2 comes out of the AGR. The energy required to isothermally compress CO_2 at 25°C from 1 bar to 110 bar is 265 kJ/kg CO_2, see equation 9-12. This drops to 242 kJ/kg CO_2 (-9%) if the CO_2 is available at 1.5 bar. At 2 bar the energy consumption is 226 kJ/kg CO_2 (-15%).

$$E=n*R*T_0*\ln\left(\frac{p_{out}}{p_{in}}\right) \tag{9-12}$$

where: E = the required energy (kW_e)
n = the molar flow of the gas (kmol/s)
R = the gas constant (8.314 kJ/(kmol*K))
T_0 = inlet temperature (K)
p = the pressure of the ingoing and outgoing gas (bar)

9.7.1 Shock wave compression

Current CO_2 compressors are limited to compression ratios below 2:1 per stage. A main reason is the need to stay under the sonic velocity due to mechanical limitations. Therefore, compressing from 1 to 110 bar requires 7-8 stages. Ramgen Power Systems LLC is developing a supersonic shock wave based compression able to achieve compression ratios exceeding 10:1 per stage, lowering the required number of stage to 2. Currently, a 1 MW_e LP compressor and an 8 MW_e HP compressor are being tested. The HP compressor is large enough to compress all the captured CO_2 from a 200 MW_e coal power plant. The technology is expected to be available for 1000 MW_e power plants around 2015.[293,294,295,296]

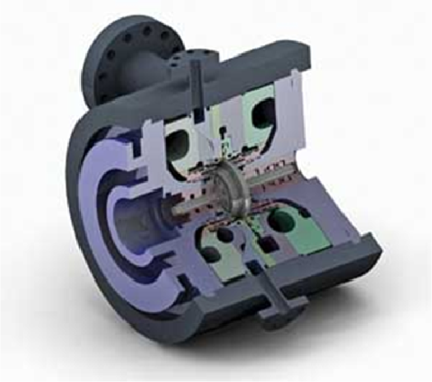

Figure 9.6 Scale model of a Ramgen shock wave compressor.[293]

The main advantages of the Ramgen compressor are an expected 50-60% reduction in capital costs and an increase in recoverable heat as the gas is discharged at 230-260°C after each stage instead of <100°C. Ramgen claims electric consumption is almost similar to conventional compression.[293,294,295,296] Recent independent research, however, indicates that the Ramgen compressor could increase specific energy consumption by 11-20%.[189] Compensating for the pressures used in this study (1.5 bar to 110 bar) results in an increase in specific energy consumption of about 24%. The same research also indicates capital costs could reduce by 25-38%.[189] In this study, a reduction of 30% is assumed.

9.7.2 Electrolytic compression

A new technology is electrolytic compression. Currently, experimental work is only being conducted with H_2. A proton-permeable membrane separates the low and high pressure areas. An external current ionises H_2 at the low pressure side. The protons move through the membrane and recombine at the high pressure area. The main advantage of this technique is the higher efficiency compared to mechanical compression as no energy is lost heating the gas.[297]

Although this technique is being developed for hydrogen compression, it could theoretically also be used for CO_2 compression using molten carbonate membrane systems. The drawback is that when CO_2 migrates through the membrane as CO_3^{2-} it carries an additional O-atom. This O-atom will recombine to O_2, resulting in a high pressure O_2/CO_2 stream. The high O_2 partial pressure can give problems during transport and storage of the CO_2. Therefore, this option is not considered any further here.

9.8 Electricity production

The main processes in the electricity production section are the gas and steam turbines and the heat recovery steam generators (HRSG). In the gas turbine syngas is combusted and a part of the chemical energy is converted into mechanical energy, which is used to generate electricity. The HRSG is responsible for the heat integration of the different processes. Excess heat is transferred to steam and supplied to heat demanding processes. Any remaining heat is used to generate electricity.

9.8.1 Syngas combustion

In conventional IGCC facilities, syngas is mixed with air and combusted in a gas turbine. To reduce NO_X formation, the syngas is diluted with nitrogen. Modern gas turbines are designed for natural gas, which has a low H_2-content. Differences in flame temperature, flame speed, flammability range, elemental composition, etc., affect gas turbine efficiency, reliability and emission when combusting H_2-rich syngas. To reduce NOx emissions, the H_2 is diluted by adding significant amounts of steam or compressed nitrogen from the ASU. Depending on the facility, this addition of a diluent can even improve economics by increasing net power output.[298]

9.8.1.1 Gas turbine

Gas turbines are expected to become more efficient over time. This can be achieved by increasing the pressure ratio and turbine inlet temperature (TIT). IG-PG facilities have a combined cycle for electricity production. For maximum efficiency of the combined cycle the temperature of the flue gas exiting the gas turbine must be 50°C higher than the highest steam temperature used in the steam cycle. At a fixed TIT and flue gas temperature, there is a specific pressure ratio which results in the highest efficiency of the combined cycle. As the TIT increases, so does the optimal pressure ratio. In the past 50 years, advancement in metallurgy, cooling techniques and thermal coatings have increased the TIT to over 1600°C (see Figure 9.7). This, combined with more efficient compressors and expanders have increased gas turbine single cycle efficiency from under 15% in the 1950's to over 40% today.[180,299,300,301,302] Recently, Mitsubishi Heavy Industries (MHI) has introduced a 1600°C TIT turbine, claiming that it will boost NGCC combined cycle efficiency to over 60%$_{HHV}$.[303] Besides more efficient gas turbines, H_2-turbines, able to burn high H_2 concentration, are being developed and should become available in the mid to long term, increasing both the availability and efficiency of facilities burning H_2-rich gas.

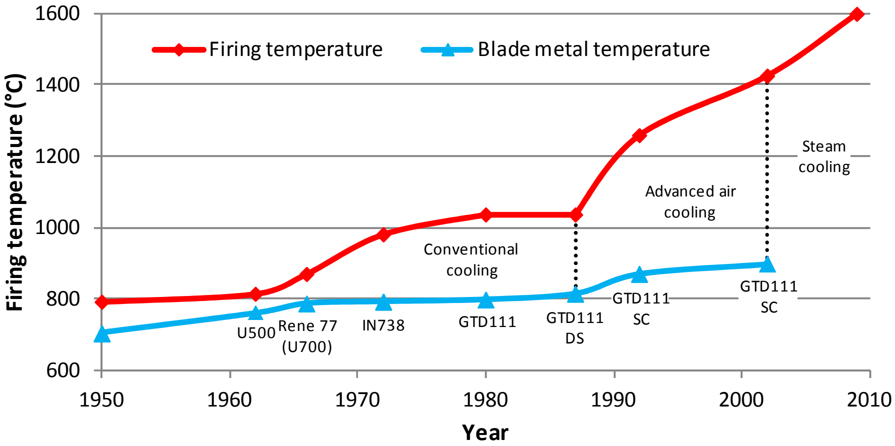

Figure 9.7 Development of gas turbine firing temperature.[180,304,305,306]

It is assumed that the TIT will increase from 1325°C now to 1700°C in the long term. Besides an increase in TIT, the gas turbine compressor and expander are expected to become more efficient. Another R&D emphasise is reducing the efficiency penalty when operating at part-load conditions. GE has developed a gas turbine which can reach a combined cycle efficiency above 60% at 87% part-load.[302]

9.8.1.2 Solid oxide fuel cell (SOFC)

An advanced form of combustion takes place in a SOFC. The main benefit of SOFCs over gas turbines is that chemical energy is directly converted into electrical energy. An anode and cathode are separated by an electrolyte of a solid, non-porous metal oxide. Above 600°C, this electrolyte becomes conductive for oxygen ions. At the cathode, molecular oxygen dissociates and reduces to O^{2-}. The oxygen ions migrate through the electrolyte. At the anode hydrogen reacts with the oxygen ions and form water. The electrons move from the anode to the cathode through an external circuit (see Figure 9.8).[307,308] As not the fuel but oxygen migrates through the membrane, a SOFC is also able to combust CO and methane. The current operating temperature for SOFCs is around 1000°C. This has the advantage of high conversion, high temperature waste heat, internal fuel reforming and protection from CO-poisoning. Main drawback of the high temperature is finding materials which are cheap to manufacture and are chemically and mechanically stable and have a high electrical conductivity at the operating conditions. The SOFC is susceptible to sulphur poisoning. However, the sulphur content of sweet syngas is most likely to have a small effect on the SOFC performance only.

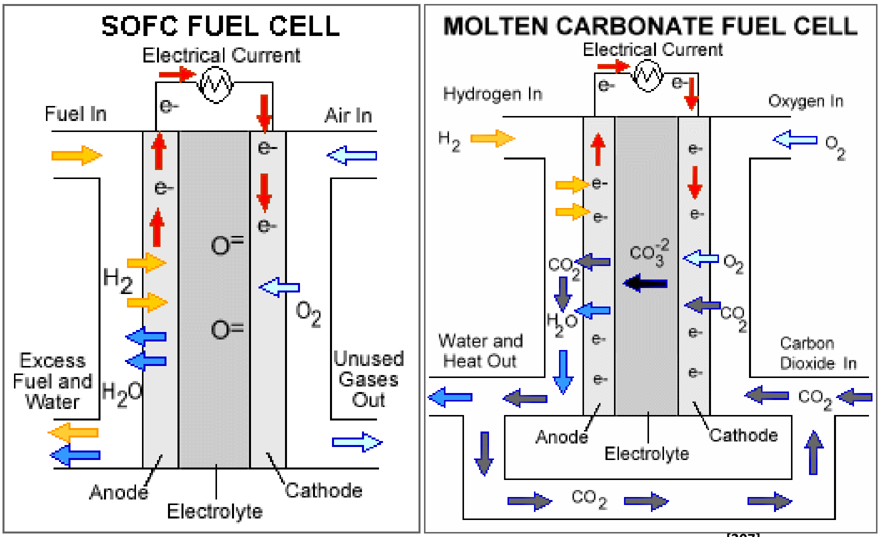

Figure 9.8 Schematic function of SOFC (left) and MCFC (right).[307]

Upscaling SOFCs is difficult and is currently only possible by installing more modules. Fuel conversion efficiency of a SOFC is expected to around 85% of the H_2. SOFC costs for 2050 are difficult to predict. For this study the updated costs expectation from the NETL of 0.50 M€/MWe was used. These costs include fuel cell stack, anode and cathode heaters, anode steam generator and reheat, syngas expander, cathode air compressor, anode and cathode expanders, inverter, catalytic oxidizer and oxygen boost compressor, condensate knockout, and foundations and are assumed to include all direct costs. Indirect costs were assumed to be 20%.[58,59]

9.8.2 Steam turbines and heat recovery steam generators

The two main variables determining steam cycle efficiency cycle are the steam inlet pressure and temperature (see Table 9.2). The higher these are, the more efficient the system. Modern IGCC operate their HRSG at subcritical pressures. Supercritical steam pressures are more used in pulverised coal power plants. In 2008 the electricity generation capacity by supercritical pulverised coal power plants exceeded 1500 TWh per year.[27] A major drawback of supercritical steam cycles is the high material cost of exotic alloys, needed to withstand the high temperatures. Current R&D focuses on developing less expensive alloys with high temperature resistant and on lowering production costs of exotic alloys.

Table 9.2 General overview of PC power plant steam cycle conditions.[309,310]

	Steam temperature (°C)	Steam pressure (bar)	Thermal cycle efficiency ($\%_{HHV}$)	Remarks
Subcritical	< 550	≈ 170	33-35	
Supercritical	< 600	220-275	36-40	1 Reheat
Ultra-supercritical	> 600	> 275	42-45	1 Reheat
Advanced USC	> 705	> 275	46-50	2 Reheat

A modern IGCC without CCS generates around 40% of its gross electricity production from the steam cycle. The improvement in efficiency listed above relate to PC plants, which have a much larger steam cycle. The improvements in the steam cycle will, therefore, have a smaller impact for IGCC. Also, a PC plant only has one main heat exchanger, which is located in the boiler. An IGCC has much more heat exchangers, making an upgrade of the steam cycle to supercritical more expensive. Therefore, it is assumed that only in the long term will it be economic to use a supercritical steam cycle in gasification facilities.

9.9 FT-synthesis

During FT-synthesis, syngas is catalytically converted into a mixture of linear alkanes and 1-alkenes with different chain length via reaction 9-13.[77,150] The chain length of the hydrocarbons can roughly be described using a modified Anderson-Schulz-Flory distribution, see equation 9-14.[77,150] Purifying and upgrading this mixture to transportation fuels consumes energy and capital. It is expected that the main improvement in this process is the development of a catalyst which has a higher selectivity, thereby reducing the separation and upgrading processes of the hydrocarbon mixture. It is assumed that in the long term catalysts are available which have a different chain growth probability (α), depending on the chain length.

$$2n\ H_2 + n\ CO \rightarrow (\text{-}CH_2)_n\text{-} + n\ H_2O \qquad (9\text{-}13)$$

$$\alpha_{Cn} = \alpha^{n-1} * (1-\alpha) \qquad (9\text{-}14)$$

The performance of the FT-synthesis section depends largely on the performance of the catalyst. Current R&D focuses on improving catalysts lifetime and increasing catalyst activity and selectivity. Improving the reactor can also reduce production costs by e.g., increasing conversion rates, catalyst activity or selectivity.[77,183,184,185,186] Note that reactor design and catalyst performance are closely linked to each other. No cost and performance projections were found in public literature. In this study it is assumed that in the long term diesel selective catalysts will become available. These catalysts result in a different α depending on the chain length of the hydrocarbon. Reactor size, costs and operating conditions are assumed to remain constant.

9.10 Availability

An important parameter of large-scale facilities is their availability. Not only does a high availability reduce production costs of electricity and FT-liquids, a high availability is also mandatory for the operation of commercial large-scale base load gasification facilities, especially power plants. For IG-PG facilities, increasing the availability from 80% to 85% decreases production costs by around 4%.[58,59,190] Initially, availability was low (see Figure 9.9).[190] This was mainly due to new technologies as well as using the facilities as testing facilities. Currently, an availability of 80% is already possible.[47,190,311,312] Availability is expected to gradually increase to 90% because of new technologies, advanced sensors and control systems, identification of critical failure-prone equipment and improved operating experience.[47,58,59,190] In this study it is assumed that the availability throughout the lifetime of the facility remains constant and that there are no start-up problems.

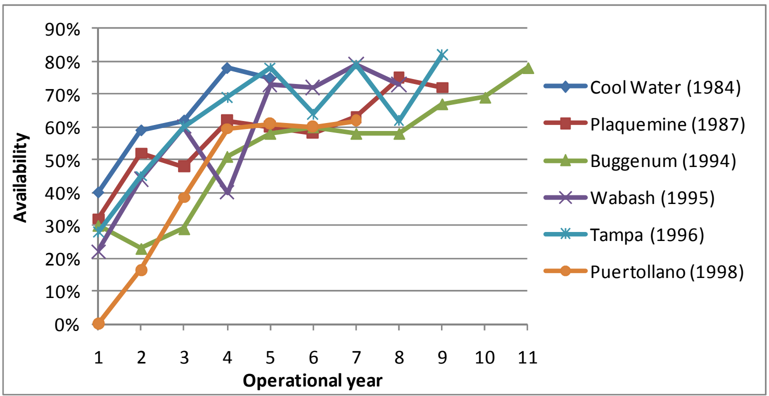

Figure 9.9 Availability of 1st generation coal-based IGCC.[190]

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Curriculum Vitae

Hans Meerman was born in Hellevoetsluis on August 3rd 1981. He studied chemical engineering at the Eindhoven University of Technology. In 2005 he performed a six-month internship at the Shell Refinery at Pernis on the mechanism of carbon deposits in the hydrogenation reactor. In 2006 he graduated on the possibility of adjusting the chirality of inorganic catalysts using acid/base reactions. In the same year he started working as a PhD student for the group Science, Technology and Society at Utrecht University, now called Energy & Resources, part of the Copernicus Institute. Here he investigated the technical and economic possibilities and limitations of altering feedstock or end-products in gasification facilities. The work was carried out under the CapTech programme and resulted among others in the thesis lying in front of you. In 2010 he joined the Utrecht Centre for Energy research (UCE), now integrated in the Utrecht Sustainability Institute (USI). In 2012 he was appointed as junior teacher at Utrecht University.

Nomenclature

α	Alpha: chain growth probability
a.r.	As received
ADIP-X	Mixture of MDEA, piperazine and water
AGR	Acid gas removal
ARA	Amsterdam, Rotterdam, Antwerp
ASU	Air separation unit
BaU	Business as usual
bbl	Oil barrel (0.159 m^3)
CCS	Carbon dioxide capture, transport and storage
CERA	Cambridge energy research associate
CGE	Cold gas efficiency
CHP	Combined heat and power
CO_2-eq.	CO_2-equivalent
COE	Cost of electricity
COP	Conference of the Parties
d.a.f.	dry, ash-free
DA	Direct action
DCP	Delayed climate policy
d_{max}	Maximum diameter
DSRP	Direct sulphur recovery plant
EF	Entrained flow
EGR	Enhanced gas recovery
EOR	Enhanced oil recovery
EP	Eucalyptus pellets
EPCCI	European power capital costs index
ER	Energy (r)evolution
E_x	Input or output of commodity x (GJ/yr or kt/yr)
FEP	Feedstock equivalent price
FGD	Flue gas desulfurization
FT	Fischer-Tropsch
FW	FosterWheeler
F_x	Annual flow of commodity x (GJ/yr or kt/yr)
GE	General Electric
GEA	Global energy assessment
GHG	Greenhouse gases
GT	Gas turbine
HHV	Higher heating value
HP	High pressure
HRSG	Heat recovery steam generation
I	Total investment costs (M€)
IEA	International energy agency
IF	Installation factor
IGCC	Integrated gasification combined cycle
IG-FT	Integrated gasification Fischer-Tropsch
IG-PG	Integrated gasification poly-generation
IP	Intermediate pressure
ITM	Ion transfer membrane
L	Economic lifetime (yr)

LC	Levelised production costs
LHV	Lower heating value
LNG	Liquefied natural gas
LP	Low pressure
LPG	Liquefied petroleum gas
LPMeOH	Liquid phase methanol
MDEA	Methyl diethanolamine or 2-N-Bis(2-hydroxyethyl)-methylamine
MeOH	Methanol
NETL	National energy technology laboratory
NG	Natural gas
NGCC	Natural gas combined cycle
NPV	Net present value
O&M	Operation and maintenance (M€/yr)
OGP	Other gasification plants
PBL	Netherlands environmental assessment agency (Planbureau voor de leefomgeving)
PC	Pulverised coal
PCCI	(American) power capital costs index
PEP	Production equivalent price
ppm (v)	Parts per million on volume basis
PPS	Procede process simulator
PR	Progress ratio
PSA	Pressure swing adsorption
P_x	Price of commodity x (€/GJ or €/kt)
r	Discount rate (%)
R&D	Research and development
RDD&D	Research, development, demonstration and deployment
RPS	Rotating particle separator
SCOHS	Selective catalytic oxidation of hydrogen sulphide
SCOT	Shell Claus offgas treatment
SCR	Selective catalytic reduction
SCWG	Supercritical water gasification
SEWGS	Sorption enhanced water-gas shift
SI	Système international d'unités
SMR	Steam methane reforming
SOFC	Solid oxide fuel cell
SOTA	State-of-the-art
SRT	Sulphur recovery treatment
ST	Steam turbine
t	tonne (1000 kg)
TCI	Total capital investments
TDS	Transport desulphurization
TIT	Turbine inlet temperature
TOPS	Torrefied woodpellets
TSA	Temperature swing adsorption
UNFCCC	United Nations Framework Convention on Climate Change
Vent	Without CCS
VLE	Vapour-liquid equilibrium
WEO	World Energy Outlook

WETO	World energy technology outlook
WGS	Water-gas shift
XtY	Facilities or systems where feedstocks are gasified and converted into products. The X is substituted if a specific feedstock is used: biomass (**BtY**), torrefied biomass (**TtY**) or coal (**CtY**). The Y is substituted if a specific output is produced: electricity (**XtP**), FT-liquids (**XtL**), methanol (**XtM**) or urea (**XtU**).
ZEP	European technology platform for zero emission fossil fuel power plants

k(ilo)	10^{3}
M(ega)	10^{6}
G(iga)	10^{9}
T(era)	10^{12}
P(eta)	10^{15}
E(xa)	10^{18}