

A comparison of electricity and hydrogen production systems with CO₂ capture and storage. Part A: Review and selection of promising conversion and capture technologies

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Abstract

We performed a consistent comparison of state-of-the-art and advanced electricity and hydrogen production technologies with CO₂ capture using coal and natural gas, inspired by the large number of studies, of which the results can in fact not be compared due to specific assumptions made. After literature review, a standardisation and selection exercise has been performed to get figures on conversion efficiency, energy production costs and CO₂ avoidance costs of different technologies, the main parameters for comparison. On the short term, electricity can be produced with 85–90% CO₂ capture by means of NGCC and PC with chemical absorption and IGCC with physical absorption at 4.7–6.9 €/ct/kWh, assuming a coal and natural gas price of 1.7 and 4.7 €/GJ. CO₂ avoidance costs are between 15 and 50 €/t CO₂ for IGCC and NGCC, respectively. On the longer term, both improvements in existing conversion and capture technologies are foreseen as well as new power cycles integrating advanced turbines, fuel cells and novel (high-temperature) separation technologies. Electricity production costs might be reduced to 4.5–5.3 €/ct/kWh with advanced technologies. However, no clear ranking can be made due to large uncertainties pertaining to investment and O&M costs. Hydrogen production is more attractive for low-cost CO₂ capture than electricity production. Costs of large-scale hydrogen production by means of steam methane reforming and coal gasification with CO₂ capture from the shifted syngas are estimated at 9.5 and 7 €/GJ, respectively. Advanced autothermal reforming and coal gasification deploying ion transport membranes might further reduce production costs to 8.1 and 6.4 €/GJ. Membrane reformers enable small-scale hydrogen production at nearly 17 €/GJ with relatively low-cost CO₂ capture.

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Keywords: CO₂ capture; Hydrogen; Electricity; Technology overview; Techno-economic analysis

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Nomenclature

ASU	air separation unit	HSD	hydrogen separation device
ATR	auto thermal reforming	IDC	interest during construction
AZEP	advanced zero emission power plant	IGCC	integrated gasification combined cycle
CCS	CO ₂ capture and storage	ITM	ion transfer membrane
CG	coal gasification	LHV	lower heating value
CHAT	cascaded humidified air turbine	MCFC	molten carbonate fuel cell
CLC	chemical looping combustion	MCM	mixed conducting membrane
COE	cost of electricity	MDEA	methyl diethanolamine
COH	cost of hydrogen	MEA	monoethanolamine
FGD	flue gas desulphurisation	MR	membrane reformer
GT	gas turbine	NGCC	natural gas combined cycle
HAT	humid air turbine	O&M	operating and maintenance
HRSG	heat recovery steam generator	OCM	oxygen conducting membrane

PC	pulverised coal-fired power plant	SMR	steam methane reforming
PEMFC	proton exchange membrane fuel cell	SOFC	solid oxide fuel cell
POX	partial oxidation	ST	steam turbine
PPC	process plant costs	TCR	total capital requirement
PSA	pressure swing adsorption	TIT	turbine inlet temperature
SCR	selective catalytic reduction	TPC	total plant costs
SER	sorption enhanced reforming	WGS	water gas shift
SETS	sorbent energy transfer system	ZEC	zero emission coal
SEWGS	sorption enhanced water gas shift		

1. Introduction

In an energy future driven by greenhouse gas constraints and fossil fuels retaining their dominant role in the world energy supply in the coming decades, CO₂ capture and storage (CCS) enables the production of electricity and hydrogen with strongly reduced CO₂ emissions [1–4]. CCS can be defined as the separation and capture of CO₂¹ produced at stationary sources, followed by transport and storage in geological reservoirs or the ocean in order to prevent its emission to the atmosphere. CCS generally also encompasses fixation of CO₂ in the form of minerals or utilisation for industrial purposes. Many studies have been performed with the purpose to identify promising CCS technologies for fossil fuel conversion into electricity and (to a lesser extent) hydrogen. The level of detail and data quality of these studies varies considerably. Due to the large variety in technologies covered, methodologies used, scales considered and assumptions made on technical performance and economic factors, the results of these studies can in fact not be compared. Few studies compare *series* of conversion technologies with CO₂ capture with respect their energetic and economic performance on a common basis, also accounting for potential developments in state-of-the-art conversion and capture technologies. David and Herzog [5], Simbeck [6], Göttlicher [7,8] and Rubin et al. [9] compared various coal and natural gas-fired power plants with CO₂ capture. Parsons et al. [10] focused on advanced power systems. In the recently accomplished CCP project, various advanced CCS technologies (mainly natural gas-fired) have been analysed in detail, including potential improvements in current capture technology [11]. Hydrogen production from various fossil fuel routes with CO₂ capture have been compared

in [12–14]. However, no complete review of state-of-the-art and advanced technologies for both electricity and hydrogen production with CO₂ capture is available that allow for a comparison of these options on a common basis.

The main objective of this study is to make a consistent techno-economic comparison of electricity and hydrogen production technologies with CCS using coal and natural gas. Therefore, an analysis based on literature review has been performed to collect data on energetic and economic performance of new-built plants and identify promising options. We consider state-of-the-art technologies that in principal could be operated on commercial scale (not considering legal and financial barriers) on the short term (around 2010) and advanced technologies that may become mature for market introduction on the longer term (around 2030).

Options with promising prospects in terms of energy efficiency and costs are further analysed in a chain analysis, incorporating CO₂ transport and storage and distribution/use of energy carriers to the end-user. A chain analysis is the most appropriate methodology to compare energy production costs and CO₂ avoidance costs in a consistent matter, because energy and CO₂ distribution costs can differ strongly among technologies and can be decisive for the performance of a technology. The analysis is performed for the Netherlands and North Sea area, where CCS may play a significant role in reducing GHG emissions, as this region has large geological storage potential and large CO₂ point sources.

Our analysis is presented in two papers. This paper (part A) summarises the review and data normalisation of electricity and hydrogen production technologies and possible combinations with CO₂ capture. On that basis, we discuss promising options and highlight gaps and weak spots in information, which can be useful to set R&D priorities. Paper B continues with the chain analysis of promising options selected in paper A.

¹ Thermal cracking of natural gas into carbon and hydrogen followed by storage or useful application of carbon can also be considered as CCS.

2. Methodology

2.1. Data standardisation and selection

An extensive literature review has been performed to assess conversion efficiency figures and costs of electricity and hydrogen production technologies with and without CO₂ capture. Since each study contains specific assumptions affecting conversion efficiency and capital/energy production costs, data need to be standardised in order to make a fair comparison. We performed the following procedure:

1. *Normalising CO₂ and H₂ pressure.* Conversion efficiency and capital cost figures are corrected for standard CO₂ and H₂ pressure of 110 and 60 bar, respectively. This is done using regression fits of compression energy versus pressure difference derived from Aspen^{plus} flowsheet modelling software. Capital costs of compression as function of capacity are derived from [15].
2. *Indexation.* Costs are converted to €₂₀₀₃ using GDP deflators [16] and annual currency exchange rates (www.oanda.com).²
3. *Normalisation of capital cost figures.* Total capital requirement (TCR) or turnkey project costs consists of various components [17]:
 - Total plant costs (TPC) or overnight construction costs represent the cost to erect the plant (process plant costs or PPC) plus engineering and contingencies.
 - Owners costs are the costs to develop and start up the plant (royalties, preproduction costs, inventory capital, cost of land and site preparation).
 - Interest during construction (IDC)

Not every study includes the same components. If specified, generally TPC is reported and some studies also include owners costs and IDC. For those studies that did not include contingencies and/or owners costs, we used average figures based on data for equal technologies from other studies (contingencies: 10% of PPC, owners costs: 7% of PPC). IDC was calculated considering a construction period of 4 years for coal-fired power plants and 2 years for gas-fired power plants and equal annual expenses. For hydrogen production, we considered a 3 years construction period for large-scale natural gas-fired plants and 4

years for gasification plants, with equal annual expenses.

4. *Scaling of capital cost figures.* Since capital costs depend strongly on the capacities considered, which may vary substantially among studies, we standardised TPC applying a generic scaling relation (Eq. (1)). A base scale of 600 MW_e³ is considered for central power production (with and without capture) and 20 MW_e for decentralised power production with natural gas-fired fuel cells. For hydrogen production, central facilities of 1000 MW_{H2} (LHV basis) and facilities at refuelling stations of 2 MW_{H2} are considered.

$$\text{Costs } A/\text{Costs } B = (\text{Capacity } A/\text{Capacity } B)^R \quad (1)$$

in which R = scaling factor for different conversion technologies

After standardisation, the literature sources are selected on a number of criteria in order to get figures on conversion efficiency and costs. Obviously, preference is given to recent, high-detail and transparent studies, ideally including data on conversion and capture efficiency, investment and O&M cost. For state-of-the-art technologies often a multitude of high-quality references is available, which makes it rather difficult to make a selection on these criteria. Other criteria are then applied to get a value or range in conversion efficiency and investment costs, among which technological maturity (technologies need to be commercially available within considered timeframe) or plant specifications. Less mature, advanced concepts are generally studied in less detail. In these cases, we often have to rely on efficiency forecasts and even less certain cost projections. Investment costs for advanced technologies are generally forecasted values based on technology improvements and do not explicitly account for the effect of technological learning.

2.2. Key parameters for comparison

The different technologies with CO₂ capture are compared on net efficiency (LHV basis), energy production costs and CO₂ avoidance costs. Hydrogen production systems often comprise various energy inputs and outputs: feed, fuel, hydrogen, steam and electricity. The conversion efficiency of these systems

² Most data in literature are presented in US\$. We use the average exchange rate of 2003: 0.885 €/US\$.

³ This capacity represents a typical Dutch coal-fired power plant and a large gas-fired power plant.

is defined as quotient of the sum of energy outputs by the sum of energy inputs, all in primary terms

$$\eta = \frac{E_{\text{hydrogen}} + \frac{E_{\text{el,out}}}{\eta_e} + \frac{E_{\text{th,out}}}{\eta_{\text{th}}}}{E_{\text{feed}} + E_{\text{fuel}} + \frac{E_{\text{el,in}}}{\eta_e} + \frac{E_{\text{th,in}}}{\eta_{\text{th}}}} \quad (2)$$

in which η_e = electric efficiency (0.5) and η_{th} = thermal efficiency (0.9)

Levelised energy production costs (COE and COH for electricity and hydrogen, respectively) are calculated by dividing the sum of annual capital, O&M and fuel costs by the annual energy production. For hydrogen production, also electricity costs or revenues are accounted for. Annual capital costs are derived from the total capital requirement by means of a capital recovery factor, which is a function of the discount rate and economic lifetime. O&M costs generally include costs related to maintenance, (operating) labour, supervision and administration, chemicals and consumables and waste disposal. Some studies include insurance, local taxes and by-product credits (sulphur), but these are not included here. We express O&M costs as a share of TPC.

CO₂ avoidance costs differs from CO₂ capture costs as it accounts for the fact that the plant with CO₂ capture is less efficient (and hence produces additional CO₂)

$$\text{CO}_2 \text{ avoidance costs} = \frac{\text{COE}_{\text{cap}} - \text{COE}_{\text{ref}}}{m_{\text{CO}_2,\text{ref}} - m_{\text{CO}_2,\text{cap}}} \quad (3)$$

in which m = CO₂ emission factor (g/kWh or g/GJ hydrogen) of capture plant (cap) and reference plant (ref). These emission factors include indirect (avoided) CO₂ emissions by electricity use or production.

The choice of the reference system has a significant impact on CO₂ avoidance costs. In paper B, the impact of various reference systems is considered more extensively. In this paper, we apply the most common approach comparing identical plants with and without CO₂ capture (i.e. the baseline varies per technology). This approach gives a clear insight which technologies enable inherent low-cost CO₂ capture.

3. Electricity production technologies

There are many combinations of energy conversion technologies and CO₂ capture technologies conceivable (in some options CO₂ capture is inherent to conversion technology), which urges the need for a transparent classification. The most common classification of CO₂ capture strategies distinguishes post-combustion capture, pre-combustion capture and oxyfuel combustion (see Fig. 1). The principle of post-combustion capture is to remove CO₂ from flue gas after combustion at low pressure and low CO₂-content. In pre-combustion schemes, CO₂ is captured from shifted syngas

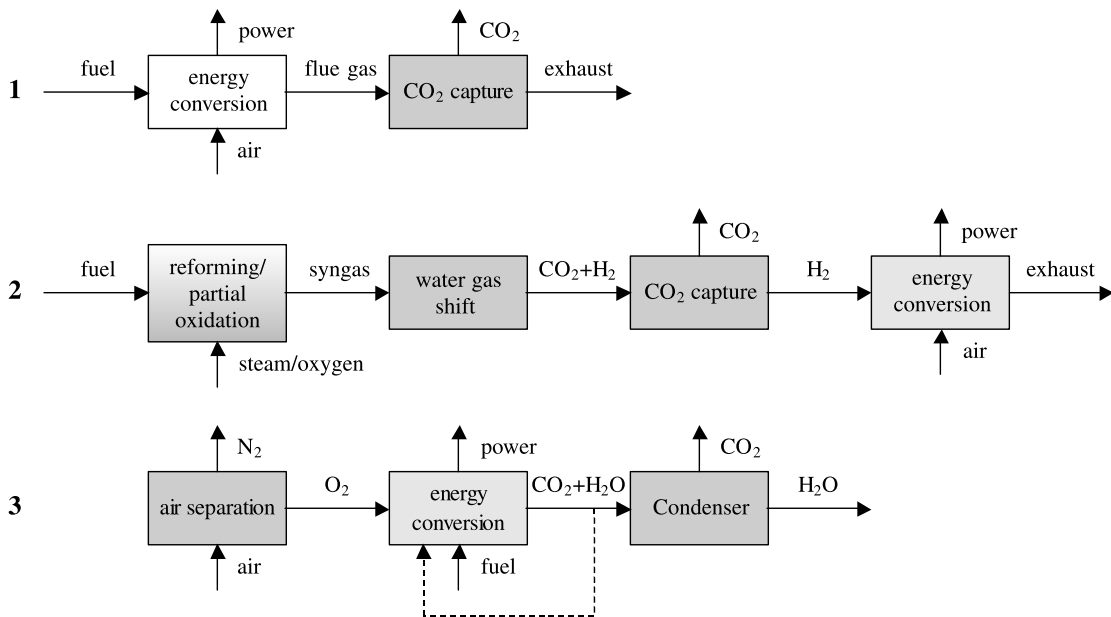


Fig. 1. Principle of post-combustion capture (1), pre-combustion capture (2) and oxyfuel combustion (3). White components are equal to the configuration without capture, light grey indicates modifications to the configuration without capture, dark grey indicate new components.

(consisting predominantly of H_2 and CO_2 at high pressure) produced by reforming (SMR) or partial oxidation (POX) of natural gas or gasification of coal. Hydrogen rich gas is subsequently combusted in adapted turbines. In oxyfuel processes, fuel is combusted in an atmosphere of oxygen with recycled CO_2 and/or steam, resulting in a concentrated stream of CO_2 and steam, which can be separated easily by condensation.

In this study, we discuss these capture options from the perspective of power production technologies, being pulverised coal-fired power plants (PC), natural gas combined cycle (NGCC), integrated coal gasification combined cycle (IGCC) and (high-temperature) fuel cells.

3.1. Pulverised coal-fired power plants

In PC plants, pulverised coal is combusted with excess air in a boiler to produce high-pressure/temperature steam, which is expanded in a steam turbine to generate electricity. The efficiency strongly depends on steam pressure and temperature, which showed an increasing trend over the last decades. Supercritical plants⁴ are state-of-the-art technology, reaching net electrical efficiencies in the range of 41–46% [18–20]. Developments in materials make use of higher steam pressures and temperatures to ultra super critical conditions possible. Efficiency forecasts for 2010 and 2025 are 48–50% and 50–55%, respectively [21].

CO_2 can be captured by means of post-combustion capture and oxyfuel combustion.

3.1.1. Post-combustion capture

The leading technology in post-combustion capture is chemical absorption of CO_2 using monoethanolamine (MEA). This technology is commercially applied to produce high-purity CO_2 from the exhaust of coal and gas-fired boilers, furnaces and turbines. The main applications for high-purity CO_2 are enhanced oil recovery, urea production and in the food/beverage industry. At present, the largest operating unit has a capacity of 800 t CO_2 /day [22]. To put this into perspective, a 500 MW_e PC unit produces circa 8000 t CO_2 /day. MEA is suited for low CO_2 partial pressures as occurring in flue gas. First, the flue gas is cooled and impurities are removed. NO_x and SO_x removal to low concentrations prior to CO_2 capture is

essential, since these components form heat stable salts. In the absorption column, CO_2 reacts chemically with the amine. The CO_2 -rich absorbent is pumped to the desorber, where CO_2 is released by increasing temperature up to 120 °C, for which steam is extracted from the steam cycle. The regenerated absorbent is recycled to the absorber and CO_2 is dried and compressed for transport conditions (typically between 100 and 150 bar).

The major bottleneck is the large steam requirements for CO_2 regeneration (circa 4.2 MJ/kg CO_2 for conventional Econamine FG process [23]) and high investment costs of the capture unit. However, existing processes are improved and new processes are researched/developed. The Econamine FG+ process, which includes a split-flow configuration, an improved solvent formulation and better heat integration, achieves significantly lower energy use [21]. A new solvent based on sterically hindered amines (KS-1) has been commercialised and adopted in an ammonia plant to capture CO_2 from furnace exhaust. Recovery energy of KS-1 has been reduced to 3.2 MJ/kg CO_2 [24]. Further improvements are foreseen in the Econamine and KS-1 processes (stripping steam requirements and amine loading). A recent study indicates that precipitating amino-acid salt solutions with higher CO_2 loading and lower heat of desorption might reduce energy requirements to 2.3 MJ/kg CO_2 , a reduction of 45% in comparison to MEA technology [25]. Investment costs of absorption units can be further reduced by further optimisation of absorber design, e.g. omitting the cooler by integrating this element in the flue gas desulphurisation unit [21]. The application of membrane contactors, serving as contact between flue gas and absorption solvent, enables the application of smaller absorption columns [26].

Alternative options to capture CO_2 from flue gasses are adsorption, low temperature distillation and membranes. Energy requirements of conventional pressure (and temperature) swing physical adsorption, in which CO_2 binds to a solid surface (e.g. zeolites), are prohibitively large [27]. A relatively new concept is the use of high-temperature solid compounds, which react with CO_2 to form a carbonate. In a different reactor, the sorbent is regenerated, producing pure CO_2 . The impact on the conversion efficiency and costs is not quantified at the time this article was written. Low temperature distillation uses a freezing technique to capitalise on the difference between the freezing point of CO_2 and the rest of the flue gas. Göttlicher [8] indicated that CO_2 avoidance costs of distillation are higher than for chemical absorption. The driving force for CO_2

⁴ Steam pressure higher than critical pressure of 220 bar and steam temperatures higher than 560 °C.

separation using membranes depends on the partial pressure difference of CO₂ between permeate and retentate side of the membrane. Using commercially available polymeric membranes results in relatively large energy requirements and CO₂ avoidance costs in comparison to chemical absorption, due to the low driving force as a consequence of the low CO₂ partial pressure in flue gas [7,28,29].

3.1.2. Oxyfuel combustion

In a power plant with oxyfuel combustion, pulverised coal is combusted using (nearly) pure oxygen, which is produced by a cryogenic air separation unit (ASU), although new technologies such as ion transport membranes (ITM) are being worked on. The flue gas, containing primarily CO₂ (the rest being H₂O, Ar, O₂, N₂, NO_x and SO₂) is partially recycled to the boiler to control the combustion temperature. The remaining flue gas is cooled to remove water, compressed, followed by separation of non-condensable gasses (Ar, O₂ and N₂) from CO₂, and is finally boosted to pipeline pressure. The advantage of oxyfuel combustion is that it enables nearly 100% CO₂ capture. In addition, the selective catalytic reduction (SCR) unit and the flue gas desulphurisation (FGD) unit might be omitted as oxyfuel combustion results in low NO_x emissions and the remaining NO_x and SO₂ present in the flue gas could in principal be left for co-storage with CO₂ or could be separated easily⁵ [31,32].

Although there is experience with oxygen fuel combustion in the glass melting and the steel and aluminium industry, the concept of oxyfuel combustion has not been applied in power plants yet. So far only some demonstration tests have been/are conducted. These tests aim to study the effects on plant operation (combustion behaviour, heat transfer⁶) and plant maintenance (fouling, slagging, corrosion). After these issues are resolved, full-scale burners need to be demonstrated successfully. Therefore this option is not considered to be viable for implementation on the short term.

⁵ NO_x and SO₂ 'co-transport and storage' with CO₂ might have serious implications for transport pipelines and storage reservoirs (e.g. enhanced oil recovery) [30]. If required, NO_x and SO₂ can be separated easily from CO₂ by means of an additional distillation unit in the CO₂ purification system [31].

⁶ A recent study indicated that the heat transfer within the furnace is improved as the emissivity of the flue gas in oxyfuel combustion systems is higher due to the higher CO₂ and H₂O content, which results in a higher boiler efficiency [31].

3.1.3. Key performance data

Table 1 shows the range in efficiency and investment costs of all PC options with CO₂ capture. We use the efficiency and investment costs of a typical, modern supercritical coal plant for North western European conditions, integrated with the latest generation post-combustion capture process (Econamine FG+) [21]. The capture penalty of this process is optimistic in comparison to the majority of studies, in which a conventional MEA is applied. Industry projections for a 2020 plant, including improvements in conversion and capture (amine) technology, results in investment cost reductions of 20 and 39%, respectively [21]. Annual O&M costs for PC with capture by means of MEA are between 4 and 6% of TPC [18,20,21].

The table seems to indicate that the efficiency (penalty) and capital costs of PC with oxyfuel combustion are comparable to PC with post-combustion capture. Few studies were accomplished which considered both capture strategies, and these studies show considerable discrepancy in the results. Both Göttlicher [7] and Parsons et al. [10] conclude that the efficiency penalty and additional costs of oxyfuel combustion schemes are lower in comparison to post-combustion capture schemes. Simbeck analysis indicate that PC with oxyfuel combustion causes a higher efficiency penalty, but lower additional costs in comparison to post-combustion capture [6]. IEA GHG studies indicate that the efficiency penalty of both capture options is comparable and the additional capital costs are higher for the oxyfuel combustion plant [21, 31]. Hence, it is unclear whether oxyfuel combustion proves to be a more cost-effective capture option than post-combustion strategies.

3.2. Natural gas combined cycle

In NGCC plants, natural gas is combusted and the hot flue gas is expanded in the turbine, driving the air compressor and a generator. The heat of the gas turbine exhaust is used to raise steam in a waste heat boiler, which is expanded in a steam turbine. The efficiency depends principally on the turbine inlet temperature (TIT). The most advanced H-class turbines reach a TIT exceeding 1400 °C, employing closed loop steam cooling of the gas turbine first two expansion stages and steam reheating [35]. These turbines, reaching net electrical efficiencies of 60%, are currently being demonstrated, but are not expected to operate in fully commercial plants until 2008 [21]. Gas turbine technology is likely to progress further in the coming decades. Systems with higher temperature exhaust

Table 1
Key parameters of PC plants (using bituminous coal) with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Net electric efficiency ^a (%)	Efficiency penalty (% points)	CO ₂ capture efficiency (%)	TCR (€/kW _e) ^b	Sources
(ultra) super-critical PC ^c	Post-combustion (amines)	30–35	8–13	85–90	1720–2490	[6,7,9,18,20,21]
(ultra) super-critical PC 2020	Improved post-combustion (amines)	40	9	85	1520	[21]
(ultra) super-critical PC ^d	Oxyfuel combustion (ASU)	33–36	9–12	90–100	1830–2220	[6,7,31]

^a Including CO₂ compression to 110 bar.

^b Investment costs have been scaled to 600 MW_e assuming a scaling factor of 0.75, derived from PC investment costs of various scales based on engineering economic models to evaluate fossil fuel power plants [9].

^c Many studies consider CO₂ capture at subcritical plants, which are not considered here as state-of-the-art PC plants in North western Europe are generally supercritical.

^d Several other studies have been performed on oxyfuel combustion applied at lignite-fired power plants [33,34]. Lignite generally has a higher moisture content and lower heating value than bituminous coal, which has a significant impact on efficiency and capital costs of the power plant [34]. As our focus is on the Netherlands, where bituminous coal is used for power generation, CO₂ capture at lignite-fired power plants is not further considered.

temperature enable the application of supercritical, once-through heat recovery steam generator (HRSG) units, which will increase the efficiency and reduce investment costs [21]. Other techniques (being) commercialised to increase efficiency are sequential combustion (reheat), recuperation and intercooling to reduce power demand of air compression [36]. These techniques are (partly) applied in (cascaded) humid air turbines ((C)HAT), in which compressed air is humidified in an air saturator and heated using turbine exhaust prior to combustion. In these concepts, the HRSG and steam cycle can be omitted. A recent study shows that the efficiency of a large scale HAT cycle was slightly lower than a standard G-class turbine and investment costs are significantly higher, indicating that this cycle is not advantageous to NGCC (when not integrated in other cycles) [10].

CO₂ capture in NGCC can be performed by means of post-combustion capture, pre-combustion capture and oxyfuel combustion.

3.2.1. Post-combustion capture

Due to the low partial pressure of CO₂ in the flue gasses of a NGCC (about 0.04 bar), chemical absorption using amines (MEA) is the preferred capture technology. Again, technological options can be implemented to reduce the energy penalty caused by conventional chemical absorption, which have been studied extensively for NGCC in two recent studies [21, 37]. These options can be classified into process simplification, process integration and solvent improvement. Cost reductions can be achieved by omitting the flue-gas cooler, using different (cheaper) heat

exchangers and pumps and structured column packing. Process integration includes recycling part of the flue gas to the air compressor to enhance the CO₂ concentration and inserting the amine reboiler tube bundles directly in the HRSG [37]. Integrated designs require further developments in equipment. Flue gas recycling can be performed in existing turbines, but the combustor and fuel and control systems need some modifications [38]. The most optimistic system considered in the CCP study, in which KS-1 solvents are combined with innovative design engineering to achieve high integration between power cycle and capture unit, could be commercially introduced for large-scale application before the end of this decade if aggressively pursued [39].

Apart from absorption, also alternative technologies have been studied [37]. Adsorption processes are relatively expensive due to the low CO₂ loading and cryogenic CO₂ separation is also considered less attractive than absorption. Capital costs savings of a membrane contactor combined with KS-1 solvent integrated versus a conventional absorption and desorption unit are within the uncertainty of the estimates.

3.2.2. Pre-combustion capture

Pre-combustion capture systems encompass syngas production by reforming⁷ and/or partial oxidation⁸ of natural gas (see Section 4 for more details), subsequent water gas shift (WGS⁹), separation of CO₂ and

⁷ CH₄ + H₂O ↔ CO + 3H₂ (ΔH₂₉₈ = 206 kJ/mol).

⁸ CH₄ + ½O₂ ↔ CO + 2H₂ (ΔH₂₉₈ = –38 kJ/mol).

⁹ CO + H₂O ↔ CO₂ + H₂ (ΔH₂₉₈ = –41 kJ/mol).

hydrogen by physical or chemical absorption and combustion of hydrogen in a gas turbine. The main problem with combusting hydrogen in current turbines is that it will result in increased NO_x emissions due to an increased flame temperature. The current technology in gas turbines to reduce NO_x emissions using premixed combustors cannot be deployed for hydrogen combustion due to its combustion characteristics [38,40]. SCR is an alternative, but costs are considered too high for reducing NO_x emissions from levels that arise when combusting pure hydrogen [40]. This makes fuel dilution with steam (from the steam cycle) or nitrogen the most feasible option. In systems where hydrogen is produced by POX using oxygen, nitrogen from the ASU can be injected in the combustor to reduce flame temperature. If nitrogen is present in the hydrogen-rich fuel (when using air as oxidizing agent), changes in the combustor may be smaller than for burning pure hydrogen [38]. A thermodynamic analysis has indicated that the efficiency of combined cycles is only moderately decreased when diluting the fuel with nitrogen or steam [40].

Various levels of process integration are possible in order to decrease energy losses and/or costs:

1. CO shift and CO_2 capture can be performed simultaneously in the so-called sorption enhanced water gas shift (SEWGS) process, in which a WGS catalyst and a special high temperature CO_2 adsorbent are combined in a reactor [41]. As CO_2 is removed from the gas phase, the WGS reaction is shifted to the product side. Since no cooling is required, the energy penalty for cooling/reheat is avoided and the steam is preserved in the hot hydrogen stream, contributing to power generation. This technology is expected to be commercially ready after 2010 [42].
2. In the membrane reformer (MR-CC¹⁰), steam methane reforming is integrated with a hydrogen separation membrane. The membrane, placed inside the reforming tube, continuously withdraws hydrogen from the reaction zone (retentate side). As hydrogen permeates through the membrane, the chemical equilibrium of the reforming and WGS reaction is shifted to the product side according to Le Chatelier's principle, thereby increasing methane conversion. On the permeate site, nearly

pure hydrogen is collected. By using steam/nitrogen as sweep gas, the partial pressure difference across the membrane increases, thereby increasing the hydrogen flux. Simultaneously, the nitrogen or steam in the gas will depress NO_x formation in the combustion chamber. The retentate gas stream leaving the reactor consists mainly of CO_2 and water vapour at feed pressure. CO_2 can be captured easily by condensation of the water vapour.

Membrane material can be metal (composite) such as Pd or Pd alloy, microporous (SiO_2 , C, zeolite) and dense mixed conducting membranes¹¹ (perovskites) [43,44]. Membranes with sufficient selectivity, stability and lifetime still require serious development efforts. MR technology is expected to be commercially ready after 2015 [42].

3.2.3. Oxyfuel combustion

In NGCC systems with oxyfuel combustion, natural gas is combusted with (nearly) pure oxygen. Cycles with a conventional ASU to produce pure oxygen can be distinguished from novel oxyfuel power cycles such as advanced zero emission power plants (AZEP) and chemical looping combustion (CLC), employing membranes and oxygen transport particles to enable stoichiometric combustion with oxygen. Various oxyfuel cycles with ASU and $\text{CO}_2/\text{H}_2\text{O}$ recycling have been proposed and evaluated, among which the Graz, Water and Matiant (combined) cycle. The Graz cycle consists of a high-pressure combustor with steam injection and a recuperated gas turbine integrated with a steam cycle. Both CO_2 and steam are recycled to the combustion chamber [45]. The Water cycle is an oxygen-fired cycle with steam recycle, producing a high-pressure, superheated mixture of mainly H_2O and CO_2 in a gas generator, which is then expanded in a series of (advanced) turbines. A 20 MW_{th} gas generator fuelled with natural gas has been tested and a 500 kW_e power plant is being developed by Clean Energy Systems (CES) [46,47]. In the Matiant or oxyfuel combined cycle, natural gas is combusted in an O_2/CO_2 atmosphere. The reaction product, consisting principally of CO_2 and H_2O , is expanded in an adapted gas turbine and the heat of the turbine exhaust is used to generate steam, which is expanded in a steam cycle [48].

Oxyfuel combustion in Brayton cycles requires gas turbines that use $\text{CO}_2/\text{H}_2\text{O}$ as working fluid, which has different expansion characteristics than nitrogen used in

¹⁰ This concept integrates a membrane reformer with a combined cycle. We use the abbreviation MR-CC to distinguish from a membrane reformer to produce pure hydrogen, which is not (directly) converted in a combined cycle (see Section 4.1.3).

¹¹ Material with high ionic and electronic conductivity.

conventional turbines. The optimal pressure ratio for CO₂ turbines is higher than for conventional turbines, which implies that the efficiency of a CO₂ turbine for typical pressure ratios nowadays is lower than the efficiency of conventional turbines [8,38,48]. Existing gas turbines cannot simply be adapted for CO₂ as working fluid; hence complete new gas turbines (combustor and compressor) need to be developed. Such a development is not likely to occur as long as gas turbine manufacturers do not foresee a significant market to arise [38].

3.2.3.1. Advanced zero emission power plant. The key of the AZEP concept is substitution of the conventional combustion chamber in a gas turbine by a mixed conducting membrane (MCM) reactor, which combines oxygen production, fuel combustion and heat transfer. Compressed air enters the MCM reactor, where oxygen passes through the membrane and is transported to the combustion chamber. The heat of combustion is transferred to oxygen-depleted air, which is expanded in a conventional turbine. Steam is generated from the turbine exhaust and the CO₂/H₂O stream generated in the combustion chamber, which is then expanded in a steam turbine. The use of a conventional turbine is essentially the advantage of AZEP over oxyfuel concepts using gas turbine with CO₂/H₂O as working fluid. The AZEP concept combines well with current technology, requiring minor adaptations in gas turbine (working fluid has lower oxygen content) and HRSG. Obviously, there are still technical challenges in the MCM development.

The AZEP concept is now being developed by a consortium of energy companies. The gas turbine selected for test phase is a 64 MW_e GTX100, with 53% efficiency in CC mode. Also a 400 MW_e V94.3A turbine with an efficiency of 57.9% has been studied, which is further considered in this study. Due to the limited temperature in the MCM reactor (circa 1200 °C), TIT is lower than TIT of most advanced gas turbines, which exceed 1400 °C. By installing an afterburner to increase TIT and accepting a lower CO₂ capture efficiency of 85%, the efficiency loss can be reduced [49].

3.2.3.2. Chemical looping combustion. CLC is an oxyfuel concept, which is still in an early phase of development. The reactor technology is currently being tested in a 10 kW prototype [50]. CLC is based on fuel combustion by means of two separate reactors in order to separate nitrogen from the combustion products. In the reduction reactor,¹² fuel is oxidised by an oxygen

carrier, generally a metal oxide such as iron/nickel oxide. The reduced metal oxide is then returned (in particulate form, using a fluidized bed concept) to the oxidation reactor, where it reacts with oxygen to close the cycle. The oxidation of the metal is highly exothermic and provides high temperature exhaust air for power generation. The metal oxide supplies heat to the endothermic reduction reaction. Both natural gas and syngas can be used as fuel, and the technology can be integrated in various power cycles. In a concept studied by Brandvoll and Bolland, a humid air turbine is used to expand (saturated) air from the oxidation reactor [51]. The metal oxide also supplies heat to the exhaust stream of the reduction reactor (CO₂ and H₂O) in a gas-solid heat exchanger. The exhaust is expanded in a CO₂/H₂O turbine. Yu et al. considered the so-called sorbent energy transfer system (SETS), a CLC concept designed to interface with an existing NGCC with minimal modifications [52]. A comparative study, in which concepts with and without CO₂ expansion turbine have been analysed, indicates that the efficiency is similar [53].

A crucial issue for CLC systems is the maximum reactor temperature at which sorbent stability is not endangered. This temperature in turn determines TIT and thus efficiency. Brandvoll and Bolland assume TIT of 1200 °C in their thermodynamic analysis [51]. Yu et al. [52] set base case reactor temperatures at 900 °C, which requires supplemental gas burning to achieve the specified TIT of a F-class turbine, reducing CO₂ capture efficiency to 50%. Wolf assumes 1200 °C to be the maximum reactor temperature, although it is uncertain whether the oxygen carrier particles can withstand such high temperatures [53].

3.2.4. Key performance data

Table 2 shows the range in efficiency and investment costs of all natural gas-fired electricity options with CO₂ capture. Many extensive studies have been performed on NGCC with post-combustion capture. The observed range in costs and efficiency data is primarily caused by turbine type and capture process considered. Annual O&M costs for NGCC with post-combustion capture lie between 3 and 6% of TPC [10, 18,20,21,42].

For the short term, we use data on performance and costs from a recent, detailed study considering two GE 9FA frame turbines integrated with chemical absorption using Econamine FG + solvent [21]. On the longer term, (advanced) post-combustion, pre-combustion and oxyfuel combustion technology are candidates for high-efficient, low-cost electricity production with strongly

¹² Reduction/oxidation refers to the state of the oxygen carrier.

Table 2
Key parameters of natural gas-fired power plants with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Net electric efficiency ^a (%)	Efficiency penalty ^b (% points)	CO ₂ capture efficiency (%)	TCR (€/kW _e) ^c	Sources
<i>Post-combustion capture</i>						
NGCC (F-class) ^d	Chemical absorption	43–50	6–12	85–90	700–1010	[9,18,20,21,55,61]
NGCC (G/H-class) ^d	Chemical absorption	48–53	5–11	85–90	780–1770	[6,10,18,21,42,48,62]
NGCC 2020	Improved chemical absorption	55	6	85	650	[21]
<i>Pre-combustion capture</i>						
SMR/POX/ATR-CC ^e	Chemical/physical absorption	43–50	8–13	85–90	900–1470	[6,10,20,48,55,62–64]
ATR-CC	SEWGS	47–48	9–10	90	980–1060	[42]
MR-CC ^f	Hydrogen membrane	51–53	5–6	100	940	[42,55]
<i>Oxyfuel combustion</i>						
Graz cycle ^g	ASU	39–45	13–19	100	–	[54,55]
Water cycle ^h	ASU	40–62	10–13	100	820–880	[10,47,54,55,65,66]
Oxyfuel (Matiant) CC	ASU	44–48	11–13	90–100	1060–1250	[6,31,48,55]
AZEP ⁱ	Oxygen membrane	50–55	2–8	85–100	~700 to 900	[49,55,57–59]
CLC-CC + CO ₂ turbine ^j	Oxygen carrier	51–54	3–5	100	~900 to 1300	[51,53,55,60]
CLC-CC ^k	Oxygen carrier	50–53	3–5	50–100	~500	[52,53,67]

^a Including CO₂ compression to 110 bar.

^b The efficiency penalty refers to the difference with a standard NGCC.

^c As gas turbines are available in certain sizes, the scale for NGCC cycles is determined by the number of turbines installed and can in fact not simply be readjusted to the scale desired. However, capacities of NGCC cases studied are generally between 300 MW_e (single turbine) and 800 MW_e (2 turbines), which cause significant cost differences. Therefore, investment costs have been scaled to 600 MW_e using a scaling factor of 0.9 derived from turnkey NGCC prices for different capacities given in [68,69].

^d The range in efficiency forecasts is depending primarily on the capture process, ranging from conventional MEA (high efficiency penalty) to KS-1 process (low efficiency penalty). The highest level of integration is represented by the so-called BIT option (best integrated technology), combining process simplification and integration with use of KS-1 solvent. This results in an efficiency penalty of 5% points [42]. Note that the TCR estimate of 1770 €/kW_e is atypical.

^e These figures are representative for a set of pre-combustion configurations, integrating various syngas production technologies with WGS, CO₂ capture by means of absorption and H₂ combustion in various turbine classes.

^f The highest efficiency is forecasted for a MR-CC with a dense hydrogen mixed conducting ceramic membrane (HMCM) and nitrogen/steam as sweep gas (costs given in table refer to this configuration) [43]. Simulations of a Pd based MR-CC with steam extracted from the steam turbine for sweep gas application results in a net electric efficiency of 51% [55].

^g The efficiency range in brackets is caused by the assumed condenser pressure, ranging from 0.1 to 0.9 bar [54].

^h Many Water cycle configurations have been studied. Generally, gas is first expanded in a high-pressure, high-temperature steam turbine. After expansion, the gas mixture is reheated and expanded in an intermediate and low-pressure turbine (which can be either a steam or gas turbine). The large range in efficiency is mainly caused by the assumed TIT of the turbines, which can be increased by using special alloys and steam cooling techniques applied in gas turbines. The lowest value represents a state-of-the-art Rankine power cycle with high-pressure TIT of 900 °C and intermediate-pressure TIT of 1328 °C [54]. The most optimistic configuration assumes a high-pressure TIT of 816 °C and intermediate and low pressure TIT of 1649 °C (double reheat) and expansion of high-pressure N₂ from the ASU [47]. Such turbines, based on turbine development goals of the US-DOE Vision 21 program, require considerable R&D efforts. The efficiency penalty and costs are representative for a configuration with single reheat and intermediate-pressure TIT of 1200–1400 °C [10,55,66].

ⁱ Represents efficiency projections of a 400 MW_e system. Additional investment costs estimations range between 400 €/kW_e in the short term to less than 250 €/kW on the long term [57]. A more detailed techno-economic evaluation of 30–40 MW_e simple cycle AZEP concepts shows that the specific investment costs for AZEP systems are 120–170% higher than for the reference GT [58]. The concepts with afterburner are in the low range, although CO₂ avoidance costs of systems with and without afterburners are very close. The costs related to CO₂ capture are lower for the original combined cycle configuration AZEP [58], also considering the difference in scale of the 30–40 MW_e simple cycle systems and 400 MW_e combined cycle system.

^j The high efficiency forecast represents a 50 MW_{th} natural gas-fired cycle with HAT turbine and CO₂/steam turbine to expand the exhaust of the reduction chamber. Crucial issues on oxygen carrier durability, mechanical properties and fuel conversion have been idealised [51]. A confidential thermo-economic analysis revealed that a similar CLC concept of 250 MW_e might reach a net electric efficiency of 50% when using natural gas containing 15% nitrogen [70]. The capital costs are rough estimates derived from individual costs of reactors, compressors and turbines [60,70].

^k The net electric efficiency of the SETS cycle (=CLC-CC) ranges from 53% for a capture efficiency of 50 to 50% for 100% CO₂ capture [52]. Investment costs of the SETS system are estimated to be 10% (!) higher than the total investment costs of an NGCC [67].

reduced CO₂ emissions. For the post-combustion options, we consider an advanced turbine ($\eta_e \sim 62\%$) integrated with further improved processes based on Fluor's Econamine FG+ [21]. Further cost reductions in that study in both NGCC and capture unit are assumed to be 20 and 32% of state-of-the-art figures, respectively. Pre-combustion capture systems comprising syngas production, WGS and CO₂ capture by absorption do not seem to offer efficiency and costs advantages in comparison to post-combustion schemes. However, advanced pre-combustion technologies such as MR-CC seem promising. We use efficiency and costs forecasts from a MR-CC with hydrogen mixed conducting ceramic membranes [42,43]. The oxyfuel concepts do not show a clear winning option at present due to the large range in conversion efficiency and lack of accurate costs estimations. The conversion efficiency of Matiant (oxyfuel) and Graz cycle with conventional ASU are clearly lower than AZEP and CLC. The Water cycle entails large technological uncertainty, which causes the wide range in Table 2. It may reach conversion efficiencies over 50%, provided that high temperature turbine machinery is developed. Results of thermodynamic analyses comparing various gas-fired options show that the Water cycle (with moderate assumptions on turbine development and oxygen production by ASU) does not offer higher conversion efficiency than advanced oxyfuel concepts such as AZEP and CLC, nor post and pre-combustion strategies¹³ [10,54,55]. Parsons et al. [10] show that the Water cycle has higher investment costs than NGCC with chemical absorption, although current estimations quoted by CES are somewhat lower [56]. Since various studies indicate that the Water cycle do not seem to offer energetic nor economic advantages above NGCC with chemical absorption and efficiency and investment costs of advanced systems are highly uncertain, the water cycle is not further considered. We consider AZEP and CLC as the most promising advanced oxyfuel technologies. For AZEP, we use conversion efficiencies calculated for the configuration with 100% CO₂ capture [49]. The capital costs are estimated from a cost quotation [57] and results of cost calculations [58,59]. For CLC, we assume a conversion efficiency of 51% based on [55] and use a rough cost estimation quoted in [60].

¹³ When comparing the Water cycle with AZEP, it would be fair to take into account the efficiency improvement in oxygen production by membranes in the Water cycle.

3.3. Integrated coal gasification combined cycle

An IGCC comprises gasification of solid or liquid fuels (coal, oil residues, biomass) to produce raw synthesis gas, which is then cleaned to remove contaminants such as particulate matter and sulphur. The resulting fuel gas is used to fire a combined cycle. Gasifiers applied for IGCC fuelled with (bituminous) coal are typically oxygen-blown entrained flow gasifiers, operating at pressures between 30 and 70 bar and temperatures from 1000 up to 1500 °C. Several technologies are available (E-gas, Prenflo, Shell, Texaco), the main differences being the coal feed (either dry or slurry feed), number of gasifier stages (one or two) and syngas cooling (water quench or heat exchangers). IGCC is not as well developed as PC; today only four coal-fired units are in operation, although there are several other IGCC units using coke and residual oil as feedstock and more units are planned [71]. The existing coal-fired IGCC units have a capacity between 250 and 335 MW_e and a net electric efficiency between 38 and 45% [72]. A major bottleneck for large-scale commercial application is the relatively high investment costs in comparison to PC. Moreover, IGCC is a technically more complex system than PC, leading to a relatively low availability and poor operational flexibility. In order to increase plant availability to a level of 85–90%, a spare gasifier might be required, which will increase total investment costs.

CO₂ capture at IGCC plants can be performed by means of post-combustion capture, pre-combustion capture and oxyfuel combustion.

3.3.1. Pre-combustion capture

In this concept, shift reactor(s) are installed after the gas cleaning section to convert CO present in syngas into CO₂ and H₂ by adding steam. Hydrogen is mixed with nitrogen/steam and fed to an adapted gas turbine. The high partial pressure of CO₂ after the shift reactor makes physical absorption the most appropriate (commercially available) CO₂ capture technology, because this process is less energy intensive than a chemical absorption process. In physical absorption processes, CO₂ is recovered from the absorbent (e.g. Selexol) by reducing the pressure in flash drums, which is less energy-intensive than stripping CO₂ using heat as applied in chemical absorption.

Various options have been proposed to improve the energetic and/or economic performance of the capture process. Alternatives for physical absorption are SEWGS or inorganic membranes for separation of CO₂ (or CO) and hydrogen. Membranes could

eventually be integrated with the shift reactor (water gas shift membrane reactor). Although not further considered in this study, cost reductions can be achieved by co-capture and storage of CO₂/H₂S ('acid gas'), a strategy applied in parts of North America to dispose CO₂ and H₂S present in natural gas. In an IGCC, H₂S is normally removed from the syngas prior to CO₂ removal and sent to a sulphur recovery system (Claus and SCOT unit), where it is converted into elemental sulphur. In a co-capture/storage system, the sulphur recovery unit is omitted and H₂S and CO₂ are compressed and stored together.

3.3.2. Post-combustion capture and oxyfuel combustion

In a configuration with post combustion capture, a chemical absorption unit is installed after the gas turbine. The advantage above the pre-combustion strategy is that no shift reactor and no modifications to the gas turbine are required. The oxyfuel strategy encompasses a larger ASU, which also provides oxygen for syngas combustion. This requires the development of new turbines designed to operate on CO₂/H₂O as working fluid. The Water cycle fuelled with syngas (IG-Water) can be considered as an IGCC oxyfuel strategy, although this technology is strictly spoken a Rankine cycle.

3.3.3. Key performance data

The key performance data for IGCC concepts with various CO₂ capture strategies are shown in Table 3. Various studies comparing IGCC with post, pre and oxyfuel combustion clearly show that pre-combustion by physical absorption is the most efficient and least capital-intensive CO₂ capture technology [6,7,73], which will be further considered here. The large range in efficiency and capital costs for IGCC can be explained mainly by the different IGCC configurations, i.e. gasifier technology (slurry/dry feed), syngas cooling (quench/heat exchangers), gas turbines (F,G and H-class), individual performance of each element and level of integration between the elements. Also the coal type (rank) has a significant impact on cold-gas efficiency and capital costs, with decreasing efficiencies and increasing costs with lower coal ranks [74].

We use a range in investment costs and performance data representing Texaco and Shell gasification technology integrated with two GE 9FA turbines¹⁴ using Australian bituminous coal as feedstock [75]. The

Shell technology represents high-performance, capital-intensive dry feed gasifiers with a heat recovery boiler, whereas the Texaco concept represents less efficient and less capital-intensive slurry feed gasifiers with product gas cooling by water quench.¹⁵ The efficiency figures are relatively conservative in comparison to other IGCC forecasts, which might be explained by the relatively high production of ungasified carbon and the level of heat integration [75]. However, a comparison with other studies considering the same IGCC configuration (same gasifier, syngas cooling and turbines) without CO₂ capture revealed very similar results [74], so the data used here are considered reliable and representative.

On the longer term, significant improvements are foreseen in the gasifier, gas turbine and air separation unit. We consider an advanced IGCC envisioned for the year 2020 by Foster Wheeler, featuring dry-feed, two stage entrained flow gasification, an improved (H) turbine, and ion transfer membrane air separation replacing conventional, energy-intensive cryogenic ASU [75]. This configuration does not account for further improvements in CO₂ capture technology and might therefore prove slightly conservative. A case study on a coke gasification IGCC plant indicates that a novel cryogenic technology for CO₂ separation promises 16% reduction in CO₂ avoidance costs versus a base case with Selexol [77]. Various studies on the techno-economic impact of high-temperature membranes to separate CO₂ from H₂, eventually integrated with water gas shift reactor, indicate that the IGCC conversion efficiency can be increased and investment costs reduced in comparison to physical absorption [7,29,78].

The Water cycle fuelled with syngas might prove a promising alternative to IGCC with pre-combustion capture. We use data for a configuration integrating an E-gas gasifier with a set of advanced turbines and single reheat. The intermediate pressure turbine has a TIT of 1427 °C, the rated temperature of H-turbines [10]. Although the efficiency is somewhat conservative in comparison to other projections [46,47], we choose to

¹⁴ H-class turbines for hydrogen rich gas as in IGCC are not expected to be commercially available on the short term [75].

¹⁵ Shell specified 2×50% capacity gasifiers and ChevronTexaco specified 4×33% capacity gasifier in order to achieve a capacity factor of 85% [75]. In various studies on IGCC costs, Texaco-based IGCC are designed with a spare gasifier in order to achieve 90% target availability [74]. Shell's membrane cooled gasifiers have a higher availability than refractory lined Texaco gasifiers and might therefore not need a spare gasifier. In case an additional spare gasifier is required for the Shell case, TPC (€/kWe) increase with circa 10% [74]. According calculations performed by Kreutz et al. [76] for a plant with quench cooling, COE increase with 10–12% when accounting for a spare gasifier.

Table 3
Key parameters of IGCC plants (using bituminous coal) with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Net electric efficiency ^a (%)	Efficiency penalty ^b (% points)	CO ₂ capture efficiency (%)	TCR (€/kW _e) ^c	Sources
IGCC (E-gas, Prenflo, Shell, Texaco) ^d	Pre-combustion (physical absorption)	32–43	5–9	85–90	1500–2170	[6,7,9,10,18,20,73,75,76,79,80]
IGCC (Prenflo)	Pre-combustion (MWGS)	40–44	7–8	80	1440–1500	[7,78]
IGCC 2020	Pre-combustion (physical absorption)	43	6	85	1500	[75]
IGCC (Texaco)	Post-combustion (chemical absorption)	37–38	8–10	90	2290–2430	[6,73]
IGCC (Texaco)	Oxyfuel combustion (ASU)	39	8	90	2280	[73]
IG-Water (E-gas, Texaco) ^e	Oxyfuel combustion (ASU)	32–55	8	100	1420–1550	[10,46,47]

^a Including CO₂ compression to 110 bar.

^b The efficiency penalty refers to the difference with a standard IGCC.

^c Due to the limited number of IGCC plants constructed, it is hard to assess a scaling factor for IGCC. We assume a scaling factor of 0.8 used by Kreutz et al. [76] to scale investment costs to 600 MW_e.

^d An IGCC plant with a Texaco gasifier and F-class turbine represents the lower value in the efficiency range. The upper value in efficiency range represents E-gas/Shell gasifiers integrated with G/H turbines. The efficiency penalty and additional investment costs caused by CO₂ capture for Texaco gasifiers are lower in comparison to Shell gasification technology, since no additional steam must be taken from the steam cycle required for CO shift, as the syngas itself contains sufficient steam [75].

^e With current technology (series of conventional steam turbines), a net electric efficiency of 32% can be achieved [46]. When applying a high-pressure, high-temperature steam turbine (103 bar, 816 °C) followed by an intermediate and low pressure, high-temperature turbine with relatively high TIT (26 bar, 1427 °C and 3 bar, 1427 °C), net efficiencies in the order of 48–51% can be reached, a decrease of 3% points in comparison to Water Cycle without CO₂ compression. These turbines are not commercially available for power generation, but have been used in aerospace applications. The most advanced concept, assuming TIT of 1649 °C based on turbine development goal of the US-DOE Vision 21 program and expansion of high-pressure nitrogen from the ASU, may reach an efficiency of 53–55% [46,47]. Further improvements can be realised by use of oxygen transport membranes. The efficiency and investment costs of the oxygen plant (representing 10–15% of IGCC investment costs) can be reduced by 37 and 35%, respectively, resulting in IGCC efficiency increase of 2% and investment cost decrease of 7% [75,81].

use data from this study, as it contains transparent data on both efficiency and investment costs.

Annual O&M costs are between 3.5 and 5% of TPC for IGCC with physical absorption [10,18,20,75] and approximately 3.5% for IG-Water plants [10,46].

3.4. Fuel cells

Fuel cells convert chemical energy of a gaseous fuel directly into electricity and heat. Fuel is oxidised electrochemically, which imply lower energy losses than direct combustion. Fuel cells consist of an electrolyte layer between anode and cathode, functioning as a membrane permeable for ions. The anode and cathode are connected to close the circuit, i.e. electrons generated at the anode flow to the cathode. Fuel cells are generally classified by the electrolyte material and operating temperature. Low-temperature fuel cells include the alkaline fuel cell (AFC), phosphoric acid fuel cell (PAFC) and proton exchange membrane fuel cell (PEMFC). Molten carbonate fuel cell (MCFC) and solid oxide fuel cell (SOFC) are high-temperature fuel

cells. MCFC and SOFC are most appropriate for stationary power production (from a few hundred kilowatt up to a few megawatt), due to their high electrical (and cogeneration) efficiencies¹⁶ [83,84]. The efficiency of SOFC and MCFC can be further increased by integration with gas turbines (MCFC or SOFC-GT hybrids).

Various MCFC and SOFC demonstration units with typical capacities of a few hundred kilowatt up to a few megawatt have been operated successfully and more tests are planned. In 1997 a pilot project started with a 100 kW SOFC cogeneration plant in the Netherlands, which achieved 46% electrical efficiency and a 25%

¹⁶ SOFC (hybrids) is the ideal technology for decentralised power (and heat) generation, due to their high efficiency at modest capacities, as a consequence of their modular design. This makes SOFC also suited for offshore power generation with CO₂ capture and injection into (nearly) depleted oil or gas fields or aquifers [82]. Another possibility is to install SOFC (hybrids) on top of a natural gas field, extracting gas to fuel the SOFC and injecting CO₂ into the gas field to enhance gas recovery [68]. In this way, the need for natural gas and CO₂ transport is avoided.

thermal efficiency. In California, the world's first SOFC-GT hybrid (220 kW) has demonstrated 53% electrical efficiency [84]. Several companies are at the threshold of introducing MCFC and SOFC modules at the market.

High-temperature fuel cells can be fuelled with hydrogen, syngas or natural gas. Due to the high temperature at which SOFC and MCFC are operated, reforming of natural gas and the subsequent shift can be performed in the fuel cell itself. In SOFC/MCFC configurations fuelled with natural gas or syngas, CO₂ capture can be performed after the fuel cell ('post-fuel cell capture'). Alternatively, hydrogen can be produced by reforming/partial oxidation of natural gas or coal gasification upstream the fuel cell.¹⁷ CO₂ can be removed after syngas is shifted by means of physical solvents, membranes or adsorbents, also referred to the 'pre-fuel cell CO₂ capture' strategy. Below, we discuss various capture options accounting for the specific features of MCFC and SOFC. Note that CO₂ capture at fuel cells are in fact oxyfuel concepts, since fuel is converted in an atmosphere without nitrogen.

3.4.1. Molten carbonate fuel cell

The molten carbonate fuel cell has an electrolyte of alkali carbonates, which are fluid at the fuel cell operating temperature of ~650 °C. At the cathode, O₂ and CO₂ and 2 electrons form CO₃²⁻ ions. These ions are transported across the electrolyte to the anode, where they react with hydrogen to form H₂O, CO₂ and 2 electrons. CO₂ is generally recycled from the anode to the cathode. A possible configuration to integrate CO₂ capture is to burn the anode exhaust with air, recirculate the combustion flue gas rich in CO₂ to the cathode and use the heat of the cathode exhaust (which contains only ~5 mol% CO₂) to generate steam in a HRSG required for CO₂ capture by chemical absorption [85, 86]. The disadvantage of this configuration is that CO₂ rich anode gas is diluted with mainly nitrogen. This can be avoided by installing a MCFC downstream to a gas turbine or coal-fired power plant, in which CO₂ from the exhaust is used as reactant at the cathode and transferred across the electrolyte, resulting in a concentrated CO₂ stream on the anode side [85,87]. CO₂ is then captured from the anode exhaust and the remaining H₂, CO (and unconverted CH₄) can be burned in a catalytic burner with cathode exhaust or

redirected to the power cycle upstream of the MCFC. These MCFC hybrids are not further considered here as the energetic and/or economic performance of a greenfield plant has not been assessed.

3.4.2. Solid oxide fuel cell

Solid oxide fuel cells have ceramic electrolytes (generally yttria-stabilised zirconia), which require temperatures around 1000 °C to achieve sufficient ionic conductivity. At the cathode, oxygen atoms are ionised and transported by the electrolyte to the anode, where oxygen ions react with hydrogen. The cathode exit stream is oxygen-depleted air and the anode exit contains principally CO₂, CO, H₂O, and H₂ (when using natural gas as fuel). In many of the proposed schemes, the fuel cell is integrated with a gas turbine. In these hybrid configurations, compressed air is fed to the fuel cell cathode and the hot cathode exhaust is expanded in a gas turbine.

High CO₂ concentration in the anode exhaust makes SOFC an interesting target for CO₂ capture. If CO₂ capture is required, the anode and cathode exhaust need to be separated by means of a seal in order to prevent CO₂ rich anode exhaust to be diluted with nitrogen. CO₂ can be captured from the anode offgas by means of cryogenic separation, chemical absorption or shifting the anode exhaust for physical absorption. The H₂(/CO) rich syngas can be recirculated to the anode inlet or can be burned with cathode outlet e.g. to drive a gas turbine. Another possibility is to oxidise the anode exhaust gas with pure oxygen in an afterburner, thereby increasing CO₂ concentration for capture. The afterburner can exist of a burner fuelled with pure oxygen [88], an oxygen conducting membrane (OCM) reactor [89], a second SOFC (electrochemical afterburner) [82], or a water gas shift membrane reactor (WGSMR) afterburner [90]. In an OCM afterburner, oxygen ions and electrons are simultaneously transported over the membrane, whereas in an SOFC the electrons are transported in an external circuit, thereby generating additional power. In a WGSMR, H₂ in the anode exhaust permeates through the membrane to be oxidised with the cathode exhaust.

The ZEC (zero emission coal) process is a highly innovative cycle, which combines a hydrogasifier, carbonation–calcination reactors and a pressurised SOFC. Coal is gasified into a gas consisting of mainly methane, using hydrogen and water. Methane is reformed to produce hydrogen through the use of a carbonation reaction, which converts CaO and CO₂ into CaCO₃ and supplies energy required to drive the reforming reaction. Hydrogen is partly recycled to the

¹⁷ Sulphur content in syngas produced by coal gasification needs to be reduced further after the gas clean up section (by means of a ZnO bed) to acceptable levels for use in fuel cell [10].

gasifier and the rest is used to produce electricity in an SOFC. The process heat from the SOFC is used to regenerate CaO by calcination of CaCO₃, thereby producing CO₂ that can be stored [91]. Although very high efficiencies are claimed, this technology still requires large developments to achieve the process integration that is crucial for the technical and economical performance.

3.4.3. Key performance data

Table 4 gives the techno-economic data of all MCFC and SOFC configurations with CO₂ capture. There is a large variation in both conversion efficiency and cost projections of fuel cell (hybrids). For fuel cell hybrids

fired with coal, we consider the IG-SOFC-GT analysed in the NETL study [10], as this is the most extensive study with most updated cost estimations of fuel cells. Although the ZEC concept promise a high-efficient technology, we do not further include it in our analysis, as projected capital costs are still relatively high. Hence forecasted COE are not structurally lower than COE for other advanced gasification concepts (even at higher coal prices of 3 €/GJ). Natural gas-fired SOFC-GT systems are mainly foreseen for decentralised power generation. We use data for a 20 MW_e system with a 2nd SOFC and afterburner assessed by Lokurlu et al. [92], the only analysis available on decentralised SOFC-GT systems with capture containing data on

Table 4
Key parameters of SOFC and MCFC plants with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Net electric efficiency ^a (%)	CO ₂ capture efficiency (%)	TCR (€/kW _e) ^b	Sources
<i>Fuel cells fired with coal</i>					
IG-MCFC-ST	Pre-fuel cell (shift, physical absorption)	35	68	3080	[86]
IG-MCFC-ST	Post-fuel cell (chemical absorption)	36	76	2720	[86]
IG-SOFC-GT ^c	Various strategies	45–50	90	1670–1760	[6,10]
ZEC (IG-SOFC-ST)	Pre-fuel cell (chemical adsorption)	71	100	1910	[91]
<i>Fuel cells fired with natural gas</i>					
MCFC-ST	Pre-fuel cell (shift, physical absorption)	49	58	2060	[86]
MCFC-ST	Post-fuel cell (chemical absorption)	55	91	1870	[86]
SOFC-GT ^d	Post-fuel cell (chemical absorption)	46–69	90	990–1600	[6,10,85,86]
SOFC-GT	Post-fuel cell (shift, physical absorption)	69	90	–	[85]
SOFC-GT ^e	Post-fuel cell (afterburner)	59–67	80–100	1530–1870	[55,85,92–94]

^a Including CO₂ compression to 110 bar.

^b The capacity of coal-fired systems is in the range 500–750 MW_e, whereas natural gas-fired systems range from 10–500 MW_e. For the coal-fired systems, we apply a scale factor of 0.9 (due to modular character of fuel cells, the scaling relation is expected to be less strong as IGCC). Investment costs of natural gas-fired systems are not corrected for scale, as the scaling relation for such hybrid systems over such a wide capacity range is unknown and the cost-breakdown of investment costs is usually not given.

^c Parsons et al. [10] consider a coal gasifier, which produces syngas for the SOFC anode and a gas turbine. Syngas to the gas turbine enters a hydrogen separation device (membrane), where CO is shifted and CO₂ removed, after which H₂ is sent to the gas turbine. Anode exhaust is combined with CO₂ from the HSD and burned in a catalytic combustor. This configuration results in 50% efficiency versus 56% without capture.

^d Recent study performed by Parsons I&TG indicated that the efficiency of a 560 MW_e cascaded humidified advanced turbine-SOFC hybrid dropped from 66% without CO₂ capture to 46% when CO₂ is recovered from the flue gas by means of MEA [79], which is considerably lower than the forecasted 60% for a similar SOFC-gas turbine configuration with post-fuel cell CO₂ capture analysed in earlier studies [6,86] or 69% forecasted in [85]. The main difference lies in the amount of natural gas that is routed to the duct burner to generate steam for CO₂ regeneration, which is much higher in the Parsons study.

^e Efficiency data cover all afterburner concepts (2nd SOFC, OCM and WGSMT). A recent thermodynamic analysis of these three afterburner technologies integrated with a 20 MW_e SOFC-GT revealed that applying a 2nd SOFC results in the highest efficiency. This configuration requires a larger surface area to oxidise the remaining fuel than an OCM afterburner [93]. This is confirmed by an analysis performed for a 20 MW_e hybrid system, which also accounted for costs [92]. By installing an additional burner after the 2nd SOFC to increase TIT, capital costs are minimised at the expense of a lower efficiency and CO₂ recovery [92], resulting in lower COE in comparison to the standard configuration.

Table 5
Key parameters of electricity production technologies with CO₂ capture

Feedstock	Conversion technology	CO ₂ capture technology	Net electric efficiency (%)	CO ₂ capture efficiency (%)	TCR (€/kW)	O&M (%)
<i>State-of-the-art technologies</i>						
Coal	PC	Post-comb (MEA)	35	88	2080	5.8
	IGCC	Pre-comb (Selexol)	32–35	85	1770–2170	4.8–5.2
Natural gas	NGCC	Post-comb (MEA)	47	85	920	4.3
<i>Advanced technologies</i>						
Coal	Adv. PC	Improved post-comb (MEA)	40	85	1520	6.5
	Adv. IGCC	Pre-comb (Selexol)	43	85	1500	5
	IG-Water	Oxyfuel (ASU)	41	100	1530	3.7
	IG-SOFC-GT	Various (membrane/cat. combustor)	50	90	1760	3.3
Natural gas	Adv. NGCC	Improved post-comb (MEA)	55	85	650	4.8
	MR-CC	Pre-comb (HMCM)	53	100	940	4 ^a
	CLC	Oxyfuel (separate combustion)	51	100	900	4 ^b
	AZEP	Oxyfuel (oxygen membrane)	50	100	900	4 ^b
	SOFC-GT (20 MW _e)	Oxyfuel (electrochemical afterburner)	59	80	1530	3

^a O&M costs for the MR-CC are 5.2% of TPC versus 6% for NGCC with advanced post-combustion capture [42]. We corrected the O&M costs for the MR-CC so that it corresponds in proportion to the value we adopted for advanced NGCC with post-combustion capture.

^b The uncertainty in investment costs of AZEP and CLC justifies the use of a more aggregate estimate of O&M costs, which is set at 4% of TPC. Variable O&M costs for single cycle AZEP were set at 4% of TPC [58].

both efficiency and costs. Annual O&M costs for SOFC-GT systems are between 3 and 3.5% of TPC [10,79,86].

3.5. Overview electricity production technologies

Table 5 summarises key data of the most promising state-of-the art and advanced electricity options with CO₂ capture and compression to 110 bar.

4. Hydrogen production technologies

Steam methane reforming is the most common and developed technology used for hydrogen production at large scales. It is likely to remain a dominant large-scale hydrogen production technology in the nearby future. Partial oxidation is the most appropriate technology to produce hydrogen from heavier feedstocks such as heavy oil residues and coal [95].

4.1. Hydrogen production from natural gas

4.1.1. Steam methane reforming

In fired tubular reformers, steam and natural gas are reacted over a nickel-based catalyst to produce syngas at temperatures of about 800–900 °C. In order to sustain

this endothermic reaction, heat is supplied to the reforming reactor by burning part of the natural gas in a furnace. Syngas is cooled and then shifted in the WGS reactor. In older plants, CO₂ is subsequently removed by means of a chemical absorption unit. Modern hydrogen plants apply pressure swing adsorption (PSA) to separate hydrogen from the other components, which produces higher quality hydrogen (99.999% against 95–98% for scrubbing systems) at feedstock pressure (circa 25 bar). Purge gas from the PSA is generally redirected to the furnace. Depending on the design, steam or electricity is imported/exported. The conversion efficiency of large-scale SMR varies between 74 and 85%¹⁸ [12,13,95,96].

Relatively few studies have been performed to assess efficiency loss and costs associated with CO₂ capture at SMR plants. CO₂ can be captured from the furnace off-gas by means of a MEA unit ('post-combustion' strategy), which would require large steam consumption due to the low CO₂ partial pressure. Alternatively, CO₂ can be captured by installing an MDEA unit between the shift reactor and the PSA unit. Although this has not been applied in hydrogen plants

¹⁸ All electricity/steam input/outputs converted into primary terms.

to date, MDEA is proven technology. When high-purity hydrogen is not required (when applied in conventional turbines or heating purposes), the PSA unit can be omitted (representing the older hydrogen plants).

4.1.2. Autothermal reforming

Steam methane reforming can be combined with partial oxidation, a process referred to as autothermal reforming. Basically, steam and oxygen (or air) are injected into the reformer, causing the reforming and oxidation reaction to occur simultaneously. The exothermic POX reaction, in which the feedstock reacts (catalytically) with oxygen in sub-stoichiometric conditions, provides heat to endothermic SMR reaction. Hence, the use of ATR allows for higher CO₂ recovery in comparison to SMR, as CO₂ emissions from the furnace at a SMR plant are relatively uneconomic to capture due to the low partial pressure in furnace exhaust. Investment costs of oxygen-blown ATR are about 15–25% lower than fired tubular reformers, but the costs of oxygen supply make ATR less attractive than SMR even for large-scale plants (650 MW_{H2}) [97]. The integration of ceramic ion transport membranes for oxygen production with an ATR reactor opens new possibilities for high-efficient and low-cost hydrogen production with CO₂ capture on the long term [98].

4.1.3. Membrane reforming

Large-scale hydrogen production by means of membrane reforming could reduce investment cost versus conventional SMR with MDEA (by omission of shift reactor, absorption unit and reduced CO₂ compression requirement). However, CO₂ avoidance costs are nearly equal, as the MR case requires electricity import (all excess steam is required for sweep), whilst the conventional system has a large electricity export. In order to become feasible for low-cost hydrogen production, membrane costs need to be reduced by applying thinner membranes without compromising selectivity [99].

MR might prove more interesting on a small scale. A recently accomplished study demonstrates that membrane reformers enable hydrogen production at refueling stations combined with relatively inexpensive CO₂ capture [100].¹⁹

¹⁹ CO₂ capture at conventional small-scale reforming units is prohibitive due to relatively high costs of CO₂ capture at such scales [101].

4.1.4. Key performance data

Table 6 provides the ranges in efficiency and investment costs for the different routes of hydrogen production from natural gas with CO₂ capture. For the short term,²⁰ we consider large-scale SMR with a chemical absorption unit (MDEA) after the shift using data from IEA GHG study [12]. This is technically feasible and more efficient and less costly than installing a MEA unit after the furnace. On the longer term, we consider both large-scale ATR-ITM with a chemical/physical absorption unit after the shift [14] and small-scale MR [100].

Annual O&M costs for SMR/ATR systems with CO₂ capture are between 3 and 6% of TPC [12–14]. Annual O&M costs for small-scale MR are estimated at 9% of TPC, of which 5% for membrane replacement [100].

4.2. Coal gasification

Coal gasification (CG) for hydrogen production has large similarities to IGCC with CO₂ capture. The main difference is the addition of a PSA unit to produce high-purity hydrogen. Syngas is produced in an oxygen-blown, entrained flow gasifier, cooled and cleaned before entering the WGS reactor. H₂S and CO₂²¹ are removed by means of physical absorption and the clean, shifted syngas is purified in a PSA unit to deliver 99.999% pure hydrogen at feed pressure. The PSA could be omitted to produce fuel-grade hydrogen (~93%) suitable for combustion, reducing hydrogen production costs with circa 5% [76]. PSA purge gas can be used to superheat steam or can be combusted for power generation in a combined cycle or Rankine cycle (co-production systems). By lowering the steam-to-carbon ratio or directing a part of the syngas to the gas turbine (by-passing the PSA), the hydrogen-to-electricity ratio can be adapted without major thermodynamic consequences, allowing these co-production systems²² to match their output to electricity and hydrogen demands [80].

²⁰ Note that short-term hydrogen production is generally foreseen by means of small scale SMR or electrolysis and that CO₂ capture at large SMR units is foreseen on the longer term, if sufficient demand exists. The term short term in this study must be considered in terms of technological maturity.

²¹ Systems without CO₂ capture do not necessarily need a physical absorption unit to remove CO₂, but doing so has the advantage of reducing the flow to the PSA and increasing the heating value of the purge gas to a point that makes it feasible for use in a gas turbine [80].

²² For simplicity, we allocate all CO₂ emissions to hydrogen production.

Table 6
Key parameters of hydrogen production from natural gas with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Fuel + feed input (GJ/GJ _{H2})	Electricity input (GJ _e /GJ _{H2})	Conversion efficiency ^a (%)	CO ₂ capture efficiency (%)	TCR (€/kW _{H2}) ^b	Source
SMR	MEA after furnace	1.53	0.06	61	90	550	[98]
SMR	MDEA after shift	1.37	0	73	85	550	[12]
Advanced	Physical solvent	1.27	0.001	79	95	390	[98]
ATR-ITM		1.28	0.03	74	90	280	[14]
Small-scale MR ^c	Pd membrane	1.26	0.13	65	70	610	[100]

^a Including CO₂ and H₂ compression to 110 and 60 bar, respectively (except MR).

^b The plants considered here have a large capacity (600–2000 MW_{H2}) and are scaled to 1000 MW_{H2}. A scaling factor of 0.6 can be derived from SMR investment costs of 70 and 700 MW_{H2} [101]. We assumed a scaling factor of 0.75, which is used to scale SMR with larger capacity (> 700 MW_{H2}) [14].

^c Represents 1.7 MW_{H2} membrane reactor using sweep-gas to maximise CO₂ concentration in the retentate flow (i.e. minimise CH₄ and H₂ concentration) to enable direct transport and storage. H₂ is pressurised to 480 bar (in contrast to other hydrogen production plants, which have been standardised to 60 bar) to enable fast cascade-filling into vehicle tanks at 350 bar [100].

In advanced coal gasification systems, ITM might be applied for oxygen production and the Selexol and PSA unit could be replaced by inorganic membranes to separate hydrogen from CO₂. Both metallic (Pd and Pd alloys) and ceramic membranes are being investigated. A techno-economic study performed by Kreutz [102] assessing hydrogen production with Pd alloy membranes versus Selexol and PSA indicates that cost advantages of membrane separators/reactors are moderate. Ceramic membranes have the advantage of being cheaper, but these membranes are less selective, resulting in lower hydrogen purity (>99.5%) [13].

Hydrogen of this purity cannot be applied in PEMFC, as CO poisons the catalyst.

Table 7 shows the range in conversion efficiency and investment costs of hydrogen production by means of coal gasification with CO₂ capture. For the short term, we use the data from the analysis performed by Chiesa and Kreutz et al. [76,80]. This is the most detailed and transparent study, in which all optimisation strategies with commercially ready technology have been assessed. The efficiency figures are relatively high in comparison to other estimates, which is explained by the more optimistic assumptions in gasifier, power generation, PSA

Table 7
Key parameters of hydrogen production from (bituminous) coal with CO₂ capture from literature review

Conversion technology	CO ₂ capture technology	Fuel input (GJ/GJ _{H2})	Electricity in/output (–) (GJ _e /GJ _{H2})	Conversion efficiency ^a (%)	CO ₂ capture efficiency (%)	TCR (€/kW _{H2}) ^b	Source
CG (Texaco/E-gas) ^c	Selexol	2.30	0	43	97	1640	[12]
		1.94	0.03	50	92	810	[13]
		1.48	0.11	59	90	790	[14]
		1.74	–0.04	62	91	840	[76,80]
CG (Texaco)	Pd-membrane	1.68	–0.02	62	100	680	[102]
CG (E-gas) ^d	Ceramic membrane	1.43	0.04	67	94	610	[13]
		1.80	–0.05	61	94	780	
Advanced CG ^e	Selexol	1.35	0.05	69	90	600	[14]

^a Including CO₂ and H₂ compression to 110 and 60 bar, respectively.

^b Plants considered have capacities in the range from 430–1030 MW_{H2} and are scaled to 1000 MW_{H2}. A scaling factor of 0.8 is assumed, as applied for IGCC [76] and hydrogen production by coal gasification [14]. An economic analysis evaluating the impact of scale on investment costs of hydrogen plants by biomass gasifiers, showing large similarity to plants considered here, revealed that the scaling factors for plants larger than 400 MW_{th} approach 0.9 [103].

^c Klett et al. [13] consider a configuration with net electricity export, but do not include H₂ and CO₂ compression, so that it becomes a net importer if this is accounted for. We do not further consider these data, nor the atypical data presented in [12].

^d Hydrogen separation device operating at 600 °C. Two configurations are considered. The most efficient, low-cost option configuration is designed to recover 95% of hydrogen. However, the electricity generated from the remaining purge-gas burned is not sufficient for CO₂ and H₂ compression. The sub-optimal (more conservative) represents a configuration with 80% H₂ recovery. The power produced from the purge gas is sufficient produced for CO₂ and H₂ compression.

^e Includes improvements in gasifier technology (high-pressure) and ITM for oxygen production.

Table 8
Key parameters of hydrogen production technologies with CO₂ capture

Conversion technology	CO ₂ capture technology	Fuel+feed input (GJ/GJ _{H2})	Electricity in/output (-) (GJ/GJ _{H2})	Conversion efficiency (%)	CO ₂ capture efficiency (%)	TCR (€/kW _{H2})	O&M (%)
<i>State-of-the-art technologies</i>							
SMR	MDEA	1.37	0	73	85	550	4
CG	Selexol	1.74	-0.04	62	91	840	4
<i>Advanced technologies</i>							
Adv. ATR	MDEA	1.28	0.03	74	90	280	4
Adv. CG	Selexol	1.35	0.05	69	90	600	4
MR (2 MW _{H2})	Pd membrane	1.26	0.13	65	70	610	9

performance and overall integration [80]. On the longer term, overall efficiencies close to 70% can be achieved. We use the values for the advanced system with ITM [14]. As the forecasted efficiencies of a system with ceramic membranes are slightly lower and hydrogen is not sufficiently pure for use in PEM fuel cells, this configuration is not further considered here. However, it might be an interesting option to produce fuel grade hydrogen, as plants with ceramic membranes are more efficient and less capital-intensive than conventional plants with PSA and Selexol [13].

Annual O&M costs for hydrogen production by coal gasification with CO₂ capture are between 3.5 and 6% of TPC [13,14,76].

4.3. Overview hydrogen production technologies

Table 8 summarises the key data of promising state-of-the-art and advanced hydrogen production options

with CO₂ capture and compression to 110 bar and hydrogen delivered at 60 bar. We use average annual O&M of 4% of TPC for large-scale hydrogen production from coal and natural gas.

5. Results

5.1. Electricity production

Fig. 2 shows the electricity production costs, assuming a discount rate of 10%, 20 years lifetime and a capacity factor of 85%. We used coal and natural gas price projections for 2020 for large industrial users of 1.7 and 4.7 €/GJ, respectively, including commodity, transmission+distribution costs, taxes and VAT [104]. The impact of fuel price on COE will be assessed by means of sensitivity analyses in the discussion. We did not account for carbon taxes or credits in our calculations.

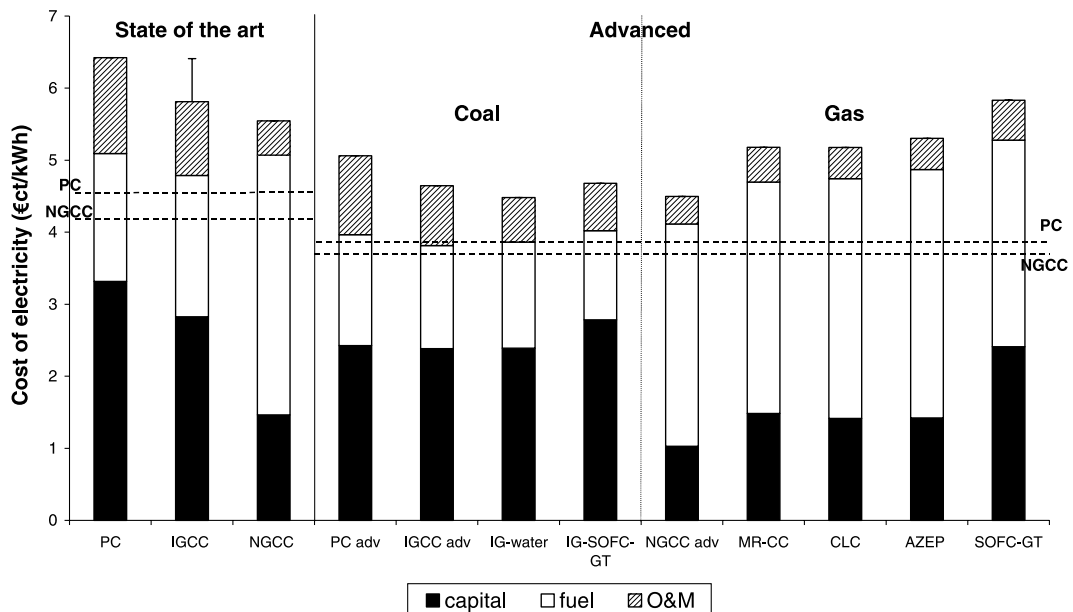


Fig. 2. Cost of electricity (COE). The dashed horizontal lines represent COE of PC and NGCC without CO₂ capture.

When CO₂ capture is applied on the short-term, COE will increase strongest at PC (from 4.6 to 6.4 €/ct/kWh), followed by NGCC (from 4.2 to 5.6 €/ct/kWh) and IGCC (from 4.8 to 5.8 €/ct/kWh). Incremental COE at NGCC is lower than for PC, as less CO₂ is captured per kWh. More optimistic NGCC designs, in which H turbines are integrated with state-of-the-art MEA [21] and G turbines that are heavily integrated with a chemical absorption unit using KS-1 solvent [37] may reach COE levels of 5.2–5.3 €/ct/kWh. Even though IGCC technology without CO₂ capture is relatively capital-intensive in comparison to PC without CO₂ capture, CO₂ capture at IGCC seems to result in lower or comparable COE than PC with capture. This is due to the relatively small efficiency penalty and modest costs of physical absorption in comparison to chemical absorption. The range in IGCC represents the variation in efficiency and investment costs of Shell/Texaco gasification technology. Texaco gasifiers are less capital-intensive and less efficient than Shell gasifiers. CO₂ capture at systems with Texaco gasifiers is more efficient than at systems with Shell gasifiers, i.e. energy penalty is lower, although the electric efficiency remains lower. The additional costs due to capture are also lower for Texaco systems, resulting in lower COE. For a more optimistic design of an IGCC with a Texaco gasifier and CO₂ capture [76,80], COE might even reach levels slightly below 5 €/ct/kWh.

On the longer term, state-of-the-art technologies (PC, IGCC and NGCC) show potential for efficiency improvement and investment cost reduction. The investment costs of the PC/NGCC plant are assumed to decrease with 20% versus current plants and the investment costs of the capture unit are assumed to decrease with 39% for PC and 32% for NGCC after [21]. The reduction in PC/NGCC technology, but also in the capture technology, may be rather optimistic. A detailed study on the potential to reduce investment cost of current post-combustion technology for NGCC, indicated less than 20% reduction in investment costs [42]. As IGCC is a less mature technology, the potential for efficiency improvement and cost reduction is expected to be larger than for PC and NGCC.

Electricity costs for the Water cycle and SOFC-GT fuelled with syngas are comparable to those for advanced IGCC, although the cost and efficiency projections for these two technologies are relatively uncertain. Note that the most advanced developments for the Water cycle, comprising turbines with higher TIT and ITM for oxygen production were not accounted for.

Fig. 2 suggests a ranking of natural gas-fired technologies on the longer term. However, there are large discrepancies and uncertainties in investment costs. First, there is no uniform basis for NGCC baseline costs. The low COE for the advanced NGCC with post-combustion capture can be explained by the optimistic costs reduction in NGCC (and capture) technology. Future cost reductions in NGCC are not explicitly considered in the studies for MR-CC and AZEP. This makes a fair comparison between advanced gas-fired technologies impossible at this stage. Efficiency forecasts of the advanced natural gas-fired technologies, which strongly affect COE, are considered more reliable.

Although the investment costs of SOFC-GT hybrids are relatively high, the calculated COE are only slightly higher than for current decentralised power generation without capture.²³

Fig. 3 presents the amount of CO₂ emitted and CO₂ captured. The sum, the total amount of CO₂ produced, is a measure of the net efficiency. Generally, the amount of CO₂ captured and emitted for coal-fired power plants are more than twice as large than natural gas-fired power plants. This is explained by the higher conversion efficiency of natural gas-fired plants and lower CO₂ emission factor of natural gas versus coal (56 and 95 kg CO₂/GJ, respectively). Total CO₂ produced and emitted is highest at state-of-the-art IGCC with (conservative) electric efficiencies between 32 and 35%. The efficiencies of advanced gas-fired options are all in the same range. Note that MR-CC, CLC, AZEP (and also SOFC-GT) enable 100% capture at high efficiency,²⁴ whereas NGCC with post combustion capture achieves only 85–90% capture.

CO₂ avoidance costs of electricity production are presented in Fig. 4. It shows that CO₂ avoidance costs for NGCC are higher than for PC (and IGCC), which is due to the higher volume of flue gas and lower CO₂ content. IGCC obviously has lowest avoidance costs, as the additional investment cost and energy penalty are relatively modest. This technological insight is basically the same on the long term. Less energy and capital-intensive technologies are able to reduce CO₂ avoidance costs to levels below 20 and 40 €/t CO₂ for

²³ A 20 MW_e NGCC is assumed as reference with an efficiency of 40% and TCR of 800 €/kWe [69].

²⁴ By installing an afterburner, the efficiency is increased at the expense of a lower CO₂ recovery, with the exception of the SOFC-GT, where the efficiency is decreased.

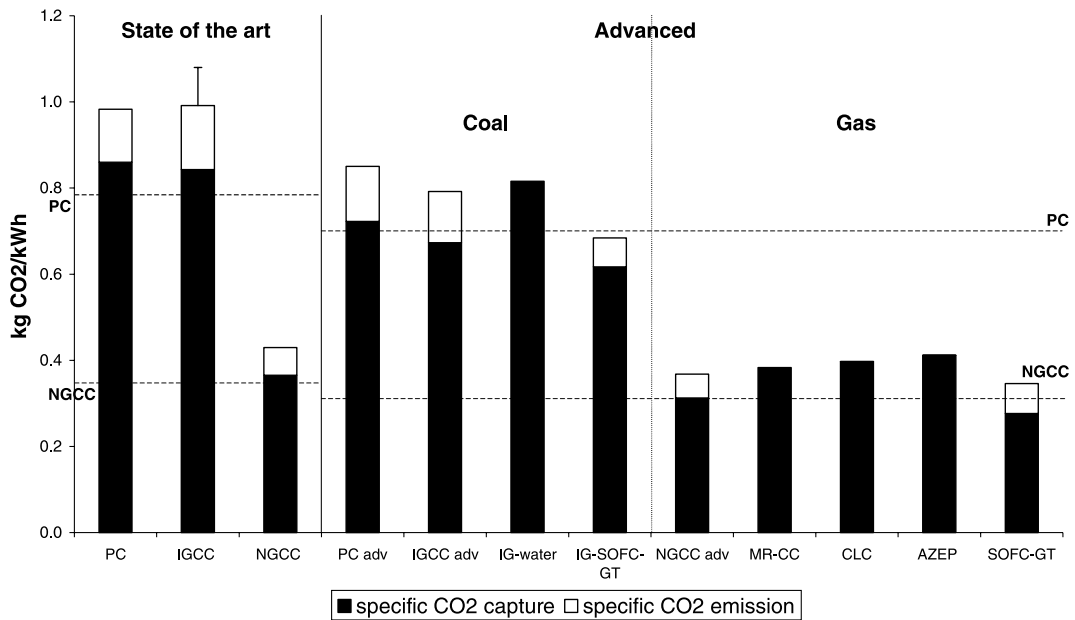


Fig. 3. Specific CO₂ capture and emission rates for electricity production. The dashed horizontal lines represent emissions of PC and NGCC without CO₂ capture.

coal and gas-fired technologies, respectively. Note that the avoidance costs for AZEP and CLC might be reduced when CO₂ capture efficiency is reduced to 85%. SOFC-GT has even lower avoidance costs, which is primarily caused by the low conversion efficiency of the reference system.

5.2. Hydrogen production

The costs of hydrogen production are presented in Fig. 5, assuming the same economic parameters as for electricity. Natural gas prices for small industrial users (MR) are estimated at 6.3 €/GJ. Electricity price

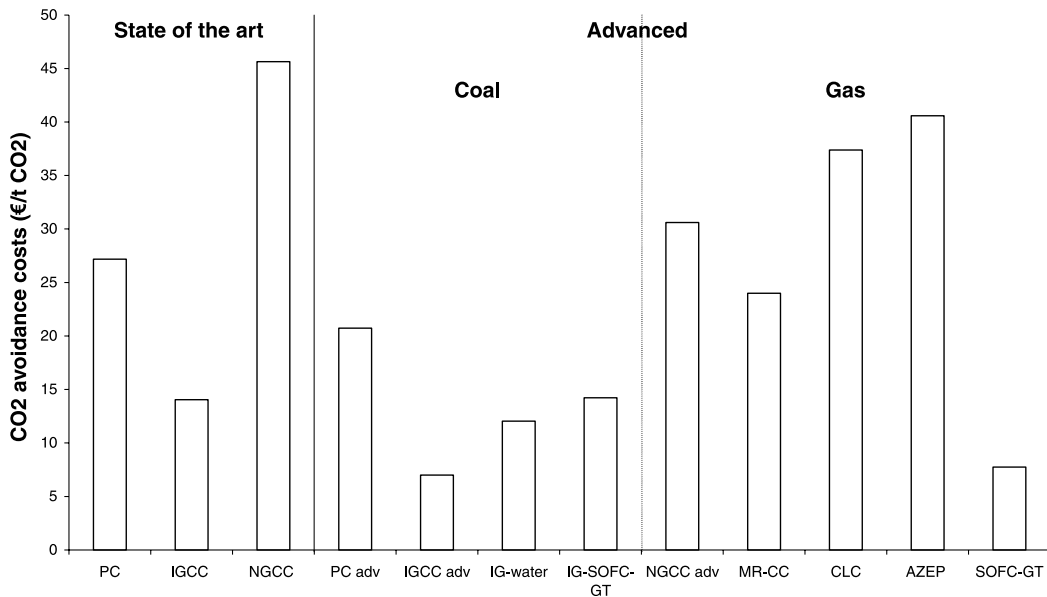


Fig. 4. CO₂ avoidance costs for electricity production (reference: identical technology without CO₂ capture. For the advanced natural gas-fired options and IG-Water/SOFC, we use NGCC and IGCC as reference, respectively).

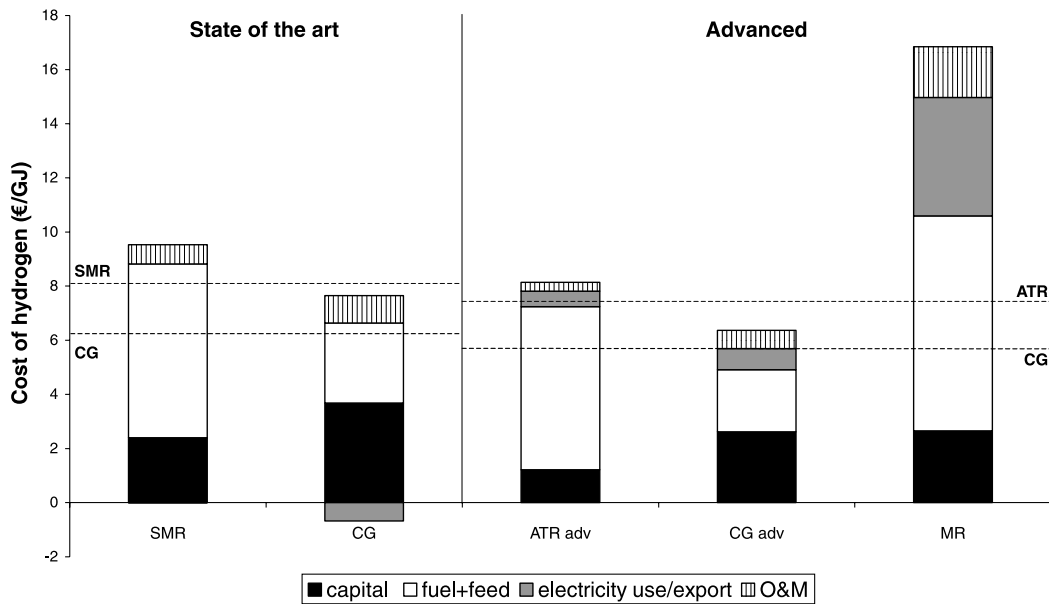


Fig. 5. Cost of hydrogen (COH). The dashed horizontal lines represent COH of SMR/ATR and CG without CO₂ capture.

projections for 2020 of 0.06 and 0.12 €/kWh for large and small industrial users, respectively, are used (both for selling and buying) [104].

Costs of hydrogen production with CO₂ capture by means of state-of-the-art CG are lower than for SMR (9.5 versus 7 €/GJ_{H₂}). The low costs for CG can be explained by the relatively small additional investment costs and low energy penalty and the rather optimistic assumptions in gasification, power generation, PSA and overall integration of the CG plant. For a more conservative configuration, in which electricity is imported [14], COH are significantly higher (9 €/GJ).

On the longer term, advanced ATR and CG systems might produce hydrogen with strongly reduced CO₂ emissions at circa 8.1 and 6.4 €/GJ, respectively. COH for small-scale MR are much higher than for large-scale SMR, although the former include costs related to hydrogen compression to 470 bar. The costs associated with additional hydrogen infrastructure (buffer storage, pipeline, compressor, etc.), which are significant and should therefore be included for a fair comparison, are considered in paper B. Provided that the same tariffs for natural gas and electricity could be negotiated as large-scale users, COH might drop below 13 €/GJ.

Fig. 6 illustrates that the amount of CO₂ emitted for the considered SMR plant is slightly higher than for the CG plant. This can be explained by the lower capture efficiency at SMR as CO₂ produced in the furnace of the SMR is emitted and the emissions avoided by electricity export at CG (which replace CO₂ emissions

generated by central power production). The capture efficiency of the advanced CG plant is equal to the state-of-the-art CG plant, but electricity is imported, causing additional CO₂ emissions. The CO₂ emission rate is largest for the small-scale MR, which is designed to recover only 70% of the produced CO₂, and additionally, electricity is imported.

The relatively small incremental COH caused by CO₂ capture expresses itself in relatively low CO₂ avoidance costs (see Fig. 7) in comparison to CO₂ capture at electricity production, which is explained by the fact that syngas needs to be shifted anyhow. CO₂ avoidance costs for CG plants are very low, since CO₂ is generally also separated from the shifted syngas in configurations without CO₂ capture. The additional investment requirements are a CO₂ compressor only. The ATR system with ITM would make CO₂ capture from hydrogen production more attractive in comparison to traditional SMR, due to the low (additional) investment costs and higher CO₂ recovery.

6. Discussion and conclusion

6.1. Electricity production with CO₂ capture

Electricity production costs of NGCC with CO₂ capture are generally lower than for PC/IGCC with CO₂ capture, but CO₂ avoidance costs for NGCC are higher than for PC/IGCC. This outcome is confirmed by most other studies comparing PC, NGCC and IGCC, with the

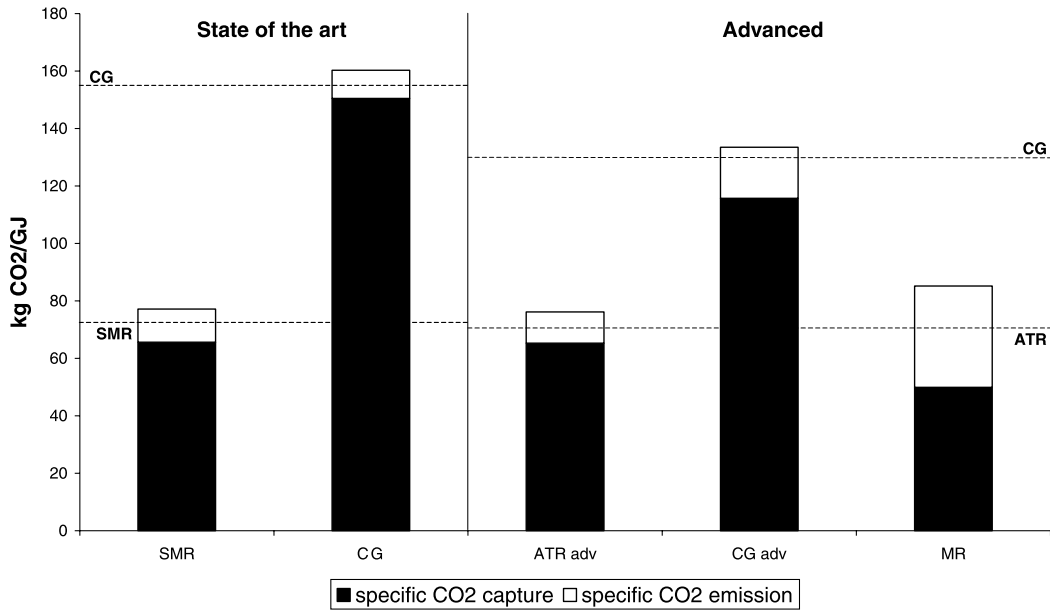


Fig. 6. Specific CO₂ capture and emission rates of hydrogen production. The dashed horizontal lines represent emissions of SMR/ATR and CG without CO₂ capture.

exception of an IEA GHG study [20], in which avoidance costs are higher for PC. Incremental COE and CO₂ avoidance costs at PC (and NGCC) presented in this study are relatively low in comparison to other studies, as we apply data representing improved chemical absorption technology with significantly lower energy consumption and costs in comparison to current generation MEA plants considered in most

other studies. IGCC offers a more efficient and cost-effective possibility for CO₂ capture in comparison to PC (i.e. COE and CO₂ avoidance costs are lower). It should be noted that uncertainty in costs and efficiency is larger for IGCC in comparison to PC. For the former we have a limited data set for first-of-a-kind plants, whereas real costs are available for PC plants (without CO₂ capture).

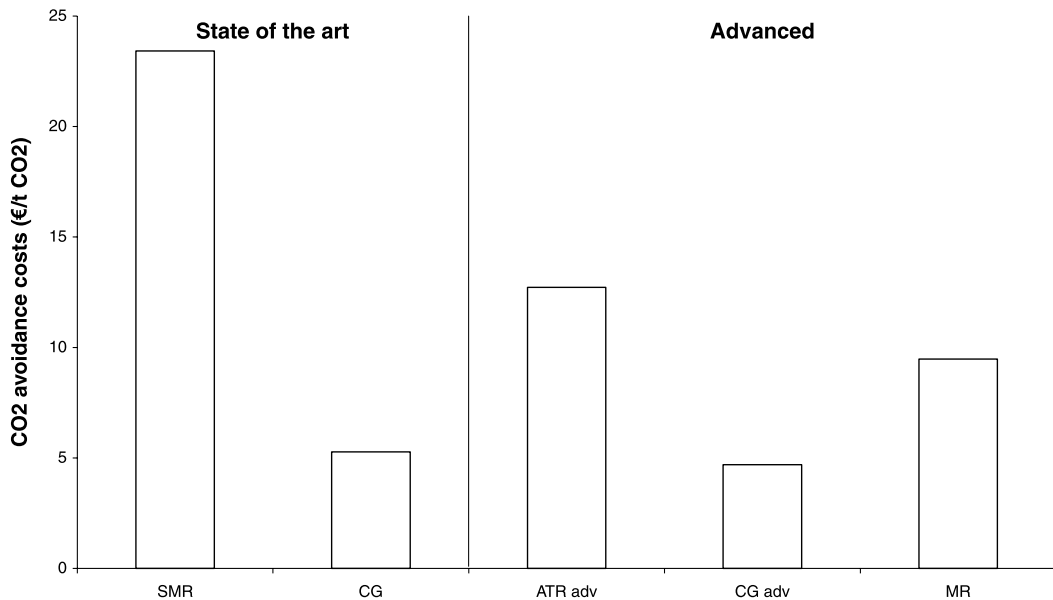


Fig. 7. CO₂ avoidance costs for hydrogen production (reference: identical technology without CO₂ capture).

It is unclear whether oxyfuel combustion is more cost-effective than post-combustion capture at PC plants. On a somewhat longer term, ITM technology could be deployed for oxygen production instead of the conventional ASU. This could make a difference, since the ASU is very energy-intensive and represents circa 20% of total investment costs [31].

Both PC with post-combustion capture and IGCC with pre-combustion capture still have significant improvement potential. IGCC maintains its position as low-cost CO₂ capture and electricity production technology in comparison to PC on the longer term. Electricity production costs in the same order as for current PC without CO₂ capture (circa 4.6 €/ct/kWh) may be achievable, provided that a net electric efficiency over 43% and investment costs of 1500 €/kW prove realistic. COE of the Water cycle fuelled with syngas and IG-SOFC-GT are comparable to IGCC. However, the technological status of these technologies is less mature. Significant R&D efforts are needed to develop and commercialise fuel cells and turbines suited for conditions in Water cycles.

Advanced natural gas-fired options such as MR-CC, AZEP, CLC and SOFC-GT promise power generation with high efficiencies and 100% CO₂ capture. However, material restrictions (membranes/oxygen particles) in these concepts limit the reactor/combustion chamber temperature to a level lower than TIT of state-of-the-art gas turbines. This can be resolved by firing additional natural gas to increase TIT, thereby reducing CO₂ capture efficiency. The trade-off between costs and CO₂ capture efficiency requires further attention.

The economic performance of these advanced power production technologies is still uncertain. This is mainly due to the uncertainties pertaining to investment and O&M costs and lack of detailed cost-engineering studies for certain concepts (such as AZEP and CLC). The technologies considered contain specific elements (membranes, fuel cells, adapted turbines) for which little data are available on future commercial cost levels when integrated in a power cycle. Possibly, different/cheaper materials might be applied in future designs, which might have significant impact on performance and costs.

A recently accomplished techno-economic study evaluating amongst others advanced post and pre-combustion CO₂ capture at NGCC plants shows that CO₂ avoidance costs of MR-CC are slightly lower than NGCC with a highly integrated post-combustion capture unit [42]. However, this study did not account for efficiency improvements and cost reductions in NGCC technology itself; only improvements in capture

technology and integration were studied. An important question remains which capture strategy will benefit most strongly from autonomous developments in NGCC technology and to what extent improved turbines can be integrated in advanced power cycles with CO₂ capture. Therefore, more detailed techno-economic analyses of advanced natural gas-fired electricity options are required, especially for AZEP and CLC. In such analyses, efficiency and cost improvement potential in state-of-the-art and advanced conversion and capture technology should be considered. The impact of technological learning should be accounted for in cost assessments, which has not been done so far for this kind of technologies.

6.2. Hydrogen production with CO₂ capture

CO₂ capture at hydrogen plants causes a relatively small increase in production costs in comparison to electricity plants, which makes hydrogen production more attractive for CO₂ capture. Especially coal gasification enables CO₂ capture with marginal additional investment costs and efficiency losses. Hydrogen production costs (with CO₂ capture) can be reduced further by application of ITM in advanced ATR and CG systems. If ITM membranes prove to operate under commercial conditions, ATR would be the preferable way to produce hydrogen from natural gas, as it allows for higher recovery and less costly CO₂ capture. On a small scale, MR allows for hydrogen production with low additional costs for CO₂ capture. However, significant R&D is still required to enable (mass) production of thin and stable membranes with sufficient lifetime.

There are other advanced, potential low-cost hydrogen production schemes that are not further considered here mainly due to a lack of data. In sorption enhanced reforming (SER), reforming is conducted in the presence of an adsorbent that removes CO₂, thereby shifting the reaction to the product site and allowing the reaction to occur at low temperatures (400–500 °C) [105]. Small-scale laboratory tests demonstrated the principle. The HyGenSys process integrates a novel reformer concept with a jet engine heat generator to supply heat for the reforming reaction [106]. Plasma reforming involves natural gas heating by means of an electric arc to a temperature that a plasma state is reached, allowing the reaction to occur without catalysts [107]. This technology is still in experimental phase (in the kilowatt range). Alternatively, natural gas can be cracked into carbon and hydrogen by means of the plasma cracking process,

Table 9
Impact of fuel prices on electricity and hydrogen production costs

Fuel price (€/GJ)	1	2	3	4	5	6
<i>COE (€/ct/kWh)</i>						
PC	5.7	6.7	7.8	–	–	–
IGCC	5.0	6.2	7.3	–	–	–
NGCC	–	–	4.2	5.0	5.8	6.5
Adv. PC	4.4	5.3	6.2	–	–	–
Adv. IGCC	4.1	4.9	5.7	–	–	–
IG-Water	3.9	4.7	5.6	–	–	–
IG-SOFC-GT	4.2	4.9	5.6	–	–	–
Adv. NGCC	–	–	3.4	4.0	4.7	5.4
MR-CC	–	–	4.0	4.7	5.4	6.1
CLC	–	–	4.0	4.7	5.4	6.1
AZEP	–	–	4.1	4.8	5.5	6.3
SOFC-GT	–	–	4.8	5.4	6.0	6.6
<i>COH (€/GJ_{H2})</i>						
SMR	–	–	7.2	8.6	9.9	11.3
CG	5.7	7.5	9.2	–	–	–
Adv. ATR	–	–	6.0	7.2	8.5	9.8
Adv. CG	5.4	6.8	8.1	–	–	–
MR (2 MW _{H2})	–	–	12.7	13.9	15.2	16.5

thereby avoiding CO₂ production (and hence separation). Several process variants are being developed and tested [108]. The HyPr-RING process integrates a novel gasification technology to produce hydrogen from coal with CO₂ capture using CaO in a single high-pressure reactor [109].

6.3. The impact of fuel prices

Energy production costs are strongly correlated to fuel prices, which have fluctuated significantly in the past (especially the price of natural gas, which is coupled to the oil price in the Netherlands) and are likely to increase in the future. Coal prices have been traditionally stable after 1990 at a level below 2 €/GJ, although prices exceeded 3 €/GJ in the 1980s. Natural gas prices for large industrial users were stable at circa 3 €/GJ in the 1990s, but increased to a level over 5 €/GJ around 2000 [110]. To account for the impact of uncertain fuel prices, we assess COE and COH for coal prices between 1 and 3 €/GJ and natural gas prices between 3 and 6 €/GJ (see Table 9). Capacity factors of NGCC plants may decrease as gas prices increase strongly, putting PC and IGCC (with CO₂ capture) in a more favourable perspective, as has been assessed by Rubin et al. [9]. However, in this analysis, we compare power generation on an equal load base, as the assessment of capacity factors for various technologies requires a dynamic model for the electricity sector to assess plant dispatch as a function of fuel/CO₂ prices and load curves and configuration of the entire system.

The table indicates that the ranking among coal and natural-gas fired options does not change significantly with changing fuel prices, as the impact of fuel price on coal-fired options is rather limited and conversion efficiencies of the advanced gas-fired options are quite close. The break-even points of state-of-the-art NGCC with CO₂ capture versus PC/IGCC with CO₂ capture lie at a natural gas price of 5.8 and 5 €/GJ, respectively (at a constant coal price of 1.7 €/GJ and equal capacity factors²⁵). This would imply that CO₂ capture at IGCC might become competitive with current trend of high gas prices, provided that capital cost as assumed in this study prove realistic.

Just as for electricity production, coal-based hydrogen production seems more competitive with current fuel prices and is likely to remain more competitive in the future, which is confirmed in a study performed by Williams [111].

Within the range of fuel prices studied, CO₂ mitigation costs of PC (and IGCC) with CO₂ capture remain lower than NGCC with CO₂ capture. Based on the results of an electricity market dispatch model, Johnson and Keith [112] conclude that CCS technologies entering the market as a consequence of high CO₂

²⁵ Based on varying operating costs of technologies with and without capture as a function of fuel prices and a relation between actual capacity factors and fuel prices, Rubin et al. [9] recalculated COE. Assuming a capacity factor of 50 and 75% for NGCC and PC with CO₂ capture, respectively, and coal/gas price of 1.2 and 4 €/GJ, COE of PC and NGCC with CO₂ capture become very close.

prices, are characterised by higher capacity factors in comparison to base-case technologies. As a consequence, the mitigation costs of CCS technologies decreases.

6.4. Overall conclusion

In this study, an extensive literature review has been performed to make a comparison of coal and natural gas-fired electricity and hydrogen production technologies with CO₂ capture. We considered both state-of-the-art and advanced technologies and accounted for developments in existing technologies. Although the focus was on large-scale facilities, also decentralised electricity and hydrogen production technologies that enable low-cost CO₂ capture were considered. Energetic and economic data derived from literature were standardised and selected on a number of criteria to get figures on conversion efficiency, energy production costs and CO₂ avoidance costs, needed to compare the technologies in a consistent matter.

In the short term (around 2010), (ultra) supercritical PC and NGCC plants equipped with chemical absorption units and IGCC plants with a shift reactor and a physical absorption unit are the most likely candidates for electricity production with strongly reduced CO₂ emission (i.e. 85–90% capture efficiency). Net electric efficiencies are 30–35% for PC, 32–40% for IGCC and 43–50% for NGCC. Investment costs of 600 MW_e units lie between 1700–2500 €/kW_e for PC, 1500–2200 €/kW_e for IGCC and 700–1100 €/kW_e for NGCC. Calculated electricity production costs are between 5.4–6.9 €/ct/kWh for PC, 4.7–6.3 €/ct/kWh for IGCC and 4.9–6.3 €/ct/kWh for NGCC, assuming a discount rate of 10%, 20 years lifetime, capacity factor of 85%, coal and natural gas prices of 1.7 and 4.7 €/GJ. CO₂ capture results in the lowest COE when applied to NGCC, although CO₂ capture at IGCC and PC could become competitive with increasingly high natural gas prices foreseen for the future. IGCC has significant lower CO₂ avoidance costs versus an identical plant without capture than NGCC and PC (14 €/t CO₂ for IGCC versus 46 €/t CO₂ for NGCC and 27 €/t CO₂ for PC).

In the longer term (around 2030), both new power cycles with integrated CO₂ capture and significant improvements in existing technology are foreseen. Advanced PC with improved post-combustion CO₂ capture might achieve a net electric efficiency of 40% and investment costs of 1500 €/kW_e. Advanced concepts based on gasification might reach net efficiencies of 43% for IGCC (with ITM) up to 50%

for IG-SOFC-GT. Investment costs are estimated between 1500 €/kW_e for IGCC and 1750 €/kW_e for IG-SOFC-GT. COE of advanced coal-fired power cycles, assuming identical economic assumptions as for the short term, are estimated at 4.5–4.7 €/ct/kWh for concepts deploying gasification and 5.1 €/ct/kWh for PC. CO₂ avoidance costs might be reduced to a level of 7 €/t CO₂ (for IGCC) up to 21 €/t CO₂ (for PC). These numbers clearly indicate that the advanced options based on coal gasification promise more efficient and cost-effective CO₂ capture than improved PC with post-combustion capture, despite the improvement potential for chemical absorption (and PC technology).

Advanced natural gas-fired combined cycle systems with post, pre (MR-CC) and oxyfuel (AZEP, CLC) combustion strategies may reach net electric efficiencies between 50 and 55% for CO₂ capture efficiencies between 85 and 100%. Estimated investment costs vary between 650 and 900 €/kW_e, resulting in COE between 4.5 and 5.3 €/ct/kWh and CO₂ avoidance costs in the range of 25–40 €/t CO₂. Decentralised SOFC-GT hybrids (20 MW_e) enable power production with CO₂ capture at a net electric efficiency close to 60% and investment costs of 1500 €/kW_e, resulting in COE of 5.8 €/ct/kWh and CO₂ avoidance costs of 8 €/t CO₂ versus a decentralised NGCC as baseline.

In contrast to the coal-fired cases, for which pre-combustion strategies seem more promising than post-combustion strategies and oxyfuel combustion with cryogenic air separation, no winning capture strategy among advanced natural gas-fired electricity production technologies can be appointed at this stage. Mainly the uncertainty in investment costs makes it impossible to make a clear ranking of these options. It is important to realise, though, that existing technology, i.e. NGCC with post-combustion capture, still has considerable development potential and may remain competitive with advanced pre-combustion and oxyfuel combustion concepts in the coming decade. Various studies comparing advanced power cycles with state-of-the-art power cycles with post-combustion capture pass by the fact that conventional gas turbine and amine scrubbing technology will further improve.

With current technologies, hydrogen can be produced by means of large-scale SMR and CG with CO₂ capture from the shifted syngas, resulting in a CO₂ capture efficiency of 85–90%. The conversion efficiency, including electricity in/outputs in primary terms, is 73% for SMR and 59–62% for CG. Investment costs are circa 550 and 840 €/kW_{H₂} for 1000 MW_{H₂} SMR and CG plants, respectively. Costs of hydrogen produced by SMR are dominated by fuel and feed costs,

which puts CG in more favourable perspective considering increasingly high gas prices. Hydrogen production costs for SMR are estimated at 9.5 €/GJ and an optimal designed CG plant with electricity export may reach 7 €/GJ. CO₂ avoidance costs versus identical plants without CO₂ capture are 23 and 5 €/t CO₂ for SMR and CG, respectively.

Advanced large-scale ATR and CG systems with ITM enable hydrogen production with 90% CO₂ capture at an efficiency of 73% for ATR and 69% for CG. Investment costs of these systems are estimated at nearly 300 and 600 €/kW_{H₂}, resulting in COH of 8.1 and 6.4 €/GJ and CO₂ avoidance costs of 13 and 5 €/t CO₂ for ATR and CG, respectively. MR enable small-scale hydrogen production with relatively low-cost CO₂ capture. A 2 MW_{H₂} plant may reach efficiencies of 65% and investment costs around 600 €/kW_{H₂} (including hydrogen compression to 480 bar), resulting in COH of nearly 17 €/GJ, considering gas and electricity prices for small industrial users.

This study is useful in the sense that it gives a complete overview of electricity and hydrogen production technologies with CO₂ capture and standardised ranges in energetic and economic performance. Where possible, the causes of the variation in conversion efficiency and costs, which can be substantial for certain technologies, are explained on the basis of specific technological or economic assumptions, which enables us to choose data according to what we consider most representative. By considering advanced technologies and improvements in current technologies, we also give insight in efficiency improvements and cost reductions that might be achievable on the longer term and technological barriers that have to be surmounted. However, the standardisation procedure has its limits. In spite of our endeavours to standardise cost and performance data, a certain variability remains due to differences in capital and labour costs, technical performance of individual components, fuel specifications, cooling water temperature, etc. between the studies considered. Capital costs may vary strongly due to assumptions with respect to different cost components. Another uncertainty is what costs do refer to exactly; they can represent a plant built today, a first-of-a-kind plant built in the future or a mature advanced technology (including technological learning). Uncertainty obviously is highest in advanced concepts (e.g. CLC and AZEP). The conversion of different currencies (mainly dollars to euros) by means of GDP deflators and annual currency exchange rates introduces

uncertainty as well. All these factors limit the use of the results to indicative figures for what may be achievable.

In order to compare CO₂ mitigation in the electricity sector versus hydrogen production and deployment, we need to consider real avoidance costs, which are directly related to the reference system. An IGCC equipped with a CO₂ capture unit does not necessarily replace an IGCC plant without CO₂ capture, as this choice is also a function of policies and market choices. ‘CO₂ neutral’ hydrogen should be compared to the fuel it substitutes, which can be either gasoline or diesel in the transport sector, natural gas in households and industry or hydrogen produced without CO₂ capture applied currently in the industry. In paper B, we will assess the technologies considered here in a chain analysis, including the complete chain of CO₂ transport and storage and energy carrier distribution and end-use. Such analysis enables a comparison of CCS technologies in their complete context and also allows us to assess in which sector CO₂ capture proves to be most efficient, both from energetic and economic perspective.

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