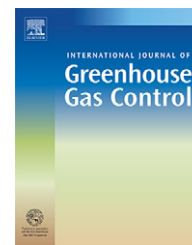


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Techno-economic analysis of natural gas combined cycles with post-combustion CO₂ absorption, including a detailed evaluation of the development potential

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ABSTRACT

We performed a detailed analysis of the potential future costs and performance of post-combustion CO₂ absorption in combination with a natural gas combined cycle (NGCC). After researching state-of-the-art technology, an Excel model was created to analyze possible developments in the performance of energy conversion, CO₂ capture, and CO₂ compression. The input variables for the three time frames we used were based on literature data, product information, expert opinions, and our own analysis. Using a natural gas price of 4.7 €/GJ, we calculated a potential decrease in the costs of electricity from 5.6 €/kWh in the short term to 4.8 €/kWh in the medium term and 4.5 €/kWh in the long term. The efficiency penalty is calculated to decline from 7.9%-points LHV in the short term to 4.9%-points and 3.7%-points in the medium and long terms, respectively. In combination with NGCC improvements, this may cause an improvement in the net efficiency, including CO₂ capture, from 49% in the short term to 55% and 58% in the medium and long terms, respectively. The total capital costs including capital costs of the NGCC were calculated to decline from 880 in the short term to 750 and 690 €/kW in the medium and long terms, respectively, with a decline in the incremental capital costs due to capture from 350 in the short term to 270 and 240 €/kW in the medium and long terms, respectively. Finally, the avoidance costs may decline from 45 €/tCO₂ in the short term to 33 €/tCO₂ in the medium term and 28 €/tCO₂ in the long term.

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1. Introduction

In a world with a need to deeply reduce greenhouse gas (GHG) emissions, carbon capture and storage (CCS) may become an important approach (IPCC, 2001, 2005; Pacala and Socolow, 2004; UNDP, 2004). CCS generally implies CO₂ capture from large point sources, CO₂ transport in a compressed form, and CO₂ storage in, e.g., deep geological formations. As reviewed by a special report of the IPCC (2005), a wide range of options is available for both capture and storage.

CO₂ capture is relatively cost- and energy-intensive. According to the IPCC (2005), state-of-the-art CO₂ capture technology based on post-combustion absorption by amines implies at present an energy penalty of approximately 8%-points when applied to a natural gas combined cycle (NGCC) and, excluding the required CO₂ infrastructure, almost a doubling of the capital costs. Moreover, combined with incremental O&M costs, the costs of electricity (COE) would increase by nearly 40%. This clearly illustrates the need for the further development of CO₂ capture technologies in order to

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decrease capital costs and energy requirements. In this respect, a comparison between state-of-the-art technology and advanced technology is important in order to select promising options for further development.

Comparisons between the various CO₂ capture technologies for power plants have been performed by several authors (Damen et al., 2006; David and Herzog, 2000; Göttlicher and Pruscek, 1997; PITGI, 2002; Simbeck, 1999). In addition, various projects have been conducted to sketch the improvement potential of specific CO₂ capture technologies (IEA GHG, 2004; Rao et al., 2004; Thomas, 2005). Until now, however, no comparisons of CO₂ capture technologies have been performed that were based on a very detailed evaluation of improvement as a function of time. This paper describes a detailed and transparent analysis of the potential future performance of post-combustion CO₂ absorption in combination with an NGCC. It analyzes the impact of individual improvement options and presents a detailed uncertainty analysis. Carrying out the same type of study on other promising CO₂ capture options will make an extensive comparison of these technologies possible.

Because the application of post-combustion CO₂ absorption technology has a large impact on both the costs and the efficiency of electricity production, we were interested in the potential improvements in the impact that CO₂ capture and compression have on COE and energy efficiency. With respect to CO₂ abatement strategy and comparability to other abatement options, it is important to study CO₂ avoidance costs. In our goal to comprehensively analyze possible future developments, we first analyzed the state-of-the-art technology and the interrelations between parameters influencing both energy consumption and capital costs as well as the O&M costs (Section 2). Then, we investigated the development potential of the technology with respect to process parameters, process layout, and component improvements based on data from the literature, product information, expert opinions, and our own analysis (Section 3). These values were subsequently used in an Excel-based model, describing the relationship between the most important process parameters and their influence on capital costs as described in Section 2. The model produced values for a breakdown of (incremental) capital costs, energy penalty, energy efficiency, and (incremental) COE for the different time frames. The time frames we investigated are shown in Table 1. The Excel model was also used to analyze the impact of the uncertainty in the development of each parameter on the energy penalty. We also performed a sensitivity analysis of the COE and avoidance costs with regard to the natural gas price, the number of operating hours, the interest rate, and the lifetime of the plant. The

results of the analysis of the improvements in energy efficiency are described in Section 4, the results of the economical analysis in Section 5, and a discussion of the results in Section 6.

Using this approach, we obtained detailed insight into future possible improvements, the time frame in which the developments can be realized, the relative importance of the various technological improvements in improving performance, and the R&D efforts that are required to achieve the required developments.

2. The process layout

This section describes the state-of-the-art technology of the natural gas combined cycle (NGCC), the chemical absorption unit, and the CO₂ compression train.

2.1. Electricity production using a natural gas combined cycle

In an NGCC, air is first compressed and fed to the natural gas combustion chamber. The pressurized hot combustion gases are then expanded in a number of stages in a gas turbine (GT) to produce work, which is converted to electricity by the generator. In this way, the GT off-gas loses its pressure, but not all of its heat, retaining a typical temperature around 500 °C. This heat is used to produce superheated high-pressure steam in a 'Heat Recovery Steam Generator' (HRSG). This steam is then expanded in the steam turbine (ST). The efficiency of a simple cycle in state-of-the-art GTs ranges from 34% to 39% LHV, whereas combined cycles based on the same turbines have efficiencies ranging at present from 55% to 58% LHV.¹ Main development items for efficiency improvements include increasing the combustion temperature and pressure in the GT and the ST, as will be described in Section 3.1.

2.2. Chemical absorption plant

The process of the chemical absorption of CO₂ from flue gas is shown schematically in Fig. 1. It is based on the reversible character of the reaction of CO₂ and other acid gases with alkaline absorbents. These absorbents are usually amines in an aqueous solution (solvents). The flue gas of a power cycle needs to be cooled before it is brought into contact with the solvent. The flue gas is pumped through an absorption column where the CO₂ binds to the absorbent at temperatures between 40 and 60 °C. The flue gas is then washed to remove water and solvent droplets/vapor. The CO₂-rich solvent is subsequently transferred to the top of a stripper column. In this column, heat is used to free the CO₂. The regeneration of the solvent takes place at a temperature between 100 and 140 °C. This heat is generated in a reboiler from steam extracted from the power cycle. The pressure of the regeneration process is nearly atmospheric. The gas stream from the stripper is a CO₂/H₂O mixture. The steam is recovered by a condenser, after which the CO₂ is pressurized for transport. Heat from the CO₂-lean solvent is then transferred

Table 1 – Definitions of the time frame in relationship to the development stage of technologies chosen for this study

	Current technology status	Year
Short term	Available technology	2010
Medium term	Pilot plant stadium	2020
Long term	Laboratory stadium	2030

¹ The <http://www.siemenswestinghouse.com>.

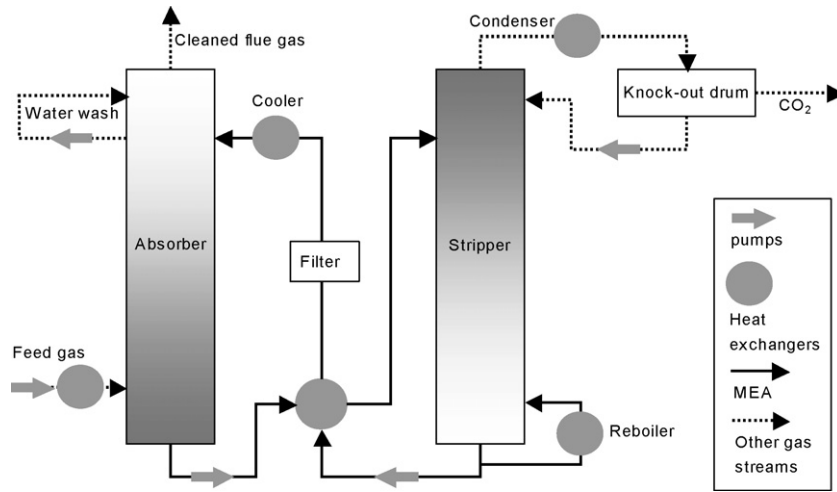


Fig. 1 – Schematic representation of a flue-gas CO₂ absorption process.

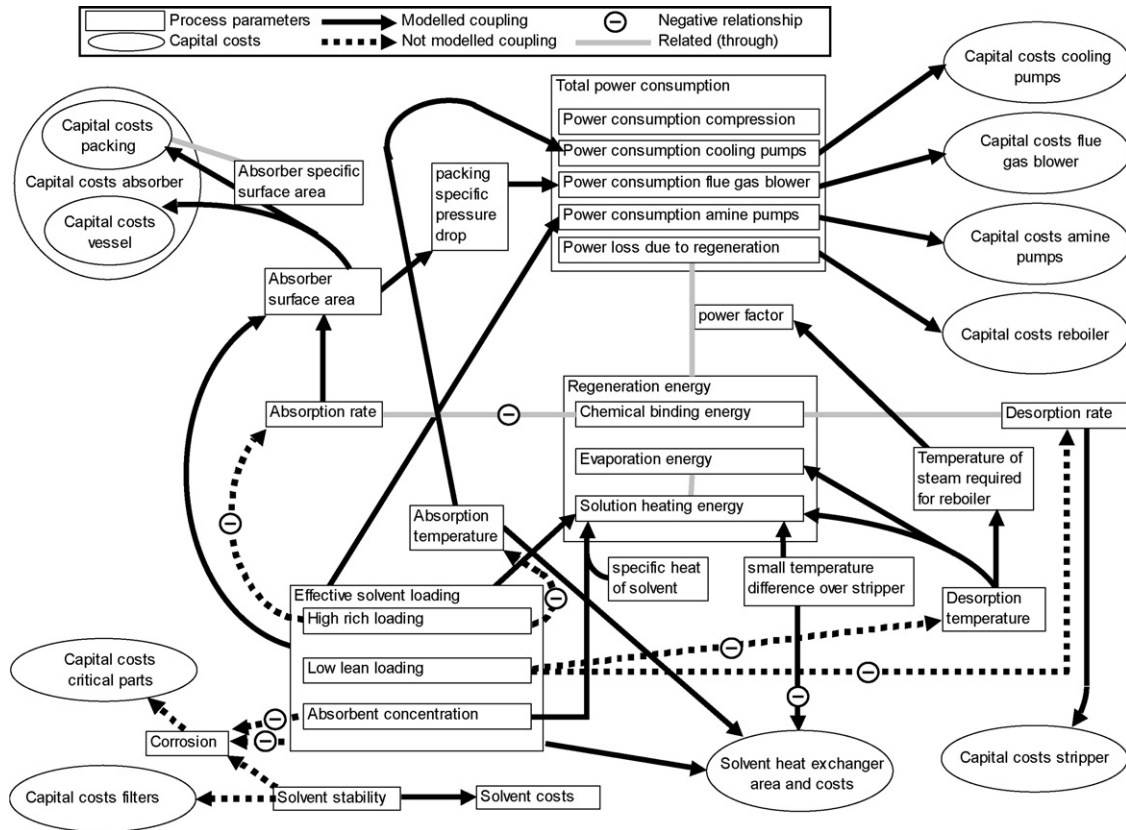


Fig. 2 – Relations of process parameters with performance and capital costs. The relationships shown in solid lines are modeled in Excel. A coupling means that a change in one parameter directly affects the other parameter negatively or positively. A relationship means that two parameters have a coupled behavior, but do not influence each other in a direct manner.

to the CO₂-rich solvent in a heat exchanger, further referred to as lean-rich HEX.

The energy consumption and capital costs of an absorption plant largely depend on solvent and absorber characteristics. The relationships between the various process parameters of the capture plant are shown in Fig. 2.

2.2.1. Solvents

The absorption characteristics of the solvent are very determining for the energy use, capital costs, and O&M costs of the capture process. The most important characteristics are the chemical binding energy, the absorption rate, the solvent loading, and absorption and desorption temperatures. The

stability of the solvent and the specific solvent costs are important characteristics with respect to O&M costs and the impact on the environment.

When CO₂ compression is included, *chemical binding energy* is responsible for over 40% of the regeneration energy requirements and about 25% of the energy penalty in state-of-the-art monoethanolamine (MEA)-based processes. In general, a lower value for the chemical binding energy will result in a lower absorption rate, causing increased size and capital costs of the absorber. The desorption rate, on the other hand, is positively effected. This is, however, a much less important variable because the desorption rate is much higher than the absorption rate. The chemical binding energy is a solvent characteristic that can only be altered by using other absorbents or by changing the solvent formulation. The most important development aim is to lower the chemical binding energy without affecting the absorption rate.

The *absorption rate* determines the required absorber surface area. Because the capital costs of absorbers make up about 40% of the total costs of an amine plant (IEA GHG, 2004), there is a strong coupling between absorption rate and the total capital costs. Moreover, a change in absorber size will directly effect the energy requirements and capital costs of the flue gas blower, which is responsible for about 10% of both the energy penalty and the overall absorption plant costs. Although the dependency of both capital costs and energy requirements on the absorption rate is high, this in itself is not considered a major development item in the available literature. It is, however, important that the absorption rate be conserved when the other parameters are changed, especially the chemical binding energy of the solvent.

The effective loading of the solvent is the difference between lean and rich loading in terms of moles of CO₂ per kilogram solvent. A high *effective solvent loading* will cause a smaller solvent flow in the amine plant, which will influence the dimensions of the solvent heat exchanger, the amine pumps, the reboiler, the absorber, and the amine piping equipment. It will also lower the electricity consumption of the amine pumps and the energy required for solvent heating. In addition, a smaller absorber will lower the flue gas blower energy requirements and dimensions, as already described. The amine pumps have a minor share in the energy penalty (3%), but a much larger share (12%) in the capital costs of the capture plant. The costs of the lean-rich HEX are relatively small, about 2% of the absorption plant costs. For a specific solvent, the optimal value of the effective loading is determined by the combined effect of the following process parameters: absorption temperature, and desorption temperature, sorbent concentration.

The *absorption temperature* determines the costs and the energy requirements of the flue gas coolers. Bringing the absorption temperature closer to the desorption temperature will decrease the costs of the solvent heat exchanger. The costs of the solvent heat exchanger and flue gas coolers, as well as the energy consumption of the flue gas cooler, are relatively low. Because of the logarithmic nature of the solvent-loading curve, a large decrease in the absorption temperature at the optimal loading will cause a minor increase in effective solvent loading, but a large increase in the size and

capital costs of the flue gas coolers and the solvent heat exchanger.

The *desorption temperature* determines the temperature of the steam that should be extracted from the steam turbine. As shown by Bolland and Undrum (2003), there is a strong dependency between the energy output of the steam cycle and the steam temperature. Moreover, a smaller difference between absorption and desorption temperatures will lower the capital costs of the heat exchanger. For specific solvents, lowering the desorption temperature will mean a higher lean loading and thus a smaller effective loading, as discussed.

Because sorbents are corrosive, there are limitations with regard to *sorbent concentration*. Maximum sorbent concentrations can, however, be increased when corrosion inhibitors are used. It should be noted that there is, at present, no consensus whether the main cause of corrosiveness is the concentration of CO₂ (Kohl and Nielsen, 1997) or the degradation products of the amines (Chakravarti et al., 2001). Whatever the cause, high corrosion leads to high O&M costs and higher equipment costs.

Solvent stability influences the solvent replacement costs. These costs are, in turn, strongly related to the *solvent costs*. The share of the solvent replacement costs is about 8% of the incremental COE. Degeneration losses are responsible for a relatively small portion of the solvent losses (15–25%), the main part being caused by evaporation losses in the stripper and absorber. More important is that the degradation products are thought to play an important role in corrosion (Chakravarti et al., 2001). The corrosive effects particularly affect the strippers, reboilers, and lean-rich heat exchangers. Lower degradation rates will thus imply lower O&M costs and lower capital costs of critical parts.

2.2.2. Absorber

The absorber is a packed column. As stated above, the required surface area of the absorber is proportional to the absorption rate of the solvent. Absorber parameters that have a large influence on costs and energy use of the entire absorption plant are predominantly the specific surface area and the pressure drop. The *specific surface area* is the absorber area per volume of packing. In principle, a higher specific surface area means a smaller vessel size, which lowers the capital costs. Random column packings were used in the first amine plants. For new technology (e.g., Econamine FG), however, these packings are exchanged for structured packings, which have a higher specific area. A new type of packing with an even higher specific surface area has been proposed for the future: membrane contactors (DeMontigny et al., 2004; Feron and Jansen, 2002; Gronvold et al., 2005). The increase in specific area will result in a lower *pressure drop* in the absorber. This, then, results in lower energy consumption by the flue gas blower (6% of the energy penalty in our calculations) and lower capital costs of the flue gas blower (3% of the capital costs in our calculations).

2.2.3. Modelled relationships

The relationships between the process parameters and the capital cost parameters of the capture process that were not used in our model calculations are represented in Fig. 2 by the dotted lines. Below, we present the arguments behind

the simplifications. We also discuss the assumptions we made about the relationships that were not included in the model.

- We combined lean loading and rich loading to one parameter: the effective absorbent loading. We think that making assumptions for optimal lean and rich loadings is too specific for hypothetical solvents.
- Due to the simplification of the loading, the parameters coupled to both lean and rich loadings cannot be quantified in our model. Assumptions for the values of absorption and desorption temperatures and absorption and desorption rates are described in more detail in Section 3.2.
- The coupling between the absorption and desorption rates and the chemical binding energy cannot be determined because hypothetical solvents were used. These parameters were treated as stand-alone input variables in our model.
- The couplings to corrosiveness are covered by assuming that corrosion will remain constant in the future, even though higher CO₂ concentrations will be applied.
- The specific pressure drop of the packing in the absorber is incorporated in the overall improvement factor describing the pressure drop per CO₂ flux over the absorber area.

2.3. Compression

The compression train is an independent part of the system in which CO₂ is compressed from near atmospheric pressure to 110–150 bar. The compression train is composed of a successive series of compressors and intercoolers. The intercoolers are situated between the compression steps because high-temperature compression is very energy-intensive. The use of more cooling steps will, in general, lower the energy requirement, but increase the complexity and hence the capital costs. This implies that there is an optimal configuration: a configuration with three or four compression steps at capacities applicable for NGCC power-plant capture units (Wildenborg et al., 1999).

Constant energy consumption in terms of kWh per kg CO₂ is assumed for the compression cycle. In the future, part of this consumption can be regained by using the heat from the intermediate coolers for the CO₂ stripping process. The amount of electricity saved in the steam extraction process is subtracted from the power consumption of the compression cycle in our calculations.

3. Input data: present status and development potential

3.1. NGCC

The efficiency of the NGCC is largely dependent on the temperature and pressure of the gas entering the gas turbine. State-of-the-art power cycles have a combustion temperature of 1475 °C (Matta et al., 2000). These temperatures are constrained by both the construction material characteristics and the formation of more NO_x at higher temperatures. The introduction of new materials and new firing technologies is required to allow temperatures up to 1850 °C (Rao et al., 2002). Cooling is required in the combustion chambers and the first-

stage nozzle and results in a large temperature drop before the gas enters the GT. The turbine inlet temperature (TIT) of large state-of-the-art turbines ranges from 1280 °C (Siemens-Westinghouse W501G) to 1320 °C (General Electric F-type). A TIT as high as 1430 °C can be achieved at the same firing temperature with improved cooling techniques (Matta et al., 2000). This technology is expected to be commercially available in the H-type turbine around 2008. Projected developments in future turbine inlet temperatures are in the range of 1700–2000 °C (Rao et al., 2002), although no time horizon has been given for these developments. Another important parameter for efficiency is the compression factor. This factor in state-of-the-art turbines varies from 1:15 (F-type) to 1:19 (W501G) and will be around 1:23 in the new H-type turbine.

A higher TIT will cause the gas turbine exhaust temperature to increase. According to the IEA (IEA GHG, 2004), this will make the application of the so-called once-through design for the HRSG possible. The once-through design is structurally simpler than the design of conventional HRSGs because no drum is required (Alstom, 2002). Moreover, it will lead to reductions in the capital costs. This design also makes the use of supercritical steam turbines possible, further increasing the combined cycle efficiency. According to Rao et al. (2002), a combined cycle based on gas turbines with a firing temperature around 1850 °C may reach an LHV efficiency of around 65%. In the IEA 2020 case (IEA GHG, 2004), a net efficiency of 61.3% was assumed with a possible range of 60–65%. We used the state-of-the-art Siemens Westinghouse W501G and General Electric F-type turbines for our short-term calculations (efficiency: 56.9%), the General Electric H-type turbine with a cycle efficiency of 60% for the medium-term calculations (IEA GHG, 2004), and the same turbine but with a further improved efficiency (62%) (IEA, 2004) for the long-term calculations.

3.2. Chemical absorption unit

Table 2 shows a subdivision of the incremental COE due to CO₂ capture by post-combustion chemical absorption from an NGCC. We calculated that, in the short term, energy losses cause about 50% of the incremental COE, incremental capital costs 25%, and the O&M costs the remaining 25% when the compression train was included. Roughly 60% of the energy penalty is caused by regeneration energy requirements, more than 25% by the energy used by the compression cycle, 7% by the flue gas blowers, and 8% by amine pumps and cooling water pumps.

3.2.1. Regeneration energy

About 60% of the energy penalty is caused by the regeneration energy requirements of state-of-the-art technology, including CO₂ compression. Based on conventional MEA, Chakma and Tontiwachwuthikul (1999) claimed that the regeneration energy contributes 35–40% to the incremental COE of the scrubbing systems. The energy consumption for regeneration can be split into three parts:

- Energy required to reverse the chemical binding of CO₂ to the solvent.
- Energy required to heat the solution.
- Water evaporation.

Table 2 – Development items classified by cost item

Cost items	Development items	Secondary effects	
Energy costs (~50%)			
Regeneration energy (~60%)	CO ₂ binding energy	Solvent formulation	'Random' effects on many solvent properties (+/-)
	Solution heating	ΔT	Lowers cost of heat exchangers & reboiler
		CO ₂ loading	Lower circulation rate lowers costs and energy requirements of amine pumps
		Specific heat	This is already a secondary effect of changing solvent formulation
	Water evaporation	Lower desorption temperature	
	Solvent heating efficiency	Reboiler design Heat exchanger design Layout/heat integration	Capital cost of reboiler Capital cost of heat exchanger Capital cost of equipment, O&M costs, etc.
		Temperature level of regeneration process	Lean loading
Parasitic power loss (~15%)	Pumps	CO ₂ loading	Lower regeneration energy requirement
	Blower	Absorption rate Absorber developments	Capital cost of absorber Capital cost of absorber
Compression (~25%)	Heat integration with stripping process		Capital cost of piping
Capital costs (~25%)	Absorption rate	Lowers capital costs of absorber due to smaller required contact area	Energy requirements and capital costs of blower
	CO ₂ loading	Lowers capital costs of pumps and stripper due to smaller solvent circulation	Energy requirements of pumps
	Corrosive effects	Lowers capital costs of critical parts	Lowers operating costs of critical parts
	Regeneration energy	Lowers capital costs of heat exchangers and reboiler and of additional power production capacity	
	Capital costs absorber	Use of different materials, e.g., concrete instead of steel, or different designs	
	Capital costs of other components	Specific developments in component technology (pumps, blowers, heat exchangers, reboiler, etc.)	
O&M costs (~25%)	Solvent losses	Lowers operating costs of solvent losses and filters in case of solvent degradation	
	Solvent costs	Lowers operating costs of solvent losses	
	Corrosive effects	Lowers operating costs of critical parts	Lowers capital costs of critical parts
	Solvent degradation rates	Lowers costs of filters	Lowers capital costs of reclaimers

The incremental costs of electricity (COE) can be subdivided into the incremental costs due to the energy penalty, the capital costs, and the O&M costs of the capture plant. The shares of these cost items in the short term are shown between parentheses. The energy penalty can be further broken down into regeneration energy, parasitic power loss, and CO₂ compression energy. The shares of these parts within the energy costs are also shown. The most important development items for each part are shown in this table. Moreover, the possible secondary effects (positive and negative) of the development items are included.

Lowering the energy requirements of each of these parts requires different improvements, as is described below. Many potential improvement figures have been presented in the literature. However, most of the references (Chakma and Tontiwachwuthikul, 1999; Gibbins et al., 2004; IEA GHG, 2004; Mimura et al., 2001) do not categorize the possible future

development of regeneration energy requirements into specific process improvements.

3.2.1.1. CO₂ binding energy. The energy required for the reversed reaction of CO₂ binding depends entirely on the solvent formulation. Assuming conventional MEA solvents as

a baseline, Rao et al. (2004) used 1.9 MJ/kg CO₂ as the energy requirement for this reaction. Several attempts have been made to improve solvent formulations or to develop new solvents focusing on this point. Examples include KS-1 (Mimura et al., 1995, 1999, 2001, 2003), PSR (Chakma, 1995; Chakma and Tontiwachwuthikul, 1999), Praxair solvents (Chakravarti et al., 2001), DECAB solvents (Feron and ten Asbroek, 2004), and blends of piperazine and potassium (Cullinane et al., 2004). Cullinane et al. specifically refer to lowered desorption energy requirements: 0.9–1.6 MJ/kg CO₂ compared to 1.9–2.1 MJ/kg CO₂ for MEA. Feron and ten Asbroek (2004) provide the same type of figures. The CO₂ binding in the DECAB process requires 1.5 MJ/kg CO₂ of the total regeneration energy of 2.3 MJ/kg CO₂.

In the present study, we use the CO₂ binding energy of MEA, 1.9 MJ/kg CO₂ for the short-term. For the medium-term we consider the CO₂ binding energy of the DECAB process (1.5 MJ/kg CO₂) feasible. The values reported by Cullinane et al. (2004), may be reached in the long term. Therefore, we assume 1.0 MJ/kg CO₂ for the long-term.

3.2.1.2. Heating of the solution. The general equation that is applicable for the amount of energy required to heat the solution is shown in Eq. (1):

$$Q = c m_s \Delta T \quad (1)$$

where Q is the heat added, c the specific heat of the solution, m_s the mass of solution, and ΔT is the change in temperature. Because we were specifically interested in the energy requirements per amount of CO₂, we needed to adjust the equation to the one shown here:

$$Q_{\text{CO}_2} = c \frac{m_s}{m_{\text{CO}_2}} \Delta T \quad (2)$$

From Eq. (2), it is clear that the most important parameter is the amount of CO₂ handled per kilogram solution in each cycle. The possible developments in this parameter are described in Section 3.2.3.

The temperature difference in the absorber depends on solvent characteristics and the efficiency of the solvent heat exchanger. The ΔT in state-of-the-art MEA processes is around 15 K. To lower this figure, ΔT in the solvent heat exchanger should also be reduced. The required heat exchanger area increases exponentially when ΔT is reduced. We assumed a ΔT of 15 K in the short-term calculations, 10 K in the medium term, and 5 K in the long term, based on expert opinions (Feron, 2006). It should be noted that ΔT is not an independent process parameter. Lowering the ΔT in the stripper will generally imply lower effective loadings when using the same solvent. This effect can be reduced by using split-flow configurations or multiple strippers. New stripper designs, however, are necessary in order to reach the long-term ΔT values without large HEX area requirements. An example can be found in the DECAB process, where a stripper-integrated heat exchanger is proposed (Feron, 2006).

Additionally, the specific heat of a solution will also change when the composition is changed. As illustrated in Table 3, there is a large range in the values of the specific heat of the different absorbents. Because a relatively large part of the

Table 3 – Specific heat of common ethanolamines and water

	Specific heat (J g ⁻¹ K ⁻¹)
Monoethanolamine (MEA) ^a	2.74
Diethanolamine (DEA) ^a	2.47
Triethanolamine (TEA) ^a	2.34
Methyldiethanolamine (MDEA) ^a	1.72
Piperazine ^b	2.57
Water	4.19

It should be noted that future solvents may include compounds other than ethanolamines.

^a Source: http://www.kadakiagroup.com/NIL_EXP1.htm.

^b Source: http://www.hejia-china.com/cgi/search-en.cgi?f=company_en_1_+product_en_1_&id=384126&t=product_en_1_.

solution is amine (30 wt% in state-of-the-art MEA processes), the choice of absorbent can have significant effects on the specific heat of the solution. With increasing absorbent concentrations, the importance of the specific heat of the solvent grows. Since specific heat has less influence on energy consumption than other solvent characteristics, it is not considered to be a development item at present. As a result, we assumed this figure to remain around the level of MEA.

A process improvement called split flow, which is included in the new Econamine FG+ process (IEA GHG, 2004), involves a semi-lean stream taken from the stripper column. This semi-lean stream is introduced half way down the absorber column. Because this stream is not as hot as the lean stream, less solvent heating energy is required in the process. According to Aroonwilas (2004), a ~30% reduction in the regeneration energy requirements could then be achieved. It should be noted that this process development causes a larger complexity and larger amine flows, which will add capital costs and parasitic power loss. When new solvents with lower regeneration energy requirements are introduced in the future, the balance between regeneration energy improvements and incremental capital costs, O&M costs, and parasitic power loss will make this configuration less interesting. Therefore, and because we could not find more detailed information on the split-flow process, we did not evaluate this development in our further analysis.

Using Eq. (2) and the input data summarized in Section 3.4, we calculated a value of 1.09 MJ/kg CO₂ for solution heating in the short term, 0.40 MJ/kg CO₂ in the medium term, and 0.16 MJ/kg CO₂ in the long term.

3.2.1.3. Water evaporation. The amount of water evaporated per mole CO₂ extracted in the stripper is affected by CO₂ loading and the stripping temperature. Although water evaporation in the stripper is responsible for a relatively large share of the energy requirements, hardly any information has been presented in the literature. In 1995, Feron and Jansen (1995) stated that, in state-of-the-art MEA processes, 1–2 moles of water are evaporated for each mole CO₂ captured, which corresponds to an energy consumption of 0.93–1.85 MJ/kg CO₂. We used the average of these values as the short-term number. The number of moles of water evaporated per mole CO₂ in an IEA study (IEA GHG, 2000a) was between 0.72 and 0.76, corresponding to an energy consumption of about 0.7 MJ/kg CO₂. We used this number as our medium-term value.

Feron and ten Asbroek (2004) claimed that this figure is only 0.41 MJ/kg CO₂ in the DECAB process, which we assumed to be achievable in the long term.

3.2.2. Heat integration

The regeneration energy requirements as described in the previous paragraphs are expressed in MJ/kg CO₂. To translate this value into a number expressed in kWh energy loss per kg CO₂, insight is required into the energy transfer efficiency in the reboiler and the steam turbine efficiency at the specifications of the steam used for regeneration.

The energy transfer efficiency is already near optimal. It includes factors like temperature difference between steam and stripper temperatures and temperature losses occurring during the transport of the steam from the ST to the reboiler. This efficiency could be improved by a better integration of the reboiler into the HRSG. Dividing the energy output per mole of steam by the energy content of the steam at the same condition gives the so-called power factor. Bolland and Undrum (2003) showed the dependency of this factor on steam temperature. As shown in Table 4, a relatively wide range of figures can be found in the literature. Because no large developments in low temperature steam turbines are expected, the main influencing parameter is steam temperature. The required steam temperature is, in fact, dependent on the regeneration temperature and the temperature difference between steam temperature and regeneration temperature. The temperature in the stripper affects the lean loading for each specific solvent. This means that an optimum must be found. Solvent developments can,

however, lower the temperature required in the stripper. For example, the KS-1 solvent preserves energy consumption by its lower regeneration temperature (110 °C instead of 120 °C) (Mimura et al., 1995).

We used the figures of Bolland and Undrum (see Table 4) in our further analysis. We assumed that solvent improvements allow a reduction in required steam temperature from 130 to 120 °C in the medium term and 110 °C in the long term. Using the relationship from Bolland and Undrum (2003), this assumption results in the following values for the power equivalent factor: 0.203 for the short term, 0.186 for the medium term, and 0.167 for the long term. The figure estimated by Feron (2005) as well as the figures recalculated from Hendriks (1994) and Mimura et al. (1995) showed much larger improvements. These values were included in our uncertainty analysis.

3.2.3. CO₂ loading

The total effective CO₂ loading is a product of the sorbent concentration in the solvent and the working capacity of that specific solvent. The maximum concentration depends on the sorbent used, as shown in Table 5. Specific absorber developments and the addition of corrosion inhibitors can, however enable higher solvent concentrations. With regard to CO₂ absorption using MEA, solvent concentrations of about 20% were feasible in the 1970s; today, concentrations of 30–35% are feasible. The difference between lean loading and rich loading is called the working capacity and is expressed in moles CO₂/moles amine.

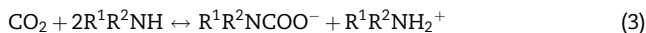
Table 4 – Values of the power factor cited in the literature

Factor	T _{reg} (°C)	Source	Status
0.203	130	Bolland and Undrum (2003)	Calculated figure
0.186	120	Bolland and Undrum (2003)	Calculated figure
0.167	110	Bolland and Undrum (2003)	Calculated figure
0.094		Mimura et al. (1997) Plant 1	Own recalculation
0.098		Mimura et al. (1997) Plant 2	Own recalculation
0.123		Mimura et al. (1997) coal-based	Own recalculation
0.107	133	Hendriks (1994) MEA optimized	Own recalculation
0.097	128	Hendriks (1994) DEA optimized	Own recalculation
0.14		Rao et al. (2004) new plant, based on 0.09–0.19 from literature: (Bolland and Undrum, 2003; Hendriks, 1994; Mimura et al., 1997)	Literature-based average
0.22		Rao et al. (2004) retrofit, based on 0.20–0.25 from literature	Literature-based average
0.329	131	Buchanan (2000) F-type, steam temperature level 326.2 °C	Calculated figure
0.316	131	Buchanan (2000) H-type, steam temperature level 326.2 °C	Calculated figure
0.25		Feron (2005) 1995 figure	Rough estimation
0.20		Feron (2005) 2005 figure	Rough estimation
0.15		Feron (2005) 2015 figure	Rough projection
0.219	122	Alie (2004) page 82, steam temperature level is 132 °C	Calculated figure

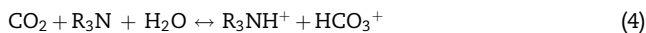
Table 5 – Characteristics of several amines (Chakma and Tontiwachwuthikul, 1999)

Amine	Maximum sorbent concentration (wt%)	Moles of CO ₂ per mole of amine		
		Lean loading	Rich loading	Working cap
MEA	30	~0.15	~0.45	~0.30
DEA low pressure	40	~0.07	~0.40	~0.30
DEA high pressure	40	~0.10	~0.80	~0.70
DGA	60	~0.10	~0.45	~0.35
DIPA high pressure	40	~0.05	~0.50	~0.45
MDEA high pressure	50	~0.05	~0.50	~0.45

The loading for primary and secondary amines has a theoretical maximum of 0.5 mole CO₂/mole amine due to the overall carbamate formation reaction presented in Eq. (3):



where R² is hydrogen in case of a primary amine. The reaction of tertiary amines as well as sterically hindered amines, however, is a base-catalyzed hydration of CO₂ as shown in Eq. (4)



This allows a theoretical maximum loading of up to 1 mole CO₂/mole amine. As solvent formulations are confidential for most new solvents, the loading figures of these solvents are presented in the literature in terms of amount of CO₂ per volume of solvent. MHI's solvents, PSR solvents, and TNO's DECAB solvent all have a higher possible loading than state-of-the-art MEA-based solvents. Feron and ten Asbroek (2004) reported a solvent loading of 92 kg/m³ for the DECAB process compared to a solvent loading of 55 kg/m³ for MEA. Chakma and Tontiwachwuthikul (1999) claim that the PSR solvent has 120–180% the solvent-loading capacity of state-of-the-art MEA-based solvents and Mimura et al. (1997) reported that the KS-1 solvent has a 'much higher' solvent-loading capacity than state-of-the-art MEA-based solvents. A few also reported future projections of the solvent concentration as a separate variable: Rao et al. (2004) for example, reported that experts project a relative improvement in sorbent concentration of, on average, 23% (best guess) to 80% (most optimistic) for 2015; no figures were given on effective solvent loading. The IEA report (IEA GHG, 2004) mentioned a lowering in the circulation rate by one third for 2020, which would mean a 50% improvement in CO₂ loading.

The short-term value of the CO₂ loading was calculated from a 30% MEA solvent with a working capacity of 0.24, which results in a 5.2 wt% effective CO₂ loading. This is close to the value mentioned by Feron and ten Asbroek (2004) (5.5 wt%) for MEA-based processes. With regard to the medium term, we adopted Feron and ten Asbroek's (2004) 9.2 wt% as the value for DECAB solvent. This is close to the 80% maximum improvement claimed by Chakma and Tontiwachwuthikul (1999) for PSR solvents. A 50% improvement in CO₂ loading as suggested by the IEA (IEA GHG, 2004) would result in an 8.3 wt% loading if Feron and ten Asbroek's value is used as the basis. We use the latter value as the lower value in our uncertainty analysis. For the long term, we considered a sorbent concentration of 54%. This was based on 80% improvement, which is the average of the most optimistic values given by the experts interviewed by Rao et al. (2004). We combined this with a future working capacity of 0.3 to yield a cyclic CO₂ concentration of 11 wt%.

3.2.4. Absorption rate

Improvements in absorption rate can be achieved with new solvents. According to Cullinane et al. (2004), the absorption rate of a piperazine/potassium carbonate solvent at 60 °C is 1.5–2 times as high at the partial pressure range in the absorber. At lower absorption temperatures, this ratio is larger, reaching 5:1 at 40 °C. Chakma and Tontiwachwuthikul

(1999) posed that a 'hypothetical perfect solvent' could score a mass transfer coefficient of 3.79 kmole h⁻¹ m⁻³ kPa⁻¹ in an absorber filled with random packing. In comparison, MEA processes have mass transfer coefficient values in the range of 0.4–0.6 kmole h⁻¹ m⁻³ kPa⁻¹ with modern random packings (Aroonwilas et al., 1999). Based on this limited data set, we can assume a doubling in the absorption rate in the medium term and another doubling in the long term.

3.2.5. Pressure drop in the absorber

Because gas blowers can be considered well developed, the entire improvement potential for the energy requirements (and capital costs) of the flue gas blower lies in reducing the pressure drop in the absorber. The IEA assumes that it is possible to reduce the pressure drop by 50% by the year 2020 (IEA GHG, 2004). A decline in pressure drop can be induced by an increased absorption rate of the solvent, which lowers the required surface area of the packing, and by specific packing developments, which lower the pressure drop per theoretical stage. Possible improvements in the absorption rate have been described in Section 3.2.4.

The developments in packing design towards reduced pressure drops primarily focus on the switch from random packings to structured packings and, in the future, possibly to membrane-based packings. This development will also lower the volume of the packings and, hence, the costs of the absorber vessel. Due to the higher volumetric costs of structured packings and membrane packings, the total costs of the packings will in general be higher, which is at present the limiting factor for membrane gas absorption systems. Structured packings are used instead of random packings in the new Econamine FG+ concept (Roberts et al., 2004). According to the range of estimates from various sources, this should improve the absorption rate coefficient per volume absorber packing from 0.2–0.6 to 1.0–1.4 kmole h⁻¹ m⁻³ kPa⁻¹ (Aroonwilas et al., 1999; DeMontigny et al., 2004; Nishikawa et al., 1995). In the future, it might also be possible to apply membrane contactors in the absorbers. This might increase the absorption rate to 3.0 kmole h⁻¹ m⁻³ kPa⁻¹ (DeMontigny et al., 2004) or even 27 kmole h⁻¹ m⁻³ kPa⁻¹ (Nishikawa et al., 1995). The only figure we could find in the literature with regard to the developments in pressure drop per meter packing was in Yagi et al. (2004), who reported an improvement from 0.076 kPa/m for random packings to 0.294 kPa/m for a structured packing.

Improvements in pressure drop in the absorber comprise a combination of the solvent-specific absorption rate, the effective absorption area per volume packing, and the pressure drop per meter packing. According to Chapel et al. (1999), the maximum diameter of the absorbers is more or less set, which means that no additional reduction in power loss can be achieved by applying larger column diameters. The improvement factors for the parameters are shown in Table 6.

We assumed a short-term pressure drop of 48 mbar in the absorber, taken from Rao et al. (2004). This value might be relatively conservative when one considers Yagi et al.'s (2004) value for the operational KS-1-based plant (8.8 mbar for random packing and 2.3 mbar for structured packing, assuming a 30 m high packing) and the fact that structured packings are used in the near commercial Econamine FG+ system

Table 6 – Improvement factors for parameters influencing a pressure drop in the absorber

	Improvement factor in the medium term	Improvement factor in the long term
Solvent-specific absorption rate	2	4
Effective absorption area per packing volume	1.2	1.6
Pressure drop per meter of packing	3.9	3.9
Maximum potential combined effect on pressure drop	9.3	24.8
Based on references cited in Sections 3.2.4 and 3.2.5.		

(Gibbins et al., 2004). We used an improvement factor of 9 as the medium-term value, which resulted in a pressure drop of 5 mbar, and an improvement factor of 25 as the long-term value, which resulted in a pressure drop of 2 mbar. The medium- and long-term values were close to the value for structured packings reported by Yagi et al. (2004). The application of membrane gas absorption technology might reduce these figures even more.

3.2.6. Flue gas recycle

The high energy requirements of CO₂ capture from NGCC flue gas is mainly caused by the low partial pressure of CO₂. One way to tackle this problem is to increase the CO₂ concentration in flue gas by recycling the flue gas. This can be done by feeding part of the flue gas stream into the combustion chamber of the GT. When half of the flue gas is recycled, the CO₂ concentration will be doubled. Because of flame-stabilization problems at low oxygen concentrations in the combustion chamber, this is the maximum amount of recycling that is acceptable with current GT technology

(IEA GHG, 2000a). The IEA (IEA GHG, 2000b) also reported that higher recirculation rates (up to 60%) are possible with limited adaptations. Although the ST output will increase by around 10% due to lower regeneration energy requirements, the overall efficiency improvement is only 2.5% (1.2% points) (IEA GHG, 2000a). This is due to a lower output of the GT, the power required for the recycle flow fan, and an increase in the cooling water flow necessary to cool the recycle stream. Although adaptations are required for the combustor, the influence of recycling on the capital costs of the GT would be marginal and the capital costs of the capture system could decrease by almost 30%, mainly because of lower column costs and a smaller reboiler size. The overall specific capital cost reduction (€/kWe) due to 50% flue gas recycling would be around 9% and the incremental capital costs would decrease by about 19% (IEA GHG, 2000a). We quantified the possible effects of the application of flue gas recycling by applying the percentile reductions in incremental capital costs and energy penalty reported by the IEA (IEA GHG, 2000a). The impact of flue gas recycling may be very different with future improved solvents. Moreover, the impact on performance using future GT technology is relatively unsure. Therefore, we present the possible future impact of applying flue gas recycling in separate tables instead of implementing the impact in the basic results.

3.3. Compression train

With regard to the energy required for compression, we refer to the relationship between end pressure and compression energy presented by Bolland and Undrum (1999) with figures similar to those reported by Wildenborg et al. (1999). The relationship is also similar to the one used by Damen et al. (2006). The IEA uses a figure 7% above the relationship used by Bolland and Undrum (IEA GHG, 2004) and EPRI 12% below that line (Buchanan, 2000). Remarkably, the relationship

Table 7 – Input parameters for the energy part of the Excel model

Parameter	Dimension	Short term (2010)	Medium term (2020)	Long term (2030)
1. Regeneration energy ^a (calculated)	MJ/kg CO ₂	4.4	2.6	1.6
1.1 Chemical binding energy	MJ/kg CO ₂	1.90	1.50	1.00
1.2 Solution heating ^b (calculated)	MJ/kg CO ₂	1.09	0.40	0.16
1.2.1 Specific heat of amine ^c	J g ⁻¹ K ⁻¹	2.74	2.74	2.74
1.2.2 Solvent concentration	wt%	30%	37%	54%
1.2.3 Effective solvent loading	Fraction	0.24	0.35	0.50
Effective solvent loading	wt%	5.2%	9.2%	11.0%
1.2.4 ΔT in stripper	K	15	10	5
1.3 Evaporation energy	MJ/kg CO ₂	1.39	0.70	0.41
2. Steam temperature level	°C	130	120	110
3. Power equivalent factor	Factor	0.203	0.186	0.167
4. Pressure drop in absorber	mbar	48	15	8
5. Compression energy	kWh/kg CO ₂	0.112	0.112	0.112
6. Efficiency of NGCC	%LHV	56.9	60.0	62.0

Calculated figures are shown in bold italics.

^a The values shown here were calculated using the parameters 'chemical binding energy', 'solution heating', and 'evaporation energy'.

^b Solution-heating figures were calculated using the figures 'specific heat of amine', 'solvent concentration', 'effective solvent loading', and 'ΔT in stripper'.

^c Because the specific heat of the solvent is not a specific development item, we used the specific heat of MEA as the constant. It should be noted that solvents with a lower specific heat might be used in the future.

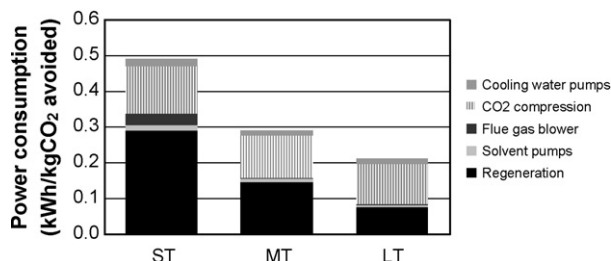


Fig. 3 – Developments in power consumption. Figures for the short-term (ST), medium-term (MT) and long-term (LT).

presented by Bolland and Undrum in a later publication (Bolland and Undrum, 2003) is 17% lower than in their 1999 paper.

In a recent IEA report (IEA GHG, 2004), no developments for compression energy were assumed in the time frame of that study (2005–2020). Referring to Koss (2005), the compression train itself is nearly optimized and only integration issues play a role, which is confirmed by Fisher et al. (2005). According to Gibbins et al. (2004), the heat cannot be used in the NGCC because it is lower than the lowest pinch temperature; therefore, the heat can only be used within the capture process. The integration of the compression train with the capture plant will reduce the steam requirements for the reboiler, but not the energy requirements for the compression. Nevertheless, we present the reduction in energy penalty due to heat integration with the compression cycle as a reduction in compression energy requirements.

Based on the Aspen^{plus} simulation of the compression cycle, we calculated that 25% of the heat can be reused in the medium term calculations. This number increases to 35% for long-term calculations due to lower regeneration temperatures. Because the actual reduction in power loss due to heat integration of the compression train with the capture plant is also determined by the power equivalent factor, the resulting figures show a 5% reduction in the medium term and a 6%

reduction in the long term. These numbers should be considered as maximum values, because part of this heat might be required for the further removal of water from the CO₂.

3.4. Summary of the input for the Excel model

The input parameters described in this chapter are summarized in Table 7.

4. Results of performance calculations

4.1. Results

The results of our cycle performance model are presented in this section. Fig. 3 shows the calculated power loss per amount of avoided CO₂ emissions for the three time frames used in the study. The calculated figure is actually built up from the power loss of the ST due to steam extraction for the regeneration process and parasitic power loss from the cooling water pumps, CO₂ compression, flue gas blower, and solvent pumps compared to the short-term calculations, we calculated a 41% reduction in the medium term and a 57% reduction in the long term. The reduction in power loss due to regeneration has a very large share in the possible overall reduction due to the large initial share (59%) and the large future reductions (49% in the medium term and 73% in the long term). The significant reductions in the energy requirements of the solvent pumps and flue gas blower have smaller effects due their lower initial share in the energy requirements.

The projected reduction of the energy penalty is given in %-points in Fig. 4a. We found that the energy penalty may decrease from 7.9%-points in 2010 to 4.9%-points in 2020 and 3.7%-points in 2030. Fig. 4b shows the projected improvements in net cycle efficiency including capture. Based on an NGCC efficiency of 56.9%, we calculated a net efficiency of 49.0% for 2010. The 2020 efficiency was 55.1%, based on an NGCC efficiency of 60.0%, and the 2030 efficiency was 58.3%, based on an NGCC efficiency of 62.0%. The error bars in both graphs

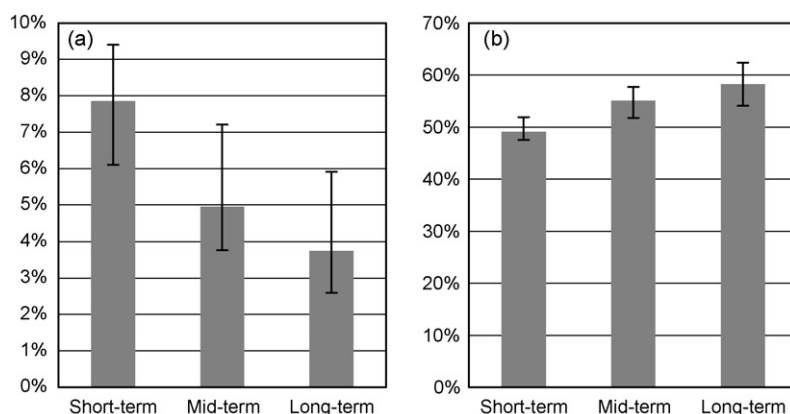


Fig. 4 – Future reductions in energy penalty (a) and improvements in the net cycle efficiency including capture (b). The calculated data are shown in %-points LHV and %LHV, respectively. The error bars represent the spread in results using extreme values of all parameters in both negative and positive directions.

Table 8 – Input parameters for the uncertainty analysis

	Short term (2010)			Medium term (2020)			Long term (2030)		
	Lower limit	Standard value	Upper limit	Lower limit	Standard value	Upper limit	Lower limit	Standard value	Upper limit
Chemical binding energy (MJ/kg CO ₂) ^a	1.90	1.90	1.90	1.90	1.50	1.40	1.50	1.00	0.90
Specific heat amine (J g ⁻¹ K ⁻¹) ^b	2.74	2.74	2.74	3.00	2.74	1.72	3.00	2.74	1.72
Effective solvent loading (%) ^c	5.2	5.2	6.1	8.3	9.2	11	8.3	11	19
ΔT stripper (K) ^d	20	15	15	15	10	8	15	5	2
Evaporation energy (MJ/kg CO ₂) ^e	1.85	1.39	0.7	1.4	0.70	0.41	0.93	0.41	0.2
Power equivalent factor ^f	0.22	0.203	0.186	0.203	0.186	0.15	0.186	0.167	0.094
Pressure drop (mbar) ^g	70	48	48	48	5	2.3	48	2	2
Compression energy (kWh/kg CO ₂) ^h	0.12	0.112	0.095	0.115	0.107	0.090	0.114	0.106	0.089
Efficiency NGCC (%LHV) ⁱ	56.9	56.9	58.0	59.0	60.0	61.5	60.0	62.0	65.0

The upper limit values resulted in lower energy penalties, the lower limit values in higher efficiency penalties.

^a MEA was the absorbent for the short-term calculations. For the medium term, the lower limit was based on MEA (Cullinane et al., 2004; Rao et al., 2004) and the upper limit on the average between the upper limits of the short and the long term. For the long term, the lower limit was based on the DECAB solvent and the upper limit on the upper limit for blends of piperazine and potassium reported by Cullinane et al. (2004).

^b We assumed the use of MEA for the short term. Other absorbents were used for the medium- and long-term calculations. We used the value of MDEA for the upper limit and a rough value above the MEA value for the lower limit.

^c The short-term values were based on a loading of 0.24 and MEA concentrations of 30–35%. The 50% CO₂ loading improvement from the IEA (IEA GHG, 2004) was used as the lower limit for the medium term. The upper limit was based on an 80% improvement (Chakma and Tontiwachwuthikul, 1999) of the upper limit of the short term. We used the lower limit of the medium term for the long term as well. The upper limit was based on an effective loading of 0.5, an absorbent concentration of 54%, and an 80% improvement (Rao et al., 2004).

^d The short-term variation was based on the variation in ΔT from Aspen^{plus} calculations by Mohammad Abu Zahra from TNO, Apeldoorn, the Netherlands. The medium- and long-term variations came from expert opinions. The upper limit for the long term was based on hypothetical values for membrane gas absorption.

^e The short-term lower limit was based on Feron and Jansen (1995), the upper limit on an IEA report (IEA GHG, 2000a). For the medium term, the lower limit was based on the average value from Feron and Jansen (1995), the upper limit on the value for the DECAB process (Feron and ten Asbroek, 2004). The upper limit from Feron and Jansen (1995) was used as the lower limit for the long term and a rough estimation was made for the upper limit.

^f For the short term, we used the relationship presented by Bolland and Undrum (2003) with a steam temperature spread of 120–140 °C. For the medium term, we used the same relationship for the lower limit and a steam temperature of 130 °C. The upper limit is taken from Feron (2005). The lower limit for the long term was based on the relationship presented by Bolland and Undrum using a steam temperature of 120 °C and the upper limit was based on Mimura et al. (1997).

^g We used the value from Won et al. (1999) as the lower limit for the short-term. The value from Rao et al. (2004) was used as the short-term's upper limit and the long-term lower limit. The value for structured packings in Yagi et al. (2004) was used as the upper limit for the medium term. The limits for the long term were given in Rao et al. (2004) and the same value divided by an improvement factor of 25 described in Section 3.2.5.

^h The variation in compression energy was based on the variation found in several references, with the upper limit coming from the IEA (IEA GHG, 2004) and the lower from Bolland and Undrum (2003). The reduction in energy requirements due to heat integration of the compression cycle and capture unit was applied to the limits.

ⁱ The short-term lower limit was based on a Westinghouse G turbine, the upper limit on F turbine technology (IEA GHG, 2000a). The latter value was also used as the lower limit for the medium term. The medium-term value was based on GE H technology (Buchanan, 2000; Matta et al., 2000). The long-term values and range were based on several efficiency improvement estimations found in the literature (Hendriks et al., 2004; IEA GHG, 2004; Rao et al., 2002).

depict the range of possible values based on the extremes of all parameters. The assumptions for the uncertainty analysis are shown in Table 8.

Fig. 5 shows the impact of each single parameter on the conversion efficiency of the plant, including capture. The light grey bars show the short-term values. The mid-tone grey bars show the net efficiency improvement when only the medium-term development of a specific parameter is taken into account and the other parameters are left unchanged. The dark grey bars are calculated in the same manner using the long-term value of the particular parameter. It should be noted that the power reduction due to regeneration is calculated by multiplying the regeneration energy requirements with the power equivalent factor. This implies that positive developments in one of the regeneration energy parameters (evaporation energy, ΔT stripper, effective solvent loading, specific heat of amine, and chemical binding energy) will lower the net

positive effects of a lower power equivalent factor and vice versa.

The figure also clearly shows that the anticipated increase in the efficiency of the NGCC power plant has the largest effect on the improvements in the total cycle efficiency. Because post-combustion absorption technology does not affect the combustion and gas turbine process parameters, advances in NGCC technology can be fully implemented. It should be noted, however, that these improvements in NGCC technology will not positively affect the absolute energy penalty. If the low-pressure steam-cycle efficiency can be improved, it will have negative effects due to a higher power equivalent factor. In contrast, the relative energy penalty will be positively affected by this parameter. As can be seen, the power equivalent factor, the evaporation energy, the effective solvent loading, chemical binding energy, and ΔT in the stripper are very important factors

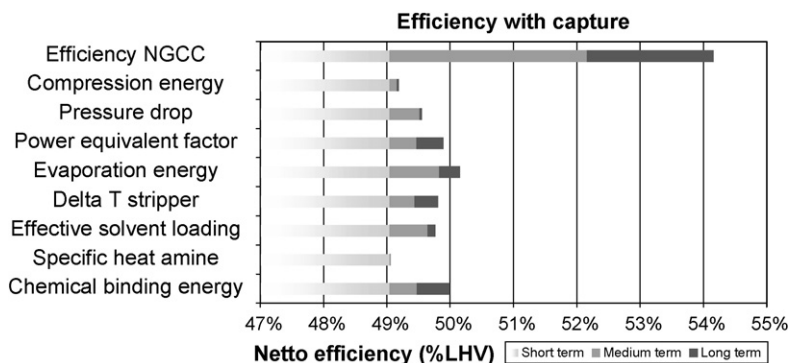


Fig. 5 – Effects on the efficiency of the entire cycle induced by the individual process parameter improvements. The lighter grey bars represent the basic efficiency in the short term; the middle grey-tone bar shows the effect of medium-term improvements of the individual parameter; and the darker grey-tone bar shows the additional effect in the long term.

Table 9 – The possible impact of flue gas recycle on energy performance

	2010	2020	2030
Energy penalty (%point LHV)			
Baseline	7.9	4.9	3.7
Flue gas recycle	6.8	4.3	3.3
Net cycle efficiency (%LHV)			
Baseline	49.0	55.1	58.3
Flue gas recycle	50.1	55.7	58.8

Based on percentile reductions of incremental capital costs and energy penalty taken from IEA (IEA GHG, 2000a).

for energy penalty reductions. The decline in the pressure drop in the absorber can play an important role in the medium term as well by applying structured packing. In the long term, membrane absorbers may lower the energy penalty even further. This is not, however, included in the figures.

No flue gas recycling was taken into account in the results described above. The possible impact of the application of 50% flue gas recycling in the short, medium, and long terms on the energy penalty and the cycle efficiency including capture is shown in Table 9.

4.2. Uncertainty analysis

The full range of possible values for the net calculated efficiency inflicted by the uncertainties in each separate parameter is shown in Fig. 6. The input values for this graph as well as for the uncertainty bars in Fig. 4 can be found in Table 8. The input values are based on the range found in the literature. Because the parameters are interrelated, an equal variation in a single parameter at a later moment in time in many cases results in a smaller range for the net efficiency. The range caused by the uncertainty in the future NGCC efficiency is the largest. Because CO₂ compression should be applied in all CO₂ capture options, the uncertainty in this parameter is of minor importance when comparing CO₂ capture options. It can be concluded from the results that the most important uncertainties in the input data are the uncertainties in NGCC efficiency, the power equivalent

factor, the chemical-binding energy, and the evaporation energy.

5. Results of cost calculations

In our cost calculations, we assumed an interest rate of 10% and a lifetime of 20 years. This results in a capital charge rate of 0.11746. We considered a natural gas price of 4.7 €/GJ and our baseline case assumption for the load was 7600 operating hours at full capacity per year.

5.1. Capital costs

The capital costs of our analysis were calculated from the relationships between the component dimensions and the capital costs reported in the literature (Table 10). To calculate the installed equipment costs, we used the installation factors suggested by Peters et al. (2003), with specific values for each

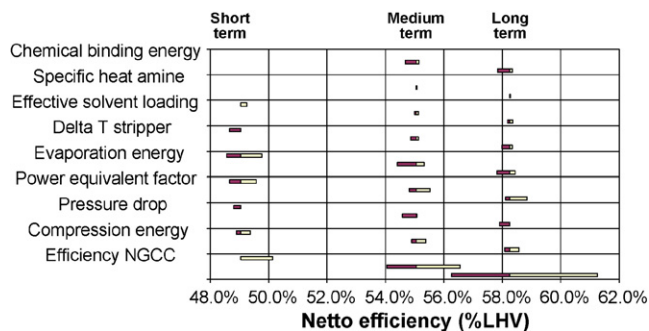


Fig. 6 – Effects on the entire cycle efficiency of the uncertainty of each parameter. For each parameter, the left bar represents the short-term values, the middle bar the medium-term values, and the right bar the long-term values. The dark bar represents the maximum deviation in the negative direction relative to the standard value, the lighter bars the maximum deviation in the positive direction. The standard value for the entire cycle is represented by an imaginary vertical line through the junctions between the dark and light bars.

Table 10 – The source of the relationship between capital costs and component dimensions used in this study

Component	Cost basis	Cost calculation	Based on
Absorber random packing ^a	Packing volume	1740 €/m ³	2-in. Rashig rings (Loh and Lyons, 2002)
Absorber structured packing ^b	Packing volume	3637 €/m ³ (MT), 3273 €/m ³ (LT)	Factor 2.1 between random and structured packing (UOP, 2000)
Absorber column ^c	Column area	1227 €/m ²	Aspen Icarus
Compression cycle ^d	Power for compression (W)	$(5.59 + 5.09P - 0.006P^2) \times 10^6$	IEA GHG (2002)
Gas coolers ^e	Cooler area (m ²)	82.54A + 132,708	Fisher et al. (2005)
Heat exchangers ^f	Heat exchanger area (m ²)	92.0 €/m ²	Fisher et al. (2005)
Blowers	Flue gas volume	$7.536 \times (m^3)^{0.765}$	Aspen Icarus, centrifugal fans
Solvent pumps	Solvent flow (m ³ /h)	Separate formula motor costs and pump costs	Aspen Icarus
Reboiler and reclaimer	Heat requirement/area	317 €/m ²	Aspen Icarus, Kettle type reboiler
Stripper	Solvent flow	3.25 M€ at 167,700 kg/h	Fisher et al. (2005); capacity factor 0.7
NGCC	Cumulative build capacity	Learning rate of 10%	Colpier and Cornland (2002)

^a Other values found in the literature for SS 2-inch rings were 29% higher (Pall rings in (Loh and Lyons, 2002)) and 44% lower (UOP, 2000). The values we used lie in the range between the values for 3.5 and 1-in. packings published by Murratti (2002).

^b Medium-term value was based on the factor 2.1 between structured and random packings reported in UOP (UOP, 2000). The value of this factor in (Murratti, 2002) lies between 2 and 4. We assumed a cost decrease of 10% from the medium to the long term.

^c The specific absorber column costs vary with the size of the column. We used the specific costs in terms of costs per area as suggested by Murratti (2002). The relationship calculated with the use of Aspen Icarus was a close match to the one in Murratti (2002), which was based on data provided by six vendors. Using the specific cost dependency on column volume yielded similar results.

^d The costs calculated by Fisher et al. (2005) are in agreement with the equation given by the IEA. The range of compression powers in Fisher et al. (2005) however, was 34.8–77.8 MW, whereas we considered compression cycles to be in the range of 16.8–18.5 MW.

^e The area of the coolers considered in this study lay well in the range of the cooler areas presented by Fisher et al. (2005).

^f It should be noted that the areas of the heat exchangers considered by Fisher et al. (2005) lay in the range of 25,000–28,000 m², whereas we considered heat exchangers with an area between 12,000 and 17,000 m². Because the calculated relationship has a zero intercept, we considered the cost figures to be reasonably reliable for this range.

type of equipment. It should be noted that no cost developments were assumed in the capital costs of most of the capture plant components. The only component for which a small decrease in costs was assumed was the structured packing of the absorber, because this was the only part that could be considered to be in a relative early stage of development.

The IEA assumed a sharp reduction in the costs (more than 20%) of the NGCC for 2020 (IEA GHG, 2004). For our study, we assumed a more conservative development of the specific costs of the NGCC: 530 €/kW total capital requirements (TCR) in the short term (Damen et al., 2006) to 480 €/kW in the medium term and 450 €/kW in the long term. The cost reductions were based on the forecasted growth of NGCC capacity from 2010 to 2020 (50% growth) and 2030 (100% growth) in the world energy outlook (IEA, 2002). Combined with power replacement figures (IEA, 2004) and our own estimations of the NGCC share in the 2000 production figures, we assumed a 100% growth in the cumulative NGCC production for 2020 and a 200% growth in the production for 2030. We calculated the corresponding costs using a learning rate of 10% (Colpier and Cornland, 2002).

The results of the capital cost calculations are shown in Fig. 7. The data presented here concern installed equipment costs of the capture plant per kW net power of the power plant, including capture. Because the efficiency improvement assumptions of a power plant play a large role in the specific capital cost decline, we report figures that include and exclude these developments. The specific costs were calculated to decrease from 76.0 to 63.2 €/kW in the medium term and 58.6 €/kW in the long term. Excluding the NGCC developments, the medium- and long-term values were 68.6 and 65.0 €/kW, respectively, which means that NGCC efficiency

improvements had a 42% share in the medium term-specific capital cost improvements of the capture plant and 37% in the long term-specific capital cost improvements. Looking at the basic cost calculations including NGCC improvements, the largest potential for capital cost reductions can be expected from the absorber. The relatively large share in the capital cost reduction of the compression cycle was largely due to the lower CO₂ output per kW. The reboiler and reclaimer had a share of about 15% in the cost decline, the stripper a little above 10%, and the gas cooling and the amine pumps just below 10%. Due to the lower assumed temperature difference over the stripper, the capital costs of the heat exchanger

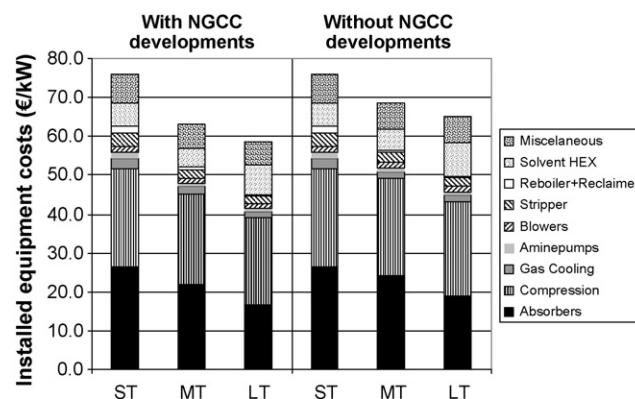


Fig. 7 – Potential development in the specific installed equipment costs of the capture plant. The specific capital cost-reducing effects of the projected NGCC efficiency improvements have been omitted from the right-hand bars.

Table 11 – Basis of the calculation of capital cost factors

Cost factor	Assumed value	Based on source
Installation costs	Component-specific	Peters et al. (2003)
Instrumentation and control	20% of purchased equipment costs	Peters et al. (2003)
Piping	40% of purchased equipment costs	Peters et al. (2003)
Electrical	11% of purchased equipment costs	Peters et al. (2003)
Building and building services	5% of purchased equipment costs	Peters et al. (2003)
Yard improvements	10% of purchased equipment costs	Peters et al. (2003)
Service facilities	25% of purchased equipment costs	Peters et al. (2003)
Land	5% of purchased equipment costs	Peters et al. (2003)
Engineering	7% of total direct costs	Rao and Rubin (2002)
Construction expense	10% of total direct costs	Rao and Rubin (2002)
Contractors fee	0.5% of total direct costs	Rao and Rubin (2002)
Contingency	20% of total direct costs	Rao and Rubin (2002)
Working investment	25% of fixed capital investment	Asselbergs (2005)
Start-up costs + MEA costs	10% of fixed capital investment	Asselbergs (2005)

increase in the long term. The application of a split-flow configuration will lower these costs at the expense of additional costs due to increased complexity.

For the analysis of the COE and the avoidance costs of CO₂ emissions, we applied the assumptions on capital cost factors shown in Table 11.

5.2. O&M costs of the capture plant

5.2.1. Solvent losses

Losses of solvent occur through the degradation of solvents and via vapor/mist from the absorber and the stripper. The proportion of each of these losses is shown in Table 12. The largest part of the lost amines is vented from the absorber together with the flue gas. To reduce the solvent-replacement costs, the largest benefit could, therefore, be expected from improvements to the absorber mist eliminators.

According to Chakma (1995), degeneration losses form a relatively small proportion of the amine losses (25% for MEA, 16% for PSR). There are, however, secondary costs involved. These include O&M and material costs due to the increased corrosive effects, filter systems that capture the degradation products, and the costs of storing the degradation products. It is hard to estimate the cost of corrosion due to degeneration, but we assumed it to be relatively low. It should be noted that the solvent conditions can also influence the losses, as reported for KS-1 by Mimura et al. (1999) and for PSR by Chakma (1995). Experts interviewed by Rao et al. (2004) expected the sorbent losses to decrease by 49% (best guess) to 76% (most optimistic) and the costs of solvent per kilogram sorbent to increase by 48% (best guess) to 3% (most optimistic). The short-term values supplied by the experts ranged from 0.5

to 3.1 kg/tCO₂. The baseline case value assumed by Rao et al. (2004) was 1.5 kg/tCO₂, which we used in this study.

According to NETL (2002), solvent costs are responsible for about 0.12 €/kWh of the COE, which equals about 2% of the total COE of an NGCC plant with CO₂ capture. The contribution to the incremental COE in this study was 7%. The reduction in solvent costs, therefore, can lower the costs only moderately. The costs for more advanced solvents like KS-1 and the DECAB solvent will logically be higher than state-of-the-art MEA. However, the lower losses will compensate largely for the higher costs. According to the IEA (IEA GHG, 2004), the costs due to solvent losses are indeed equal for the cheaper Econamine FG+ solvent and the KS-1 solvent. It should, however, be noted that new, more sophisticated solvents have more potential for cost reductions than MEA, which is a bulk chemical with a more-or-less fixed price. For this study, we used the average of the values for solvent losses and costs given by the experts. For the solvent losses, we used the average of the best-guess values for the medium term and the average of the most optimistic values for the long term. We used the best-guess average for both the medium and the long term for the solvent costs, because we think it is not realistic to combine the lowest cost estimation with the highest developed solvent.

5.2.2. Corrosive effects

The corrosive effects of amine-based solvents increase the capital and O&M costs of a capture system. According to Chakma and Tontiwachwuthikul (1999), corrosion problems can occur throughout a system, particularly in the strippers, reboilers, and rich-lean heat exchangers. Kohl and Nielsen (1997) reported CO₂ as the most corrosive agent, followed by O₂. In their view, the amines were, in principle, not corrosive, but did cause corrosiveness by absorbing CO₂. They presented a ranking in which primary amines cause the highest corrosive effects, followed by secondary amines, and finally tertiary amines. Chakravarti et al. (2001), however, claimed amine degradation products to be very important corrosive agents as well. Chakma (1995) claimed the degradation rate for MEA towards carbon steel to be 18 mpy, while the formulated amine PSR had a corrosion rate of only 2 mpy. The numbers stated by Mimura et al. (1995) show a much higher corrosive effect of MEA: 76–93 mpy. They claimed their solvent KS-1

Table 12 – Solvent losses (kg/tCO₂) in state-of-the-art MEA processes and PSR process

Solvent	Absorber losses	Stripper losses	Degradation losses	Total losses
PSR	0.46	0.08	0.11	0.65
MEA	2.07	0.23	0.80	3.1

Taken from Chakma (1995).

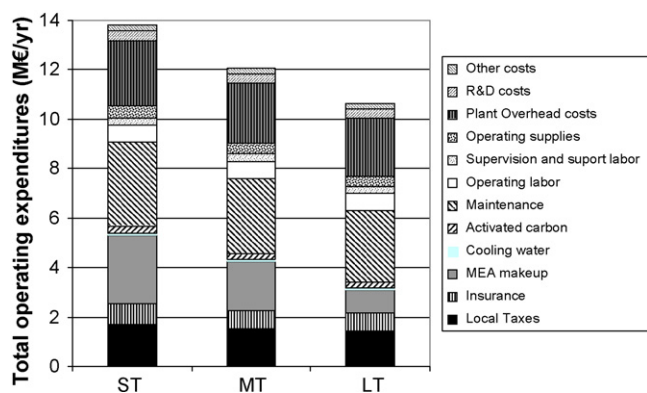


Fig. 8 – Potential development in the O&M costs of the capture plant.

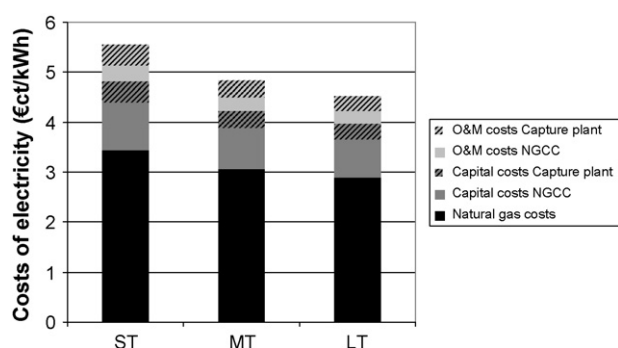


Fig. 9 – The costs of electricity (COE) including capture. The COE are broken down into natural gas costs, capital costs, and O&M costs. The shaded parts represent the capital costs and the O&M costs of the capture plant.

(sterically hindered amine) to have a corrosion rate of 3.1–3.6 mpy. According to Roberts et al. (2004), the corrosion rates of MEA with inhibitors, as formulated for the Econamine FG process as well as the KS-1 solvents, allow equipment to be constructed with carbon steel.

5.2.3. O&M cost developments

The calculated figures for the O&M costs showed a decline of 13% in the medium term and 23% in the long term. The breakdown of the O&M costs is shown in Fig. 8. The decline in the costs of solvent consumption is the most important factor in future reductions, causing 44% of the medium-term and 58% of the long-term cost reductions.

5.3. Cost of electricity and CO₂ avoidance costs

5.3.1. Results of the calculations

The results presented in the previous paragraphs can be included in the calculations of the COE and the CO₂ avoidance costs. The potential development of the COE including CO₂ capture is shown in Fig. 9. The figure shows a 13% decrease from 5.6 €/ct/kWh in the short term to 4.8 €/ct/kWh in the medium term and a 18% decrease to 4.5 €/ct/kWh in the long term. The reductions due to the energy-efficiency improvements of the NGCC power plant play an important role in these

figures. Omitting these developments yields a value of 5.3 €/ct/kWh for the medium term and 5.0 €/ct/kWh for the long term. Consequently, 60% of the reduction in the COE over time is caused by improvements in NGCC efficiency. To study the cost-reducing effects, it is also interesting to look at the incremental costs of electricity production and the CO₂ avoidance costs (Fig. 10). We calculated a potential decline in incremental COE from 1.49 €/ct/kWh to 1.03 €/ct/kWh in the medium term and 0.85 €/ct/kWh in the long term. Excluding the NGCC developments, the medium- and long-term values are 1.31 and 1.13 €/ct/kWh, respectively. Around two thirds of the reductions in the incremental COE and avoidance costs are caused by reductions in the energy penalty and are due to the relatively large initial share (44%) and the large reduction potential (48% in the medium term and 64% in the long term). The reduction potential of the capital costs and O&M costs are much lower, but still relevant. Capital costs of the capture plant may be reduced by 16% in the medium term and 22% in the long term and O&M costs by 18% in the medium term and 30% in the long term. For the incremental COE, 60% of the medium-term and 43% of the long-term reductions are caused by improvements in NGCC efficiency. This share is larger for the efficiency penalty (medium term: 70%; long term: 50%) than for the capital costs part (medium term: 44%; long term: 38%) and the O&M cost part (medium term: 37%; long term: 24%).

No flue gas recycling data were included in the results presented above. Based on data by the IEA (IEA GHG, 2000a), quantification of the possible impact of the application of a 50% flue gas recycling on the (incremental) capital costs, the COE and the avoidance costs was calculated as shown in Table 13.

5.3.2. Sensitivity analysis

The sensitivities of the COE and the avoidance costs to the number of operating hours, the natural gas price, the

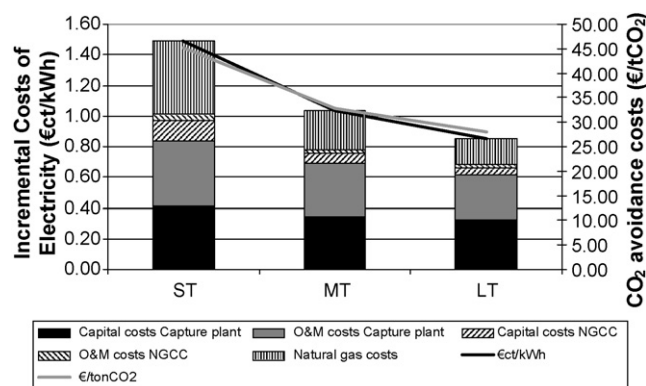


Fig. 10 – Incremental costs of electricity (COE) due to CO₂ capture. The COE are broken down into the costs due to the capital costs of the capture plant, the O&M costs of the capture plant, and the energy penalty. The energy penalty is broken down further into the incremental capital costs of the NGCC per kW net output, the incremental O&M costs of the NGCC per kW net output, and the additional natural gas costs. The avoidance costs in terms of €/tCO₂ avoided are also shown.

Table 13 – The possible impact of flue gas recycle on costs

	2010	2020	2030
Incremental capital costs (€/kWe)			
Baseline	352	266	237
Flue gas recycle	283	216	192
Total capital requirements (€/kWe)			
Baseline	882	746	687
Flue gas recycle	813	696	642
Costs of electricity (€/kWh)			
Baseline	5.6	4.8	4.5
Flue gas recycle	5.4	4.7	4.4
Avoidance costs (€/kg CO₂)			
Baseline	45	33	28
Flue gas recycle	39	29	25

Based on percentile reductions of incremental capital costs and energy penalty taken from IEA (IEA GHG, 2000a).

interest rate, and lifetime are shown for the short, medium, and long terms in Figs. 11 and 12. It should be noted that in these graphs, the parameters were changed in both the reference plant and the plant with capture. The interest rate and lifetime of the power plant and capture plant could be varied independently as well. Variations in these two parameters for the capture plant and the power plant separately with regard to the COE yielded less steep lines. Both the COE and the avoidance costs show a large sensitivity to the annual number of operating hours. The sensitivity to the gas price was much higher for the COE than for the avoidance costs due to the much larger share of NG costs in the COE (short term: 62%; medium term: 63%; long term: 64%) than in the avoidance costs (short term: 32%; medium term: 25%; long term: 21%). The sensitivities to the

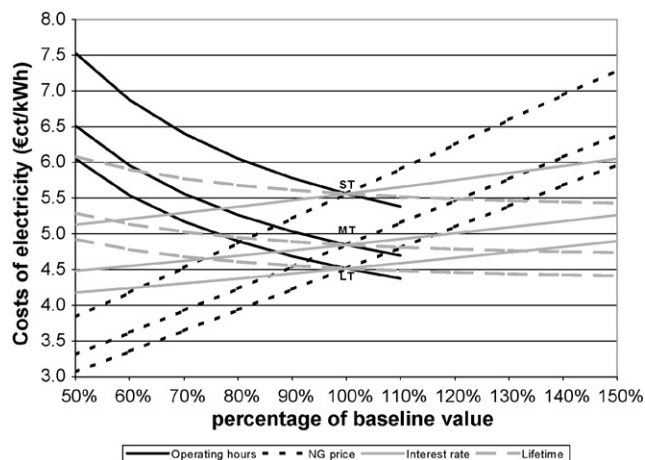


Fig. 11 – Sensitivity analysis of the costs of electricity (COE) including CO₂ capture. The sensitivity of the COE to variations in operating hours, natural gas price, interest rate, and lifetime is shown for the short, medium, and long terms. These parameters are varied for both the reference plant and the power plant including capture. The baseline values are 7600 operating hours per year, a natural gas price of 4.7 €/GJ, an interest rate of 10%, and a lifetime of 20 years.

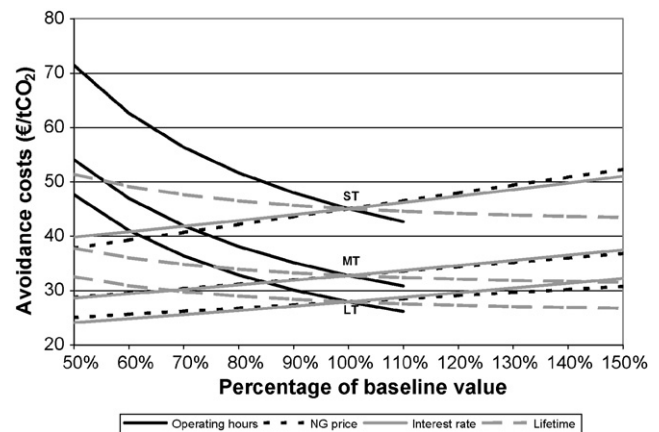


Fig. 12 – Sensitivity analysis of avoidance costs. The sensitivity of the avoidance costs to variations in operating hours, natural gas price, interest rate, and lifetime is shown for the short, medium, and long terms.

variations in lifetime and interest rate were smaller, but still relevant.

6. Discussion and conclusions

In this study, we thoroughly analyzed the current technical performance and costs of post-combustion CO₂ absorption from NGCC power plants as well as possible future developments. Due to the level of detail in our analysis, we were able to identify the level of importance of individual improvements in the potential future performance of the capture technology.

6.1. Development potential of NGCC with post-combustion absorption of CO₂

It can be concluded from this study that large improvements with respect to costs and energy efficiency can be achieved for post-combustion absorption technology at NGCC power plants (Table 14). A large reduction in CO₂ avoidance can be realized by a large decrease in the energy penalty and substantial reductions in the O&M costs and capital costs of the capture plant.

The improvement in NGCC efficiency plays a major role in the reduction of avoidance costs. The NGCC efficiency improvements affect O&M and capital costs, but have the largest impact on the energy penalty costs. The fact that technology is applied downstream of the NGCC principally allows all NGCC improvements to be implemented straight away. Compared with more integrated technologies, this might be the important advantage of this type of technology. The important contribution of the improved conversion efficiency of future NGCC power plants in the cost reduction of CO₂ capture can be an argument to postpone the implementation of capture technology. On the other hand, it should be noted that major improvements in the capture plant will be postponed as well.

The reductions in the energy penalty are primarily induced by solvent developments and process optimization. Solvent

Table 14 – Summary of the results of this study

	2010	2020	2030
COE (€/kWh)	5.6	4.8	4.5
Incremental COE (€/kWh)	1.5	1.0	0.9
Netto efficiency (%LHV)	49.0	55.1	58.3
Energy penalty (%point LHV)	7.9	4.9	3.7
Capital costs (€/kW)	882	746	687
Incremental capital costs (€/kW)	352	266	237
CO ₂ avoidance costs (€/tCO ₂)	45	33	28

developments can reduce the energy requirements needed to desorb CO₂ from the solvent. Solvents with higher effective solvent loadings can have large effects on the energy penalty as may solvents with lower desorption temperatures, which lower the evaporation energy losses and the power equivalent factor, because lower grade steam can be used. Furthermore, the optimization of desorption and absorption temperatures will have a reducing effect on the above-mentioned parameters. The reduction of the pressure drop in the absorber has a smaller, but still relevant impact on the energy penalty.

The solvent developments mentioned above also have important effects on the capital costs of the power plant. Solvents with higher absorption rates will decrease absorber size. Higher solvent concentrations will result in a smaller stripper size; a lower regeneration energy requirement will reduce the size of the reboiler needed; and higher solvent loadings will reduce the size of the amine pumps. The absorber cost reductions, the main contributor to the capital cost reduction potential, are predominantly caused by solvent developments and the substitution of random packing by structured packing. Reductions in compression energy are

induced by heat integration of the cooling process. Other important options for cost reductions (in order of importance) are the reboiler and reclaimer, the stripper, and the solvent pumps.

6.2. Limitations of this study

Because this analysis dealt with long time frames and hypothetical solvents, it was necessary to apply a number of simplifications in the calculations as described in Section 2.2.3. This implies a relatively independent treatment of the interrelated parameters, which caused an increased uncertainty in our results. Whether all of the parameters can meet the projected improvements at the same time is subject to debate. In particular, characteristics like chemical binding energy, absorption and desorption rates, and absorption and desorption temperature levels are solvent-specific. In our analysis, we chose for the optimal development in chemical binding energy and a suboptimal development in, for example, evaporation energy and power equivalent factor (illustrated in Fig. 6). This should minimize the risk of an overestimation of the improvement potential. Fig. 4 shows the maximum range of the results due to the uncertainty in the parameters. When assuming that, at future optimal improvements of the chemical binding energy, all other developments perform at the minimum level and the uncertainty of the compression energy is omitted, the maximum range of the relative energy penalty figures decreases by one third.

Because of the uncertainty, it is valuable to compare our results with values from the literature. For the entire regeneration energy, reference values for state-of-the-art technology and several future developments listed in the literature are presented in Table 15.

Table 15 – Regeneration energy of several solvents and processes from the literature and from this study

Source	Solvent/process	Status	Regeneration energy (MJ/kg CO ₂)
Gibbins et al. (2004)	MEA	State-of-the-art	4.00–4.25
Feron and ten Asbroek (2004)	MEA	State-of-the-art	4.20
Rao et al. (2004)		Present value	4.3 (range: 3.2–5.9)
This study		2010	4.4
Gibbins et al. (2004)	Econamine FG+	Near commercial	3.00–3.25
Morrisson (2004)	Econamine FG+	Near commercial	79% of state-of-the-art EFG (coal) ^{a,b}
IEA GHG (2004)	Econamine FG+	Near commercial	3.24
Mimura et al. (1995)	KS-1	Near commercial	85–90% of MEA/2.93 ^b
Mimura et al. (2001)	KS-1	Near commercial	2.96 (14% CO ₂), 3.54 (3.6% CO ₂)
Gibbins et al. (2004)	KS solvents	Near commercial	3.00–3.25
IEA GHG (2004)	KS solvents	Near commercial	2.80–3.20
Rao et al. (2004)		'Best guess' 2015	3.3 (range: 2.4–4.9)
This study		2020	2.6
Chakma (1995)	PSR	R&D phase	70–80% of MEA ^a
Chakma and Tontiwachwuthikul (1999)	PSR	R&D phase	55–85% of MEA ^a
Feron and ten Asbroek (2004)	DECAB	R&D phase	2.30
Cullinane et al. (2004)	PZ/K ₂ CO ₃	Lab-scale tests	70–80% of MEA ^a
Rao et al. (2004)		'Most optimistic' 2015	2.4 (range: 1.5–3.2)
This study		2030	1.6

^a No actual figures are given in these sources, only a percent decline compared with MEA-based processes. Because the reference value is unclear, the figures have a relatively large uncertainty range.

^b These sources describe absorption after coal-based electricity generation. Because these processes generate flue gas with higher partial CO₂ pressures, the energy requirements are lower. An example of this dependency is given by Mimura et al. (2001).

One interesting reference was the expert interview performed by Rao et al. (2004). The short-term value we calculated for the regeneration energy was 4.4 MJ/kg CO₂. This value is similar to the one given by the experts in Rao et al.'s (2004) report and the values for the state-of-the-art processes shown in Table 15. Our medium-term value (2.6 MJ/kg CO₂) lies between the 'best guess' and 'most optimistic' values for 2015 given in Rao et al. (2004). It also lies in the range between the 'near commercial' and 'R&D phase' processes shown in Table 15. Our long-term value (1.6 MJ/kg CO₂) is within the most optimistic 2015 range presented by Rao et al. (2004). Values for 25 years in the future were not found in literature. Based on the match between our calculated values and the reference values, we can conclude that the coupling of the parameters in our model generated realistic results.

The avoidance costs listed in the IEA report (IEA GHG, 2004), normalized to the gas price, discount rate, and lifetime we used, are comparable to the results found in the present study. The normalized IEA value for the short term was 46 €/tCO₂ avoided (see Damen et al. (2006)) and is comparable to our result of 45 €/tCO₂. The 2020 figure of 31 €/tCO₂ avoided was a little lower than our 33 €/tCO₂.

6.3. Comparison to other CCS technologies

In order to make a detailed comparison between post-combustion CO₂ absorption from NGCC power plants and other CCS technologies, the same type of analysis should be performed. By referring to proper values of other promising CCS technologies, however, we were able to put the data from our analysis in the context of the other technologies.

Damen et al. (2006) reviewed short- and long-term CCS technologies, generating standardized values for COE and avoidance costs. The same assumptions were used in that study for the natural gas prices, discount rate, lifetime, and final CO₂ pressure as in our study. Therefore, our results can be compared to these reference values. It should be noted that the figures presented by Damen et al. (2006) are not explicitly linked to a certain time frame. Moreover, the different technologies had different capture efficiencies than assumed in our study, namely 85% (Advanced Zero Emission Power Plant; AZEP & Chemical Looping Combustion; CLC) and 100% (Membrane Reforming; MR).

Table 16 compares the net efficiency and avoidance costs for AZEP, CLC, and MR. Note that a NGCC power plant with

Table 16 – Comparison of the conversion efficiency and CO₂ avoidance figures with other CO₂ capture technologies

	Conversion efficiency including CO ₂ capture (%LHV)	CO ₂ avoidance costs (€/tCO ₂)
This study 2010	49 (range: 47–52)	45
This study 2020	55 (range: 52–58)	33
This study 2030	58 (range: 54–62)	28
AZEP	53	29
CLC	56	36
MR	53	39

The data for the advanced zero emission power plant (AZEP), chemical looping combustion (CLC), and membrane reforming (MR) are based on Damen et al. (2006).

state-of-the-art post-combustion technology has a lower conversion efficiency than the advanced plants of the near future. Post-combustion capture may score comparable efficiencies in the medium term and higher conversion efficiencies in the long term. Similar figures are shown for the avoidance costs. It should be noted that there is a large uncertainty with regard to the capture cost figures of the advanced options. We used our medium-term figures on the NGCC as a baseline for the calculation of avoidance costs for the other options. With regard to the efficiency figures, we can conclude that post-combustion CO₂ capture may move towards competing values with the advanced technologies in the medium (CLC and MR) and long (AZEP) term.

The comparison of our results with coal-based electricity production with CCS requires an assumption for the coal price as well as a choice in reference system. A comparison with the data for post-combustion capture from pulverized coal (PC) power plants is shown in Table 17. Because of the high CO₂ concentration in PC plants, the avoidance costs per tonne of CO₂ were 25–40% lower. The impact on cycle efficiency and COE, however, was larger, i.e., a 6–15% higher COE for PC plants with post-combustion capture than for NGCC plants with post-combustion capture, using the assumed fuel prices.

6.4. Recommendations for further research

A number of potentially cost-declining developments were not taken into account in this study. These developments

Table 17 – Comparison of the key results of this study with results on post-combustion capture after a pulverized (PC) coal power plant

	NGCC with post-combustion capture (this study)			PC with post-combustion capture (Damen et al.)	
	Short term	Medium term	Long term	PC	Advanced PC
COE (€/kWh)	5.6	4.8	4.5	6.4	5.1
Net efficiency (%LHV)	49	55	58	35	40
Avoidance costs: NGCC/PC (€/tCO ₂)	45	33	28	27	21
Avoidance costs: NGCC (€/tCO ₂)	45	33	28	94	38

Taken from Damen et al. (2006). The numbers are based on a coal price of 1.7 €/GJ. The avoidance costs were calculated against equal technology without capture (NGCC/PC) and against an NGCC.

should be included in future research. Suggestions for further research are:

1. *Split-flow configurations*: This option lowers regeneration energy by introducing a semi-lean amine stream. The higher complexity of the system, however, is a negative side effect. Aroonwilas (2004) claimed a possible 30% decrease in the regeneration energy of an MEA-based process using such configurations. These improvements can be achieved at the expense of additional capital and O&M costs and parasitic power loss, items that were not further quantified in the study. This improvement option is, however, less attractive for future hypothetical solvents with lower regeneration energy.
2. *Introduction of membrane gas absorption*: This technique will decrease the size of the absorber column and the pressure drop in the absorber. Cheaper membranes are required to make this option viable.
3. *Precipitating solvents*: Such solvents, as proposed by TNO, will require new process layouts. They may have the potential to substantially lower the regeneration energy.
4. *Flue gas recycle*: This recycle may have a large impact on the performance of post-combustion absorption as shown in Tables 8 and 13.
5. *Technology comparison*: For a detailed comparison between post-combustion chemical CO₂ absorption in combination with an NGCC power plant and other promising CO₂ capture technologies, the same type of analysis should be performed.

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